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ASSESSMENT OF CONTROL TECHNOLOGIES FOR REDUCING EMISSIONS OF SO_2 AND NO_x FROM EXISTING COAL-FIRED UTILITY BOILERS

FINAL REPORT

by

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16. ABSTRACT The report reviews available information and estimated costs on 15 emission control technology categories applicable to existing coal-fired electric utility boilers. The categories include passive controls such as least emission dispatching, conventional processes, and emerging technologies still undergoing pilot scale and commercial demonstration. The status of each technology is reviewed relative to four elements: (1) Description--how the technology works; (2) Applicability--its applicability to existing plants; (3) Performance--the expected emissions reduction; and (4) Costs--the capital cost, busbar cost, and cost per ton of SO2 and NOx removed. Costs are estimated for new and retrofit applications for various boiler sizes, operating characteristics, fuel qualities, and boiler retrofit difficulties. Capital costs vary from $2/kW for overfire air to $2800/kW for integrated gasification combined cycle in 1988 dollars. (NOTE: A major objective of the National Acid Precipitation Assessment Program is to evaluate alternative methods for reducing SO2 and NOx emissions from combustion sources and to identify options which appear most promising from both an emissions reduction and cost standpoint. Part of this overall effort is to develop up-to-date generic assessments of commercial, near-commercial and emerging emission control technology categories applicable to these utility boilers.)

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Utilities	Cost Effectiveness	Stationary Sources		14A
Boilers	Gasification		13A	14H, 07A
Coal	Emission		21D	14G
Combustion			21B	
Sulfur Dioxide			07B	

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ABSTRACT

A major objective of the National Acid Precipitation Assessment Program is to evaluate alternative methods for reducing SO_2 and NO_x emissions from combustion sources and to identify those options which appear most promising from both an emissions reduction and cost standpoint. Part of this overall effort is to develop up-to-date generic assessments of commercial, near-commercial, and emerging emission control technologies applicable to existing coal-fired electric utility boilers. This report reviews available information and estimated costs on 15 technology categories, including passive controls such as least emission dispatching, conventional processes, and emerging technologies still undergoing pilot scale and commerical demonstration.

The status of each technology is reviewed relative to the following four elements:

- Description--how does the technology work?
- Applicability--what is its applicability to existing plants?
- Performance--what is the expected emissions reduction?
- Cost--what is the capital cost, busbar cost, and cost per ton of SO_2 and NO_x removed?

Cost estimates are presented for new and retrofit applications for various boiler sizes, operating characteristics, fuel qualities, and boiler retrofit difficulty. Capital costs vary from $2 per kilowatt for Overfire Air to $2,800 per kilowatt for Integrated Gasification Combined Cycle in 1988 dollars.

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ABBREVIATIONS AND SYMBOLS

ABBREVIATIONS

ADVACATE	--	advanced silicate process
AEERL	--	Air and Energy Engineering Research Laboratory
AFBC	--	atmospheric fluidized bed combustion
AFDC	--	allowance for funds during construction
ASTM	--	American Society for Testing and Materials
AUSM	--	advanced utility simulation model
CCR	--	Conoco Coal Research
CCTF	--	coal cleaning test facility
CF	--	capacity factor
CS/B +$5	--	coal switching and blending at $5 coal cost differential
CS/B +$15	--	coal switching and blending at $15 coal cost differential
DM	--	dense media
DOE	--	(U.S.) Department of Energy
DSD	--	duct spray drying
EER	--	Energy and Environmental Research Corporation
EPRI	--	Electric Power Research Institute
ESP	--	electrostatic precipitator
FBC	--	fluidized bed combustion
FF	--	new fabric filter
FPD	--	fuel price differential
FSI	--	furnace sorbent injection
GRI	--	Gas Research Institute
HALT	--	hydrate addition at low temperature
HRSG	--	heat recovery steam generator
IAPCS	--	integrated air pollution control system
IGCC	--	integrated gasification combined cycle
kW	--	kilowatt
L/LS FGD	--	lime/limestone flue gas desulfurization
LED	--	least emissions dispatch
LNC	--	low NOx combustion
LNB	--	low NOx burners
LTSI	--	low-temperature sorbent injection
LSD	--	lime spray drying
MHI	--	Mitsubishi Heavy Industries
MW	--	megawatt
NAPAP	--	National Acid Precipitation Assessment Program
NESCAUM	--	Northeast States for Coordinated Air Use Management
NGR	--	natural gas reburn
NSPS	--	new source performance standards

ABBREVIATIONS
(Continued)

NSR	--	normalized stoichiometric ratio
OFA	--	overfire air
PCC	--	physical coal cleaning
PM	--	particulate matter
PFBC	--	pressurized fluidized bed combustion
SCA	--	specific collection area
SCR	--	selective catalytic reduction
S.G.	--	specific gravity
STAR	--	state acid rain
U.S. EPA	--	U.S. Environmental Protection Agency

CONVERSION TO METRIC VALUES

Btu	--	1,054.8 Joules
Btu/lb	--	2,325.445 Joules/Kg
ft^2/1000 acfm	--	0.3048 m^2/actual m^3/sec
gallon	--	0.0038 m^3
pound (lb)	--	0.45359 Kg
lb/10^6Btu	--	0.0043 Kg/Joules
ton	--	907.18 Kg

INTRODUCTION AND SUMMARY

BACKGROUND AND PURPOSE

One of the objectives of the National Acid Precipitation Assessment Program is to evaluate the potential performance and cost of alternative methods for reducing SO_2 and NO_x emissions from combustion sources. Part of this overall effort is to develop up-to-date generic information on commercial, near-commercial, and emerging emission control technologies applicable to coal-fired electric utility boilers. This report presents a review of available information on the technologies shown on Table 1-1. Because the various acid rain regulatory proposals focus on reduction of SO_2 and NO_x in the eastern half of the United States, the report focuses primarily on each technology's potential for retrofit onto existing boilers in the eastern U. S. burning medium and high sulfur coals.

ORGANIZATION

The technology reviews are divided into three major sections covering technologies which are commercial (Section 2), near-commercial (Section 3), and emerging (Section 4). These three classes are respectively defined, as follows: technologies routinely used by U. S. electric utilities, technologies undergoing large-scale demonstration by U. S. utilities or commercially used in Japan or Europe, and those still undergoing laboratory or pilot-scale testing. Designation of a technology to one of these three classes is based on the technology's demonstrated status on low and high sulfur coals.

Within each major section, the technologies are presented in the following order: passive controls, pre-combustion controls, combustion controls, post-combustion controls, and combined systems. The term "passive controls" refers to technologies which in many cases require little or no

TABLE 1-1. CONTROL TECHNOLOGIES REVIEWED

Section	Technology	Potential Emission Reductions (%) SO_2	NO_x
Commercial:			
2.1	Fuel Switching and Blending	50-80	0-10
2.2	Least Emissions Dispatch	0-90	0-40
2.3	Physical Coal Cleaning	20-50	0
2.4	Low NO_x Burners	0	30-50
2.4	Overfire Air	0	15-30
2.5	Lime/Limestone FGD	90-95	0
2.5	Additive Enhanced FGD	90-95	0
2.5	Dual Alkali FGD	90-95	0
2.6	By-Product Recovery FGD	90-95	0
2.7	Spray Drying	70-90	0
Near Commercial:			
3.1	Integrated Gasification Combined Cycle	90-95	90-95
3.2	Fluidized Bed Combustion	80-90	>50
3.3	Selective Catalytic Reduction	0	80-90
3.4	Furnace Sorbent Injection	50-70	0
3.5	Low-Temperature Sorbent Injection	50-70	0
3.6	Reburning	15-20	35-50
Emerging:			
4.1	Advanced Coal Cleaning	45-60	0
4.2	Electron Beam Irradiation	80-95	55-90
4.2	Copper Oxide FGD	90-95	90-95

FGD = Flue Gas Desulfurization.

capital expenditure (i.e., hardware), but which will require changes in a utility's operating methods. The status of each technology is reviewed relative to the following four elements:

- Description--how does the technology work?
- Applicability--what is its applicability to existing plants burning low and high-sulfur coals?
- Performance--what is the expected emissions reduction?
- Cost--what are the capital cost, busbar cost, and cost per ton of SO_2 and NO_x removed?

Because of the importance of consistent treatment of each technology, a consistent set of economic procedures was used for most technologies to allow comparisons. The methodology used for this purpose is discussed below.

METHODOLOGY

Because of the diversity of plant sizes and designs, operating characteristics, fuel quality, and financing arrangements found throughout U. S. utilities, it was necessary to define a uniform methodology for use in the report. These procedures can be divided into two major categories: boiler design and economic assumptions. Base case, high, and low values were selected for boiler design and economic parameters. The range in values was evaluated to present boiler conditions which may favor the selection of one technology option over another. Table 1-2 presents the range of boiler design and economic assumptions selected.

