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Muscle Shoals, AL 35660

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Technical Review of Dry FGD Systems and Economic Evaluation of Spray Dryer FGD Systems

Interagency Energy/Environment R&D Program Report

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Technical Review of Dry FGD Systems and Economic Evaluation of Spray Dryer FGD Systems

by

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ABSTRACT

The report gives results of an extensive study of dry flue gas desulfurization (FGD) systems, involving dry injection of absorbents or spray drying. (The study was undertaken because they appear to have both process and economic advantages over wet FGD.) Design concepts (e.g., type of absorbent and atomizer, approach to flue gas saturation temperature, and particulate collection method) remain to be demonstrated at full scale. Most vendors prefer a lime slurry system with rotary atomizers and fabric filter particulate collection, while all systems now under contract to utilities apply to low-sulfur coal. SO_2 removal efficiencies sufficient for high-sulfur coal applications at stable operating conditions and economically feasible absorbent utilization rates have not yet been demonstrated. In conceptual design cost comparisons based on a new 500-MW utility power generation unit, a lime spray dryer/fabric filter combination had lower capital investments and annual revenue requirements for 0.7% sulfur western coal and both 0.7% and 3.5% sulfur eastern coal than a wet limestone scrubbing process. With lignite fuel, similar cost advantages were evident for dry (relative to wet) FGD. The capital investment advantage of dry over wet FGD increased with increasing coal sulfur content.

CONTENTS

FIGURES

FIGURES (continued)

TABLES

TABLES (continued)

TABLES (continued)

TABLES (continued)

ABBREVIATIONS AND CONVERSION FACTORS

ABBREVIATIONS

aft^3	actual cubic feet
Btu	British thermal unit
°C	degrees Celsius
dia	diameter
ESP	electrostatic precipitator
°F	degrees Fahrenheit
FD	forced draft
FGD	flue gas desulfurization
ft	feet
ft/sec	feet per second
g	gram
gal	gallon
gpm	gallons per minute
gr	grain
hr	hour
ID	induced draft
in.	inch
kg	kilogram
kℓ	kiloliter
kW	kilowatt
kWh	kilowatthour
lb	pound
k	thousand (kilo)
L/G	liquid to gas ratio
M	million (mega)
min	minute
mol	mole
MW	megawatt (electrical)
ppm	parts per million (volume)
sec	second
sft^3	standard cubic feet
SIP	State Implementation Plan
SO_x	sulfur oxides
vol	volume
wt	weight
yr	year

CONVERSION FACTORS

To convert from English units	To metric units	Multiply by
acres	hectares	0.405
British thermal units	kilocalories	0.252
degrees Fahrenheit minus 32	degrees Celsius	0.5556
feet	centimeters	30.48
square feet	square meters	0.0929
cubic feet	cubic meters	0.02832
cubic feet per minute	cubic meters per second	0.000472
gallons (U.S.)	liters	3.785
gallons per minute	liters per second	0.06308
grains per cubic foot	grams per cubic meter	2.288
horsepower	kilowatts	0.746
inches	centimeters	2.54
pounds (mass)	kilograms	0.4536
pounds per cubic foot	kilograms per cubic meter	16.02
pounds (force) per square inch	Pascals (Newton per square meter)	6895
miles	meters	1609
standard cubic feet per minute (60°F)	normal cubic meters per hour (0°C)	1.6077
tons (short)[a]	metric tons	0.9072

a. All tons, including tons of sulfur, are expressed in short tons.

ACKNOWLEDGEMENTS

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TECHNICAL REVIEW OF DRY FGD SYSTEMS AND ECONOMIC EVALUATION OF SPRAY DRYER FGD SYSTEMS

EXECUTIVE SUMMARY

INTRODUCTION

Dry scrubbing technology for flue gas desulfurization (FGD), particularly that phase in which spray dryers are used, is currently receiving considerable attention in the utility industry. An alkaline solution or slurry is atomized in the flue gas and evaporates to dryness while reacting with the SO_x and HCl. The resulting reaction products are collected, usually along with fly ash, and disposed of as a dry waste. This method has several potential advantages over wet scrubbing FGD because it eliminates the complexity, operating problems, and liquid wastes associated with the large volume of scrubbing liquid used in wet scrubbing. Conversely, high removal efficiencies are more difficult to attain and a highly reactive (i.e., nonlimestone) absorbent must be used.

In the past few years a number of companies and consortia have entered the spray dryer FGD field with pilot studies, and several have contracted to build commercial units. Most of the pilot studies and all of the utility units are for lignite or low-sulfur western coal applications, where removal efficiencies and absorbent consumption are usually lower than with high-sulfur coals and, in some cases, the high alkalinity of the fly ash can supplement the absorbent. The rapid growth of spray dryer FGD is in part a result of the increasing use of western coal, but its rapid growth also owes, in some degree, to its derivation from the proven industrial technologies of spray drying and particulate matter collection. The development of fabric filter fly ash collection for utility use has been particularly advantageous. In many cases, companies active in spray dryer FGD have backgrounds in either particulate collection or spray drying.

Spray dryer FGD is related, and in some aspects evolved from, earlier efforts in dry injection of absorbents. Although many of these studies were disappointing in terms of SO_2 removal efficiency, absorbent utilization, and availability of economical absorbents, interest in such uncomplicated approaches to FGD has continued. The development of spray dryer FGD, its general technical aspects, and its status through mid-1980 are discussed as a portion of this study. In addition, the history and current status of dry injection processes are reviewed.

An important aspect of spray dryer FGD is its economics in relation to limestone wet scrubbing processes, in which the absorbent is less expensive. Although various vendors have made economic comparisons, there have been no independent economic comparisons applicable to general utility applications. As a portion of this study, economic comparisons of a lime spray dryer process and a limestone scrubbing process are made for a lignite, a low-sulfur western coal, and a low- and a high-sulfur eastern coal applications. A soda ash spray dryer process is also evaluated for the low-sulfur western coal case.

BACKGROUND

Dry absorption of SO_2 received considerable attention during the early 1970's because it appeared to have several technical advantages over wet scrubbing. Of the potential absorbents, only sodium-based materials were found to be sufficiently reactive and, of these, nahcolite proved the most effective. However, due to economic and environmental considerations, commercial mining of the large nahcolite reserves did not (and has not) occurred, and questions about the widespread availability of nahcolite in the future forced a search for other absorbents. Because other more readily available raw materials are either too expensive or not sufficiently reactive, development of dry absorption FGD slowed in the mid-1970's, and primary emphasis focused on the technology to make the readily available reactants more reactive without losing the potential significant advantages of the dry FGD system. This search (along with the simultaneous development of the regenerable aqueous carbonate process) led to the development of the spray dryer FGD process.

Recently interest in dry absorption has resurfaced with several pilot-plant programs. For absorbents, nahcolite is still the primary focus but trona is also being evaluated. Trona, unlike nahcolite, has the advantage that it is already being mined in commercial quantities for the production of soda ash. Although early results appear promising, at least for applications where only 70% SO_2 removal is required, significant amounts of development work remain to be completed.

Most spray dryer designs for FGD are direct adaptations of the standard designs so widely used in other industries. Typically in these uses, a hot gas passes downward through a cylindrical vessel, mixing with a solution or slurry atomized by either rotary or nozzle atomizers. The liquid is evaporated while the droplets are in suspension, and the particles are collected in the conical bottom, in external collection equipment, or both. Complete evaporation in suspension is important and is achieved by suitable design and control of operating conditions. In FGD applications, the latitude of these controls is limited. The flue gas temperature is fixed by boiler efficiency requirements and may vary as the boiler load changes and the absorbent rate is controlled by SO_2 removal requirements. In addition, it is economically important that absorbent consumption is minimized. These limitations may complicate applications in which high SO_2 removal efficiencies are required.

The reactions of SO_2 and HCl with the absorbent proceed rapidly while surface liquid is present, but more slowly when the absorbent is dry. An important design consideration is, therefore, that the saturation temperature be approached as closely as possible and that the particles remain in contact with the flue gas as long as possible. Whatever these conditions, however, effective reaction rates require a reactive absorbent. Limestone has not proven satisfactory, leaving soda ash and lime as the only economically practical absorbents generally available in sufficient quantities. Soda ash is more reactive but also more expensive, and the soluble waste of sodium sulfites and sulfates produced has raised questions of its practicality for disposal in areas of high rainfall. Lime is less reactive and more difficult to handle because it must be slaked and then handled as a slurry. It is, however, less expensive and it produces a relatively insoluble waste of calcium sulfites and sulfates. At higher SO_2 removal efficiencies the absorbent must sometimes be used at high stoichiometric ratios, particularly if lime is used. Utilization can sometimes be increased by reslurrying and recycling the waste. Also, if a highly alkaline fly ash is produced, as is usual with western coals, the fly ash alkalinity can supplement the absorbent. Coal moisture content is also a factor. Flue gas produced by low-rank, high-moisture coals has a higher saturation temperature, limiting the amount of water that can be added, compared with high-rank coals. This may place restrictions on absorbent liquid concentrations that affect the SO_2 removal efficiency.

The methods by which vendors treat these considerations differ. In most cases, a conventional spray dryer design, rotary atomizers, lime absorbent, and fabric filter baghouse particulate collection are used. The approach to saturation is regulated by controlling water addition rates. Some warm (300°F) flue gas from the air heater may be bypassed around the spray dryers and recombined with the cleaned gas for reheating before the flue gas enters the baghouse. In extreme cases for high SO_2 removal efficiencies some hot (700°F) flue gas may be bypassed around both the air heater (at the expense of boiler thermal efficiency) and the spray dryers to attain sufficient reheating.

Important exceptions to the above design exist. One major vendor uses two-fluid nozzles and ESP collection and one commercial unit will use soda ash. Absorbent utilization is increased in some cases by recycling the waste, while in other cases it is not. The degree to which saturation temperature is approached (and thus the possibility of wet upsets) varies somewhat between vendors. In general, the vendor's approach to design reflects his experience and the requirements of the particular application. The technical and economic advantages of such design variations as type of atomizer, waste recycle, and ESP or baghouse collection remain, in large part, to be demonstrated in further investigation and application.

In general, based on current practice and trends, soda ash processes will not use waste recycle, will rarely use hot (700°F) gas bypass, and will use warm gas bypass (bypass around the spray dryer) only for very high SO_2 removal efficiencies under unusual flue gas conditions. Lime processes will use waste recycle, will usually use warm gas bypass, and will occasionally use hot gas bypass (particularly when high (over 85%) SO_2 removal efficiencies are required).

DEVELOPMENT AND CURRENT STATUS

The first concerted spray dryer FGD studies in the United States were begun in the early 1970's by Rockwell International. These, however, were part of a regenerable process study rather than the waste-producing nonregenerable processes that are the subject of this study. It was not until 1977 that extensive spray dryer FGD studies began. These received considerable impetus when dry injection studies at the Basin Electric Power Cooperative's Leland Olds Station were expanded to include spray dryers. Four companies subsequently operated spray dryer FGD pilot plants there and at other Basin Electric power plants as a bid qualification requirement for FGD units on new Basin Electric construction.

Subsequently, other organizations became active in spray dryer FGD studies. In mid-1980 ten companies or consortia (shown in Table S-1) were active in spray dryer FGD investigations and six had contracted for units for ten utility and five industrial installations. These contract awards are shown in Table S-2. All of the commercial utility applications are for low-sulfur lignite or western coal, as has been the preponderance of pilot-scale studies (shown in Table S-3).

TABLE S-1. COMPANIES ACTIVE IN SPRAY DRYER-BASED FGD SYSTEMS FOR UTILITY APPLICATIONS

Babcock & Wilcox
Buell-Envirotech/Anhydro, Inc.
Carborundum Environmental Systems
Combustion Engineering, Inc.
Ecolaire Environmental Corporation
Flakt, Inc.
Joy Manufacturing/Niro Atomizer, Inc.
Research-Cottrell
Rockwell International
Wheelabrator-Frye, Inc.

Of the contracted utility units, all will use lime as the alkali raw material except the Coyote Station in Beulah, North Dakota. All vendors except Babcock & Wilcox (B&W) use atomizers in spray dryers of

Table S-2. CONTRACT AWARDS FOR SPRAY DRYER-BASED FGD SYSTEMS

Installation	Size, gross MW	Fuel type (% S)	SO_2 removal, %	Alkali raw material	Startup date	Vendor
Utility Boiler						
Coyote Unit 1	456	Lignite (0.78)	70	Soda ash	4/81	RI/WF[a]
Laramie River Unit 3	575	Subbituminous (0.54)	85	Lime	4/82	B&W[c]
Antelope Valley Unit 1	440	Lignite (0.68)	62	Lime	4/82	Joy/Niro[b]
Shiras Unit 3	44	Subbituminous (1.5)	80	Lime	9/82	Buell/Anhydro
Stanton Unit 2	63	Lignite (0.77)	73	Lime	9/82	R-C[d]
Craig Unit 3	447	Bituminous (0.70)	87	Lime	4/83	B&W
Holcomb Unit 1	319	Subbituminous (0.30)	80	Lime	6/83	Joy/Niro
Rawhide Unit 1	260	Subbituminous (0.29)	80	Lime	12/83	Joy/Niro
Springerville Unit 1	350	Subbituminous (0.69)	61	Lime	2/85	Joy/Niro
Springerville Unit 2	350	Subbituminous (0.69)	61	Lime	9/86	Joy/Niro
Industrial Boiler						
Strathmore Paper Co.	14[e]	Bituminous (2.0-2.5)	75	Lime	7/79	Mikropul
Celanese Fibers Co.	22[e]	Bituminous (1.0-3.5)	70-80	Lime	1/80	RI/WF
Calgon	17[e]	-	75	Soda ash	6/81	Niro/Joy
University of Minnesota	83[e]	Subbituminous (0.6-0.7)	70	Lime	9/81	Carborundum
Argonne National Lab.	29[e]	Bituminous (3.5)	80	Lime	9/81	Niro/Joy

Based on contact with vendors representing the status of announced contracts through October 1980.

a. Rockwell International/Wheelabrator-Frye.

b. Western Precipitation Division of Joy Manufacturing Company/Niro Atomizer, Inc.

c. Babcock & Wilcox.

d. Research-Cottrell.

e. Based on 2,900 aft^3/MW.

TABLE S-3. SPRAY DRYER PILOT PLANTS AND DEMONSTRATION UNITS FOR FGD

Company	Size, kaft^3/min	Fuel type (% S)	Primary alkali tested
Babcock & Wilcox			
Alliance Research Center	1.5	Various coals	Lime
W. J. Neal Station Unit 2 (Basin Electric)	8.0	Lignite (0.4)	Lime
Jim Bridger Unit 3 (Pacific Power and Light)	120.0	Subbituminous (0.6)	Lime
Buell-Envirotech/Anhydro, Inc.			
Copenhagen Anhydro Laboratory	3.0	-[a]	Lime, soda ash
Martin Drake Unit 6 (City of Colorado Springs)	20.0	Subbituminous (0.5)	Lime, trona
Carborundum Environmental Systems			
Carborundum Knoxville Laboratory	0.1	-[a]	Lime, NH_3, $NaHCO_3$, and nahcolite
Carborundum Knoxville Laboratory	1.0	Bituminous (0.5)	Lime, Na_2CO_3, and fly ash
Leland Olds Station Unit 1 (Basin Electric)	15.0	Lignite (0.6)	Lime, NH_3, and soda ash

(continued)

TABLE S-3 (continued)

Company	Size, kaft^3/min	Fuel type (% S)	Primary alkali tested
Combustion Engineering			
Sherburne County Unit 1 (Northern States Power)	20.0	Subbituminous (1.0)	Lime
Gadsden Unit 1 (Alabama Power)	20.0	Bituminous (1.8)	Lime
Ecolaire Environmental Corporation			
Gerald Gentleman Unit 1[b] (Nebraska Public Power)	10.0	Subbituminous (0.3)	Lime
Joy Manufacturing/Niro Atomizer, Inc.			
Niro Laboratory Copenhagen	3.0	–[a]	Lime, MgO, and soda ash
Hoot Lake Unit 2 (Ottertail Power)	20.0	Lignite	Lime, soda ash
Riverside Station Units 6 & 7 (Northern States Power)	640.0	Subbituminous (1.0) Petroleum coke (4.0)	Lime

(continued)

TABLE S-3 (continued)

Company	Size, kaft^3/min	Fuel type (% S)	Primary alkali tested
<u>Research-Cottrell, Inc.</u>			
Big Brown Unit 2 (Texas Utilities)	10.0	Lignite (1.0)	Lime
Comanche Unit 2 (Public Service of Colorado)	10.0	Subbituminous (0.5)	Lime
<u>Rockwell International/Wheelabrator-Frye, Inc.</u>			
Stork-Bowen Engineering Laboratory	5.0	–[a]	Lime, soda ash
Leland Olds Station (Basin Electric)	10.0	Lignite (0.6)	Lime
Joliet Station (Commonwealth Edison)	5.0	Subbituminous (0.5)	Lime
Sherburne County Unit 3[b] (Northern States Power)	5.0	Subbituminous (0.8)	Lime
Jim Bridger[b] (Pacific Power and Light)	5.0	Subbituminous (0.6)	Lime

Based on contacts with vendors and representing information through June 1980.
a. Propane burner with SO_2 spiking.
b. Mobile unit.

conventional design and fabric filter baghouse collection. B&W uses two-fluid nozzles, evolved from boiler oil burners, in horizontal chambers and prefers ESP's for particulate collection. B&W manufactures all of its FGD equipment. The other vendors are either consortia which include a spray dryer manufacturer or they have an exclusive agreement with a spray dryer manufacturer.

B&W began spray dryer FGD studies in 1977, initially with a commercial spray dryer/reactor and subsequently with their own two-fluid nozzle atomizer and horizontal reactor design they call a dry scrubbing reactor. Steam was first used as the atomizing fluid; more recently air has been favored. B&W uses ESP's for particulate collection, believing them to be a more developed technology and more amenable to wet upset. At first B&W did not generally favor waste recycle, however the results of their demonstration unit at Jim Bridger Station have pursuaded them that waste recycle should be used.

Close approach to the flue gas saturation temperature in the absorber and particle size reduction to attain efficient absorbent utilization are also emphasized in their design. In addition to their two pilot units, B&W is conducting continuing studies at their Alliance, Ohio, research center. B&W has been awarded contracts for two utility applications.

The Buell Emission Control Division of Envirotech Corporation and Anhydro, Inc., of Copenhagen, Denmark, are currently developing a spray dryer FGD system as a joint venture. Buell is a designer and marketer of particulate control equipment while Anhydro is a designer and marketer of spray dryers. The pilot unit at the Martin Drake Station uses a single rotary atomizer and baghouse particulate collection. In addition, proprietary waste recycle systems are being evaluated. Buell/Anhydro has been awarded the contract for one utility system.

Carborundum Environmental Systems is a subsidiary of Kennecott Copper Corporation based in Knoxville, Tennessee. Carborundum has recently signed a licensing agreement with Kochiwa Kakohki Company, Inc., a Japanese spray dryer manufacturer. The spray dryers for the Carborundum system will be manufactured in the United States, while the rotary atomizers may be manufactured in either Japan or the United States. Baghouses for the spray dryer FGD system will be designed and built by Carborundum. Much of the development work for Carborundum's spray dryer FGD system was done at a 100 ft^3/min bench-scale unit at their test facility at the University of Tennessee in Knoxville. The initial pilot studies were made to qualify for bids on Basin Electric units. The present design uses a conventional spray dryer with three rotary atomizers and baghouse particulate collection with no waste recycle. Carborundum has been awarded a contract for an industrial boiler system.

While Combustion Engineering has been installing limestone-based FGD systems for several years, they first entered into the development of spray dryer FGD systems in 1978. Construction of their first pilot-plant unit began in February 1979. During 1979 an exclusive agreement was concluded with James Howden Holida BV (The Netherlands) for use of their baghouse technology. The current design consists of a conventional spray dryer with multiple nozzle atomizers and baghouse particulate collection. Compressed air is the atomizing fluid. In addition to their pilot plant at Gadsden Station, a 30-MW demonstration unit is currently planned.

Ecolaire Environmental Corporation is the subsidiary of Ecolaire Corporation that markets Ecolaire's spray dryer FGD process. Although other subsidiaries in the Ecolaire Corporation have been supplying equipment to the electric utility industry for many years, before the design and construction of their mobile demonstration unit (MDU) in 1979, Ecolaire had very little experience in the design and operation of FGD systems. The MDU was erected in 1979 at a Nebraska power plant. The unit has a conventional spray dryer using either a rotary or two-fluid nozzle atomizer and fabric filters for particulate collection.

The Western Precipitation Division of The Joy Manufacturing Company, that markets fabric filter baghouses, and Niro Atomizer, Inc., that markets spray dryers, have an exclusive agreement to market a spray dryer FGD system. Niro began FGD studies in Denmark in 1975. The first pilot plant was operated in 1978 to qualify to bid for Basin Electric units. The Joy/Niro design consists of a spray dryer with one rotary atomizer and a manifold that introduces flue gas both above and below the atomizer. The particulate matter is collected both in the bottom of the spray dryer and in baghouses. Waste recycle, using the large particles from the bottom of the spray dryer, is used for most applications. Current development work is being conducted in Niro's Copenhagen laboratory. The demonstration unit at the Riverside Power Station will provide the facilities for future (at least until 1983 when the agreement expires) large-scale testing. Joy/Niro has been awarded five utility contracts and two industrial contracts.

The Research-Cottrell system uses Komeline-Sanderson spray dryers of conventional design with either a single or multiple rotary atomizers. Part of the particulate matter is collected in the bottom of the spray dryer and the rest is collected in the baghouse. Waste recycle is usually used. Most details and test results are proprietary and little published information is available. The pilot plant at the Comanche Station is partially funded by EPA, so results of these tests will be available in 1981. Research-Cottrell has been awarded one contract for a utility boiler.

Until early 1980 Rockwell International and Wheelabrator-Frye, Inc., had agreements to market spray dryer FGD systems based on Rockwell International's experience in spray dryer FGD and Wheelabrator-Frye's fabric filter technology. This joint venture was dissolved in 1980 and each will now market its own system. The first Rockwell International and Wheelabrator-Frye spray dryer pilot unit was operated at the Leland Olds Station in 1977. This and other pilot studies have provided considerable data on the system. The design consists of a conventional spray dryer with three rotary atomizers and baghouse particulate collection. Waste recycle is used if the conditions warrant it. Rockwell International and Wheelabrator-Frye have been awarded two commercial contracts, one for a utility system and one for an industrial application.

DESIGN AND ECONOMIC PREMISES

The economic evaluations are based on flue gas cleaning (FGD and fly ash) systems to meet the 1979 NSPS for a new 500-MW pulverized-coal-fired utility boiler. The FGD systems are designed with one redundant scrubber train, a redundant feed preparation area, 50% emergency flue gas bypass, and are costed as proven technology with no adjustments for estimated stage of development. The power plant is assumed to have a 30-year, 165,000-hour life and to operate 5,500 hours the first year. Flue gas compositions are based on a 0.9% sulfur, 7.2% ash, 6,600 Btu/lb lignite; a 0.7% sulfur, 9.7% ash, 9,700 Btu/lb western coal; a 0.7% sulfur, 16% ash, 10,700 Btu/lb eastern coal; and a 3.5% sulfur, 16% ash, 10,700 Btu/lb eastern coal. A Northern Great Plains location is used for the lignite and the western coal cases; a midwestern location is used for the eastern coal cases. The spray dryer designs are generic and are based on vendor information. The limestone scrubbing system is based on data from the EPA Shawnee test facility and general industry information. Design data for the absorbers are shown in Table S-4.

Raw materials consist of commercial-grade soda ash at $145/ton, pebble lime at $102/ton in the West and $75/ton in the East, and limestone at $8.50/ton (all costs are in 1984 dollars). The waste disposal sites are one mile from the FGD facility. They consist of a clay-lined pond for the soda ash spray dryer process and landfills for the lime spray dryer and limestone scrubbing wastes.

The economics consist of study-grade capital investments, first-year annual revenue requirements, and levelized annual revenue requirements. The capital investments are based on major equipment costs developed from flow diagrams and material balances and factored costs for installation and ancillary equipment. These capital investments are estimated to have an absolute accuracy of -20% to +40%. However since the same estimation methods were used for each evaluation, the accuracy for comparison is much better, i.e., ±10%. Capital investment costs are based on mid-1982 costs.

First-year annual revenue requirements consist of raw material, operating, and overhead costs and levelized capital charges. Revenue requirements are based on 1984 costs. The levelized annual revenue requirements are factored to account for a 10% discount and a 6% inflation rate over the life of the power plant.

TABLE S-4. FGD SYSTEM DESIGN CONDITIONS

	Lignite		Low-sulfur western coal			Low-sulfur eastern coal		High-sulfur eastern coal	
	Lime spray dryer	Limestone scrubbing	Soda ash spray dryer	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing
Absorbent stoichiometry[a]	1.2	1.1	1.0	1.2	1.1	1.3	1.1	1.8	1.3
Bypass, %	22.5	28.1	0	22	28.1	19	25.2	4[b]	0
Total FGD ΔP, in. H_2O	12	8.6	12	12	8.4	12	8.5	12	9.5
Absorber									
Removal efficiency, %	83.5	90	70	83	90	83	90	89	90
Absorbent liquid, % solids[c]	22.5	60	0	22.5	60	3	60	17	60
ΔP, in. H_2O	2	2	2	2	2	2	2	2	2
L/G, gal/kaft^{3b}	0.3	80	0.13	0.2	80	0.3	80	0.3	90
Exit gas, wt % liquid	0	0.1	0	0	0.1	0	0.1	0	0.1
Effluent, % solids	100	15	100	100	15	100	15	100	15
Recombined gas, $^{\circ}F$	170	180	170	170	170	170	160	170	127
Reheat, $^{\circ}F$	-	-	-	-	-	-	10	-	43

a. Defined as mol Ca/mol SO_2 absorbed for both the spray dryer and the limestone processes.
b. Hot gas bypass.
c. Excludes dilution water and recycle loop if used.

SYSTEMS ESTIMATED

For the lignite case a lime spray dryer process and a limestone scrubbing process are evaluated. For the low-sulfur western coal case, the soda ash and lime spray dryer processes and the limestone scrubbing process are evaluated. For the eastern coal cases, only the lime spray dryer and limestone scrubbing processes are evaluated because of the economically indefinable problems likely to occur with soluble sodium wastes in high rainfall regions. Process designs include all the equipment and material needed to transfer the flue gas from the boiler to the stack plenum. All requirements for both fly ash collection and disposal and SO_2 removal and disposal are included in the costs.

The soda ash and lime spray dryer processes consist of four or five trains of cylindrical spray dryers, each with three rotary atomizers. One train is included as a nonoperating spare in all cases. Emergency hot gas (700°F) bypass of flue gas around the air heater and spray dryers is provided for the low-sulfur coal cases although under normal operating conditions it will not be needed. A continuous 4% hot gas bypass is used in the lime spray dryer process for the high-sulfur eastern coal case. Warm gas (300°F) bypass (i.e., around only the spray dryer) rates of 22.5%, 22%, and 19% are used in the lime spray dryer process for the lignite, the low-sulfur western, and the low-sulfur eastern coal cases respectively. No flue gas bypass is used for the soda ash spray dryer process under normal operating conditions. The higher reactivity of the soda ash (versus lime) does not require a close approach to the saturation temperature and flue gas bypass is not necessary to ensure dry conditions in the baghouse. In all cases the processes are designed for a flue gas stack temperature of 175°F (including the 5°F to 10°F provided by adiabatic compression in the ID fan). A single baghouse is used for particulate collection. The waste is pneumatically conveyed to storage silos and trucked to a disposal site one mile away. An earthen-diked, clay-lined pond is used for the soda ash spray dryer process and a landfill is used for the lime spray dryer process. For the lime spray dryer process in the lignite, the low-sulfur western coal case, and the high-sulfur eastern coal case, waste recycle is used to reduce absorbent consumption. Waste is not recycled in the other spray dryer processes.

The limestone scrubbing process consists of four or five trains of spray tower absorbers, one of which is a spare, using a 15% solids slurry of pulverized limestone. The slurry is oxidized by sparging air into the recirculation tank to produce $CaSO_4 \cdot 2H_2O$. A purge stream is dewatered by thickening and filtering to 80% solids. The waste is trucked one mile to a landfill and disposed of with fly ash collected in ESP's upstream of the FGD system. For the lignite and low-sulfur western and eastern coal cases, 28%, 28%, and 25% warm gas bypass is used, respectively, and the scrubbing efficiency is 90% to obtain an overall 70% SO_2 reduction. Full scrubbing at an 89% SO_2 reduction efficiency is used for the high-sulfur eastern coal case. The flue gas bypass eliminates the need of flue gas reheat in the lignite and the low-sulfur western coal case and substantially reduces it in the low-sulfur eastern coal case. Full reheat is used in the high-sulfur eastern coal case. In all cases a flue gas stack temperature of 175°F is used.

ECONOMIC EVALUATION

Capital investments, first-year annual revenue requirements, and levelized annual revenue requirements were developed based on the processes described in the systems estimated section and the conditions described in the premises.

Capital Investment

The capital investments for the soda ash and lime spray dryer processes and the limestone scrubbing process are shown in Table S-5.

In comparing the soda ash and lime spray dryer processes for the low-sulfur western coal case, the overall capital investments of $158/kW and $154/kW, respectively, differ only slightly. The largest cost area, particulate collection, is the same for both. Similarly, the total of the gas handling and SO_2 absorption areas differ little, suggesting little cost difference between partial bypass and full scrubbing. Material handling, feed preparation, and particulate handling costs are lower for the soda ash spray dryer process, but this is more than offset by the higher disposal site construction and land costs because a pond is used for the sodium wastes. Waste recycle in the lime process approximately doubles the particulate handling costs, but costs in this area are small compared with other cost areas. The comparisons suggest that neither bypass nor waste recycle is an important capital investment consideration for low-sulfur coal applications.

More significant differences emerge in the comparisons of the lime spray dryer and limestone processes. In overall capital investment the limestone scrubbing process is 12% to 23% higher than the lime spray dryer process, the difference being greatest for the lignite case and least for the low-sulfur western coal case. The major cost area for the limestone scrubbing process is SO_2 absorption, representing nearly one-third of the direct costs. In addition, this cost increases about 44% as the coal sulfur content increases from 0.7% to 3.5%. In contrast, the SO_2 absorption area costs for the lime spray dryer process are about one-half those of the limestone scrubbing process for the low-sulfur coals, and they increase only 23% in going to the 3.5% sulfur case. These SO_2 absorption costs are the major cause of the capital investment cost differences between the processes.

In other areas, the two processes have similar costs. The limestone scrubbing process has moderately higher gas handling costs, very slightly higher costs for solids separation (thickening and filtering) compared with particulate handling (pneumatic conveying and silo storage) and slightly lower disposal costs because of the higher bulk density of the gypsum waste. Materials handling costs for the limestone scrubbing

TABLE S-5. CAPITAL INVESTMENT SUMMARY

	Lignite		Low-sulfur western coal			Low-sulfur eastern coal		High-sulfur eastern coal	
	Lime spray dryer	Limestone scrubbing	Soda ash spray dryer	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing
Direct Costs									
Material handling	1,778	1,291	461	1,691	1,009	1,762	1,011	5,014	2,518
Feed preparation	765	2,406	91	680	1,923	909	1,944	2,438	4,618
Gas handling	10,665	13,249	9,088	10,030	11,646	9,770	11,665	11,456	13,653
SO_2 absorption	7,336	17,357	9,208	7,366	15,054	7,336	15,597	9,018	21,625
Stack gas reheat	-	-	-	-	-	-	1,225	-	3,325
Particulate collection	12,091	15,076	11,523	11,523	11,688	11,523	11,688	11,235	9,998
Particulate handling	2,163	-	750	2,057	-	753	-	2,114	-
Solids separation	-	2,268	-	-	1,828	-	1,846	-	3,350
Total, k$	34,798	51,647	31,121	33,347	43,148	32,053	44,976	41,275	59,087
Other Costs									
Solids disposal	867	790	725	719	616	855	743	1,443	1,007
Disposal site construction	3,756	3,690	7,228	2,520	2,158	2,939	2,625	4,899	3,441
Land	960	920	1,146	770	670	905	795	1,520	1,070
Other capital costs	42,246	50,313	39,228	39,757	41,472	38,551	43,478	50,959	57,348
Total, k$	82,627	107,360	79,448	77,113	88,064	75,303	92,617	100,096	121,953
Total, $/kW	165.25	214.72	158.90	154.23	176.13	150.61	185.23	200.19	243.91

Basis: TVA Design and Economic Premises

process are lower because the limestone can be simply stockpiled. Limestone grinding costs greatly exceed lime slaking costs, however, making the sum of costs for handling and preparing absorbents similar.

Annual Revenue Requirements

First-year and levelized annual revenue requirements are shown in Table S-6. Levelized costs are actual first-year costs adjusted by a factor of 1.886, an adjustment that takes into account inflation and the cost of money over the 30-year life of the installation.

The lime spray dryer has lower first-year annual revenue requirements than the limestone scrubbing process (7.61 mills/kWh versus 9.57 mills/kWh) in the lignite case. With the exception of lime costs, which are significantly higher than the limestone costs, and fuel costs, where the differences are insignificant, the lime spray dryer process has lower annual costs in each category, compared with the limestone scrubbing process. The much higher costs for maintenance, overheads, and levelized capital charges for the limestone scrubbing process easily overcome the absorbent cost advantage of using limestone.

In the low-sulfur western coal case, the soda ash spray dryer process has first-year annual revenue requirements of 7.42 mills/kWh, compared with 6.92 and 7.90 for the lime spray dryer and limestone scrubbing processes respectively. For the spray dryer processes the difference is almost entirely the result of absorbent costs, almost 1 mill/kWh for soda ash and 0.4 mill/kWh for lime. Other minor differences account for the remaining cost difference. For the limestone scrubbing process, absorbent costs are minor, less than 0.1 mill/kWh, but maintenance costs are, in general, more than 50% higher than those of the spray dryer processes. The indirect costs, overheads, and levelized capital charges account for the remaining cost difference between the limestone scrubbing process and the spray dryer processes.

For the low-sulfur eastern coal case a similar relationship prevails. Most costs differ insignificantly from those of the low-sulfur western coal case, in spite of the different flue gas bypass conditions. The lime spray dryer costs are slightly lower, primarily because of the lower lime cost in the East. The limestone scrubbing process costs are slightly higher, a result of general cost increases stemming from the lower flue gas bypass ratio. The small amount of flue gas reheat required for the limestone scrubbing process does not significantly affect process costs.

Somewhat different conditions prevail for the high-sulfur eastern coal case. The difference in cost between the lime spray dryer process and the limestone scrubbing process decreases from a 12% to 21% advantage for the lime spray dryer process in the low-sulfur cases, to only about 2% for the high-sulfur case. The increase in cost for the limestone scrubbing process in going from the low-sulfur eastern coal case to the high-sulfur eastern coal case is about 42% while the increase for the lime spray dryer process is about 70%. The salient cost factor is

TABLE S-6. ANNUAL REVENUE REQUIREMENTS SUMMARY

	Lignite		Low-sulfur western coal			Low-sulfur eastern coal		High-sulfur eastern coal	
	Lime spray dryer	Limestone scrubbing	Soda ash spray dryer	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing
Direct Costs									
Absorbent	1,663	227	2,661	1,030	150	848	156	8,430	1,127
Operating labor and supervision	1,025	1,212	846	972	1,140	1,022	1,175	1,202	1,341
Fuel	335	297	265	262	215	329	242	653	407
Electricity	1,617	1,986	1,523	1,464	1,508	1,458	1,518	1,582	2,428
Steam	-	-	-	-	-	-	234	456[a]	1,330
Other utilities	20	20	10	12	17	21	18	20	26
Maintenance	2,232	3,851	1,863	2,136	3,219	2,058	3,355	2,649	4,404
Analysis	88	70	88	88	70	88	70	89	105
Total direct, k$	6,980	7,663	7,256	5,964	6,319	5,824	6,768	15,081	11,168
Indirect Costs									
Overheads	1,794	2,872	1,475	1,717	2,467	1,688	2,563	2,097	3,289
Total O and M, k$	8,774	10,535	8,731	7,681	8,786	7,512	9,331	17,178	14,457
Capital charges	12,146	15,782	11,679	11,336	12,945	11,070	13,615	14,714	17,927
Total, k$	20,920	26,317	20,410	19,017	21,731	18,582	22,946	31,892	32,384
Total, mills/kWh	7.61	9.57	7.42	6.92	7.90	6.76	8.34	11.60	11.78
Levelized									
Total, k$	28,694	35,651	28,146	25,822	29,515	25,238	31,213	47,111	45,193
Total, mills/kWh	10.43	12.96	10.23	9.39	10.73	9.18	11.35	17.13	16.43

a. Boiler heat loss in lieu of reheat

absorbent cost. Absorbent costs for the limestone process increase by a factor of seven. Absorbent costs for the lime spray dryer process increase about tenfold. For the lime spray dryer process, however, this increase results in absorbent costs totaling about 27% of the total first-year annual revenue requirements; for the limestone scrubbing process only 3%. Other costs increase little in comparison and in general the increases are similar for both processes. A significant requirement for flue gas reheat also appears in both processes, one in the form of steam, the other in the form of hot flue gas.

The previously discussed first-year annual revenue requirements do not include the effects of inflation or the time-value of money on the annual direct costs (such as raw materials, operating labor, etc.). The levelized annual revenue requirements (shown in Table S-6), however, do take these factors into consideration. As is apparent from Table S-6, levelizing the annual revenue requirements results in a significant increase in the magnitude of the costs. For the lignite and the low-sulfur coal cases where annual direct costs are minor relative to the capital charges, levelizing the revenue requirements does not change the relative economics of the lime spray dryer and the limestone scrubbing processes. However, for the high-sulfur coal case, where the direct costs for the lime spray dryer process are significantly higher than those for the limestone scrubbing process, levelizing the annual revenue requirements results in a reversal whereby the lime spray dryer becomes 4% higher in cost than the limestone scrubbing processes. In fact using the results of this study over the 30-year life of the FGD system, the limestone scrubbing process is $60M less expensive than the lime spray dryer process.

CONCLUSIONS

The development of spray dryer FGD has been rapid and several vendors will soon have full-scale systems in commercial operation. The technical and economic feasibility of the vendors' approaches to design features such as type atomizer, degree of approach to saturation temperature, the particulate collection method, and waste recycle remain to be demonstrated. Current trends suggest the majority will use a lime slurry with rotary atomizers, partial flue gas bypass, waste recycle, and fabric filter collection. For the near future utility spray dryer FGD systems will probably be limited largely to low-sulfur coal applications where sufficient pilot-plant data have been generated to allow detailed design of full-scale systems.

Interest in dry injection continues but development of these processes is hindered by the lack of an economical absorbent. Nahcolite, the most promising candidate, is unlikely to be available in sufficient quantities for several years. The development of processes using other absorbents, including limestone, is only beginning and the practicality of such processes remains to be demonstrated.

The spray dryer processes are similar in cost and both are substantially lower in capital investment and annual revenue requirements than the limestone scrubbing process at least for low-sulfur applications. The lime spray dryer process is more cost effective than the limestone scrubbing process for all three of the low-sulfur coal cases studied. For the high-sulfur coal case the lime spray dryer has a lower capital investment than the limestone scrubbing process but the first-year annual revenue requirements are essentially equivalent, given the uncertainties associated with a study-grade estimate. The differences are largely the result of lower spray dryer equipment costs, compared with wet scrubbers, and lower utility and maintenance costs. Offsetting these advantages, absorbent costs become substantial for the spray dryer processes at high coal sulfur levels. The relationship of equipment costs is unlikely to change substantially. Operating and absorbent costs could, however, require adjustment as more operating experience is gained, particularly in the high-sulfur coal case where important design considerations (stoichiometry, etc.,) are not well defined.

TECHNICAL REVIEW OF DRY FGD SYSTEMS

AND ECONOMIC EVALUATION OF SPRAY DRYER FGD SYSTEMS

INTRODUCTION

One of the recent developments in flue gas desulfurization (FGD), the so-called dry scrubbing technology using a concentrated alkali solution or slurry in a spray dryer, is currently receiving extensive attention. Much of this interest is due to some potentially significant technical and economic advantages of the spray dryer over conventional wet scrubbing FGD technology. In particular, the process design is relatively simple, operating costs are low, and a dry waste, rather than a wet sludge, is produced.

Spray dryer FGD in the U.S. evolved in part from FGD studies begun in the 1960's in which absorbents were injected into the flue gas as dry powders and collected in fabric filter baghouses or ESP's. Although many of these dry absorption studies were initially disappointing, the potential benefits of such uncomplicated approaches to FGD have maintained interest in this technology. Recently, these studies have been expanded to include coinjection of limestone with the coal in pulverized coal boilers. The evolution and current status of these dry absorption processes are reviewed as a portion of this study.

Quite rapidly in the past few years, spray dryer FGD technology has expanded such that numerous companies and consortia are active in pilot studies and several commercial-sized units are under contract. In part this growth is a result of the potential technical and economic attractiveness of the method and to the increased use of western coals by utilities, for which spray drying is particularly suited. Also important is the broad base of spray dryer and particulate collection technology from which these processes can be directly evolved. Quite frequently a company experienced in one of these technologies has joined with one experienced in the other to market a spray dryer FGD process. The development of this phase of the FGD industry, its general technical considerations, and its status in the early months of 1980 are discussed as a portion of this study.

These dry scrubbers have one significant disadvantage--the need to use a highly reactive absorbent to achieve acceptable sulfur removal rates. This requires an expensive (relative to limestone) absorbent such as lime or soda ash. If the savings in capital charges and operating and maintenance costs by using the spray dryer systems are larger than the raw material cost penalty for lime or soda ash, however, the spray dryer FGD systems will remain economically competitive with the wet scrubbing systems.

This is one of the reasons that the first commercial applications are on utility boilers fired with western coals. Since these coals are normally low in sulfur, the amount of sulfur to be removed, and hence the consumption of absorbent, is low compared with eastern bituminous coal. Another advantage for the spray dryer processes for western coals is the relatively high calcium content of this coal. Not only does this alkalinity react with sulfur in the boiler and thus reduce the amount of sulfur removal required in the FGD system, the alkalinity in the fly ash can also be used to remove SO_2 from the flue gas in the spray dryer by recycling the fly ash with the absorbent. Recycling thereby decreases consumption of alkali raw material.

Although capital investments and revenue requirements for these dry scrubbing processes have been estimated by various process vendors and compared with a conventional wet limestone scrubbing process, no independent economic comparisons have been published. In addition to carefully describing the technology, the major purpose of this study is to make an economic comparison of the spray dryer processes (similar to the majority of contracted commercial units) with a conventional state-of-the-art wet limestone scrubbing process using the same design and economic premises for four alternative coals. Two of these coals, a lignite and a low-sulfur (0.7% S) western coal, reflect the current utility market for these spray dryer systems. The other two coals are eastern bituminous, one a low-sulfur (0.7% S) coal and the other a high-sulfur (3.5% S) coal. In addition to these base-case evaluations, a sensitivity analysis is included for each of the four coal cases. In this sensitivity analysis the annual revenue requirements are calculated for various absorbent costs and stoichiometries.

CONCLUSIONS

The development of spray dryer FGD has been rapid and processes by several vendors will soon be in commercial operation. The technical and economic feasibility of the vendors' approach to design features such as type of atomizer, degree of approach to saturation temperature, the particulate collection method, and waste recycle remain to be demonstrated. Current trends suggest the majority will use a lime slurry with rotary atomizers, partial flue gas bypass, and fabric filter collection. In addition, most of these spray dryer FGD units will probably be limited to low-sulfur coal applications, at least for the near future, since very little development work has been done on high-sulfur coal applications.

Interest in dry injection continues but development of these processes is hindered by the lack of an economical absorbent. Nahcolite, the most promising candidate, is unlikely to be available in sufficient quantities for several years. The development of processes using other absorbents, including limestone, is only beginning and the practicality of such processes remains to be proved.

The spray dryer processes are similar in cost and both are substantially lower in capital investment and annual revenue requirements than the limestone scrubbing process, at least for low-sulfur applications. The lime spray dryer process is more cost effective than the limestone scrubbing process for all three of the low-sulfur coal cases studied. For the high-sulfur coal case the lime spray dryer has a lower capital investment than the limestone scrubbing process but the first-year annual revenue requirements are essentially equivalent, given the uncertainties associated with a study-grade estimate. The differences are largely the result of lower spray dryer equipment costs, compared with wet scrubbers, and lower utility and maintenance costs. Offsetting these advantages, absorbent costs become substantial for the spray dryer processes at high coal sulfur levels. In fact, significant variations (i.e., those outside the range selected for the sensitivity analysis) in raw material costs or raw material stoichiometry could reverse these results and make the limestone slurry process more cost effective in terms of annual revenue requirements for high-sulfur coal cases. For low-sulfur applications, the annual revenue requirements are relatively insensitive to both the raw material cost and the raw material stoichiometry. The relationship of equipment costs is unlikely to change substantially. Operating and absorbent costs could, however, require adjustment as more operating experience is gained, particularly in the high-sulfur coal applications where important design considerations (stoichiometry, etc.) are not well defined.

SPRAY DRYER FGD TECHNOLOGY

Spray dryers have been used for years for a wide range of drying, reaction, and purification processes in the chemical industry. However, essentially all of these applications involve smaller scale spray dryers than those which will be typical in FGD. The uses and the technology for industrial applications are amply documented (1,2). The technique is most useful for materials difficult to dry by other methods, for heat-sensitive materials requiring a rapid drying rate, for drying to a particular particle configuration, and for rapid, intimate mixing of reactants. In the most common spray dryer system the heated gas enters a cylindrical, conical-bottom vessel through a manifold at the top and leaves through a side or bottom opening. The atomized liquid is sprayed into the gas stream from the upper part of the vessel. Drying occurs while the liquid is suspended in the gas flow, and the solid material is collected at the bottom of the dryer or is carried out in the gas stream and collected in external equipment, or both. The liquid is atomized by numerous designs of rotary disk atomizers or high-pressure or two-fluid nozzles that produce droplets in the size range of one to several hundred micrometers, depending on the material and conditions.

In the most common spray dryer operations (i.e., those that produce solid particles) important requisites are that the particles dry in suspension without impinging on and sticking to the chamber surfaces and that the gas does not reach its saturation temperature so that no wet particles are produced. In most applications this is achieved by proper design configurations and control of operating conditions. For example, gas temperatures and liquid concentrations and rates can be adjusted to control drying conditions and exit conditions. In FGD applications the latitude of these controls is considerably restricted. The quantity of the flue gas is determined by boiler operating requirements. More importantly, the flue gas temperature leaving the boiler is fixed by the necessity of extracting as much heat from it as possible and returning the heat to the boiler in the combustion air, a necessity dictated by a high boiler efficiency requirement. The flue gas leaving the boiler has been cooled to a temperature just above the saturation point of the sulfuric acid (i.e., formed by reaction of flue gas SO_3 and H_2O). Condensation of sulfuric acid would cause intolerable corrosion in the air heater and ducts. This temperature is usually about 300°F, depending on the sulfur content of the coal and the combustion conditions. Furthermore, the amount of absorbent added is fixed by SO_2 removal requirements. Effective spray dryer FGD operation is thus restricted to a design configuration with little latitude in specification of optimum operating conditions.

BACKGROUND

Although there are several vendors developing spray dryer FGD systems, the process design and principles of operation for all systems are generally similar. Differences that do exist are based to a large extent on vendor experience and are relatively difficult, from a technical viewpoint, to compare. They are based on each vendor's optimization of the technology, and there is no commercial FGD experience to verify the results. A prime example of such process design differences is the type and number of atomizers used in the spray dryer. Thus, in the following sections the spray dryer concept will be discussed in the form of a broad overview, and the process differences between vendors will be outlined in later sections where the processes are discussed in more detail.

In its general form, a spray dryer FGD system consists of two basic units, a spray dryer and a particulate collection device. The spray dryers most frequently used are of conventional design and use rotary atomizers. The collection system is either a utility-type ESP or, more frequently, a utility-type fabric filter baghouse. The spray dryer is typically situated downstream of the boiler air heater and is followed by a baghouse. Some flue gas may bypass the spray dryer, depending on SO_2 removal requirements and design considerations, but all of the flue gas must pass through the baghouse.

Technical Comparison of Spray Dryer and Wet Scrubbing FGD

Spray dryer FGD systems have several potential advantages over wet scrubbing FGD systems, as well as several technical and economic limitations that could restrict these advantages in some applications. Spray dryer FGD does not have the large volume of liquid scrubbing medium recirculating through the absorber with its high energy requirements for pumping and the corrosion and scaling problems that have often been a vexation of wet scrubbing processes. Only a relatively small stream of liquid is pumped into the spray dryer, and this liquid does not contact the spray dryer surfaces. Expensive corrosion-resistant materials are unnecessary and the operating expenses associated with high corrosion rates and plugging are eliminated. High flue gas pressure drops associated with wet scrubbing are also reduced. Since the flue gas does not reach its saturation temperature, mist eliminators and flue gas reheaters may not be necessary. These are a frequent source of plugging and corrosion problems in wet scrubbing systems, and reheat is also an important energy requirement. Collection of a dry waste instead of a sludge that may require extensive treatment is also a major advantage. Equally important, the collection of fly ash may be combined with collection of the dry FGD waste, eliminating the need for the separate fly ash collection and handling facilities that have been found necessary with many wet scrubbing systems.

On the other hand, there are certain disadvantages and limitations associated with the reaction conditions in spray dryer FGD. The flue gas-absorbent contact time and efficiency of this contact are limited by the mechanics of the process. Absorbents which are more reactive and expensive than limestone are required, and these must sometimes be used at higher stoichiometric ratios than is common in wet scrubbing. The cost of these absorbents is a major consideration in spray dryer FGD. There is also a limit to the quantity of water, and hence the quantity of absorbent, that can be added to the flue gas. For some high-sulfur coals, this could limit SO_2 removal.

The overall effects of these advantages and limitations have not been fully defined, particularly the extent to which spray dryer FGD will be technically and economically applicable to high-sulfur coal. It is obviously most suitable to low-sulfur coal applications, especially if the fly ash alkalinity can be used to supplement the absorbent.

Process Chemistry

Although the basic chemistry of SO_2 removal in spray dryer FGD (i.e., the reaction mechanisms and rate-controlling steps) is not completely defined, additional work is continuing toward a better understanding of the process chemistry. With additional work it is hoped that further improvements in spray dryer performance can be achieved.

The flue gas normally enters the spray dryer at about 300°F. The absorbent liquid is sprayed across the direction of gas flow. Typical L/G ratios are about 0.3 gal/kaft^3. The SO_2 and HCl in the flue gas are absorbed into the liquid and react as shown below for a lime slurry process.

$$Ca(OH)_2 + SO_2 \rightarrow CaSO_3 + H_2O \uparrow \quad (1)$$

$$Ca(OH)_2 + SO_3 \rightarrow CaSO_4 + H_2O \uparrow \quad (2)$$

$$Ca(OH)_2 + 2HCl \rightarrow CaCl_2 + 2H_2O \uparrow \quad (3)$$

In addition to these primary reactions, the following secondary reaction also occurs:

$$CaSO_3 + 1/2O_2 \rightarrow CaSO_4 \quad (4)$$

All of the water evaporates in the spray dryer, forming particles composed of the reaction products and, usually, unreacted absorbent. Any remaining SO_2 continues to react, although at a much slower rate, with the dry absorbent in the particles during their passage to the collection device. If a fabric filter is used, passage of the flue gas through the built-up particulate cake provides valuable additional contact time for the reaction of SO_2 and any remaining absorbent. Some process vendors claim that 10% to 20% of the overall SO_2 removal may occur in this manner if a baghouse is used.

Spray dryer technology has some inherent advantages over the dry phase reactions (3). Since the water in the atomized solution in spray dryers evaporates from the surface of the droplet inward, the initial deposition of solids in the droplet occurs at the outer edges when a soluble absorbent such as soda ash is used. The rate of diffusion of the solid back into the solution is slower than the flow of water from the interior to the surface, and therefore the solids tend to accumulate at the surface. The resulting particulate matter is very porous and, depending on the drying rate, may even be hollow. Since one of the requirements for good raw material utilization is high porosity (which gives a large available surface area for reaction), the particles formed from solutions in spray dryers have a high reactivity.

Although a somewhat similar situation prevails in the lime-slurry-based spray dryer FGD system, the low solubility of lime results in the formation of a highly porous outer layer surrounding a relatively dense inner particle. As the water in the atomized droplets evaporates, lime particles are compressed until a single particle is formed. In contrast to the particles formed in the soda ash system, which are very porous throughout, the lime particles tend to be porous only in their outer layers and have a solid inner core. The evaporation of the remaining moisture in the particle is controlled by the diffusion of the water through these pores to the surface. The final moisture content is a function of the diffusion rates, vapor pressures, and temperature differences between the particles and the flue gas, and the residence time.

The previous discussion of the drying stage of the particles is further complicated by reactions with the flue gas. Previous investigations (4) have indicated that the reaction of SO_2 with lime particles causes the outer layer of the particle to expand as calcium sulfite is formed. This expansion constricts the pore openings to the inner regions of the particle. Thus, the central regions of the particle remain relatively unreacted, and the overall raw material utilization may be significantly less than the 100% conversion present at the surface of the particle. In other words, the reaction tends to seal the core and prevent additional reaction. These problems can be minimized by grinding the makeup lime as small as economically feasible, recycling the collected FGD waste, or adding an inert material to prevent the pores in the particles from becoming sealed.

For sodium- (or other highly soluble alkalis) based systems this pore closure problem is not as significant since the central core of the atomized solution has no solid alkali particles and in fact may even be hollow due to the drying mechanism. In terms of SO_2 absorption, the highly soluble alkali metal reagents such as soda ash not only have the advantage of being more chemically reactive than calcium compounds, but they also expose more surface area to the SO_2-laden flue gas during the drying stage. Thus, these sodium-based alkalis have an inherent advantage over calcium compounds in terms of raw material utilization.

The reactions proceed rapidly as long as there is surface moisture and absorbent available. Therefore, an efficient process design is one in which as much water as possible enters the spray dryer and the exit flue gas temperature closely approaches the flue gas saturation temperature. Another important operating requirement for an efficient process using lime is that the lime is ground as small as economically feasible. Thus the lime particles have the highest possible surface area for reaction, and they retain their surface moisture as long as possible. This also minimizes the amount of lime unable to react due to coating with reaction products. In actual practice, however, process economics determine the optimum particle size, and there is also a need for a safety margin between the actual flue gas temperature and the flue gas saturation temperature so that condensation or wet operation does not occur.

Fly Ash Composition

The composition of the fly ash is also important in spray dryer FGD chemistry. Coals contain varying amounts of reactive metals, predominately calcium, that form reactive alkaline ash components (5). These components react to some extent with SO_2 and HCl in the flue gas and can, in some cases, supplement the FGD absorbent. In the past this ash property has not been generally important. Historically almost all utility coal was mined in the Appalachian or Central basins. These coals are typically low in these metals. In contrast, western coals that are coming into increasing use by utilities are typically higher in these metals. Fly ash from western coals and lignites often contains appreciable reactive alkalinity and is used as an FGD absorbent or absorbent supplement (6). A comparison of the ash compositions of eastern and western fly ashes is shown in Table 1.

TABLE 1. FLY ASH ANALYSIS COMPARISON

Fly ash component	Western fly ash, wt %	Eastern fly ash, wt %	Lignite fly ash, wt %
SiO_2	32.2	50.8	23.0
Al_2O_3	17.4	20.6	11.5
Fe_2O_3	6.0	16.9	8.6
CaO	20.0	2.0	21.6
MgO	4.7	1.0	6.0
Na_2O	1.7	0.4	5.9
K_2O	0.5	2.6	0.5
TiO_2	1.0	2.5	0.5
SO_3	15.3	2.4	19.2
Other	1.2	0.8	3.2
Total	100.0	100.0	100.0

Note: Based on published analyses that were used to develop the design premises.

The use of alkaline fly ash as an absorbent is made somewhat difficult by the physical and chemical nature of the fly ash particles, which renders much of the alkalinity relatively unreactive or physically inaccessible to the SO_2. During its passage through the spray dryer FGD system, for example, alkaline fly ash appears to contribute relatively little, compared with its total alkalinity, to SO_2 removal. This is thought to be due to the difficulty associated with a gas-solid reaction at these temperatures. If, however, it can be slurried with water and recycled through the spray dryer, it will be much more reactive and can be used as a supplement to reduce absorbent consumption. Its use is facilitated in spray dryer FGD by the fact that fly ash is collected with the FGD particulate waste. Depending on the fly ash alkalinity and the amount of unreacted absorbent in the waste it may be economically desirable to recycle and reuse a portion of the waste.

Importance of Coal Characteristics

Several factors that are important in the spray dryer FGD technology depend on the type of coal being burned. For example, the previously mentioned fly ash composition can be a major factor in the design of the FGD system. With low-sulfur eastern bituminous coal (in this study we define low-sulfur coal as a coal that when burned requires only 70% SO_2 removal), which typically contains very little available alkalinity in the fly ash, recycle of the waste material would probably not be economically attractive; in addition, absorbent utilization would approach 100% without recycle. For high-sulfur coals (defined as all coals requiring more than 70% SO_2 removal), recycle could be required regardless of the coal rank since the process economics would probably dictate a higher absorbent utilization than could be achieved by a once-through waste-producing system.

For low-sulfur western coals and lignites, which typically have a relatively high amount of available alkalinity in the fly ash, waste recycle could provide substantial amounts of alkali for SO_2 absorption. In fact in a recent evaluation for EPA (7), the available alkalinity in the recylced FGD - fly ash waste accounted for nearly 20% of the total alkalinity injected into the spray dryer. Considering the high unit cost for lime and soda ash, this represents a substantial cost savings in annual revenue requirements.

The rank, and hence the heating value, of the coal being burned in the boiler also affects the spray dryer FGD system in the degree of SO_2 removal required. Lignite, because of its low heating value, must have a sulfur content of less than 0.7% (and for lower quality lignites even less than 0.5% sulfur) in order to qualify for the 70% SO_2 removal emission regulation. Subbituminous coals, on the other hand, can approach 1.0% sulfur and still only require 70% SO_2 removal. Bituminous coal can approach 1.2% sulfur before needing more than 70% SO_2 removal.

Additionally, the rank of the coal being burned affects the design of the spray dryer system because the lower rank coals typically have higher moisture contents. For example, on an as-fired basis, moisture levels for bituminous coals are normally less than 10%, moisture levels for subbituminous coals range from 10% to 30%, while those for lignite can approach 40%. These higher moisture levels for lower rank coals result in higher moisture levels in the flue gas and therefore a higher flue gas saturation temperature. The amount of water that can be injected into the spray dryer decreases as the flue gas saturation temperature increases. Thus for the same inlet SO_2 concentration and SO_2 removal efficiency, the absorbent liquid to the spray dryer must have a higher solids content for flue gas from a lignite-fired boiler than for a bituminous-coal-fired boiler. For 70% SO_2 removal in a low-sulfur coal application, this is not significant; however, for a higher sulfur coal with a higher SO_2 removal requirement and waste recycle rate, absorbent liquid-handling problems could present potential problems.

Since the spray dryer FGD processes are most cost effective when sulfur removal efficiencies are as low as possible (i.e., 70% removal) to minimize consumption of expensive alkali raw materials, the best possible application would be either a low-sulfur lignite or subbituminous coal. In addition, the fly ash should be analyzed for available alkalinity, since recycle of highly alkaline fly ash could make a higher sulfur lignite or subbituminous coal more cost effective than a lower sulfur fuel with a low-alkalinity ash.

Comparison of Absorbents

Potential absorbents can be classified depending on their solubility in aqueous solutions--sodium-based absorbents and calcium-based absorbents. Economically practical sodium-based absorbents are soda ash, and the naturally occurring ores of $NaHCO_3$ and Na_2CO_3, nahcolite and trona. Economically practical calcium-based absorbents are essentially limited to lime.

Sodium-based absorbents are extremely soluble in water. This means that these absorbents are very reactive in the spray dryer environment, giving a high SO_2 removal efficiency and simultaneously, a high raw material utilization. Another advantage due to their high solubility is a lower maintenance cost and their compatibility with less expensive materials of construction. However, this high solubility also means that the waste generated in these systems is difficult and expensive to dispose of in an acceptable manner for most applications. The other drawback to sodium-based systems, at least for eastern locations, is the high cost of soda ash, the primary sodium-based raw material. The soda ash currently being marketed is processed to a high purity because of the requirements in other applications; however, for these FGD applications this high purity is not required. Therefore, various vendors are considering other potentially lower cost sodium sources such as nahcolite and trona. In fact, even wastes from the production of high-purity soda ash are being considered (8).

In contrast to the sodium-based compounds, lime has a relatively low solubility in water and hence is used in the form of an aqueous slurry. This results in a lower reactivity in the spray dryer and a significantly lower utilization rate, particularly at high SO_2 removal efficiencies. These difficulties, however, can be partially overcome by modifications in the process design. Since the lime is used as a slurry, maintenance costs are somewhat higher for the lime-based system and more expensive materials of construction are required. A slaker and milling equipment are required, and the slurry is more abrasive in pumps and atomizers than a solution.

These disadvantages of calcium-based systems must be balanced against the significant advantages associated with the insoluble calcium-based wastes. Since the waste material is removed as a dry material and is relatively insoluble, it can be disposed of in a landfill rather than a more expensive lined pond required for the sodium-based wastes. In fact, due to the unreacted lime available in the waste material, it is claimed by some vendors that with the addition of the proper amount of water this waste will form a solid material similar to a low-grade concrete. Thus leaching of water soluble salts, such as calcium chloride and calcium sulfite, and fly ash components would be minimized.

Two-Fluid Nozzle and Rotary Atomization

A main difference between the various spray dryer FGD processes is the type of atomizer used. Most process developers use rotary atomizers but a few are using two-fluid nozzle atomizers. Both have been used in other types of spray drying, and each has its own particular advantages and disadvantages.

A typical rotary atomizer, as shown in Figure 1 uses a rapidly spinning disk (up to 20,000 rpm) to produce the fine droplet mist in the spray dryer. The absorbent liquid flows down through the vertical, spinning shaft of the rotary atomizer unit to the rotating disk at the base of the shaft and out into the internal chamber of the disk. Centrifugal force moves the liquid out through silicon carbide inserts, which are abrasion-resistant inserts running from the internal chamber of the disk to the outer periphery. As the liquid moves to this outer, rapidly spinning edge of the disk it forms a thin layer on the face of the disk. Additional liquid moving out through the inserts thickens this layer until the centrifugal force acting on the outer layers overcome the forces holding the liquid on the disk, and the outer layers are sheared off into the gas stream. This liquid has a horizontal velocity which results in the formation of a large umbrella spray pattern characteristic of a rotary atomizer.

In spray dryer applications the most important factor is the size of the droplet. The smaller the droplet, the larger the surface area per unit volume and therefore the larger the area that is available for SO_2 absorption. However there are also practical limits on the smallness of the droplet. In order to achieve smaller droplets, the energy consumed in the atomizer must increase, and at some point the increased energy consumption is not economically justified by the increased SO_2 removal. In addition, the close approach to saturation could result in these smaller droplets remaining entrained in the flue gas to the baghouse (most vendors provide either warm or hot gas bypass immediately after the spray dryer to prevent this problem).

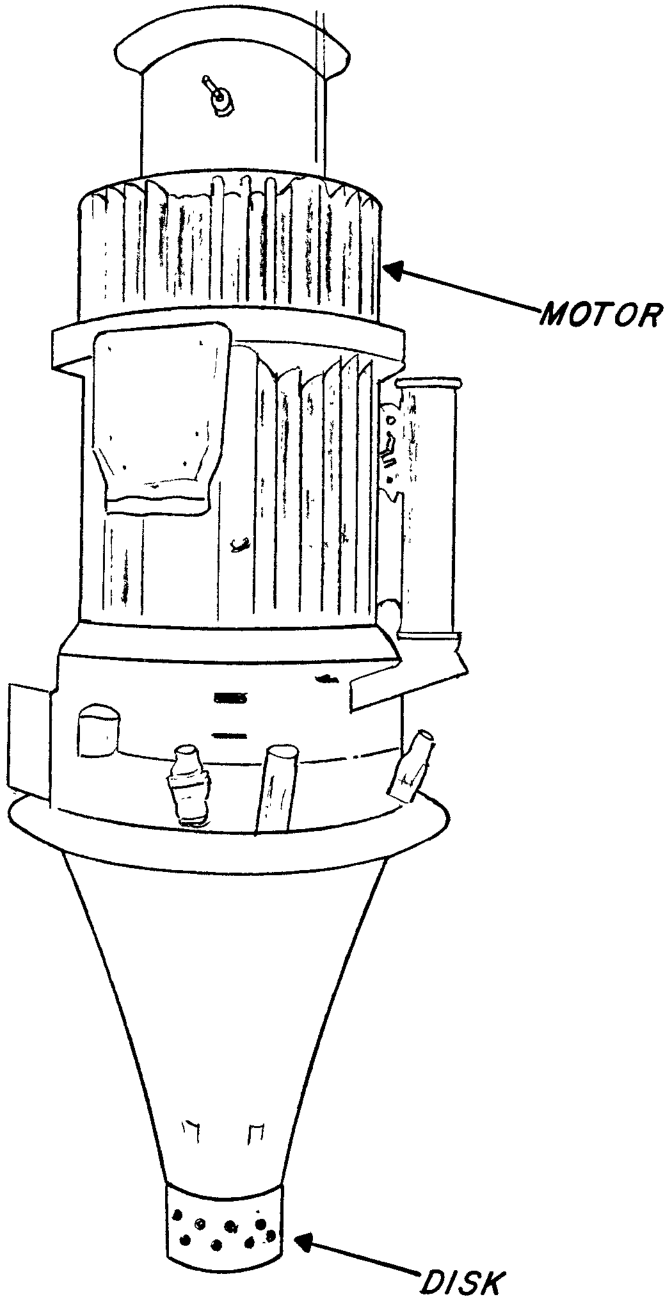

Figure 1. Design of rotary atomizer (9)

In the rotary atomizer system, the droplet size is determined by the velocity of the rotating disk and the liquid characteristics and is reasonably independent of the feed rate to the atomizer. The higher the velocity of the rotating disk, the smaller the resulting droplet size is. This factor can be a significant advantage in FGD applications where the flue gas rate and the inlet SO_2 levels fluctuate. Since the SO_2 removal efficiency required in the spray dryer remains constant regardless of inlet SO_2 levels and this removal efficiency is dependent on the droplet size, it is an advantage to be able to maintain the same droplet size regardless of the alkali feed rate required.

The primary disadvantage of rotary atomizers is that they tend to be mechanically complex in comparison with nozzle atomizers. Since rotary atomizers depend on a high velocity rotation, maintenance shutdowns could be more frequent. A second potential disadvantage, particularly with the lime-based systems, is plugging of the atomizer disk. Many developers minimize this problem through the design of the disk, but plugging may still occur. Operating experience with the full-scale utility FGD units will help to quantify the scope of these potential disadvantages.