For the technologies addressed in this report, order of magnitude cost estimates are presented. Cost estimates presented in the text are based on a range of boiler and coal parameters. Cost of many technologies is very site-specific and varies significantly depending on the boiler and coal characteristics.

The Integrated Air Pollution Control Systems (IAPCS) (1) cost model, which is currently being updated to include more technologies, was used to develop the cost estimates for some of the technologies in this report.

TABLE 1-2. BASES FOR COST ESTIMATES[a]

Parameter Descriptions	Base Case Value	High Case Value	Low Case Value
Boiler and Coal Characteristic Assumptions			
Unit:			
Size in MW	300	100	700
Capacity factor, %	50	10	90
Specific collection area	300	200	400
Coal Characteristics:			
Sulfur content, %	2.0	1.0	4.0
Switched fuel sulfur content, %	0.9	0.9	0.9
Ash content, %	10.0	5.0	15.0
High heating value, Btu/lb	11,000	9,000	13,000
Economic Assumptions			
Capital Cost Indirects:			
General facilities, %	10%		
Engineering, %	10%		
Project contingencies, %	30%		
Process contingencies, %	0-10%, commercial technologies 10-30%, developing technologies		
Retrofit factor (for FGD or SCR)	1.3	1.5	1.0
Economic Life	20	15	30
Carrying Charge Factor	0.189	0.205	0.175
O&M Levelizing Factor	1.57	1.45	1.75
Operating Costs:			
Fuel price differential, $/ton	10	15	5
Operating labor, $/hr	21.4		
Steam, $/1000 lb	7.0		
Electricity, mills/kWh	57.0		
Lime, $/ton	60.0		
Limestone, $/ton	16.0		
Organic acid, $/ton	1,725.0		
Ammonia, $/ton	150.0		
SCR catalyst, $/ton	20,300.0		
Waste disposal, $/ton	10.0		
Water, $/1000 gal	0.65		
Natural gas, $/$10^6$ Btu	2.0		
Sulfur, $/ton	65.0		

[a]It is EPA's policy to use metric units. However, for the convenience of the reader non-metric units are used in this report. Conversion factors to metric units are given in the table of Abbreviations and Symbols.

The cost/performance assumptions are the same as used under the National Acid Precipitation Assessment Program site-specific retrofit cost study under which the costs of retrofitting SO_2 and NO_x controls at 200 coal-fired utility power plants are being estimated (2). The IAPCS cost model was used to develop cost estimates for the following control technologies.

- Coal switching and blending (CS/B),
- Furnace sorbent injection (FSI),
- Lime spray drying with reuse of the existing electrostatic precipitators (LSD+ESP),
- Lime spray drying with a new fabric filter (LSD+FF),
- Lime/limestone FGD (L/LS FGD),
- Natural gas reburn (NGR),
- Low NO_x burner (LNB),
- Overfire air (OFA),
- Selective catalytic reduction (SCR),
- Integrated gasification combined cycle (IGCC), and
- Atmospheric fluidized bed combustion (AFBC).

For the other technologies addressed in this report, costs are from referenced publications. These costs are not included in this section for comparison, since other cost model assumptions were used in generating costs which may not be consistent with assumptions used in the IAPCS cost estimates.

Economic Assumptions

Cost estimates are presented in 1988 dollars using both current and constant dollar procedures. The Electric Power Research Institute's (EPRI) general costing procedures were used to incorporate inflation, cost of capital, and levelization of future expenses (3). The cost of replacement power or lost capacity while a plant is out-of-service during retrofit is not included in the analysis. Downtime replacement power costs depend on the duration of the downtime period and the cost differential between the purchased or replaced electricity and the cost of power generated by the out-of-service unit. For example, assuming a power cost differential of 10.0 mills/kWh for three different downtime periods of 1, 3, and 6 months

with a capacity factor of 50 percent, the following additional capital investments would be required:

Downtime Period (Months)	Downtime Replacement Power Costs ($/kW)
1	4
3	11
6	22

New coal-fired plant cost of power would be approximately 60 mills/kWhr with half the cost being fixed costs and half the cost being fuel and consumable costs. For post combustion technologies the downtime replacement power cost is less of a factor than for in-situ technologies. Constant dollar calculations are based on standard return on investment (i.e., annuity) calculations without consideration of tax incentives (e.g., accelerated depreciation, investment tax credits) or allowance for funds used during construction (AFDC). The cost calculations include a state and federal income tax rate of 38 percent.

The costs presented in the appendices are in current 1988 dollars and a 30-year book life. To approximate the total levelized busbar cost of power in constant dollars, divide the current dollar costs by 1.75.

SUMMARY OF RESULTS

Table 1-3 and Figures 1-1 through 1-3 summarize for each technology the range of cost estimates developed in Table 1-2 using the high and low case values. The most representative value, the base case, is shown on the figures for each technology as a mid-way point on the bar graphs. This is to show the technology sensitivity to variation in boiler and coal characteristics and that there is no single "winner" for all retrofit applications.

Only those costs which were developed using the IAPCS cost model were presented in this section for consistency. Cost estimates for other technologies which were obtained from other references are presented in the respective technology sections.

Sensitivity case cost estimates developed using the IAPCS cost model are also presented in the appendices. The major cost parameters were varied for the sensitivity analysis. Major cost parameters differ for the different technologies. For example, FGD costs are very sensitive to retrofit factors, coal sulfur content, capacity factor, and boiler size, while coal switching is mainly a function of fuel price differential and percent reduction required. A list of sensitivity case parameters for different technologies is summarized in Table 1-3.

Figures 1-1 through 1-3 present cost estimates for both low and high cases for capital, levelized annual, and unit costs. Costs as well as pollutant removal efficiencies vary for different technologies. These two factors should be balanced in choosing one technology over another and determining the cheapest technology for meeting acid gas removal requirements for a given boiler and coal characteristics.

In this study both high and low sulfur coals are switched to a 0.9 percent West Virginia bituminous coal. Therefore, for high sulfur coal (low case) over 80 percent SO_2 removal is achieved, while for low sulfur coal (high case) the removal value is less than 10 percent. Because of a very low removal efficiency due to switching from one low sulfur coal to another low sulfur coal with less than 10 percent SO_2 removal, the unit cost (dollar per ton of SO_2 removed) resulted in a very large number (the division denominator was a very small value for tons of SO_2 removal). The AFBC and IGCC costs presented are for new systems. The costs for these two technologies are much higher than for other technologies presented in this report because pulverized coal boiler costs (equivalent to AFBC and IGCC) are not included with the other technologies. For FSI, it is assumed that 70 percent SO_2 removal can be achieved with humidification and that existing ESPs are adequate in size and can be reused. Therefore the major cost items are sorbent preparation and modification of the existing furnace for sorbent injection.

SCR costs are much greater than other NO_x removal technologies. This is mainly due to the initial as well as the replaced catalyst cost. However, unlike other NO_x removal technologies, SCR can achieve more than 80 percent NO_x removal.

TABLE 1-3. SUMMARY OF COST RESULTS - CONSTANT 1988 DOLLARS

Technology	Emission Reduction Percent		Capital Costs ($/kW)			Levelized Annual Costs mills/kWh)			Costs Per Unit of Pollutant Removed ($/ton)			Most Sensitive Parameters
	SO_2	NO_x	Low	Base	High	Low	Base	High	Low	Base	High	
Commercial												
Fuel Switching and Blending	2-80	0	20	28	30	3	6	13	350	600	24,000	% S, CF, FPD, SCA
Lime/Limestone FGD	90	0	120	240	520	5	16	150	800	980	3,600	MW, RF, CF, %S
Lime Spray Drying with reuse of existing ESP	76	0	70	170	540	3	10	130	550	750	4,000	MW, RF, CF, %S, SCA
Lime Spray Drying with new fabric filter	86	0	140	240	620	5	13	150	730	850	4,000	MW, RF, CF, %S
Low NO_x Burners	0	50	8	13	25	<1	<1	5	60	190	1,600	% Reduction
Overfire Air	0	25	2	3	6	<1	<1	1	40	130	1,050	MW, CF, % Reduction
Near Commercial												
Add-on Controls:												
Furnace Sorbent Injection	70	0	25	50	110	2	6	40	390	460	1,220	MW, CF, %S, SCA
Natural Gas Reburn	15	60	10	18	28	2	4	9	730	1,400	2,600	FPD
Selective Catalytic Reduction	0	80	90	130	190	3	8	70	950	2,100	14,000	RF, MW, CF, Catalyst life
Advanced Combustion Systems:												
Integrated Gasification Combined Cycle*	95	60-70	1,710	2,100	2,800	44	91	605	--	--	--	MW, CF, heat rate
Atmospheric Fluidized Bed Combustion*	90	20-50	1,360	1,680	2,250	40	80	480	--	--	--	MW, CF, heat rate

MW - size in megawatts, %S - coal sulfur content, CF - capacity factor, SCA - specific collection area of electrostatic precipitator, FPD - fuel price differential, RF - retrofit factor.