The typical two-fluid nozzle atomizers shown in Figure 2, which are used by several process developers, depend on impinging an atomizing gas onto the liquid. This high pressure gas is typically either steam or air (although recent demonstration unit tests have persuaded one vendor previously using steam to use compressed air instead). This compressed air furnishes the energy to break up the liquid into a fine mist of small droplets.

The liquid is pumped through a central tube in the atomizer, and compressed gas is blown around the central tube into annular space. At the end of the central tube, there is an orifice that emits the liquid into the compressed gas stream at a high velocity. The compressed gas shears and further breaks up the droplets, which are then blown out of the nozzle and into the flue gas. The small diameter nozzles tend to concentrate the mist into a much narrower spray than the rotary atomizers, and therefore for large gas volumes (such as utility FGD systems) multiple nozzle atomizer arrangements are required.

The droplet size from a nozzle atomizer depends on several factors: relative gas velocity, gas density, and ratio of gas-to-liquid. Typically the relationship between these factors (and of course the geometry of the gas-liquid interaction) is complex, and actual experimental data must be used to predict the average droplet size. The average droplet size is used since, instead of producing a single, narrow range of droplet sizes as is the case with rotary atomizers, these two-fluid atomizers tend to produce a range of droplet sizes.

Since the design of the spray dryer for a specific SO_2 removal efficiency and raw material utilization depends on knowing the droplet size from the nozzle atomizer, initial full-scale testing of the nozzle atomizer is required. Even after the nozzle has been tested, the droplet

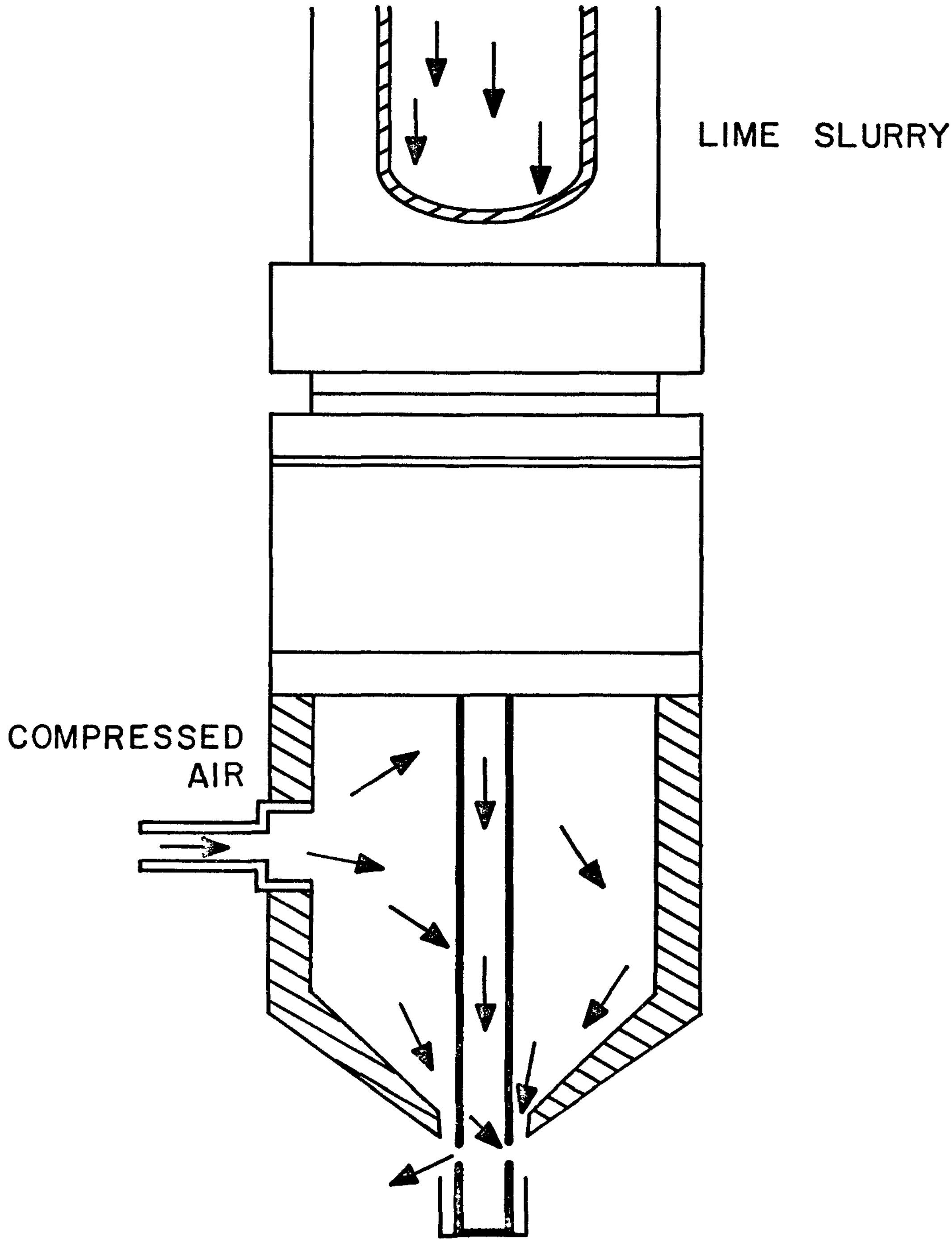

Figure 2. Design of two-fluid nozzle atomizer (10).

size may vary depending on the nozzle operating conditions (gas velocity, gas density, ratio of gas-to-liquid). This is not of overriding importance in most types of spray dryer applications where feed rates and temperature are generally well controlled. The utility FGD system, however, is one in which inlet temperature and SO_2 concentrations change, and therefore for economical operation the amount of absorbent atomized into the spray dryer must change. Changing the absorbent liquid feed rate changes the droplet size, which in turn can change the SO_2 removal efficiency. This potential problem is somewhat minimized by having multiple nozzle atomizers.

The primary disadvantage of the nozzle atomizer is that it appears to be more prone to operating problems. In fact, the first two demonstration units using nozzle atomizers are reported to have had operating problems with their atomizers. The advantages of the nozzle atomizers include the absence of moving parts and the opportunity to use large liquid passages to minimize plugging. As with the rotary atomizers, operating experience at the full-scale utility FGD units is needed to quantify the scope of the potential advantages and disadvantages.

BASIC PROCESS DESIGN CONSIDERATIONS

Once the basic conditions of the boiler application have been specified by the utility (coal and ash composition, flue gas rate and composition, SO_2 removal efficiency, etc.), other important design and operating factors that tend to characterize the current spray dryer FGD technology are usually specified by the process developer. These factors include the type of absorbent used, the design practices by which control of flue gas temperature is obtained, and the methods by which efficient use of absorbent is attained. The manner in which these factors are employed depends in large part on the type of coal burned, but they also vary with vendor preferences.

An important consideration in the design and operation of spray dryer FGD systems is the overall energy balance around the spray dryer. The temperature of the flue gas entering the particulate collection device must be high enough to insure that condensation does not occur in a fabric filter when used or the downstream flue gas ducts or stack. Once this flue gas temperature and the flue gas temperature at the boiler air heater exit have been specified, the ΔT (flue gas temperature drop) or "spray down" temperature in the spray dryer, and hence the amount of water that can be injected into the spray dryer, is fixed. SO_2 removal efficiency is controlled by the amount of alkali raw material injected into the spray dryer. The more concentrated the alkali solution or slurry, the higher the SO_2 removal efficiency. Counterbalancing this higher SO_2 removal efficiency is a lower raw material utilization, i.e., more of the absorbent passes through the spray dryer unreacted. Three potential design alternatives are available to increase the raw material utilization: waste recycle, warm gas bypass, and hot gas bypass.

Flue Gas Temperature

Most vendors agree that the following occurs in the spray dryer. As the alkaline solution or slurry is atomized and sprayed into the flue gas, the SO_2 and HCl dissolve into the alkaline solution or liquid surrounding the slurry particle and react with the absorbent. The reactions proceed at acceptable rates as long as there are surface water and reactants available. Therefore, an efficient process design is one in which as much water as possible enters the spray dryer and the exit flue gas temperature closely approaches the flue gas saturation temperature. This close approach to the flue gas saturation temperature maximizes the time surface water remains on the particles. Another important operating requirement for an efficient process using lime is that the lime is ground as small as economically feasible. Thus the lime particles have the highest possible surface area for reaction and they retain their surface moisture longer. This also minimizes the amount of lime unable to react because of coating with reaction products. In actual practice, however, process economics determine the optimum particle size and the need for a safety margin between the actual exit flue gas temperature and the flue gas saturation temperature so that condensation or wet operation does not occur. Wet operation could result in caking on fabric filters or loss of ESP efficiency. Caking could result in high pressure drops and difficulties in dislodging the FGD waste from the filters. In addition, a close approach to flue gas saturation without reheat could also result in sulfuric acid formation and corrosion in the collection equipment, ducts, and stack.

The degree of approach to flue gas saturation varies among the vendors. Typically, those using an ESP for particulate control recommend a closer approach to the saturation temperature than vendors using fabric filters. Although it is claimed by those using fabric filters that no significant operating problems occur during upset conditions when wet material passes through to the filters (i.e., when normal conditions return the filters simply dry out), these vendors design more conservatively than ESP-using vendors to minimize this problem. Those vendors recommending ESP's believe that ESP's are more moisture-tolerant collection devices and are not harmed in any way by an occasional wet upset. Therefore they typically design for a closer approach to saturation temperature and have a lower safety margin.

Although it would seem reasonable to assume that the flue gas bypass around the spray dryer would contain sufficient heat to evaporate any liquid carryover from the spray dryer before it reached the collection device, this may not in fact happen. The residence time in the ducts may not be long enough to evaporate the liquid carryover during the upset conditions (either due to the short residence time or the larger droplet size that may result from wet operation).

Stoichiometry and Absorbent Utilization

The raw material stoichiometry for the dry scrubbing systems can be misleading, particularly to those most familiar with wet FGD systems, because of the different meaning of the term stoichiometry in the dry and in the wet FGD systems. The raw material stoichiometry for the dry scrubbing systems is typically defined as the mols of absorbent/mol of SO_2 inlet to the scrubber whereas for the wet scrubbing systems stoichiometry is defined as the mols of absorbent/mol of SO_2 absorbed. Since this report compares the relative economics of both a dry and a wet FGD system, raw material stoichiometry in this report is arbitrarily defined as mols of absorbent/mol of SO_2 absorbed in all cases.

As was previously discussed in the process chemistry section, the highly soluble sodium-base scrubbing solutions tend to generate particles that are extremely porous. This high porosity results in a large surface area available for reaction and, when combined with the higher reactivity of the sodium-based compounds, normally results in a more efficient system.

For low-sulfur coal applications where only 70% SO_2 removal is required, the alkali raw material is sufficiently reactive that close approach to the flue gas saturation temperature is not required. Absorbent stoichiometry (defined as mols of absorbent/mol of SO_2 absorbed) closely approaches 1.0 and the absorbent utilization approaches 100%. Even as the SO_2 removal efficiency required increases substantially above 70%, absorbent stoichiometry remains around the theoretically required 1.0 and the raw material utilization remains high.

The process chemistry for the calcium- or lime-based systems dictates a less efficient spray dryer system because of the relatively insoluble nature of lime in aqueous solution. Although the outer regions of the particle are porous, the central core tends to be dense and unreactive, particularly after the outer regions become blinded with reaction products. Since the central core tends to remain unreacted, a higher stoichiometry is required in the lime-based system than in the sodium-based system.

For low-sulfur coals where only 70% SO_2 removal is required, absorbent stoichiometry (defined as mols of absorbent/mol of SO_2 absorbed) can approach 1.0 and the absorbent utilization can approach 100%. This is particularly true if the flue gas approaches its saturation temperature and waste recycle is employed. As the required SO_2 removal efficiency increases, the absorbent stoichiometry increases significantly above 1.0 and the absorbent utilization decreases for lime-based spray dryer systems. For high SO_2 removal efficiencies, the poor utilization of an expensive alkali raw material can lead to significant economic penalties for lime-based processes.

Two alternatives are available for overcoming this potential problem--waste recycle or operating closer to the flue gas saturation temperature. The various process vendors disagree on which method to use as well as when to use it. The vendor designs seem to run the gamut from no recycle at all to recycle for every application. One vendor apparently does not believe in recycling at all under most conditions but rather in designing sufficient flue gas bypass to allow the spray dryer to be operated very near the flue gas saturation temperature. In some designs this may even require bypassing the gas around the boiler air heater (with the associated energy penalty in the boiler heat rate) to achieve sufficient flue gas reheat downstream of the spray dryer. Some vendors believe in recycling the waste material for all applications, while others recommend recycle only when a high SO_2 removal efficiency is required or when the fly ash is highly alkaline.

Figure 3 illustrates two of the basic tenets of the lime spray dryer FGD system. The most important implication is that regardless of whether FGD waste recycle is used or not, approaching closer to the flue gas saturation temperature in the spray dryer significantly increases the SO_2 removal efficiency at constant raw material stoichiometry. For example at a stoichiometric ratio of 2.0 in a once-through system, a 70°F approach to flue gas saturation temperature results in 55% SO_2 removal while a 30°F approach to flue gas saturation temperature gives 80% SO_2 removal. The other important implication is that recycling waste material (which does not count toward the stoichiometric ratio) increases the SO_2 removal efficiency at the same approach to the flue gas saturation temperature. For example at a 30°F approach to flue gas saturation temperature and a stoichiometric ratio of 2.0, SO_2 removal for a once-through system is 80% while the recycle system approaches 98% removal. (This figure is based on one vendor's pilot-plant data and is not meant to imply these results could be achieved in a full-scale unit but rather to show the effect of the degree of approach to flue gas saturation temperatures.)

Waste Recycle--

The technical reasons for using waste recycle are obvious. One of the major annual costs for these spray dryer FGD processes is the cost of the alkali raw material. Since this raw material cost is a significant factor in the process economics, anything that reduces the makeup raw material requirements deserves serious consideration.

Waste recycle can reduce this makeup raw material requirement in two ways. First, the fly ash from the boiler contains varying amounts of alkalinity depending on the coal. Some coals (especially western coals) produce a fly ash that can have up to 30% wt CaO, although most of this CaO is not freely available for reaction as lime. However, if only 10% of this CaO (or 3% wt of the fly ash) is available for reaction, this can amount to significant quantities of "free" alkalinity for the FGD process. Since the fly ash from the boiler enters the spray dryer dry, the rate of the reaction between the SO_2 in the flue gas and the available alkalinity in the fly ash is very slow and most of this alkalinity passes through the system unreacted to end up in the process waste. The second way in which waste recyle can reduce makeup raw material requirements is that lime is added in greater than stoichiometric amounts (i.e., the amount theoretically needed to achieve the required SO_2 removal) to the spray dryer. This extra lime is provided both because the spray dryer (like any chemical reactor) is not a perfect system and because a safety margin is needed. Thus some excess alkalinity is intentionally added to the system and passes through unreacted to end up in the process waste. This total excess lime added as a safety factor varies depending on the vendor and the SO_2 removal efficiency

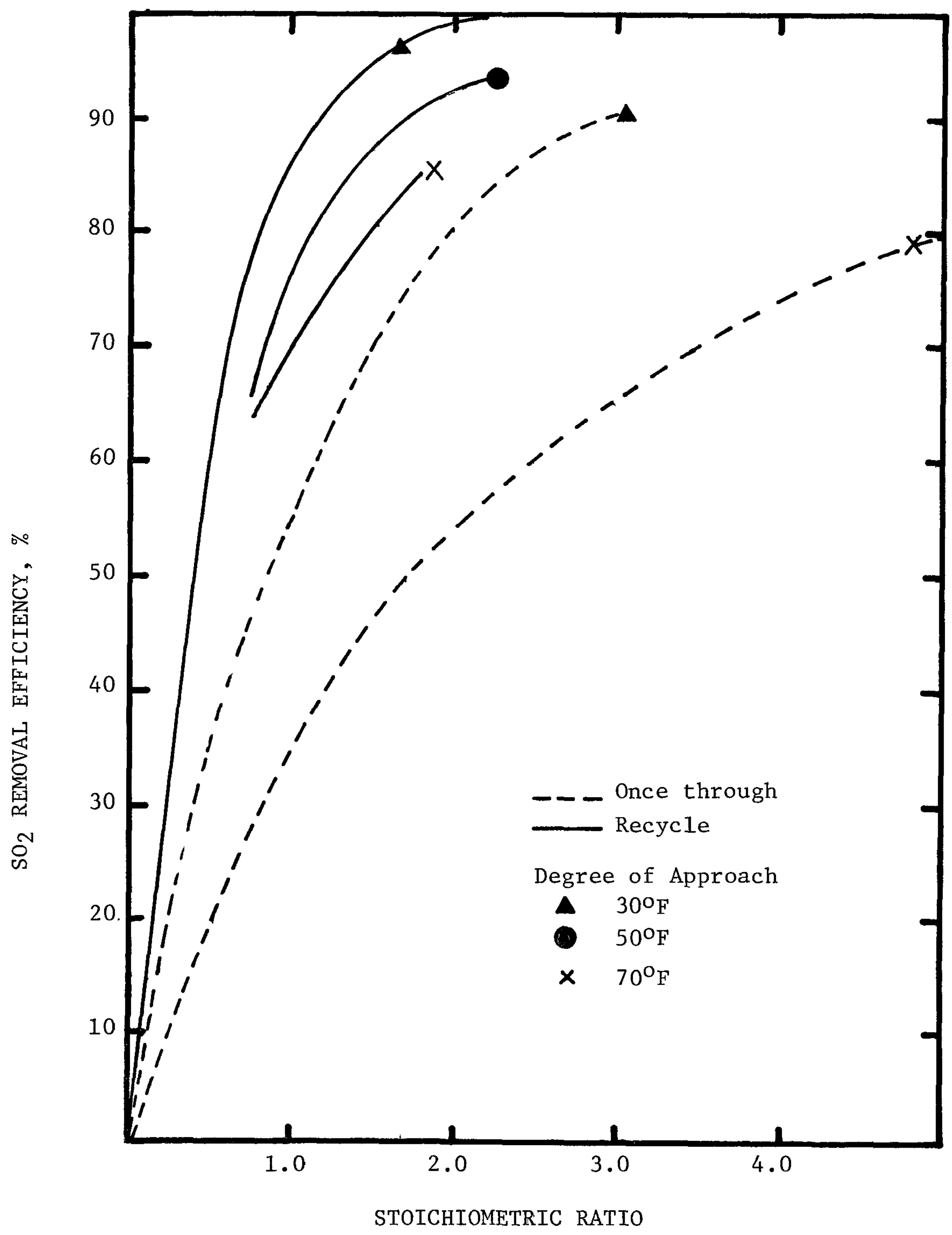

Figure 3. The effects of recycle and degree of approach to flue gas saturation temperature on SO_2 removal efficiency and raw material stoichiometry for a lime spray dryer FGD system (11).

required. However, typically, the higher the SO_2 removal efficiency the larger this safety factor will be and thus the more excess alkalinity in the waste. Thus there can be two significant sources of alkalinity in the waste from these spray dryer processes, fly ash alkalinity and unreacted raw material.

Since the equipment to recycle the process waste involves an additional capital expenditure (on the order of \$2/kW for a low-sulfur coal case), initially most of the vendors believed that the only time it would be economically attractive to recycle the waste was when the fly ash was highly alkaline or when a high SO_2 removal efficiency was required. Thus by using waste recycle the makeup raw material stoichiometry in the spray dryer could be reduced and the overall raw material utilization could be improved. There are still several vendors designing their FGD systems based on this philosophy.

Recently several of the vendors have turned from this earlier philosophy to one in which they recycle part of the FGD waste in all applications. Pilot-plant data seem to indicate that recycling the waste, even if it is low in available alkalinity, will increase the raw material utilization in the spray dryer. The current theory is that the larger particles of recycled waste provide a base on which the smaller, fresh lime particles can expose more of their surface area for reaction. Thus the makeup lime achieves a higher utilization rate than would be the case if there was no recycle. Reslurrying and recycling the FGD waste also makes any alkalinity in the waste available for reaction with SO_2, but this is currently being seen by some vendors as being relatively minor compared with the increased utilization of the makeup lime. At the present time sufficient information about test results is not available to determine which theory is a better description of this complex interaction.

Warm Gas Bypass--

When optimizing the design of spray dryer FGD systems, the best process design would involve treating the entire flue gas stream for the particular SO_2 emission regulation. For example, if 70% overall SO_2 removal efficiency is required, treating all of the flue gas for 70% SO_2 removal would be the best possible design for minimizing both costs and operating problems. For soda ash applications, particularly at low SO_2 removal efficiencies (about 70%), this is typically the case. However, if the desired SO_2 removal efficiency cannot be achieved at reasonable raw material utilizations, as is quite often the case with lime-based systems (even at 70% SO_2 removal), warm gas bypass is used. Warm gas bypass involves bypassing some of the 300°F flue gas from the boiler air heater exit around the spray dryer and returning it to the flue gas ducts upstream of the particulate removal device.

By using warm flue gas bypass and having 300°F flue gas available for reheat, the spray dryer can be operated so that the treated flue gas more closely approaches the flue gas saturation temperature. As the flue gas approaches saturation temperature, the alkali droplets retain their moisture longer and the liquid phase residence time for SO_2

absorption is increased. This results in a better raw material utilization (as well as a higher SO_2 removal efficiency) in the spray dryer. Although there is an additional capital investment for the flue gas bypass ductwork, this is offset by both the lower capital investment for the spray dryers and the lower annual cost for absorbent.

Hot Gas Bypass--

When higher SO_2 removal efficiencies (>85%) are required, most of the flue gas must be treated in the spray dryer, and there might not be sufficient heat available using warm gas bypass to reheat the flue gas. Under these conditions, hot gas bypass may be considered. Hot gas bypass involves removing some of the 700°F flue gas upstream of the boiler air heater, bypassing it around both the boiler air heater and the spray dryer, and returning it to the flue gas ducts upstream of the particulate collection device. The technological reason for using hot gas is the same as for using warm gas bypass--it allows the spray dryers to be operated closer to the flue gas saturation temperature and thus increases raw material utilization and improves SO_2 removal efficiency. In addition to the higher capital investment for the required ductwork as in the case of warm gas bypass, hot gas bypass incurs economic penalties because it affects the heat rate and operation of the boiler.

Particulate Matter Collection

The choice of particulate control device usually depends on the vendor (sometimes the buyers specify their preference). Most favor a fabric filter baghouse but at least one suggests an ESP. Although baghouses for utility applications have only been recently demonstrated on a commercial scale (12) and are more expensive than an ESP in traditional applications, the recent promulgation of a 0.03 lb/MBtu particulate emission standard and increasing use of western coals has increased their attractiveness. Not only is the baghouse capable of achieving high removal efficiency regardless of the coal being burned, but it also removes a higher proportion of the submicrometer particulate matter in the fly ash which is difficult to remove in an ESP. This ability to remove very fine particulate matter may become more important in the future because these particles may present more of a respiratory health hazard than the larger particles which have received the primary emphasis in the past (13). This fine particulate matter may also present an opacity problem for new coal-fired boilers, which are restricted to a 20% opacity requirement by the 1979 revised NSPS. In addition, the fly ash from low-sulfur western coals often have high resistivities, making collection by ESP's more difficult. Increasing use of western coals has thus increased interest in baghouses for particulate matter collection (14).

The location of the particulate collection device downstream of the spray dryer in spray dryer FGD systems has a moderating effect on their capital costs, as compared with the traditional location of the particulate control device in other FGD systems. Since the capital investments for particulate control devices are proportional to the flue gas volume, cooling and humidifying the flue gas in the spray dryer decreases the volume of flue gas to be treated in the particulate collection device.

The capital investment for the particulate collection device is lower than would be the traditional case (treating the 300°F flue gas from the boiler air heater).

A second benefit of having the particulate collection device located downstream of the spray dryer is particularly advantageous for the ESP. Cold ESP's have had trouble achieving high particulate collection efficiencies on low-sulfur coal applications where the resistivity of the fly ash tends to be very high. The resistivity versus temperature curve for the fly ash, however, is bell shaped--from a maximum near 300°F the resistivity of the fly ash tends to fall with decreasing temperature to more optimum values for fly ash collection. By locating the ESP downstream of the spray dryer in the cooler temperature region, the spray dryer is claimed to make particulate matter collection easier.

Summary

The specific design of the spray dryer FGD system depends on many factors. These include: SO_2 emission regulations, the type of coal being burned, the available alkalinity in the fly ash, and the alkali raw material selected and its delivered price. The final design will be the net result of optimizing all process variables for each specific application. Based on the limited amount of information currently available, the following generalizations can be made.

For soda ash applications:

1. Waste recycle will be used for few if any applications.

2. Hot gas bypass will be used for few if any applications.

3. Warm gas bypass will be used only for high SO_2 removal efficiencies (about 90%).

For lime applications:

1. Waste recycle will depend on the vendor. Some vendors will always recycle and some only when the available alkalinity in the fly ash is high or when the required SO_2 removal efficiency is high.

2. Warm gas bypass will probably be used even for moderate SO_2 removal (70%).

3. Hot gas bypass will be used only as a last resort and only for high SO_2 removal (over 85%).

DRY ABSORPTION TECHNOLOGY

BACKGROUND

Dry absorption, or "dry sorption," has been described as "any process that directly produces a dry product . . ." (15). Included in this definition are processes using direct injection of dry absorbents into the boiler or flue gas as well as spray dryer processes. Dry absorption processes have several potential economic and operational advantages over wet FGD processes that have been long recognized by those concerned with SO_2 emission control. These advantages have been widely discussed (15, 16, 17). Paramount are the simplicity and operational flexibility of the process, reduced energy requirements, production of a dry product, and the opportunity for simultaneous collection of fly ash and sulfur-salt waste.

Direct injection of lime was investigated in the early part of this century (18). Widespread interest, however, paralleled the development of practical particulate matter control devices for flue gas in the 1960's, particularly fabric filters. The use of fabric filters for collection has the advantage of providing additional contact of flue gas and unreacted absorbent as it passes through the built-up solids on the filters. Bechtel (16) has summarized early electric utility investigations of direct injection of dry adsorbents and the development of particulate matter control technology in utility applications.

The first modern investigations of dry adsorption, beginning with a 320-MW installation at Southern California Edison's Alametos Station in 1965, consisted of dry injection of the adsorbent followed by baghouse collection. From 1967 to 1969 Wheelabrator-Frye, Inc., evaluated dry injection methods at Public Service of Indiana's Edwardsport Station. In 1968 and 1969 Air Preheater Company made similar evaluations at the Mercer Station of the Public Service Electric and Gas Company of New Jersey.

During the early 1970's Superior Oil conducted pilot-plant studies of fixed-bed dry absorption as well as dry injection, using an ESP for particulate matter collection. Wheelabrator-Frye conducted further dry injection - baghouse collection studies at the Nucla Station of the Colorado Ute Electrical Association in 1974.

A number of absorbents were evaluated during these studies, including dolomite, limestone, quicklime, hydrated lime, sodium bicarbonate, soda ash, and nahcolite. Bechtel (16) in summarizing these studies concluded that only sodium-based absorbents were sufficiently reactive to warrant further investigation in dry injection FGD processes. Of the absorbents tested, nahcolite (a naturally occurring sodium bicarbonate) proved most effective, exceeding both soda ash and manufactured sodium bicarbonate in removal efficiency (up to 90% at 90% conversion). Commercial sodium bicarbonate is prohibitively expensive for use as a nonregenerated adsorbent. Soda ash is less expensive although considerably less efficient than sodium bicarbonate. Nahcolite has, therefore, attracted much attention for use in dry absorption FGD processes.

NAHCOLITE

Nahcolite, a naturally occurring sodium bicarbonate ($NaHCO_3$) mineral, is one of several sodium-based compounds that occur as evaporite deposits. Although ore-grade deposits are uncommon, extensive deposits are found in the Green River formation of Colorado, Wyoming, and Utah. The Unitas and Green River formations, fluvial and lacushrine basin deposits of the Eocene Age, constitute the much-discussed oil shale areas of that region (19). In the Green River Basin in Wyoming a flourishing captive mining industry produces trona ($Na_2CO_3 \cdot NaHCO_3 \cdot 2H_2O$) from underground mines in the Green River formation for soda ash manufacture (20). Yet another sodium mineral, dawsonite [$NaAl(OH)_2CO_3$] also occurs in the Green River formation in greater tonnage and wider areal extent than nahcolite, although in a more dispersed form. This dawsonite has been suggested as a potential alumina source in the future.

The major nahcolite deposits are located in the portion of the Green River formation lying in the Piceance Creek Basin of Northwestern Colorado. The deposits of most immediate economic interest occur as horizontal beds up to 20 feet thick at 1500 feet or more beneath the surface. These nahcolite deposits are of remarkably high quality, some containing up to 85% $NaHCO_3$. Other nahcolite occurs as lenses, nodules, and disseminated particles. Although proven reserves are sufficient for many years of high-volume mining, commercial mining of nahcolite has not developed. Bechtel (16) and Lutz et al. (21) discuss the status of nahcolite mining in detail. The intimate association of nahcolite with oil shale and dawsonite and the resulting complexities of Federal and State lease restrictions, problems with overlying incompetent strata, dusting and explosive conditions, and uncertain demand (only widespread power plant FGD use offers a market sufficient to justify mine development costs) have all acted to delay nahcolite mining development. Several mining companies are actively pursuing mine development plans, however (22, 23). Lutz et al. (21) estimate nahcolite suitable for FGD can be produced for $20 to $30 per ton at the mine.

Process Chemistry

One of the main disadvantages in using a nahcolite injection system for FGD is its required operating temperature for high SO_2 removal efficiencies. $NaHCO_3$ decomposes to Na_2O, H_2O, and CO_2 at about 260°C (500°F). Thus, if nahcolite is injected at relatively high flue gas temperatures, this decomposition reaction leads to porous particles of

Na_2O as the gases are expelled from the interior of the particles. This significantly increases the surface area of the particles that is available for reaction with SO_2 and minimizes pore plugging and the presence of unreacted Na_2O in the core of the particles, thus maximizing absorbent utilization.

Nahcolite reactions include the following:

$$2NaHCO_3 \cdot Na_2CO_3 \cdot 2H_2O \rightarrow 2Na_2CO_3 + CO_2 + 3H_2O$$

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2$$

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4$$

At temperatures less than about 260°C (500°F) very little of the $NaHCO_3$ decomposes and SO_2 removal is achieved only by the reaction of absorbed SO_2 with $NaHCO_3$. Since none of the explosive release of CO_2 occurs at these temperatures, internal pores in the particle tend to be small and easily plugged. This decreased surface area for reaction tends to result in poor raw material utilization since only the outer layers are converted by reaction with SO_2.

From the limited pilot-plant data available, it appears that the SO_2 removal efficiency increases as the flue gas temperature increases with temperature up to about 500°F. At this temperature the SO_2 removal efficiency approaches 90% in an optimum design. Unfortunately in most boilers this flue gas temperature occurs inside the boiler air heater (the flue gas enters the air heater at about 700°F and leaves at 300°F). At higher temperatures (those at the economizer exit) the SO_2 removal efficiency declines. Nahcolite reaches a maximum reactivity for SO_2 removal at about 550°F.

Thus, in order to achieve the optimum flue gas temperature of 500°F, one of three expensive alternatives needs to be selected. The boiler design could have twice as many air heaters, each with about half the surface area of a conventional air heater, so that the first air heater only cools the flue gas to the range of 500°F to 550°F while the second (directly after the FGD system) cools the flue gas to the normal air heater exit temperature of 300°F. Not only would the capital expense for the boiler increase because of the increased number of boiler air heaters, but the thermal efficiency of the boiler would decline since the FGD system and the second air heater would result in additional heat losses. A second alternative would be to simply bypass some of the flue gas around the boiler air heater to achieve the 500°F flue gas temperature to the FGD system. Again, however, this could substantially decrease the thermal efficiency of the boiler. The third alternative would be to preheat the incoming 300°F flue gas with an auxilary burner.

None of the three alternatives is economically attractive. Thus at the present time most of the development work concerns those applications in which significantly less than 90% SO_2 removal is required (i.e., the 70% standard under the 1979 NSPS). From the limited pilot-plant data now available, it appears that nahcolite injection systems may be able to average the 70% SO_2 removal which is potentially possible at conventional boiler air heater exit temperatures. Although this SO_2 removal is technically feasible, whether the low nahcolite utilization and other economic factors will allow an economic advantage over these systems is uncertain.

The inability to achieve 90% SO_2 removal may limit the widespread application of nahcolite injection systems. Low-sulfur coals that require only 70% SO_2 removal (i.e., those coals where the 70% control would result in an emission rate of less than 0.6 lb SO_2/MBtu heat input) will for the most part be western bituminous and subbituminous coals that have less than about 1.2% sulfur (based on 11,700 Btu/lb). Although lignites typically contain less than 1.0% sulfur, they have such a low heating value that the upper limit on sulfur level is in the range of 0.76% (based on 9,500 Btu/kWh and 6,500 Btu/lb). In either case, since the range of sulfur levels in the coal from a particular seam may vary, only those applications where the average sulfur levels are somewhat less than these values could consider nahcolite injection (barring a technical development which would significantly improve the SO_2 removal efficiency at these temperatures).

The SO_2 removal efficiency as a function of stoichiometric ratio is dependent on the nahcolite injection method used, as discussed below. As shown in Figure 4, in order to maintain the required 70% SO_2 removal efficiency required by the 1979 NSPS, the nahcolite:SO_2 stoichiometric ratio would range from approximately 1.0:1.0 for the batch injection system to 1.5:1.0 for the continuous injection system. For the semibatch injection method, the SO_2 removal efficiencies would be expected to be nearly the same (although slightly lower) as those of the batch method.

Dry injection systems will typically involve injection of a finely powdered absorbent to maximize the surface area available for gas-solid contact. Results from a previous study (24) indicate that under the same reaction conditions decreasing the particle size from 70% through 200 mesh to 70% through 400 mesh increases the SO_2 removal efficiency (at a stoichiometric ratio of 0.8) from 58% to 75%. Of course decreasing the particle size involves higher capital and operating costs and thus involves an economic tradeoff between higher annual capital charges versus increased operating costs for raw materials.

Injection Systems

As has been noted in previous studies, there are several methods of injecting nahcolite into the flue gas stream (21): continuous, batch, or semibatch. In the continuous injection system, the nahcolite is

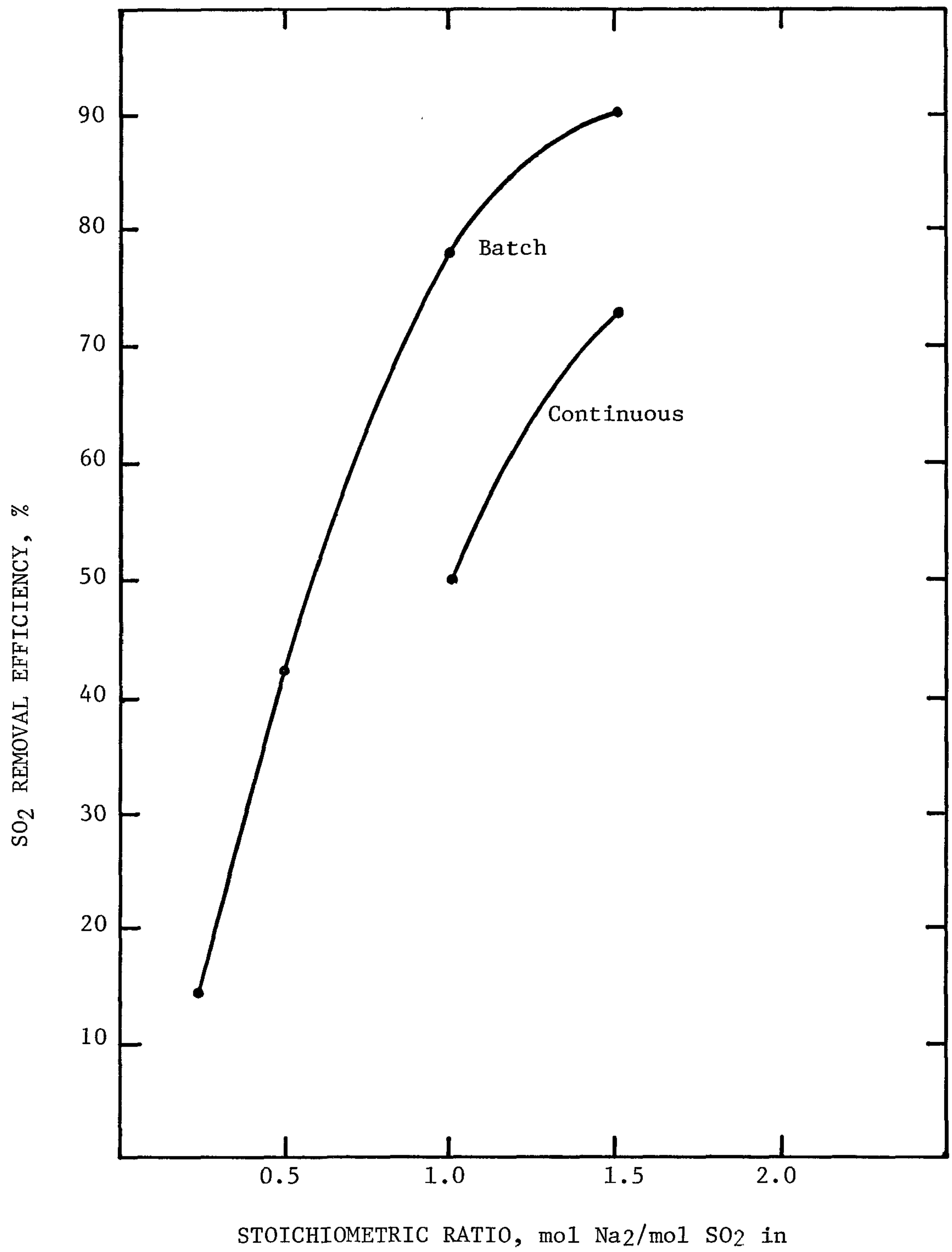

Figure 4. SO_2 removal efficiency as a function of stoichiometric ratio and injection method for nahcolite (24).

added upstream of the baghouse and travels, entrained in the flue gas, to the baghouse where it is captured to form a coating on the fabric filters. Advantages of this system include a longer residence time in contact with the flue gas, a low initial pressure drop which gradually increases as the cake builds up, potentially less operating manpower since the nahcolite can be added by process equipment, lower capital investment for the injection equipment since only one injection point per flue gas duct is required, and less spare baghouse capacity since a normal cleaning cycle can be used and the baghouse compartment can theoretically come back on-line immediately. The primary disadvantages include the potential loss of SO_2 removal in the clean compartment until an alkali cake has been built up and the potential for poor distribution of the nahcolite resulting in areas of the bags having no cake so that the SO_2 passes through unreacted.

Since most of the SO_2 removal may occur as the flue gas passes through the nahcolite cake built up on the fabric filter, a continuous injection system (where the nahcolite is injected into the flue gas ducts and is carried to the bags entrained in the flue gas) results in very little SO_2 removal immediately after the cleaned compartment comes back on-line. As the nahcolite cake builds up the instantaneous (as opposed to average) SO_2 removal efficiency increases until a maximum SO_2 removal efficiency occurs immediately before the compartment comes off-line for cleaning. Thus, the integrated, or average, SO_2 removal efficiency over the entire baghouse cycle is significantly lower than this maximum instantaneous SO_2 removal efficiency. For example, during one series of pilot tests a maximum SO_2 removal efficiency of 67% at a stoichiometric ratio of 1 was obtained. However, the average SO_2 removal over the complete cycle was only 50%. For an average 70% SO_2 removal over the complete cycle (as required by the 1979 NSPS) the stoichiometric ratio had to be increased to 1:5 (24).

The batch injection system involves precoating the bags with a layer of nahcolite following each cleaning cycle. Advantages of this system include better control over the distribution of the nahcolite on the filters and the presence of nahcolite on the bags as soon as the compartment comes back on-line. Disadvantages include the higher capital investment for multiple injection points, a higher initial pressure drop after bag cleaning, additional off-line time for each compartment as it is precoated, and additional delay time for high SO_2 removal to begin since the nahcolite must achieve reaction temperatures before efficient SO_2 removal begins.

SO_2 removal can begin as soon as the bags come back on-line since there is a nahcolite cake already present. In fact, to give the same maximum SO_2 removal previously given for the continuous injection (67%) requires a stoichiometric ratio of only 0.9:1, and furthermore, the average SO_2 removal over the baghouse cycle was 66% versus only 50% for the continuous injection system.

Since high SO_2 removal (>90%) at acceptable raw material utilization rates is difficult to achieve under the best conditions (even without suffering almost no SO_2 removal for the startup period), the compromise or semibatch system is currently being favored. Immediately following compartment cleaning, an initial precoat is added before the compartment is brought on-line. Nahcolite is also continuously injected upstream of the baghouse. Although requiring more equipment than either the continuous or the batch system, the semibatch method has the advantage of providing a continuous supply of nahcolite in the baghouse for SO_2 removal. Granted that the initial SO_2 removal efficiency (i.e., before the precoat achieves reaction temperature) may be lower, some SO_2 removal will occur as soon as the compartment comes back on-line. Some will be achieved by the entrained nahcolite and some by the relatively cool nahcolite precoat. Thus, there is not a short time period after the compartment cleaning when essentially no SO_2 removal occurs as is the case with both the continuous and the batch injection system.

Waste Disposal

Waste disposal for nahcolite or trona dry injection FGD systems is a significant problem due to the water-soluble nature of the sodium compounds. Unless treated to reduce their solubility the sodium-based wastes could easily leach into surface or underground water. Thus, unless the FGD system is located near a trona mining operation or a dry lake bed of sodium salts or some other unique situation where the dumping of the sodium waste would seem to have very little adverse environmental effect, the disposal of these wastes could be prohibitively expensive.

This environmental problem, in fact, is receiving considerable attention since without a method to acceptably dispose of these FGD wastes, nahcolite, or any other sodium-based system, may not be commercially practical. Three methods, which are described below, are currently being discussed for these sodium wastes and one, insolubilization, is being evaluated on a pilot scale by Battelle Columbus Laboratories.

Insolubilization--

The waste from the baghouse containing both FGD waste and fly ash is mixed with lime and the resulting mix is sintered at high temperature to yield what is said to be inert, insoluble material. The major process currently being evaluated is the Sinterna® process (24) which was patented by Industrial Resources, Inc. Although this process has been claimed to render the waste inert, no detailed, long-term technical evaluation has been completed. In addition the process economics--particularly the apparent high energy consumption to sinter the waste--need to be independently estimated.

Clay-Lined Isolation Cells (24)--

This procedure involves excavating a large pit and lining the floor and walls with a thick (12-24 inch) layer of impermeable clay. Within this protected boundary, individual cells, each probably sized to handle approximately one days' supply of FGD waste (and fly ash) would be built. All of the walls would be made of clay and at the end of each day a clay cap would be laid on top of each cell. This method would minimize the amount of water entering the cell and thus prevent the waste from leaching into the surrounding area. The primary unresolved problem associated with this method is the relative cost of this disposal system. Other uncertainties include how long the disposal area needs to be monitored for leaching and for integrity of the clay cap.

Regeneration of the Sodium Compounds--

Since calcium-sulfur compounds are relatively insoluble, it would be technically feasible to regenerate sodium carbonate by the reaction of the FGD waste with limestone. This of course is similar to double alkali FGD systems (25) wherein wet-scrubber sodium salts absorb SO_2 and the sodium solution is regenerated with limestone. Since the sodium salts will be a mixture of sulfites and sulfates, the FGD waste would have to be dissolved in aqueous solution and then oxidized to sulfate to avoid the disposal problems associated with sulfite-sulfate sludges. The oxidized sodium solution could be reacted with limestone to precipitate gypsum ($CaSO_4 \cdot 2H_2O$) which would be filtered from solution and disposed of in a landfill.

None of these options appear to be economically attractive alternatives. No other alternatives appear to exist at the present time, however.

TRONA

In addition to nahcolite, the other dry injection alkali that is receiving attention is trona. Since $NaHCO_3$ is more chemically active than Na_2CO_3, it is not unexpected that at the same stoichiometric ratio nahcolite (about 80% $NaHCO_3$) can achieve a higher SO_2 removal than trona (about 30% $NaHCO_3$). Trona, however, has an advantage over nahcolite in that it is already being commercially mined and is thus potentially available in large quantities for FGD systems. Since the active chemical in trona is the same as that in nahcolite most of the process chemistry is essentially identical to that given previously for a nahcolite FGD system. The primary difference is that much larger quantities of trona must be used. These larger trona requirements may be due to its lower $NaHCO_3$ content. The only major reaction difference between nahcolite and trona is that trona appears to react rapidly upon injection in the pilot-plant tests.

From the limited pilot-plant data currently available (26), trona appears to increase in reactivity for SO_2 removal as the temperature increases whereas nahcolite reactivity reaches a maximum at about 550°F. Trona also appears to require a higher stoichiometric ratio to achieve the same SO_2 removal efficiency. SO_2 removal efficiency for nahcolite and trona for continuous injection at 250°F is shown below.

	Moles Na_2/moles SO_2 in	SO_2 removal efficiency, %
Nahcolite	1.0	50
	1.5	70
Trona	1.0	40
	3.0	70

This comparison indicates that to achieve a minimum 70% SO_2 removal a substantially higher feed rate is required for trona than for nahcolite.

Part of the EPA-sponsored pilot-plant work involves upgrading the raw trona to approximately 92% $NaHCO_3$. This upgraded trona would reduce the stoichiometric ratio required and, depending on the relative cost of upgrading, might make trona a more economically practical alternative to nahcolite for dry injection FGD systems.

PAST STUDIES AND CURRENT STATUS

Numerous bench-scale and pilot-plant studies were carried out in the early 1970's as part of efforts to develop a technically feasible and cost effective dry FGD system for utility boilers. In these test programs numerous alkali raw materials for dry injection FGD systems were evaluated, including limestone, lime, soda ash, sodium bicarbonate, nahcolite, and trona. These evaluations came to the same conclusion: nahcolite was the best alkali for a dry injection FGD system. Some of these test programs are summarized below.

Past Studies

Nucla Station--

In July 1974 Wheelabrator-Frye, Inc., began a pilot-plant program at the Nucla Station of the Colorado Ute Electric Association. This unit is an 11-MW (gross) spreader-stoker-fired boiler burning an 0.8% sulfur subbituminous coal. Inlet flue gas typically contained 480 ppm SO_2, although when higher sulfur coal (1.1% sulfur) was burned on occasion, inlet SO_2 approached 900 ppm. The nahcolite was supplied by Superior Oil Company and averaged 60% $NaHCO_3$.

The baghouse used in the test program was originally designed and installed for fly ash removal only. Most of the tests involved batch operation of the nahcolite injection system (i.e., precoating the bags). The flue gas rate was 44,000 aft^3/min at 285°F. SO_2 removal efficiencies ranged from 50% to 70% depending on the raw material stoichiometry. Raw material utilization was approximately 56% at 70% SO_2 removal.

Leland Olds Station--

During the first quarter of 1977, Wheelabrator-Frye operated a pilot plant at Basin Electric Power Cooperative's Leland Olds Station

Unit 2 to demonstrate a nahcolite injection FGD system. Leland Olds Unit 2 is a 440-MW cyclone boiler burning 0.8% sulfur North Dakota lignite. The primary purpose of this test facility was to demonstrate the technical feasibility of using a nahcolite-based FGD system for the Ottertail Power Company's planned Coyote Station, which would be a design similar to Leland Olds. However, Wheelabrator-Frye also wanted to evaluate the effects of several operating conditions, including method of injection (continuous, batch, or semibatch), stoichiometric ratio, and inlet SO_2 concentration. An SO_2 spiking system was provided for the evaluation of the removal versus the inlet SO_2 concentration.

Figure 5 shows the SO_2 removal efficiency as a function of stoichiometric ratio for optimum operating conditions. However, for tests carried out at less than optimum conditions, removal efficiencies were significantly lower. Unfortunately, due to the proprietary nature of the test work, neither were the less-than-optimum conditions identified nor were the actual removal efficiencies under these less-than-optimum conditions quantified. In this pilot-plant system, which was designed specifically to test nahcolite injection for a utility FGD application, raw material utilization showed some improvement over the results obtained at the Nucla test facility. For example, 77% utilization at 83% SO_2 removal and 60% utilization at 90% SO_2 removal were claimed versus 56% utilization at 69% SO_2 removal at Nucla. However, even 60% utilization means (as noted by the vendors) that significant amounts of excess nahcolite must be injected to achieve the high SO_2 removal efficiencies that may be required at commercial utility boilers.

Development work on nahcolite injection for FGD at Leland Olds was terminated in late March 1977 due to nahcolite supply problems. (In fact, this is essentially the current status of the nahcolite injection technology.) Since there did not appear to be any hope for these supply problems to abate in the near future, investigations were undertaken for alternative alkali raw materials. However, lime, limestone, and soda ash had been previously demonstrated to be insufficiently reactive for dry injection at these temperatures, and when commercial $NaHCO_3$ was found to be too expensive for a commercial utility application, the search for an acceptable dry raw material was terminated and the development work switched to modifying the process conditions so that commercially available alkali raw materials (lime, soda ash, and trona) could be used. This, of course, led to the development of the spray dryer based FGD system.

Current Status

The attraction of a nahcolite injection process for FGD control--less equipment and a totally dry system--has led to continuing interest by various groups which continues even today. Various bench-scale units and small pilot plants are currently operating.

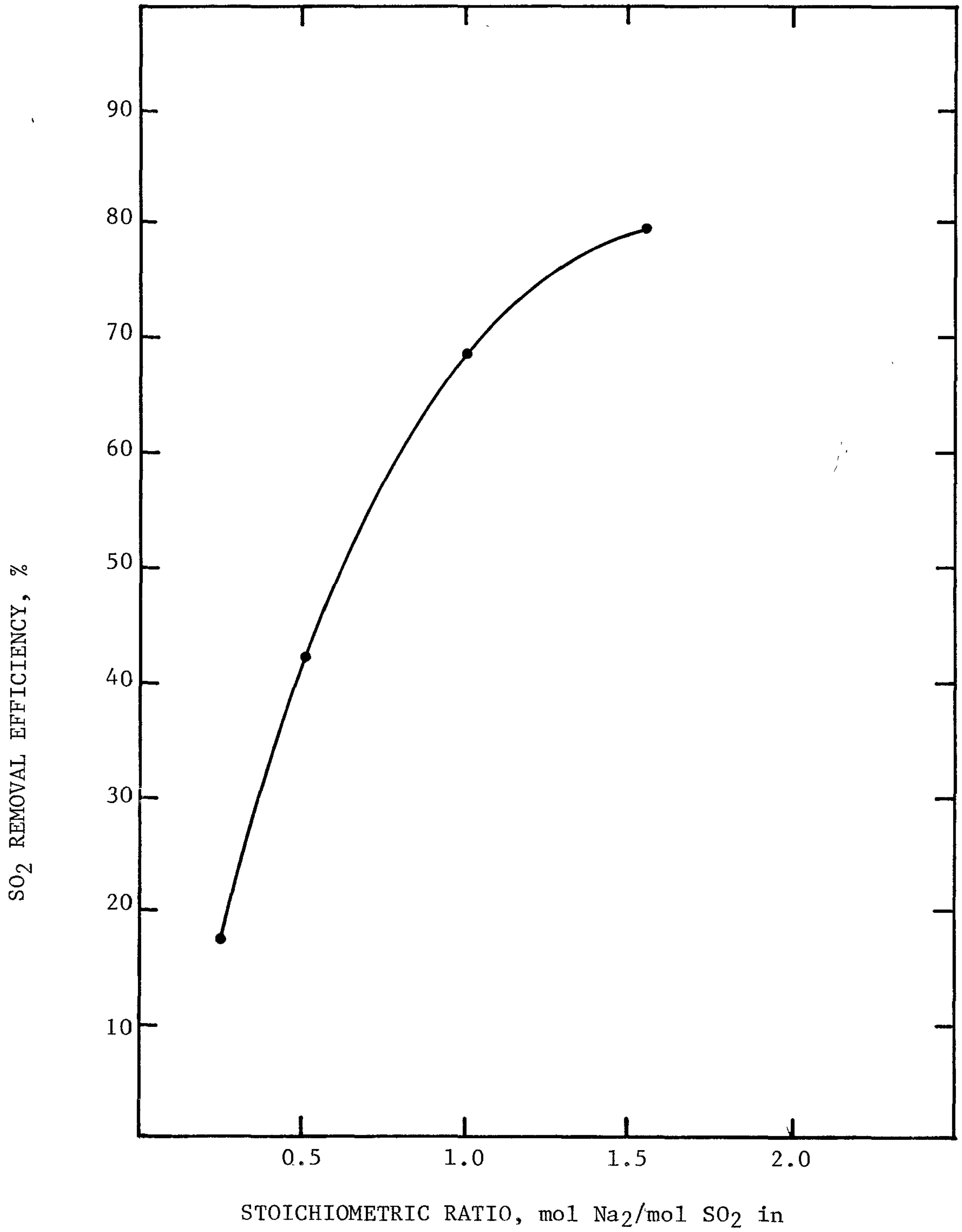

Figure 5. SO_2 removal efficiency as a function of stoichiometric ratio for nahcolite injection (24).

KVB, Inc.--

KVB, Inc. has been carrying out bench-scale tests under contract with EPRI since late 1977. The primary purpose of this work is to evaluate various alternative sodium-based raw materials including trona, nahcolite, and commercial $NaHCO_3$ at different flue gas temperatures, sorbent residence times, and stoichiometric ratios. Flue gas is provided by a coal-fired bench-scale combustor generating 725 sft^3/min.

Preliminary results using nahcolite indicate that maximum SO_2 removal efficiency is achieved at about 290°C (550°F) with removal efficiency declining as the flue gas temperature increases above or decreases below 290°C (550°F). For example, at a stoichiometric ratio of 1:0 (mol Na_2/mol SO_2 in), SO_2 removal was 67% at 150°C (300°F), 80% at 290°C (550°F), and 60% at 425°C (800°F). (These are maximum instantaneous SO_2 removal efficiencies--the integrated SO_2 removal over the total cycle is significantly less.)

Carborundum (Now Carborundum Environmental Systems)--

Since 1976 Carborundum has been operating both a small (100 aft^3/min) bench-scale unit and later a larger (1000 aft^3/min) bench-scale unit in Knoxville, Tennessee. The primary purpose is to evaluate various raw materials including commercial $NaHCO_3$, nahcolite, and ammonia sorbents. Due to the in-house nature of this research, essentially no data are available.

Martin Drake Station--

Buell has undertaken a pilot-plant study of a nahcolite injection system for FGD control at the Martin Drake Station owned by the City of Colorado Springs. This program is being partially funded by EPA and hence, perhaps for the first time, the actual pilot-plant data will be available to the public (27).

This pilot plant is treating a 3000 aft^3/min slipstream of flue gas from the Unit 6 boiler. This unit is an 85-MW boiler burning a mixture of three Colorado bituminous coals which average 0.5% sulfur and 12,000 Btu/lb. The particulate collection device for these nahcolite tests was the pilot baghouse and an existing baghouse, originally designed for fly ash control only and built by Buell. The nahcolite for the test program is being supplied by the Bureau of Mines from their existing stockpile. Other dry sorbents to be evaluated include raw trona (about 19% $NaHCO_3$) and an upgraded trona (92% $NaHCO_3$).

The pilot plant was started up in late 1979 and parametric testing was completed in May 1980. The various parameters which were evaluated included stoichiometric ratio, inlet flue gas temperature, and inlet SO_2 concentration.

In addition to evaluating these sodium-based dry sorbents, Buell is also undertaking an EPA-funded waste disposal study based on the Sinterna process. This patented process was developed by Industrial Resources, Inc., in which the waste material is sintered to make the material less soluble. The actual evaluation is being carried out by Battelle Columbus Laboratories under a subcontract with Buell.

DEVELOPMENT AND CURRENT STATUS OF SPRAY DRYER FGD PROCESSES

Most of the current development on spray dryer FGD has as its basis the work done during the early 1970's with dry injection of nahcolite for low-sulfur western coal applications. Although the use of nahcolite as a dry absorbent for SO_2 removal appeared promising from a technical viewpoint, questions about its future availability in large enough quantities for widespread FGD application led to the search for other alkali raw materials (16).

Other alkaline raw materials which did not have the availability problems of nahcolite were limestone, lime, and soda ash. Dry limestone injection into the flue gas ducts had been previously attempted but with disappointing results in terms of SO_2 removal and sorbent utilization. Although both soda ash and lime are more reactive than limestone, these raw materials could not simultaneously achieve the necessary high SO_2 removal efficiencies and the high utilizations required for economical operation when injected into the flue gas (or precoated on filters) as a dry powder (16).

At approximately the same time, the use of a spray dryer as an SO_2 absorber in the regenerable Rockwell International aqueous carbonate process (ACP) was underway. The use of a spray dryer, although more expensive than simple dry injection, offered many of the same advantages as dry nahcolite injection, i.e., a relatively simple absorption system with no need for recirculating large volumes of erosive slurry and the production of a dry waste material. Furthermore, by using the soda ash and lime in the form of an aqueous solution or slurry, respectively, both high SO_2 removal efficiencies (70% to 90%) and high absorbent utilization (70% to 100%) could be achieved.

The first use of a sodium carbonate solution in a spray dryer was in 1972 (28) but this pilot-plant work was aimed at developing the spray dryer as an absorber for the regenerable ACP process. The first applications of the spray dryer for a waste-producing FGD system were not until 1977. Several upper midwestern electric utilities who had been involved in the earlier nahcolite testing work had become interested in the application of spray dryers for FGD. Several companies built and operated small pilot plants (about 10,000 aft^3/min) to prequalify as bidders for the first commercial installations (9).

As a result of this and other pilot-plant development work, over a dozen contracts have been awarded for commercial applications of spray dryer FGD systems. Five of these contracts are for industrial boiler applications and the remainder are for utility boilers, as shown in Table 2. Wet lime and limestone FGD systems have been selected over the spray dryer systems for at least two low-sulfur western coal applications. These are the particular applications where the spray dryer FGD systems are claimed to have a significant economic advantage over wet lime and limestone scrubbing, and the specifics of why the spray dryer FGD systems were not selected have not yet been revealed.

In addition to these commercial utility applications, numerous companies are currently building or operating pilot-plant units ($\leq$10,000 aft^3/min) to further develop spray dryer technology. Those on which data were available through June 1980 are shown in Table 3. At least two of these pilot plants are mobile units that are set up on trucks and can be moved to different utility sites, thereby giving these vendors an opportunity to evaluate various site-specific conditions before preparing bids. These mobile units can also be moved to other sites where they can be used for more general studies of the technology.

For the most part, these pilot-plant units have been operated on low-sulfur, highly alkaline coals (North Dakota lignites and western subbituminous) in applications for which these processes appear to be most economically attractive (i.e., only 70% SO_2 removal is required). In fact most of the initial pilot plants listed for each company in Table 3 were built in order to qualify as bidders for the utility FGD systems listed in Table 2.

The Joy/Niro unit at Riverside Station listed in Table 3 is more correctly a 100-MW demonstration unit. This unit will have a single train consisting of a full-scale spray dryer (for a large plant multiple spray dryers of this size would be used). For particulate collection either an existing ESP or a baghouse can be used. The boiler is currently firing a blend of low-sulfur (<1%) Montana subbituminous coal and high-sulfur (4%) petroleum coke. Thus testing a full range of sulfur levels is potentially available.

On the following pages the history and current status, through June 1980, is discussed for each organization that is or has been active in spray dryer FGD technology. Most of the information was developed by direct contacts with company representatives and site visits during 1979 and 1980. It should be pointed out that the determination as to whether a spray dryer facility is considered a pilot-plant unit, a demonstration unit, or a commercial unit was based on the following arbitrary rules: (1) if a utility or company purchased the equipment for the FGD system, the unit was considered commercial unit regardless of its size; (2) other units $\leq$10 MW_e were considered as pilot-plant units; and (3) other units >10 MW_e but $\leq$125 MW_e were considered as demonstration units.

TABLE 2. CONTRACT AWARDS FOR SPRAY DRYER-BASED FGD SYSTEMS

Installation	Size, gross MW	Fuel type (% S)	SO_2 removal, %	Alkali raw material	Startup date	Vendor
Utility Boiler						
Coyote Unit 1	456	Lignite (0.78)	70	Soda ash	4/81	RI/WF[a]
Laramie River Unit 3	575	Subbituminous (0.54)	85	Lime	4/82	B&W[c]
Antelope Valley Unit 1	440	Lignite (0.68)	62	Lime	4/82	Joy/Niro[b]
Shiras Unit 3	44	Subbituminous (1.5)	80	Lime	9/82	Buell/Anhydro
Stanton Unit 2	63	Lignite (0.77)	73	Lime	9/82	R-C[d]
Craig Unit 3	447	Bituminous (0.70)	87	Lime	4/83	B&W
Holcomb Unit 1	319	Subbituminous (0.30)	80	Lime	6/83	Joy/Niro
Rawhide Unit 1	260	Subbituminous (0.29)	80	Lime	12/83	Joy/Niro
Springerville Unit 1	350	Subbituminous (0.69)	61	Lime	2/85	Joy/Niro
Springerville Unit 2	350	Subbituminous (0.69)	61	Lime	9/86	Joy/Niro
Industrial Boiler						
Strathmore Paper Co.	14[e]	Bituminous (2.0-2.5)	75	Lime	7/79	Mikropul
Celanese Fibers Co.	22[e]	Bituminous (1.0-3.5)	70-80	Lime	1/80	RI/WF
Calgon	17[e]	-	75	Soda ash	6/81	Niro/Joy
University of Minnesota	83[e]	Subbituminous (0.6-0.7)	70	Lime	9/81	Carborundum
Argonne National Lab.	29[e]	Bituminous (3.5)	80	Lime	9/81	Niro/Joy

Based on contact with vendors representing the status of announced contracts through October 1980.

a. Rockwell International/Wheelabrator-Frye.

b. Western Precipitation Division of Joy Manufacturing Company/Niro Atomizer, Inc.

c. Babcock & Wilcox.

d. Research-Cottrell.

e. Based on 2,900 aft^3/MW.

TABLE 3. SPRAY DRYER PILOT PLANTS AND DEMONSTRATION UNITS FOR FGD

Company	Size, kaft^3/min	Fuel type (% S)	Primary alkali tested	Operating period
Babcock & Wilcox				
Alliance Research Center	1.5	Various coals	Lime	Ongoing
W. J. Neal Station Unit 2 (Basin Electric)	8.0	Lignite (0.4)	Lime	June 1978-May 1979
Jim Bridger Unit 3 (Pacific Power and Light)	120.0	Subbituminous (0.6)	Lime	August 1979-?
Buell-Envirotech/Anhydro, Inc.				
Copenhagen Anhydro Laboratory	3.0	-[a]	Lime, soda ash	Ongoing
Martin Drake Unit 6 (City of Colorado Springs)	20.0	Subbituminous (0.5)	Lime, trona	December 1979-Fall 1980
Carborundum Environmental Systems				
Carborundum Knoxville Laboratory	0.1	-[a]	Lime, NH_3, $NaHCO_3$, and nahcolite	1976-1977
Carborundum Knoxville Laboratory	1.0	Bituminous (0.5)	Lime, Na_2CO_3, and fly ash	Ongoing
Leland Olds Station Unit 1 (Basin Electric)	15.0	Lignite (0.6)	Lime, NH_3, and soda ash	April 1978-October 1978

(continued)

TABLE 3. (continued)

Company	Size, kaft^3/min	Fuel type (% S)	Primary alkali tested	Operating period
Combustion Engineering				
Sherburne County Unit 1 (Northern States Power)	20.0	Subbituminous (1.0)	Lime	July 1979–January 1980
Gadsden Unit 1 (Alabama Power)	20.0	Bituminous (1.8)	Lime	May 1980–?
Ecolaire Environmental Corporation				
Gerald Gentleman Unit 1[b] (Nebraska Public Power)	10.0	Subbituminous (0.3)	Lime	January 1980–?
Joy Manufacturing/Niro Atomizer, Inc.				
Niro Laboratory Copenhagen	3.0	–[a]	Lime, MgO, and soda ash	Ongoing
Hoot Lake Unit 2 (Ottertail Power)	20.0	Lignite	Lime, soda ash	February 1978–April 1978 September 1978–December 1978
Riverside Station Units 6 & 7 (Northern States Power)	640.0	Subbituminous (1.0) Petroleum coke (4.0)	Lime	September 1980–August 1983

(continued)

TABLE 3. (continued)

Company	Size, $kaft^3/min$	Fuel type (% S)	Primary alkali tested	Operating period
Research-Cottrell, Inc.				
Big Brown Unit 2 (Texas Utilities)	10.0	Lignite (1.0)	Lime	June 1979-Early 1980
Comanche Unit 2 (Public Service of Colorado)	10.0	Subbituminous (0.5)	Lime	May 1980-February 1981
Rockwell International/Wheelabrator-Frye, Inc.				
Stork-Bowen Engineering Laboratory	5.0	-[a]	Lime, soda ash	Ongoing
Leland Olds Station (Basin Electric)	10.0	Lignite (0.6)	Lime	May 1977-September 1978
Joliet Station (Commonwealth Edison)	5.0	Subbituminous (0.5)	Lime	July 1979-July 1980
Sherburne County Unit 3[b] (Northern States Power)	5.0	Subbituminous (0.8)	Lime	May 1979-July 1979
Jim Bridger[b] (Pacific Power and Light)	5.0	Subbituminous (0.6)	Lime	August 1979-September 1979

Based on contacts with vendors and representing information through June 1980.
a. Propane burner with SO_2 spiking.
b. Mobile unit.

BABCOCK & WILCOX (29)

Babcock & Wilcox (B&W) entered the spray dryer FGD field in early 1977. The initial test work used a Hitachi, Ltd., (Japan) vertical spray dryer/reactor that has since been replaced by a horizontal reactor incorporating a modified B&W Y-jet nozzle atomizer. This use of a two-fluid nozzle atomizer is unique among the various spray dryer FGD vendors who have contracted for commercial units. Because of its developmental status, it has created some operating problems at B&W's pilot and demonstration units. B&W is also unique among commercial-unit vendors in recommending the use of an ESP for particulate collection for spray dryer FGD systems. B&W is also the only spray dryer vendor that builds all of the major equipment for their FGD system.

Background and Current Status of Development

Pilot-Plant Units--

William J. Neal Station--With the realization that spray dryer FGD systems were becoming an FGD alternative, particularly for western coals and lignites, B&W built a spray dryer pilot plant in late 1977 and early 1978 at Basin Electric Power Cooperative's William J. Neal Station Unit 2. Unit 2 is a 20-MW pulverized coal boiler burning North Dakota lignite that averages 0.4% sulfur and has a 7,900 Btu/lb heating value. The flue gas conditions at this boiler were similar to the projected design conditions for the Antelope Valley Unit 1 boiler (which B&W subsequently bid) and, in fact, this pilot plant was built primarily to demonstrate B&W's spray dryer FGD system and thereby qualify to bid on the FGD system for Antelope Valley.

Initially the pilot plant used a Hitachi, Ltd., vertical spray dryer/reactor but the test results did not appear promising. Instead of continuing with the Hitachi spray dryer/reactor, B&W modified a Y-jet nozzle (which B&W had originally designed for use in an oil-fired boiler) for use as an atomizer for their spray dryer/reactor. The atomizers were arranged in a matrix on one wall of a horizontal reactor. Flue gas entered through registers (again similar to a boiler design) around each nozzle.