Greenfield plant costs. Repowered plant costs would be 10-20% lower in capital cost for IGCC and 25-40% lower in capital cost for AFBC. IGCC repowered plant/unit size would be 2-3 times larger than the existing plant/unit size when the repowered plant/unit is sized based on reuse of the existing steam turbine for combined cycle power generation.

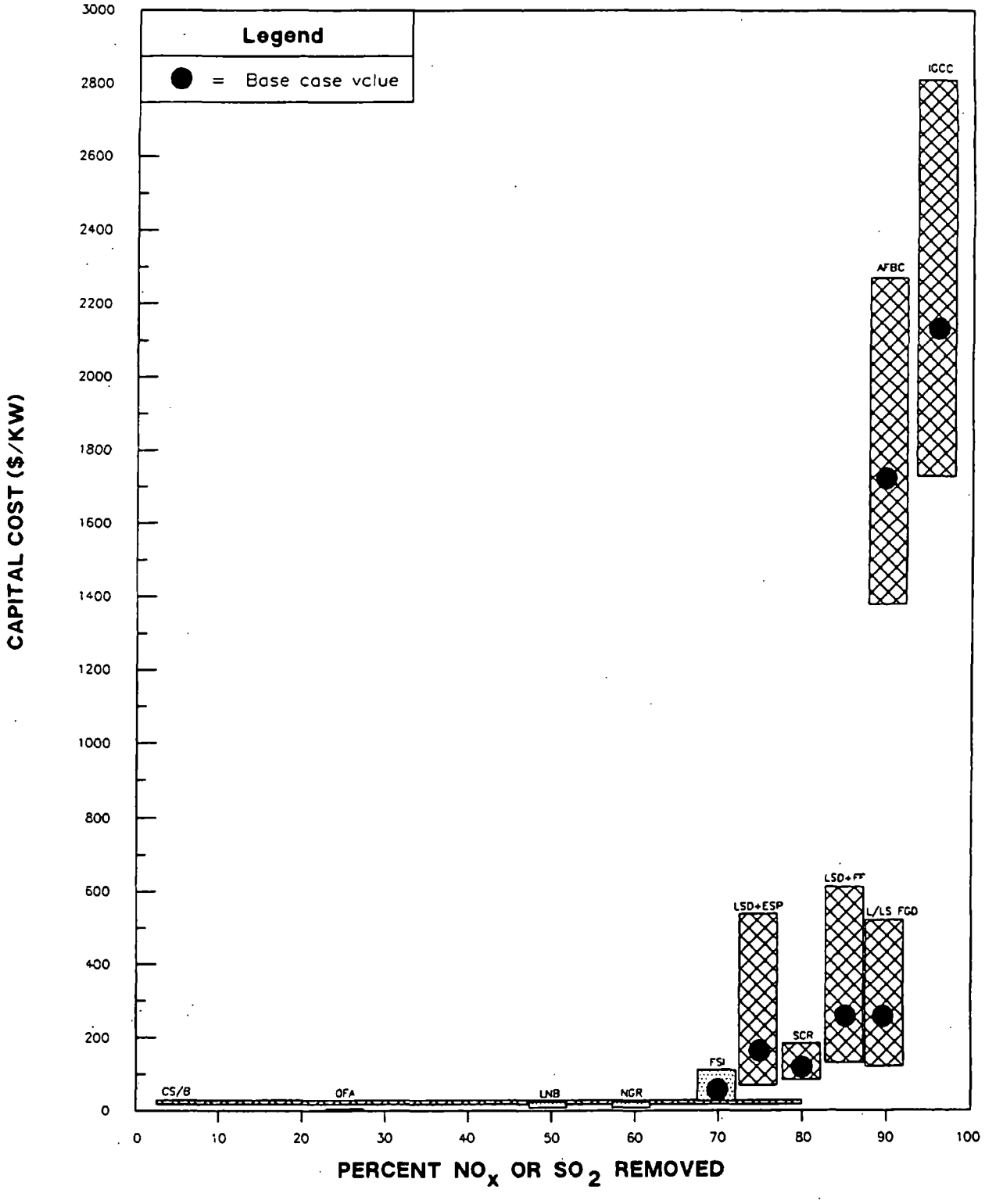


Figure 1-1. Capital Costs - Constant 1988 Dollars

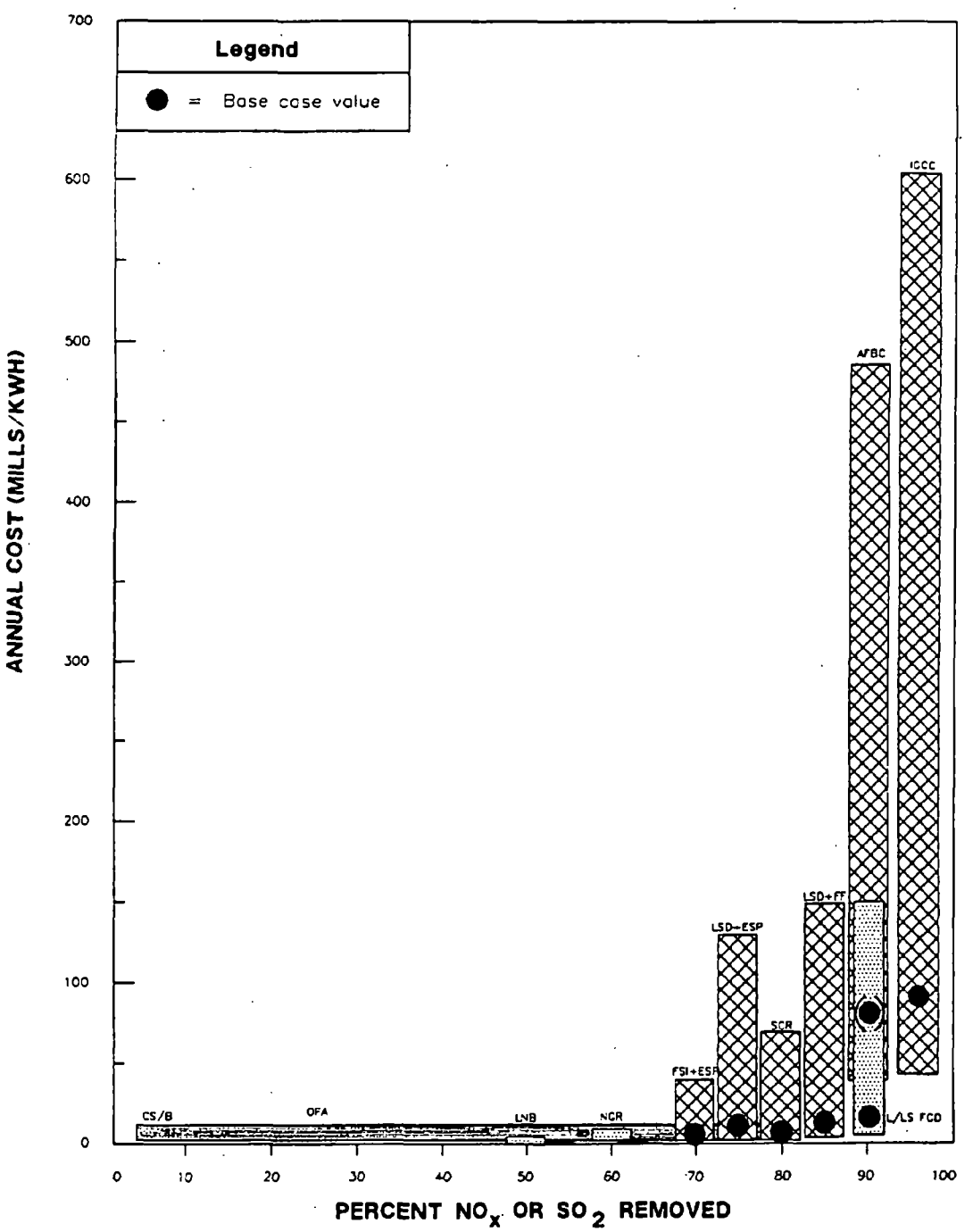


Figure 1-2. Levelized Annual Cost - Constant 1988 Dollars

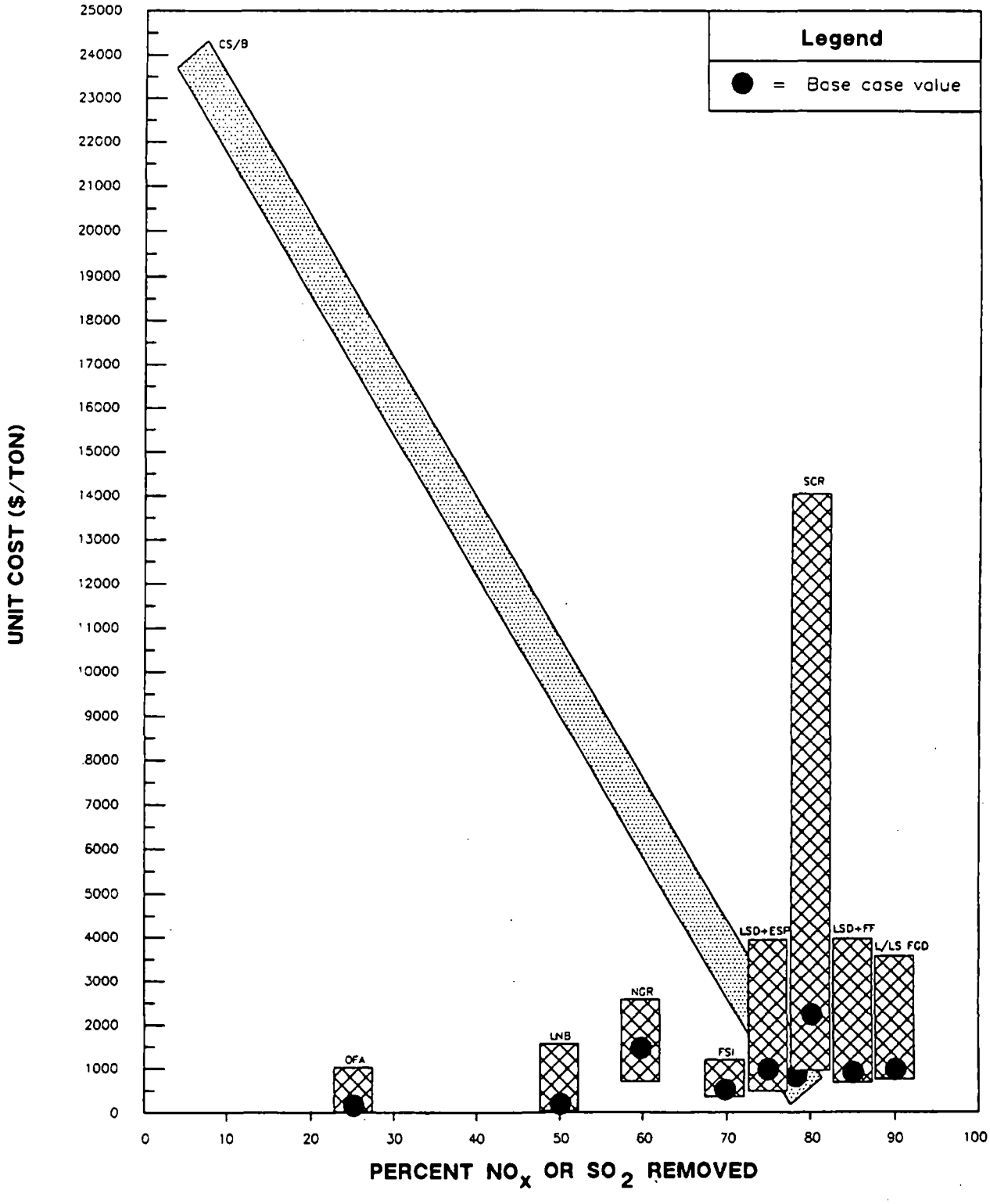


Figure 1-3. Unit Cost - Constant 1988 Dollars

References

1 Maibodi, M., A. L. Blackard, and R. J. Page. Integrated Air Pollution Control System. Draft Report. U. S. Environmental Protection Agency, Research Triangle Park, North Carolina. February 1990.

2. Emmel, T. E. and M. Maibodi. Retrofit Costs for SO_2 and NO_x Control Options at 200 Coal-Fired Plants. Draft Report. U. S. Environmental Protection Agency, Research Triangle Park, North Carolina. December 1989.

3. TAG - Technical Asesssment Guide. (Volume 1). EPRI Report P-4463-SR. Electric Power Research Institute, Palo Alto, California, 1986.

SECTION 2

COMMERCIAL TECHNOLOGIES

INTRODUCTION

A variety of commercial technologies are available for reducing SO_2 and NO_x emissions from existing power plants. These technologies include both passive (primarily associated with technologies which generally require little or no new equipment) and active (based on installation of new equipment) controls. For purposes of presentation, these controls are divided into four main groupings and represent the following six technology areas:

- Passive Controls
  - Fuel Switching and Blending (Section 2.1)
  - Least Emissions Dispatch (Section 2.2)

- Pre-Combustion Controls
  - Physical Coal Cleaning (Section 2.3)

- Combustion Controls
  - Combustion Modifications (Section 2.4)

- Post-Combustion Controls
  - Throwaway Wet Flue Gas Desulfurization (FGD) (Section 2.5)
  - By-product Recovery FGD (Section 2.6)

In addition, a discussion of combining these technologies to further reduce total plant emissions is presented in Section 2.7.