This modified spray dryer/reactor, rated at 8,000 aft^3/min, was used for the remainder of the test program. Flue gas from the spray dryer/reactor was then passed through an ESP for particulate collection. An SO_2 spiking system and lime preparation equipment were also included. Testing with this system continued from June 1978 until May 1979. During these tests three different coals were fired in the Unit 2 boiler. In addition to the North Dakota lignite normally burned, a Wyoming subbituminous coal from the Powder River Basin (0.5% sulfur) and a Montana subbituminous coal (1.2% sulfur) were used. (The Wyoming subbituminous coal is similar to that which will be burned in the Laramie River Station Unit 3 where B&W ultimately was awarded a contract for the spray dryer FGD system.) Although lime was the chosen absorbent and was used for most of the tests, other alkalis were also evaluated. These included hydrated lime, limestone, soda ash, magnesia, and ammonia. Even though all three coals burned at the Neal Station produce an alkaline fly ash, B&W apparently did not run tests using waste recycle. The pilot plant was dismantled in mid-1979.

Alliance Research Center--With the closing of the pilot plant at the William J. Neal Station, B&W constructed a small pilot plant at their Alliance Research Center to continue the development and testing of their process. This pilot plant uses a small coal-fired combustor producing 1,500 aft^3/min of flue gas. In contrast to the large units using multiple arrays of nozzle atomizers, the spray dryer/reactor at Alliance has a single nozzle atomizer. A cyclone is used for particulate collection. An ESP and a baghouse may be evaluated in the future.

This pilot plant is being used to test a range of spray dryer/reactor operating conditions, initially for the lime-based system but probably for several alternative absorbents later. B&W also intends to use the pilot plant to increase the understanding of the SO_2 removal reaction mechanism and therefore determine how they can optimize their reactor system. In addition, the performance of the lime-based FGD system will be monitored while about five different coals are burned. These coals include North Dakota lignite, western subbituminous, as well as eastern low-sulfur, medium-sulfur, and high-sulfur coals.

Demonstration Units--

Jim Bridger Station--Early in 1979 B&W, Pacific Power & Light, and Idaho Power & Light reached an agreement under which B&W built a demonstration-size reactor for treating 120,000 aft^3/min of the flue gas from Jim Bridger Station Unit 3. Unit 3 is a 500-MW boiler burning a Wyoming subbituminous coal that averages 0.6% sulfur and has a heating value of 9,500 Btu/lb. Both B&W and the owners of Jim Bridger were interested in demonstrating the spray dryer technology. B&W was interested in demonstrating (and obtaining the data necessary for proving) their full-scale spray dryer/reactor design. The owners of Jim Bridger, on the other hand, have a power unit that must be retrofitted with an FGD system. Since this boiler unit burns a low-sulfur coal and has a relatively high-alkaline fly ash, a spray dryer FGD system appeared to be an attractive alternative.

The demonstration unit treats about 120,000 aft^3/min in a single, spray dryer/reactor containing six nozzle atomizers. Most of the flue gas from the reactor reenters the boiler ductwork and passes through the existing boiler ESP's. However, a 5,000 aft^3/min slipstream is routed through a pilot baghouse for particulate collection. The lime preparation system uses a closed-loop, wet ball mill slaker.

Design and construction of this demonstration unit was completed in July 1979, and it was started up in August. Although the demonstration plant has been able to follow boiler load changes without any significant problems, there have been some operating problems in the spray dryer/reactor. Troubles with poor atomization and flue gas distribution have led to wet operation, plugging of the atomizers, and inability to closely approach flue gas saturation temperatures. B&W has recently switched from steam to compressed air for the atomizing fluid to alleviate some of these operating problems. Since this demonstration plant was intended to demonstrate the design for B&W's commercial FGD system at Basin Electric Power Cooperative's Laramie River Station Unit 3, any significant design changes at Jim Bridger will probably be incorporated into the Laramie River design.

B&W is now testing only lime absorbent although they hope to evaluate ammonia, soda ash, and a sodium-based waste liquor in the future. Tests for the lime-based slurry were originally projected to continue for 4 to 6 months (i.e., until January 1980) however, atomization problems have delayed the schedule somewhat.

Commercial Units--

Laramie River Station--In November 1978 B&W was awarded a contract for a spray dryer FGD system for Basin Electric Power Cooperative's Laramie River Station Unit 3. Unit 3, currently scheduled for startup in April 1982, will be a 575-MW boiler burning a low-sulfur Wyoming subbituminous coal from the Powder River Basin that averages 0.54% sulfur and has a heating value of 8,000 Btu/lb. The FGD system is designed to treat 2,300,000 aft^3/min (286°F) containing 530 ppm SO_2. The system is designed and guaranteed for 85% SO_2 removal under average coal conditions and 90% SO_2 removal for the worst-case coal sulfur level (i.e., designed to meet the stringent Wyoming SIP of 0.20 lb SO_2/MBtu heat input to the boiler).

The FGD system for the Laramie River Station Unit 3 will consist of four spray dryer/reactors, each with 12 nozzle atomizers arranged in a 3 by 4 array. Although originally designed with steam as the atomizing fluid, after the recent test work at Jim Bridger the atomizing system has been modified to use compressed air. Under full-boiler load only three of the spray dryer/reactors will be operating with the other as an in-line spare. Each reactor will be followed by an ESP (i.e., four parallel trains of reactors and ESP's) for particulate collection.

Although this unit was originally designed without provision for waste recycle, B&W has modified its design philosophy and this installation will now use waste recycle. The FGD system is also designed for hot gas bypass. Approximately 3% of the total flue gas from the boiler economizer will be bypassed around the spray dryer/reactor to achieve a 15°F to 20°F reheat. Waste disposal will be in an existing FGD sludge pond. (Units 1 and 2 at the Laramie River Station have limestone slurry FGD systems with pond disposal.)

Craig Station--In March 1980, B&W was awarded a contract for a lime spray dryer system for Colorado-Ute Electric Association's Craig Station Unit 3. Unit 3 will be a 447-MW boiler burning a Colorado bituminous coal that averages 0.7% sulfur and has a heating value of 10,000 Btu/lb. The FGD system is designed for 87% SO_2 removal under average coal conditions. Projected startup date is April 1983.

The FGD system will be designed essentially the same as the one for Laramie River Unit 3. The system will consist of four spray dryer/reactors, each with 12 nozzle atomizers. As at Laramie River, three reactors will be operating and one will be an in-line spare. At the utilities request particulate removal will be achieved by using a baghouse after the reactors. FGD waste recycle will be used and hot gas bypass provisions will be incorporated.

Conceptual Design

The following conceptual design (see Figure 6) for a 500-MW coal-fired boiler is based on B&W's recent full-scale contract awards. The design assumes a lime system for a low-sulfur western coal application requiring a 70% SO_2 removal efficiency.

Flue gas from the boiler air heaters at about 300^{o}F enters a common plenum that distributes the gas to the four trains of spray dryer/reactors. During normal full-boiler load operation only, three trains are operating and the fourth is an in-line spare. Each reactor has 12 horizontal nozzle atomizers arranged in a 3 by 4 array. The atomizing fluid is compressed air.

B&W's original design philosophy was to operate the spray dryer/reactor so that the flue gas closely approached its saturation temperature, and thereby maximizing SO_2 removal efficiency and absorbent utilization. However, B&W has changed its design so that the approach to saturation is now 15^{o}F to 20^{o}F and some type of waste recycle will be included in their spray dryer FGD systems.

Each spray dryer/reactor is followed by an ESP which collects the FGD waste (B&W will provide a baghouse if it is specified). Although some additional SO_2 removal is usually achieved in a baghouse and an ESP probably would not provide this additional removal, the ESP is claimed to be more adaptable to upset conditions, and there is also more experience with operating an ESP than a baghouse. Flue gas from the ESP passes through an ID fan and into the stack plenum. FGD waste from the ESP hoppers is periodically moved to intermediate storage hoppers before being trucked to a landfill for disposal.

Pebble lime is received by rail and stored in concrete silos onsite. This lime is removed as needed, conveyed to a ball mill slaker, slurried, and pumped to the spray dryer.

Technical Considerations

The B&W spray dryer/reactor is based on using their Y-jet slurry atomizer, shown in Figure 7, which was originally developed for use as a fuel oil atomizer in oil-fired power units. Although this nozzle-type atomizer was to use steam as the atomizing fluid, recent test work has indicated that compressed air gives better results. Each of the full-scale spray dryer/reactors is designed with 12 of these Y-jet slurry atomizers arranged in a 3 by 4 array. The atomizers are mounted horizontally in one wall of the spray dryer/reactor (see Figure 8). Flue gas

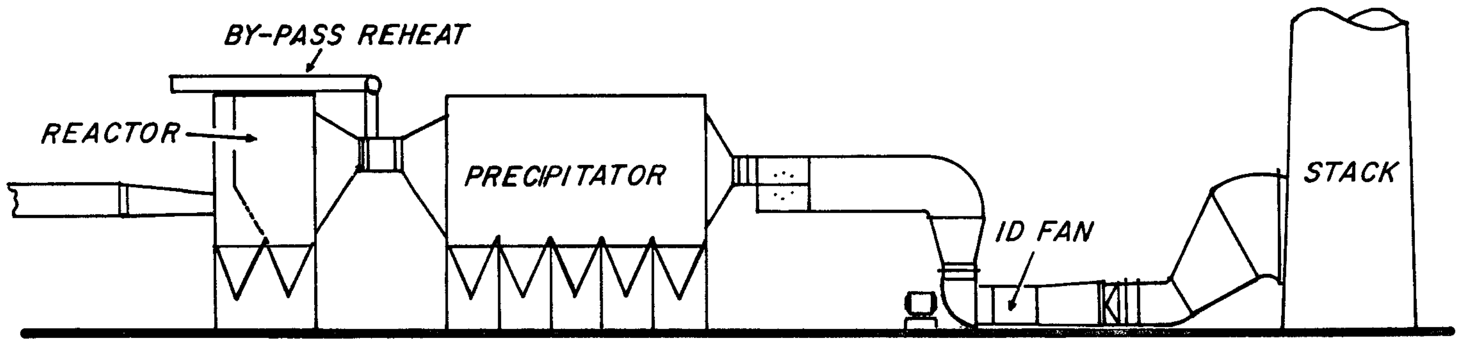

Figure 6. Conceptual design for the Babcock & Wilcox spray dryer FGD process (9).

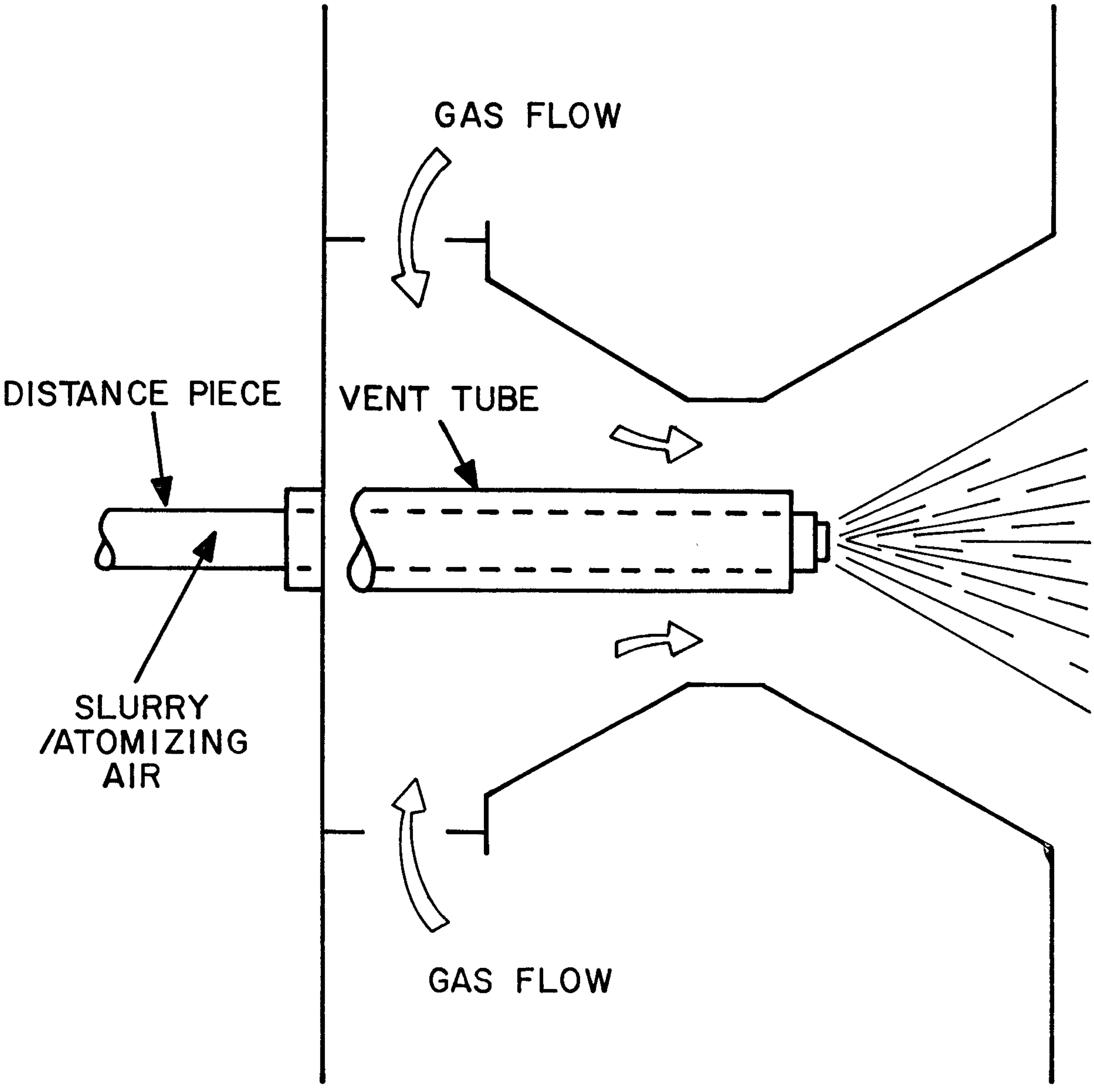

Figure 7. Design of two-fluid nozzle atomizer for the Babcock & Wilcox spray dryer FGD process (9).

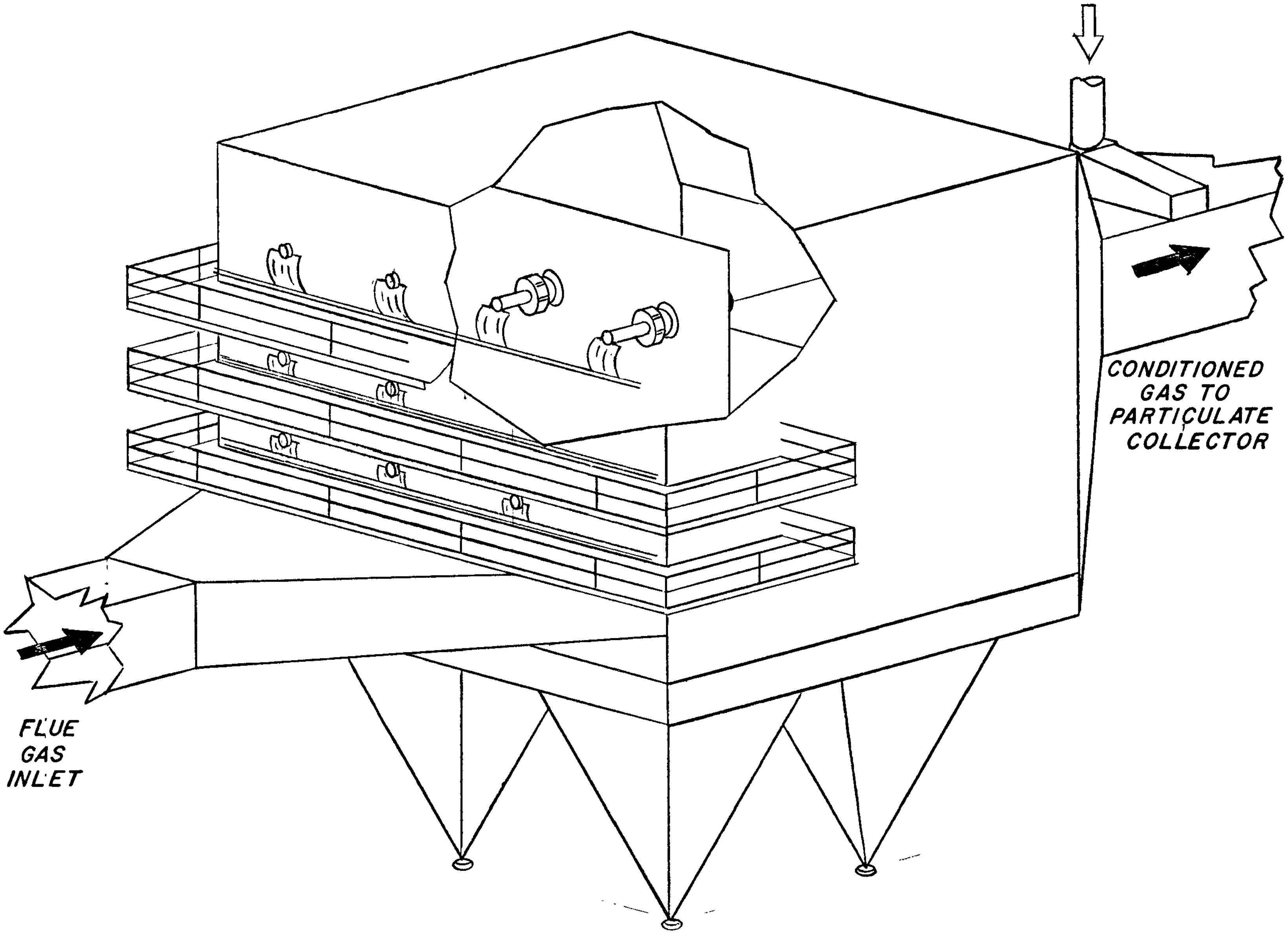

Figure 8. Spray dryer/reactor design for the Babcock & Wilcox FGD process (29).

enters a manifold surrounding the atomizer array and passes out around the barrel of the atomizer. Intimate mixing of the slurry and flue gas occurs just inside the spray dryer/reactor wall and continues throughout the length of the reactor. The B&W design philosophy is to closely approach (within 20^{o}F) the flue gas saturation temperature to obtain a high SO_2 removal efficiency and absorbent utilization in the spray dryer/reactor.

B&W is using an ESP for particulate collection in their FGD system even though baghouses tend to increase SO_2 removal as the flue gas passes through the absorbent-based waste coating the bags. ESP's are not likely to achieve as much additional SO_2 removal since most of the flue gas passing through the ESP does not come into intimate contact with the FGD waste. However, since the amount of SO_2 removed in the baghouse is reduced at higher absorbent utilization (which B&W claims for their spray dryer/reactor), B&W has decided to use an ESP for particulate collection in most applications. The other reason for using the ESP is that it is thought to be more tolerant of upset conditions in the spray dryer/reactor.

The ESP can be used even in low-sulfur coal applications where fly ash resistivity has been a problem because the spray dryer FGD system is claimed to condition the flue gas. There are several reasons for believing that this is the case for most applications. The spray dryer cools and humidifies the flue gas. Since the resistivity curve tends to fall off from its peak at the air heater temperature, cooling the flue gas should enhance the collection efficiency. Humidifying the flue gas also tends to increase precipitator efficiency. The third reason for better than expected collection efficiency in the ESP handling fly ash from low-sulfur coal is that not only does the FGD waste tend to produce larger diameter particles, they may trap fly ash particles which are captured by the lime slurry.

B&W also designs their lime slaking system with a ball mill slaker. The ball mill slaker is used to grind the pebble lime before the actual slaking occurs. This tends to result in very fine lime particles which indicates the emphasis in the B&W system is on maximizing the absorbent utilization in the first pass through the spray dryer/reactor.

BUELL-ENVIROTECH/ANHYDRO, INC. (30)

The Buell Emission Control Division of Envirotech Corporation and Anhydro, Inc., of Copenhagen, Denmark, are currently developing a spray dryer FGD system as a joint venture. Buell is a designer and marketer of particulate control equipment (fabric filters, ESP's, and cyclones) while Anhydro is a designer and marketer of spray dryers. Together these companies have a background in the two major types of equipment for spray dryer FGD systems. Buell will be the actual bidder on any spray dryer FGD system and will subcontract the spray dryers to Anhydro under their exclusive agreement.

Background and Current Status of Development

Pilot-Plant Units--

Copenhagen Laboratory--Initial pilot-plant testing of the spray dryer as an FGD absorber was completed in Anhydro's Copenhagen laboratory. This test facility has a 6-1/2-ft-diameter spray dryer treating a simulated flue gas from a propane burner. The spray dryer is rated at 3,000 aft^3/min and has a single, direct-drive, rotary atomizer.

The first FGD tests consisted of injection of dry lime into the flue gas ducts and then dry soda ash as a second attempt; neither gave good SO_2 removal at acceptable raw material utilization rates. This, of course, led to the use of the spray dryer.

Martin Drake Station--During 1979 Buell and Anhydro reached agreement with the City of Colorado Springs to install a pilot plant at Unit 6 of the Martin Drake Station. In addition to the Buell/Anhydro spray dryer work, Buell is also doing some nahcolite injection work for EPA at Martin Drake at the same time (25). Unit 6 is an 85-MW pulverized-coal-fired boiler and burns a mixture of three Colorado bituminous coals which average about 0.5% sulfur and have heating values of about 12,000 Btu/lb. It also has an existing baghouse, designed and built by Buell. Startup of this pilot plant began in December 1979, and the spray dryer operations are funded for at least six months.

This test unit has a 12-1/2-ft-diameter spray dryer (rated at 20,000 aft^3/min) containing a single, rotary atomizer. It treats a slipstream from the boiler and uses a full-size test baghouse (15,000 aft^3/min) for particulate collection. In situ-resistivity tests were also performed to evaluate an ESP for particulate collection. An SO_2 spiking system was installed to allow tests with up to 2,500 ppm SO_2 in the inlet flue gas. Absorbents evaluated in the spray dryer system include limestone and adipic acid, trona, lime, and dolomitic lime. Various recycle schemes are currently being evaluated for commercial applications. A detention slaker is used for the lime slurry generation. The effects of bypass reheat, varied absorber inlet temperatures, and prequenching before the absorber inlet have been tested.

Demonstration Units--

No plans have been announced as yet for a demonstration unit.

Commercial Units--

Shiras Station--In October 1980 Buell-Envirotech/Anhydro, Inc., received a contract to provide a spray dryer FGD system for the City of Marquette's Shiras Station Unit 3. This unit is a 44-MW pulverized-coal-fired boiler burning a western subbituminous coal averaging about 1.5% sulfur and having a heating value of 7,700 Btu/lb. The FGD system is designed to treat 226,100 aft^3/min (at 265°F) of flue gas containing 1,500 ppm SO_2. Design SO_2 removal efficiency is 80%. The projected startup date for the FGD system is September 1982.

The system will contain a single 35-1/2-ft-diameter spray dryer. The spray dryer will have one rotary atomizer. A single eight compartment reverse air baghouse will be used to collect the FGD waste and fly ash. FGD waste recycle and hot gas bypass will be incorporated into the system. A paste-type slaker will be used with lime as the alkali raw material. Waste disposal will be handled by the City of Marquette.

Conceptual Design

The following conceptual design for a 500-MW coal-fired boiler is based on the pilot-plant data currently available. As their process becomes further developed, this process design may undergo some minor design modifications, but the overall process description equipment, and operating conditions are not expected to undergo any significant change. The design is based on a lime system for a low-sulfur western coal application requiring a 70% SO_2 removal efficiency.

Flue gas from the boiler air heaters at about 300°F enters a common plenum that distributes the gas to the six spray dryers. During normal full-boiler-load operations all six trains are operating. Each spray dryer reactor contains a single rotary atomizer. The spray dryers are insulated and enclosed in a simple, shell-type building. The six spray dryers feed a single baghouse. The flue gas from the baghouse passes through four ID fans and into the stack plenum.

The compartments in the baghouse are periodically emptied, and the waste is pneumatically conveyed to storage hoppers. No data have been released on the method of waste recycle to be used in the Buell/Anhydro system (i.e., where the waste will be removed, or how the waste will be recycled).

Pebble lime is received by rail and stored in concrete silos onsite. This lime is removed as needed, conveyed to a slaker, slurried and pumped to the spray dryer as needed. In general, a paste slaker will be used.

Technical Considerations

The Anhydro spray dryer contains a single rotary atomizer regardless of how large the spray dryer is.Although this rotary atomizer is relatively large and must handle significant quantities of lime slurry, this design (one atomizer/spray dryer) is typical of non-FGD applications. Flue gas enters the spray dryer through a scroll-type duct and passes in a concentric ring around the single atomizer. Flue gas leaves the side of the spray dryer after making a 90-degree turn. This design causes a decrease in the flue gas velocity at the duct and most of the larger particles entrained in the flue gas drop out in the bottom of the spray dryer. A rotary valve at the base of the spray dryer allows the removal of this fly ash and FGD waste for recycle. This design also allows for the rapid cleaning of the spray dryer during any upset conditions when the particulate matter remains wet at the flue gas exit in the spray dryer. Since the particulate matter would be heavier when wet than when dry, it would have a tendency to fall out at the base of the spray dryer rather than passing on to the baghouse.

Although the waste collected in the baghouse can be recycled, most of the waste used in the recycle stream comes from the spray dryer. The recycle waste is pneumatically conveyed to the slurry mixing tank where makeup water is added. This resulting slurry is pumped through a classifier (to separate the larger particles) to the spray dryers.

Although the standard Anyhdro spray dryer for utility applications has not been specified, each application normally consists of multiples of a standard size. The desired gas residence time (10 to 12 seconds) is obtained by varying the height of the drying chamber. Each spray dryer is insulated and the spray dryer area may be enclosed in a simple, shell-type building. The design usually includes one baghouse with a design air-to-cloth ratio (gross) of 2:1.

CARBORUNDUM ENVIRONMENTAL SYSTEMS (31)

Carborundum Environmental Systems is a subsidiary of Kennecott Copper Corporation and is based in Knoxville, Tennessee. During the initial development and pilot-plant testing of their spray dryer-based FGD system, Carborundum had an exclusive agreement with the De Laval Separator Company to supply the spray dryers. However, this agreement has recently been terminated, and Carborundum has recently (1979) signed a licensing agreement with Kochiwa Kakohki Company, Inc., a Japanese spray dryer manufacturer. The spray dryers for the Carborundum system will be manufactured in the United States, while the rotary atomizers may be manufactured in either Japan or the United States. Baghouses for the spray dryer FGD system will be designed and built by Carborundum.

Background and Current Status of Development

Pilot-Plant Units--

Much of the development work for Carborundum's spray dryer FGD system was done at a 100 ft^3/min bench-scale unit at their test facility, which was built in 1976, at the University of Tennessee in Knoxville. Although initially designed to evaluate ammonia, sodium bicarbonate, and nahcolite injection for FGD, it was later used to develop the lime spray dryer FGD system in late 1977.

Leland Olds Station--Initial pilot-plant work was done at Unit 1 of Basin Electric Power Cooperative's Leland Olds Station. Unit 1 is a 215-MW cyclone boiler burning a North Dakota lignite which typically averages 0.6% sulfur and has a 7,000 Btu/lb heating value. This pilot plant started up in the spring of 1978 and operated approximately six months.

The test unit typically treated a slipstream of about 7,000 to 9,000 aft^3/min in an 8-ft-diameter spray dryer, although both the spray dryer and the baghouse were sized to handle 15,000 aft^3/min. The pilot unit also contained an SO_2 spiking system. Primary absorbents evaluated during the test program were lime and soda ash.

The goals of this test program were twofold. The initial goal was to complete tests to optimize the spray dryer system. Parameters evaluated included inlet SO_2 concentration (600 to 2,500 ppm), inlet flue gas temperature (275°F to 350°F), degree of approach to flue gas saturation temperature, and raw material stoichiometric ratio. Tests with recycled waste were also made. The other goal for the pilot-plant tests at the Leland Olds Station was to qualify to bid on the FGD system for Basin Electric's Antelope Valley Station.

University of Tennessee--Additional pilot-plant work of the spray dryer FGD system will be completed with a 1,000 aft^3/min unit at a University of Tennessee laboratory facility in Knoxville. This pilot plant will treat a slipstream from a spreader stoker boiler burning an eastern bituminous coal that averages 0.5% sulfur and has a 10,000 Btu/lb heating value. Startup date for this pilot unit was June 1980, and the test program will continue at least through 1980.

Primary absorbents to be evaluated include lime, soda ash, and fly ash. Additional testing is also expected, but since this development work is being internally funded, the actual test program and any results are considered proprietary information.

Demonstration Unit--

No plans for a demonstration unit have been announced by Carborundum.

Commercial Unit--

University of Minnesota--During the first quarter of 1980 Carborundum Environmental Systems received a contract to provide the FGD system for two coal-fired boilers at the University of Minnesota. This will be a retrofit installation on existing boilers that are being reactivated as part of a cogeneration system to provide both steam and electricity for the University. These boilers burn a western coal averaging 0.6% to 0.7% sulfur (maximum sulfur is 0.73%) and having a heating value of about 9,500 Btu/lb. Each boiler generates 120,000 aft^3/min (at 375°F) of flue gas containing 630 ppm SO_2. The FGD system is designed to meet the NSPS (i.e., 70% overall SO_2 removal and 0.03 lb of particulate matter/MBtu heat input) and is currently scheduled for startup in September 1981.

Each of the boilers will have a separate, but similar, FGD system. One FGD system will have a 24-1/2-ft-diameter spray dryer containing a single rotary atomizer. The other will have a 27-1/2-ft-diameter spray dryer containing 3 rotary atomizers. Flue gas from these spray dryers will not be reheated before entering the baghouses. No provisions have been included for waste recycle. Pebble lime will be prepared in a single lime preparation area (serving both FGD systems). A paste-type slaker will be used. Waste disposal will be handled by the University of Minnesota.

Conceptual Design

Although Carborundum has not been awarded any contracts for a full-scale commercial utility FGD system, the following process description for a 500-MW coal-fired power unit is based on their conceptual design (Figure 9) and early pilot-plant experience. Additional pilot-plant experience in the future may therefore cause some minor design modifications, but the overall process description, equipment, and operating conditions are not expected to undergo any significant change.

Flue gas from the boiler air heater at approximately 300°F enters a common plenum which distributes the gas to the four trains of spray dryers (three operating and one spare). The flue gas enters the spray dryer from the top and passes downward around each of the rotary atomizers. (Each spray dryer has three rotary atomizers.) As the flue gas passes the rotating atomizer, the lime slurry is sprayed as a fine mist into the hot gas. The atomized slurry and flue gas remain intimately mixed throughout their residence time in the spray dryer, where the SO_2 and HCl are absorbed in the droplets and react with the lime slurry. The water present in the spray dryer evaporates and the mixed calcium-sulfur

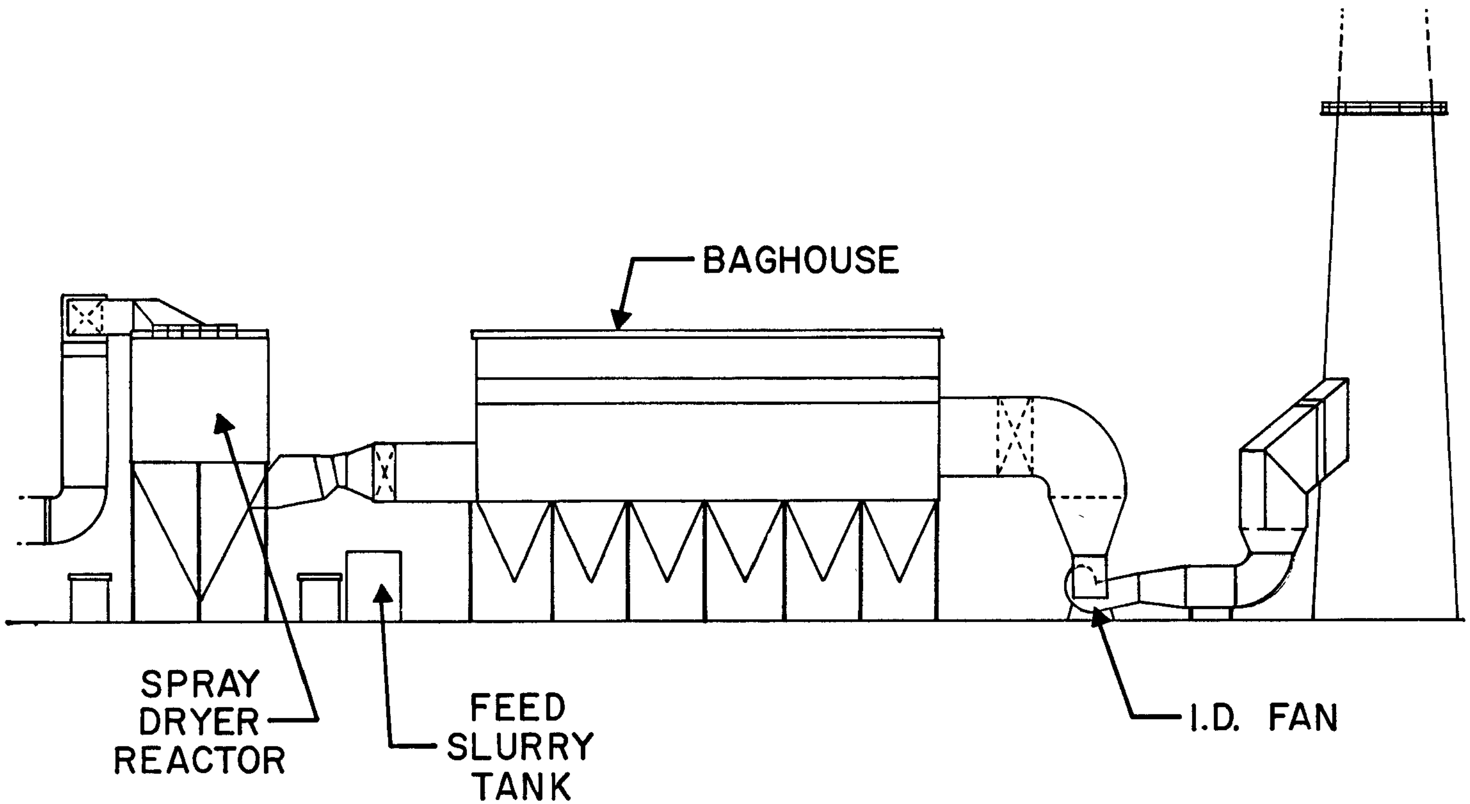

Figure 9. Conceptual design for the Carborundum spray dryer FGD process (31).

salts and the fly ash flow to the bottom of the spray dryer. At the bottom of the spray dryer, the flue gas makes a 180-degree turn. As the superficial gas velocity decreases, the larger particles of entrained particulate matter drop into a hopper from which they are periodically removed. The flue gas passes to the baghouse where the remaining solids are removed and then through an ID fan to the stack. The fly ash and FGD waste from the spray dryer and the baghouse are pneumatically conveyed to a temporary storage silo from which it is trucked to a disposal area.

Pebble lime is received by truck or rail and is stored onsite in a silo. Lime is removed daily to fill the intermediate feed bin which in turn feeds the ball mill slakers. The pebble lime is slaked and pumped first to the mills product tank and then to the slurry feed tank. The makeup lime slurry is pumped continuously from the slurry feed tank to the rotary atomizers in the spray dryer. Dilution water is also provided continuously to the rotary atomizers.

Technical Considerations

The Kochiwa Kakohki spray dryer is designed with three rotary atomizers for full-scale utility applications. The use of three rotary atomizers decreases the size of each atomizer since each must atomize only one-third of the total absorbent fed to the spray dryer. This design is claimed to result in a better mix of flue gas and atomized absorbent and thus lead to better raw material utilization and a better temperature profile through the spray dryer. However, this use of multiple atomizers also complicates the flue gas distribution to the atomizers.

Another claimed advantage for using multiple atomizers is that when they are designed properly and have sufficient excess capacity, a spare spray dryer is not necessary. The vendors believe that if one of the rotary atomizers fails, the other two can maintain sufficient feed to, and mixing in, the spray dryer so that SO_2 removal will not be seriously affected for short periods of time (i.e., until the boiler load declines and the spray dryer can be taken off-line). They also believe that the rotary atomizer can be changed while the spray dryer is on-line and operating, if necessary. Unfortunately (since the cost of the spare spray dryer and its associated equipment is a fairly significant component of the total capital investment), design of an FGD system without a spare in-line absorber is not a totally accepted practice in the utility industry.

COMBUSTION ENGINEERING (32)

Although Combustion Engineering has been installing limestone-based FGD systems for several years, they first entered into the development of spray dryer FGD systems in June 1978 with their first conceptual design. Engineering design and planning continued throughout the remainder of 1978 and construction of their first pilot-plant unit began in February 1979. During 1979 an exclusive agreement was reached with James Howden Holima BV (The Netherlands) for the use of their baghouse technology. Combustion Engineering will use their own spray dryer technology.

Background and Current Status of Development

Pilot-Plant Units--

Sherburne County Station--Initial pilot-plant work on Combustion Engineering's spray dryer FGD system was begun in June 1979 with the construction of a 20,000 aft^3/min test unit at the Sherburne County Station of Northern States Power Company. This pilot plant was started up in July 1979 and treated a slipstream from Unit 1. This unit is a 700-MW pulverized-coal-fired boiler burning a Montana subbituminous coal averaging about 1.0% sulfur and having a heating value of 8,500 Btu/lb. The pilot plant was shut down in January 1980.

This pilot plant consisted of a 9-ft-diameter spray dryer followed by a small (about 1,500 aft^3/min) baghouse and ESP (about 7,000 aft^3/min) operating in parallel. (Thus, the pilot unit was effectively limited to 8,500 aft^3/min.) The spray dryer contained a single two-fluid nozzle atomizer with air as the atomizing fluid. An SO_2 spiking system and lime preparation equipment were also included in the pilot plant. The major purposes of this pilot plant were to evaluate operating conditions and to evaluate the two types of particulate control devices. Parameters evaluated include stoichiometric ratio, inlet SO_2 concentration, flue gas temperature, and waste recycle. The primary absorbent tested was lime.

Gadsden Station--A second pilot plant was recently built at Alabama Power Company's Gadsden Station. This pilot plant, which started up during August 1980, treats a slipstream of about 20,000 aft^3/min from Unit 1 of the Gadsden Station. This unit is a 69-MW boiler burning an Alabama bituminous coal averaging about 1.8% sulfur and having a heating value of 12,500 Btu/lb. No shutdown date has been announced. This pilot unit has the same 9-ft-diameter spray dryer used at Sherburne County followed by a baghouse for particulate collection. The primary absorbent being evaluated is lime. Tests at higher SO_2 levels ($\leq$2,000 ppm) will be used to confirm the results achieved at Sherburne County. Additional test work will evaluate solids recirculation and nearness of the approach to flue gas saturation.

Demonstration Units--

Although no startup date has been established and no detailed design work has been completed, Combustion Engineering plans to build a 100,000 aft^3/min (about 30-MW) demonstration unit early in 1981. The specific site for this demonstration unit has not been announced. The demonstration unit will contain a 20-ft-diameter spray dryer. Particulate collection will be provided by an existing boiler baghouse. Other equipment will include a lime preparation system (silos, ball mill slaker, tanks, and pumps). The primary purpose of this demonstration unit is to test equipment components and to confirm the operation of the system on a larger scale. This demonstration unit will be a module about one-third the size that would be used in a full-scale utility application.

Commercial Units--

No contracts have been awarded to Combustion Engineering for a commercial, full-scale utility spray dryer FGD system at the present time. Combustion Engineering has, however, qualified and submitted bids on three full-scale utility systems.

Conceptual Design

The following conceptual design for a 500-MW coal-fired boiler is extrapolated from data recently published by Combustion Engineering for a 250-MW unit. The design is based on a lime system for a low-sulfur western coal application requiring a 70% SO_2 removal efficiency.

Flue gas from the boiler air heater enters a common plenum which distributes the gas to the four trains of spray dryers (three of which are operating and the other is a spare). The flue gas enters the top of the spray dryer and passes through the swirl vanes surrounding the nozzle atomizers in a concentric ring. Each spray dryer contains four nozzle atomizers. The atomizing fluid is compressed air. The spray dryers are insulated and enclosed in a simple, shell-type building. The spray dryers feed a single baghouse designed for an air-to-cloth ratio (gross) of 1.8:1. The flue gas from the baghouse passes through four ID fans and into the stack.

The compartments in the baghouse are periodically emptied and the waste is pneumatically conveyed to storage hoppers. The spray dryers are designed such that the larger particles drop out of the flue gas in the spray dryer. This material from the spray dryer and part of the material collected in the baghouse are available for recycle if recycle is economically justified.

Pebble lime is received by rail and stored in concrete silos onsite. This lime is removed as needed, conveyed to a ball mill slaker, slurried, and pumped to the spray dryer as needed.

Technical Considerations

The Combustion Engineering spray dryer (Figure 10) typically has four nozzle-type atomizers (Figure 11) and uses compressed air as the atomizing fluid. These atomizers are mounted in the top of the spray dryer. Flue gas enters the top of the spray dryer and passes down around the atomizer through a concentric swirl ring. The swirl ring provides more intimate mixing of the flue gas and slurry in the spray drying chamber.

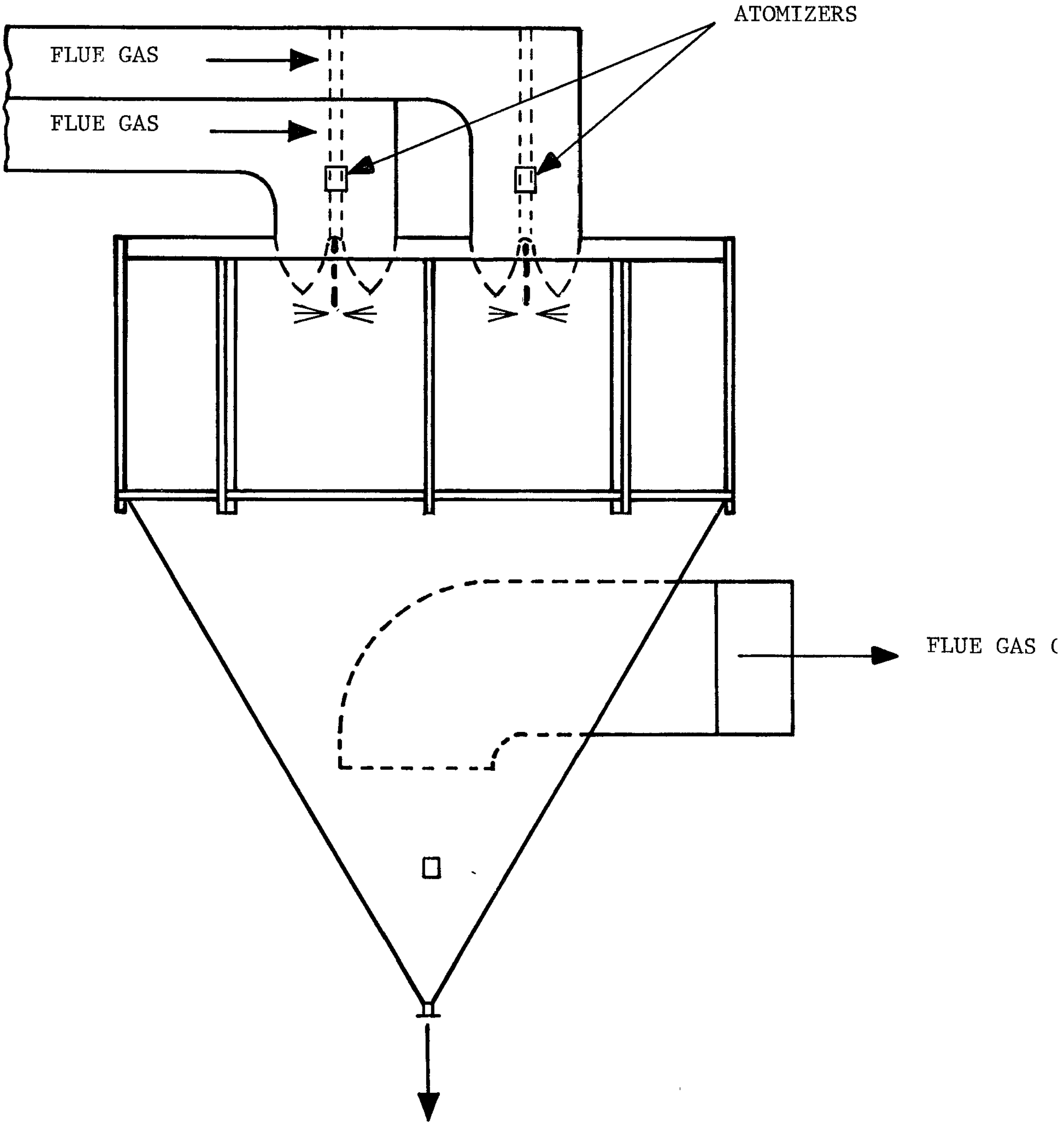

Figure 10. Spray dryer design for the Combustion Engineering FGD process (10).

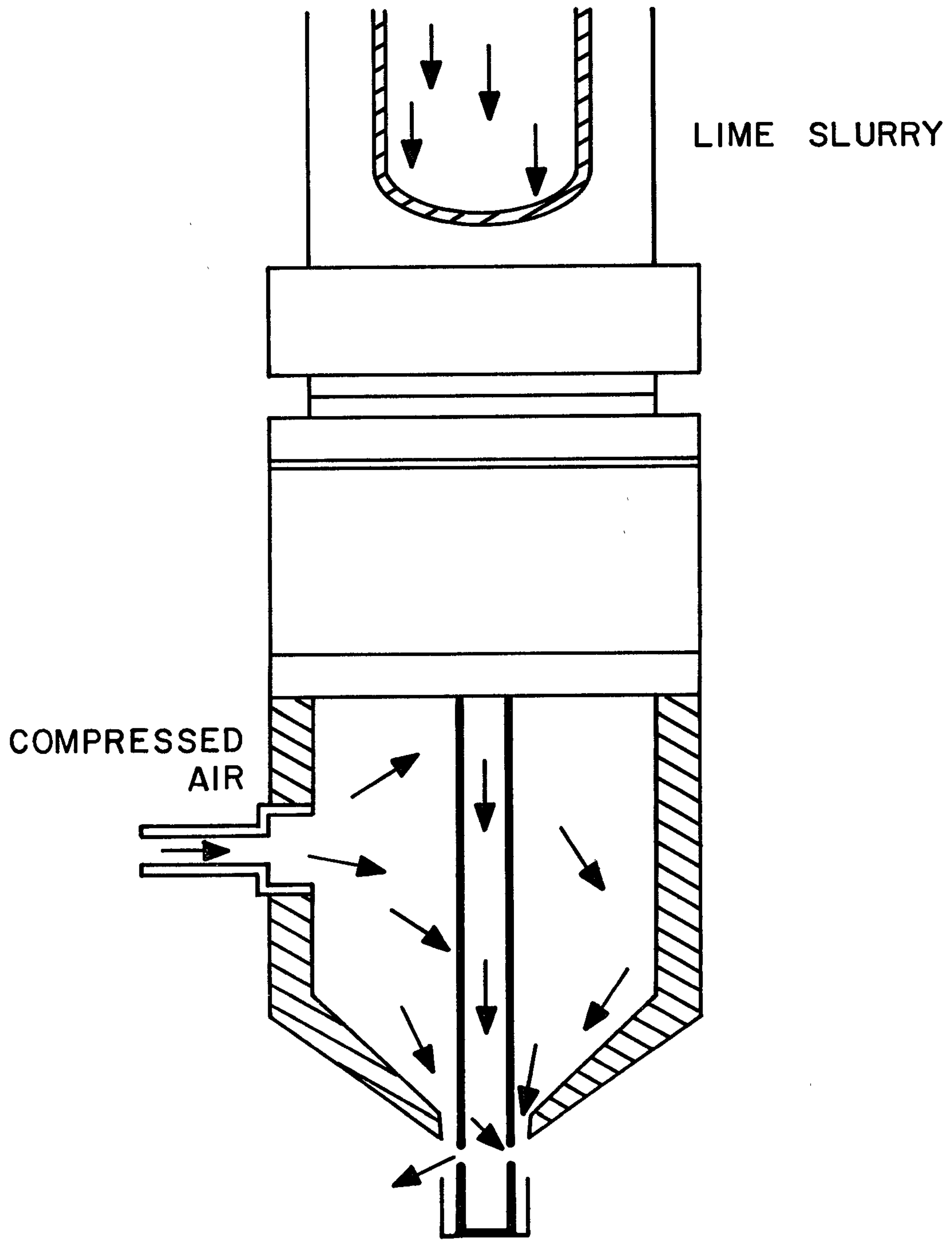

Figure 11. Design of two-fluid nozzle atomizer for the Combustion Engineering spray dryer FGD process (10).

An advantage claimed for using multiple atomizers is that when they are designed properly and have sufficient excess capacity, a spare spray dryer is not necessary. The vendors believe that if one of the nozzle atomizers fails, the others can maintain sufficient feed to and mixing in the spray dryer so that SO_2 removal will not be seriously affected for short periods of time (i.e., until the boiler load declines and the spray dryer can be taken off-line). They also believe that the nozzle atomizer can be changed while the spray dryer is on-line and operating, if necessary.

Flue gas leaving the spray dryer must make a 180-degree turn near the base of the spray dryer to enter the outlet duct. This design causes a decrease in flue gas velocity at the duct and most of the larger particles entrained in the flue gas drop out in the bottom of the spray dryer. A rotary valve at the base of the spray dryer allows the removal of this fly ash and FGD waste for recycle. This design also allows for the rapid cleaning of the spray dryer during any upset conditions when the particulate matter remains wet at the flue gas exit in the spray dryer. Since the particulate matter would be heavier when wet than when dry, it would have a tendency to fall out at the base of the spray dryer rather than passing on to the baghouse.

Although the waste collected in the baghouse can be recycled, most of the waste used in the recycle stream comes from the spray dryer. The recycle waste is pneumatically conveyed to the slurry mixing tank where it is added to the makeup lime slurry from the slaker. The resulting combined slurry is pumped to the spray dryers.

ECOLAIRE ENVIRONMENTAL CORPORATION (33)

Ecolaire Environmental Corporation is the subsidiary of Ecolaire Incorporated which markets Ecolaire's spray dryer FGD process. Other Ecolaire companies have been supplying equipment to the electric utility industry for many years. One is Ecolaire Environmental Company (EEC), formerly Industrial Clean Air, which designs and builds baghouses for the electric utility industry. This in-house baghouse experience could be an advantage for the Ecolaire spray dryer FGD system.

Background and Current Status of Development

Pilot-Plant Units--

Although Ecolaire's EEC subsidiary has been designing and installing baghouses for the utility industry for several years, prior to the design and construction of their mobile demonstration unit (MDU) in 1979, Ecolaire had very little experience in FGD systems. The MDU is a pilot plant consisting of mobile semitrailer modules, a spray dryer, and a baghouse, all of which can be trucked from site to site. It is essentially a self-contained unit requiring only foundations to be built at the site and access to flue gas, electricity, steam, and water.

Gerald Gentleman Station--Fabrication of the MDU was completed during late summer of 1979, and the unit was erected during November 1979 at Nebraska Public Power District's (NPPD) Gerald Gentleman Unit 1. This 650-MW pulverized-coal-fired boiler burns a Powder River subbituminous coal containing 0.31% sulfur and having a heating value of 8,900 Btu/lb. Uncontrolled SO_2 emissions are approximately 0.70 lb/MBtu and the SO_2 concentration in the flue gas is about 300 ppm. (An SO_2 spiking system has been installed to allow testing in the 300 to 1,800 ppm SO_2 range.) Completion of this initial test phase is scheduled for September 1980.

The MDU consists of four semitrailers, one containing the lime slurry preparation area (hoppers, slaker, tanks, and pumps), a second containing the analysis laboratory, a third containing the process control board, and a fourth containing the ID fan, stack, and storage area. The spray dryer (originally built by Niro Atomizer) is rated at 10,000 aft^3/min (about 3 MW) and has a drying chamber measuring 10 feet in diameter and 15 feet high. It is equipped with either one rotary atomizer or one two-fluid nozzle atomizer. The rotary atomizer is rated at 150 hp and can be operated at 1,000 to 23,000 rpm. The flue gas leaves the spray dryer through a down-turned vent forcing the gas to turn 180 degrees. This allows some of the particulate matter to drop out at the base of the spray dryer and be removed through a valved opening.

The flue gas from the spray dryer passes through a four-compartment baghouse before entering the ID fan and the stack. The bags in three of these compartments are the standard utility size (12 inches in diameter by 35 feet long) while the fourth compartment contains experimental bags (16 inches in diameter by 35 feet long).

Since this is the initial installation, a four- to five-month test program was scheduled in which changes in various operating parameters were to be evaluated. Although data acquisition and analysis are continuing, much of these data are proprietary in nature. However, parameters which are being evaluated include: inlet flue gas temperature, approach to flue gas, saturation temperature, stoichiometric ratio, inlet SO_2 concentration, and alkali raw material. Raw materials being tested include hydrated lime, pebble lime, high-calcium lime, and dolomitic lime. The applicability of waste recycle is also being tested.

Demonstration Units--

No plans have been announced as yet for a demonstration unit.

Commercial Units--

No contracts have been awarded to Ecolaire for a commercial, full-scale utility spray dryer FGD system at the present time. Ecolaire, however, has qualified and submitted at least one bid on a utility system.

Conceptual Design

Although Ecolaire has not been awarded any contracts for a full-scale, commercial utility FGD system, the following process description for a 500-MW coal-fired power unit is based on their conceptual designs and early pilot-plant experience. Additional pilot-plant experience in the future may result in some minor design modifications, but the overall process description, equipment, and operations are not expected to undergo any significant change.

Flue gas from the boiler air heater at approximately 300°F enters a common plenum which distributes the gas to the five trains of spray dryers (four operating and one spare). Most of the flue gas enters the spray dryer from the top and passes downward around the rotary atomizers. As the flue gas passes the atomizer, the lime slurry is sprayed as a fine mist into the hot gas. The atomized slurry and flue gas remain intimately mixed throughout the spray dryer where SO_2 and HCl are absorbed and react with the lime slurry. The water present in the spray dryer evaporates and the mixed calcium-sulfur salts and the fly ash entrained in the gas pass into the flue gas ducts upstream of the baghouse.

The remaining flue gas, which bypasses the spray dryer, may reenter the flue gas ducts to reheat the flue gas from the spray dryer before it enters the baghouse. The fly ash-FGD waste mixture is removed in the baghouse and temporarily stored before being recycled and/or trucked to an onsite landfill for disposal. The clean flue gas passes through an ID fan before entering the stack.

Technical Considerations

Due to the lack of published information concerning their process, it would not be appropriate to discuss any potential advantages or disadvantages of the Ecolaire spray dryer FGD process.

JOY MANUFACTURING/NIRO ATOMIZER, INC. (34)

The Western Precipitation Division of Joy Manufacturing Company, which designs and builds dust collection equipment, and Niro Atomizer, Inc., which designs and builds spray dryers and spray absorption systems, have an exclusive agreement to market a spray dryer FGD system. The initial two-year agreement was recently extended to November 1984 (with provisions for future extension of the agreement). In this partnership arrangement Joy Manufacturing will market the FGD system to the utility industry while Niro Atomizer will supply the industrial market.

Background and Current Status of Development

Pilot-Plant Units--

Hoot Lake Station--Although the original spray dryer FGD work was begun by Niro Atomizer, Inc., in Denmark in early 1975, it was not until Joy/Niro signed their exclusive agreement in late 1977 that actual pilot-plant operations of the spray dryer FGD system began. This first pilot plant was built at Ottertail Power Company's Hoot Lake Station, Unit 2 during January and February 1978. The primary reason for building this 20,000 aft^3/min plant was to prequalify to bid on Basin Electric Power Cooperative's Antelope Valley Unit 1 (for which Joy/Niro eventually was awarded the contract), but it was also built to gain additional operating experience on the flue gas from North Dakota lignite-fired boilers (as well as from other fuels which were burned at this boiler). Hoot Lake Unit 2 is a 53-MW cyclone-fired boiler burning lignite and emitting approximately 800 ppm SO_2 and 2 gr/aft^3 of particulate matter. Both the boiler and the resulting flue gas are somewhat similar to what will be encountered at Antelope Valley Unit 1.

This initial pilot plant consisted of an 11-ft-diameter spray dryer (rated at 20,000 aft^3/min) and a four-compartment baghouse containing utility-size bags (12 inches in diameter by 30 feet long). Both acrylic- and teflon-coated fiberglass bags were tested. This baghouse was designed for 15,000 aft^3/min. A small ESP (rated at 5,000 aft^3/min) was also used as a particulate collection device in some of the tests to evaluate its potential use in the spray dryer FGD system. In addition to the major equipment items and their associated equipment such as slakers, tanks, and pumps, a particulate recirculation loop was installed and operated to evaluate the effects on raw material utilization and stoichiometry. The pilot plant also contained SO_2 spiking equipment capable of maintaining 4,000 ppm SO_2 in the flue gas.

The initial operation of the pilot plant from mid-February 1978 to May 1978 was used for parametric tests, longer verification tests, and two 100-hour endurance tests. Both soda ash and lime were evaluated as absorbents and 90% SO_2 removal was demonstrated with each. In addition, recirculation tests with the lime waste were performed. (Niro Atomizer has a patent pending on this recirculation process.) When Joy/Niro had been awarded the contract for the full-scale Antelope Valley Unit 1 FGD system, the Hoot Lake pilot plant was reactivated for a series of verification tests from September through December 1978. This pilot plant has since been dismantled and removed.

Copenhagen Laboratory--The only active pilot plant now in the Joy/Niro organization is a recently completed (January 1979) 5,000 aft^3/min unit in Copenhagen, Denmark. This unit uses a propane burner, shipped in fly ash, and bottled SO_2 gas to simulate the flue gas at any potential site. The initial test work includes evaluation of a pulse jet and a reverse air baghouse as well as an ESP. Rather than being a research facility this pilot plant is used primarily to evaluate the actual fly ash prior to offering proposals for specific projects.

Demonstration Unit--

Riverside Station--Joy/Niro and Northern States Power Company have recently concluded an agreement in which Joy/Niro will build a 650,000 aft^3/min (about 100-MW) spray dryer FGD demonstration unit at Units 6 and 7 of Northern States Power's Riverside Station. This demonstration unit will duplicate one of the modules to be built at Antelope Valley Unit 1. Joy/Niro projects that this unit will be on-line by late fall of 1980. It is expected that the FGD system will operate for about three years. Since the baghouse will not be completed until December 1980 the existing ESP's will be used initially to collect the FGD waste.

Units 6 and 7 at Riverside were designed to burn high-sulfur (>3.0%) Illinois coal and were equipped with ESP's for particulate control. However, the plant was converted to low-sulfur (<1.0%) Montana coal. Because of the low-sulfur coal, the ESP's did not perform well and the plant did not meet its opacity standard. To increase the particulate removal efficiency and bring the plant back into compliance, these boilers are now burning a mixture of 85% to 90% low-sulfur Montana coal and 10% to 15% high-sulfur (4%) petroleum coke. Since Joy/Niro will be installing a baghouse designed for the flue gas from Units 6 and 7, these boilers are expected to meet the opacity standard without using high-sulfur fuel.

The single spray dryer will be 46 feet in diameter and contain a single rotary atomizer. Flue gas from the spray dryer will pass to one of two baghouses--each rated at 50% of total capacity. Design air-to-cloth ratio (gross) is 2:1. FGD waste from the spray dryer will be recycled. Pebble lime will be prepared in a ball mill slaker and pumped to the spray dryer.

Commercial Units--

Calgon--Niro/Joy was awarded a contract by Calgon Corporation in December 1979 for an FGD system at their plant at Big Sandy, Kentucky. The FGD system will treat 56,900 aft^3/min (at 1,700°F) of off-gas from a Herreshoff multiple hearth furnace (a tray-type tower typically fired with natural gas that is used for calcining solids). This high inlet temperature is not expected to cause significant problems for the spray dryer system. Additional water will be added in the spray dryer to cool and humidify this gas before it reaches the baghouse. The off-gas from the furnace will average 1,100 ppm SO_2 and 4,000 ppm HCl. Design SO_2 removal efficiency is 75%. HCl removal will be 99%. Startup is currently scheduled for June 1981.

This FGD system will use soda ash at Calgon's request. Waste recycle is not being considered and waste disposal will be handled by Calgon. A single 22.5-ft-diameter spray dryer containing one rotary atomizer will treat the total gas stream (there will be no spare spray dryer and no gas bypass). Flue gas from the spray dryer will pass to a single pulse-jet baghouse containing 360 bags in four compartments.

Antelope Valley Station--Joy/Niro has been awarded a turnkey contract for a full-scale spray dryer FGD system for Basin Electric Power Cooperative's Antelope Valley Station Unit 1. This FGD system will treat 2,055,000 aft^3/min of flue gas from the 440-MW lignite-fired cyclone boiler. The sulfur content in the North Dakota lignite is expected to range from 0.3% to 1.2%. Process design for the FGD system on Unit 1 is complicated since the required SO_2 removal efficiency will increase when Antelope Valley Unit 2 comes on-line. Although the FGD system will only be required to remove up to 78% of the SO_2 initially, it will be designed for the 89% (maximum) SO_2 removal that will be required when Unit 2 is on-line. (The contract for the FGD system for Unit 2 has not yet been awarded.) Commercial operation for Unit 1 is currently scheduled for April 1982.

The Antelope Valley FGD system will use lime as the absorbent and will use waste recycle to improve the lime utilization. It will also use hot gas bypass to maximize lime utilization. Five 46-ft-diameter spray dryers will be used, each containing a single rotary atomizer. During normal operation all five spray dryers will be operating although the FGD system can operate at full capacity with only four in use. The spray dryers will be insulated and enclosed in a simple shell-type building. Flue gas from the spray dryers will feed two baghouses, each rated at 50% of capacity. The design air-to-cloth ratio (gross) is 2.19:1 and there will be 28 compartments in the two baghouses. Since this is a mine-mouth plant, waste disposal will be in the mine. Pebble lime will be prepared in a ball mill slaker.

Springerville Station--Joy/Niro has also been awarded a materials-only contract for the full-scale spray dryer FGD systems for Tucson Electric Company's Springerville Station Units 1 and 2. Each of these units is a 350-MW boiler burning a New Mexico subbituminous coal averaging about 0.69% sulfur and having a heating value of 10,500 Btu/lb. The FGD system for each unit will treat 1,664,000 aft^3/min of flue gas containing 1,400 ppm SO_2. Although both FGD systems will be designed to achieve 61.3% SO_2 removal (these units are grandfathered under the NSPS) for the average coal, up to 82% SO_2 removal may be required if higher sulfur coals are burned. Unit 1 is scheduled to be on-line in February 1985 with Unit 2 following about 18 months later.

The Springerville FGD system will be lime-based and will use waste recycle to improve the lime utilization. Since the SO_2 removal requirement is only 61.3% these units will be designed for warm gas bypass rather than the hot gas bypass in the Antelope Valley system. Three 46-ft-diameter spray dryers, each with a single rotary atomizer, will be used for each boiler (six in all).

During normal operation all three spray dryers on each unit will be operating although the FGD system can operate at full capacity with only two spray dryers per unit. (Excess capacity in both the spray dryer and the bypass ductwork can accommodate the higher flue gas flow rate when one dryer is off-line although the lime feed rate to the spray dryers will have to be increased to maintain the SO_2 removal efficiency.) The spray dryers will be insulated and enclosed in a simple shell-type building. Flue gas from the spray dryers of each unit will feed two baghouses, each rated at 50% capacity. The design air-to-cloth ratio (gross) is 1.91:1, and there will be 28 compartments in the two baghouses for each unit.

Rawhide Station--The third commercial-scale spray dryer FGD system awarded to Joy/Niro was by the Platte River Authority for its Rawhide Station Unit 1. This boiler is a 250-MW unit burning a Wyoming subbituminous coal averaging 0.29% S (0.44% S maximum) and having a heating value of about 8,500 Btu/lb. The FGD system will treat 1,352,000 aft^3/min of flue gas containing 875 ppm SO_2 (based on 0.44% S). The design SO_2 removal efficiency will be 80%. Startup for this unit is currently scheduled for December 1983.

The Rawhide FGD system will be lime-based and will use waste recycle to improve the lime utilization. Since the SO_2 removal requirement is only 80%, this unit will be designed for warm gas bypass. Three 46-ft-diameter spray dryers will be used, each with a single rotary atomizer. In keeping with Joy/Niro's standard design, all three spray dryers will be on-line during normal operations with the higher sulfur coal. When the coal sulfur level remains near the design average of 0.29% S, only two of the spray dryers will be operating. The spray dryers will be insulated and enclosed in a simple shell-type building. Flue gas from the spray dryers will feed two baghouses, each rated at 50% capacity. The design air-to-cloth ratio (gross) is 2:1, and there will be 24 compartments in the two baghouses.

Holcomb Station--In April 1980 Joy/Niro was awarded a turnkey contract for the FGD system at Sunflower Electric Cooperative's Holcomb Station Unit 1. This is a 319-MW unit burning a Wyoming subbituminous coal averaging 0.3% sulfur and having a heating value of about 8,200 Btu/lb. The FGD system will treat 1,306,000 aft^3/min of flue gas containing 1,000 ppm SO_2. The design SO_2 removal efficiency will be 80%. Startup for this unit is currently scheduled for June 1983.

The Holcomb FGD system will be lime-based and will use waste recycle. In contrast to the typical lime spray dryer design, no flue gas bypass will be used. Three 50-ft-diameter spray dryers will be used, each with a single rotary atomizer. During normal full-load operation, all three spray dryers will be operating although the FGD system can operate at full capacity with only two spray dryers. The spray dryers will be insulated and enclosed in a simple shell-type building. Flue gas from the spray dryers will feed two baghouses, each rated at 50% capacity. The design air-to-cloth ratio (gross) is 1.85:1, and there will be 28 compartments in the two baghouses.

Argonne National Laboratory--Argonne National Laboratory awarded a turnkey contract to Niro/Joy in November 1980 for a lime-based spray dryer FGD system at their onsite boiler. The renovated spreader-stoker-fired boiler burns a 12,000 Btu/lb Illinois bituminous coal averaging 3.5% sulfur. The flue gas rate to the FGD system will be about 85,000 aft^3/min (at 355^{o}F). This boiler which was originally designed for coal, was converted to oil in the early 1970's and is now being reconverted to coal. Therefore the FGD system will be a retrofit. The applicable SO_2 emission limit is 1.2 lb of SO_2 MBtu and therefore approximately 78.5%. SO_2 removal will be required for the average coal. This FGD system is currently projected to start up in September 1981.

The FGD system will be lime-based and will use waste recycle. One 27-1/2-ft-spray dryer will be used to treat the entire flue gas stream (there will be no spare). Particulate removal will be by a pulse-jet baghouse which has four compartments. Design air-to-cloth ratio will be 3:1. There will be no flue gas bypass. Waste disposal will be Argonne's obligation and various methods and possible test programs are currently being evaluated.

Conceptual Design

The following conceptual design for a 500-MW coal-fired boiler is a generalized version of Joy/Niro's recent full-scale contract awards. This design is based on a lime system for a low-sulfur western coal application requiring a 70% SO_2 removal efficiency.

Flue gas from the boiler air heater at about 300^{o}F enters a common plenum which distributes the gas to five operating spray dryers (the ductwork and spray dryers are designed with excess capacity such that four of the dryers can handle the full boiler load). Each spray dryer is 46 feet in diameter and contains a single rotary atomizer. The spray dryers are insulated and may be enclosed in a simple, shell-type building. The five spray dryers feed two baghouses, each rated at 50% of full-load capacity. The baghouses are designed for an air-to-cloth ratio (gross) of 2:1. The flue gas from the baghouse passes through ID fans and into the stack plenum.

The hoppers in the baghouse are periodically emptied, and the waste is pneumatically conveyed to storage silos. The Niro spray dryers are designed so that the larger particles of FGD waste drop out of the flue gas and fall to the bottom of the spray dryer. Some of the material from the spray dryer and part of the material collected in the baghouse are reslurried and recycled through the spray dryer.

Pebble lime is received by rail and stored in concrete silos onsite. This lime is discharged as needed, conveyed to ball mill slakers, slurried, and pumped to the spray dryer as needed.

Technical Considerations

The Niro spray dryer (as shown in Figure 12) has several readily apparent differences from those offered by most other vendors. The

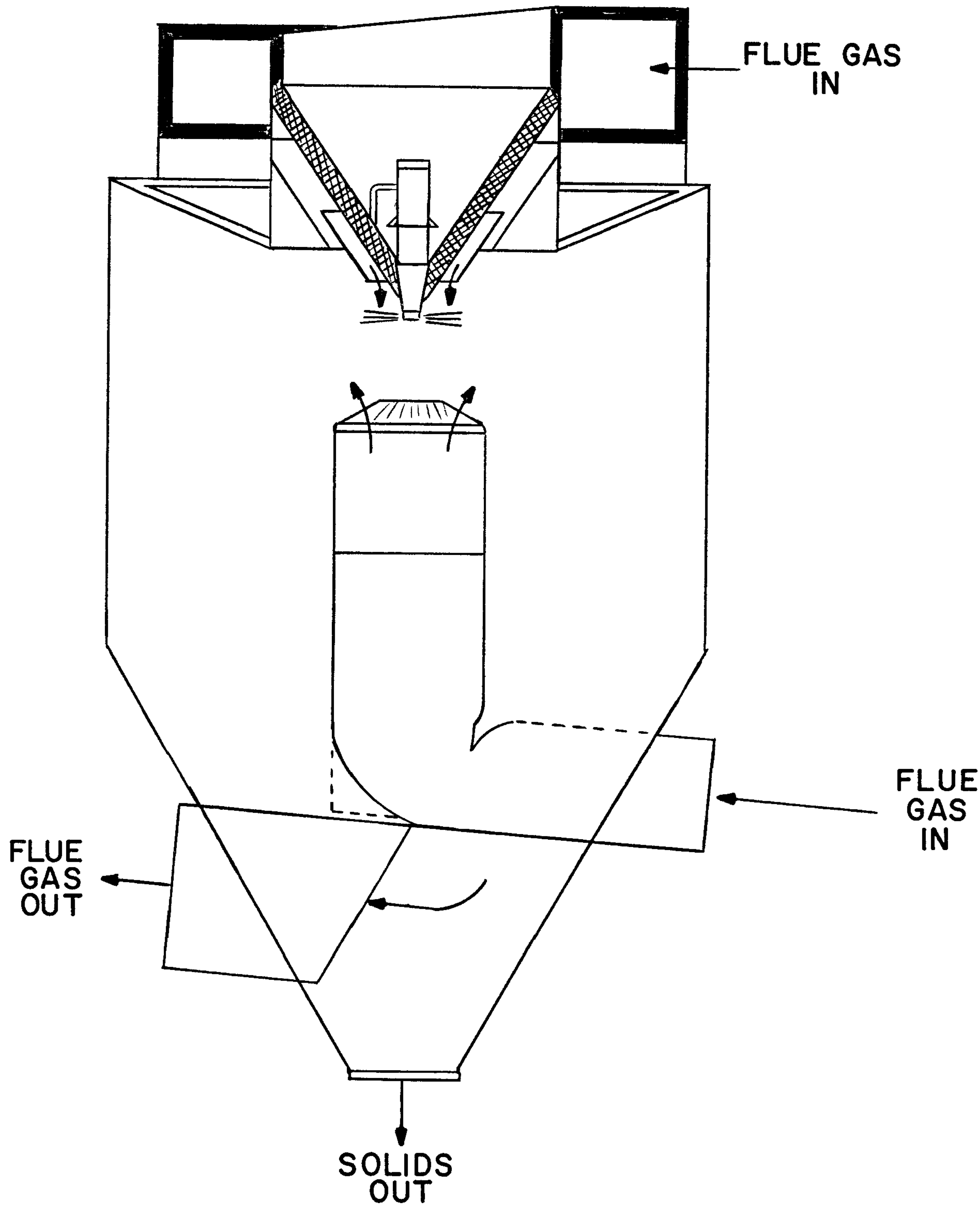

Figure 12. Spray dryer design for the Joy/Niro FGD process (11).

spray dryer contains only one rotary atomizer regardless of how large the spray dryer is (up to the maximum 50-ft-diameter size available). Although this rotary atomizer is relatively large (about 700 hp) and must handle significant quantities of lime slurry, this has been the standard spray dryer design in past non-FGD applications and has worked satisfactorily. Normally in the Joy/Niro design, all of the spray dryers, including the in-line spare, operate at full boiler load. Using the in-line spare tends to decrease the total system pressure drop and allows for better operation of the FGD system.

Another feature in the Niro spray dryer design is that the flue gas not only enters the top of the spray dryer in a concentric ring around the single atomizer, but it also enters from below the atomizer with an upward flow. This design is to assure more intimate mixing of the atomized mist and the flue gas and also to prevent possible distortion of the mist pattern at high flue gas velocities.

Flue gas leaving the spray dryer must make two turns near the base of the spray dryer, first to enter the outlet duct and then again inside the outlet duct. This design causes a decrease in flue gas velocity at the duct and most of the larger particles entrained in the flue gas drop out in the bottom of the spray dryer. A rotary valve at the base of the spray dryer allows the removal of this fly ash and FGD waste for recycle. This design also allows for the rapid cleaning of the spray dryer during any upset conditions when the particulate matter remains wet at the flue gas exit in the spray dryer. Since the particulate matter would be heavier when wet than when dry, it would have a tendency to fall out at the base of the spray dryer rather than passing on to the baghouse.

Although the waste collected in the baghouse can be recycled, most of the waste used in the recycle stream comes from the spray dryer. The recycle waste is pneumatically conveyed to the slurry mixing tank where it is added to the makeup lime slurry from the slaker. The resulting combined slurry is pumped to the spray dryers.

The standard Niro spray dryer for utility applications is 46 feet in diameter, and each application normally consists of multiples of this standard size. The desired gas residence time (10 to 12 seconds) is obtained by varying the height of the drying chamber. Each spray dryer is insulated and the spray dryer area may be enclosed in a simple, shell-type building. The Joy/Niro design usually includes two baghouses, each rated at 50% capacity. The design air-to-cloth ratio (gross) is normally 2:1.

RESEARCH-COTTRELL, INC. (35)

Research-Cottrell, Inc., which has an extensive background in conventional wet limestone scrubbing, entered the spray dryer FGD field in 1979. Research-Cottrell has an agreement with Komline-Sanderson Engineering Corporation, a spray dryer designer and manufacturer since 1946, for the exclusive use of their spray dryer technology. Since Research-Cottrell also designs and markets ESPs and baghouses the agreement with Komline-Sanderson enables Research-Cottrell to provide a complete spray dryer FGD system.

Background and Current Status of Development

Pilot Plant Units--

Big Brown Station--During the summer of 1979 Research-Cottrell and Texas Utilities reached agreement to locate a 10,000 aft^3/min spray dryer pilot plant at Unit 2 of the Big Brown Station of Texas Utilities. Unit 2 at Big Brown burns Texas lignite averaging 1.0% sulfur and 7,500 Btu/lb to produce about 593 MW of power. Pilot-plant testing began in June 1979. The primary purpose of this pilot unit was to demonstrate the technical feasibility of the process and to initiate tests to optimize the process.

This pilot plant had a single, 8-ft-diameter Komline-Sanderson spray dryer rated at about 10,000 aft^3/min and a Research-Cottrell baghouse. This two-compartment baghouse contained commercial-size bags (12 inches diameter by 30 feet long) and was designed to treat 5,000 aft^3/min. The potential for FGD waste recycle was evaluated. Absorbents included were slaked lime and fly ash. Due to the propriety nature of the development work, specific test results are not available.

Comanche Station--Research-Cottrell has recently transferred their 10,000 aft^3/min pilot plant to Unit 2 of Public Service Company of Colorado's Comanche Station. This is a 385-MW (gross) unit burning a low-sulfur western subbituminous coal averaging 0.5% sulfur and having a heating value of about 9,000 Btu/lb. Pilot-plant testing began in May 1980 and is scheduled to continue through 1980. The primary purposes of this pilot unit are twofold: to demonstrate this process on a subbituminous coal and to run tests to optimize the process for similar applications. In addition, waste disposal studies will be made. Partial funding for this pilot plant is being provided by the EPA, and therefore at least some of the data and results are expected to be available later this year.

Demonstration Units--

No plans for a demonstration unit have been announced by Research-Cottrell.

Commercial Units--

Stanton Station--Research-Cottrell has been awarded a contract for a lime spray dryer FGD system for United Power Associations's Stanton Station Unit 2. This 63-MW cyclone unit will burn a North Dakota lignite which is expected to average 0.77% sulfur and have a heating value of about 7,000 Btu/lb. Design SO_2 removal efficiency ranges from 73% for the average lignite (0.77% sulfur) to 90% for the worst-case lignite. Guaranteed SO_2 removal efficiency for worst-case lignite (1.94% sulfur) is nearly 91%. This FGD system is currently projected to startup during September 1982. The FGD system will be lime-based and will use waste recycle to improve the lime utilization. One Komline-Sanderson spray dryer containing a single rotary atomizer will be used to treat the entire flue gas stream. There will be no spare spray dryer. A Research-Cottrell baghouse will be used for particulate collection.

Conceptual Design

Although Research-Cottrell has not received any contracts for a full-scale utility FGD system, a conceptual design for this system has been prepared. For a typical 500-MW coal-fired boiler, the flue gas from the boiler air heater passes through a common plenum to six spray dryers connected in parallel (although all six would be operating, five can handle the full boiler load). Each spray dryer has three rotary atomizers. The flue gas leaves the spray dryer through an upward turning duct (as shown in Figure 13). This design forces the flue gas to make a 180-degree turn, thereby slowing the gas velocity and allowing some of the larger particles to drop to the bottom of the spray dryer.

The flue gas passes through the duct to the baghouse where the remaining entrained particulate matter is removed. The clean flue gas from the baghouse passes through an ID fan to the stack. Through proper design, the flue gas does not need further stack gas reheat before entering the stack.

Makeup pebble lime is received by rail or truck and stored in onsite storage silos. Lime is removed as needed, slaked with makeup water, and pumped to the spray dryers. For most applications now envisioned (low-sulfur western subbituminous coal or lignite applications), waste recycle will be used to achieve acceptable raw material utilization. FGD waste can be removed both from the spray dryer and the baghouse and pneumatically conveyed to a reslurrying tank for recycle through the spray dryers.

Technical Considerations

Due to the lack of published information concerning their process, it would not be appropriate to discuss any potential advantages or disadvantages of the Research-Cottrell spray dryer FGD process.

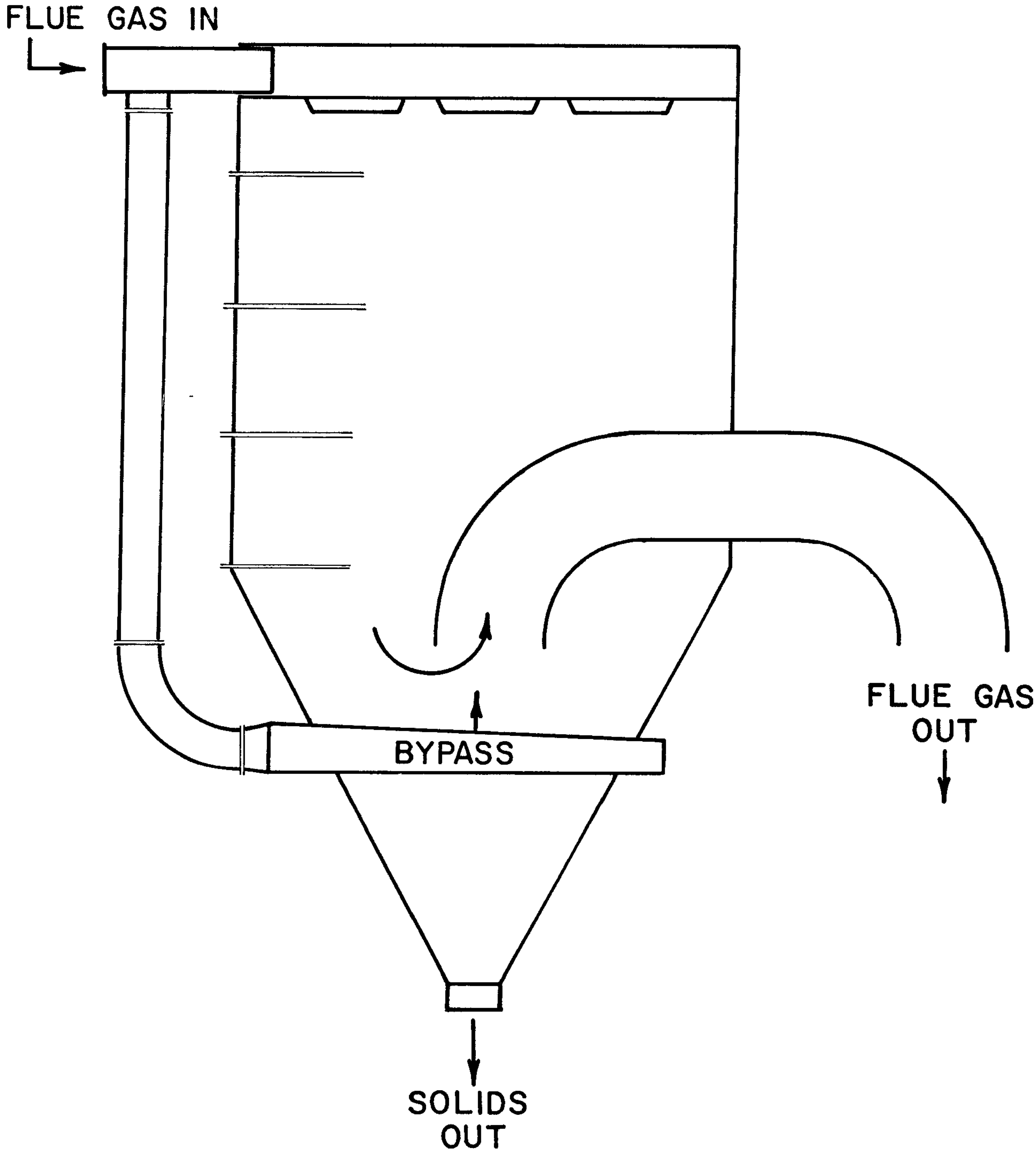

Figure 13. Spray dryer design for the Research-Cottrell FGD process.

ROCKWELL INTERNATIONAL/WHEELABRATOR-FRYE, INC. (36)

Until early 1980 Rockwell International/Wheelabrator-Frye, Inc., and Stork-Bowen Engineering, Inc., had exclusive agreements to design and build spray dryer FGD systems called the two-stage open-loop process. Rockwell International brought its previous pilot-plant experience in spray dryer FGD work from the development of the regenerable aqueous carbonate process, while Stork-Bowen and Wheelabrator-Frye provided the necessary experience in designing and marketing spray dryers and baghouses, respectively. In the past Wheelabrator-Frye has also been deeply involved in the conceptual design and pilot-plant testing of dry injection of nahcolite into flue gas ducts and baghouses.

However, the joint venture of Rockwell International and Wheelabrator-Frye was recently dissolved, with both Rockwell International and Wheelabrator-Frye continuing to design and market their own spray dryer FGD system. Stork-Bowen Engineering will continue to provide the spray dryers for the Rockwell system, but the baghouse will now be subcontracted on a competitive bid basis.

Background and Current Status of Development

Pilot-Plant Units--

The original spray drying work was done by Rockwell International and Stork-Bowen in the early 1970's and was related to the development of the regenerable, closed-loop aqueous carbonate process (while Wheelabrator-Frye was involved in dry nahcolite injection pilot-plant work), it was not until 1977 that these three companies came together to develop the dry waste-producing, open-loop aqueous carbonate process, which is now called the two-stage open-loop process.

Leland Olds Station--Initial pilot-plant work on the two-stage open-loop process was done in 1977 and early 1978 in an existing 4,000 to 5,000 aft^3/min pilot-plant unit at the Leland Olds Station of Basin Electric Power Cooperative. The baghouse section of this pilot unit was built in late 1976 to test and demonstrate an FGD system using dry nahcolite as the SO_2 absorbent for the Coyote Unit 1 to be operated by Ottertail Power Company. Leland Olds is a cyclone boiler similar to the Coyote unit and burns a similar North Dakota lignite. However, questions about the availability of nahcolite in commercial quantities led to the search for other raw materials. Based on spray dryer tests conducted at Stork-Bowen Engineering's test facility in early 1977, a spray dryer was added and the test program at Leland Olds was extended to include the use of a spray dryer/baghouse FGD system as an alternative to nahcolite injection.

The modified pilot plant consisted of a 7-ft-diameter spray dryer followed by the original two-compartment baghouse containing 11-1/2-inch-diameter by 30-ft-long filter bags. A variety of raw materials was evaluated including soda ash, trona, lime, limestone, and fly ash at various inlet SO_2 concentrations from 400 to 2,300 ppm. The test program was completed in August 1978 and the pilot plant was dismantled in September 1978.

Joliet Station--Early in 1978 Rockwell International/Wheelabrator-Frye reached agreement with Commonwealth Edison Power Company to reassemble their pilot plant at the Joliet Station in Illinois. The Joliet Station currently burns a mixture of four western subbituminous coals ranging from 0.4% to 1.0% sulfur and 9,000 to 10,000 Btu/lb in heating value. Testing began in July 1979 and is expected to last approximately two years. The test unit has a single, 7-ft-diameter spray dryer (4,000 to 5,000 aft^3/min) with the option of testing two types of baghouses as well as an ESP. In addition three types of slakers will be evaluated.

This pilot unit is currently being operated on a time-sharing basis. Both Rockwell International and Wheelabrator-Frye conduct separate test work for discrete time periods.

Mobile Pilot Plant--

In late 1978 and early 1979 Rockwell International/Wheelabrator-Frye designed and built a mobile pilot plant capable of being moved on flatbed trucks. Not only can this mobile unit be used to evaluate the effects of various flue gases from a wide variety of both coals and boiler types on the spray dryer FGD system design, but it can also be used for determining operating conditions for a specific boiler and coal in preparing a bid. This unit has a single 7-ft-diameter spray dryer (4,000 to 5,000 aft^3/min) and a pulse-jet baghouse. The mobile pilot plant was initially moved to Northern States Power Company's Sherburne County 700-MW Unit 1. This unit burns a Montana subbituminous coal containing 0.8% sulfur and having a heating value of 8,600 Btu/lb. Testing lasted through June 1979.

During July 1979 the mobile pilot plant was moved to Pacific Power and Light's and Idaho Power and Light's Jim Bridger Station. The Jim Bridger Station has four units, three of which are currently operating. Each of these three units are rated at 500 MW, and each must have a retrofitted FGD system capable of 75% SO_2 removal on-line by 1988. The test program at Jim Bridger lasted through August and September 1979.

During late 1979 the unit was moved again to the Celanese industrial boiler site in Cumberland, Maryland, for additional test work to verify the FGD system design. The mobile pilot plant is currently at Rockwell International's Santa Susanna field laboratories with the testing schedule to resume in September 1980. The system still includes a pulse-jet collector but not the same collector used in previous testing.

Demonstration Units--

No plans have been announced as yet for a demonstration unit.

Commercial Units--

Celanese--The Rockwell International/Wheelabrator-Frye joint venture received a turnkey contract in January 1979 from Celanese Corporation for a lime-based spray dryer FGD system for a coal-fired boiler at their Amcelle plant in Cumberland, Maryland. This renovated stoker-fired boiler burns an eastern bituminous coal ranging from 1.0% to 3.5% sulfur. A pulse-jet baghouse is used for particulate collection. The FGD system also incorporates warm gas bypass to increase lime utilization.

Although the spray dryer FGD system is designed to treat up to a maximum of 87,000 aft^3/min (at 420°F), under average boiler loads, the flue gas rate is typically 65,000 aft^3/min (at 380°F). The current emission regulations for the boiler are 70 lbs/hr for SO_2 and 0.01 grains/aft^3 for particulate matter in the flue gas from the stack. This SO_2 standard corresponds to about 70% SO_2 removal for a 1.0% sulfur coal and 86% SO_2 removal for a 2.0% sulfur coal.

The FGD system contains a single 20-ft-diameter spray dryer followed by a four-compartment, pulse-jet baghouse. In contrast to the large spray dryers for utility applications, which have three rotary atomizers, this smaller 20-ft-diameter model has only one. The pulse-jet baghouse used in this application is typical of smaller installations rather than the reverse-air type usually used in large utility applications. The boiler and the FGD system was put in operation in January 1980 and acceptance testing for the FGD system was completed in February.

Coyote Station--A turnkey contract was awarded to Rockwell International/Wheelabrator-Frye in April 1977 by the consortium headed by Ottertail Power Company for the Coyote Unit 1 FGD system. Unit 1 (originally scheduled for startup in April 1981 but currently ahead of this timetable) is a 456-MW (gross) cyclone-fired boiler burning North Dakota lignite averaging 0.78% sulfur and having a heating value of 7,000 Btu/lb. This spray dryer FGD application is currently the only utility system using a sodium-based absorbent (all others are lime). This system was chosen because disposal of the fly ash, which contains about 25% sodium salts, already presented a solubility problem. Final disposal of the FGD waste and the fly ash will be in an adjacent mined-out area (this is a mine-mouth power plant).

The FGD system will consist of four 46-ft-diameter spray dryers each containing three rotary atomizers and will treat about 1,890,000 aft^3/min. During normal operation all four spray dryers will be operating although the FGD system can operate at full capacity with only three in use. The spray dryers will be insulated and enclosed in a simple, shell-type building. Particulate collection will be in a baghouse. Neither waste recycle nor flue gas bypass for reheat will be used. Guarantees for the worst-case coal include 70% SO_2 removal (for $\leq$1.4% sulfur lignite) and 80% utilization of sodium carbonate.

Conceptual Design

The two-stage open-loop FGD system for a full-scale, commercial, utility boiler (about 500 MW) typically has four spray dryers in parallel. Three of these are operating at full boiler load and one is an in-line spare.

The spray dryer is designed such that all of the inlet flue gas enters a common open chamber directly above the rotary atomizers. One-third of the total inlet flue gas enters the vanes which form a concentric ring surrounding each rotary atomizer in the spray drying chamber. There are no internals within the spray dryer. The flue gas leaves at the bottom of the spray dryer through a downward sloping duct. The ductwork to the baghouse has this downward slope so that during low boiler loads (and the resulting decrease in flue gas velocity) the particulate matter that falls out will be swept forward into the baghouse hoppers. In addition, during upset conditions any wet waste will flow into hoppers in the baghouse inlet manifold rather than remaining in the spray dryer or the ductwork. When recycle of waste is required to achieve acceptable raw material utilization rates, FGD waste will be removed from the baghouse, reslurried in a separate tank, and combined with the makeup slurry at the spray dryer.

Technical Considerations

The Stork-Bowen spray dryer (as shown in Figure 14) is different from most others in that each spray dryer has three rotary atomizers. The use of three rotary atomizers decreases the size of each atomizer since each must atomize only one-third of the total absorbent fed to the spray dryer. This design (which uses state-of-the-art atomizers and vane rings) is claimed to result in a better mix of flue gas and atomized absorbent and thus lead to better raw material utilization. Conversely the use of multiple atomizers complicates the flue gas distribution to the atomizers.

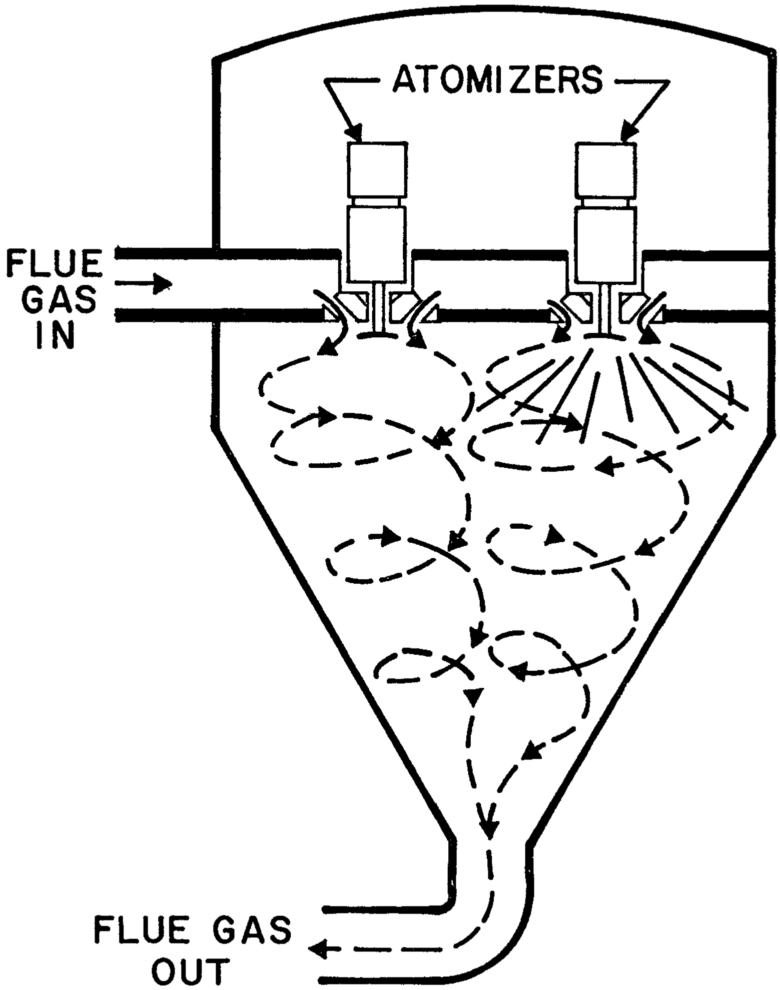

Figure 14. Spray dryer design for the Rockwell International FGD process (36).

DESIGN AND ECONOMIC PREMISES

In this study the economics of three FGD systems (two spray dryer processes--one sodium-based, the other lime-based--and a limestone scrubbing process) are compared on an equitable basis using conditions that are representative of projected industry conditions and that provide a clearly definable breakdown of costs into significant and useful divisions. The premises used in this study have been developed by TVA, EPA, and others during similar economic evaluations made since 1967. The design premises are formulated to establish efficiencies, process flow rates, and other operating and design conditions. The economic premises are designed to represent the many factors affecting costs.

DESIGN PREMISES

The utility plant design and operation is based on Federal Energy Regulatory Commission (FERC) historical data (37) and TVA experience. The conditions used are representative of a typical modern boiler for which FGD systems would most likely be considered. An Upper Great Plains and Rocky Mountain location (Wyoming, Colorado, Nebraska, and North and South Dakota) is used for the lignite and the low-sulfur western coal cases because the concentration of these low-sulfur coal supplies in this area make it representative of the segment of the power industry most active in spray dryer FGD. A midwestern location (Illinois, Indiana, and Kentucky) was selected for the low-sulfur and high-sulfur eastern coal cases for similar reasons.

In keeping with current industry practice, a redundant absorber train is provided to maintain acceptable boiler availability. In the integration of the absorber system with the boiler systems, provision for turndown and maintenance are limited to provision of a common plenum between the systems with dampers to allow individual trains to be shut down.

Emission Standards

New source performance standards (NSPS) established by EPA in 1979 (38) specify a maximum emission, based on heat input, of 0.03 lb/MBtu for particulate matter, 1.2 lb/MBtu for SO_2, and 0.6 lb/MBtu for NO_x. In addition to meeting this maximum emission limit of 1.2 lb/MBtu for SO_2, the 1979 NSPS also require that new plants must reduce the uncontrolled SO_2 emissions from 70% to 90%, depending on the uncontrolled SO_2 emission level. For the lignite and for both low-sulfur coals chosen in this study, this percentage SO_2 reduction is 70%. For the high-sulfur eastern coals the percentage SO_2 reduction is 89.6%. In addition it is assumed that the boilers in this study are designed to meet the 0.6 lb/MBtu NO_x standard. The FGD system includes all the process equipment assumed to be needed to meet both the particulate matter and the SO_2 removal standards.

Fuel

The coal premises are composites of many samples representing major coal production areas. The lignite coal is assumed to have a heating value of 6,600 Btu/lb and an ash content of 7.2% as fired and a sulfur content of 0.9%, on a dry basis. The western coal has a heating value of 9,700 Btu/lb and an ash content of 9.7% (both as fired) and a sulfur content of 0.7% (dry basis). This coal is based on coals from various western coal fields (39). The eastern bituminous coals are both assumed to have a heating value of 11,700 Btu/lb and an ash content of 15.1% as fired. The sulfur content of the low-sulfur and high-sulfur coals are 0.7% and 3.5%, respectively, on a dry basis. The composition for all four coals are shown in Table 4.

Power Plant Design

A single boiler with a 500-MW net electrical output is used. This net output does not include the power requirements for the FGD system. In contrast to some previous FGD studies by TVA, particulate matter removal and disposal are included as part of the FGD unit rather than as part of the boiler because of the nature of the spray dryer FGD processes, which collect fly ash and sulfur salts simultaneously.

Power Plant Operation

A total operating lifetime of 165,000 hours over a 30-year period is used. The boiler capacity factor is 62.8% (equivalent to full load for 5,500 hr/yr). A boiler heat rate of 11,000 Btu/kWh is assumed for the lignite unit. The boiler heat rate for the other three coal cases is 9,500 Btu/kWh.

Flue Gas Composition

Flue gas compositions are the result of boiler design, fuel, and operating conditions. Combustion and emission conditions used to determine flue gas composition are based on boiler design and average values for the sulfur content of coal. Flue gas compositions are based on combustion of pulverized coal using a total air rate equivalent to 139% of the stoichiometric requirement. This includes 20% excess air to the boiler and 19% air inleakage in the ducts and boiler air heater, which reflect operating experience with horizontal, frontal-fired, coal-burning units. It is assumed that 80% of the ash present in the coal is emitted as fly ash. It is also assumed that 85% of the sulfur in the coal is emitted as SO_x for the lignite and the western coals while 92% sulfur is emitted as SO_x for the eastern coals. In all four cases 3% of the SO_x emitted is assumed to be SO_3 and the remainder SO_2. The base-case flue gas composition and flow rates calculated for these conditions are shown in Table 5.

TABLE 4. COAL COMPOSITIONS

Component	wt % as fired Lignite	Low-sulfur western	Low-sulfur eastern	High-sulfur eastern
C	40.1	57.0	68.8	66.7
H	2.8	3.9	3.6	3.8
O	12.4	11.5	6.3	5.6
N	0.6	1.2	1.4	1.3
S	0.57	0.59	0.67	3.36
Cl	0.01	0.1	0.1	0.1
Ash	7.2	9.7	15.1	15.1
Moisture	36.3	16.0	4.0	4.0

Basis: TVA design and economic premises

TABLE 5. BASE-CASE FLUE GAS COMPOSITIONS AND FLOW RATES

Flue gas component	Lignite coal, 0.9% S Volume, %	lb/hr	Western coal, 0.7% S Volume, %	lb/hr	Eastern coal, 0.7% S Volume, %	lb/hr	Eastern coal, 3.5% S Volume, %	lb/hr
N_2	68.95	4,509,000	73.09	3,887,000	75.67	3,867,000	75.21	3,851,000
O_2	5.08	379,500	5.39	327,200	5.03	325,400	5.54	323,900
CO_2	11.92	1,225,000	12.24	1,023,000	12.76	1,024,000	12.34	992,300
SO_2	0.05	7,825	0.04	4,760	0.04	4,850	0.20	24,330
SO_3	-	302	-	184	-	187	0.01	940
NO_x	0.03	2,210	0.03	1,590	0.03	1,908	0.03	1,980
HCl	-	86	0.01	504	0.01	418	0.01	418
H_2O	13.97	587,200	9.20	314,600	6.46	212,100	6.66	219,100
Ash	-	48,130	-	38,000	-	49,000	-	49,000
Total	100.00	6,759,000	100.00	5,597,000	100.00	5,485,000	100.00	5,463,000

Basis: TVA design and economic premises

FGD System Design

The design of the spray dryer processes are based on current industry practices developed from vendor information and published data representing the most prevalent and more fully developed systems. The processes are generic and do not simulate a particular vendor design. The limestone process is based on experience at the Shawnee EPA Alkali Test Demonstration Facility, extensive power industry experience with these processes, and vendor information. The soda ash spray dryer process is evaluated only for the western coal case because it is assumed that the highly soluble waste would present economically unpredictable disposal practices in areas of high net precipitation.

The spray dryers contain three rotary atomizers and have neither presaturators nor mist eliminators. Flue gas leaves the spray dryer at about 20°F above its saturation temperature and contains no entrained liquid. Flue gas from the spray dryers and the bypass ducts is mixed and sent to the baghouse where the dry particulate matter (both fly ash and FGD waste) is removed. This waste is either recycled or trucked to the landfill.

The limestone slurry process absorbers consist of spray towers with spray presaturators to saturate the flue gas before entering the spray tower and mist eliminators to reduce the entrained flue gas moisture to 0.1% in the exiting flue gas. The slurry is oxidized to over 99% $CaSO_4 \cdot 2H_2O$ (gypsum) by air sparged into the slurry recirculation tank. The waste effluent withdrawn from the absorber circulation system is thickened from 15% to 40% solids, filtered to 80% solids, and trucked to the landfill.

In the low-sulfur coal cases, four absorber trains are used for the lime spray dryer and the limestone scrubbing process. For the soda ash spray dryer process, five trains are used. The capacities are based on one train serving as a spare at full-load boiler operation. Partial scrubbing and warm gas bypass are used for the lime spray dryer and limestone scrubbing processes. The lime spray dryers are designed for 83% SO_2 removal and the limestone scrubbers are designed for 90% SO_2 removal so that the overall 70% SO_2 removal efficiency is achieved. This is the prevalent design practice for lime spray dryer FGD and has also been found more economical for wet scrubbing (40). Full scrubbing at 70% SO_2 removal is used for the soda ash spray dryer process because the high reactivity makes a close approach to flue gas saturation temperature unnecessary. For the high-sulfur eastern coal case, the bypass for the lime spray dryer process is limited to 4% of the flue gas and it is hot gas bypass. No bypass is used for the limestone scrubbing process. In addition each of the FGD systems is designed for emergency bypass around the scrubbers. This emergency bypass duct is sized to handle 50% of the scrubbed flue gas in addition to the normal bypass gas.

No flue gas reheat is provided for the spray dryer processes, and the systems are designed for a minimum flue gas temperature of 165°F to

175°F in the baghouse. To represent typical conservative design practice, a partial flue gas bypass around the boiler air heater is provided for emergency use to maintain this temperature during abnormal periods of operation. Indirect steam flue gas reheat to 175°F is provided as necessary in the limestone scrubbing process.

Because of the temperature sensitivity of the spray dryer processes, 4 inches of insulation are provided on all the ductwork from the boiler wall to the baghouse, whereas the limestone slurry process, which is less temperature sensitive, has only 2 inches of insulation. In addition, a simple shell building houses the spray dryers, while the absorbers for the limestone slurry process are not enclosed. Induced-draft (ID) fans located upstream of the stack plenum are provided to compensate for the pressure drop through the FGD system.

Operating conditions are shown in Table 6. These conditions are used for both the base-case and case variation studies. Cost scaling factors based on gas and product rates are used to calculate values at conditions other than the base case.

Raw Materials

The raw materials used for each process are listed below. Limestone is crushed and wet ground as part of the scrubbing operation. The lime is not processed, only slaked before use.

Property	Soda ash spray dryer process	Lime spray dryer process	Limestone scrubbing process
Size as received	<100 mesh	3/4 - 1-1/4 in.	0 - 1-1/2 in.
Ground size	-	-	90% <325 mesh
Analysis	99.5% Na_2CO_3	90% CaO	90% $CaCO_3$
Bulk density, lb/ft^3	35	55	95

Waste Disposal

The waste is trucked in on-road type trucks to a disposal area one mile from the plant site. An area-fill type landfill is used for both the lime spray dryer and the limestone slurry processes. The landfill size is based on a dry waste bulk density of 50 lb/ft^3, a gypsum wet bulk density of 120 lb/ft^3, and a 30-ft fill depth. The landfill design includes dikes, clay lining, drainage ditches, top soil storage, perimeter fencing and lighting, and runoff neutralization. An earthen-diked, clay-lined pond, designed to minimize the sum of land and construction costs, is used for the soda ash spray dryer process. Pond evaporation is assumed equal to rainfall. Provisions for normal site maintenance of the pond and for normal landfill operations, including compacting, covering the waste, contouring to control runoff, and revegetation, are included. No costs are provided for monitoring or post-operation maintenance.

TABLE 6. FGD SYSTEM DESIGN CONDITIONS

	Lignite		Low-sulfur western coal			Low-sulfur eastern coal		High-sulfur eastern coal	
	Lime spray dryer	Limestone scrubbing	Soda ash spray dryer	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing
Absorbent stoichiometry[a]	1.2	1.1	1.0	1.2	1.1	1.3	1.1	1.8	1.3
Bypass, %	22.5	28.1	0	22	28.1	19	25.2	4[b]	0
Total FGD ΔP, in. H_2O	12	8.6	12	12	8.4	12	8.5	12	9.5
Absorber									
Removal efficiency, %	83.5	90	70	83	90	83	90	89	90
Absorbent liquid, % solids[c]	22.5	60	0	22.5	60	3	60	17	60
ΔP, in. H_2O	2	2	2	2	2	2	2	2	2
L/G, gal/kaft^3[b]	0.3	80	0.13	0.2	80	0.3	80	0.3	90
Exit gas, wt % liquid	0	0.1	0	0	0.1	0	0.1	0	0.1
Effluent, % solids	100	15	100	100	15	100	15	100	15
Recombined gas, °F	170	180	170	170	170	170	160	170	127
Reheat, °F	-	-	-	-	-	-	10	-	43

a. Defined as mol Ca/mol SO_2 absorbed for both the spray dryer and the limestone processes.
b. Hot gas bypass.
c. Excludes dilution water and recycle loop if used.

ECONOMIC PREMISES

The economic premises are divided into criteria for capital costs for construction of the FGD system and annual revenue requirements for its operation. The premises are based on regulated utility economics using the design premises as a costing basis. The estimates use cost information obtained from engineering-contracting, processing, and equipment companies; raw material suppliers; and published cost indexes. Spray dryer costs were obtained by scaling vendor-supplied information. Raw material costs are based on those prevailing in the Upper Great Plains - Rocky Mountain region for the lignite and the low-sulfur western coal cases and the Midwest region for the low- and high-sulfur eastern coal cases. Labor costs are assumed equivalent to those in the Midwest for all coal cases.

Capital Costs

The capital structure for the electric utility company is assumed to be:

Common stock	35%
Preferred stock	15%
Long-term debt	50%

The cost of capital is assumed to be:

Common stock	11.4%
Preferred stock	10.0%
Long-term debt	9.0%
Weighted cost of capital (based on capital costs above)	10.0%

The discount rate is 10%, the same as the weighted cost of capital.

For other economic factors needed in financial calculations, the following values are assumed:

Investment tax credit	10%
Federal and State income tax	50%
Property tax and insurance	2.5%
Annual inflation rate	6%

The levelized annual capital charge approach used in these premises is similar to that used by the Electric Power Research Institute (EPRI) (41).

Depreciation--

A 30-yr economic life and a 30-yr tax life are assumed for the utility plant. Salvage value is less than 10% and is equal to removal costs. The annual sinking fund factor for a 30-yr economic life and 10.0% weighted cost of capital is:

$$\text{Sinking fund factor} = \frac{WCC}{(1 + WCC)^n - 1} = 0.61\% \quad (1)$$

where: n = economic life (in years)
WCC = weighted cost of capital (as a decimal fraction)

The use of the sinking fund factor does not indicate that regulated utilities commonly use sinking fund depreciation. The sinking fund factor is used since it is equivalent to straight-line depreciation levelized for the economic life of the facility using the weighted cost of capital.

An annual interim replacement allowance of 0.56% is also included as an adjustment to the depreciation account to ensure that the initial investment will be recovered within the actual rather than the forecasted life of the facility. Since power plant retirements occur at different ages, an average service life is estimated. Many different retirement dispersion patterns occur. The type S-1 Iowa State Retirement Dispersion pattern is used (42). This S-1 pattern is symmetrical with respect to the average-life axis, and the retirements are represented to occur at a low rate over many years. The interim replacement allowance does not cover replacement of individual items of equipment since these are covered by the maintenance charge.

The sum of the years digits method of accelerated depreciation is used for tax purposes. On a levelized basis (using flow-through accounting), this results in a credit in the fixed charge rate as follows:

$$\text{Accelerated tax depreciation} = \frac{2CRF_B \left(n_T - \frac{1}{CRF_T}\right)}{n_T\,(n_T + 1)\,(WCC)} \quad (2)$$

where: CRF_B = Capital recovery factor (weighted cost of capital plus sinking fund factor) for the economic life (as a decimal fraction)

CRF_T = Capital recovery factor (weight cost of capital plus sinking fund factor) for the tax life (as a decimal fraction)

n_T = Tax life (in years)

$$\text{Levelized accelerated depreciation credit} = (ATD - SLD) \times \frac{ITR}{1 - ITR}$$

where: ATD = Accelerated tax depreciation (as a decimal fraction)
SLD = Straight-line depreciation (as a decimal fraction)
ITR = Income tax rate (as a decimal fraction)

For a 50% tax rate, 30-yr tax life, 30-yr book life, 10.0% weighted cost of capital, and 0.61% sinking fund factor, the annual levelized accelerated depreciation credit is 1.36% using flow-through accounting.

Investment Tax Credit--

The levelized investment tax credit is calculated as follows:

$$\text{Levelized investment tax credit} = \frac{(CRF_B)\ (\text{Investment tax credit rate})}{(1 + WCC)\ (1 - ITR)} \qquad (3)$$

where CRF_B, WCC, and ITR are the same factors previously defined in equations (1) and (2).

Using a 10.0% weighted cost of capital, 0.61% sinking fund factor, 10% investment tax credit rate, 50% income tax rate, the levelized investment tax credit is 1.92% annually.

Income Tax--

The levelized income tax is calculated as follows:

$$\text{Levelized income tax} = [CRF_B + AIR - SLD]\ [1 - \frac{\text{Debt Ratio x Debt Cost}}{WCC}]\ [\frac{ITR}{1 - ITR}] \qquad (4)$$

where: AIR = Allowance for interim replacement

Using a 10.61% capital recovery factor (weighted cost of capital plus sinking fund factor), 0.56% allowance for interim replacements, 3.3% straight-line depreciation, 50% debt ratio, 9.0% debt cost, and a 50% income tax rate, the levelized income tax rate is 4.31%.

Annual Capital Charge--

The levelized annual capital charges for a publicly owned electric utility, as shown in Table 7, are 14.7% of the total investment. The annual capital charge includes charges for the capital recovery factor, interim replacements, insurance, and property taxes, State and Federal income taxes, and credits for investment credits and accelerated depreciation.

TABLE 7. LEVELIZED ANNUAL CAPITAL CHARGES

FOR REGULATED UTILITY FINANCING

	Capital charge, %
Capital recovery factor	10.61
Interim replacements	0.56
Insurance and property taxes	2.50
Levelized income tax	4.31
Investment credit	(1.92)
Accelerated depreciation	(1.36)
Total	14.70

The annual capital charge is applied to the total capital investment. It is recognized that land and working capital (except spare parts) are not depreciable and that provisions must be made at the end of the economic life of the facility to recover their capital value. In addition, investment credit and accelerated depreciation credit cannot be taken for land and working capital (except spare parts). The cumulative effect of these factors makes an insignificant change in the annual capital charge rate and is therefore ignored.

Capital Investment Estimates

Capital investment estimates for this study represent projects beginning in early 1981 and ending in late 1983. Capital cash flows for a standard project are assumed to be 25% the first year, 50% the second year, and 25% the third year of the project life. Capital costs for fixed assets are projected to mid-1982, which represents the approximate midpoint of the construction expenditure schedule. The estimates in this study are based on a process description, flowsheet, material balance, and equipment list. These study-level estimates are considered to have a -20% to +40% range of accuracy.

The total fixed capital investment consists of direct capital costs for equipment, building, utilities, service facilities, raw material and byproduct storage, waste disposal facilities, engineering design and supervision, construction expense, contractor fees, and contingency. The total capital investment consists of the total fixed capital investment plus allowances for startup and modifications, royalties, the cost of funds during construction, plus the cost of land and working capital.

Direct Capital Investment Process--

Direct capital costs cover process equipment, piping, insulation, transport lines, foundations, structures, electrical equipment, instrumentation, raw material and byproduct storage, site preparation and excavation, buildings, roads and railroads, trucks, and earthmoving equipment. Direct investment costs are prepared using the average annual Chemical Engineering (43) cost indexes and projections as shown below:

Year	1978	1979[a]	1980[a]	1981[a]	1982[a]	1983[a]	1984[a]
Plant	218.8	240.2	259.4	278.9	299.8	322.3	344.9
Material[b]	240.6	262.5	286.1	309.0	333.7	360.4	385.6
Labor[c]	185.9	209.7	226.5	244.6	264.2	285.3	305.3

a. TVA projections.
b. Same as index in Chemical Engineering (43) for "Equipment, machinery, supports."
c. Same as index in Chemical Engineering (43) for "Construction labor."

The overtime premium for 7% overtime is included in the construction labor. Appropriate amounts for sales tax and for freight are included in the process capital costs.

Direct Capital Investment - Utilities, Services and Miscellaneous--

Necessary electrical substations and conduit, steam, process water, fire and service water, instrument air, chilled water, inert gas, and compressed air distribution facilities are included in the utilities investment. These facilities are costed as increments to the facilities already required by the power plant. Service facilities such as maintenance shops, stores, communications, security, offices, and road and railroad facilities are estimated on the basis of process requirements. Services, non-power plant utilities, and miscellaneous costs will normally be in the range of 4% to 8% of the total process capital depending on the type of process. A 6% rate is used in this evaluation for both processes.

Indirect Capital Investment--

Indirect capital investment covers engineering design and supervision, architect and engineering contractor costs, construction costs, contractor fees, and contingency. Construction facilities (which include costs for mobile equipment, temporary lighting, construction roads, raw water supply, construction safety and sanitary facilities) and other similar expenses incurred during construction are considered as part of construction expenses and are charged to indirect capital investment.

Listed below are the indirect costs used:

	% of direct investment		
	Process	Landfill	Pond
Engineering design and supervision	7	2	2
Architect and engineering contractor	2	1	1
Construction expense	16	8	8
Contractor fees	5	5	5
Total	30	16	16

A contingency of 20% is included for the spray dryer processes because projects such as costed in this report have a higher likelihood of exceeding rather than underrunning the capital estimate particularly at their current status of development. The contingency for the limestone scrubbing process itself (i.e., excluding the landfill) is 10% because there is more experience with these FGD systems and fewer uncertainties in their design. A contingency of 20% is assumed for the landfill in the limestone scrubbing process. These contingencies are calculated as a percentage of the sum of the direct and the indirect investments.

Other Capital Investment--

Startup and modification allowances are estimated at 8% to 12% of the total fixed investment depending upon the complexities of the process being studied. For the spray dryer processes, a midpoint value of 10% of the total fixed investment is assumed. For the limestone scrubbing process 8% is used.

Cost of funds during construction is 15.6% of the total fixed investment for each process. This factor is equivalent to the 10% weighted cost of capital with 25% of the construction expenditures of the first year, 50% the second year, and 25% the third year of the project construction schedule. Expenditures are assumed uniform over each year. Startup costs are assumed to occur late enough in the project schedule that there are no charges for the use of money used to pay startup costs.

For both spray dryer processes, royalty fees of 1% of the total process capital (excluding pond or landfill) are charged. No royalty fee is assessed for the limestone scrubbing process. Land cost is assumed to be $5,000 per acre.

Working capital is the total amount of money invested in raw materials, supplies, finished and semifinished products, accounts receivable, and monies on deposit for payment of operating expenses such as salaries, wages, raw materials, purchases, taxes, and accounts payable. For these premises, working capital is defined as the equivalent cost of 1 month's raw material, 1.5 months' conversion cost, and 1.5 months' plant and administrative overhead costs. In addition, it includes an amount equal to 3% of the total direct investment to cover spare parts, accounts receivable, and monies on deposit to pay taxes and accounts payable.

Annual Revenue Requirements

Annual revenue requirements use 1984 costs and are based on 5,500 hours of operation per year at full load. Annual revenue requirements are divided into direct costs and indirect costs. Both first-year and levelized annual revenue requirements are determined. Levelized annual revenue requirements are based on a 10% discount factor and a 6% inflation rate over the 30-yr life of the power unit. Direct costs consist of raw materials, labor, utilities, maintenance, and analytical costs. Indirect costs consist of levelized annual capital charges and overheads.

Direct Costs--

Projected raw material, labor, and utility costs are listed in Table 8. Unit costs for lime and soda ash are different for the western and the eastern coal applications reflecting the actual differences in delivered cost to specific areas. These differences are primarily due to the delivery charges rather than the cost of the raw material itself. Unit costs for steam and electricity are based on the assumption that the required energy is purchased from another source. Unit costs ($/kW, mills/kWh) are calculated on the basis of net power output of the unit excluding the electricity consumed by the pollution control systems. Actually, electrical use by the pollution control equipment will result in a derating of the utility plant for either a new or a retrofitted unit. To minimize iterative calculations, the pollution control equipment is charged with purchased electricity instead of derating the utility plant.

TABLE 8. PROJECTED 1984 UNIT COSTS FOR RAW

MATERIALS, LABOR, AND UTILITIES

	$/unit	
	Lignite and western coal	Eastern coals
Raw materials		
Limestone	8.50/ton	8.50/ton
Lime	102.00/ton	75.00/ton
Soda ash	145.00/ton	160.00/ton
Labor		
Operating labor	15.00/man-hr	15.00/man-hr
Analyses	21.00/man-hr	21.00/man-hr
Mobile equipment	21.00/man-hr	21.00/man-hr
Utilities		
Process water	0.14/kgal	0.14/kgal
Electricity	0.037/kWh	0.037/kWh
Steam	-	2.50/klb

Maintenance costs are estimated as a percentage of the direct investment, based on unit size and process complexity. For the limestone slurry process, non-landfill maintenance is 7% and landfill maintenance is 3%. For the lime spray dryer process, non-landfill maintenance is 6% and landfill maintenance is 3%. For the soda ash spray dryer process, non-landfill maintenance is 5% and pond maintenance is 3%.

Indirect Costs--

The levelized annual capital charges consist of a sinking fund factor, an allowance for interim replacement, property taxes, insurance, weighted cost of capital, income tax, credits for accelerated depreciation, and investment credit. The levelized annual capital charge for a regulated utility, as was shown in Table 7, is 14.7%.

Plant and administrative overhead is 60% of conversion costs less utilities. The plant and administrative overheads include plant services such as safety, cafeteria, medical, plant protection, janitor, purchasing personnel, general engineering (excluding maintenance), interplant communications and transportation, recreational facilities, and the expenses connected with management activities. Fringe benefits such as retirement, vacation, dental and medical plans are included in the base wage rates.

SYSTEMS ESTIMATED

This section describes the design of the FGD processes evaluated in this report. For the lignite case only the lime spray dryer and limestone scrubbing process are evaluated. For the low-sulfur western coal case all three processes, the soda ash spray dryer, the lime spray dryer, and the limestone scrubbing processes, are evaluated. For the low-sulfur eastern coal and the high-sulfur eastern coal, again only the lime spray dryer and limestone scrubbing processes are evaluated. The lime spray dryer and limestone scrubbing processes are essentially the same design for all four coal cases. Equipment sizes, flow rates, waste recycling, and bypass and reheat requirements differ, depending on the ash compositions and removal efficiencies required, as described in the design premises.

The simple, basic chemistry of these FGD processes has been described elsewhere (3,4). In essence, the SO_2, SO_3, and HCl in the gas react with the absorbent to form a mixture of hydrated sulfites and sulfates and chlorides. In the soda ash spray dryer process, these are the highly soluble $Na_2SO_3 \cdot nH_2O$, $Na_2SO_4 \cdot nH_2O$, and NaCl. In the lime spray dryer and limestone scrubbing processes, these are the relatively insoluble $CaSO_3 \cdot 1/2H_2O$ and $CaSO_4 \cdot 2H_2O$ and $CaCl_2$. In all three processes the sulfite species predominates. In the air-oxidation limestone process used in this study, the waste is further oxidized by sparging air through the slurry to aid dewatering. This produces a waste composed almost entirely (about 95%) of $Ca_2SO_4 \cdot 2H_2O$ (gypsum), with the original quantity of $CaCl_2$.

LIGNITE CASE

The 0.9% sulfur lignite, with a heating value of 6,600 Btu/lb and 7.2% ash, is lower in both heating value and ash and higher in moisture content than the eastern and the western coals used in this study. Equally important are the characteristics of the ash, which has a high calcium content and a high resistivity. Both of these factors are important determinants of flue gas cleaning costs, particularly in economic comparisons of systems using spray dryers and those using wet scrubbing with separate fly ash collection. The alkalinity of the ash can serve to supplement expensive (i.e., nonlimestone) absorbents. Separate fly ash collection using ESP's requires higher SCA's than would be used for high-sulfur eastern coals and, hence, more expensive ESP's. The designs of FGD systems prescribed by these factors thus differ from designs of the same process for higher sulfur eastern coals.

Lime Spray Dryer Process

The process uses four trains of spray dryers, each with warm-gas bypass and one fabric filter baghouse. Each spray dryer is equipped with emergency hot-gas bypass ducts. Four ducts, each with an ID booster fan, connect the baghouse to the stack plenum. Because of the alkalinity of the fly ash, provision is made for recycling some of the collected waste. The flow diagram and layout are shown in Figures 15 and 16, and the material balance is shown in Table 9.

Pebble lime (CaO) is received by rail and stored in a silo. The lime is removed from the silo and slaked to an absorbent slurry of about 22% solids which is pumped to the spray dryers. Three top-mounted rotary atomizers are used, each with separate recycled waste and water additions. The SO_2 content of the cleaned flue gas controls the absorbent addition rate, and the temperature of the cleaned flue gas (after recombination with the bypassed flue gas) controls the water addition rate. These provide both SO_2 emission and flue gas temperature controls. About 22% of the flue gas bypasses the spray dryers and enters the ducts to the baghouse. The remaining flue gas enters the three operating spray dryers at $300^{\circ}F$ through manifolds around each atomizer and mixes with the atomized absorbent. The particle-laden flue gas leaves the bottom of the spray dryer and passes down an inclined duct where it mixes with the bypassed flue gas. The recombined flue gas enters the baghouse at $165^{\circ}F$. The SO_x in the flue gas continues to react with the absorbent particles until it passes through the cake on the fabric filters. The cake is periodically dislodged by reverse air flow and falls into hoppers from which it is periodically removed and pneumatically conveyed to storage and recycle silos. The recycle waste is mixed with water to form a 40% solids slurry and pumped to the spray dryers. About 65% of the material collected is recycled. The waste from the storage silo is trucked to the landfill site.

To facilitate cost determinations and comparisons, the lime spray dryer process is divided into six sections. The equipment list, giving the description and cost of each equipment item by section, is shown in Table 10. These costs do not include the investment required for foundations, structures, electrical components, piping, instruments and controls, etc. Each of these processing sections is described below.

Material Handling--

This and the following section, feed preparation, compose the raw material receiving and preparation section. The material handling section includes all of the equipment to receive pebble lime by rail and to maintain a supply of lime to the weigh feeders. It includes a lime storage silo with a 30-day capacity and two lime feed bins each having a 12-hour capacity.

Feed Preparation--

The feed preparation section includes the equipment necessary to convert the lime into a 22% solids slurry for SO_2 absorption. Two trains of lime preparation equipment (feeders, slakers, tanks, and

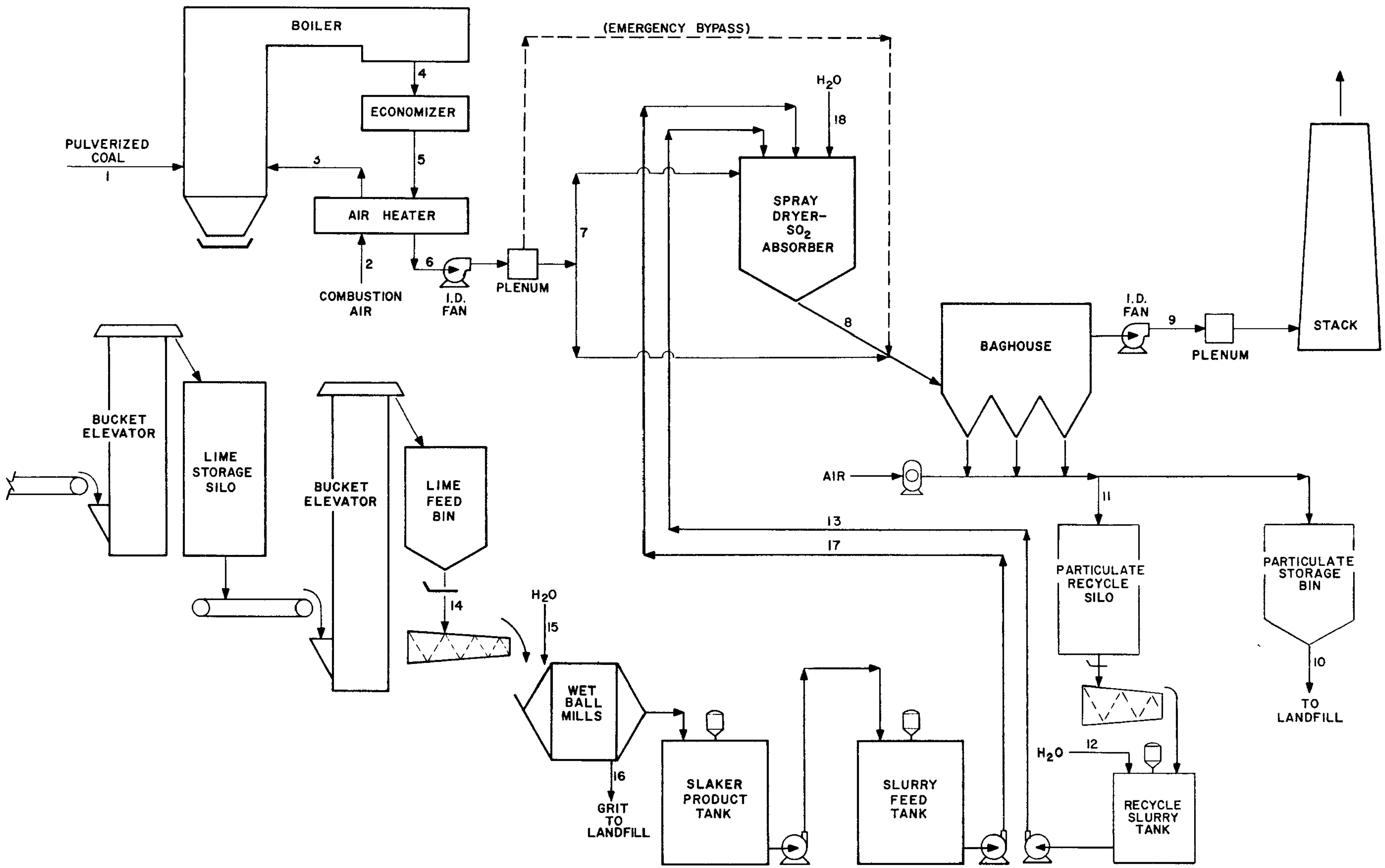

Figure 15. Lignite case. Lime spray dryer process. Flow diagram.

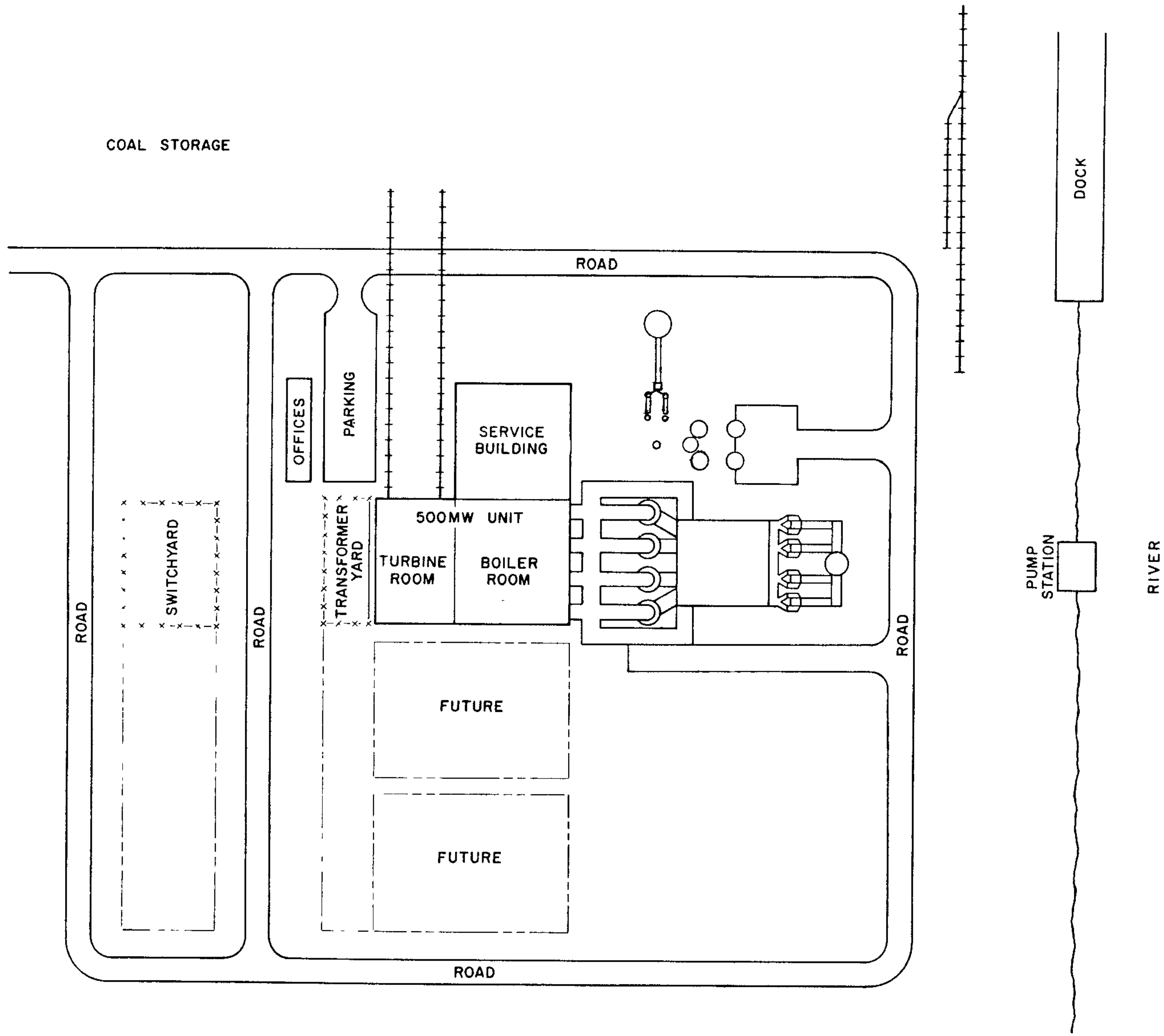

Figure 16. Lignite case. Lime spray dryer process. Plot plan.

TABLE 9. LIGNITE CASE

LIME SPRAY DRYER PROCESS

MATERIAL BALANCE

	Stream	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	833,300	5,958,000	5,124,000	5,925,000	5,925,000
2						
3	Flow rate, sft^3/min @60°F		1,315,000	1,131,000	1,290,000	1,290,000
4	Temperature, °F		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to FGD system	Gas to spray dryer	Gas from spray dryer	Gas to stack[a]	Waste to landfill
1	Total stream lb/hr	6,759,000	5,238,000	5,559,000	6,977,000	57,640
2						
3	Flow rate, sft^3/min @60°F	1,474,000	1,142,000	1,215,000	1,560,000	
4	Temperature, °F	300	300	155	175	
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Waste to recycle particulate silo	Makeup water to recycle slurry tank	Recycle slurry to spray dryer	Makeup lime to slaker	Makeup water to slaker
1	Total stream, lb/hr	110,200	165,400	275,600	5,910	18,270
2						
3	Flow rate, sft^3/min @60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm		332			37
7	Specific gravity					
8	pH					
9	Undissolved solids, %			40		
10						

	Stream No.	16	17	18		
	Description	Grit to landfill	Lime slurry to spray dryer	Dilution water to spray dryer		
1	Total stream, lb/hr	590	23,600	19,890		
2						
3	Flow rate, sft^3/min @60°F					
4	Temperature, °F			60		
5	Pressure, psig					
6	Flow rate, gpm			40		
7	Specific gravity					
8	pH					
9	Undissolved solids, %		22.5			
10						

a. Includes air inleakage.