2.1 FUEL SWITCHING AND BLENDING

Description

Switching--

Coal switching represents one of several passive (non-hardware) control methods for emissions reduction. The sulfur content of coals used in the U. S. ranges from less than 0.5 percent up to roughly 6 percent. Thus, one alternative for reducing SO_2 emissions from coal combustion is for power plants to switch from burning high-sulfur coal to lower-sulfur coals.

Most coal-burning facilities are designed to burn coals within a specified range of heating value, ash content, and other physical and chemical properties. Largely as a result of empirical experience gained over time, boiler manufacturers have developed specialized knowledge of how to design a boiler for a given coal. As shown in Figure 2.1-1, significant variations in boiler size and configuration exist among boilers designed for coals with different fuel characteristics. Switching an existing boiler to a different coal, an area in which boiler manufacturers have had relatively little experience, can result in mismatching plant capabilities and may require major modifications to the plant. Primary areas of concern include coal handling and pulverization, combustion kinetics, ash deposition on heat transfer surfaces, particulate collection, and ash handling. Major coal characteristics and hardware systems which must be considered in evaluating coal switching are listed in Table 2.1-1.

The impact of these factors on a given boiler is coal and boiler specific. In general, switching to a coal of the same rank (i.e., conversion from a high-sulfur bituminous to a low-sulfur bituminous) will cause fewer problems than conversion to a lower rank coal. However, other coal properties may still restrict fuel switching in individual boilers. For example, most low-sulfur coals have high fusion temperature ashes which do not slag at normal furnace temperatures, effectively precluding their use in cyclone and wet-bottom boilers.

Power plant systems other than the boiler may also be affected. Of concern in evaluating fuel switching as an SO_2 control option is that low-

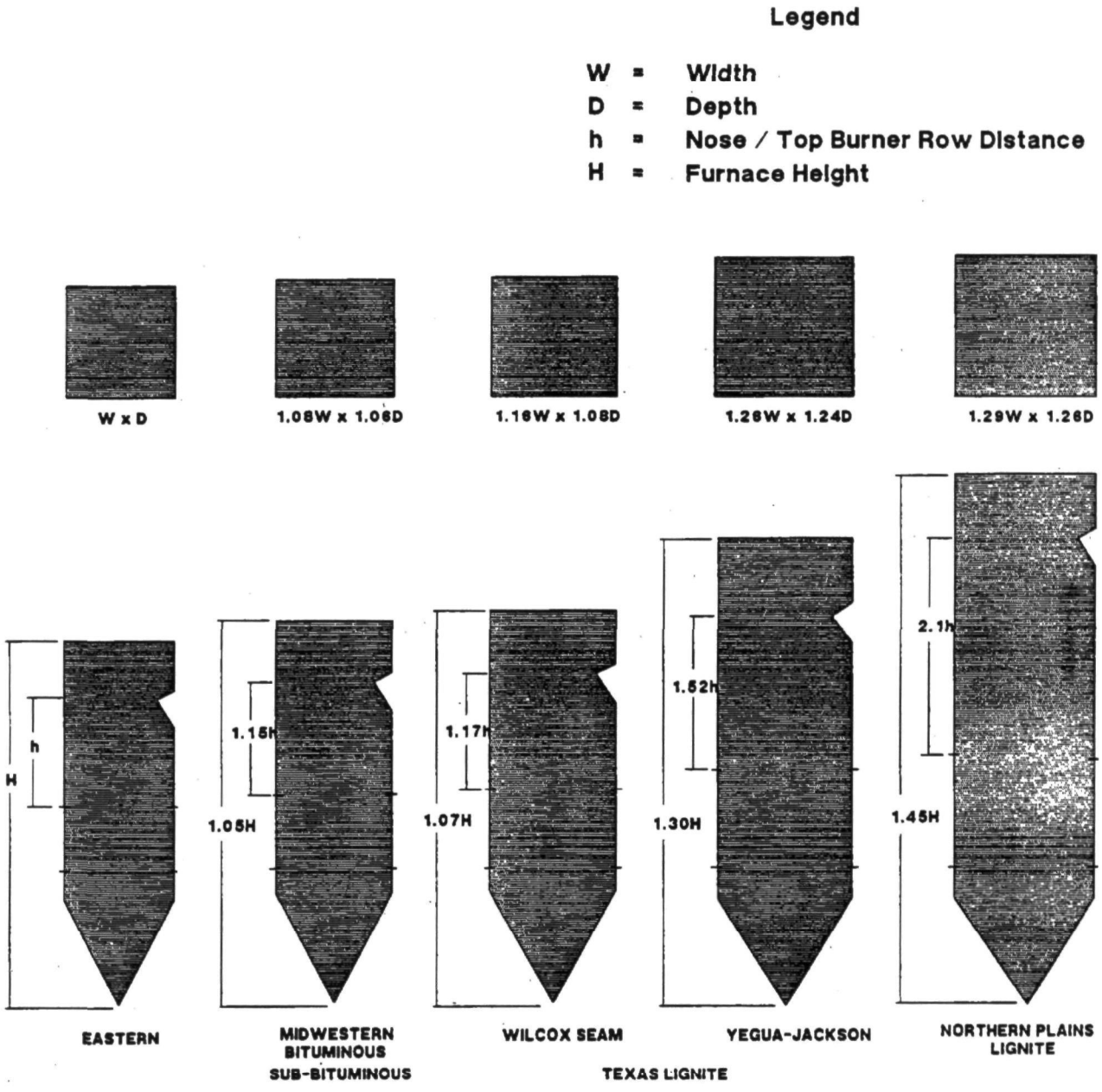


Figure 2.1-1. Effect of coal rank on furnace sizing (1).