TABLE 10. LIGNITE CASE

LIME SPRAY DRYER PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Car shaker	1	Top mounted with crane, 20 hp	74,700	11,300
2.	Car puller	1	25 hp with 5 hp return	64,600	24,200
3.	Hopper, unloading	1	12 ft x 12 ft x 2 ft bottom, 20 ft deep, carbon steel	3,700	2,600
4.	Pump, pit sump	3	Centrifugal, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	10,600	3,300
5.	Conveyor, lime unloading (enclosed)	1	Belt, 24 in. wide, 200 ft long, 2 hp, 100 tons/hr	14,600	1,000
6.	Elevator, storage silo	1	Continuous bucket, 16 in. x 8 in. x 11-3/4 in., 75 ft lift, 15 hp, 100 tons/hr, 160 ft/min	33,600	3,400
7.	Silo, lime storage	1	45 ft dia x 50 ft straight side, 79,500 ft^3, 45° slope, carbon steel	109,600	99,200
	Vibrators	1	Bin activator, 10 ft dia	14,500	2,400
8.	Conveyor, live lime feed	1	Belt, 14 in. x 100 ft long, 2 hp, 16 tons/hr, 100 ft/min	11,400	1,000
9.	Elevator, live lime feed	2	Continuous bucket, 8 in. x 5-1/2 in. x 7-3/4 in., 35 ft lift, 2 hp, 16 tons/hr, 150 ft/min	16,800	2,700

(continued)

TABLE 10 (continued)

Area 1--(continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
10. Bin, lime feed	2	12 ft dia x 12 ft straight side, 1,360 ft^3, 60^o slope, w/cover, carbon steel	13,400	11,500
11. Dust collecting system	1	Bag filter, polypropylene bag, 2,200 ft^3/min, 7-1/2 hp	8,000	1,600
Subtotal			375,500	164,200

Area 2--Feed Preparation

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Feeder, lime bin discharge	2	Vibrating, 3-1/2 hp, carbon steel	8,300	600
2. Feeder, lime feed	2	Screw, 6 in. dia x 12 ft long, 1 hp, 3 tons/hr	4,000	3,300
3. Slaker	2	Ball-mill type, 37.5 hp slaker, 1 hp classifier, 3 tons/hr	213,000	29,300
4. Tank, slaker product	2	7 ft dia x 9 ft high, 2,600 gal, open top, four 7 in. baffles, agitator supports, carbon steel, neoprene lined	5,800	4,300
5. Agitator, slaker product tank	2	2 turbines, 28 in. dia, 3 hp, neoprene coated	15,400	1,600

(continued)

TABLE 10 (continued)

Area 2 (continued)

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
6.	Pump, slaker product tank	3	Centrifugal, 43 gpm, 50 ft head, 1-1/2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	6,700	2,500
7.	Tank, slurry feed tank	1	11 ft dia x 15 ft high, 10,670 gal, open top, four 11 in. baffles, agitator supports, carbon steel, neoprene lined	11,500	8,600
8.	Agitator, slurry feed tank	1	2 turbines, 44 in. dia, 7-1/2 hp, neoprene coated	14,900	1,100
9.	Pump, slurry feed tank	12	Centrifugal, 43 gpm, 100 ft head, 5 hp, carbon steel, neoprene lined (6 operating, 6 spare)	30,000	9,000
10.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 ft^3/min, 7-1/2 hp (1/2 cost in materials handling area)	3,900	600
	Subtotal			313,500	60,900

Area 3--Gas Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Fan	4	Induced draft, 473,000 aft^3/min, 12 in. static head, 1,500 hp, fluid drive, double width, double inlet (4 operating)	2,895,500	56,100
	Subtotal			2,895,500	56,100

(continued)

TABLE 10 (continued)

Area 4--SO_2 Absorption

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Spray dryer	4	48 ft dia x 54 ft high, with 3 rotary atomizers, carbon steel (3 operating, 1 spare)	4,324,000	567,200
Subtotal			4,324,000	567,200

Area 5--Particulate Removal

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Baghouse	1	Automatic fabric filter, 28 compartments, 2.5 air-to-cloth ratio	8,973,000	3,227,000
Subtotal			8,973,000	3,227,000

Area 6--Particulate Handling and Recycle

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Conveyor, particulate feed to bin	1	Pneumatic, pressure vacuum, 250 hp	243,100	78,200
2. Bin, particulate storage	2	25 ft dia x 30 ft straight side, 14,800 ft^3, 60^o slope, w/cover, carbon steel	56,500	51;100
Vibrators	2	Bin activator, 10 ft dia	28,900	4,800

(continued)

TABLE 10 (continued)

Area 6--(continued) Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
3. Silo, particulate storage	2	28 ft dia x 44 ft straight side, 27,100 ft^3, 60° slope, w/cover, carbon steel	85,000	72,400
4. Feeder, particulate	2	Vibrating, 3-1/2 hp, carbon steel	8,400	800
5. Feeder, recycle slurry tank	2	Screw, 14 in. dia x 12 ft long, 5 hp, 60 tons/hr	30,800	4,500
6. Tank, recycle slurry	1	26 ft dia x 28 ft high, 110,200 gal, open top, four 26 in. baffles, agitator supports, carbon steel, neoprene lined	50,100	39,000
7. Agitator, recycle slurry tank	1	104 in. dia, 40 hp, neoprene lined	71,200	3,400
8. Pump, recycle slurry tank	12	Centrifugal, 153 gpm, 100 ft head, 20 hp, carbon steel, neoprene lined (6 operating, 6 spare)	53,300	12,500
Subtotal			627,300	266,700

agitators) are used. Each train is sized to handle 100% of the full load capacity. A slurry feed tank with a 4-hour capacity is provided.

Gas Handling--

Includea in this area is an inlet plenum that supplies the flue gas ducts to the scrubber trains. This area also includes the ductwork from the boiler to the inlet plenum, the bypass ducting around the spray dryers, the emergency bypass ductwork around the boiler air heaters and spray dryers, the ductwork from the spray dryers to the baghouse, and the ductwork from the baghouse to the stack plenum. It also includes the four ID booster fans between the baghouse and the stack plenum.

SO_2 Absorption--

Four spray dryers are provided (three operating and one spare); each is sized to handle one-third of the total flue gas volume being scrubbed.

Particulate Removal--

A single baghouse containing 28 compartments and the associated equipment is provided.

Particulate Handling and Recycle--

A single train of equipment to store, reslurry, and recycle the waste material is provided. Two particulate storage bins are included to provide a 24-hr capacity for waste material to be landfilled.

Limestone Scrubbing Process

The process uses four trains of flue gas process equipment (three operating and one spare). Each train consists of a cold-side ESP, the spray tower absorber system with warm-gas bypass, and an ID booster fan. The purge streams from the absorbers are dewatered in a single train consisting of a thickener and a rotary filter. The flow diagram and layout are shown in Figures 17 and 18 and the material balance is shown in Table 11.

Fly ash is collected in 99.8% efficient cold-side ESP's located upstream of the boiler ID fans. Because of the high resistivity of lignite fly ash, the ESP's are sized for a SCA of 700 $ft^2/kaft^3$. Standard pressure pneumatic conveying equipment is used to convey the fly ash to storage silos from which it is trucked to the landfill.

Flue gas from the ESP's enters the inlet plenum and is distributed to the three absorber trains and the bypass ductwork. About 28% of the flue gas in the plenum is bypassed around the absorbers to the stack plenum for reheat purposes. The remaining flue gas enters one of the three absorber trains and is cooled to about 140°F in a presaturator using sprays of absorbent liquid and passes upward through the absorber and mist eliminator, emerging at about 130°F with an entrained moisture content of 0.1%. This scrubbed flue gas passes through the booster ID fan to the stack plenum, where it is mixed with the bypassed flue gas. The resulting mixture, at 175°F and with an overall SO_2 reduction of 70%, enters the stack.

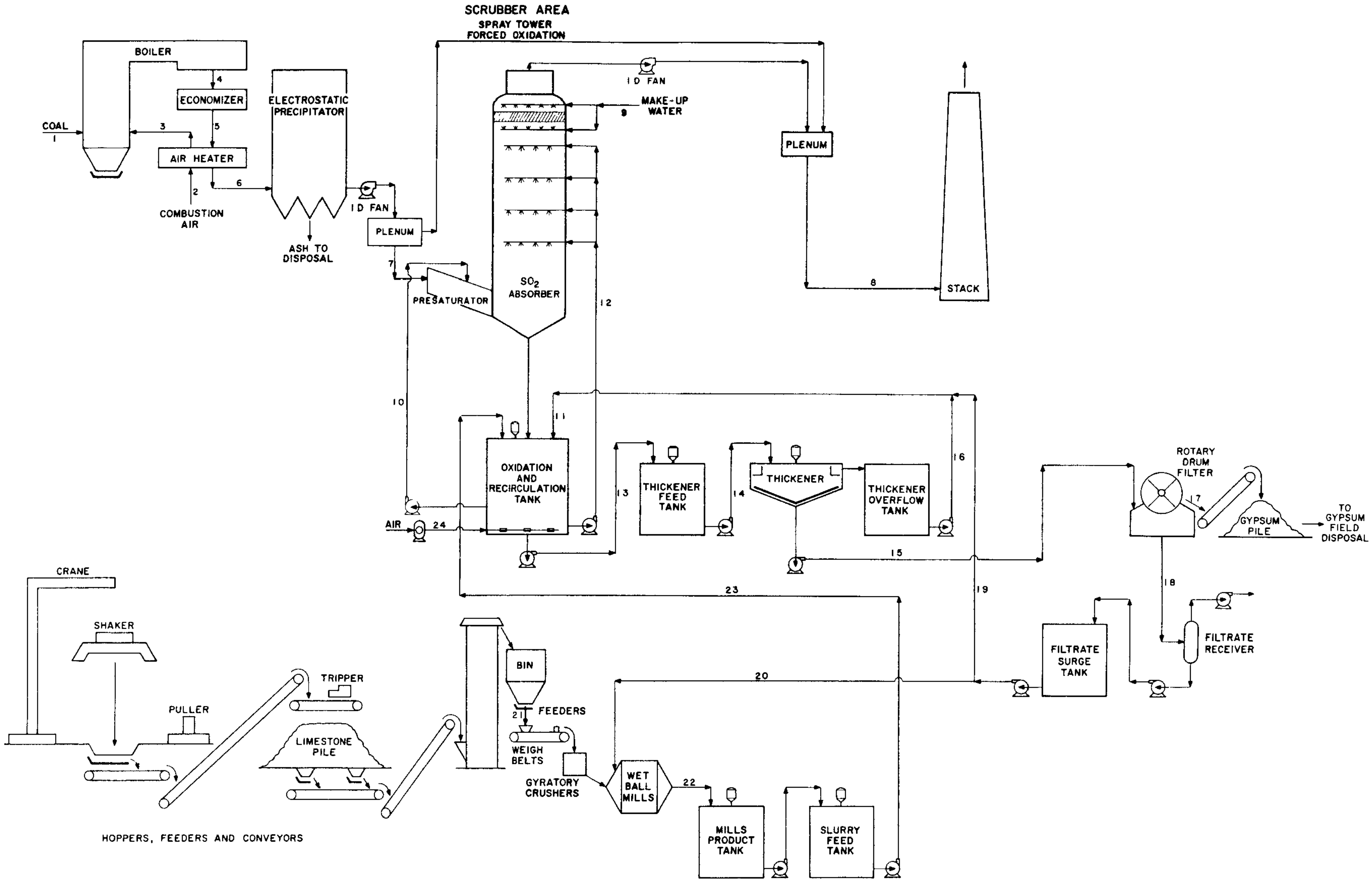

Figure 17. Lignite case. Limestone scrubbing process. Flow diagram.

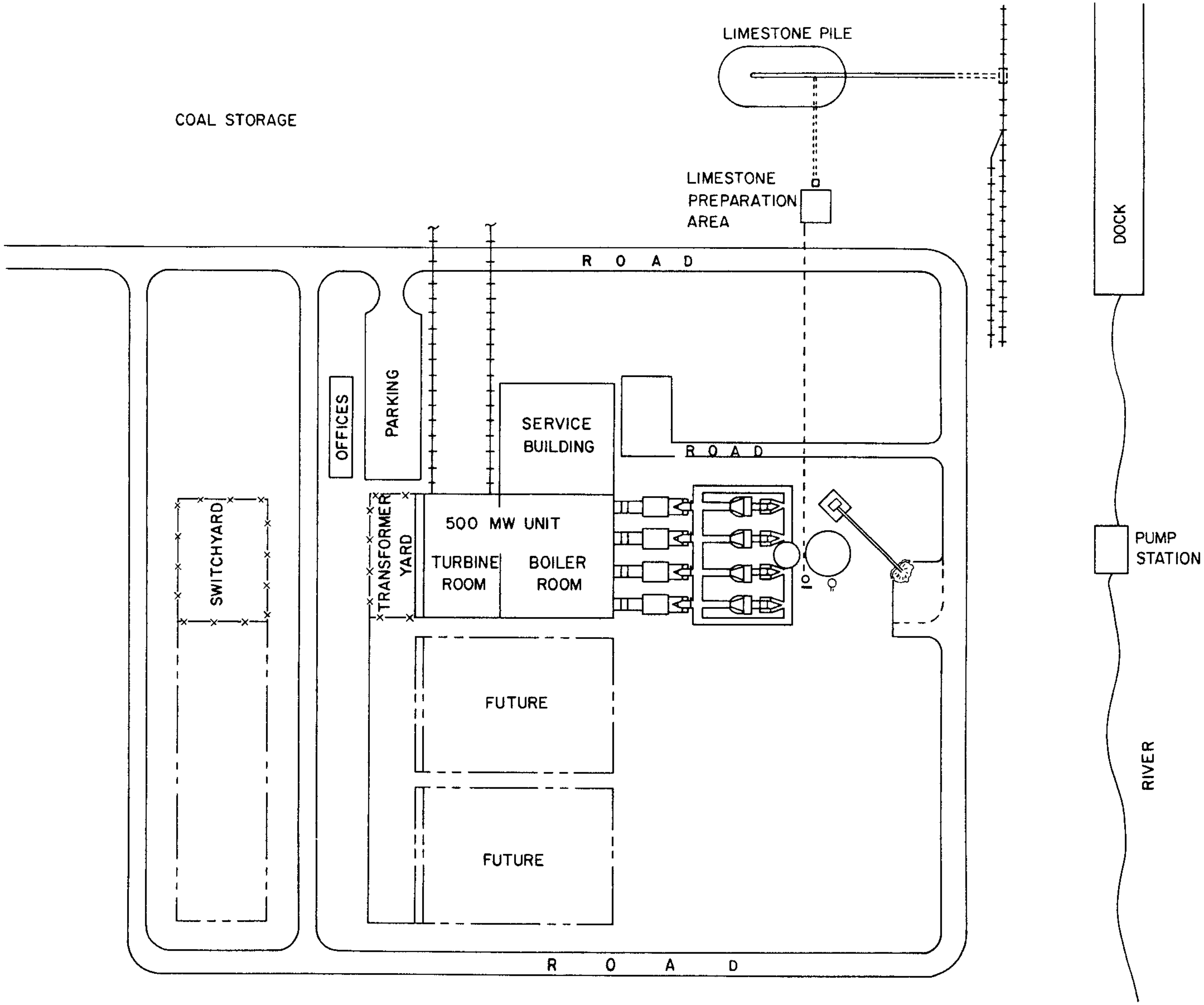

Figure 18. Lignite case. Limestone scrubbing process. Plot plan.

TABLE 11. LIGNITE CASE

LIMESTONE SCURBBING PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	833,300	5,958,000	5,124,000	5,925,000	5,925,000
2						
3	Flow rate, sft^3/min @60^{o}F		1,315,000	1,131,000	1,290,000	1,290,000
4	Temperature, ^{o}F		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Gas to spray tower	Gas to stack	Makeup water to spray tower	Recycle slurry to presaturator
1	Total stream, lb/hr	6,759,000	4,826,000	6,913,000	213,200	2,866,000
2						
3	Flow rate, sft^3/min @60^{o}F	1,474,000	1,060,000	1,545,000		
4	Temperature, ^{o}F	300	300	185		
5	Pressure, psig					
6	Flow rate, gpm				427	5,205
7	Specific gravity					
8	pH					
9	Undissolved solids, %					15
10						

	Stream No.	11	12	13	14	15
	Description	Supernate to oxidation-recirculation tank	Recycle slurry to spray tower	Slurry to thickener feed tank	Slurry to thickener	Thickener underflow to filters
1	Total stream, lb/hr	73,470	57,320,000	103,100	103,100	38,680
2						
3	Flow rate, sft^3/min @60^{o}F					
4	Temperature, ^{o}F					
5	Pressure, psig					
6	Flow rate, gpm	747	104,110	187	187	58
7	Specific gravity					
8	pH					
9	Undissolved solids, %		15	15	15	
10						

	Stream No.	16	17	18	19	20
	Description	Thickener over-flow to oxidation recirculation tank	Gypsum filter cake to disposal	Filtrate to filtrate surge tank	Filtrate to oxidation-recirculation tank	Filtrate to ball mills
1	Total stream, lb/hr	60,600	19,340	19,340	12,870	6,470
2						
3	Flow rate, sft^3/min @60^{o}F					
4	Temperature, ^{o}F					
5	Pressure, psig					
6	Flow rate, gpm	121		39	26	13
7	Specific gravity					
8	pH					
9	Undissolved solids, %		80			
10						

(continued)

TABLE 11 (continued)

	Stream No.	21	22	23	24	
	Description	Limestone to weigh feeders	Limestone slurry to mills product tank	Limestone slurry to oxidation-recirculation tank	Air to oxidation-recirculation tank	
1	Total stream, lb/hr	9,700	16,200	16,200	14,080	
2						
3	Flow rate, sft^3/min @60^{o}F				3,067	
4	Temperature, ^{o}F				60	
5	Pressure, psig					
6	Flow rate, gpm		20	20		
7	Specific gravity					
8	pH					
9	Undissolved solids, %		60	60		
10						

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Makeup limestone is fine crushed and wet ball milled to 90% less than 325 mesh and stored as a 60% solids slurry. This slurry, along with recycled water from the waste dewatering system, is added to the absorbers to maintain 15% solids in the absorber recirculating slurry. The recirculating slurry is pumped through the spray tower, from which it drains into a recirculation-oxidation tank. Air is sparged into the tank to oxidize the $CaSO_3 \cdot 1/2H_2O$ to $CaSO_4 \cdot 2H_2O$ (gypsum). A purge stream is withdrawn from each tank and pumped to the dewatering system.

The dewatering system consists of a thickener in which the purge stream is thickened to 40% solids and a rotary filter in which it is dewatered to 80% solids. The resulting cake is piled in a holding area from which it is trucked to the landfill.

To facilitate cost determinations and comparisons, the limestone slurry process is divided into six processing sections. The equipment list, giving the description and cost of each equipment item by section, is shown in Table 12. Each of these processing sections is described below.

Material Handling--

This area includes all of the facilities needed for receiving the raw limestone, areas for a 30-day storage stockpile, and a 24-hour in-process limestone storage area.

Feed Preparation--

Three trains (two operating and one spare) of gyratory crushers and wet ball mills to convert the raw limestone to a 90% minus 325 mesh; 60% solids slurry is included in this area. It also contains a product storage tank with an 8-hour capacity.

Particulate Removal--

Four 99.8% efficient ESP units sized for lignite are included in this area.

Gas Handling--

Included in this area is an inlet plenum supplying the four flue gas ducts to the absorbers. This area also includes the ductwork from the boiler air heater to the ESP and from the ESP to the inlet plenum, the emergency bypass ductwork around the absorber area, the ductwork from the absorbers to the stack plenum, and four ID fans to compensate for the pressure drop in the FGD system.

SO_2 Absorption--

Four trains of spray tower absorbers with presaturators, mist eliminators, recirculation tanks, and recirculating pumps are included. Oxidation air blowers and sparging rings for each recirculation tank are also included. Each absorber train is sized to handle one-third of the total flue gas volume being scrubbed.

TABLE 12. LIGNITE CASE

LIMESTONE SCRUBBING PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Mobile equipment	1	Bucket tractor	76,000	-
2.	Hopper, reclaim	1	7 ft x 4-1/4 ft x 2 ft deep, carbon steel	1,200	800
3.	Feeder, live limestone storage	1	Vibrating pan, 5 hp	5,500	500
4.	Pump, tunnel slump	1	Vertical, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined (1 operating, 1 spare)	2,400	800
5.	Conveyor, live limestone feed	1	Belt, 30 in. wide x 100 ft long, 2 hp, 100 tons/hr, 60 ft/min	22,900	1,400
6.	Conveyor, live limestone feed (inclined)	1	Belt, 30 in wide x 190 ft long, 40 hp, 35 ft lift, 100 tons/hr, 60 ft/min	60,300	3,700
7.	Elevator, live limestone feed	1	Continuous bucket, 12 in. x 8 in. x 11-3/4 in., 75 hp, 90 ft lift, 100 tons/hr, 160 ft/min	57,800	6,700
8.	Bin, crusher feed	3	13 ft dia x 21 ft high, w/cover, carbon steel	43,300	24,100
9.	Conveyor, feed belt	1	Belt, 30 in. wide x 60 ft long, 7.5 hp, 100 tons/hr, 60 ft/min	20,500	1,400
10.	Tripper	1	1 hp, 30 ft/min	27,200	9,100
11.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system	7,800	2,600
	Subtotal			324,900	51,100

(continued)

TABLE 12 (continued)

Area 2--Feed Preparation					
	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Feeder, crusher	3	Weigh belt, 18 in. wide x 14 ft long, 2 hp, 3 tons/hr	49,600	2,300
2.	Crusher	3	Gyratory, 0 x 1-1/2 to 3/4 in., 75 hp, 3 tons/hr	297,100	6,500
3.	Ball mill	3	Wet, open system, 150 hp, 3.0 tons/hr	543,800	65,300
4.	Tank, mills product	3	10 ft dia x 10 ft high, 5,500 gal, open top, four 10 in. baffles, agitator supports, carbon steel, flakeglass lined	13,700	11,000
5.	Agitator, mills product tank	3	36 in. dia, 10 hp, neoprene coated	22,900	5,500
6.	Pump, mills product tank	3	Centrifugal, 12 gpm, 60 ft head, 1 hp, carbon steel, neoprene coated (2 operating, 1 spare)	7,700	2,700
7.	Tank, slurry feed	1	12 ft dia x 12 ft high, 10,800 gal, open top, four 12 in. baffles, agitator supports, carbon steel, flakeglass lined	6,500	5,400
8.	Agitator, slurry feed tank	1	48 in. dia, 16 hp, neoprene coated	13,100	1,100
9.	Pump, slurry feed tank	6	Centrifugal, 7 gpm, 60 ft head, 1/4 hp, carbon steel, neoprene lined	15,000	5,500
10.	Dust collecting system	3	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system	23,300	7,800
	Subtotal			992,700	113,100

(continued)

TABLE 12 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
Area 3--Particulate Removal				
1. ESP	4	99.8% removal efficiency SCA = 700	9,223,000	4,542,700
2. Conveyor, fly ash to particulate bin	1	Pneumatic, pressure-vacuum, 125 hp	84,000	30,500
3. Bin, particulate	2	28 ft dia x 28 ft high, w/cover, carbon steel	66,000	64,800
4. Vibrator	2	Bin activator, 10 ft dia	28,900	4,800
Subtotal			9,401,900	4,642,800
Area 4--Gas Handling				
1. Fans	4	Induced draft, 472,900 aft^3/min, 8.6 in. static head, 900 hp, fluid drive, double width, double inlet, Inconel	3,063,400	52,600
Subtotal			3,063,400	52,600
Area 5--SO_2 Absorption				
1. SO_2 Absorber	4	Spray tower, 27 ft long x 27 ft wide x 40 ft high, 1/4 in. carbon steel, neoprene lined; FRP spray header, 316 stainless steel chevron vane entrainment separator and nozzles (3 operating, 1 spare)	5,144,600	418,500
2. Tank, effluent-oxidation	4	42 ft dia x 42 ft high, 429,500 gal, open top, four 42 in. wide baffles, agitator supports, carbon steel, flakeglass lined (4 operating, 1 spare)	346,500	280,000

(continued)

TABLE 12 (continued)

Area 5 (continued) Item		No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
3.	Agitator, effluent-oxidation tank	4	168 in. dia, 100 hp, neoprene coated (3 operating, 1 spare)	474,000	155,600
4.	Pump, slurry recirculation	12	Centrifugal, 17,400 gpm, 100 ft head, 100 hp, carbon steel, neoprene lined (6 operating, 6 spare)	1,343,400	116,800
5.	Pump, presaturator recycle	8	Centrifugal, 1,735 gpm, 100 ft head, 80 hp, carbon steel, neoprene lined (3 operating, 5 spare)	90,600	28,500
6.	Pump, oxidation bleed	6	Centrifugal, 62 gpm, 60 ft head, 2.0 hp, carbon steel, neoprene lined (3 operating, 3 spare)	16,800	5,500
7.	Air blower, oxidation	4	1,022 sft^3/min, 125 hp (3 operating, 1 spare)	66,200	3,100
8.	Sparger, oxidation	4	21 ft dia ring (3 operating, 1 spare)	52,100	31,000
9.	Pump, makeup water	2	Centrifugal, 3,253 gpm, 200 ft head, 300 hp, carbon steel (1 operating, 1 spare)	31,700	3,600
10.	Soot blower	32	Air fixed	89,500	83,500
	Subtotal			7,655,400	1,126,100

(continued)

TABLE 12 (continued)

Area 6--Solids Separation

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Tank, thickener feed	1	21 ft dia x 42 ft high, 107,200 gal, open top, agitator supports, four 19 in. baffles, carbon steel, flakeglass lined	34,500	28,500
2.	Agitator, thickener feed tank	1	84 in. dia, 60 hp, neoprene coated	40,300	3,300
3.	Pump, thickener feed	2	Centrifugal, 187 gpm, 60 ft head, 5 hp, carbon steel, neoprene lined (1 operating, 1 spare)	8,100	2,700
4.	Thickener	1	Stainless steel tank, 33 ft dia x 5 ft high; concrete basin, 4 ft high	46,600	49,300
5.	Pump, thickener overflow	2	Centrifugal, 121 gpm, 75 ft head, 4 hp, carbon steel, neoprene lined (1 operating, 1 spare)	8,900	1,000
6.	Tank, thickener overflow	1	8-1/2 ft dia x 5 ft high 2,000 gal, open top, carbon steel	1,200	800
7.	Pump, thickener underflow	2	Centrifugal, 58 gpm, 6 ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	4,900	1,800
8.	Tank, filter feed	1	5-1/2 ft dia x 5-1/2 ft high, 963 gal, open top, carbon steel, flakeglass lined	1,300	1,100
9.	Agitator, filter feed tank	1	22 in. dia, 3 hp, neoprene coated	1,200	100
10.	Pump, filter feed slurry	3	Centrifugal, 29 gpm, 50 ft head, 1 hp, carbon steel, neoprene lined	8,000	2,700

(continued)

TABLE 12 (continued)

Area 6 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
11. Filter	3	Rotary vacuum, 5 ft dia x 5 ft face, 150 total hp	185,800	50,400
12. Pump, filtrate	4	Centrifugal, 19 gpm, 20 ft head, 1 hp, carbon steel, neoprene lined (2 operating, 2 spare)	16,500	1,900
13. Tank, filtrate surge	1	5 ft dia x 5 ft high, 638 gal, carbon steel	600	400
14. Pump, filtrate surge tank	2	Centrifugal, 39 gpm, head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	8,500	1,000
15. Conveyor, gypsum disposal	1	Belt, 14 in. wide x 75 ft long, 100 ft inclined, 1-1/2 hp, 11 tons/hr, 70 ft/min	37,100	3,500
Subtotal			403,500	148,500

Note: These costs represent equipment costs only. Costs for piping, electrical equipment, instruments, foundations, and other installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

Most equipment cost estimates are based on informal vendor quotes and TVA information.

Solids Separation--

This area includes a single train of equipment for recovering the gypsum from the absorber system effluent slurry. The major equipment consists of a thickener and a filter but other minor equipment such as tanks, pumps, agitators, and a conveyor are also included.

LOW-SULFUR WESTERN COAL CASE

The 0.7% sulfur western coal, with a heating value of 9,700 Btu/lb and 9.7% ash, is lower in both heating value and ash than the eastern coals used in this study. Equally important are the characteristics of the ash, which has a high calcium content and a high resistivity. Both of these factors are important determinants of flue gas cleaning costs, particularly in economic comparisons of systems using spray dryers and those using wet scrubbing with separate fly ash collection. The alkalinity of the ash can serve to supplement expensive (i.e., nonlimestone) absorbents. Separate fly ash collection using ESP's requires higher SCA's than would be used for high-sulfur eastern coals and, hence, more expensive ESP's. The designs of FGD systems prescribed by these factors thus differ from designs of the same process for eastern coals.

Soda Ash Spray Dryer Process

The process uses five trains of spray dryers without warm-gas bypass and one fabric filter baghouse. Each spray dryer is equipped with emergency hot-gas bypass ducts. Four ducts, each with an ID booster fan, connect the baghouse to the stack plenum. The flow diagram and layout are shown in Figures 19 and 20, and the material balance is shown in Table 13.

Dry bulk soda ash is received in self-unloading rail cars and stored as the monohydrate under a saturated 32% solution, a common industry practice that reduces storage volume. The solution is withdrawn and diluted to about 6% for use in the spray dryers. Three top-mounted rotary atomizers are used in each spray dryer. In addition to the absorbent solution, water is supplied to the atomizers. The absorbent feed rate is controlled by the SO_2 content of the cleaned flue gas, and the water rate is controlled by the temperature of the cleaned flue gas. In this manner both SO_2 emissions and stack temperature are controlled. Flue gas enters the spray dryers at about 300°F through manifolds in the top and passes downward, mixing with the atomized absorbent. The particle-laden flue gas leaves the spray dryer at about 165°F and passes down an inclined duct to the baghouse. Reaction of SO_2 and dry absorbent continues until the flue gas passes through the cake on the fabric filters. Essentially all of the soda ash is utilized. The accumulated sulfur-salt and fly ash cake on the filters is periodically dislodged by reverse air flow and falls into hoppers from which it is pneumatically conveyed to storage silos at a rate of about 46,000 lb/hr (900 ft^3/hr). It is trucked dry to the landfill empoundment.

To facilitate cost determinations and comparisons, the process is divided into six processing sections. The equipment list, giving the

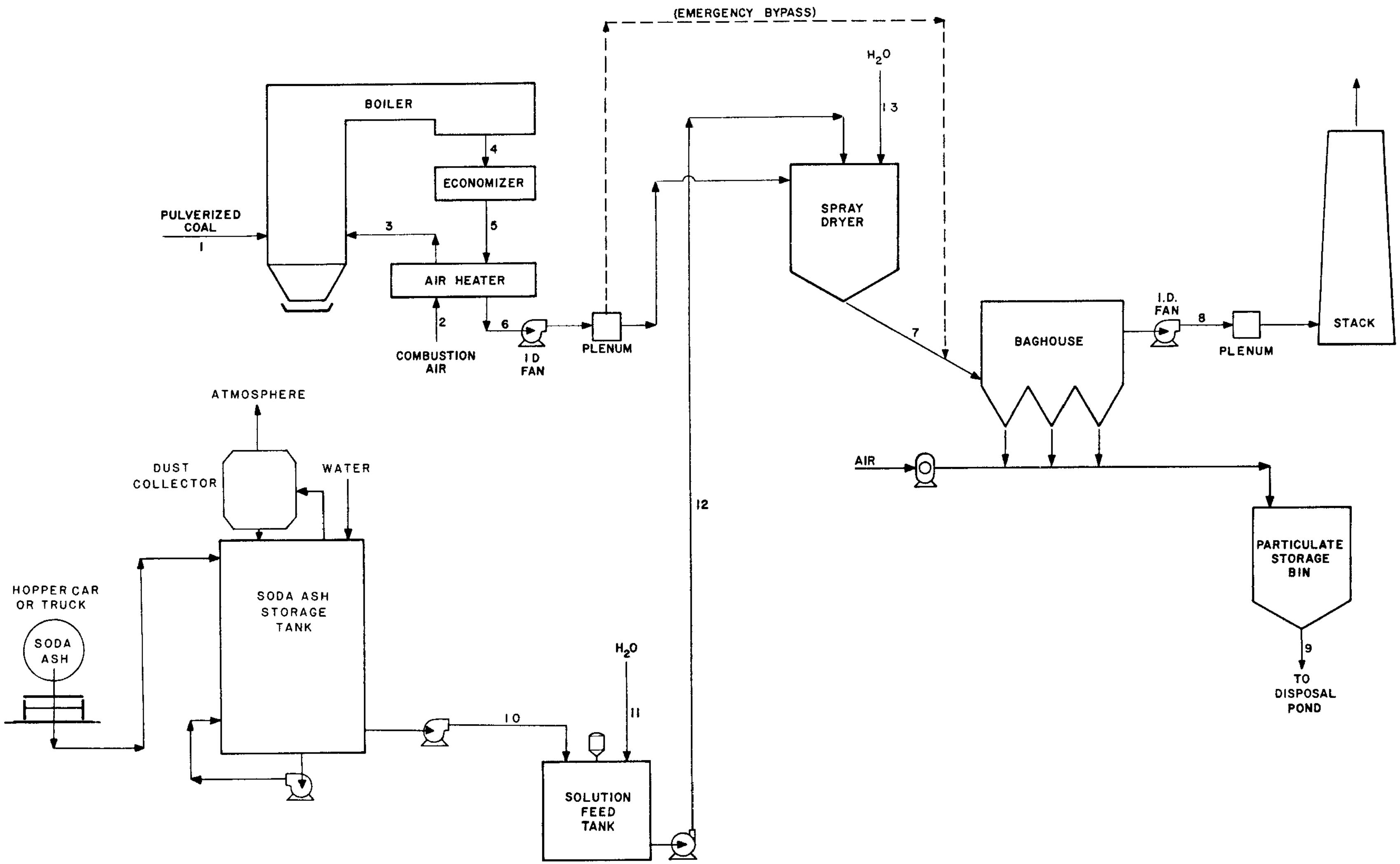

Figure 19. Low-sulfur western coal case. Soda ash spray dryer process. Flow diagram.

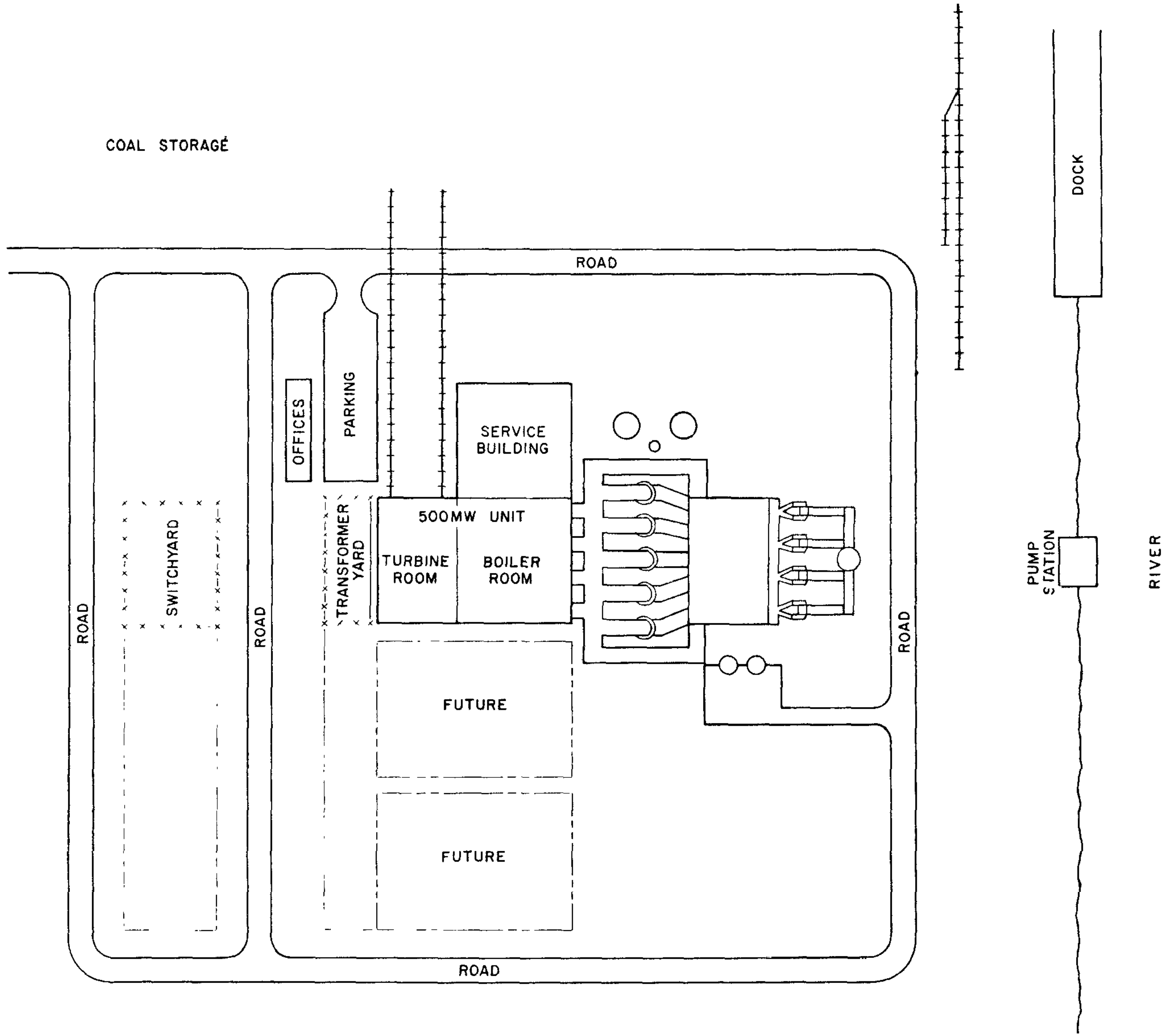

Figure 20. Low-sulfur western coal case. Soda ash spray dryer process. Plot plan.

TABLE 13. LOW-SULFUR WESTERN COAL CASE

SODA ASH SPRAY DRYER PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	489,700	5,119,000	4,419,000	4,897,000	4,897,000
2						
3	Flow rate, sft^3/min@60^{o}F		1,131,000	975,300	1,045,000	1,045,000
4	Temperature, ^{o}F		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to FGD system	Gas from spray dryer[a]	Gas to stack[a]	Waste to pond	Soda ash solution to mixing tank
1	Total stream, lb/hr	5,597,000	5,746,000	5,727,000	45,500	20,000
2						
3	Flow rate, sft^3/min@60^{o}F	1,200,000	1,246,000	1,252,000		
4	Temperature, ^{o}F	300	170	175		
5	Pressure, psig					
6	Flow rate, gpm					30
7	Specific gravity					1.34
8	pH					
9	Undissolved solids, %				100	
10						

	Stream No.	11	12	13		
	Description	Makeup water to mixing tank	Soda ash solution to spray dryer	Dilution water to spray dryer		
1	Total stream, lb/hr	81,100	101,100	19,890		
2						
3	Flow rate, sft^3/min@60^{o}F					
4	Temperature, ^{o}F			60		
5	Pressure, psig					
6	Flow rate, gpm	160	190	40		
7	Specific gravity	1.0	1.05	1.0		
8	pH					
9	Undissolved solids, %					
10						

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a. Includes air inleakage

description and cost of each equipment item by section, is shown in Table 14. These costs do not include the investment required for foundations, structures, electrical components, piping, instruments and controls, etc. Each of these processing sections is described below.

Material Handling--

The material handling section includes all of the equipment to receive soda ash by rail and to maintain a supply of soda ash solution to the solution feed tank. It includes an insulated soda ash storage tank with a 30-day capacity and dust collection, temperature control, and makeup water facilities.

Feed Preparation--

The feed preparation section includes a single train of the equipment required to dilute the saturated soda ash solution received from storage and pump the resulting solution to the spray dryers. A solution feed tank with a 4-hour capacity is provided.

Gas Handling--

Included in this area is an inlet plenum supplying the flue gas ducts that feed the spray dryer trains. This area also includes the ductwork for the hot-gas bypass, the ductwork from the air heater to the inlet plenum, the emergency bypass around the spray dryers, the ductwork from the spray dryers to the baghouse, and the ductwork from the baghouse to the stack plenum. Four ID fans are provided between the baghouse and the stack to compensate for the pressure drop through the FGD system.

SO_2 Absorption--

Five spray dryers are provided (four operating and one spare); each is sized to handle one-fourth of the total flue gas volume. The spray dryers are 48 feet in diameter and 54 feet high with individual flue gas manifolds for each of the three rotary atomizers.

Particulate Removal--

A single baghouse containing 28 compartments and the associated equipment is provided.

Particulate Handling--

Two particulate storage bins and the associated pneumatic conveyors are included to provide a 24-hour storage capacity for the waste.

Lime Spray Dryer Process

The lime spray dryer process for the low-sulfur western coal is exactly the same as the lignite. The only differences between the processes are the equipment sizes resulting from different flow rates. The low-sulfur western coal case burns 41% less coal than the lignite case, resulting in a 17% reduction in flue gas volume. The low-sulfur western coal also produces 23% less waste than the lignite because of the lower flow rates.

The flow diagram and plot plan for the lime spray dryer process are shown in Figures 21 and 22, and the material balance shown in Table 15.

TABLE 14. LOW-SULFUR WESTERN COAL CASE

SODA ASH SPRAY DRYER PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Car shaker	1	Top mounting with crane, 20 hp shaker, 7-1/2 hp crane	25,300	4,700
2.	Tank, soda ash solution storage	2	42-1/2 ft dia x 45 ft high, 423,000 gal, w/ cover, carbon steel, insulated	127,800	86,200
3.	Heater, solution	1	Coil type, 100 ft^2, carbon steel	900	300
4.	Pump, soda ash solution recirculating	2	Centrifugal, 30 gpm, 50 ft head, 1 hp, carbon steel, insulated (1 operating, 1 spare)	2,300	1,100
5.	Pump, soda ash solution feed	2	Centrifugal, 30 gpm, 50 ft head, 1 hp, carbon steel, insulated (1 operating, 1 spare)	2,300	1,100
6.	Dust collecting system	1	Bag filter, polypropylene bag, 4,000 aft^3/min, automatic shaker system	15,500	5,200
	Subtotal			174,100	98,600

Area 2--Feed Preparation

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Tank, solution feed	1	20 ft dia x 20 ft high, 47,000 gal, w/cover, four 20 in. baffles, carbon steel	15,000	9,900

(continued)

TABLE 14 (continued)

Area 2 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
2. Agitator, solution storage tank	1	80 in. dia, 20 hp, carbon steel	23,300	2,400
3. Pump, solution feed pump	6	Centrifugal, 48 gpm, 200 ft head, 5 hp, carbon steel (4 operating, 2 spare)	8,500	3,600
Subtotal			46,800	15,900

Area 3--Gas Handling

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Fan	4	Induced draft, 382,000 aft^3/min, 12 in. static head, 875 rpm, 1,250 hp, fluid drive, double width, double inlet, carbon steel (4 operating)	2,260,800	49,700
Subtotal			2,260,800	49,700

Area 4--SO_2 Absorption

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Spray dryer	5	48 ft dia x 54 ft high, 3 rotary atomizers, carbon steel (4 operating spray dryers and 1 spare)	5,405,000	709,000
Subtotal			5,405,000	709,000

(continued)

TABLE 14 (continued)

Area 5--Particulate Removal

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Baghouse	1	Automatic fabric filter, 28 compartments, 2.5 air-to-cloth ratio	8,262,000	2,971,500
	Subtotal			8,262,000	2,971,500

Area 6--Particulate Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Conveyor, particulate feed	1	Pneumatic, pressure-vacuum, 250 hp	243,100	78,200
2.	Bin, particulate storage	2	27-1/2 ft dia x 30 ft high, 17,800 ft^3, 60^o cone, w/cover, carbon steel	63,200	42,300
	Vibrator	2	Bin activator, 10 ft dia	28,900	4,800
	Subtotal			335,200	125,300

Basis: Most equipment cost estimates are based on informal vendor quotes and TVA information. The only exception is the cost for the spray dryers which is based on information supplied by the vendors.

These costs represent equipment costs only. Costs for piping, electrical equipment, instruments, foundations, and other installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

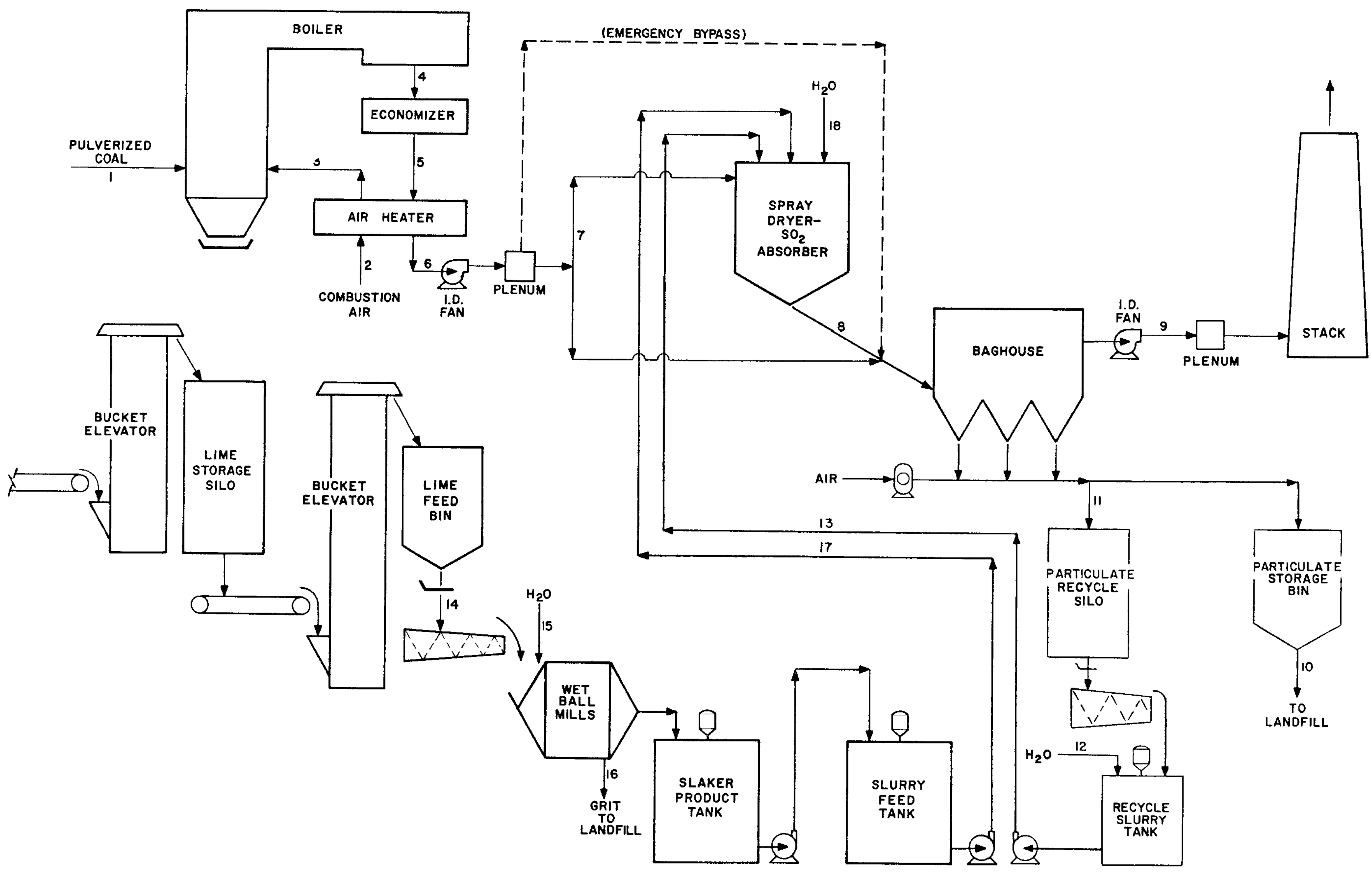

Figure 21. Low-sulfur western coal case. Lime spray dryer process. Flow diagram.

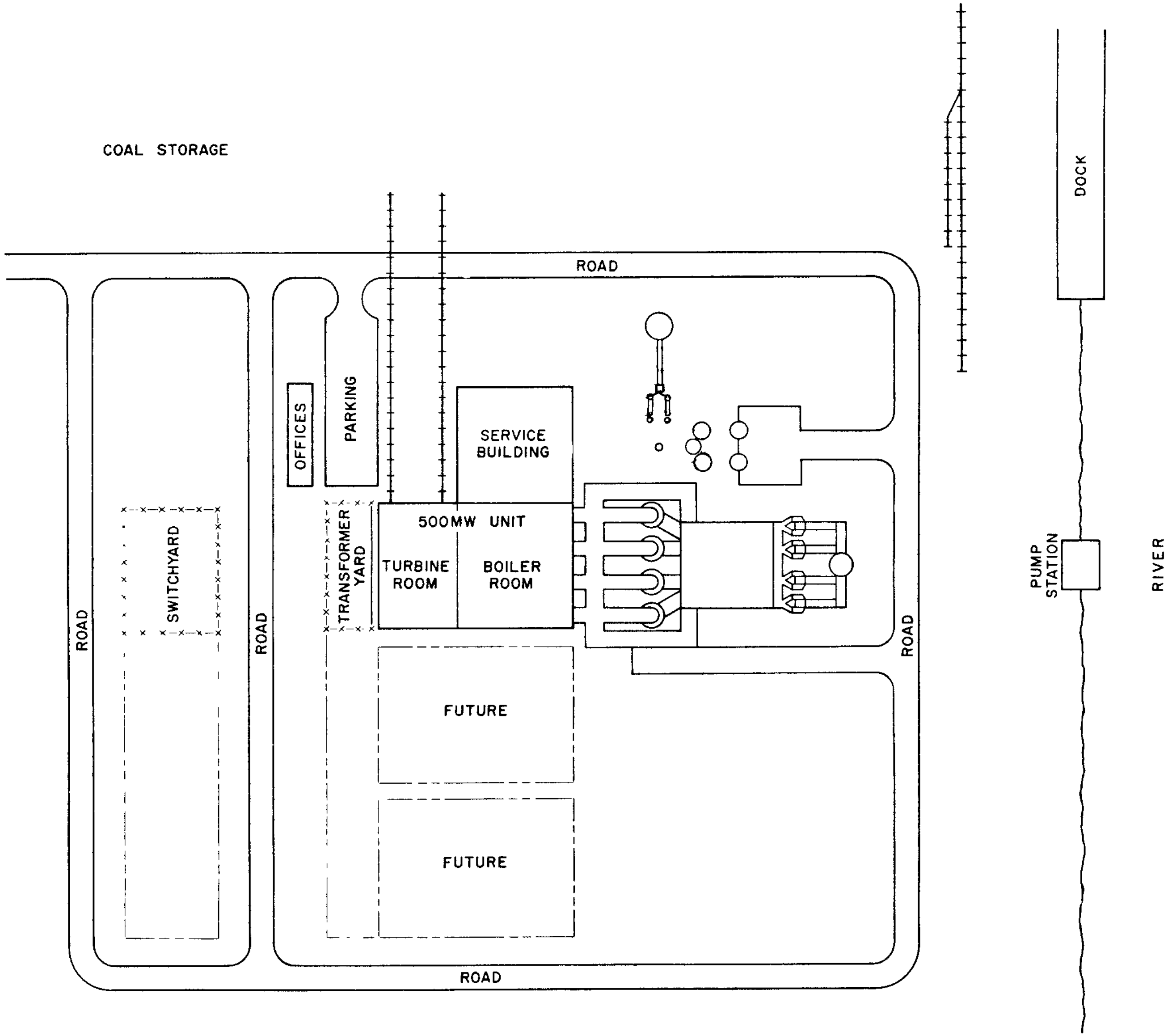

Figure 22. Low-sulfur western coal case. Lime spray dryer process. Plot plan.

TABLE 15. LOW-SULFUR WESTERN COAL CASE

LIME SPRAY DRYER PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	489,700	5,119,000	4,419,000	4,897,000	4,897,000
2						
3	Flow rate, sft^3/min@60^{o}F		1,131,000	975,300	1,045,000	1,045,000
4	Temperature, ^{o}F		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to FGD system	Gas to spray dryer	Gas from spray dryer[a]	Gas to stack[a]	Waste to landfill
1	Total stream, lb/hr	5,597,000	4,547,000	4,743,000	5,715,000	44,650
2						
3	Flow rate, sft^3/min@60^{o}F	1,200,000	975,000	1,020,000	1,250,000	
4	Temperature, ^{o}F	300	300	160	175	
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Waste to recycle particulate silo	Makeup water to recycle slurry tank	Recycle slurry to spray dryer	Makeup lime to slaker	Makeup water to slaker
1	Total stream, lb/hr	55,450	83,150	138,600	3,661	11,350
2						
3	Flow rate, sft^3/min@60^{o}F					
4	Temperature, ^{o}F					
5	Pressure, psig					
6	Flow rate, gpm		166			23
7	Specific gravity					
8	pH					
9	Undissolved solids, %			40		
10						

	Stream No.	16	17	18		
	Description	Grit to landfill	Lime slurry to spray dryer	Dilution water to spray dryer		
1	Total stream, lb/hr	366	14,650	19,890		
2						
3	Flow rate, sft^3/min@60^{o}F					
4	Temperature, ^{o}F			60		
5	Pressure, psig					
6	Flow rate, gpm			40		
7	Specific gravity					
8	pH					
9	Undissolved solids, %		22.5			
10						

a. Includes air inleakage.

In the low-sulfur western coal case, three of the four trains process 81% of the flue gas at about 83% SO_2 removal efficiency. When the clean flue gas is combined with the bypass flue gas, it produces an overall removal efficiency of 70% and a flue gas temperature of 165°F in the baghouse and 175°F emitted to the stack. In the low-sulfur western coal case 55% of the waste from the baghouse is recycled while the other 45% is stored in silos and trucked to the landfill.

The equipment list shown in Table 16 is divided into the same six processing areas used in lignite case. The area-by-area description for the lignite process applies to this process.

Limestone Scrubbing Process

The same basic design used for the limestone scrubbing process in the lignite case is used for the low-sulfur western coal case. The only difference is the equipment sizes due to the reduction in flow rate in the low-sulfur western coal case. Also, 22% of the flue gas is bypassed to the absorber outlet producing an overall 70% removal efficiency.

The flow diagram and plot plan for the low-sulfur western coal case limestone scrubbing process are shown in Figures 23 and 24, and the material balance is shown in Table 17. The equipment list is shown in Table 18. The same division by processing section and the same description used in the lignite case are used for the low-sulfur western coal case.

LOW-SULFUR EASTERN COAL CASE

The 0.7% sulfur eastern coal, with a heating value of 10,700 Btu/lb and 16% ash, is representative of eastern bituminous coals. In comparison with the western coal, it has, in addition to a higher heating value, a lower moisture content and a lower calcium content, both of which are important in spray dryer FGD. It thus provides a direct comparison at the same coal sulfur content of western and eastern coals.

Lime Spray Dryer Process

The lime spray dryer process for the low-sulfur eastern coal is very similar in most design aspects to the design for the low-sulfur western coal. The primary design difference is the omission of waste recycle for the low-sulfur eastern coal process because the available alkalinity in the fly ash is too low to justify the added expense. The other differences between the processes are equipment size differences resulting from the different flow rates. Using the rates for the low-sulfur western coal process as the basis, the low-sulfur eastern coal process burns 17% less coal but this results in only a 2% reduction in flue gas volume. On the other hand, the low-sulfur eastern coal process uses 12% more lime because the absorbent is not supplemented by fly ash alkalinity. The low-sulfur eastern coal process also produces 24% more waste because of the higher ash content of the coal.

TABLE 16. LOW-SULFUR WESTERN COAL CASE

LIME SPRAY DRYER PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Car shaker	1	Top mounted with crane, 20 hp	74,700	11,300
2.	Car puller	1	25 hp with 5 hp return	64,600	24,200
3.	Hopper, unloading	1	12 ft x 12 ft x 2 ft bottom, 20 ft deep, carbon steel	3,700	2,600
4.	Pump, pit sump	3	Centrifugal, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	10,600	3,300
5.	Conveyor, lime unloading (enclosed)	1	Belt, 24 in. wide, 200 ft long, 2 hp, 100 tons/hr	14,600	1,000
6.	Elevator, storage silo	1	Continuous bucket, 16 in. x 8 in. x 11-3/4 in., 75 ft lift, 15 hp, 100 tons/hr, 160 ft/min	33,600	3,400
7.	Silo, lime storage	1	40 ft dia x 50 ft straight side, 62,800 ft^3, 45^o slope, carbon steel	94,300	86,000
	Vibrators	1	Bin activator, 10 ft dia	14,500	2,400
8.	Conveyor, live lime feed	1	Belt, 14 in. x 100 ft long, 2 hp, 16 tons/hr, 100 ft/min	11,400	1,000
9.	Elevator, live lime feed	2	Continuous bucket, 8 in. x 5-1/2 in. x 7-3/4 in., 35 ft lift, 2 hp, 16 tons/hr, 150 ft/min	16,800	2,700
10.	Bin, lime feed	2	11 ft dia x 12 ft straight side, 1,140 ft^3, 60^o slope, w/cover, carbon steel	11,600	10,700

(continued)

TABLE 16 (continued)

Area 1 (continued)

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
11.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 ft^3/min, 7-1/2 hp (1/2 cost in feed preparation area)	8,000	1,600
	Subtotal			358,400	150,200

Area 2--Feed Preparation

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Feeder, lime bin discharge	2	Vibrating, 3-1/2 hp, carbon steel	8,300	600
2.	Feeder, lime feed	2	Screw, 6 in. dia x 12 ft long, 1 hp, 2 tons/hr	4,000	3,300
3.	Slaker	2	Ball-mill type, 25 hp slaker, 1 hp classifier, 2.0 tons/hr	107,100	13,500
4.	Tank, slaker product	2	6 ft dia x 8 ft high, 1,700 gal, open top, four 6 in. baffles, agitator supports, carbon steel, neoprene lined	6,800	5,500
5.	Agitator, slaker product tank	2	2 turbines, 24 in. dia, 3 hp, neoprene coated	15,800	1,900
6.	Pump, slaker product tank	3	Centrifugal, 40 gpm, 50 ft head, 1-1/2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	5,300	2,400
7.	Tank, slurry feed	1	10 ft dia x 12 ft high, 7,100 gal, open top, four 10 in. baffles, agitator supports, carbon steel, neoprene lined	8,700	7,000
8.	Agitator, slurry feed tank	1	40 in. dia, 7-1/2 hp, neoprene coated	15,300	1,300

(continued)

TABLE 16 (continued)

Area 2 (continued)

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
9.	Pump, slurry feed tank	12	Centrifugal, 40 gpm, 100 ft head, 5 hp, carbon steel, neoprene lined (6 operating, 6 spare)	30,000	9,300
10.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 ft^3/min, 7-1/2 hp (1/2 cost in material handling area)	3,900	600
	Subtotal			205,200	45,400

Area 3--Gas Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Fan	4	Induced draft, 382,000 aft^3/min, 12 in. static head, 875 rpm, 1,250 hp, fluid drive, double width, double inlet (4 operating)	2,260,800	49,700
	Subtotal			2,260,800	49,700

Area 4--SO_2 Absorption

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Spray dryer	4	48 ft dia x 54 ft high, with 3 rotary atomizers, carbon steel (3 operating, 1 spare)	4,324,000	567,200
	Subtotal			4,324,000	567,200

(continued)

TABLE 16 (continued)

Area 5--Particulate Removal

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Baghouse	1	Automatic fabric filter, 28 compartments, 2.5 air-to-cloth ratio	8,262,000	2,971,500
	Subtotal			8,262,000	2,971,500

Area 6--Particulate Handling and Recycle

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Conveyor, particulate feed to bin	1	Pneumatic, pressure-vacuum, 250 hp	243,100	78,200
2.	Bin, particulate storage	2	24 ft dia x 25 ft straight side, 11,300 ft^3, 60^o slope, w/cover, carbon steel	49,800	49,300
	Vibrator	2	Bin activator, 10 ft dia	28,900	4,800
3.	Silo, particulate recycle	2	25 ft dia x 30 ft straight side, 14,700 ft^3, 60^o slope, w/cover, carbon steel	56,500	52,900
4.	Feeder, particulate	2	Vibrating, 3-1/2 hp, carbon steel	8,400	800
5.	Feeder, recycle slurry tank	2	Screw, 12 in. dia x 12 ft long, 5 hp, 50 tons/hr	30,800	4,700
6.	Tank, recycle slurry	1	21 ft dia x 23 ft high, 55,400 gal, open top, four 21 in. baffles, agitator supports, carbon steel, neoprene lined	37,200	29,500
7.	Agitator, recycle slurry tank	1	84 in. dia, 30 hp, neoprene coated	42,100	2,700

(continued)

TABLE 16 (continued)

Area 6 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
8. Pump, recycle slurry feed	12	Centrifugal, 80 gpm, 100 ft head, 10 hp, carbon steel, neoprene lined (6 operating, 6 spare)	49,100	11,800
Subtotal			545,900	234,700

Basis: Most equipment cost estimates are based on informal vendor quotes and TVA information. The only exception is the cost for the spray dryers which is based on information supplied by the vendors.

These costs represent equipment costs only. Costs for piping, electrical equipment, instruments, foundations, and other installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

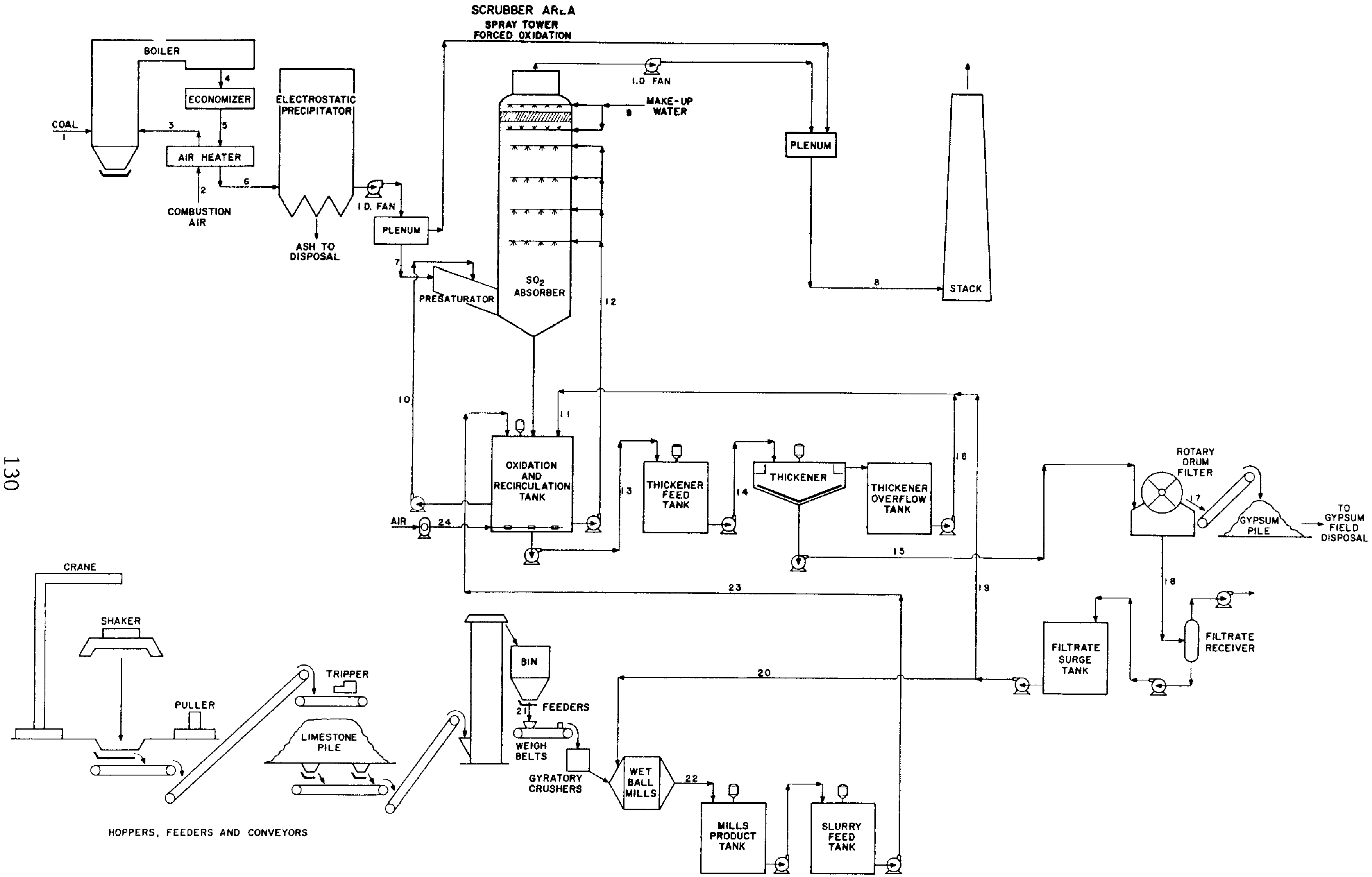

Figure 23. Low-sulfur western coal case. Limestone scrubbing process. Flow diagram.

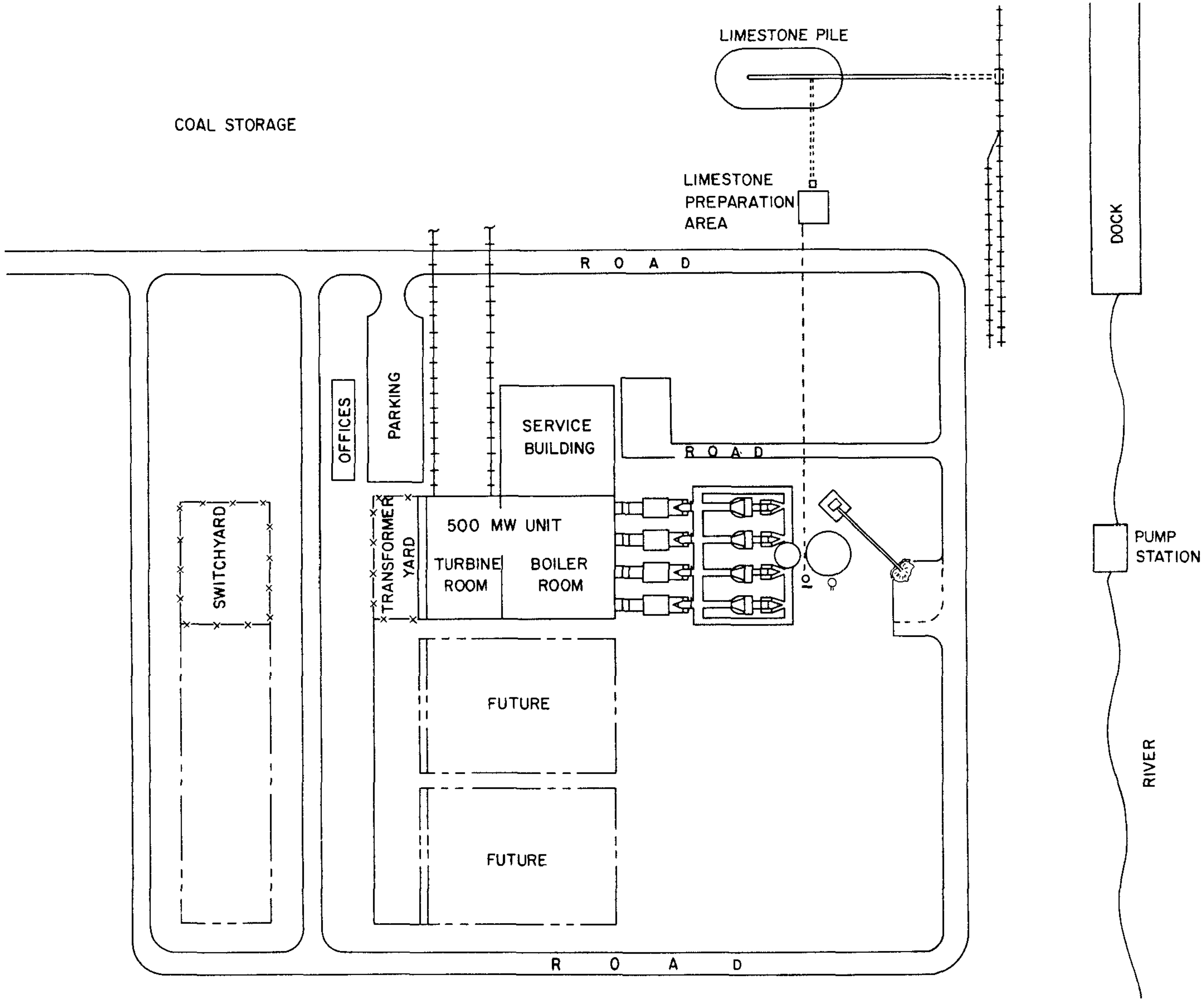

Figure 24. Low-sulfur western coal case. Limestone scrubbing process. Plot plan.