TABLE 2.1-1. TECHNICAL FACTORS AFFECTING COAL SWITCHING

Change In Coal Characteristics	Resulting Potential Operating Problems	Possible Solutions
Lower heating value	Insufficient coal handling capacity Unable to achieve design steam output	Enlarge coal handling equipment Derate capacity
Higher moisture	Longer/cooler flames Lower furnace exit temperature Higher gas flow	Derate capacity Increase/modify boiler heat transfer surface area Increase fan capacity
Higher ash	Increased particulates in flue gas Increased solid waste	Increase soot blowing Modify boiler convective section heat transfer area Increase ESP plate area Modify ash handling and disposal systems
Lower sulfur	Lower particulate collection efficiency	Increase ESP plate area Install flue gas conditioning
Higher ash fusion	Incompatible with cyclone and wet-bottom furnaces	Use different coal
Higher sodium and iron content in ash	Increased slagging and fouling	Increase soot blowing Use ash additives Accept higher forced outage rates
Higher volatility	Changed heat transfer characteristics Heating and potential fires in coal handling equipment	Change boiler tube distribution in furnace and convective sections Modify pulverizers, silos, and other coal handling equipment
Harder grindability	Insufficient pulverizer capacity	Increase pulverizer capacity Derate boiler capacity

sulfur coals generally have higher ash resistivity. As a result, ESP performance is likely to be poorer than with high-sulfur coals. This may be especially critical in boilers equipped with older ESP's which are only marginally in compliance with existing particulate standards.

Blending--

Coal blending to reduce the average sulfur content of coals is another strategy for compliance with existing SO_2 emission limits. Depending on plant-specific requirements, "blending" can denote either the use of multiple coals by a single plant even though each boiler at the plant may burn a single coal or the use of multiple coals in a single boiler. In the first case, several distinct coal piles are maintained for supplying individual boilers. A central coal receiving facility and certain other common equipment may be used to handle all coals used at a plant, but most of the equipment connecting the coal pile to the boiler is unit specific. The major limitation on retrofit application of this form of coal blending is the availability of space for additional coal piles.

The second form of blending involves the use of two or more coals in a single boiler. If the individual coals have significantly different combustion or ash characteristics (e.g., reactivity, ash fusion temperature, or base/acid ratio), failure to adequately blend the coals into a uniform mixture can result in rapid fluctuations in fuel composition being fed to the boiler and cause major plant operational problems. Although a number of power plants receive coals from different suppliers, in most cases coals purchased by an individual plant come from the same coal producing area and have similar combustion and ash characteristics. In these cases, blending operations may be fairly simple and relatively inexpensive.

When significant differences in fuel quality exist or when continuous monitoring requirements on plant emissions impose short averaging times, additional coal-handling equipment (stack-out equipment, storage piles, reclaim hoppers, conveyors, and belt scales) must be added to the plant to assure adequate mixing of the individual coals. Capital and operating costs of this equipment can be significant, as evidenced by Detroit Edison's $350 million

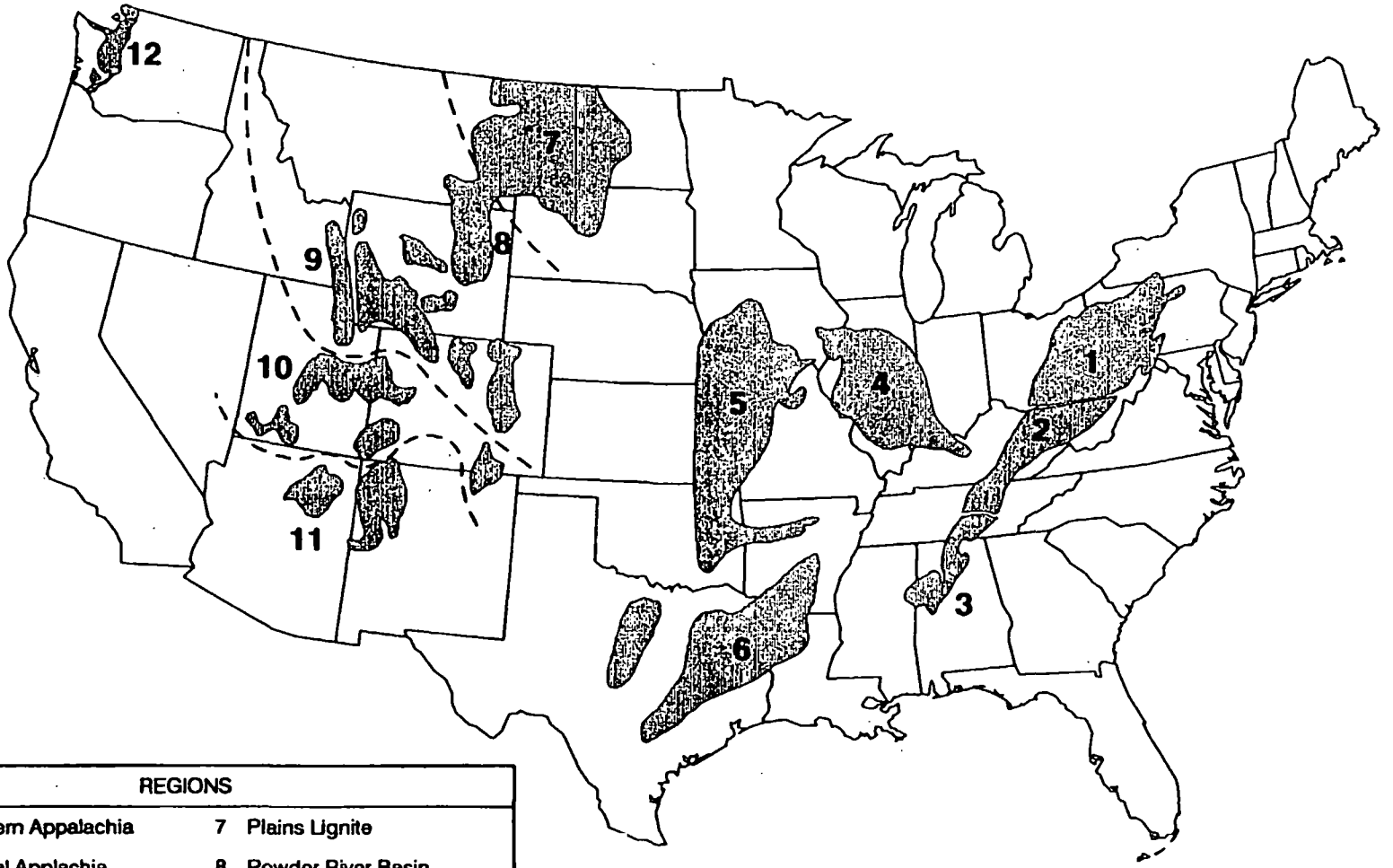
coal blending facility at its Monroe Station. The facility has the capability of blending up to 10,000 tons/hr of high-sulfur bituminous and low-sulfur subbituminous coals.

Combustion characteristics of coal blends cannot be accurately estimated as simple linear averages of the individual coals used. This is especially true for ash slagging, fouling, and resistivity which are influenced by eutectics formed from the mineral matter in the individual coals in the blend. The only way to be sure a given coal or coal blend can be satisfactorily burned in a given boiler is to conduct a test burn of several days to several weeks in the actual boiler. Preliminary testing of candidate blends in a pilot-scale combustor can be useful in better defining potential problems prior to full-scale tests.

Applicability

Low-sulfur coals ($\leq$1 percent sulfur) are available in several regions of the U. S.: central and southern Appalachia, Powder River Basin, southern Wyoming, and Colorado/Utah (Figure 2.1-2). Low-sulfur coals from southern Appalachia are expensive to mine and limited in quantity, and are therefore unlikely to be used as an alternative to high-sulfur coal in existing boilers. Basic coal characteristics for the remaining low-sulfur coals and for high-sulfur coals from northern Appalachia and the Illinois Basin are presented in Table 2.1-2. On a pounds of SO_2 per million Btu basis, low-sulfur coals generally have one-quarter to one-half the sulfur content of the high-sulfur coals.

Central Appalachia coals are used widely in the eastern U. S., accounting for virtually all of the coal used in Virginia, North Carolina, South Carolina, southern West Virginia, eastern Kentucky, and Tennessee. Significant amounts of central Appalachian coal are also used in Ohio and Michigan to meet SIP requirements for low-sulfur coal. Estimated reserves of low-sulfur coal in the region are 23 billion tons. Production in 1982 was 118 million tons. Coal seams in the region are generally thin (2.5 to 7 feet thick). Mining and reclamation in the region is difficult due to the area's mountainous topography and result in high mining costs. Coal prices FOB mine over the last few years have ranged from $30-40/ton (2,3).



REGIONS			
1	Northern Appalachia	7	Plains Lignite
2	Central Applachia	8	Powder River Basin
3	Southern Applachia	9	Southern Wyoming
4	Illinois Basin	10	Colorado/Utah
5	Western Interior	11	Arizona/New Mexico
6	Gulf Lignite	12	Washington

SOURCE: Energy Ventures Analysis, Inc., COALCAST Quarterly Report Service, 1983.

Figure 2.1-2. Coal supply regions (1).

TABLE 2.1.2 TYPICAL DELIVERED COAL PROPERTIES FOR SELECTED SUPPLY REGIONS[a]

Property		Northern Appalachia	Central Appalachia	Illinois Basin	Powder River Basin	Southern Wyoming	Colorado/ Utah
Rank		Bituminous	Bituminous	Bituminous	Subbituminous	Subbituminous	Bituminous
Higher Heating Value, Btu/lb.	range	11,000-13,000	11,500-13,500	10,000-12,000	8,200-9,700	9,000-10,500	10,000-12,500
	average	12,138	12,252	11,114	N/A	N/A	11,130
Sulfur, %	range	0.7-5.0	0.6-3.5	1.0-3.5	0.3-0.7	0.4-1.0	0.3-0.8
	average	2.5	1.1	2.8	N/A	N/A	0.5
Moisture, %		5-10	5-10	5-15	20-30	15-20	7-20
Ash, %	range	7-15	6-15	8-15	5-9	7-11	6-13
	average	12.8	10.9	10.5	N/A	N/A	9.3
Grindability[b]		50-100	40-90	50-70	40-55	40-55	45-50
Ash Fusion Temperature, Fluidity, °F		2,100-2,800	2,100-2,800	1,900-2,400	2,100-2,300	2,100-2,400	2,200-2,600

[a]See Reference 1.
[b]Hardgrove grindability index (ASTM D-409).

Western low-sulfur coals, especially those from the Powder River Basin, are currently being shipped to utilities in the western, south-central, and midwestern U. S. Estimated reserves of low-sulfur coal exceed 140 billion tons. Total production in 1982 was 157 million tons. Coal seams range up to over 100 feet thick in the Powder River Basin where mining is by large pit-type surface mines. Seams in southern Wyoming, Colorado, and Utah are generally thinner and are recovered by both surface and underground mines. Coal prices FOB mine have ranged from less than $6/ton in the Powder River Basin up to $30/ton in Utah (1,3).

Two major questions exist regarding the future price of low-sulfur coals: the impact of acid rain policies on 1) minemouth low-sulfur coal prices (especially in central Appalachia) and 2) transportation rates from the mine to the consumer. Several studies based on computer modeling have estimated that a major acid rain control program could increase the demand for central Appalachia low-sulfur coal by up to 50 million tons over otherwise projected levels and increase minemouth prices by as much as $15 per ton (4). Minemouth costs of low-sulfur western coals are not expected to increase as much due to region's current excess mining capacity and larger reserves. However, regardless of coal prices FOB mine, railroad transportation rates may increase the delivered cost of low-sulfur coal and decrease the attractiveness of fuel switching. Other analysts argue that recent structural and technological changes in the coal and railroad industries are not captured in these modeling efforts, and that price increases will be less than projected (1). For example, increases in labor productivity within the coal industry and the declining demand for metallurgical coal by the steel industry have both worked to keep the cost of coal from central Appalachia below forecast prices.

Performance

In 1985 the 31 states adjoining or east of the Mississippi River contained 230,000 MW of coal-fired capacity. Of this total, roughly 171,000 MW had SO_2 emissions in excess of 1.2 lbs/million Btu. Total annual SO_2 emissions from these units were 12.6 million tons. An estimated 85,000 MW of this total are believed to be technically capable of converting to low-sulfur coal. The remaining 86,000 MW include cyclone and wet-bottom furnaces unable to use low-

sulfur coals due to high ash-fusion temperatures and minemouth plants with limited transportation alternatives. If all of these units are converted to lower sulfur coals, annual SO_2 emissions would decrease by roughly 4-7 million tons depending on the sulfur contents of the coal fired in the units prior to being converted and utility dispatch decisions. The resulting increase in the demand for lower sulfur coals would be roughly 125 million tons per year.

The actual extent of fuel switching will partially depend on the nature of the SO_2 reduction program mandated. Key factors impacting fuel switching benefits at a given plant include the delivered price of low-sulfur coal, the sulfur and Btu content of the new fuel, the relative capital cost of FGD retrofit versus plant modifications for fuel switching, plant-specific considerations such as plant layout and projected capacity factors, and the expected ease of implementing a major SO_2 control program throughout an integrated utility system. Potential SO_2 reductions at a typical plant range from 50-80 percent.

Cost

Fuel switching and blending are generally assumed to have low capital cost. While in many instances the capital cost for conversion between two coals may be small, this assumption is not true in every instance. Unfortunately, many analyses of SO_2 retrofit costs have ignored the technical limitations and capital costs associated with coal switching.

The technical and economic feasibility of switching fuels in a given boiler are specific to the boiler design and location. Where switching is technically feasible, economics will depend on the cost of coal delivered to the plant, the capital and O&M costs associated with necessary plant modifications, the cost of replacement power associated with plant derating when a different coal is used, and the comparative sulfur contents of the existing and candidate coals. In this study using the IAPCS cost model (5), estimated capital costs are for particulate control improvements and fuel price differential (FPD). Particulate control impacts were evaluated by considering the altered particulate loading and particulate resistivity associated with the replacement coal as compared to original conditions. ESP upgrades were costed

by calculating additional plate area in the presence of SO_3 gas conditioning to maintain current PM emission rates. The premiums of $5 and $15 per ton were used to span the range of fuel cost increased associated with low sulfur coal demand. Using the EPRI cost procedures (6), capital costs include preproduction costs estimate as 25 percent of full capacity fuel cost for one month. This results in a higher capital costs for $15 FPD versus the $5 FPD. The IAPCS costs do not include the cost impact of additional coal receiving/storage/handling facilities (if needed) and the cost impact of boiler derate due to pulverizer capacity and boiler fouling, slagging, and erosion. The majority of the capital costs are due to coal inventory costs (preproduction costs).

The following table provides the range of values for capital cost, annualized cost, and cost per ton of pollutant removed estimated using the IAPCS cost model. The lower capital and annualized costs are for a large unit with high capacity factor and $5 FPD, while the higher capital and annualized costs are for a small unit with low capacity factor and $15 FPD. Figures 2.1-3 through 2.1-5 show these costs as a function of boiler size and fuel price differential. Appendix A contains tables for estimating coal switching costs as a function of size, coal sulfur content, capacity factor and FPD.

	Range
Capital Cost ($/kW)	19.1 to 43.1
Annualized Cost (mills/kWh)	5.4 to 17.2
Cost Per ton of SO_2 Removed ($/ton)	213.7 to 20,354.1

The delivered cost of low-sulfur coals used in these calculations may not include sufficient fuel premiums to account for rapid increases in low-sulfur coal demand as a result of "acid rain" legislation.

Table 2.1-3 summarizes key factors influencing the decision to switch coal versus retrofit a wet scrubbing system or some other form of active emission control technology.