TABLE 17. LOW-SULFUR WESTERN COAL CASE

LIMESTONE SCRUBBING PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	489,700	5,119,000	4,419,000	4,897,000	4,897,000
2						
3	Flow rate, sft^3/min@60°F	.	1,131,000	975,300	1,045,000	1,045,000
4	Temperature, °F		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Gas to spray tower	Gas to stack	Makeup water to spray tower	Recycle slurry to presaturator
1	Total stream, lb/hr	5,597,000	4,333,000	5,743,000	191,600	2,305,000
2						
3	Flow rate, sft^3/min@60°F	1,200,000	934,800	1,265,000		
4	Temperature, °F	300	300	175		
5	Pressure, psig					
6	Flow rate, gpm				383	4,189
7	Specific gravity					1.1
8	pH					
9	Undissolved solids, %					15
10						

	Stream No.	11	12	13	14	15
	Description	Supernate to oxidation-recirculation tank	Recycle slurry to spray tower	Slurry to thickener feed tank	Slurry to thickener	Thickener underflow to filters
1	Total stream, lb/hr	46,760	46,090,000	62,690	62,690	23,510
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm	93	83,763	114	114	35
7	Specific gravity		1.1	1.1	1.1	1.33
8	pH					
9	Undissolved solids, %		15	15	15	40
10						

	Stream No.	16	17	18	19	20
	Description	Thickener overflow to oxidation-recirculation tank	Gypsum filter cake to disposal	Filtrate to filtrate surge tank	Filtrate to oxidation-recirculation tank	Filtrate to ball mills
1	Total stream, lb/hr	39,190	11,750	11,750	7,568	4,182
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm	78		23	23	8
7	Specific gravity		2.0			
8	pH					
9	Undissolved solids, %		80			
10						

(continued)

TABLE 17 (continued)

	Stream No.	21	22	23	24	
	Description	Limestone to weigh feeder	Limestone slurry to mills product tank	Limestone slurry to oxidation-recirculation tank	Air to oxidation-recirculation tank	
1	Total stream, lb/hr	6,590	10,460	10,770	8,577	
2						
3	Flow rate, $sft^3/min@60^oF$				1,893	
4	Temperature, ^oF					
5	Pressure, psig					
6	Flow rate, gpm		19	19		
7	Specific gravity		1.1	1.1		
8	pH					
9	Undissolved solids, %		60	60		
10						

1						
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TABLE 18. LOW-SULFUR WESTERN COAL CASE

LIMESTONE SCRUBBING PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Mobile equipment	1	Bucket tractor	76,000	-
2.	Hopper, reclaim	1	7 ft x 4-1/4 ft x 2 ft deep, carbon steel	1,200	300
3.	Feeder, live limestone storage	1	Vibrating pan, 5 hp	5,500	500
4.	Pump, tunnel sump	1	Vertical, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined (1 operating, 1 spare)	2,400	800
5.	Conveyor, live limestone feed	1	Belt, 30 in. wide x 100 ft long, 2 hp, 100 tons/hr, 60 ft/min	22,900	1,400
6.	Conveyor, live limestone feed (incline)	1	Belt, 30 in. wide x 190 ft long, 40 hp, 35 ft lift, 100 tons/hr, 60 ft/min	60,300	3,500
7.	Elevator, live limestone feed	1	Continuous bucket, 12 in. x 8 in. x 11-3/4 in., 75 hp, 90 ft lift, 100 tons/hr, 160 ft/min	66,500	6,700
8.	Bin, crusher feed	2	13 ft dia x 21 ft high, w/ cover, carbon steel	28,900	16,000
9.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system	7,800	2,600
	Subtotal			271,500	32,500

(continued)

TABLE 18 (continued)

Area 2--Feed Preparation

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Feeder, crusher	2	Weigh belt, 18 in. wide x 14 ft long, 2 hp, 3.0 tons/hr	33,100	1,600
2.	Crusher	2	Gyratory, 0 x 1-1/2 to 3/4 in., 75 hp, 3.0 tons/hr	198,100	4,300
3.	Ball mill	2	Wet, open system, 200 hp, 3.0 tons/hr	404,000	47,200
4.	Tank, mills product	2	10 ft dia x 10 ft high, 5,500 gal, open top, four 10 in. baffles, agitator supports, carbon steel, flakeglass lined	9,200	7,300
5.	Agitator, mills product tank	2	36 in. dia, 10 hp, neoprene coated	15,300	3,600
6.	Pump, mills product tank	2	Centrifugal, 14 gpm, 60 ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	5,200	1,900
7.	Tank, slurry feed	1	11 ft dia x 11 ft high, 7,000 gal, open top, four 11 in. baffles, agitator supports, carbon steel, flakeglass lined	4,900	4,100
8.	Agitator, slurry feed tank	1	44 in. dia, 15 hp, neoprene coated	9,800	800
9.	Pump, slurry feed tank	6	Centrifugal, 5 gpm, 60 ft head, 1/4 hp, carbon steel, neoprene lined (3 operating, 3 spare)	14,700	5,500
10.	Dust collecting system	2	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system	15,500	5,200
	Subtotal			709,800	81,500

(continued)

TABLE 18 (continued)

Area 3--Particulate Removal

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	ESP	4	99.8% removal efficiency SCA = 700	7,385,300	3,821,500
2.	Conveyor, fly ash to particulate bin	1	Pneumatic, pressure-vacuum, 125 hp	84,000	30,500
3.	Bin, particulate	2	26 ft dia x 25 ft high, w/cover, carbon steel	56,500	55,400
4.	Vibrator	2	Bin activator, 10 ft dia	28,900	4,800
	Subtotal			7,554,700	3,912,200

Area 4--Gas Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Fans	4	Induced draft, 349,000 aft^3/min, 8.4 in. static head, 890 rpm, 700 hp, fluid drive, double width, double inlet, Inconel (3 operating, 1 spare)	2,672,900	46,100
	Subtotal			2,672,900	46,100

(continued)

TABLE 18 (continued)

Area 5--SO_2 Absorption

	Item	No.	Description	Total material cost, 1982 \$	Total labor cost, 1982 \$
1.	SO_2 absorber	4	Spray tower, 25 ft long x 25 ft wide x 40 ft high, 1/4 in. carbon steel, neoprene lining; FRP spray headers, 316 stainless steel chevron vane entrainment separator and nozzles (3 operating, 1 spare)	4,313,100	360,700
2.	Tank, oxidation-effluent	4	39 ft dia x 39 ft high, 344,800 gal, open top, four 39 in. wide baffles, agitator supports, carbon steel, flakeglass lined (3 operating, 1 spare)	299,400	241,900
3.	Agitator, oxidation-effluent tank	4	156 in. dia, 100 hp, neoprene coated (3 operating, 1 spare)	405,800	133,200
4.	Pump, slurry recirculation	12	Centrifugal, 13,930 gpm, 100 ft head, 700 hp, carbon steel, neoprene lined (6 operating, 6 spare)	1,190,400	106,100
5.	Pumps, presaturator, recycle	8	Centrifugal, 1,400 gpm, 100 ft head, 75 hp, carbon steel, neoprene lined (3 operating, 5 spare)	83,900	25,900
6.	Pump, oxidation bleed	6	Centrifugal, 39 gpm, 60-ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	16,200	5,500
7.	Air blower, oxidation	4	620 sft^3/min, 75 hp (3 operating, 1 spare)	48,200	3,100
8.	Sparger, oxidation	4	19-1/2 ft dia ring (3 operating, 1 spare)	49,200	31,000

(continued)

TABLE 18 (continued)

Area 5 (continued)

Item		No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
9.	Pump, makeup water	2	Centrifugal, 2,620 gpm, 200 ft head, 250 hp, carbon steel (1 operating, 1 spare)	27,200	3,100
10.	Soot blowers	32	Air, retractable	89,500	83,500
	Subtotal			6,522,900	994,000

Area 6--Solids Separation

Item		No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Tank, thickener feed	1	19-1/2 ft dia x 39 ft high, 86,100 gal, open top, agitator supports, four 19 in. baffles, carbon steel, flakeglass lined	29,800	24,600
2.	Agitator, thickener feed tank	1	2 turbines, 78 in. dia, 50 hp, neoprene coated	34,500	2,800
3.	Pump, thickener feed	2	Centrifugal, 117 gpm, 60 ft head, 3 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,800	2,200
4.	Thickener	1	Stainless steel tank, 26 ft dia x 5 ft high; concrete basin, 4 ft high	34,400	37,500
5.	Pump, thickener overflow	2	Centrifugal, 75 gpm, 75 ft head, 2 hp, carbon steel (1 operating, 1 spare)	8,700	1,000

(continued)

TABLE 18 (continued)

Area 6 (continued)

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
6.	Tank, thickener overflow	1	6 ft dia x 6 ft high, 1,230 gal, open top, carbon steel	900	600
7.	Pump, filter feed slurry	2	Centrifugal, 35 gpm, 50 ft head, 1 hp, carbon steel (1 operating, 1 spare)	5,400	1,800
8.	Filter	2	Rotary vacuum, 11 ft dia x 11 ft face, 15 total hp	134,900	34,800
9.	Pump, filtrate	2	Centrifugal, 24 gpm, 20 ft head, 1 hp, carbon steel (1 operating, 1 spare)	8,300	900
10.	Tank, filtrate surge	1	4 ft dia x 4 ft high, 390 gal, carbon steel	400	300
11.	Pump, filtrate surge tank	2	Centrifugal, 24 gpm, 85 ft head, 1 hp, carbon steel (1 operating, 1 spare)	8,400	900
12.	Conveyor, gypsum disposal	1	Belt, 14 in. wide x 50 ft long, 100 ft inclined, 1 hp, 5.9 tons/hr, 40 ft/min	32,900	3,100
	Subtotal			306,400	110,500

Note: These costs represent equipment costs only. Costs for piping, electrical equipment, instruments, foundations, and other installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

Most equipment cost estimates are based on informal vendor quotes and TVA information.

The flow diagram and plot plan for the lime spray dryer process are shown in Figures 25 and 26, and the material balance shown in Table 19. As in the low-sulfur western coal process, three of the four trains process 81% of the flue gas at 83% SO_2 removal efficiency. Thus, when the cleaned flue gas is combined with the bypassed flue gas, the process produces an overall removal efficiency of 70% and a flue gas temperature of 160°F in the baghouse and 170°F downstream of the ID booster fan. All waste collected in the baghouse is stored in silos, from which it is trucked to the landfill.

The equipment list is shown in Table 20. The same division of equipment into six processing areas as is used for the low-sulfur western coal process is used. Except that there is no waste recycle, the area-by-area description for the low-sulfur western coal process applies to this process.

Limestone Scrubbing Process

The limestone scrubbing process for the low-sulfur eastern coal case is similar in most design aspects to the low-sulfur western coal process. As in the lime spray dryer process, 17% less eastern coal is burned, compared to the western coal, and slightly less flue gas is produced. In contrast, slightly more SO_2 is emitted (a 92% SO_2 emission in the flue gas instead of 85% is used because of the low alkalinity of the fly ash) and about 30% more fly ash is produced. Because of the different flue gas composition, as compared with the western coal case, only 19% of the flue gas is bypassed and a small amount of indirect steam reheat is necessary to produce a 175°F temperature in the recombined flue gas. Other differences from the limestone scrubbing process described for the low-sulfur western coal consist of duct and equipment size differences. The flow diagram and plot plan for the low-sulfur eastern coal case limestone scrubbing process are shown in Figures 27 and 28, and the material balance is shown in Table 21. The equipment list is shown in Table 22. The same division by processing section is used as is used in the low-sulfur western coal case and the same descriptions apply.

HIGH-SULFUR EASTERN COAL CASE

The 3.5% sulfur eastern coal, with a heating value of 10,700 Btu/lb and 16% ash, is representative of eastern and midwestern coals widely used by utilities in these areas. It differs from the low-sulfur western coal in having a higher heating value and sulfur content and a lower moisture and calcium content. It provides a comparison of spray dryer FGD technology in what may be considered typical western and eastern coal applications. With the low-sulfur eastern coal it provides a comparison of different sulfur contents in otherwise similar coals.

Lime Spray Dryer Process

The lime spray dryer process for the high-sulfur eastern coal is similar to the designs for the low-sulfur western and eastern coals. As

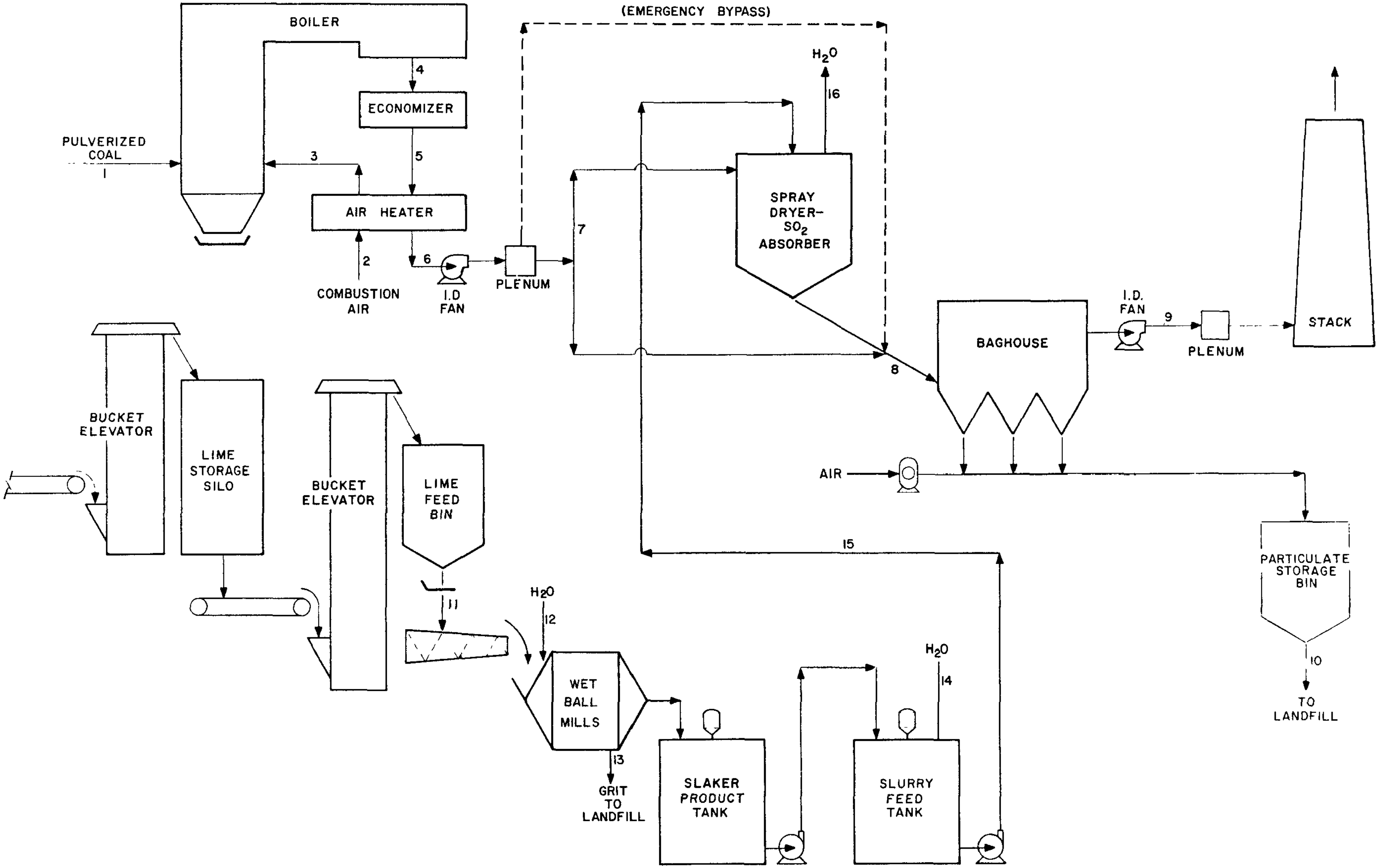

Figure 25. Low-sulfur eastern coal case. Lime spray dryer process. Flow diagram.

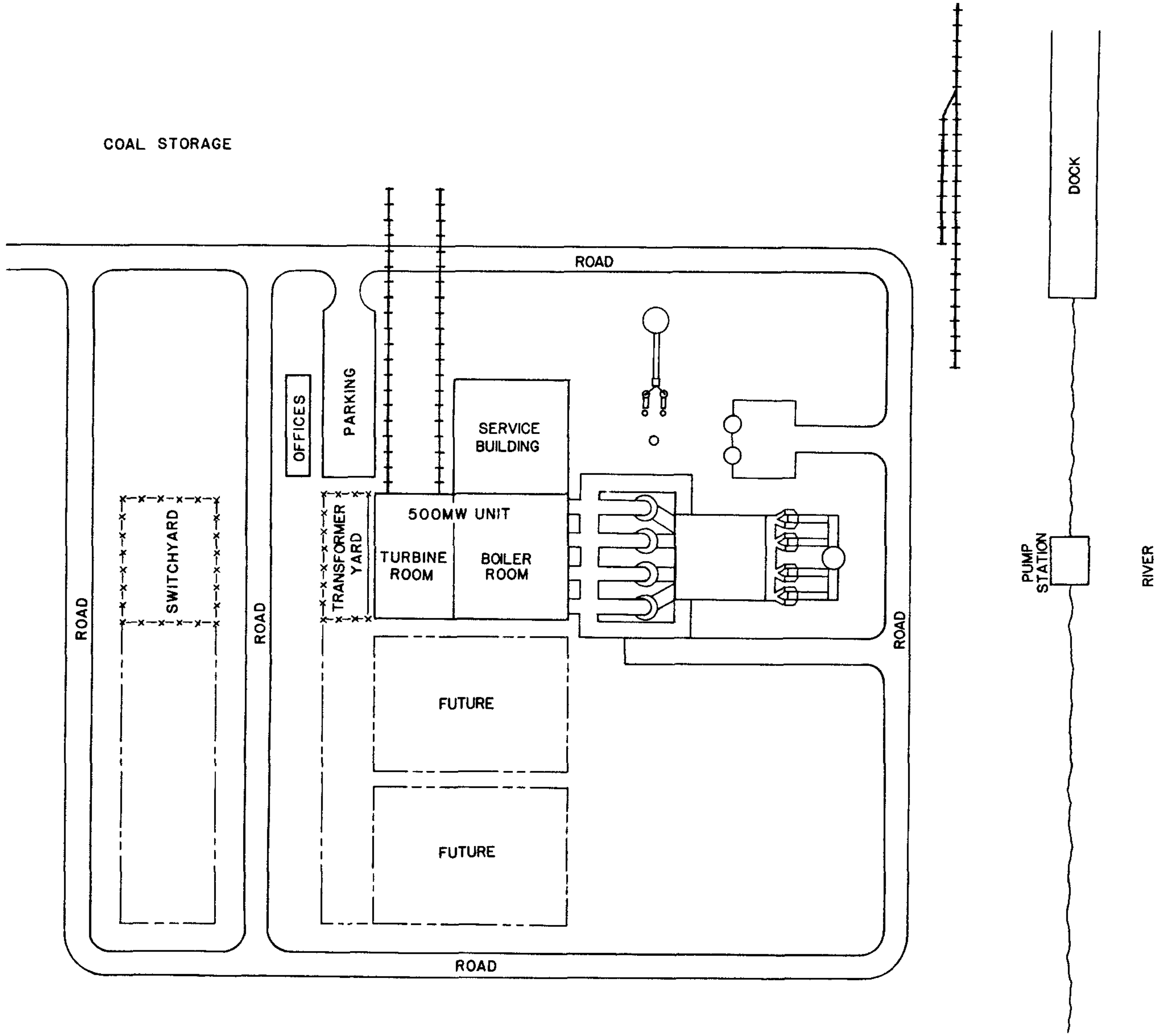

Figure 26. Low-sulfur eastern coal case. Lime spray dryer process. Plot plan.

TABLE 19. LOW-SULFUR EASTERN COAL CASE

LIME SPRAY DRYER PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,089,000	4,393,000	4,787,000	4,787,000
2						
3	Flow rate, $sft^3/min@60^oF$		1,123,000	969,400	1,005,000	1,005,000
4	Temperature, ^oF		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to FGD system	Gas to spray dryer	Gas from spray dryer[a]	Gas to stack[a]	Waste to landfill
1	Total stream, lb/hr	5,485,000	4,454,000	4,696,000	5,692,000	56,250
2						
3	Flow rate, $sft^3/min@60^oF$	1,158,000	940,000	1,023,000	1,245,000	
4	Temperature, ^oF	300	300	160	175	
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Makeup lime to slaker	Makeup water to slaker	Grit to landfill	Makeup water to slurry feed tank	Lime slurry to spray dryer
1	Total stream, lb/hr	4,110	11,400	410	102,600	117,700
2						
3	Flow rate, $sft^3/min@60^oF$					
4	Temperature, ^oF					
5	Pressure, psig					
6	Flow rate, gpm		23		205	228
7	Specific gravity					
8	pH					
9	Undissolved solids, %					3
10						

	Stream No.	16				
	Description	Dilution water to spray dryer				
1	Total stream, lb/hr	103,000				
2						
3	Flow rate, $sft^3/min@60^oF$					
4	Temperature, ^oF					
5	Pressure, psig					
6	Flow rate, gpm	205				
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

a. Includes air inleakage.

TABLE 20. LOW-SULFUR EASTERN COAL CASE

LIME SPRAY DRYER PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Car shaker	1	Top mounted with crane, 20 hp	74,700	11,300
2.	Car puller	1	25 hp with 5 hp return	64,600	24,200
3.	Hopper, unloading	1	12 ft x 12 ft x 2 ft bottom, 20 ft deep, carbon steel	3,700	2,600
4.	Pump, pit sump	3	Centrifugal, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	10,600	3,300
5.	Conveyor, lime unloading (enclosed)	1	Belt, 24 in. wide, 200 ft long, 2 hp, 100 tons/hr	14,600	1,000
6.	Elevator, storage silo	1	Continuous bucket, 16 in. x 8 in. x 11-3/4 in., 75 ft lift, 15 hp, 100 tons/hr, 160 ft/min	33,600	3,400
7.	Silo, lime storage	1	40 ft dia x 55 ft straight side, 69,000 ft^3, 45^o slope, carbon steel	100,300	90,100
	Vibrators	1	Bin activator, 10 ft dia	14,500	2,400
8.	Conveyor, live lime feed	1	Belt, 14 in. x 100 ft long, 2 hp, 16 tons/hr, 100 ft/min	11,400	1,000
9.	Elevator, live lime feed	2	Continuous bucket, 8 in. x 5-1/2 in. x 7-3/4 in., 35 ft lift, 2 hp, 16 tons/hr, 150 ft/min	16,800	2,700
10.	Bin, lime feed	2	11 ft dia x 12 ft straight side, 1,140 ft^3, 60^o slope, w/cover, carbon steel	11,600	10,700

(continued)

TABLE 20 (continued)

Area 1 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
11. Dust collecting system	1	Bag filter, polypropylene bag, 2,200 ft^3/min, 7-1/2 hp (1/2 cost in feed preparation area)	8,000	1,600
Subtotal			363,400	154,300

Area 2--Feed Preparation

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Feeder, lime bin discharge	2	Vibrating, 3-1/2 hp, carbon steel	8,300	600
2. Feeder, lime feed	2	Screw, 6 in. dia x 12 ft long, 1 hp, 1.0 tons/hr	4,000	3,300
3. Slaker	2	Ball-mill type, 25 hp slaker, 1 hp classifier, 2.0 tons/hr	107,100	13,500
4. Tank, slaker product	2	6 ft dia x 10 ft high, 2,100 gal, open top, four 6 in. baffles, agitator supports, carbon steel, neoprene lined	8,200	6,700
5. Agitator, slaker product tank	2	2 turbines, 24 in. dia, 3 hp, neoprene coated	15,800	1,900
6. Pump, slaker product tank	3	Centrifugal, 40 gpm, 50 ft head, 1-1/2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	5,300	2,400
7. Tank, slurry feed	1	22 ft dia x 22 ft high, 60,300 gal, open top, four 22 in. baffles, agitator supports, carbon steel, neoprene lined	36,600	29,600

(continued)

TABLE 20 (continued)

Area 2 (continued)

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
8.	Agitator, slurry feed tank	1	88 in. dia, 10 hp, neoprene coated	33,200	2,100
9.	Pump, slurry feed tank	12	Centrifugal, 228 gpm, 100 ft head, 10 hp, carbon steel, neoprene lined (6 operating, 6 spare)	57,900	22,900
10.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 ft^3/min, 7-1/2 hp, (1/2 cost in materials handling area)	8,000	1,600
	Subtotal			284,400	84,600

Area 3--Gas Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Fan	4	Induced draft, 380,000 aft^3/min, 12 in. static head, 875 rpm, 1,250 hp, fluid drive, double width, double inlet (4 operating)	2,260,800	49,700
	Subtotal			2,260,800	49,700

(continued)

TABLE 20 (continued)

Area 4--SO_2 Absorption

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Spray dryer	4	48 ft dia x 54 ft high, with 3 rotary atomizers, carbon steel (3 operating spray dryers, 1 spare)	4,324,000	567,200
Subtotal			4,324,000	567,200

Area 5--Particulate Removal

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Baghouse	1	Automatic fabric filter, 28 compartments, 2.5 air-to-cloth ratio	8,262,000	2,971,500
Subtotal			8,262,000	2,971,500

Area 6--Particulate Handling

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Conveyor, particulate feed to bin	1	Pneumatic, pressure-vacuum, 250 hp	243,100	78,200
2. Bin, particulate storage	2	26 ft dia x 26 ft straight side, 13,800 ft^3, 60^o slope, w/cover, carbon steel	53,500	54,800
Vibrator	2	Bin activator, 10 ft dia	28,900	4,800
Subtotal			325,500	137,800

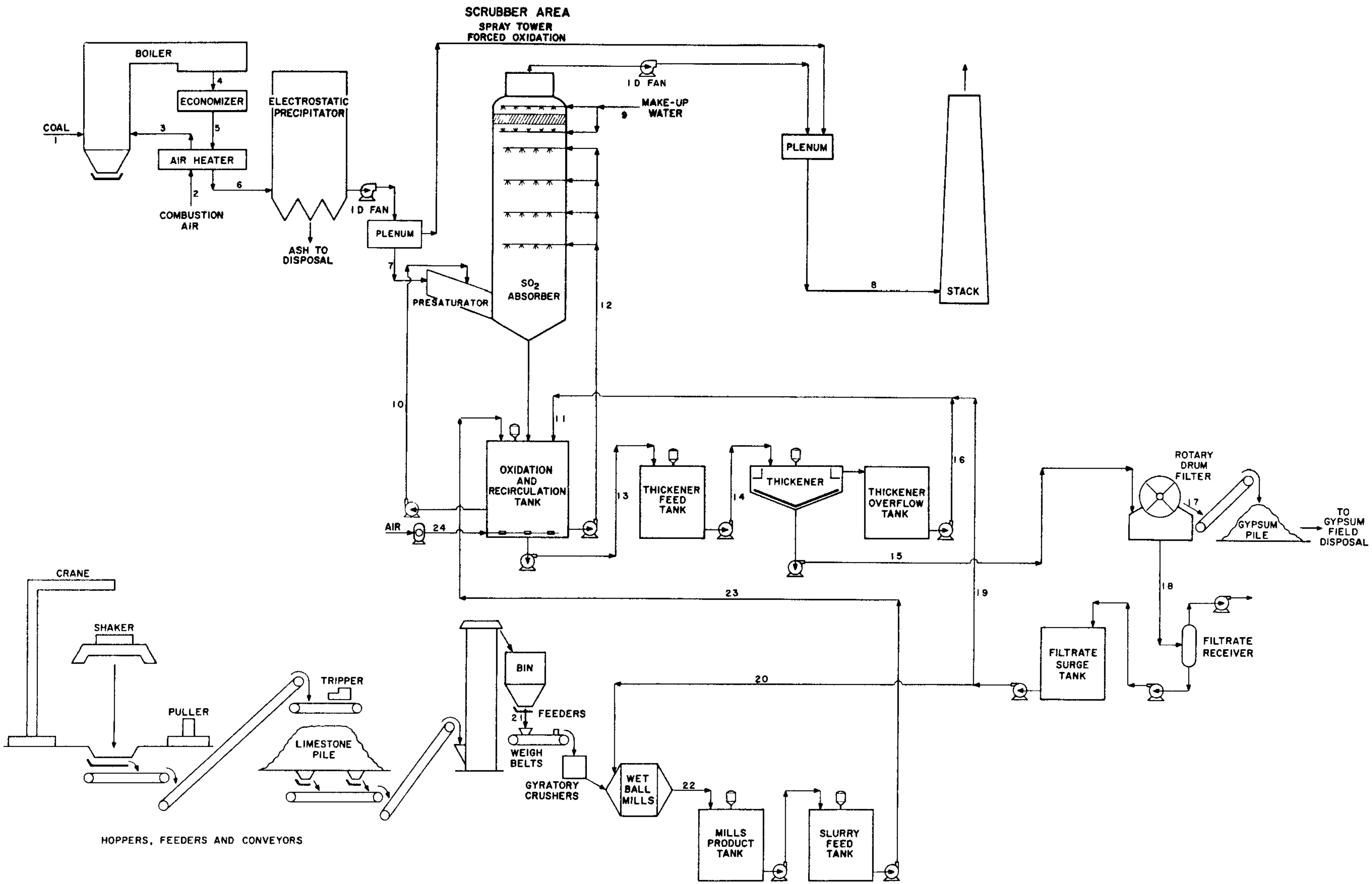

Figure 27. Low-sulfur eastern coal case. Limestone scrubbing process. Flow diagram.

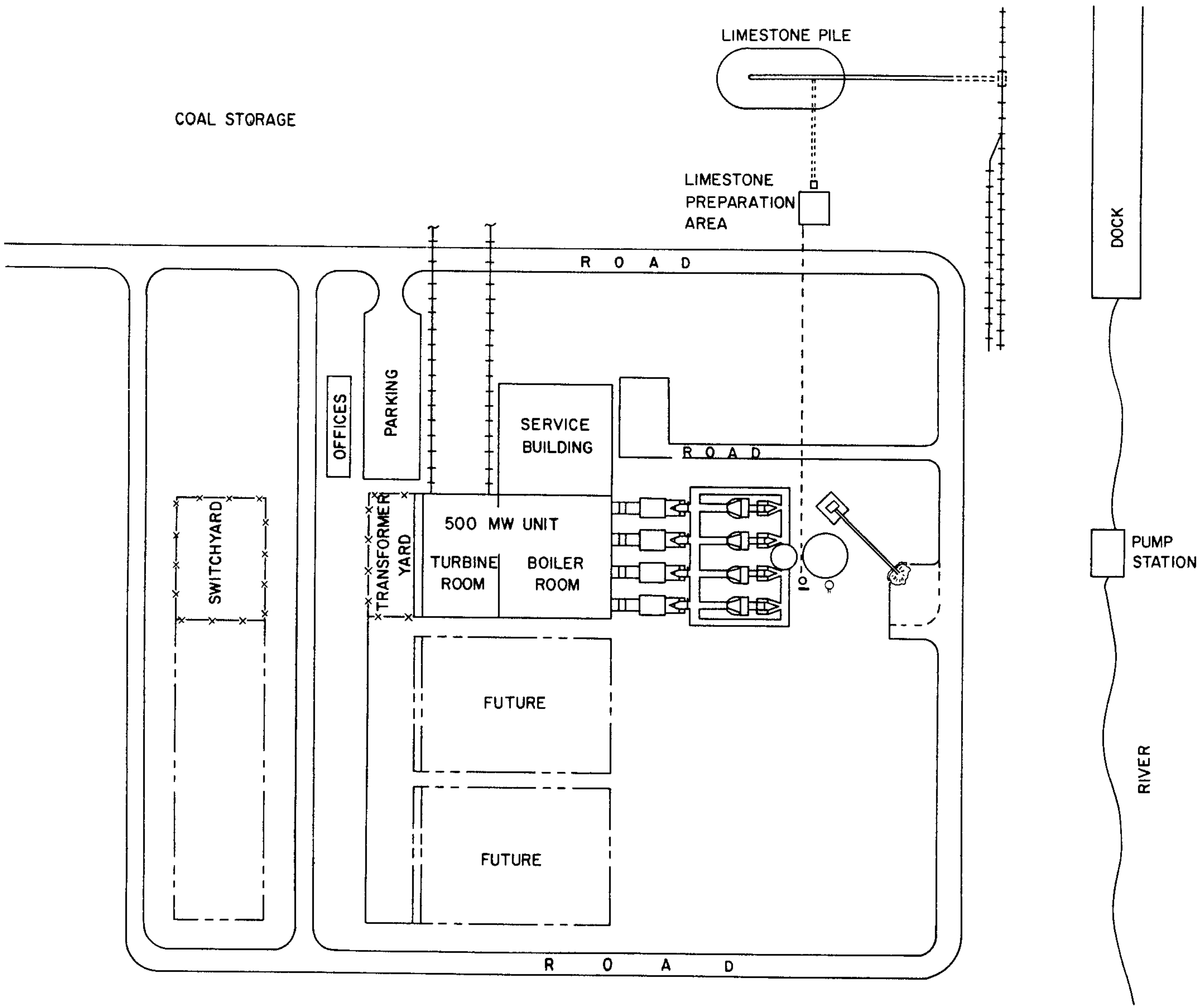

Figure 28. Low-sulfur eastern coal case. Limestone scrubbing process. Plot plan.

TABLE 21. LOW-SULFUR EASTERN COAL CASE

LIMESTONE SCRUBBING PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,089,000	4,393,000	4,787,000	4,787,000
2						
3	Flow rate, sft^3/min@60^{o}F		1,123,000	969,400	1,005,000	1,005,000
4	Temperature, ^{o}F		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Gas to spray tower	Gas to stack	Makeup wate. to spray tower	Recycle slurry to presaturator
1	Total stream, lb/hr	5,485,000	4,454,000	5,622,000	171,900	2,206,600
2						
3	Flow rate, sft^3/min@60^{o}F	1,159,000	940,700	1,225,000		
4	Temperature, ^{o}F	300	300	175		
5	Pressure, psig					
6	Flow rate, gpm				344	4,008
7	Specific gravity					1.1
8	pH					
9	Undissolved solids, %					15
10						

	Stream No.	11	12	13	14	15
	Description	Supernate to oxidation-recirculation tank	Recycle slurry to spray tower	Slurry to thickener feed tank	Slurry to thickener	Thickener underflow to filters
1	Total stream, lb/hr	46,130	44,141,000	63,800	63,800	23,940
2						
3	Flow rate, sft^3/min@60^{o}F					
4	Temperature, ^{o}F					
5	Pressure, psig					
6	Flow rate, gpm	92	80,220	116	116	36
7	Specific gravity		1.1	1.1	1.1	1.33
8	pH					
9	Undissolved solids, %		15	15	15	40
10						

	Stream No.	16	17	18	19	20
	Description	Thickener overflow to oxidation-recirculation tank	Gypsum filter cake to disposal	Filtrate to filtrate surge tank	Filtrate to oxidation-recirculation tank	Filtrate to ball mills
1	Total stream, lb/hr	38,350	11,970	11,970	7,784	4,186
2						
3	Flow rate, sft^3/min@60^{o}F					
4	Temperature, ^{o}F					
5	Pressure, psig					
6	Flow rate, gpm	77	12	24	16	8
7	Specific gravity		2.0			
8	pH					
9	Undissolved solids, %		80			
10						

(continued)

TABLE 21 (continued)

	Stream No.	21	22	23	24	
	Description	Limestone to weigh feeder	Limestone slurry to mills product tank	Limestone slurry to oxidation-recirculation tank	Air to oxidation-recirculation tank	
1	Total stream, lb/hr	6,602	10,480	10,480	8,740	
2						
3	Flow rate, sft^3/min@60°F				1,904	
4	Temperature, °F				80	
5	Pressure, psig					
6	Flow rate, gpm		19	19		
7	Specific gravity		1.1	1.1		
8	pH					
9	Undissolved solids, %		60	60		
10						

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TABLE 22. LOW-SULFUR EASTERN COAL CASE

LIMESTONE SCRUBBING PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Mobile equipment	1	Bucket tractor	76,000	-
2.	Hopper, reclaim	1	7 ft x 4-1/4 ft x 2 ft deep, carbon steel	1,200	800
3.	Feeder, live limestone storage	1	Vibrating pan, 5 hp	5,500	500
4.	Pump, tunnel sump	1	Vertical, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined (1 operating, 1 spare)	2,400	800
5.	Conveyor, live limestone feed	1	Belt, 30 in. wide x 100 ft long, 2 hp, 100 tons/hr, 60 ft/min	22,900	1,400
6.	Conveyor, live limestone feed (incline)	1	Belt, 30 in. wide x 190 ft long, 40 hp, 35 ft lift, 100 tons/hr, 60 ft/min	60,300	3,700
7.	Elevator, live limestone feed	1	Continuous bucket, 12 in. x 8 in. x 11-3/4 in., 75 hp, 90 ft lift, 100 tons/hr, 160 ft/min	66,500	6,700
8.	Bin, crusher feed	2	13 ft dia x 21 ft high, w/cover, carbon steel	28,900	16,000
9.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system	7,800	2,600
	Subtotal			271,500	32,500

(continued)

TABLE 22 (continued)

Area 2--Feed Preparation

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Feeder, crusher	2	Weigh belt, 18 in. wide x 14 ft long, 2 hp, 3.0 tons/hr	33,100	1,600
2.	Crusher	2	Gyratory, 0 x 1-1/2 to 3/4 in., 75 hp, 3.0 tons/hr	198,100	4,300
3.	Ball mill	2	Wet, open system, 190 hp, 3.0 tons/hr	410,400	47,800
4.	Tank, mills product	2	10 ft dia x 10 ft high, 5,500 gal, open top, four 10 in. baffles, agitator supports, carbon steel, flakeglass lined	9,200	7,300
5.	Agitator, mills product tank	2	36 in. dia, 10 hp, neoprene coated	15,300	3,700
6.	Pump, mills product tank	2	Centrifugal, 14 gpm, 60 ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	5,200	1,800
7.	Tank, slurry feed	2	11 ft dia x 11 ft high, 7,500 gal, open top, four 11 in. baffles, agitator supports, carbon steel, flakeglass lined	5,100	4,200
8.	Agitator, slurry feed tank	2	44 in. dia, 15 hp, neoprene coated	10,100	800
9.	Pump, slurry feed tank	6	Centrifugal, 4 gpm, 60 ft head, 1/4 hp, carbon steel, neoprene lined (3 operating, 3 spare)	14,800	5,500
10.	Dust collecting system	2	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system	15,500	5,200
	Subtotal			716,800	82,200

(continued)

TABLE 22 (continued)

Area 3--Particulate Removal

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. ESP	4	99.8% removal efficiency SCA = 700	7,385,300	3,821,500
2. Conveyor, fly ash to particulate bin	1	Pneumatic, pressure-vacuum, 125 hp	84,000	30,500
3. Bin, particulate	2	28 ft dia x 28 ft high, w/cover, carbon steel	66,000	64,800
4. Vibrator	2	Bin activator, 10 ft dia	28,900	4,800
Subtotal			7,564,700	3,921,600

Area 4--Gas Handling

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Fans	4	Induced draft, 347,000 aft^3/min, 8.5 in. static head, 890 rpm, 700 hp, fluid drive, double width, double inlet, Inconel (3 operating, spare)	2,701,400	46,500
Subtotal			2,701,400	46,500

(continued)

TABLE 22 (continued)

Area 5--SO_2 Absorption

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	SO_2 absorber	4	Spray tower, 24 ft long x 24 ft wide x 40 ft high, 1/4 in. carbon steel, neoprene lining; FRP spray headers, 316 stainless steel chevron vane entrainment separator and nozzles (3 operating, 1 spare)	4,303,900	360,100
2.	Tank, oxidation-effluent	4	39 ft dia x 39 ft high, 343,900 gal, open top, four 31-1/2 in. wide baffles, agitator supports, carbon steel, flakeglass lined (3 operating, 1 spare)	298,800	241,500
3.	Agitator, oxidation-effluent tank	4	156 in. dia, 100 hp, neoprene coated (3 operating, 1 spare)	405,100	133,000
4.	Pump, slurry recirculation	12	Centrifugal, 13,360 gpm, 100 ft head, 700 hp, carbon steel, neoprene lined (6 operating, 6 spare)	1,188,800	106,000
5.	Pumps, presaturator recycle	8	Centrifugal, 1,340 gpm, 100 ft head, 75 hp, carbon steel, neoprene lined (3 operating, 5 spare)	83,800	25,800
6.	Pump, oxidation bleed	6	Centrifugal, 41 gpm, 60 ft head, 1 hp, carbon steel, neoprene lined (3 operating, 3 spare)	16,200	5,500
7.	Air blower, oxidation	4	660 sft^3/min, 100 hp (3 operating, 1 spare)	50,000	3,100
8.	Sparger, oxidation	4	19-1/2 ft dia ring (3 operating, 1 spare)	49,200	31,000

(continued)

TABLE 22 (continued)

Area 5 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
9. Pump, makeup water	2	Centrifugal, 2,500 gpm, 200 ft head, 250 hp, carbon steel (1 operating, 1 spare)	27,200	3,100
10. Soot blower	48	Air, fixed	235,900	146,000
Subtotal			6,658,900	1,055,100

Area 6--Stack Gas Reheat

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Reheater	4	Inline steam type, 628 ft^2, Inconel 625	859,300	36,700
Subtotal			859,300	36,700

Area 7--Solids Separation

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Tank, thickener feed	1	19-1/2 ft dia x 39 ft high, 85,900 gal, open top, agitator supports, four 19 in. baffles, carbon steel, flakeglass lined	29,700	24,600
2. Agitator, thickener feed tank	1	2 turbines, 78 in. dia, 50 hp, neoprene coated	34,400	2,800
3. Pump, thickener feed	2	Centrifugal, 123 gpm, 60 ft head, 3 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,800	2,200

(continued)

TABLE 22 (continued)

Area 7 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
4. Thickener	1	Stainless steel tank, 27 ft dia x 5 ft high; concrete basin, 4 ft high	35,700	38,800
5. Pump, thickener overflow pumps	2	Centrifugal, 77 gpm, 75 ft head, 2 hp, carbon steel, neoprene lined (1 operating, 1 spare)	8,700	1,000
6. Tank, thickener overflow	1	7 ft dia x 5 ft high, 1,340 gal, open top, carbon steel	900	600
7. Pump, filter feed	2	Centrifugal, 38 gpm, 50 ft head, 1 hp, carbon steel (1 operating, 1 spare)	5,400	1,800
8. Filter	2	Rotary vacuum, 11 ft dia x 11 ft face, 15 total hp (1 operating, 1 spare)	138,800	35,300
9. Pump, filtrate	2	Centrifugal, 24 gpm, 20 ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	8,300	900
10. Tank, filtrate surge	1	4 ft dia x 4 ft high, 395 gal, carbon steel	400	300
11. Pump, filtrate surge tank	2	Centrifugal, 24 gpm, 85 ft head, 1 hp, carbon steel (1 operating, 1 spare)	8,400	900
12. Conveyor, gypsum disposal	1	Belt, 14 in. wide x 50 ft long, 100 ft inclined, 1.5 hp; 6 tons/hr, 40 ft/min	32,900	3,100
Subtotal			311,400	112,300

Note: These costs represent equipment costs only. Costs for piping, electrical equipment, instruments, foundations, and other installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

Most equipment cost estimates are based on informal vendor quotes and TVA information.

in the low-sulfur eastern coal process, no waste recycle is used. The major difference between this process and the low-sulfur coal processes is that about 89% SO_2 reduction is required, greatly reducing the amount of flue gas bypass that can be used. Bypass (excluding the emergency bypass) in this case is reduced to 4%. For comparison, the flue gas volume and composition for the high-sulfur eastern coal process is very similar, with the exception of sulfur content, to the low-sulfur eastern coal process. The design differences between the processes are the result of the higher SO_2 removal requirements. In this case, five spray dryer trains are used (four operating and one spare) to handle the additional flue gas volume passed through the spray dryers.

The flow diagram and plot plan are shown in Figures 29 and 30, and the material balance is shown in Table 23. Four of the five trains process 96% of the flue gas at an 89% SO_2 removal efficiency. The recombined flue gas passes through the baghouse at 160°F and is at 170°F downstream of the ID booster fan. All waste is stored in silos, from which it is trucked to the landfill. The equipment list is shown in Table 24. The same division of equipment by processing area is used as was used for the other coal cases. The same description of equipment and function also apply.

Limestone Scrubbing Process

The same basic design that was used in the low-sulfur cases is used for the high-sulfur eastern coal limestone scrubbing process. Since about 89% SO_2 removal is necessary, however, no flue gas bypass is used, and five absorber trains (four operating and one spare) are used to handle the additional flue gas scrubbed. Since all of the flue gas emerges from the absorbers at 127°F, full indirect steam reheat to 170°F is required. The flow diagram and plot plan are shown in Figures 31 and 32, and the material balance is shown in Table 25. With the exceptions discussed above, the process descriptions for the low-sulfur limestone scrubbing processes apply. The equipment list is shown in Table 26. The area-by-area discussion for the low-sulfur processes also apply to this process.

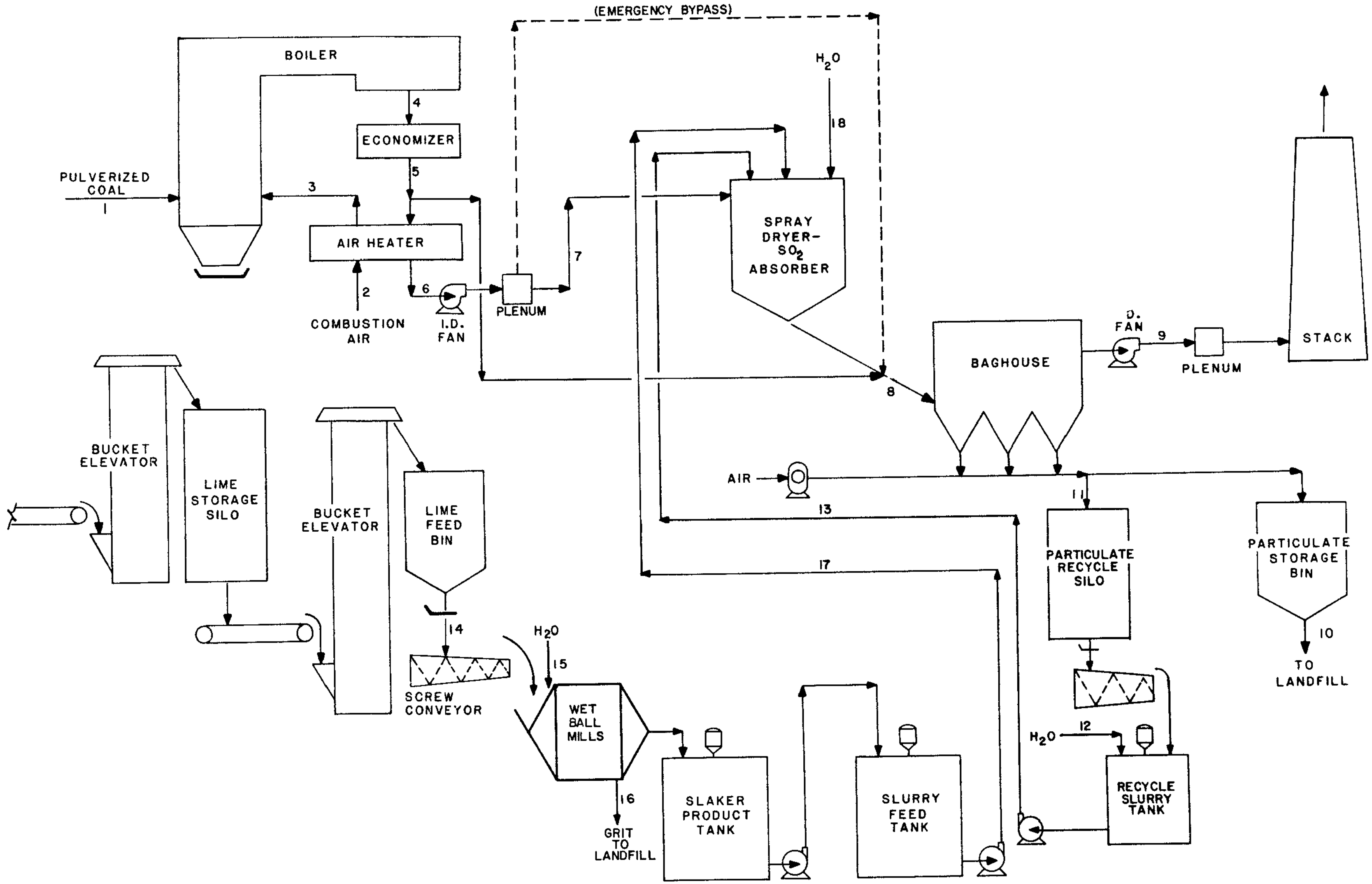

Figure 29. High-sulfur eastern coal case. Lime spray dryer process. Flow diagram.

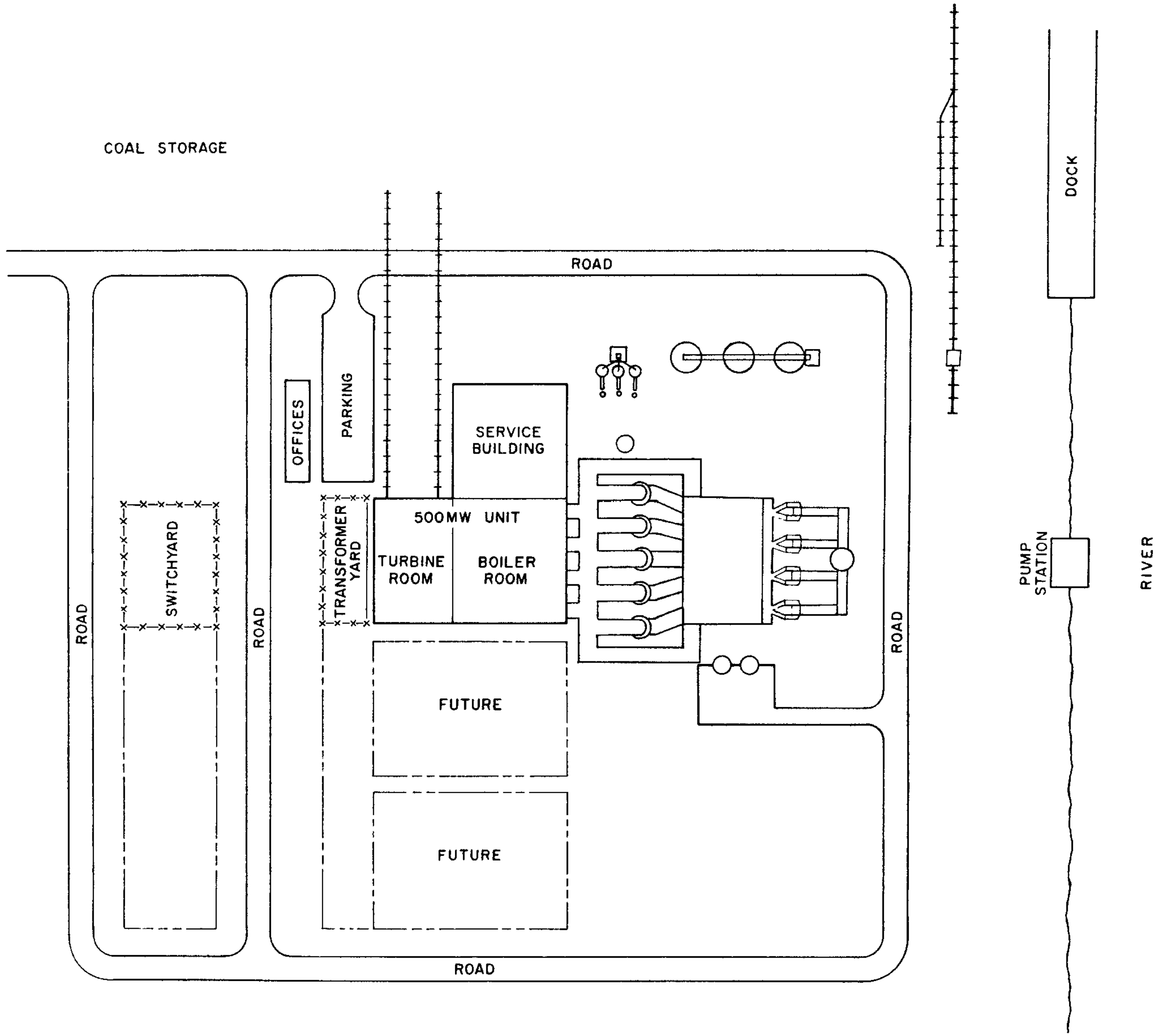

Figure 30. High-sulfur eastern coal case. Lime spray dryer process. Plot plan.

TABLE 23. HIGH-SULFUR EASTERN COAL CASE

LIME SPRAY DRYER PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,021,200	4,334,800	4,776,600	4,538,000
2						
3	Flow rate, $sft^3/min@60^oF$		1,108,000	956,600	1,011,000	960,300
4	Temperature, ^oF		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to FGD system	Gas to spray dryer	Gas to baghouse	Gas to stack	Waste to landfill
1	Total stream, lb/hr	5,224,000	5,224,000	5,723,000	5,648,000	110,100
2						
3	Flow rate, $sft^3/min@60^oF$	1,106,000	1,106,000	1,231,000	1,237,000	
4	Temperature, ^oF	300	300	170	180	
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Waste to recycle particulate silo	Makeup water to recycle slurry tank	Recycle slurry to spray dryer	Makeup lime to slaker	Makeup water to slaker
1	Total stream, lb/hr	31,200	46,800	78,000	40,900	126,700
2						
3	Flow rate, $sft^3/min@60^oF$					
4	Temperature, ^oF		60			
5	Pressure, psig					
6	Flow rate, gpm		94			253
7	Specific gravity					
8	pH					
9	Undissolved solids, %			40		
10						

	Stream No.	16	17	18		
	Description	Grit to landfill	Lime slurry to spray dryer	Dilution water to spray dryer		
1	Total stream, lb/hr	4,100	163,500	30,590		
2						
3	Flow rate, $sft^3/min@60^oF$					
4	Temperature, ^oF			60		
5	Pressure, psig					
6	Flow rate, gpm			60		
7	Specific gravity			1.0		
8	pH					
9	Undissolved solids, %		22.5			
10						

TABLE 24. HIGH-SULFUR EASTERN COAL CASE

LIME SPRAY DRYER PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Car shaker	1	Top mounted with crane, 20 hp	74,700	11,300
2.	Car puller	1	25 hp with 5 hp return	64,600	24,200
3.	Hopper, unloading	1	12 ft x 12 ft x 2 ft bottom, 20 ft deep, carbon steel	3,700	2,600
4.	Pump, pit sump	3	Centrifugal, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	10,600	3,300
5.	Conveyor, lime unloading (enclosed)	1	Belt, 24 in. wide, 200 ft long, 2 hp, 100 tons/hr, 200 ft/min	14,600	1,000
6.	Elevator, lime storage	1	Continuous bucket, 16 in. x 8 in. x 11-3/4 in., 112 ft lift, 25 hp, 100 tons/hr, 160 ft/min	42,900	5,300
7.	Conveyor, lime storage (enclosed, silo mounted)	1	Belt, 24 in. wide x 170 ft long, 3 hp, 100 tons/hr, 200 ft/min	33,300	9,600
8.	Tripper	1	5 hp, 30 ft/min	33,000	9,300
9.	Silo, lime storage	3	50 ft dia x 90 ft straight side, 178,500 ft^3, 45° slope, concrete	449,700	1,225,500
	Vibrators	3	Bin activator, 10 ft dia	43,400	7,100
10.	Conveyor, live lime feed	1	Belt, 18 in. x 215 ft long, 2 hp, 40 tons/hr, 150 ft/min	28,500	2,200

(continued)

TABLE 24 (continued)

Area 1 (continued)

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
11.	Elevator, live lime feed	3	Continuous, bucket 14 in. x 8 in. x 11-3/4 in., 70 ft lift, 5 hp, 40 tons/hr, 160 ft/min	54,900	8,100
12.	Bin, feed	3	20 ft dia x 30 ft straight side, 8,900 ft^3, 60^o slope, w/cover, carbon steel	62,200	56,100
13.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 ft^3/min, 7-1/2 hp, (1/2 cost in feed preparation area)	8,000	1,600
	Subtotal			924,100	1,366,600

Area 2--Feed Preparation

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Feeder, lime	3	Vibrating, 3-1/2 hp, carbon steel	12,500	900
2.	Conveyor, lime feed	3	Screw, 9 in. dia x 12 ft long, 2 hp, 10.2 tons/hr	7,500	6,200
3.	Slaker	3	Ball-mill type, 125 hp slaker, 3 hp classifier, 11.0 tons/hr	600,100	30,100
4.	Tank, slaker product	3	8 ft dia x 10 ft high, 3,500 gal, open top, four 8 in. baffles, agitator supports, carbon steel, neoprene lined	56,000	48,000
5.	Agitator, slaker product tank	3	32 in. dia, 3 hp, neoprene coated	23,700	2,800

(continued)

TABLE 24 (continued)

Area 2 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
6. Pump, slaker product tank	3	Centrifugal, 211 gpm, 100 ft head, 10 hp, carbon steel, neoprene lined (2 operating, 1 spare)	14,500	5,800
7. Tank, slurry feed	1	27-1/2 ft dia x 30 ft high, 131,000 gal, open top, four 27-1/2 in. baffles, agitator supports, carbon steel, neoprene lined	51,700	40,700
8. Agitator, slurry feed tank	1	108 in. dia, 40 hp, neoprene coated	73,200	3,600
9. Pump, slurry feed tank	15	Centrifugal, 211 gpm, 100 ft head, 10 hp, carbon steel, neoprene lined (8 operating, 7 spare)	74,800	30,300
10. Dust collecting system	1	Bag filter, polypropylene bag, 2,200 ft/min, 7-1/2 hp (1/2 cost in material handling area)	8,000	1,600
Subtotal			922,000	170,000

Area 3--Gas Handling

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1. Fan	4	Induced draft, 368,700 aft^3/min, 12 in. static head, 700 rpm, 1,250 hp, fluid drive, double width, double inlet	2,260,800	49,700
Subtotal			2,260,800	49,700

(continued)

TABLE 24 (continued)

Area 4--SO_2 Absorption

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Spray dryer	5	48 ft dia x 54 ft high, with 4 rotary atomizers, carbon steel (4 operating, 1 spare)	5,405,000	709,000
	Subtotal			5,405,000	709,000

Area 5--Particulate Removal

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Baghouse	1	Automatic fabric filter 26 compartments, 2.5 air-to-cloth ratio	8,032,600	2,921,100
	Subtotal			8,032,600	2,921,100

Area 6--Particulate Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Conveyor, particulate feed to bin	1	Pneumatic, pressure-vacuum, 350 hp	323,600	96,300
2.	Bin, particulate storage	2	29 ft dia x 37 ft straight side, 24,400 ft^3, 60^o slope, w/cover, carbon steel	79,800	75,500
	Vibrator	2	Bin activator, 10 ft dia	28,900	4,800

(continued)

TABLE 24 (continued)

Area 6 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
3. Silo, particulate recycle	2	20 ft dia x 24 ft straight side, 7,600 ft^3, 60° slope, w/cover, carbon steel	36,100	33,900
4. Feeder, particulate recycle	2	Vibrating, 3-1/2 hp, carbon steel	8,400	800
5. Feeder, recycle slurry tank	2	Screw, 12 in. dia x 12 ft long, 5 hp, 50 tons/hr	30,800	4,700
6. Tank, recycle slurry	1	19 ft dia x 22 ft high, 46,700 gal, open top, four 19 in. baffles, agitator supports, carbon steel, neoprene lined	26,700	21,400
7. Agitator, recycle slurry tank	1	76 in. dia, 20 hp, neoprene coated	26,300	2,700
8. Pump, recycle slurry feed	12	Centrifugal, 41 gpm, 100 ft head, 5 hp, carbon steel, neoprene lined	30,000	9,000
Subtotal			590,600	249,100

Note: These costs represent equipment costs only. Costs for piping, electrical, instruments, foundations, and other installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

Most equipment cost estimates are based on informal vendor quotes and internal TVA information. The only exceptions are the costs for the spray dryers and the baghouse which were provided by the vendors.

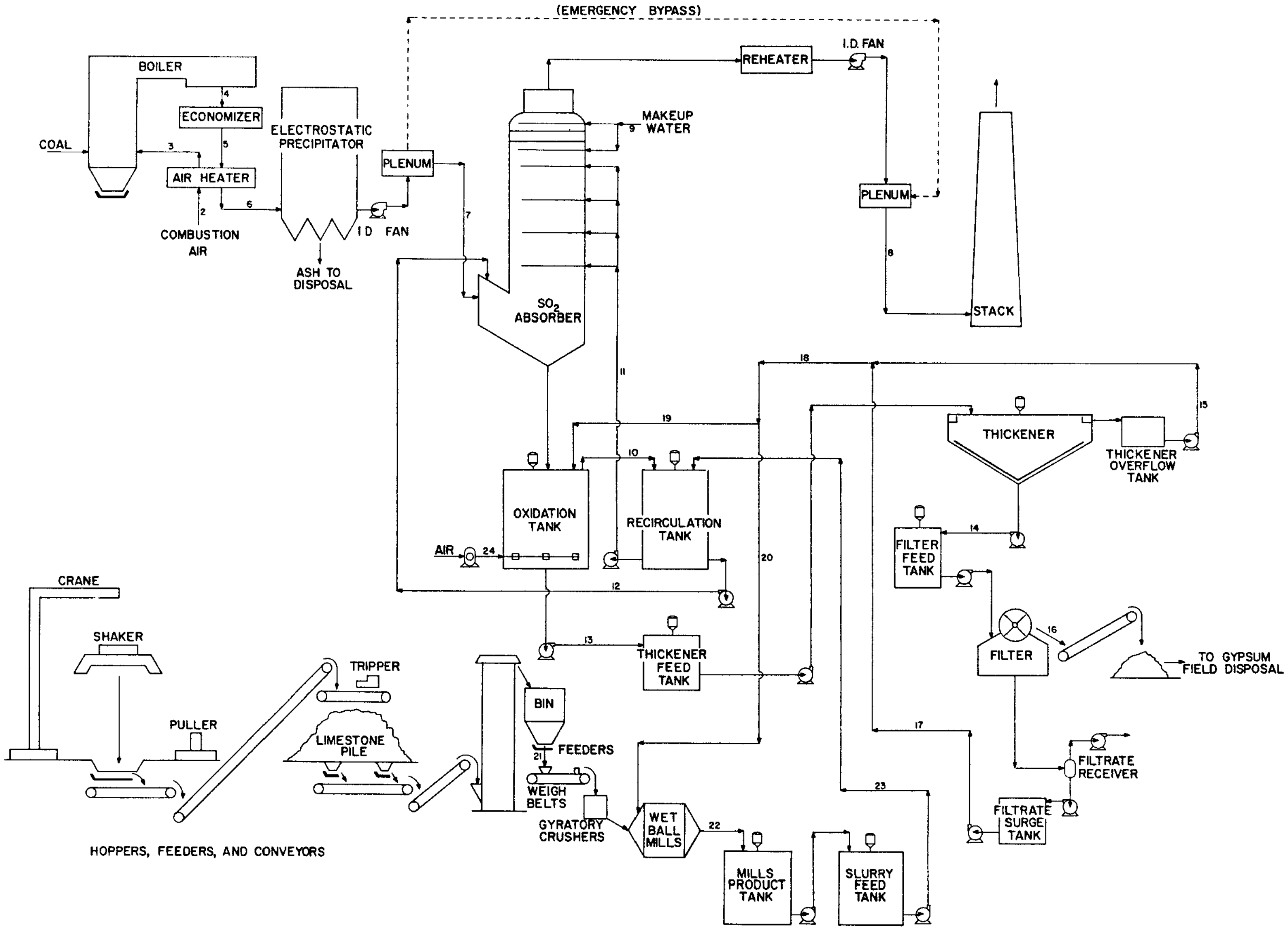

Figure 31. High-sulfur eastern coal case. Limestone scrubbing process. Flow diagram.

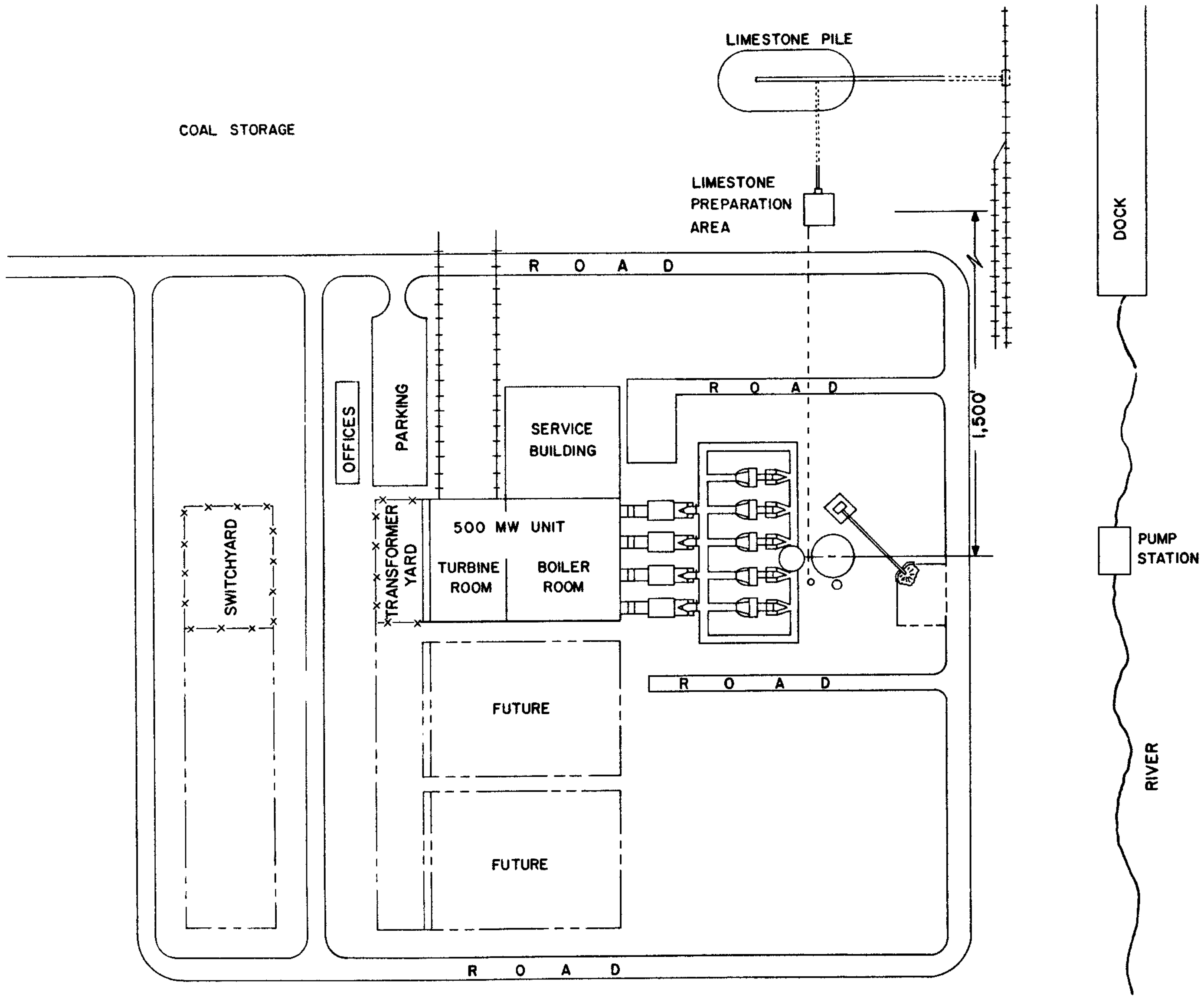

Figure 32. High-sulfur eastern coal case. Limestone scrubbing process. Plot plan.

TABLE 25. HIGH-SULFUR EASTERN COAL CASE

LIMESTONE SCRUBBING PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,021,200	4,334,800	4,776,600	4,776,600
2						
3	Flow rate, sft^3/min@60^{o}F		1,108,000	956,600	1,011,000	1,011,000
4	Temperature, ^{o}F		80	535	890	705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Gas to spray tower	Gas to stack	Makeup water to spray tower	Slurry to recirculation tank
1	Total stream, lb/hr	5,463,000	5,414,000	5,639,000	276,500	71,693,600
2						
3	Flow rate, sft^3/min@60^{o}F	1,156,000	1,156,000	1,236,000		
4	Temperature, ^{o}F	300	300	175		
5	Pressure, psig					
6	Flow rate, gpm				553	130,246
7	Specific gravity					
8	pH					
9	Undissolved solids, %					15
10						

	Stream No.	11	12	13	14	15
	Description	Recycle slurry to spray tower	Recycle slurry to presaturator	Slurry to thickener feed tank	Thickener underflow to filter feed tank	Thickener overflow to oxidation tank
1	Total stream, lb/hr	68,720,000	3,054,000	481,200	180,400	288,300
2						
3	Flow rate, sft^3/min@60^{o}F					
4	Temperature, ^{o}F					
5	Pressure, psig					
6	Flow rate, gpm	124,800	5,547	874	272	577
7	Specific gravity					
8	pH					
9	Undissolved solids, %	15	15	15		
10						

	Stream No.	16	17	18	19	20
	Description	Gypsum filter cake to disposal	Filtrate to oxidation tank	Total supernate return	Supernate to oxidation tank	Supernate to ball mills
1	Total stream, lb/hr	90,220	90,220	378,500	348,300	30,160
2						
3	Flow rate, sft^3/min@60^{o}F					
4	Temperature, ^{o}F					
5	Pressure, psig					
6	Flow rate, gpm	92	181	757	697	60
7	Specific gravity					
8	pH					
9	Undissolved solids, %	80				
10						

(continued)

TABLE 25 (continued)

	Stream No.	21	22	23	24	25
	Description	Limestone to weigh feeder	Limestone slurry to mills product tank	Limestone slurry to recirculation tank	Air to oxidation tank	Steam to reheater
1	Total stream, lb/hr	48,240	80,390	80,390	59,950	96,730
2						
3	Flow rate, sft^3/min@60^{o}F				13,060	
4	Temperature, ^{o}F				60	470
5	Pressure, psig					
6	Flow rate, gpm		101	101		
7	Specific gravity		1.6	1.6		
8	pH					
9	Undissolved solids, %		60	60		
10						

1						
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TABLE 26. HIGH-SULFUR EASTERN COAL CASE

LIMESTONE SCRUBBING PROCESS

EQUIPMENT LIST, DESCRIPTION, AND COST

Area 1--Materials Handling

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Car shaker	1	Top mounted with crane	71,900	13,000
2.	Car puller	1	25 hp puller, 5 hp return	63,100	19,600
3.	Hopper, limestone unloading	1	16 ft dia x 10 ft straight side height, includes 6 in. square grating	15,500	5,900
4.	Feeder, limestone unloading	1	Vibrating pan, 42 in. wide x 60 in. long, 3 hp, 250 tons/hr	5,500	500
5.	Conveyor, limestone unloading	1	Belt, 36 in. wide x 20 ft long, 5 hp, 250 tons/hr, 130 ft/min	11,400	1,400
6.	Conveyor, limestone stocking (incline)	1	Belt, 36 in. wide x 310 ft long, 50 hp, 250 tons/hr	85,300	4,800
7.	Dust collecting system	1	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system, w/dust hood	11,200	5,200
8.	Pump, unloading pit sump	1	Centrifugal, 60 gpm, 70 ft head, 5 hp, neoprene lined	2,400	800
9.	Conveyor, limestone stocking	1	Belt, 36 in. wide x 200 ft long, 5 hp, 250 tons/hr, 130 ft/min	73,100	3,900
10.	Tripper	1	1 hp, 30 ft/min	27,200	9,100
11.	Mobile equipment	1	Scraper tractor	141,900	-
12.	Hopper, reclaim	2	7 ft x 4-1/4 ft x 2 ft deep, carbon steel	2,400	1,600
13.	Feeder, live limestone storage	2	Vibrating pan, 3 hp	10,900	1,000

(continued)

TABLE 26 (continued)

Area 1 (continued)				
Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
14. Pump, tunnel sump	1	Vertical, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined (1 operating, 1 spare)	2,400	800
15. Conveyor, live limestone feed	1	Belt, 30 in. wide x 200 ft long, 5 hp, 100 tons/hr, 60 ft/min	40,900	2,900
16. Conveyor, live limestone feed (incline)	1	Belt, 30 in. wide x 190 ft long, 40 hp, 35 ft lift, 100 tons/hr, 60 ft/min	60,300	3,700
17. Elevator, live limestone feed	1	Continuous bucket, 12 in. x 8 in. x 11-3/4 in., 75 hp, 90 ft lift, 100 tons/hr, 160 ft/min	57,800	6,700
18. Bin, crusher feed	3	13 ft dia x 21 ft high, w/ cover, carbon steel	43,300	24,000
19. Conveyor, feed belt	1	Belt, 30 in. wide x 60 ft long, 7.5 hp, 100 tons/hr, 60 ft/min	20,500	1,400
20. Tripper	1	1 hp, 30 ft/min	27,200	9,100
21. Dust collecting system	1	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system	7,800	2,600
Subtotal			782,000	118,000
Area 2--Feed Preparation				
1. Feeder, crusher	3	Weigh belt, 18 in. wide x 14 ft long, 2 hp, 13.0 tons/hr	49,600	2,300
2. Crusher	3	Gyratory, 0 x 1-1/2 to 3/4 in., 75 hp, 13.0 tons/hr	297,100	6,500
3. Ball mill	3	Wet, open system, 700 hp, 13.0 tons/hr	1,678,000	119,300

(continued)

TABLE 26 (continued)

Item		No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
Area 2 (continued)					
4.	Tank, mills product	3	10 ft dia x 10 ft high, 5,500 gal, open top, four 10 in. baffles, agitator supports, carbon steel, flakeglass lined	13,700	11,000
5.	Agitator, mills product tank	3	36 in. dia, 10 hp, neoprene coated	22,900	5,500
6.	Pump, mills product tank	3	Centrifugal, 51 gpm, 60 ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,500	2,700
7.	Tank, slurry feed	1	20-3/4 ft dia x 20-3/4 ft high, 52,430 gal, open top, four 20-3/4 in. baffles, agitator supports, carbon steel, flakeglass lined	18,900	15,600
8.	Agitator, slurry feed tank	1	83 in. dia, 50 hp, neoprene coated	40,100	3,300
9.	Pump, slurry feed tank	8	Centrifugal, 25 gpm, 60 ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spare)	21,500	7,300
10.	Dust collecting system, ball mill	3	Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, automatic shaker system	23,300	7,800
	Subtotal			2,174,200	181,300
Area 3--Particulate Removal					
1.	ESP	4	99.8% removal efficiency SCA = 550	6,267,100	3,133,000
2.	Conveyor, fly ash, particulate bin	1	Pneumatic, pressure, vacuum, 125 hp	84,000	30,500
3.	Bin, particulate	2	26 ft dia x 25 ft high, w/cover, carbon steel	66,000	64,800

(continued)

TABLE 26 (continued)

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
Area 3 (continued)					
4.	Vibrator	2	Bin activator, 10 ft dia	28,900	4,800
	Subtotal			6,446,000	3,233,100
Area 4--Gas Handling					
1.	Fans	5	Induced draft, 368,700 aft^3/min, 9-1/2 in. static head, 890 rpm, 800 hp, fluid drive, double width, double inlet, Inconel (4 operating, 1 spare)	3,651,600	61,800
	Subtotal			3,651,600	61,800
Area 5--SO_2 Absorption					
1.	SO_2 absorber	5	Spray tower, 28 ft long x 28 ft wide x 40 ft high, 1/4 in. carbon steel, neoprene lining; FRP spray headers, 316 stainless steel chevron vane entrainment separator and nozzles (4 operating, 1 spare)	5,429,500	449,500
2.	Tank, oxidation-recirculation	5	27-1/2 ft dia x 39 ft high, 171,600 gal, open top, four 27-1/2 in. wide baffles, agitator supports, carbon steel, flakeglass lined (4 operating, 1 spare)	218,500	180,600
3.	Agitator, oxidation-recirculation tank	5	110 in. dia, 75 hp neoprene coated (4 operating, 1 spare)	309,600	127,000
4.	Tank, effluent hold	5	39 ft dia x 39 ft high, 343,200 gal, open top, four 39 in. wide baffles, agitator support, carbon steel	373,000	301,500

(continued)

TABLE 26 (continued)

Item		No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
Area 5 (continued)					
5.	Agitator, effluent hold tank	5	156 in. dia, 100 hp, neoprene coated	505,600	207,500
6.	Pump, slurry recirculation	15	Centrifugal, 15,600 gpm, 100 ft head, 800 hp, carbon steel, neoprene lined (8 operating, 7 spare)	1,581,400	139,400
7.	Pumps, presaturator, recycle	10	Centrifugal, 1,400 gpm, 100 ft head, 75 hp, carbon steel, neoprene lined (4 operating, 6 spare)	104,700	32,300
8.	Pump, oxidation bleed	8	Centrifugal, 221 gpm, 60 ft head, 7-1/2 hp, carbon steel, neoprene lined (4 operating, 4 spare)	34,300	11,500
9.	Blower, oxidation air	6	3,300 sft^3/min, 400 hp (4 operating, 2 spare)	208,400	4,700
10.	Sparger, oxidation air	5	19-1/2 ft dia ring (4 operating, 1 spare)	95,200	41,600
11.	Pump, makeup water	2	Centrifugal, 3,470 gpm, 200 ft head, 300 hp, carbon steel (1 operating, 1 spare)	33,200	3,700
12.	Soot blowers	60	Air, retractable	294,900	182,600
	Subtotal			9,188,300	1,681,900
Area 6--Stack Gas Reheat					
1.	Reheater	5	Inline steam type, 2,776 ft^2, 1/2 of tubes made of Inconel 625 and 1/2 made of Corten	2,614,700	165,900
	Subtotal			2,614,700	165,900

(continued)

TABLE 26 (continued)

Area 7--Solids Separation

	Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
1.	Tank, thickener feed	1	19-1/2 ft dia x 39 ft high, 85,700 gal, open top, agitator supports, four 19-1/2 in. baffles, carbon steel, flakeglass lined	29,700	24,600
2.	Agitator, thickener feed tank	1	2 turbines, 77 in. dia, 50 hp, neoprene coated	34,400	2,800
3.	Pump, thickener feed	2	Centrifugal, 868 gpm, 60 ft head, 25 hp, carbon steel, neoprene lined (1 operating, 1 spare)	15,900	5,300
4.	Thickener	1	Stainless steel tank, 71 ft dia x 6 ft high; concrete basin, 4 ft high	127,100	130,200
5.	Pump, thickener overflow	2	Centrifugal, 583 gpm, 75 ft head, 20 hp, carbon steel (1 operating, 1 spare)	10,400	1,200
6.	Tank, thickener overflow	1	16-1/4 ft dia x 6-1/4 ft high, 9,600 gal, open top, carbon steel	3,300	2,200
7.	Pump, thickener underflow	2	Centrifugal, 275 gpm, 10 ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,800	3,200
8.	Tank, filter feed	1	9-1/4 ft dia x 9-1/4 ft high, 4,490 gal, open top, carbon steel, flakeglass lined	3,600	3,000
9.	Agitator, filter feed tank	1	36 in. dia, 7-1/2 hp, neoprene coated	5,300	400
10.	Pump, filter feed slurry	3	Centrifugal, 135 gpm, 50 ft head, 5 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,900	3,500

(continued)

TABLE 26 (continued)

Area 7 (continued)

Item	No.	Description	Total material cost, 1982 $	Total labor cost, 1982 $
11. Filter	3	Rotary vacuum, 11 ft dia x 11 ft face, 15 total hp (2 operating, 1 spare)	372,300	68,000
12. Pump, filtrate	4	Centrifugal, 91 gpm, 20 ft head, 1 hp, carbon steel (2 operating, 2 spare)	17,200	1,900
13. Tank, filtrate surge	1	8 ft dia x 8 ft high, 3,000 gal, open top, carbon steel	1,600	1,100
14. Pump, filtrate surge tank	2	Centrifugal, 182 gpm, 85 ft head, 7-1/2 hp, carbon steel (1 operating, 1 spare)	9,200	1,000
15. Conveyor, gypsum disposal	1	Belt, 14 in. wide x 75 ft long, 100 ft inclined, 1.5 hp, 5.9 tons/hr, 40 ft/min	37,100	3,500
Subtotal			686,800	251,900

Basis: Most equipment cost estimates are based on informal vendor quotes and TVA information.

These costs represent equipment costs only. Costs for piping, electrical equipment, instruments, foundations, and other installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

ECONOMIC EVALUATION AND COMPARISON

Based on the power plant, design and economic premises, and the specific process equipment for each process described in the previous sections, study-grade capital investments, first-year revenue requirements, and levelized annual revenue requirements are prepared for the economic evaluation and comparison of the soda ash and lime spray dryer processes and a limestone scrubbing process (including an ESP). Both first-year and levelized annual revenue requirements are calculated. First-year annual revenue requirements are useful for comparing the relative cost differences between processes for their first year of operation, and they are an indicator of the magnitude of the annual revenue requirements. However, first-year annual revenue requirements are not representative of the actual cost of operating the plant since they do not consider either the time-value of money or inflation over the life of the plant. In order to reflect these costs, a levelizing factor is applied to the first-year annual revenue requirements to give a levelized annual revenue requirement. This levelizing factor is based on a 10% discount factor and a 6% inflation rate over the 30-year life of the power unit. Sensitivity analyses are also performed to evaluate the effects of varying the raw material price and stoichiometry for the lime spray dryer process.

Even though the spray dryer processes are described and costed as proven technology, the current status of development does not fully justify this assumption since none of the spray dryer processes have been operated on a commercial coal-fired boiler. However, for TVA cost estimation purposes each system is assumed to be proven technology.

ACCURACY OF ESTIMATES

The accuracy associated with these study-grade cost estimates, i.e., -20%, +40%, is defined as the relationship between the estimated costs and what the actual installed costs for the process might be. The accuracy assigned to a cost estimate is empirical and not related to variabilities in a statistical sense, but rather, it depends on both the amount and the quality of the technical data available. Accuracy ranges also reflect the numerous uncertainties surrounding estimates made using simplifying assumptions. For example, in a study-grade estimate in which only a flowsheet, material balance, and an equipment list are available--and all other indirect investments are factored--the uncertainty surrounding the investment is much greater than a preliminary-level estimate where quantities and costs for piping, electrical equipment, instruments, etc., are calculated rather than factored. However, when comparing the study-grade costs for two competing process technologies, many of the same simplifying assumptions are made for each of the processes, and therefore the comparability is greater than the accuracy of the estimates. When directly comparing two estimates of the same grade, the uncertainty ranges associated with the compared costs are estimated to be only ±10%.

LIGNITE CASE--CAPITAL INVESTMENT

Results

The capital investment for the lime spray dryer process for the lignite case is \$82.6M (\$165/kW) as shown in Table 27 and the capital investment for the limestone scrubbing process (a combined particulate-collection-limestone FGD system) is \$107.6M (\$215/kW) as shown in Table 28.

Comparison

The direct investment and the total capital investment for the two FGD systems are shown in Table 29. The lime spray dryer process is about 23% less capital intensive than the limestone scrubbing process. The major reasons for this substantial difference in capital investment between the lime spray dryer process and the limestone scrubbing process are shown in Table 30, in which the major capital investment areas for each process are compared. The major capital investment differences are in the SO_2 absorption, particulate removal, and gas handling areas. All are areas in which the costs for the limestone scrubbing process are substantially higher than the equivalent area in the lime spray dryer process. The differences in the SO_2 absorption area are primarily due to the use of a spray dryer as the absorber. In the spray dryer process presaturators, mist eliminators, forced-oxidation equipment, and large recirculating pumps and tanks are not required in the SO_2 absorption area. Eliminating this process equipment leads to a significantly lower investment for the SO_2 absorption area in the lime spray dryer process.

TABLE 29. LIGNITE CASE

DIRECT INVESTMENTS AND CAPITAL INVESTMENTS

Process	Direct investment M$	Direct investment $/kW	Total capital investment M$	Total capital investment $/kW
Lime spray dryer	41.5	83.0	82.6	165.3
Limestone scrubbing	59.2	118.4	107.4	214.7

TABLE 27. LIGNITE CASE

CAPITAL INVESTMENT

LIME SPRAY DRYER PROCESS

(500-MW new coal-fired unit, 0.9% S in coal; 70% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	1,778
Feed preparation	765
Gas handling	10,665
SO_2 absorption	7,336
Particulate removal	12,091
Particulate handling	2,163
Total process capital	34,798
Services, utilities, and miscellaneous	2,088
Total direct investment excluding landfill	36,886
Solids disposal	867
Landfill construction	3,756
Total direct investment	41,509
Indirect Investment	
Engineering design and supervision	2,657
Architect and engineering contractor	776
Construction expense	6,202
Contractor fees	2,032
Contingency	10,461
Total fixed investment	63,637
Other Capital Investment	
Allowance for startup and modifications	5,754
Interest during construction	9,793
Royalties	348
Land	960
Working capital	2,135
Total capital investment	82,627
Dollars of total capital per kW of generation capacity	165.25

Basis
- Upper Midwest plant location represents project beginning mid-1980, ending mid-1983. Average cost basis for scaling, mid-1982.
- Minimum in-process storage, redundant scrubber train, and pumps are spared.
- Landfill located one mile from power plant.
- FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.
- Only nominal construction overtime included.

TABLE 28. LIGNITE CASE

CAPITAL INVESTMENT

LIMESTONE SCRUBBING PROCESS

(500-MW new coal-fired power unit, 0.9% S in coal; 70% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	1,291
Feed preparation	2,406
Particulate removal	15,076
Gas handling	13,249
SO_2 absorption	17,357
Stack gas reheat	-
Solids separation	2,268
Total process capital	51,647
Services, utilities, and miscellaneous	3,099
Total direct investment excluding landfill	54,746
Solids disposal	790
Landfill construction	3,690
Total direct investment	59,226
Indirect Investment	
Engineering design and supervision	3,906
Architect and engineering contractor	1,132
Construction expense	9,054
Contractor fees	2,922
Contingency	7,973
Total fixed investment	84,213
Other Capital Investment	
Allowance for startup and modifications	6,263
Interest during construction	13,014
Royalties	-
Land	920
Working capital	2,950
Total capital investment	107,360
Dollars of total capital per kW of generation capacity	214.72

Basis
- Upper Midwest plant location represents project beginning in early 1981 and ending in late 1983. Average cost basis for scaling, mid-1982.
- Minimum in-process storage, redundant scrubber train and feed preparation area, pumps spared.
- Disposal area located one mile from power plant.
- FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.
- Only nominal construction overtime included.

TABLE 30. LIGNITE CASE

SUMMARY OF THE CAPITAL INVESTMENTS

Investment area	Total cost, k$ Lime spray dryer process	Limestone scrubbing process
Material handling	1,778	1,291
Feed preparation	765	2,406
Gas handling	10,665	13,249
SO_2 absorption	7,336	17,357
Particulate removal	12,091	15,076
Particulate handling	2,163	-
Solids separation	-	2,268
Solids disposal	867	790
Landfill construction	3,756	3,690
Land	960	920
All other capital costs	42,246	50,313
Total capital investment	82,627	107,360

Basis: TVA design and economic premises

The differences in investment between the lime spray dryer process and limestone scrubbing process for the gas handling area are primarily due to two factors: the larger ductwork requirements for the limestone scrubbing process and corrosion-resistant materials of construction used downstream of the SO_2 absorbers. Larger ductwork is required for the limestone scrubbing process because of the ductwork needed to neck up to the ESP and then neck down after the ESP, which is not required for the baghouse in the spray dryer process. Since the flue gas downstream of the spray dryer is not saturated, and thus corrosion is not expected to be a problem, the ductwork from the spray dryer to the baghouse and from the baghouse to the stack is made of Corten. In the limestone scrubbing process, the flue gas leaving the spray tower is saturated and contains entrained liquid (the bypassed flue gas is not mixed with the scrubbed gas until it enters the stack plenum) and corrosion could be a significant problem. Therefore, the flue gas ducts downstream of the spray towers are made of stainless steel. Stainless steel is about 3-1/2 times as expensive as Corten. About half of the increased cost for ductwork in the limestone scrubbing process is due to this higher material cost for ductwork downstream of the absorber. (For the same reasons, the ID fan in the limestone scrubbing process is made of Inconel and is thus much more expensive than the ID fan in the spray dryer process.)

The particulate collection area investment is lower for the spray dryer process even though the spray dryer process cost is based on a baghouse and the cost for the limestone scrubbing process is based on an ESP. There are several reasons for this lower cost. Since the baghouse is located downstream of the spray dryers, the flue gas has been cooled and thus the baghouse treats a much lower volume of flue gas. The other major reason for the more expensive ESP is that the fly ash is expected to have a high resistivity and require a high SCA.

The limestone scrubbing process also has the highest investment for the feed preparation and the particulate handling/solids separation areas. The feed preparation investment is higher for the limestone scrubbing process because crushers and ball mills are needed to grind the limestone. The feed preparation area in the lime spray dryer process involves only slaking to prepare the absorbent slurry. The slightly higher investment for the limestone scrubbing process in the particulate handling/solids separation area is due to the nature of the wastes. The FGD waste in the limestone scrubbing process is in the form of a dilute slurry which must be separated (by a thickener and a filter) into FGD waste and water for recycle to the process. The particulate handling area for the spray dryer process involves conveying the waste, which is collected as a dry free-flowing solid, to the waste storage bins and particulate recycle bins. This area also includes the equipment to reslurry and recycle the FGD waste through the spray dryer to increase absorbent utilization.

The limestone scrubbing process has the lowest investment in landfill construction, land, and the solids disposal area, although the differences do not completely counteract the higher investments required in other areas. These lower costs result because the gypsum from the limestone scrubbing process is much more dense than the dry FGD waste from the lime spray dryer process.

The investment for the limestone scrubbing process material handling section is less than the spray dryer process. The lime spray dryer process has higher material handling investment primarily because a large storage silo is required for a 30-day lime supply, but also because of the various large conveyors and elevators to move the lime. The limestone for the limestone scrubbing process can be stored on the ground and thus the material handling area includes only the large conveyors and elevators.

The category in Table 30, all other capital costs, includes those costs which are calculated entirely or in part as a percentage of the direct investment. Thus if the process has a higher direct investment (i.e., the limestone slurry process) it will have a proportionally higher charge for all other capital costs.

In order to facilitate comparisons of these economic results with those from other EPA-sponsored evaluations, the total capital investments are presented in a slightly different form in Table 31. In this table the capital investment is broken down into three areas: SO_2 absorption, particulate removal, and waste disposal. Each area includes not only direct investment for equipment, piping, electrical equipment, etc., but also its pro rata share of the indirect investments and all other capital charges. Previously identified direct investment areas grouped in the SO_2 absorption area are: material handling, feed preparation, gas handling, and SO_2 absorption. The particulate removal area consists only of the original particulate removal area. The waste disposal area combines the particulate handling/solids separation, solids disposal, and landfill construction.

TABLE 31. LIGNITE CASE

CAPITAL INVESTMENTS

Area	Investment, $/kW: Lime spray dryer process	Investment, $/kW: Limestone scrubbing process
SO_2 absorption	89.13	133.35
Particulate removal	51.25	57.47
Waste disposal	24.87	23.90
Capital investment	165.25	214.72

Basis: TVA design and economic premises

LIGNITE CASE--ANNUAL REVENUE REQUIREMENTS

Results

The first-year annual revenue requirements for the lime spray dryer process are $20.92 in 1984 dollars as shown in Table 32. The equivalent first-year unit revenue requirement is 7.61 mills/kWh. Levelized annual and unit revenue requirements are $28.69M and 10.43 mills/kWh respectively. The first-year annual revenue requirements for the limestone scrubbing process are $26.32M as shown in Table 33. The equivalent first-year unit revenue requirement is 9.57 mills/kWh. Levelized annual and unit revenue requirements are $35.65M and 12.96 mills/kWh respectively.

Comparison

The first-year and the levelized annual revenue requirements for each of the FGD processes are shown in Table 34. In terms of first-year

TABLE 32. LIGNITE CASE

ANNUAL REVENUE REQUIREMENTS

LIME SPRAY DRYER PROCESS

(500-MW new coal-fired power unit, 0.9% S in coal;
70% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Lime	16,300 tons	102.00/ton	1,663
Total raw material cost			1,663
Conversion costs			
Operating labor and supervision			
FGD	25,400 man-hr	15.00/man-hr	381
Solids disposal	30,675 man-hr	21.00/man-hr	644
Utilities			
Fuel	209,375 gal	1.60/gal	335
Process water	139,570 kgal	0.14/kgal	20
Electricity	43,703,473 kWh	0.037/kWh	1,617
Maintenance			
Labor and material			2,232
Analysis	4,200 man-hr	21.00/man-hr	88
Total conversion costs			5,317
Total direct costs			6,980
Indirect Costs - First-Year			
Overheads			
Plant and administrative			1,794
Total first-year operating and maintenance cost			8,774
Levelized capital charges (14.7% of total capital investment)			12,146
Total first-year annual revenue requirements			20,920
Levelized first-year operating and maintenance cost (1.886 first-year O and M)			16,548
Levelized capital charges (14.7% of total capital investment)			12,146
Levelized annual revenue requirements			28,694

	M$	Mills/kWh
First-year annual revenue requirements	20.92	7.61
Levelized annual revenue requirements	28.69	10.43

Basis
- Upper Midwest plant location, 1984 revenue requirements.
- Remaining life of power plant, 30 years.
- Power unit onstream time, 5,500 years
- Coal burned, 2,291,575 tons/yr, 9,500 Btu/kWh.
- Total direct investment, $41,509,000; total fixed investment, $63,637,000; and total capital investment $82,627,000.

TABLE 33. LIGNITE CASE

ANNUAL REVENUE REQUIREMENTS

LIMESTONE SCRUBBING PROCESS

(500-MW new coal-fired power unit, 0.9% S in coal; 70% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Limestone	26,700 tons	8.50/ton	227
Total raw material cost			227
Conversion costs			
Operating labor and supervision			
FGD	39,620 man-hr	15.00/man-hr	594
Solids disposal	29,447 man-hr	21.00/man-hr	618
Utilities			
Fuel	185,625 gal	1.60/gal	297
Process water	144,871 kgal	0.14/kgal	20
Electricity	53,683,254 kWh	0.037/kWh	1,986
Maintenance			
Labor and material			3,851
Analysis	3,335 man-hr	21.00/man-hr	70
Total conversion costs			7,436
Total direct costs			7,663
Indirect Costs - First-Year			
Overheads			
Plant and administrative			2,872
Total first-year operating and maintenance costs			10,535
Levelized capital charges (14.7% of total capital investment)			15,782
Total first-year annual revenue requirements			26,317
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			19,869
Levelized capital charges (14.7% of total capital investment)			15,782
Levelized annual revenue requirements			35,651

	M$	Mills/kWh
First-year annual revenue requirements	26.32	9.57
Levelized annual revenue requirements	35.65	12.96

Basis
- Upper Midwest plant location, 1984 revenue requirements.
- Remaining life of power plant, 30 years.
- Power unit onstream time, 5,500 hr/yr.
- Coal burned, 2,291,575 tons/yr, 9,500 Btu/kWh.
- Total direct investment, $59,226,000; total fixed investment, $84,213,000; and total capital investment, $107,360,000.

annual revenue requirements, the lime spray dryer process is substantially lower than the limestone scrubbing process (21% less). The relative ranking in terms of levelized annual revenue requirement does not change although the percentage difference does change slightly.

TABLE 34. LIGNITE CASE

FIRST-YEAR AND LEVELIZED ANNUAL REVENUE REQUIREMENTS

Process	First-year annual revenue requirements		Levelized annual revenue requirements	
	M$	Mills/kWh	M$	Mills/kWh
Lime spray dryer	20.92	7.61	28.69	10.43
Limestone scrubbing	26.32	9.57	35.65	12.96

Basis: TVA design and economic premises

Table 35 compares the various cost components of the first-year annual revenue requirements for each process. With the exception of the raw material and all other annual costs categories (which include minor costs such as process water, analysis, and waste disposal fuel), the limestone scrubbing process has the highest annual costs of the two processes. The higher annual costs for the limestone scrubbing process relative to the lime spray dryer process are primarily due to three areas: levelized capital charges, maintenance costs, and overhead costs. These areas alone are about $6.3M higher in the limestone scrubbing process than in the lime spray dryer process.

The higher levelized capital charge for the limestone scrubbing process is due to its higher capital investment, as previously discussed. Maintenance costs are much higher for the limestone scrubbing process because of the limestone handling and grinding equipment and the equipment needed to recirculate large quantities of the erosive slurry. Since the overhead charges are calculated as a percentage of the direct costs excluding utilities, the overheads for the limestone scrubbing process are substantially higher than those for the lime spray dryer process.

The raw material annual costs are the only major area in which the lime spray dryer process is higher than the limestone scrubbing process. This higher raw material cost in the lime spray dryer process is due to the use of a more reactive alkali reagent, which has a higher unit cost. The differences in annual quantities used (16,300 tons of lime, 26,700 tons of limestone) are not as significant as the differences in unit costs assumed ($102/ton for lime, $8.50/ton for limestone). The effects on the first-year annual revenue requirements of different unit costs of the raw materials and the raw material stoichiometry are discussed in the next section in the form of a sensitivity analysis.

TABLE 35. LIGNITE CASE

SUMMARY OF FIRST-YEAR ANNUAL REVENUE REQUIREMENTS

Annual cost	Total cost, k$ Lime spray dryer process	Limestone scrubbing process
Raw material	1,663	227
Operating labor and supervision	1,025	1,212
Electricity	1,617	1,986
Maintenance	2,232	3,851
Overheads	1,794	2,872
Levelized capital charges	12,146	15,782
All other	443	387
Total first-year annual revenue requirements	20,920	26,317

Basis: TVA design and economic premises

LIGNITE CASE--SENSITIVITY ANALYSIS

Sensitivity to Absorbent Prices

The sensitivities of the first-year annual revenue requirements to raw material costs for the lime spray dryer process and the limestone scrubbing process were calculated using the absorbent costs shown in Table 36. The results of this sensitivity analysis are shown in Figure 33.

TABLE 36. LIGNITE CASE

DELIVERED UNIT RAW MATERIAL COSTS ASSUMED FOR THE

SENSITIVITY ANALYSIS

Process	Raw material	Variation	$/ton[a]	% change
Lime spray dryer	Lime	Low	82.00	-20
		Base	102.00	-
		High	143.00	+40
Limestone scrubbing	Limestone	Low	7.00	-20
		Base	8.50	-
		High	12.00	+40

a. Delivered cost

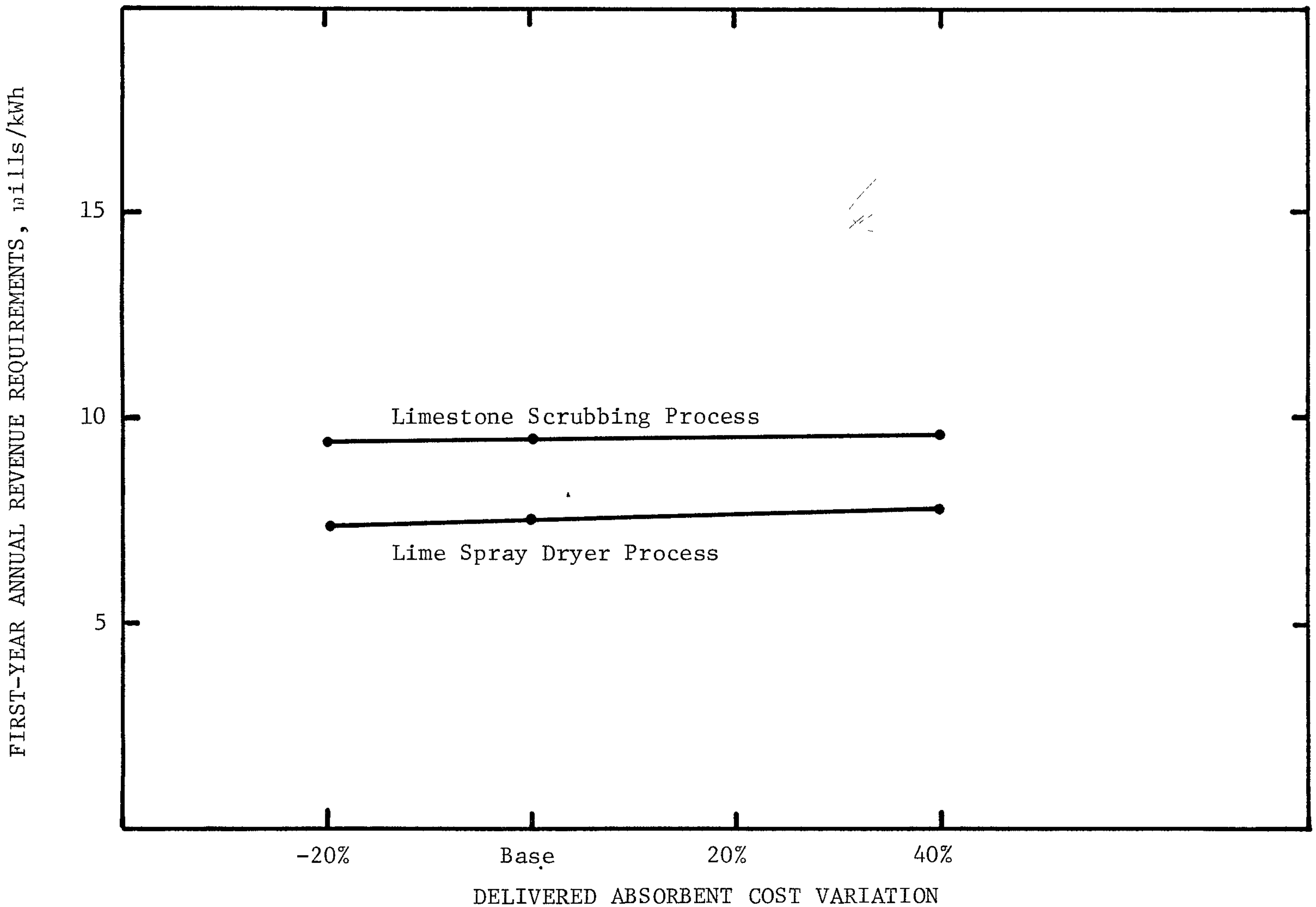

Figure 33. Lignite case--Sensitivity of the first-year annual revenue requirements to the delivered raw material cost.

Although the lime spray dryer process is more sensitive than the limestone scrubbing process to changes in the price of the absorbent, the low-sulfur nature of the lignite and the low SO_2 removal requirements preclude changes from the base-case costs from significantly changing the economic results. The lime spray dryer process has lower first-year annual revenue requirements regardless of the lime costs selected (within the -20% to +40% range). For example, a 40% increase in the delivered cost of lime results in only a 2.2% increase in the first-year annual revenue requirements for the lime spray dryer process. This is still 18% less than the first-year annual revenue requirements for the base-case limestone scrubbing process.

The limestone scrubbing process, because of the low unit cost of limestone as well as the lower sulfur level in the coal and the lower SO_2 removal requirements, is essentially insensitive to the delivered cost of limestone (a 40% increase in cost results in only a 0.3% increase in first-year annual revenue requirements).

Sensitivity to Raw Material Stoichiometry

Since the spray dryer process technology has only been demonstrated on a pilot-plant scale, the assumed stoichiometry in the spray dryer could change as the technology is developed. The required stoichiometry for coals with the same sulfur content could also change, depending on the fly ash alkalinity of the coal being burned. Therefore, a sensitivity analysis showing the changes in first-year revenue requirements as the stoichiometry in the spray dryer is changed has been included. Table 37 lists both the base-case and the alternative stoichiometries used in the sensitivity analysis. The stoichiometries given are in moles of alkali per mole of SO_2 absorbed. The range of stoichiometries shown for the lime spray dryer process is 1.1 (-10%) to 1.46 (19.7%).

The capital investments for each processing area are adjusted by using area scale factors and the ratio of flow rates through each area. Processing areas that are sized independently of the absorbent rates (gas handling and SO_2 absorption) are the same for each of the alternative stoichiometries. Many of the processing areas that are dependent on the absorbent flow rate contribute only minor amounts to the capital investment. For example, a 20% increase in absorbent flow rate increases the capital investment about 1%.

The annual revenue requirements for the lime spray dryer process are somewhat more sensitive to the absorbent stoichiometry than they are to the absorbent cost. For example, a 20% increase in the absorbent stoichiometry results in a 2.5% increase in first-year revenue requirements. However, from these results (as shown in Figure 34) it is apparent that stoichiometry changes over a wide range will have little effect on the capital investment and annual revenue requirement relationships of the two processes for this lignite case.

TABLE 37. LIGNITE CASE

COMPARISON OF CAPITAL INVESTMENT AND FIRST-YEAR UNIT

REVENUE REQUIREMENTS FOR THE LIME SPRAY DRYER PROCESS

AT VARIOUS RAW MATERIAL STOICHIOMETRIES

	Raw material stoichiometry			Total capital investment		First-year unit revenue requirements	
Process	Variation	Value[a]	% change[b]	\$/kW	% change[b]	mills/kWh	% change[b]
Lime spray	Low	1.10	-9.9	164.0	-0.79	7.50	-1.45
	Base	1.22	-	165.3	-	7.61	-
	High	1.46	19.7	167.3	1.21	7.80	2.50
Limestone scrubbing	Base	1.12	-	214.7	-	9.57	-

a. Raw material stoichiometry is defined as mols of alkali per mol of SO_2 absorbed.
b. Change is calculated relative to the base-case value.

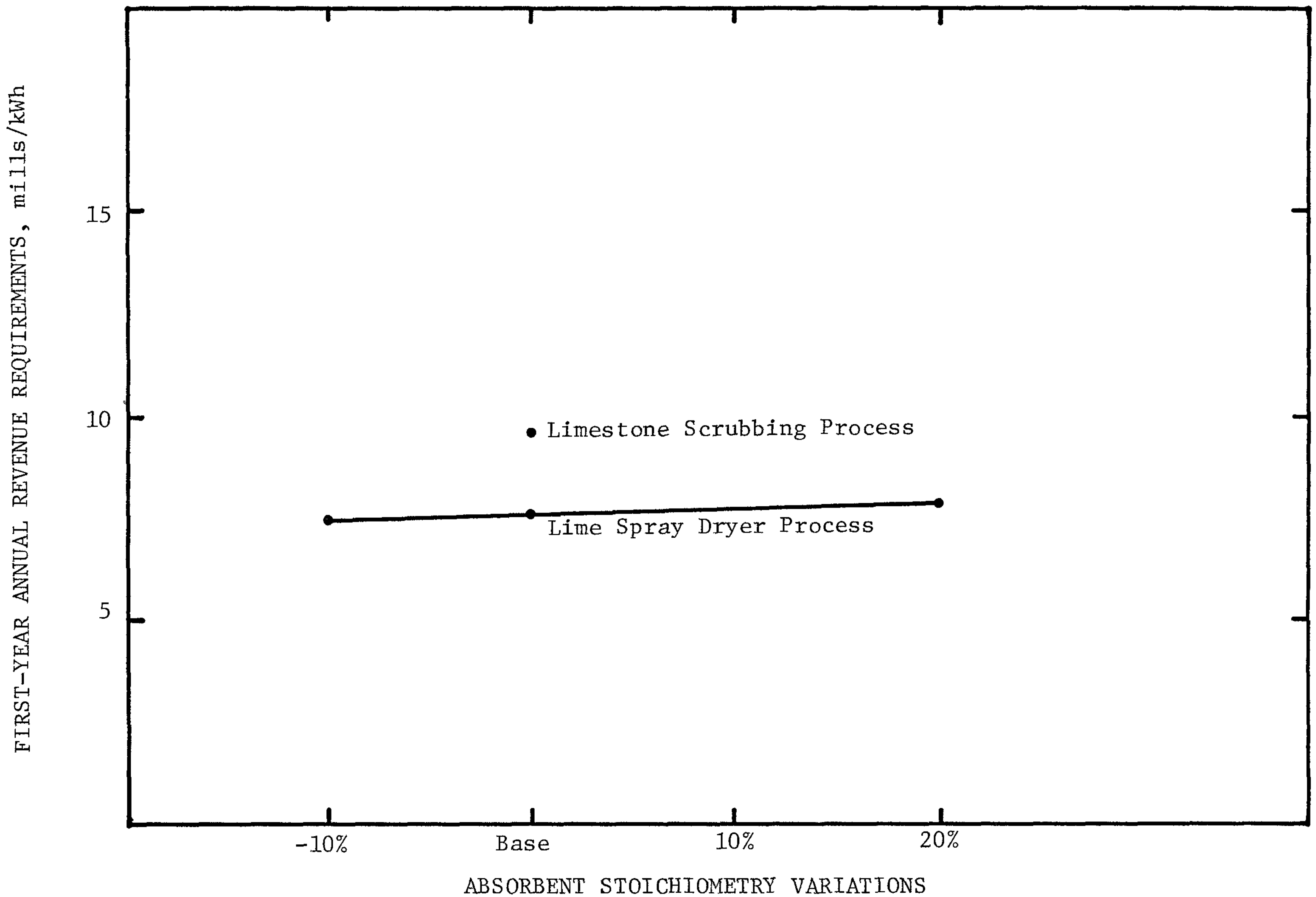

Figure 34. Lignite case--Sensitivity of the first-year annual revenue requirements to the raw material stoichiometry in the absorber.

LOW-SULFUR WESTERN COAL CASE--CAPITAL INVESTMENT

Results

The capital investment for the soda ash spray dryer process for the low-sulfur western coal case is $79.4M ($159/kW) in mid-1982 dollars as shown in Table 38. The capital investment for the lime spray dryer process is $77.1M ($154/kW) as shown in Table 39. The capital investment for the limestone scrubbing process (a combined particulate-collection-limestone FGD system) is $88.1M ($176/kW) as shown in Table 40.

Comparison

The direct investment and the total capital investment for the three FGD systems are shown in Table 41. Both the lime and the soda ash spray dryer processes are about 10% to 12% less capital intensive than the limestone scrubbing process. Although the lime spray dryer process is the least capital intensive of the three FGD systems for the low-sulfur western coal case, the difference (about 3% in capital investment) between the two spray dryer processes is not highly significant when compared with the accuracy limits associated with studies of this type.

TABLE 41. LOW-SULFUR WESTERN COAL CASE

DIRECT INVESTMENTS AND CAPITAL INVESTMENTS

Process	Direct investment M$	Direct investment $/kW	Total capital investment M$	Total capital investment $/kW
Lime spray dryer	38.6	77.2	77.1	154.2
Soda ash spray dryer	40.9	81.9	79.4	158.9
Limestone scrubbing	48.5	97.0	88.1	176.1

The major reasons for this substantial difference in capital investment between the spray dryer processes and the limestone scrubbing process are shown in Table 42, in which the major investment areas for each process are compared. The major investment differences are in the SO_2 absorption and gas handling areas. Both are areas in which the costs for the limestone scrubbing process are substantially higher than the equivalent area in the spray dryer processes. The differences in the SO_2 absorption area are primarily due to the use of a spray dryer as the absorber which eliminates the need for presaturators, mist eliminators, forced-oxidation equipment, and large recirculating pumps and tanks. Eliminating this process equipment leads to a significantly lower direct investment for the SO_2 absorption area in the spray dryer processes.

TABLE 38. LOW-SULFUR WESTERN COAL CASE

CAPITAL INVESTMENT

SODA ASH SPRAY DRYER PROCESS

(500-MW new coal-fired unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	461
Feed preparation	91
Gas handling	9,088
SO_2 absorption	9,208
Particulate removal	11,523
Particulate handling	750
Total process capital	31,121
Services, utilities, and miscellaneous	1,867
Total direct investment excluding pond	32,988
Solids disposal	725
Pond construction	7,228
Total direct investment	40,941
Indirect Investment	
Engineering design and supervision	2,454
Architect and engineering contractor	732
Construction expense	5,856
Contractor fees	2,010
Contingency	9,415
Total fixed investment	61,408
Other Capital Investment	
Allowance for startup and modifications	5,146
Interest during construction	9,467
Royalties	311
Land	1,146
Working capital	1,970
Total capital investment	79,448
Dollars of total capital per kW of generation capacity	158.90

Basis
- Upper Midwest plant location represents project beginning mid-1980, ending mid-1983. Average cost basis for scaling, mid-1982.
- Minimum in-process storage, redundant scrubber train, and pumps are spared.
- Pond located one mile from power plant.
- FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.
- Only nominal construction overtime included.

TABLE 39. LOW-SULFUR WESTERN COAL CASE

CAPITAL INVESTMENT

LIME SPRAY DRYER PROCESS

(500-MW new coal-fired unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	1,691
Feed preparation	680
Gas handling	10,030
SO_2 absorption	7,366
Particulate removal	11,523
Particulate handling and recycle	2,057
Total process capital	33,347
Services, utilities, and miscellaneous	2,001
Total direct investment excluding landfill	35,348
Solids disposal	719
Landfill construction	2,520
Total direct investment	38,587
Indirect Investment	
Engineering design and supervision	2,524
Architect and engineering contractor	732
Construction expense	5,858
Contractor fees	1,893
Contingency	9,775
Total fixed investment	59,369
Other Capital Investment	
Allowance for startup and modifications	5,514
Interest during construction	9,149
Royalties	333
Land	770
Working capital	1,978
Total capital investment	77,113
Dollars of total capital per kW of generation capacity	154.23

Basis
- Upper Midwest plant location represents project beginning mid-1980, ending mid-1983. Average cost basis for scaling, mid-1982.
- Minimum in-process storage, redundant scrubber train, and pumps are spared.
- Landfill located one mile from power plant.
- FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.
- Only nominal construction overtime included.

TABLE 40. LOW-SULFUR WESTERN COAL CASE

CAPITAL INVESTMENT

LIMESTONE SCRUBBING PROCESS

(500-MW new coal-fired power unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	1,009
Feed preparation	1,923
Particulate removal	11,688
Gas handling	11,646
SO_2 absorption	15,054
Stack gas reheat	-
Solids separation	1,828
Total process capital	43,148
Services, utilities, and miscellaneous	2,589
Total direct investment excluding landfill	45,737
Solids disposal	616
Landfill construction	2,158
Total direct investment	48,511
Indirect Investment	
Engineering design and supervision	3,245
Architect and engineering contractor	937
Construction expense	7,491
Contractor fees	2,395
Contingency	6,447
Total fixed investment	69,026
Other Capital Investment	
Allowance for startup and modifications	5,232
Interest during construction	10,672
Royalties	-
Land	670
Working capital	2,464
Total capital investment	88,064
Dollars of total capital per kW of generation capacity	176.13

Basis

Upper Midwest plant location represents project beginning in early 1981 and ending in late 1983. Average cost basis for scaling, mid-1982.

Minimum in-process storage, redundant scrubber train and feed preparation area, pumps spared.

Disposal area located one mile from power plant.

FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.

Only nominal construction overtime included.

TABLE 42. LOW-SULFUR WESTERN COAL CASE

SUMMARY OF THE CAPITAL INVESTMENTS

	Total cost, k$		
Investment area	Lime spray dryer process	Soda ash spray dryer process	Limestone scrubbing process
Material handling	1,691	461	1,009
Feed preparation	680	91	1,923
Gas handling	10,030	9,088	11,646
SO_2 absorption	7,366	9,208	15,054
Particulate removal	11,523	11,523	11,688
Particulate handling	2,057	750	-
Solids separation	-	-	1,828
Solids disposal	719	725	616
Landfill/pond construction	2,520	7,228	2,158
Land	770	1,146	670
All other capital costs	39,757	39,228	41,472
Total capital investment	77,113	79,448	88,064

Basis: TVA design and economic premises

The differences in investment between the spray dryer processes and limestone scrubbing process for the gas handling area are primarily due to two factors: the larger ductwork requirements for the limestone scrubbing process and corrosion-resistant materials of construction (stainless steel) used downstream of the SO_2 absorbers. Since the flue gas downstream of the spray dryer is not saturated, and thus corrosion is not expected to be a problem, the ductwork from the spray dryer to the baghouse and from the baghouse to the stack is made of Corten steel. In the limestone scrubbing process, the flue gas leaving the spray tower is saturated and contains entrained liquid (the bypassed flue gas is not mixed with the scrubbed gas until it enters the stack plenum) and corrosion could be a significant problem. Therefore, the flue gas ducts downstream of the spray tower are made of stainless steel. (For the same reasons, the ID fan in the limestone scrubbing process is made of Inconel and is thus much more expensive than the ID fan in the spray dryer processes.)

Due to the different raw material characteristics of the spray dryers processes, the only other area similar in both of the spray dryer processes is the particulate collection area. The particulate collection area investment is the same for both of the spray dryer processes and is essentially the same for the limestone scrubbing process even though the spray dryer process costs are based on a baghouse and the cost for the limestone scrubbing process is based on an ESP.

The limestone scrubbing process also has the highest investment for the feed preparation and the particulate handling/solids separation areas. The feed preparation investment is somewhat more expensive for the limestone scrubbing process because it is necessary to grind the limestone before preparing the absorbent slurry. The feed preparation area in the lime spray dryer process involves only slaking to prepare the absorbent slurry and the same area in the soda ash spray dryer process involves simply diluting the saturated soda ash solution to prepare the absorbent solution. The lime spray dryer and the limestone scrubbing processes have much higher particulate handling/solids separation area investments than the soda ash spray dryer process. The lime spray dryer process has the highest investment because it not only includes the equipment to convey the FGD waste to waste storage bins and recycle bins but also the equipment to reslurry and recycle the FGD waste through the spray dryer to increase absorbent utilization. For the limestone slurry process this is due to the FGD waste being in the form of a dilute slurry which must be separated (by a thickener and a filter) into FGD waste and water for recycle to the process. The particulate handling area for the soda ash spray dryer process involves only conveying the waste, which is collected as a dry free-flowing solid, to the waste storage bins.

The limestone scrubbing process has the lowest investment in landfill/pond construction, land, and the disposal area, although the differences do not balance the higher investments required in other areas. The landfill/pond construction, land, and solids disposal area costs for the soda ash spray dryer process are much higher than the equivalent areas in the other processes because the soluble nature of the waste requires a pond rather than a landfill for disposal.

The investment for the limestone scrubbing process material handling section falls between those of the spray dryer processes. The lime spray dryer process has the highest material handling investment primarily because a large storage silo is required for a 30-day lime supply, but also because of the various large conveyors and elevators to move the lime. The limestone for the limestone scrubbing process can be stored on the ground and thus the material handling area includes only the large conveyors and elevators. The soda ash spray dryer process has the lowest area investment since the soda ash is stored as a slurry in a large tank, and other than several small pumps requires no additional equipment.

In order to facilitate comparisons of these economic results with those from other EPA-sponsored evaluations, the total capital investments are presented in a slightly different form in Table 43. In this table the capital investment is broken down into three areas: SO_2 absorption, particulate removal, and waste disposal. Each area includes not only the direct investment for equipment, piping, electrical equipment, etc., but also its pro rata share of the indirect investments and all other capital charges. Previously identified direct investment areas grouped in the SO_2 absorption area are: material handling, feed preparation, gas handling, and SO_2 absorption. The particulate removal area consists only of the original particulate removal area. The waste disposal area combines the particulate handling solids separation, solids disposal, and landfill/pond construction.

TABLE 43. LOW-SULFUR WESTERN COAL CASE

CAPITAL INVESTMENTS

Area	Investment, $/kW		
	Lime spray dryer process	Soda ash spray dryer process	Limestone scrubbing process
SO_2 absorption	85.62	81.85	114.96
Particulate removal	48.84	48.84	44.56
Waste disposal	19.76	28.21	16.61
Capital investment	154.22	158.90	176.13

Basis: TVA design and economic premises

LOW-SULFUR WESTERN COAL CASE--ANNUAL REVENUE REQUIREMENTS

Results

The first-year annual revenue requirements for the soda ash spray dryer process are $20.41M in 1984 dollars as shown in Table 44. This corresponds to a first-year unit revenue requirement of 7.42 mills/kWh. Equivalent levelized annual revenue requirements for the soda ash spray dryer process are $28.15M, or 10.23 mills/kWh. The first-year annual revenue requirements for the lime spray dryer process are $19.02M as shown in Table 45. The equivalent first-year unit revenue requirement is 6.92 mills/kWh. Levelized annual and unit revenue requirements are $25.82M and 9.39 mills/kWh respectively. The first-year annual revenue requirements for the limestone scrubbing process are $21.73M as shown in Table 46. The equivalent first-year unit revenue requirement is 7.90 mills/kWh. Levelized annual and unit revenue requirements are $29.52M and 10.73 mills/kWh respectively.

Comparison

The first-year and the levelized annual revenue requirements for each of the FGD processes are shown in Table 47. In terms of first-year annual revenue requirements, the relative ranking of the three processes from lowest cost to the highest is lime spray dryer, soda ash spray

TABLE 44. LOW-SULFUR WESTERN COAL CASE

ANNUAL REVENUE REQUIREMENTS

SODA ASH SPRAY DRYER PROCESS

(500-MW new coal-fired power unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Soda ash	18,350 tons	145.00/ton	2,661
Total raw material cost			2,661
Conversion costs			
Operating labor and supervision			
FGD	16,640 man-hr	15.00/man-hr	250
Solids disposal	28,362 man-hr	21.00/man-hr	596
Utilities			
Fuel	165,725 gal	1.60/gal	265
Process water	70,236 kgal	0.14/kgal	10
Electricity	41,152,243 kWh	0.037/kWh	1,523
Maintenance			
Labor and material			1,863
Analysis	4,191 man-hr	21.00/man-hr	88
Total conversion costs			4,595
Total direct costs			7,256
Indirect Costs - First-Year			
Overheads			
Plant and administrative			1,475
Total first-year operating and maintenance costs			8,731
Levelized capital charges (14.7% of total capital investment)			11,679
Total first-year annual revenue requirements			20,410
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			16,467
Levelized capital charges (14.7% of total capital investment)			11,679
Levelized annual revenue requirements			28,146

	M$	Mills/kWh
First-year annual revenue requirements	20.41	7.42
Levelized annual revenue requirements	28.15	10.23

Basis
- Upper Midwest plant location, 1984 revenue requirements.
- Remaining life of power plant, 30 years.
- Power unit onstream time, 5,500 hr/yr.
- Coal burned, 1,116,500 tons/yr, 9,500 Btu/kWh.
- Total direct investment, $40,941,000; total fixed investment, $61,408,000; and total capital investment, $79,448,000.

TABLE 45. LOW-SULFUR WESTERN COAL CASE

ANNUAL REVENUE REQUIREMENTS

LIME SPRAY DRYER PROCESS

(500-MW new coal-fired power unit, 0.7% S in coal;
70% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Lime	10,100 tons	102.00/ton	1,030
Total raw material cost			1,030
Conversion costs			
Operating labor and supervision			
FGD	25,400 man-hr	15.00/man-hr	381
Solids disposal	28,152 man-hr	21.00/man-hr	591
Utilities			
Fuel	163,750 gal	1.60/gal	262
Process water	82,193 kgal	0.14/kgal	12
Electricity	39,571,324 kWh	0.037/kWh	1,464
Maintenance			
Labor and material			2,136
Analysis	4,191 man-hr	21.00/man-hr	88
Total conversion costs			4,934
Total direct costs			5,964
Indirect Costs - First-Year			
Overheads			
Plant and administrative			1,717
Total first-year operating and maintenance costs			7,681
Levelized capital charges (14.7% of total capital investment)			11,336
Total first-year annual revenue requirements			19,017
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			14,486
Levelized capital charges (14.7% of total capital investment)			11,336
Levelized annual revenue requirements			25,822

	M$	Mills/kWh
First-year annual revenue requirements	19.02	6.92
Levelized annual revenue requirements	25.82	9.39

Basis
- Upper Midwest plant location, 1984 revenue requirements.
- Remaining life of power plant, 30 years.
- Power unit onstream time, 5,500 hr/yr.
- Coal burned, 1,346,700 tons/yr, 9,500 Btu/kWh.
- Total direct investment, $38,587,000; total fixed investment, $59,369,000; and total capital investment, $77,113,000.

TABLE 46. LOW-SULFUR WESTERN COAL CASE

ANNUAL REVENUE REQUIREMENTS

LIMESTONE SCRUBBING PROCESS

(500-MW new coal-fired power unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Limestone	17,600 tons	8.50/ton	150
Total raw material cost			150
Conversion costs			
Operating labor and supervision			
FGD	39,620 man-hr	15.00/man-hr	594
Solids disposal	25,981 man-hr	21.00/man-hr	546
Utilities			
Fuel	134,540 gal	1.60/gal	215
Process water	120,000 kgal	0.14/kgal	17
Electricity	40,769,000 kWh	0.037/kWh	1,508
Maintenance			
Labor and material			3,219
Analysis	3,330 man-hr	21.00/man-hr	70
Total conversion costs			6,169
Total direct costs			6,319
Indirect Costs - First-Year			
Overheads			
Plant and administrative			2,467
Total first-year operating and maintenance costs			8,786
Levelized capital charges (14.7% of total capital investment)			12,945
Total first-year annual revenue requirements			21,731
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			16,570
Levelized capital charges (14.7% of total capital investment)			12,945
Levelized annual revenue requirements			29,515

	M$	Mills/kWh
First-year annual revenue requirements	21.73	7.90
Levelized annual revenue requirements	29.52	10.73

Basis
- Upper Midwest plant location, 1984 revenue requirements.
- Remaining life of power plant, 30 years.
- Power unit onstream time, 5,500 hr/yr.
- Coal burned, 1,346,700 tons/yr, 9,500 Btu/kWh.
- Total direct investment, $48,511,000; total fixed investment, $69,026,000; and total capital investment, $88,064,000.

dryer, and limestone scrubbing. The lime spray dryer process is slightly lower in cost than the soda ash spray dryer process (7% less) and substantially lower than the limestone slurry process (12% less). The relative ranking in terms of levelized annual revenue requirement does not change although the percentage difference does change slightly.

TABLE 47. LOW-SULFUR WESTERN COAL CASE

FIRST-YEAR AND LEVELIZED ANNUAL REVENUE REQUIREMENTS

Process	First-year annual revenue requirements		Levelized annual revenue requirements	
	M$	Mills/kWh	M$	Mills/kWh
Lime spray dryer	19.02	6.92	25.82	9.39
Soda ash spray dryer	20.41	7.42	28.15	10.23
Limestone scrubbing	21.73	7.90	29.52	10.73

Basis: TVA design and economic premises

Table 48 compares the various cost components of the first-year annual revenue requirements for each process. With the exception of the costs for raw material, electricity, and all other categories, the limestone scrubbing process has the highest annual costs of all three processes. The higher annual costs for the limestone scrubbing process relative to the spray dryer processes are primarily due to three areas: levelized capital charges, maintenance costs, and overhead costs. These areas alone are about $3.5M higher in the limestone scrubbing process than in the spray dryer processes.

The higher levelized capital charge for the limestone scrubbing process is due to its higher capital investment. Maintenance costs are much higher for the limestone scrubbing process because of the limestone handling and grinding equipment and the equipment needed to recirculate large quantities of the erosive slurry. Since the overhead charges are calculated as a percentage of the direct costs excluding utilities, (i.e., maintenance and operating labor), the overheads for the limestone scrubbing process are substantially higher than those for the spray dryer processes.

TABLE 48. LOW-SULFUR WESTERN COAL CASE

SUMMARY OF FIRST-YEAR ANNUAL REVENUE REQUIREMENTS

	Total cost, k$		
Annual cost	Lime spray dryer process	Soda ash spray dryer process	Limestone scrubbing process
Raw material	1,030	2,661	150
Operating labor and supervision	972	846	1,140
Electricity	1,464	1,523	1,508
Maintenance	2,136	1,863	3,219
Overheads	1,717	1,475	2,467
Levelized capital charges	11,336	11,679	12,945
All other	362	363	302
Total first-year annual revenue requirements	19,017	20,410	21,731

Basis: TVA design and economic premises

The raw material annual costs are the only major area in which the spray dryer processes are significantly higher than the limestone scrubbing process. This higher raw material cost in the spray dryer processes is due to the use of a more reactive alkali reagent, which has a higher unit cost. The differences in annual quantities used (10,100 tons of lime, 18,350 tons of soda ash, 17,600 tons of limestone) are not as significant as the differences in unit costs assumed ($102/ton for lime, $145/ton for soda ash, $8.50/ton for limestone). The effects on the first-year annual revenue requirements of different unit costs of the raw materials and the raw material stoichiometry are discussed in the next section in the form of a sensitivity analysis.

A comparison of the various cost components for the two spray dryer processes shows that the largest difference is in the raw materials cost ($1.6M). The soda ash spray dryer process has a higher raw material cost because of both the higher annual consumption (no waste recycle) and the higher unit cost of the soda ash, compared with lime. Other less significant differences are in three areas: maintenance, overheads, and levelized capital charges. The soda ash process has a higher levelized capital charge because it has a somewhat higher total capital investment. The maintenance cost (and overhead cost because it is calculated in part as a percentage of the maintenance cost) is somewhat lower for the soda ash process because the material handling equipment required to prepare the soda ash for use in the process involves only solutions rather than solids and aqueous slurries.

LOW-SULFUR WESTERN COAL CASE--SENSITIVITY ANALYSIS

Sensitivity to Absorbent Prices

The sensitivities of the first-year annual revenue requirements to raw material costs for the lime spray dryer process, the soda ash spray dryer process, and the limestone scrubbing process were calculated using the absorbent costs shown in Table 49. The results of this sensitivity analysis are shown in Figure 35.

TABLE 49. LOW-SULFUR WESTERN COAL CASE

DELIVERED UNIT RAW MATERIAL COSTS ASSUMED FOR THE

SENSITIVITY ANALYSIS

Process	Raw material	Variation	\$/ton[a]	% change
Lime spray dryer	Lime	Low	82.00	-20
		Base	102.00	-
		High	143.00	+40
Soda ash spray dryer	Soda ash	Low	116.00	-20
		Base	145.00	-
		High	203.00	+40
Limestone scrubbing	Limestone	Low	7.00	-20
		Base	8.50	-
		High	12.00	+40

a. Delivered cost

Although the lime spray dryer process and the soda ash process are more sensitive than the limestone scrubbing process to changes in the price of the absorbent, the low-sulfur nature of the coal and the low SO_2 removal requirements preclude changes from the base-case costs from significantly changing the economic results. The lime spray dryer process has lower first-year annual revenue requirements regardless of the lime costs selected (within the -20% to +40% range). For example, a 40% increase in the delivered cost of lime results in only a 2.2% increase in the first-year annual revenue requirements for the lime spray dryer process. This is still 5% less than the first-year annual revenue requirements for the base-case soda ash spray dryer process and 11% less than the first-year annual revenue requirements for the base-case limestone scrubbing process.

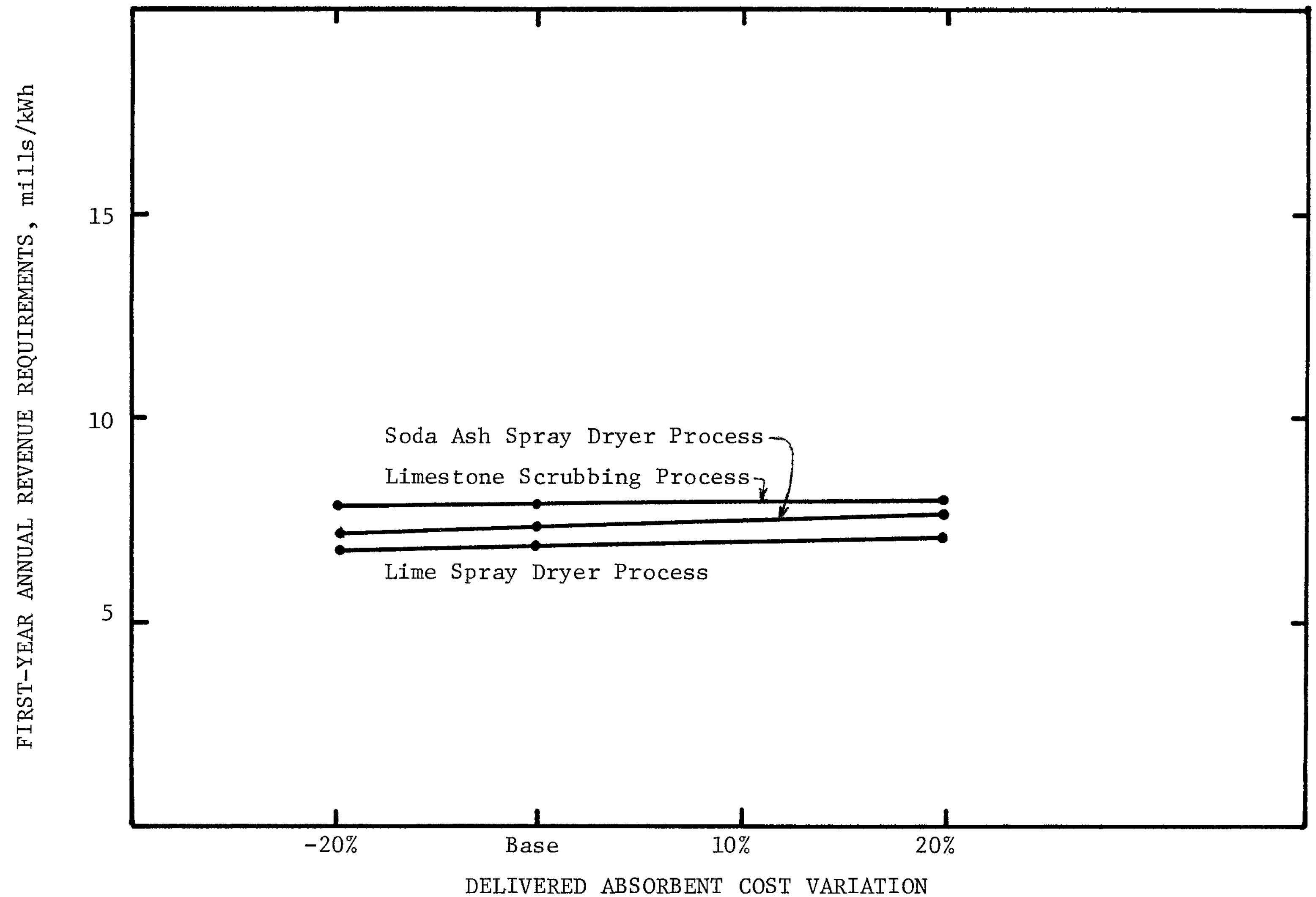

Figure 35. Low-sulfur western coal case--Sensitivity of the first-year annual revenue requirements to the delivered raw material cost.