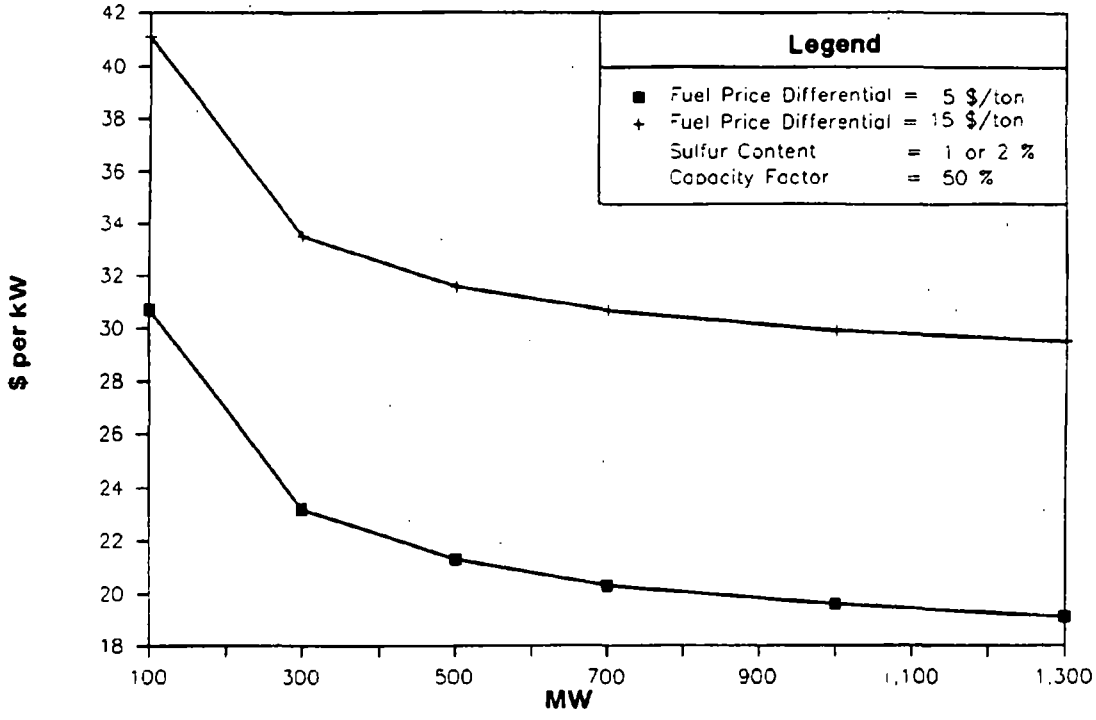


Figure 2.1-3. Coal Switching - Capital Cost, Current 1988 Dollars

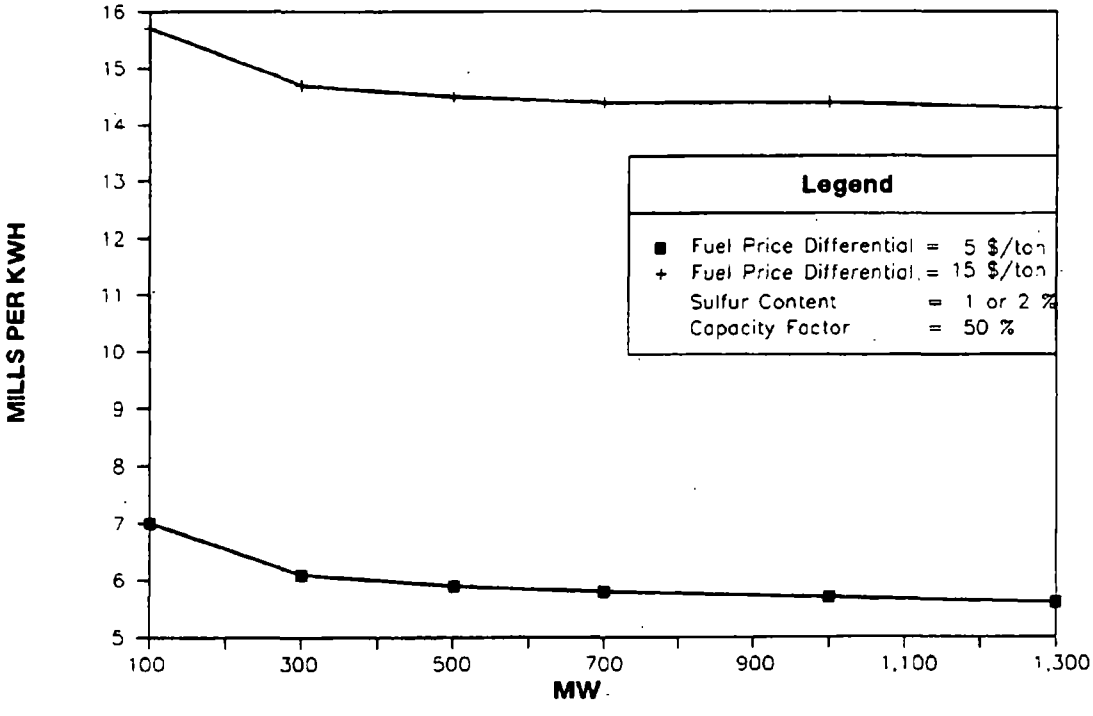


Figure 2.1-4. Coal Switching - Levelized Annual Cost, Current 1988 Dollars

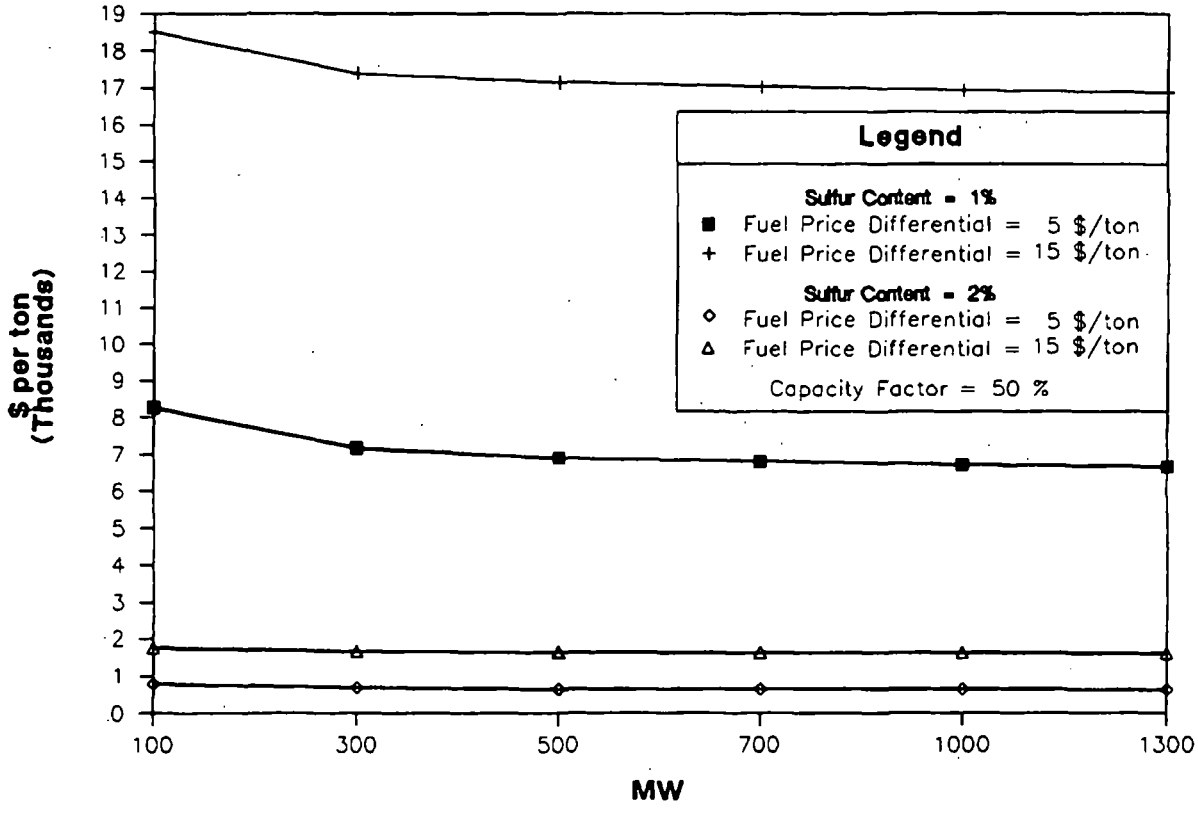


Figure 2.1-5. Coal Switching - Cost Per Ton of SO2 Removed, Current 1988 Dollars

TABLE 2.1-3. FACTORS INFLUENCING THE DECISION TO SWITCH OR SCRUB TO REDUCE SULFUR DIOXIDE EMISSIONS

Factor	Coal Switching	Retrofit Scrubbers
High-sulfur coal cost	high*	low*
Boiler type	dry bottom	cyclone
High-sulfur coal supply	not restrictive	restrictive
Low-sulfur coal transportation costs	low	high
Powerplant age	old	new
Difficulty of scrubber construction	high	low
Existing coal delivery facilities	rail, barge	truck, conveyor
Located in high-sulfur coal producing state	no	yes
Required SO_2 emission reduction	low	high
Capacity margin	high	low
System costs of derating	low	high
Load growth outlook	low	high
Utility financial condition	weak	strong

Source: Reference 1.

*This table is read in the following manner: If the current cost of high sulfur coal delivered to the plant is high, the utility is more likely to choose coal switching. If the cost of the high sulfur coal is low, the utility is more likely to choose a scrubber retrofit.