The soda ash spray dryer process is slightly more sensitive to the delivered price of the raw material because of the higher unit cost of soda ash. A 40% increase in the delivered cost of soda ash increases the first-year annual revenue requirements for the soda ash spray dryer process about 3.2%. This is still about 1% less than the limestone scrubbing process.

The limestone scrubbing process, because of the low unit cost of limestone as well as the lower sulfur level in the coal and the lower SO_2 removal requirements, is essentially insensitive to the delivered cost of limestone (a 40% increase in cost results in only a 0.3% increase in first-year annual revenue requirements).

Sensitivity to Raw Material Stoichiometry

The sensitivity analysis for the low-sulfur western coal compares only the lime spray dryer process and the limestone scrubbing process since the soda ash spray dryer process usually approaches 100% absorbent utilization. Table 50 lists both the base-case and the alternative stoichiometries used in the sensitivity analysis. The stoichiometries given are in moles of alkali per mole of SO_2 absorbed. The range of stoichiometries shown for the lime spray dryer process is 1.1 (-10%) to 1.46 (19.7%).

Many of the processing areas that are dependent on the absorbent flow rate contribute only minor amounts to the capital investment. For example, a 20% increase in absorbent flow rate increases the capital investment about 1%. The annual revenue requirements for the lime spray dryer process are somewhat more sensitive to the absorbent stoichiometry than to the absorbent cost. For example, a 20% increase in the absorbent stoichiometry results in a 2.0% increase in first-year revenue requirements. However, from these results (as shown in Figure 36) it is apparent that stoichiometry changes over a wide range will have little effect on the capital investment and annual revenue requirement relationships of the two processes for this low-sulfur western coal case.

LOW-SULFUR EASTERN COAL CASE--CAPITAL INVESTMENT

Results

The capital investment for the lime spray dryer process for the low-sulfur eastern coal application is $75.3M ($151/kW) in mid-1982 dollars as shown in Table 51. The capital investment for the limestone scrubbing process is $92.6M ($185/kW) as shown in Table 52.

Comparison

The direct investment and the total capital investment for the two FGD systems are shown in Table 53. The lime spray dryer process is substantially less (19%) capital intensive than the limestone scrubbing process for this low-sulfur eastern coal case.

TABLE 50. LOW-SULFUR WESTERN COAL CASE

COMPARISON OF CAPITAL INVESTMENT AND FIRST-YEAR UNIT

REVENUE REQUIREMENTS FOR THE LIME SPRAY DRYER PROCESS

AT VARIOUS RAW MATERIAL STOICHIOMETRIES

	Raw material stoichiometry			Total capital investment		First-year unit revenue requirements	
Process	Variation	Value[a]	% change[b]	$/kW	% change[b]	mills/kWh	% change[b]
Lime spray dryer	Low	1.1	-10	153.3	-0.58	6.85	-1.01
	Base	1.22	-	154.2	-	6.92	-
	High	1.46	19.7	156.1	1.23	7.06	2.02
Limestone scrubbing	Base	1.12	-	176.1	-	7.90	-

a. Raw material stoichiometry is defined as mols of alkali per mol of SO_2 absorbed.
b. Change is calculated relative to the base-case value.

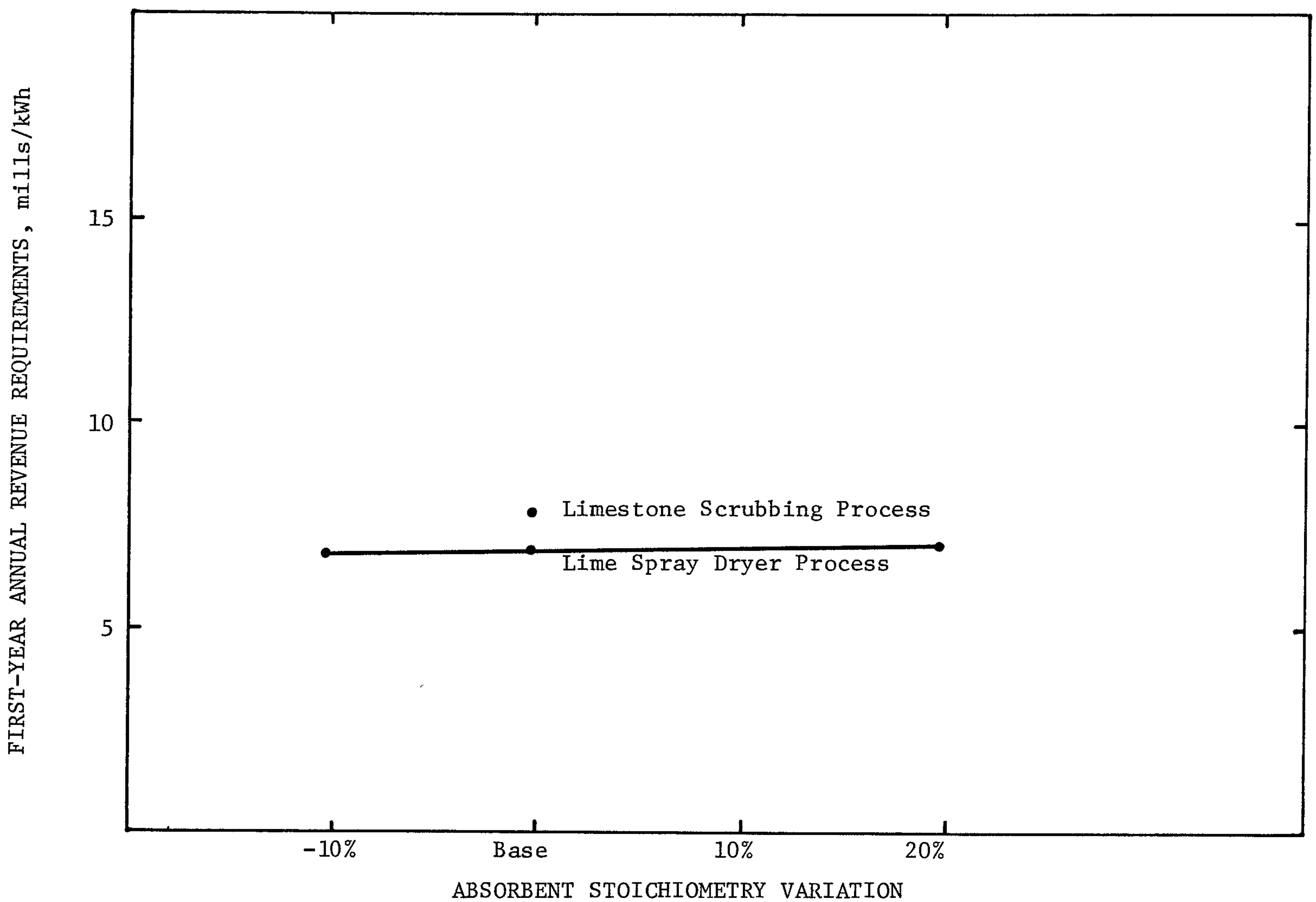

Figure 36. Low-sulfur western coal case--Sensitivity of the first-year annual revenue requirements to the raw material stoichiometry in the absorber.

TABLE 51. LOW-SULFUR EASTERN COAL CASE

CAPITAL INVESTMENT

LIME SPRAY DRYER PROCESS

(500-MW new coal-fired unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	1,762
Feed preparation	909
Gas handling	9,770
SO_2 absorption	7,336
Particulate removal	11,523
Particulate handling	753
Total process capital	32,053
Services, utilities, and miscellaneous	1,923
Total direct investment excluding landfill	33,976
Solids disposal	855
Landfill construction	2,939
Total direct investment	37,770
Indirect Investment	
Engineering design and supervision	2,437
Architect and engineering contractor	709
Construction expense	5,671
Contractor fees	1,846
Contingency	9,516
Total fixed investment	57,949
Other Capital Investment	
Allowance for startup and modifications	5,300
Interest during construction	8,906
Royalties	320
Land	905
Working capital	1,923
Total capital investment	75,303
Dollars of total capital per kW of generation capacity	150.61

Basis
- Midwest plant location represents project beginning mid-1980, ending mid-1983. Average cost basis for scaling, mid-1982.
- Minimum in-process storage, redundant scrubber train, and pumps are spared.
- Landfill located one mile from power plant.
- FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.
- Only nominal construction overtime included.

TABLE 52. LOW-SULFUR EASTERN COAL CASE

CAPITAL INVESTMENT

LIMESTONE SCRUBBING PROCESS

(500-MW new coal-fired power unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	1,011
Feed preparation	1,944
Particulate removal	11,688
Gas handling	11,665
SO_2 absorption	15,597
Stack gas reheat	1,225
Solids separation	1,846
Total process capital	44,976
Services, utilities, and miscellaneous	2,699
Total direct investment excluding landfill	47,675
Solids disposal	743
Landfill construction	2,625
Total direct investment	51,043
Indirect Investment	
Engineering design and supervision	3,390
Architect and engineering contractor	980
Construction expense	7,838
Contractor fees	2,515
Contingency	6,807
Total fixed investment	72,573
Other Capital Investment	
Allowance for startup and modifications	5,454
Interest during construction	11,205
Royalties	-
Land	795
Working capital	2,590
Total capital investment	92,617
Dollars of total capital per kW of generation capacity	185.23

Basis
- Midwest plant location represents project beginning in early 1981 and ending in late 1983. Average cost basis for scaling, mid-1982.
- Minimum in-process storage, redundant scrubber train and feed preparation area, pumps spared.
- Disposal area located one mile from power plant.
- FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.
- Only nominal construction overtime included.

TABLE 53. LOW-SULFUR EASTERN COAL CASE

DIRECT INVESTMENTS AND CAPITAL INVESTMENTS

Process	Direct investment		Total capital investment	
	M$	$/kW	M$	$/kW
Lime spray dryer	37.8	75.5	75.3	150.6
Limestone scrubbing	51.0	102.1	92.6	185.2

The major investment differences between the lime spray dryer process and the limestone scrubbing process are in the SO_2 absorption and gas handling areas as shown in Table 54. Other less significant differences are in the feed preparation, stack gas reheat, material handling, and particulate handling/solids separation areas. With the exception of the material handling area, the limestone scrubbing process costs are higher than those for the corresponding areas in the lime spray dryer process. The investment difference in the SO_2 absorption area alone is about $8.2M. The use of a spray dryer as an absorber results in a less expensive area cost since it eliminates the need for large slurry recirculating tanks and pumps, mist eliminators, and the forced-oxidation equipment which are required in the limestone scrubbing process. The gas handling area is substantially higher for the limestone scrubbing process primarily because of the requirement for additional ductwork costs for bypassing some of the flue gas around the scrubber area to the stack plenum rather than the individual scrubber bypass which is used in the lime spray dryer process. The higher investment for the solids separation area in the limestone scrubbing process, as compared with the particulate handling area in the lime spray dryer process, is due to the nature of the wastes. The FGD waste from the lime spray dryer process is collected dry and is simply conveyed to a storage silo before being trucked to the landfill (there is no waste recycle). The FGD waste from the limestone slurry is pumped through a thickener and a filter and conveyed to a storage pile before being trucked to a landfill.

The investment for the feed preparation area is slightly higher for the limestone scrubbing process while the solids disposal equipment cost, landfill construction cost, and land cost are slightly higher for the lime spray dryer process. The material handling area cost for the lime spray dryer process is higher than the equivalent limestone scrubbing area because of the need to store lime in silos rather than in a pile, as is the case with limestone.

TABLE 54. LOW-SULFUR EASTERN COAL CASE

SUMMARY OF THE CAPITAL INVESTMENTS

Investment area	Total cost, k$ Lime spray dryer process	Limestone scrubbing process
Material handling	1,762	1,011
Feed preparation	909	1,944
Gas handling	9,770	11,665
SO_2 absorption	7,336	15,597
Stack gas reheat	-	1,225
Particulate removal	11,523	11,688
Particulate handling	753	-
Solids separation	-	1,846
Solids disposal	855	743
Landfill construction	2,939	2,625
Land	905	795
All other capital costs	38,551	43,478
Total capital investment	75,303	92,617

Basis: TVA design and economic premises

In order to facilitate comparisons of the economic results with those from other EPA-sponsored evaluations, the capital investments for each process are presented in a slightly different form in Table 55. In this table the total capital investment is broken down into three areas: SO_2 absorption, particulate removal, and waste disposal. Each area includes not only the direct investment for equipment, piping, electrical equipment, etc., but also its pro rata share of the indirect investments and all other capital charges. Previously identified direct investment areas grouped in the SO_2 absorption area are: material handling, feed preparation, gas handling, and SO_2 absorption. The particulate removal area consists only of the original particulate removal area. The waste disposal area combines the particulate handling, solids separation, solids disposal, and landfill/pond construction.

LOW-SULFUR EASTERN COAL CASE--ANNUAL REVENUE REQUIREMENTS

Results

The first-year annual revenue requirements for the lime spray dryer process are $18.67M in 1984 dollars as shown in Table 56. This corresponds to a first-year unit revenue requirement of 6.79 mills/kWh. Equivalent levelized annual revenue requirements for the lime spray dryer process

TABLE 56. LOW-SULFUR EASTERN COAL CASE

ANNUAL REVENUE REQUIREMENTS

LIME SPRAY DRYER PROCESS

(500-MW new coal-fired power unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Lime	11,300 tons	75.00/ton	848
Total raw material cost			848
Conversion costs			
Operating labor and supervision			
FGD	25,400 man-hr	15.00/man-hr	381
Solids disposal	30,505 man-hr	21.00/man-hr	641
Utilities			
Fuel	205,625 gal	1.60/gal	329
Process water	150,821 kgal	0.14/kgal	21
Electricity	39,410,135 kWh	0.037/kWh	1,458
Maintenance			
Labor and material			2,058
Analysis	4,199 man-hr	21.00/man-hr	88
Total conversion costs			4,976
Total direct costs			5,824
Indirect Costs - First-Year			
Overheads			
Plant and administrative			1,688
Total first-year operating and maintenance costs			7,512
Levelized capital charges (14.7% of total capital investment)			11,070
Total first-year annual revenue requirements			18,582
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			14,168
Levelized capital charges (14.7% of total capital investment)			11,070
Levelized annual revenue requirements			25,238

	M$	Mills/kWh
First-year annual revenue requirements	18.58	6.76
Levelized annual revenue requirements	25.24	9.18

Basis
- Midwest plant location, 1984 revenue requirements.
- Remaining life of power plant, 30 years.
- Power unit onstream time, 5,500 years.
- Coal burned, 1,116,500 tons/yr, 9,500 Btu/kWh.
- Total direct investment, $37,770,000; total fixed investment, $57,949,000; and total capital investment, $75,303,000.

are \$25.40M, or 9.24 mills/kWh. The first-year revenue requirements for the limestone scrubbing process are \$22.87M as shown in Table 57. This is equivalent to a first-year unit revenue requirement of 8.34 mills/kWh. Levelized annual revenue requirements are \$31.21M, or 11.35 mills/kWh.

TABLE 55. LOW-SULFUR EASTERN COAL CASE

CAPITAL INVESTMENTS

Area	Investment, \$/kW: Lime spray dryer process	Investment, \$/kW: Limestone scrubbing process
SO_2 absorption	85.63	121.93
Particulate removal	48.84	44.56
Waste disposal	16.13	18.72
Capital investment	150.60	185.21

Basis: TVA design and economic premises

Comparison

The first-year and the levelized annual revenue requirements for both the lime spray dryer and the limestone scrubbing processes are given in Table 58. In both first-year and levelized annual revenue requirements the lime spray dryer process is substantially lower (19% lower) in cost than the limestone scrubbing process.

TABLE 58. LOW-SULFUR EASTERN COAL CASE

FIRST-YEAR AND LEVELIZED ANNUAL REVENUE REQUIREMENTS

Process	First-year revenue requirements: M\$	First-year revenue requirements: Mills/kWh	Levelized annual revenue requirements: M\$	Levelized annual revenue requirements: Mills/kWh
Lime spray dryer	18.58	6.76	25.24	9.18
Limestone scrubbing	22.95	8.34	31.21	11.35

Basis: TVA design and economic premises

TABLE 57. LOW-SULFUR EASTERN COAL CASE

ANNUAL REVENUE REQUIREMENTS

LIMESTONE SCRUBBING PROCESS

(500-MW new coal-fired power unit, 0.7% S in coal; 70% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Limestone	18,400 tons	8.50/ton	156
Total raw material cost			156
Conversion costs			
Operating labor and supervision			
FGD	40,110 man-hr	15.00/man-hr	602
Solids disposal	27,276 man-hr	21.00/man-hr	573
Utilities			
Fuel	151,250 gal	1.60/gal	242
Steam	93,780 klb	2.50/klb	234
Process water	125,097 kgal	0.14/kgal	18
Electricity	41,015,749 kWh	0.037/kWh	1,518
Maintenance			
Labor and material			3,355
Analysis	3,329 man-hr	21.00/man-hr	70
Total conversion costs			6,612
Total direct costs			6,768
Indirect Costs - First-Year			
Overheads			
Plant and administrative			2,563
Total first-year operating and maintenance costs			9,331
Levelized capital charges (14.7% of total capital investment)			13,615
Total first-year annual revenue requirements			22,946
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			17,598
Levelized capital charges (14.7% of total capital investment)			13,615
Levelized annual revenue requirements			31,213

	M$	Mills/kWh
First-year annual revenue requirements	22.95	8.34
Levelized annual revenue requirements	31.21	11.35

Basis
- Midwest plant location, 1984 revenue requirements.
- Remaining life of power plant, 30 years.
- Power unit onstream time, 5,500 hr/yr.
- Coal burned, 1,116,500 tons/yr, 9,500 Btu/kWh.
- Total direct investment, $51,043,000; total fixed investment, $72,573,000; and total capital investment, $92,617,000.

Table 59 compares the various cost components of the first-year annual revenue requirements for each process. With the exception of the raw material and the "all other" categories, the limestone scrubbing process has the higher annual cost. From Table 59 it is apparent that the limestone scrubbing process has higher revenue requirements primarily because of higher annual costs in the same three areas as the other coal cases: levelized capital charges, maintenance costs, and overhead costs. These areas alone are about $4.7M higher in the limestone scrubbing process than the lime spray dryer process.

TABLE 59. LOW-SULFUR EASTERN COAL CASE

SUMMARY OF FIRST-YEAR ANNUAL REVENUE REQUIREMENTS

	Total cost, k$	
Annual cost	Lime spray dryer process	Limestone scrubbing process
Raw material	848	156
Operating labor and supervision	1,022	1,175
Electricity	1,458	1,518
Steam	-	234
Maintenance	2,058	3,355
Overheads	1,688	2,563
Levelized capital charges	11,070	13,615
All other	438	330
Total first-year annual revenue requirements	18,582	22,946

Basis: TVA design and economic premises

The higher levelized capital charge for the limestone scrubbing process is due to its higher capital investment. Maintenance costs are much higher for the limestone scrubbing process because of the maintenance for equipment needed to handle and grind the limestone and also to handle and recirculate large quantities of the erosive slurry. Overheads are higher primarily because of the higher maintenance costs for the limestone scrubbing process.

Other somewhat higher costs for the limestone scrubbing process include operating labor, steam, and electricity. Operating labor is higher because of the operation of additional equipment. The steam consumption is for supplemental reheat (only a small reheater is required for this coal case) which the spray dryer process is claimed not to need. Electrical costs are somewhat higher because of the large pumps required to recirculate the slurry.

The raw material cost is the only major category in which the lime spray dryer process is higher than the limestone scrubbing process. This higher raw material cost in the lime spray dryer process is primarily due to the higher unit cost for lime ($75/ton for lime versus $8.50/ton for limestone) and also because a higher stoichiometry is used with the lime spray dryer process. The effects on the first-year annual revenue requirements of changing the unit cost of the raw material and also the raw material stoichiometry are discussed in the next section in the form of a sensitivity analysis.

LOW-SULFUR EASTERN COAL CASE--SENSITIVITY ANALYSIS

Sensitivity to Raw Material Prices

The sensitivity of the first-year annual revenue requirements for the lime spray dryer process and the limestone scrubbing process to the delivered absorbent cost was calculated for the range of costs listed in Table 60. The results of this sensitivity analysis are shown in Figure 37.

TABLE 60. LOW-SULFUR EASTERN COAL CASE

DELIVERED UNIT RAW MATERIAL COSTS ASSUMED FOR THE

SENSITIVITY ANALYSIS

Process	Raw material	Variation	$/ton	% change
Lime spray dryer	Lime	Low	60.00	-20
		Base	75.00	-
		High	105.00	+40
Limestone scrubbing	Limestone	Low	7.00	-20
		Base	8.50	-
		High	12.00	+40

Although the lime spray dryer process is more sensitive than the limestone scrubbing process to changes in the delivered price of the absorbent, because of the low-sulfur nature of the coal and the low SO_2 removal requirement the changes from the base case costs do not significantly change the economic results. The lime spray dryer process has lower first-year annual revenue requirements regardless of the absorbent prices selected (within the -20%, +40% range). For example, a 40% increase in the delivered cost of lime results in only a 1.8% increase in the first-year annual revenue requirements for the lime spray dryer process which is still 18% less than the first-year annual revenue requirements for the base case limestone scrubbing process.

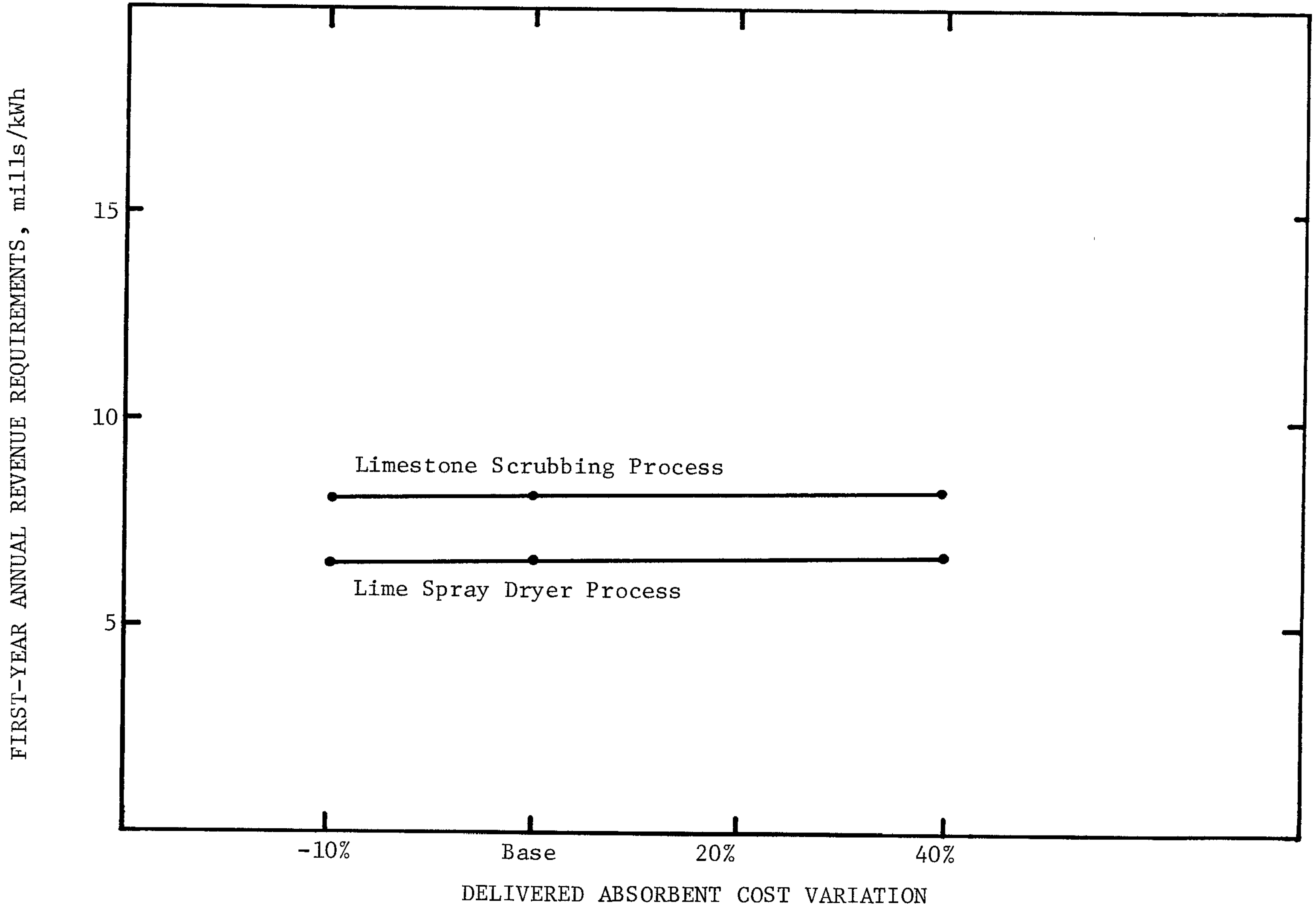

Figure 37. Low-sulfur eastern coal case--Sensitivity of the first-year annual revenue requirements to the delivered raw material cost.

The limestone scrubbing process, due to the low unit cost of limestone as well as the lower sulfur level in the coal and the lower SO_2 removal requirements, is essentially insensitive to the delivered cost of limestone. A 40% increase in the cost of limestone results in only a 0.4% increase in the first-year annual revenue requirements for the limestone scrubbing process.

Sensitivity to Raw Material Stoichiometry

Table 61 lists both the base case and the alternative stoichiometries used in the sensitivity analysis for the low-sulfur eastern coal. (The raw material stoichiometries are given as moles of alkali per mole of SO_2 absorbed.) The range of stoichiometries for the lime spray dryer process is 1.17 (-10%) to 1.56 (20%). The results are shown in Figure 38.

Since many of the processing areas that are dependent on the absorbent flow rate contribute only minor amounts to the capital investment, a 20% increase in absorbent flow rate increases the capital investment only about 1%. The annual revenue requirements for the lime spray dryer process are somewhat more sensitive to the absorbent stoichiometry than to the absorbent cost. For example, a 20% increase in the raw material stoichiometry results in a 1.9% increase in first-year revenue requirements. However, from these results it is apparent that stoichiometry changes over a wide range will have little effect on the capital investment and annual revenue requirement relationships of the two processes.

HIGH-SULFUR EASTERN COAL CASE--CAPITAL INVESTMENT

Results

The capital investment for the lime spray dryer process for the high-sulfur eastern coal application is \$100.1M (\$200/kW) in mid-1982 dollars as shown in Table 62. The capital investment for the limestone scrubbing process is \$122.0M (\$244/kW) as shown in Table 63.

Comparison

The direct investment and the total capital investment for the two FGD systems are shown in Table 64. The lime spray dryer process is substantially (about 18%) less capital intensive than the limestone scrubbing process for this high-sulfur eastern coal case.

The primary investment difference between the lime spray dryer process and the limestone scrubbing process is in the SO_2 absorption area, as shown in Table 65. As was previously discussed in the other comparison sections, the large difference in investment in the SO_2 absorption area (\$12.6M) is due to the use of the spray dryer as the absorber. Other major differences between the processes are in the material handling and particulate handling/solids separation areas. Again, the reasons for the large investment differences in these areas

TABLE 61. LOW-SULFUR EASTERN COAL CASE

COMPARISON OF CAPITAL INVESTMENT AND FIRST-YEAR UNIT

REVENUE REQUIREMENTS FOR THE LIME SPRAY DRYER PROCESS

AT VARIOUS RAW MATERIAL STOICHIOMETRIES

	Raw material stoichiometry			Total capital investment		First-year unit revenue requirements	
Process	Variation	Value[a]	% change[b]	$/kW	% change[b]	mills/kWh	% change[b]
Lime spray dryer	Low	1.17	-10	149.6	-0.69	6.69	-1.04
	Base	1.30	-	150.6	-	6.76	-
	High	1.56	20	152.7	1.37	6.89	1.92
Limestone scrubbing	Base	1.12	-	185.2	-	8.32	-

a. Raw material stoichiometry is defined as mols of alkali per mol of SO_2 absorbed.
b. Change is calculated relative to the base-case value.

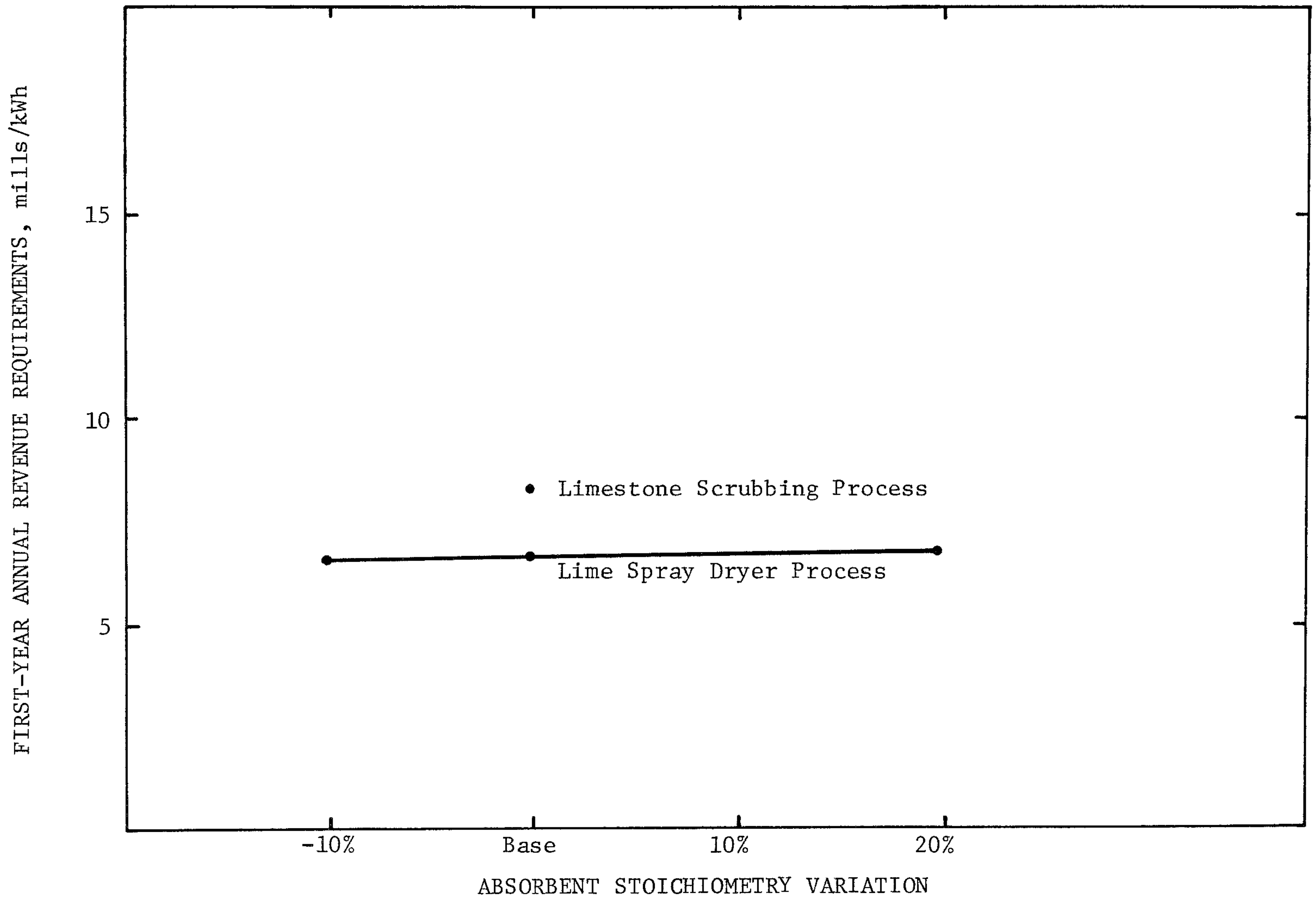

Figure 38. Low-sulfur eastern coal case--Sensitivity of the first-year annual revenue requirements to the raw material stoichiometry in the absorber.

TABLE 62. HIGH-SULFUR EASTERN COAL CASE

CAPITAL INVESTMENT

LIME SPRAY DRYER PROCESS

(500-MW new coal-fired unit, 3.5% S in coal; 88.6% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	5,014
Feed preparation	2,438
Gas handling	11,456
SO_2 absorption	9,018
Particulate removal	11,235
Particulate handling	2,114
Total process capital	41,275
Services, utilities, and miscellaneous	2,477
Total direct investment excluding landfill	43,752
Solids disposal	1,443
Landfill construction	4,899
Total direct investment	50,094
Indirect Investment	
Engineering design and supervision	3,161
Architect and engineering contractor	924
Construction expense	7,392
Contractor fees	2,433
Contingency	12,513
Total fixed investment	76,517
Other Capital Investment	
Allowance for startup and modifications	6,825
Interest during construction	11,712
Royalties	413
Land	1,520
Working capital	3,109
Total capital investment	100,096
Dollars of total capital per kW of generation capacity	200.19

Basis
- Midwest plant location represents project beginning mid-1980, ending mid-1983. Average cost basis for scaling, mid-1982.
- Minimum in-process storage, redundant scrubber train, and pumps are spared.
- Landfill located one mile from power plant.
- FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.
- Only nominal construction overtime included.

TABLE 63. HIGH-SULFUR EASTERN COAL CASE

CAPITAL INVESTMENT

LIMESTONE SCRUBBING PROCESS

(500-MW new coal-fired power unit, 3.5% S in coal; 89.6% SO_2 removal; onsite solids disposal)

	Investment, k$
Direct Investment	
Material handling	2,518
Feed preparation	4,618
Particulate removal	9,998
Gas handling	13,653
SO_2 absorption	21,625
Stack gas reheat	3,325
Solids separation	3,350
Total process capital	59,087
Services, utilities, and miscellaneous	3,545
Total direct investment excluding landfill	62,632
Solids disposal	1,007
Landfill construction	3,441
Total direct investment	67,080
Indirect Investment	
Engineering design and supervision	4,453
Architect and engineering contractor	1,287
Construction expense	10,296
Contractor fees	3,304
Contingency	8,940
Total fixed investment	95,360
Other Capital Investment	
Allowance for startup and modifications	7,165
Interest during construction	14,719
Royalties	-
Land	1,070
Working capital	3,639
Total capital investment	121,953
Dollars of total capital per kW of generation capacity	243.91

Basis
- Midwest plant location represents project beginning in early 1981 and ending in late 1983. Average cost basis for scaling, mid-1982.
- Minimum in-process storage, redundant scrubber train and feed preparation area, pumps spared.
- Disposal area located one mile from power plant.
- FGD process investment begins at boiler air heater exit. Stack plenum and stack excluded.
- Only nominal construction overtime included.

have been previously discussed. The higher landfill construction cost for the lime spray dryer process results from the lower density of the FGD waste in the landfill when compared with the gypsum waste from the limestone scrubbing process. This is the same reason that the solids disposal area investment and the land costs for the lime spray dryer process are somewhat higher than those for the limestone scrubbing process.

TABLE 64. HIGH-SULFUR EASTERN COAL CASE

DIRECT INVESTMENTS AND CAPITAL INVESTMENTS

Process	Direct investment		Capital investment	
	M$	$/kW	M$	$/kW
Lime spray dryer	50.1	100.2	100.1	200.2
Limestone scrubbing	67.1	134.2	122.0	243.9

Basis: TVA design and economic premises

TABLE 65. HIGH-SULFUR EASTERN COAL CASE

SUMMARY OF THE CAPITAL INVESTMENTS

Investment area	Total cost, k$	
	Lime spray dryer process	Limestone scrubbing process
Material handling	5,014	2,518
Feed preparation	2,438	4,618
Gas handling	11,456	13,653
SO_2 absorption	9,018	21,625
Stack gas reheat	-	3,325
Particulate removal	11,235	9,998
Particulate handling	2,114	-
Solids separation	-	3,350
Solids disposal	1,443	1,007
Landfill construction	4,899	3,441
Land	1,520	1,070
All other capital costs	50,959	57,348
Total capital investment	100,096	121,953

Basis: TVA design and economic premises

In order to facilitate comparisons with other EPA-sponsored evaluations, the capital investments for each process are presented in a slightly different form in Table 66. In this table the total capital investment is broken down into three areas: SO_2 absorption, particulate removal, and waste disposal. Each area includes the direct investment for equipment, piping, electrical equipment, etc., and also its pro rata share of the indirect investments and all other capital charges. Previously identified direct investment areas grouped in the SO_2 absorption area are: material handling, feed preparation, gas handling, and SO_2 absorption. The particulate removal area consists only of the original particulate removal area. The waste disposal area combines the particulate handling, solids separation, solids disposal, and landfill/pond construction.

TABLE 66. HIGH-SULFUR EASTERN COAL CASE

CAPITAL INVESTMENTS

Area	Investment, $/kW Lime spray dryer process	Limestone scrubbing process
SO_2 absorption	121.94	177.83
Particulate removal	47.62	38.12
Waste disposal	30.62	27.95
Capital investment	200.18	243.90

Basis: TVA design and economic premises

HIGH-SULFUR EASTERN COAL CASE--ANNUAL REVENUE REQUIREMENTS

Results

The first-year annual revenue requirements for the lime spray dryer process are $31.89M in 1984 dollars, as shown in Table 67. This corresponds to a first-year unit revenue requirement of 11.60 mills/kWh. Levelized annual revenue requirements for the lime spray dryer process are $47.11M, or 17.13 mills/kWh. The first-year annual revenue requirements for the limestone scrubbing process are $32.38M as shown in Table 68. Equivalent unit revenue requirements are 11.78 mills/kWh. Levelized annual revenue requirements are $45.34M, or 16.49 mills/kWh.

TABLE 67. HIGH-SULFUR EASTERN COAL CASE

ANNUAL REVENUE REQUIREMENTS

LIME SPRAY DRYER PROCESS

(500-MW new coal-fired power unit, 3.5% S in coal; 88.6% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Lime	112,400 tons	75.00/ton	8,430
Total raw material cost			8,430
Conversion costs			
Operating labor and supervision			
FGD	29,120 man-hr	15.00/man-hr	437
Solids disposal	36,429 man-hr	21.00/man-hr	765
Utilities			
Fuel	408,125 gal	1.60/gal	653
Process water	143,636 kgal	0.14/kgal	20
Electricity	42,767,100 kWh	0.037/kWh	1,582
Boiler heat loss	137,400 MBtu	3.32/MBtu	456
Maintenance			
Labor and material			2,649
Analysis	4,238 man-hr	21.00/man-hr	89
Total conversion costs			6,651
Total direct costs			15,081
Indirect Costs - First-Year			
Overheads			
Plant and administrative			2,097
Total first-year operating and maintenance cost			17,178
Levelized capital charges (14.7% of total capital investment)			14,714
Total first-year annual revenue requirements			31,892
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			32,398
Levelized capital charges (14.7% of total capital investment)			14,714
Levelized annual revenue requirements			47,112

	M$	Mills/kWh
First-year annual revenue requirements	31.89	11.60
Levelized annual revenue requirements	47.11	17.13

Basis
Midwest plant location, 1984 revenue requirements.
Remaining life of power plant, 30 years.
Power unit onstream time, 5,500 years.
Coal burned, 1,116,500 ton/yr, 9,500 Btu/kWh.
Total direct investment, $50,094,000; total fixed investment, $76,517,000; and total capital investment, $100,096,000.

TABLE 68. HIGH-SULFUR EASTERN COAL CASE

ANNUAL REVENUE REQUIREMENTS

LIMESTONE SCRUBBING PROCESS

(500-MW new coal-fired power unit, 3.5% S in coal; 89.6% SO_2 removal; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First-Year			
Raw materials			
Limestone	132,600 tons	8.50/ton	1,127
Total raw material cost			1,127
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	32,546 man-hr	21.00/man-hr	683
Utilities			
Steam	532,000 klb	2.50/klb	1,330
Fuel	254,562 gal	1.60/gal	407
Process water	188,194 kgal	0.14/kgal	26
Electricity	65,613,193 kWh	0.037/kWh	2,428
Maintenance			
Labor and material			4,404
Analysis	4,988 man-hr	21.00/man-hr	105
Total conversion costs			10,041
Total direct costs			11,168
Indirect Costs - First-Year			
Overheads			
Plant and administrative			3,289
Total first-year operating and maintenance costs			14,457
Levelized capital charges (14.7% of total capital investment)			17,927
Total first-year annual revenue requirements			32,384
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			27,266
Levelized capital charges (14.7% of total capital investment)			17,927
Levelized annual revenue requirements			45,193

	M$	Mills/kWh
First-year annual revenue requirements	32.38	11.78
Levelized annual revenue requirements	45.19	16.43

Basis
- Midwest plant location, 1984 revenue requirements.
- Remaining life of power plant, 30 years.
- Power unit onstream time, 5,500 hr/yr.
- Coal burned, 1,116,500 tons/yr, 9,500 Btu/kWh.
- Total direct investment, $67,080,000; total fixed investment, $95,360,000; and total capital investment, $121,953,000.

Comparison

The first-year and the levelized annual revenue requirements for each of the FGD processes are shown in Table 69. The lime spray dryer process is approximately 2% lower in cost (11.60 mills/kWh versus 11.78 mills/kWh for the first-year) than the limestone scrubbing process in both first-year costs annual revenue requirements. However in terms of levelized annual revenue requirements the lime spray dryer is about 4% higher than the limestone scrubbing process. This is due to the higher ratio of operating and maintenance costs to capital charges in the lime spray dryer process. When levelized the costs for the lime spray dryer process are increased proportionally more than those of the limestone process.

TABLE 69. HIGH-SULFUR EASTERN COAL CASE

FIRST-YEAR AND LEVELIZED ANNUAL REVENUE REQUIREMENTS

Process	First-year annual revenue requirements		Levelized annual revenue requirements	
	M$	Mills/kWh	M$	Mills/kWh
Lime spray dryer	31.89	11.60	47.11	17.13
Limestone scrubbing	32.38	11.78	45.19	16.43

Basis: TVA design and economic premises

Table 70 compares the various component costs of the first-year revenue requirements for both processes. The major cost differences are the capital charges and raw materials. However, the raw materials cost difference, in which the lime spray dryer process is about $7.3M higher, is somewhat balanced by the difference in levelized capital charges, in which the limestone scrubbing process is about $3.2M higher. Other significant differences are the costs for maintenance, overheads, steam, and electricity. The limestone scrubbing process has higher costs for all three. Maintenance costs are calculated as a percentage of direct investment (which is higher for the limestone scrubbing process). The limestone scrubbing process involves grinding the makeup limestone and handling and recirculating large quantities of an erosive limestone slurry. The lime spray dryer process involves only slaking the lime and pumping the resulting slurry to the spray dryer. There are no large recirculating pumps handling large quantities of slurry. This lack of large recirculating pumps for the lime spray dryer process is also the primary reason that the electrical cost is much lower in the lime spray dryer process. Since the overheads are charged based on the operating labor and maintenance costs and the maintenance costs for the limestone scrubbing process are much higher than for the lime spray dryer process, the overhead costs for the limestone scrubbing process are significantly higher ($1.2M). Steam costs are higher in the limestone scrubbing process because of the need for full stack gas reheat.

TABLE 70. HIGH-SULFUR EASTERN COAL CASE

SUMMARY OF FIRST-YEAR ANNUAL REVENUE REQUIREMENTS

Item	Total cost, k$ Lime spray dryer process	Limestone scrubbing process
Raw materials	8,430	1,127
Operating labor and supervision	1,202	1,341
Electricity	1,582	2,428
Steam	-	1,330
Maintenance	2,649	4,404
Overheads	2,097	3,289
Levelized capital charges	14,714	17,927
Other annual costs	1,218	538
Total first-year annual revenue requirements	31,892	32,384

Basis: TVA design and economic premises

HIGH-SULFUR EASTERN COAL CASE--SENSITIVITY ANALYSIS

Sensitivity to Absorbent Prices

The sensitivity of the first-year annual revenue requirements for the lime spray dryer process and the limestone scrubbing process to the delivered absorbent cost was calculated for the range of absorbent costs listed in Table 71. The results of this sensitivity analysis are shown in Figure 39.

The lime spray dryer process is more sensitive than the limestone scrubbing process to changes in the delivered cost of the absorbent. Depending on the lime cost assumed the lime spray dryer process can have first-year annual revenue requirements which are lower, the same, or higher than the limestone scrubbing process. For example, a 20% decrease in the delivered cost of lime results in a 5.3% decrease in the first-year annual revenue requirements for the lime spray dryer process, which are then about 7% less than the first-year annual revenue requirements for the base case limestone scrubbing process. For a 40% increase in

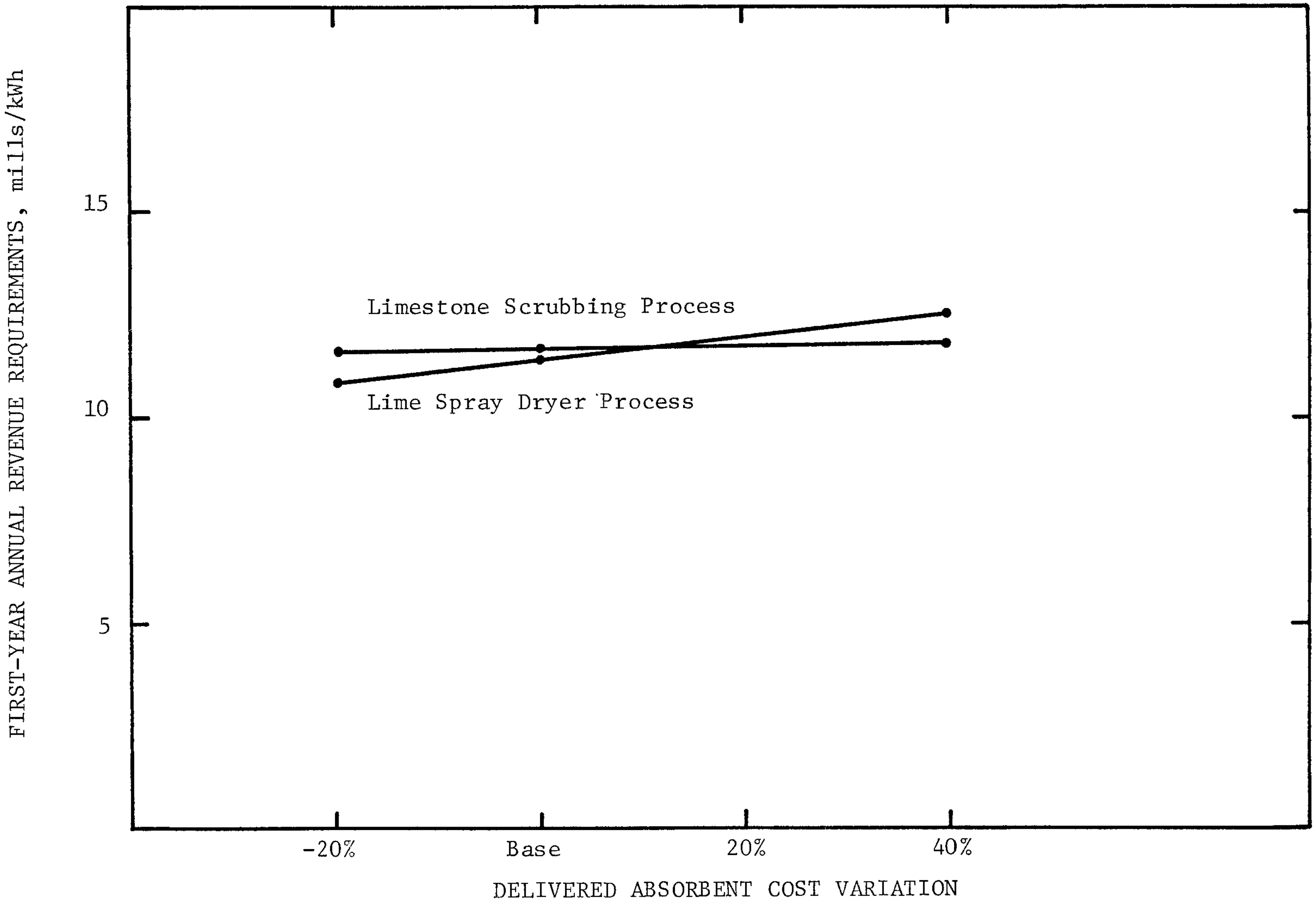

Figure 39. High-sulfur eastern coal case--Sensitivity of the first-year annual revenue requirements to the delivered raw material cost.

the delivered cost of lime, the first-year annual revenue requirements for the lime spray dryer process increase 10.7% and are about 9% higher than the limestone scrubbing process.

TABLE 71. HIGH-SULFUR EASTERN COAL CASE

DELIVERED UNIT RAW MATERIAL COSTS ASSUMED FOR THE

SENSITIVITY ANALYSIS

Process	Raw material	Variation	$/ton	% change
Lime spray dryer	Lime	Low	60.00	-20
		Base	75.00	-
		High	105.00	+40
Limestone scrubbing	Limestone	Low	7.00	-20
		Base	8.50	-
		High	12.00	+40

The limestone scrubbing process, due to the low unit cost of limestone, is essentially insensitive to the delivered cost of limestone. A 40% increase in the cost of limestone results in only a 1.4% increase in the first-year annual revenue requirements for the limestone scrubbing process.

Sensitivity to Absorbent Stoichiometry

Since the lime spray dryer process technology has only been demonstrated on a pilot-plant scale and then primarily with low-sulfur coals, the assumed stoichiometry in the spray dryer could change as the spray dryer technology is developed further. In addition, the required lime stoichiometry for two 3.5% sulfur coals could change depending on the actual coal being burned. Therefore, a sensitivity analysis, showing the changes in total first-year revenue requirements as the absorbent stoichiometry has been included.

Table 72 lists both the base case and the alternative stoichiometries used in the sensitivity analysis. (The raw material stoichiometries are given as moles of alkali per mole of SO_2 absorbed.) The range of stoichiometries for the lime spray dryer process is 1.62 (-10%) to 2.16 (20%). The results are shown in Figure 40.

Since many of the processing areas that are dependent on the absorbent flow rate contribute only minor amounts to the capital investment, a 20% increase in absorbent flow rate increases the capital investment only

TABLE 72. HIGH-SULFUR EASTERN COAL CASE

COMPARISON OF CAPITAL INVESTMENT AND FIRST-YEAR UNIT

REVENUE REQUIREMENTS FOR THE LIME SPRAY DRYER PROCESS

AT VARIOUS RAW MATERIAL STOICHIOMETRIES

	Raw material stoichiometry			Total capital investment		First-year unit revenue requirements	
Process	Variation	Value[a]	% change[b]	\$/kW	% change[b]	mills/kWh	% change[b]
Lime spray dryer	Low	1.62	-10	196.9	-1.65	11.18	-3.62
	Base	1.80	-	200.2	-	11.60	-
	High	2.16	20	206.6	3.20	12.47	7.50
Limestone scrubbing	Base	1.30	-	243.9	-	11.78	-

a. Raw material stoichiometry is defined as mols of alkali per mol of SO_2 absorbed.
b. Change is calculated relative to the base-case value.

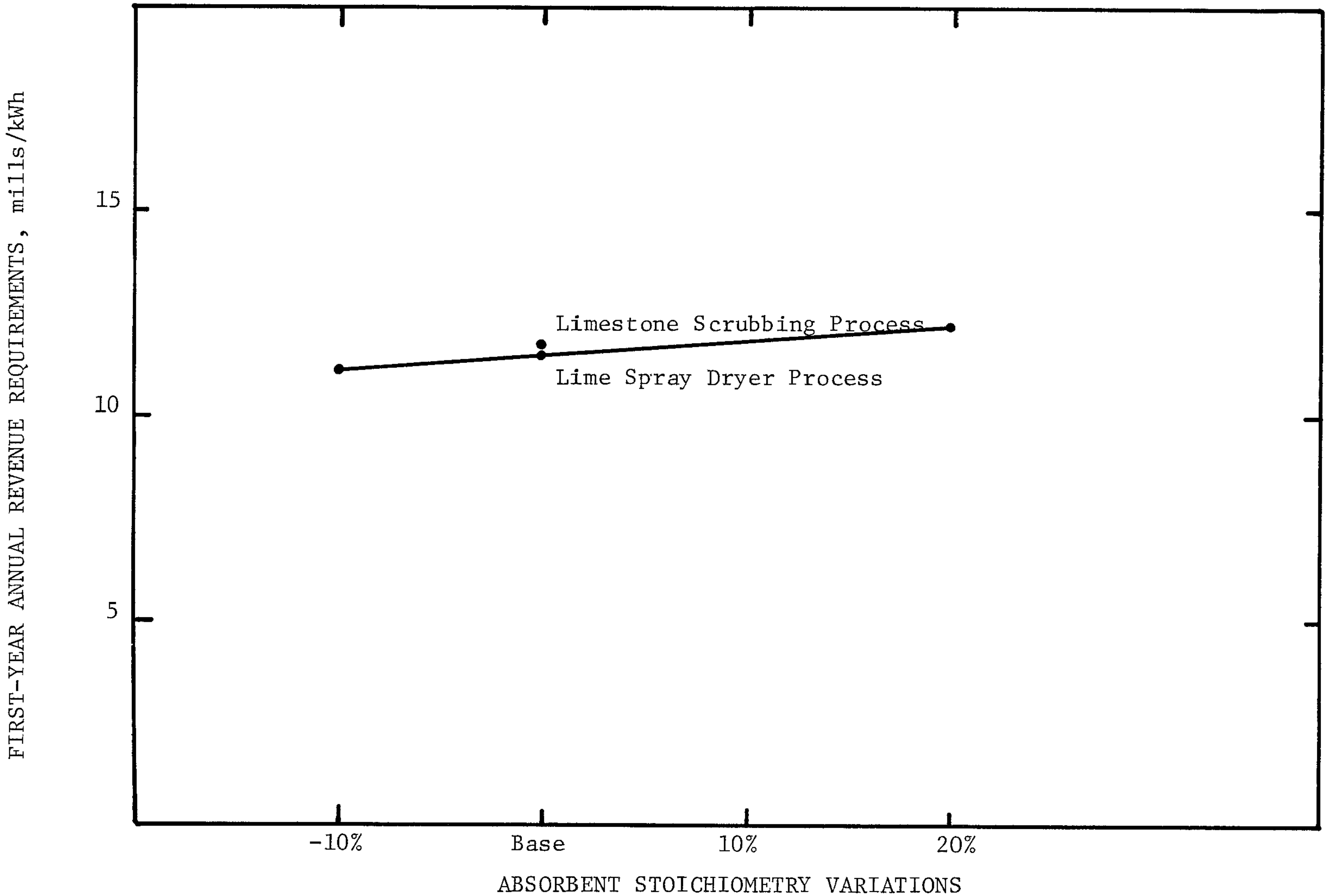

Figure 40. High-sulfur eastern coal case--Sensitivity of the first-year annual revenue requirements to the raw material stoichiometry in the absorber.

about 3%. The annual revenue requirements for the lime spray dryer process are somewhat more sensitive to the stoichiometry than to the absorbent cost. For example, a 20% increase in the stoichiometry results in a 7.5% increase in first-year revenue requirements, which are then 5.5% more than the base case limestone scrubbing process.

DISCUSSION OF RESULTS

The capital investments for the soda ash and lime spray dryer processes and the limestone scrubbing process are summarized in Table 73. The corresponding first-year and levelized annual revenue requirements are summarized in Table 74.

The soda ash process is evaluated for only the low-sulfur western coal case. This application is chosen because it is expected to be the only potential application where the technical problems (i.e., the high solubility of the sodium waste) could be minimized and the economic factors would be optimum (i.e., the delivery cost for soda ash would be low). As is apparent from these tables, the soda ash spray dryer process economics are not favorable even under the most optimistic conditions. In fact in a previous evaluation for a high-sulfur coal application (44), the process economics showed the soda ash spray dryer process to be prohibitively expensive.

The lime spray dryer and the limestone scrubbing processes are evaluated for all four coal applications. In overall capital investment the lime spray dryer process is 12% to 23% lower than the limestone scrubbing process, the difference being greatest for the lignite case and least for the low-sulfur western case. The major cost area for the limestone scrubbing process is SO_2 absorption, representing nearly one-third of the direct capital costs. This cost increases about 44% as the coal sulfur content increases from 0.7% to 3.5%. In contrast, the SO_2 absorption costs for the lime spray dryer process are about one-half those of the limestone scrubbing process for the low-sulfur coal cases and they increase only 23% in going to the 3.5% sulfur case. These SO_2 absorption costs are the major cause of the capital investment cost differences between the processes.

In other areas, the two processes have similar capital costs. The limestone scrubbing process has moderately higher gas handling costs, very slightly higher costs for solids separation (thickening and filtering) compared with particulate handling (pneumatic conveying and silo storage) and slightly lower disposal costs because of the higher bulk density of the gypsum waste. Materials handling costs for the limestone scrubbing process are lower because the limestone can be simply stockpiled. Limestone grinding costs greatly exceed lime slaking costs, however, making the sum of costs for handling and preparing absorbents similar.

TABLE 73. CAPITAL INVESTMENT SUMMARY

	Lignite		Low-sulfur western coal			Low-sulfur eastern coal		High-sulfur eastern coal	
	Lime spray dryer	Limestone scrubbing	Soda ash spray dryer	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing
Direct Costs									
Material handling	1,778	1,291	461	1,691	1,009	1,762	1,011	5,014	2,518
Feed preparation	765	2,406	91	680	1,923	909	1,944	2,438	4,618
Gas handling	10,665	13,249	9,088	10,030	11,646	9,770	11,665	11,456	13,653
SO_2 absorption	7,336	17,357	9,208	7,366	15,054	7,336	15,597	9,018	21,625
Stack gas reheat	-	-	-	-	-	-	1,225	-	3,325
Particulate collection	12,091	15,076	11,523	11,523	11,688	11,523	11,688	11,235	9,998
Particulate handling	2,163	-	750	2,057	-	753	-	2,114	-
Solids separation	-	2,268	-	-	1,828	-	1,846	-	3,350
Total, k$	34,798	51,647	31,121	33,347	43,148	32,053	44,976	41,275	59,087
Other Costs									
Solids disposal	867	790	725	719	616	855	743	1,443	1,007
Disposal site construction	3,756	3,690	7,228	2,520	2,158	2,939	2,625	4,899	3,441
Land	960	920	1,146	770	670	905	795	1,520	1,070
Other capital costs	42,246	50,313	39,228	39,757	41,472	38,551	43,478	50,959	57,348
Total, k$	82,627	107,360	79,448	77,113	88,064	75,303	92,617	100,096	121,953
Total, $/kW	165.25	214.72	158.90	154.23	176.13	150.61	185.23	200.19	243.91

Basis: TVA Design and Economic Premises

TABLE 74. ANNUAL REVENUE REQUIREMENTS SUMMARY

	Lignite		Low-sulfur western coal			Low-sulfur eastern coal		High-sulfur eastern coal	
	Lime spray dryer	Limestone scrubbing	Soda ash spray dryer	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing
Direct Costs									
Absorbent	1,663	227	2,661	1,030	150	848	156	8,430	1,127
Operating labor and supervision	1,025	1,212	846	972	1,140	1,022	1,175	1,202	1,341
Fuel	335	297	265	262	215	329	242	653	407
Electricity	1,617	1,986	1,523	1,464	1,508	1,458	1,518	1,582	2,428
Steam	-	-	-	-	-	-	234	456[a]	1,330
Other utilities	20	20	10	12	17	21	18	20	26
Maintenance	2,232	3,851	1,863	2,136	3,219	2,058	3,355	2,649	4,404
Analysis	88	70	88	88	70	88	70	89	105
Total direct, k$	6,980	7,663	7,256	5,964	6,319	5,824	6,768	15,081	11,168
Indirect Costs									
Overheads	1,794	2,872	1,475	1,717	2,467	1,688	2,563	2,097	3,289
Total O and M, k$	8,774	10,535	8,731	7,681	8,786	7,512	9,331	17,178	14,457
Capital charges	12,146	15,782	11,679	11,336	12,945	11,070	13,615	14,714	17,927
Total, k$	20,920	26,317	20,410	19,017	21,731	18,582	22,946	31,892	32,384
Total, mills/kWh	7.61	9.57	7.42	6.92	7.90	6.76	8.34	11.60	11.78
Levelized									
Total, k$	28,694	35,651	28,146	25,822	29,515	25,238	31,213	47,111	45,193
Total, mills/kWh	10.43	12.96	10.23	9.39	10.73	9.18	11.35	17.13	16.43

a. Boiler heat loss in lieu of reheat

In the lignite case, the lime spray dryer has lower first-year annual revenue requirements than the limestone scrubbing process (7.61 mills/kWh versus 9.57 mills/kWh). With the exception of absorbent costs, where the lime costs are significantly higher than the limestone costs, and the fuel charges, where the differences are insignificant, the lime spray dryer process has lower annual costs in each category than the limestone scrubbing process. The much higher costs for maintenance, overheads, and levelized capital charges for the limestone scrubbing process easily overcome the absorbent cost advantage of using limestone.

In the low-sulfur western coal case, the soda ash spray dryer process has first-year annual revenue requirements of 7.42 mills/kWh, compared with 6.92 and 7.90 for the lime spray dryer and limestone scrubbing processes respectively. For the spray dryer processes the difference is almost entirely the result of absorbent costs, almost 1 mill/kWh for soda ash and 0.4 mill/kWh for lime. Other minor differences account for the remaining cost difference. For the limestone scrubbing process, absorbent costs are minor, less than 0.1 mill/kWh, but maintenance costs are, in general, more than 50% higher than those of the spray dryer processes. The indirect costs, overheads and levelized capital charges, account for the remaining cost difference between the limestone scrubbing process and the spray dryer processes.

For the low-sulfur eastern coal case a similar relationship prevails. Most costs differ insignificantly from those of the low-sulfur western coal case, in spite of the different flue gas bypass conditions. The lime spray dryer costs are slightly lower, primarily because of the lower lime cost in the East. The limestone scrubbing process costs are slightly higher, a result of general cost increases stemming from the lower flue gas bypass ratio. The small amount of flue gas reheat required for the limestone scrubbing process has little effect on the costs. Overall, for each process some cost differences occur for the low-sulfur western and eastern coal processes as a result of different flue gas bypass rates and raw material costs.

For the high-sulfur eastern coal case, somewhat different conditions prevail. The difference in cost between the lime spray dryer process and the limestone scrubbing process decreases from a 12% to 21% advantage for the lime spray dryer process, to only about 2% for the high-sulfur eastern coal case. The increase in cost for both the lime spray dryer process and the limestone scrubbing process in going from the low-sulfur eastern coal case to the high-sulfur eastern coal case is about 42% while the increase for the lime spray dryer process is about 70%. The salient cost factor is absorbent cost. Absorbent costs for the limestone scrubbing process increase about sevenfold. Absorbent costs for the lime spray dryer process increase about tenfold. For the lime spray dryer process, however, this increase results in absorbent costs totaling 27% of the total first-year annual revenue requirements; for the limestone scrubbing process only 3%. Other costs increase little in comparison and in general the increases are similar for both processes. A significant requirement for flue gas reheat also appears in both processes, one in the form of steam, the other in the form of hot flue gas.

The previously discussed first-year annual revenue requirements do not include the effects of inflation or the time-value of money on the annual direct costs (such as raw materials, operating labor, etc.). The levelized annual revenue requirements (shown in Table S-6), however, do take these factors into consideration. As is apparent from Table S-6, levelizing the annual revenue requirements results in a significant increase in the magnitude of the costs. For the lignite and the low-sulfur coal cases where annual direct costs are minor relative to the capital charges, levelizing the revenue requirements does not change the relative economics of the lime spray dryer and the limestone scrubbing processes. However, for the high-sulfur coal case, where the direct costs for the lime spray dryer process are significantly higher than those for the limestone scrubbing process, levelizing the annual revenue requirements results in a reversal whereby the lime spray dryer becomes 4% higher in cost than the limestone scrubbing processes. In fact using the results of this study over the 30-year life of the FGD system, the limestone scrubbing process is $60M less expensive than the lime spray dryer process.

The lime spray dryer process economics are relatively insensitive to both the cost and stoichiometry of lime for the low-sulfur coal applications and only moderately sensitive for the high-sulfur coal case. For a 40% increase in the delivered price of lime, the first-year revenue requirements increase about 2% for the low-sulfur coal cases and nearly 11% for the high-sulfur coal case. For a 20% increase in the raw material stoichiometry, the first-year revenue requirements increase about 2% for the low-sulfur coal cases and about 7% for the high-sulfur coal case.

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16. ABSTRACT The report gives results of an extensive study of dry flue gas desulfurization (FGD) systems, involving dry injection of absorbents or spray drying. (The study was undertaken because they appear to have both process and economic advantages over wet FGD.) Design concepts (e.g., type of absorbent and atomizer, approach to flue gas saturation temperature, and particulate collection method) remain to be demonstrated at full scale. Most vendors prefer a lime slurry system with rotary atomizers and fabric filter particulate collection, while all systems now under contract to utilities apply to low-sulfur coal. SO2 removal efficiencies sufficient for high-sulfur coal applications at stable operating conditions and economically feasible absorbent utilization rates have not yet been demonstrated. In conceptual design cost comparisons based on a new 500-MW utility power generation unit, a lime spray dryer/fabric filter combination had lower capital investments and annual revenue requirements for 0.7% sulfur western coal and both 0.7 and 3.5% sulfur eastern coal than a wet limestone scrubbing process. With lignite fuel, similar cost advantages were evident for dry (relative to wet) FGD. The capital investment advantage of dry over wet FGD increased with increasing coal sulfur content.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Slurries	Pollution Control	13B	
Flue Gases	Dust	Stationary Sources	21B	
Desulfurization	Aerosols	Dry Processes	07A,07D	
Spray Drying	Filtration	Rotary Atomizers	13H	
Sorbents	Fabrics	Particulate	11G	11E
Calcium Oxides	Coal	Fabric Filters	07B	21D
	Calcium Carbonates			

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