2.2 LEAST EMISSIONS DISPATCH

Description

One of the key operational decisions facing a utility (or group of interconnected utilities) on day-to-day basis is which plants to operate or "dispatch" to satisfy the demand for electricity at any given time. Presently, utilities dispatch plants in accordance with state regulatory guidelines to minimize the cost of generation, subject to physical constraints such as individual plant availability, flexibility in adjusting to variations in electrical demand, and plant maintenance requirements. If environmental regulations were changed to limit system-wide emissions of SO_2 or if emissions were taxed to reflect the "cost" of emissions, the economics of plant dispatch would be changed and a different dispatch pattern would potentially occur. This alternate dispatch pattern has been referred to in various publications as "least emissions dispatch" (LED) and "environmental dispatch".

Several different LED concepts have been described in the literature. Key differences in these concepts are the time period (temporary vs. permanent) and geographic area affected. One proposal is to use LED to reduce annual emissions over a large area. In this approach, maximum use is made of nuclear and hydroelectric capacity and low-emitting fossil-fuel plants (natural gas, low-sulfur oil and coal, and plants with scrubbers for SO_2 reduction) in preference to plants burning high-sulfur fuels. Interregional transfers of power are also used, including the importation of electricity from Canada.

The other major LED approach would reduce SO_2 emissions only during periods of adverse air quality in environmentally sensitive areas. For example, LED could be used during periods of air stagnation in the Adirondacks. The objective of this approach is to reduce ambient concentrations of SO_2 in selected areas as opposed to reducing total emissions of all plants in a large region. To be effective, this approach requires knowledge of source/receptor relationships affecting pollutant transport and transformation. The only known use of episodic LED has been in the Los Angeles air basin where power plant operations were controlled to reduce NO_x emissions during periods of poor air quality. This effort is reported to have been a failure and was discontinued (7).

Applicability and Performance

Because every utility and region of the U.S. is different, assessment of the potential and cost effectiveness of LED for reducing SO_2 is utility specific. For utilities operating both large, thermally efficient plants equipped with FGD systems and smaller, less efficient plants without pollution control systems, least cost and least emissions dispatching may result in operation of the same plants. For utility systems which can import power from other geographic regions or which depend on plants that are not equipped with pollution control equipment for base-load power generation, LED may significantly alter dispatch decisions. Also, the utility's capacity factor will influence whether LED can be effective. Utilities with low capacity factors may be able to alter dispatching significantly and thus appreciably reduce SO_2 emissions. Utilities with high capacity factors have limited opportunity to shift generation between plants. To the extent low-sulfur coal plants are dispatched in preference to high-sulfur coal plants, the impact of LED in coal markets will be similar to coal switching.

The greatest potential for LED is in a large utility or interconnected utility pool having multiple plants, using a variety of fuels, and having a relatively low system-wide capacity factor. Although the concept of LED has been presented in a number of studies, very little analytical work has been done to quantify its cost effectiveness or the magnitude of potential emission reductions. Given the requirements of state regulatory agencies to minimize the cost of electricity, such analyses will be required before LED can be incorporated into or replace the existing dispatch process.

Cost

LED does not involve capital expenditures and only the differentials in fuel and variable O&M costs need to be considered in estimating the cost of LED. For example, the cost differential
per kilowatt-hour from dispatching a plant burning $41/ton central Appalachia coal with 0.7% sulfur versus a $35/ton northern Appalachia coal with 2.4% sulfur (assuming a 10,000 Btu/kWh heating value and equal O&M costs for both

plants) is $0.0025 ($6 coal price differential divided by 24 million Btu per ton). This equates to a 30-year current 1988 dollar cost per ton of SO_2 removed of $334.

Estimating the cost-effectiveness of LED for an entire utility network is analytically similar to "bubbling" regional emissions. The major analytical difference with LED is that the focus is on altered dispatch to minimize SO_2 rather than to minimize cost. As with fuel switching, the cost of the first ton of SO_2 reduction using LED is small. As the number of altered dispatches increases, the incremental cost of LED can increase significantly.

Two studies have been published on the cost-effectiveness of LED on a regional basis. These studies were conducted by the Northeast States for Coordinated Air Use Management (NESCAUM) (7) and the Wisconsin Public Service Commission (8) as part of EPA's State Acid Rain (STAR) program.

The NESCAUM study examined SO_2 and NO_x control alternatives for the region containing New York, New Jersey, and the six New England states. This analysis, based on a regionalized version of EPA's Advanced Utility Simulation Model (AUSM), concluded that the region's utilities could reduce SO_2 emissions by 35 percent (158,000 tons) and NO_x emissions by 41 percent (51,900 tons) through use of LED at an average current 1988 dollar cost of $4,893/ton.

The Wisconsin study estimated that if 1) all power plants in the state were centrally dispatched and 2) significant conservation efforts are implemented (resulting in relatively low system capacity factors), LED could reduce SO_2 emissions by roughly 15 percent (73,400 tons) at an average current dollar cost of $494/ton of SO_2. At higher system capacity factors, the SO_2 reduction potential of LED is reduced.

The large difference in projected cost from these two studies indicates the regional- and utility-specific differences in LED effectiveness, as well as the significant difference in cost associated with various emission reduction targets (15 percent in Wisconsin versus 35 percent in New England).

2.3 PHYSICAL COAL CLEANING

Description

Run-of-mine coal consists of individual particles containing varying amounts of organic and inorganic matter. The inorganic materials are largely sedimentary rock (e.g., shale, claystone, sandstone, and limestone) and pyritic sulfur. Physical coal cleaning separates the combustible organic material (i.e., coal) from the non-combustible impurities. Although the primary economic impetus for most existing coal cleaning is removal of ash, significant benefits also accrue from the removal of sulfur. Of the total coal consumed by power plants, about 42 percent is cleaned using coal cleaning processes (9).

For most medium- and high-sulfur bituminous coals, pyritic sulfur which can potentially be removed by physical separation methods accounts for 30 to 70 percent of the total sulfur. Most of the remaining sulfur is associated with the organic structure of the coal and is not separable by physical means. The upper limit on sulfur removal depends on the physical liberation of pyrite from the organic structure during coal crushing prior to cleaning. Sulfur contents of some Northern Appalachia coals (e.g., the Freeport and Kittaning seams of Pennsylvania) can be reduced by 50-60 percent using currently available cleaning methods, but removals of 30 percent or less are common for most coals.

Most commercial techniques rely on differences in specific gravity to accomplish separation.(10) For example, clean bituminous coal has a specific gravity (s.g.) of 1.2-1.3; shale, 2.0-2.7; and pyrite, 4.8-5.0. Selective adhesion, magnetism, and surface tension are also used.

In cleaning processes based on specific gravity differences, the raw coal is separated into float and sink fractions by submerging it in a fluid with a specific gravity between that of the coal and the inorganics. In some technologies, such as jigging and tabling, the differences in specific gravity are combined with mechanical separation techniques to enhance the recovery of a clean coal fraction. However, the clean coal float will always contain some impurities, and the rejected sink fraction will contain some coal. There are two reasons for this. First, inherent inefficiencies in separating processes cause a fraction of heavy particles to be misplaced in the float instead of the

sink, and vice versa for some light particles. Second, the organic and inorganic matter cannot be totally separated because they frequently occur in the same particles.

Yields and qualities of clean coal achievable for different specific gravities and particle top sizes can be estimated from laboratory washability data. Figure 2.3-1 shows washability data for two coals. The coal shown on the left is easily cleaned because, as shown on Curve A, the ash content of incremental amounts of coal recovered increases rapidly as the specific gravity of separation exceeds a given level, in this case roughly 1.5 (Point X). As shown on Curve B, the cumulative ash content of the recovered coal at this specific gravity is less than 10 percent versus an original ash content of roughly 30 percent (as plotted on Curve B at 100 percent recovery). The cumulative ash content of the sink material (Curve C) at s.g.=1.5 is roughly 65 percent. The right side of Figure 2.3-1 represents a difficult-to-clean coal for which there is no easily definable point of separation. Commercial cleaning equipment is most efficient when separations are easily made.

Because particles of different sizes behave differently when suspended in a fluid, technologies have been developed to clean each size fraction separately. These technologies are based on three general size classifications:

Fraction Classification	Typical Size Range
Coarse	greater than 3/8 inch
Intermediate	28 mesh (0.6 mm) to 3/8 inch
Fine	0 to 28 mesh (0.6 mm)

Jigs and dense media (DM) vessels are the most common equipment used for coarse coal cleaning. A jig feeds raw coal through a column of pulsating water. These pulsations suspend the lighter coal particles at the top of the jig while allowing the dense ash and pyrite particles to sink. Dense media vessels use a suspension of fine magnetite particles (Fe_3O_4) and water to float all material below a certain s.g. After separation, the refuse and clean coal are screened and rinsed with water to remove the magnetite. Magnetite in the rinse water is reconcentrated using a magnetic separator. Figure 2.3.1 Coarse coal cleaning accounts for one-half to two-thirds of all coal cleaned in the U. S.

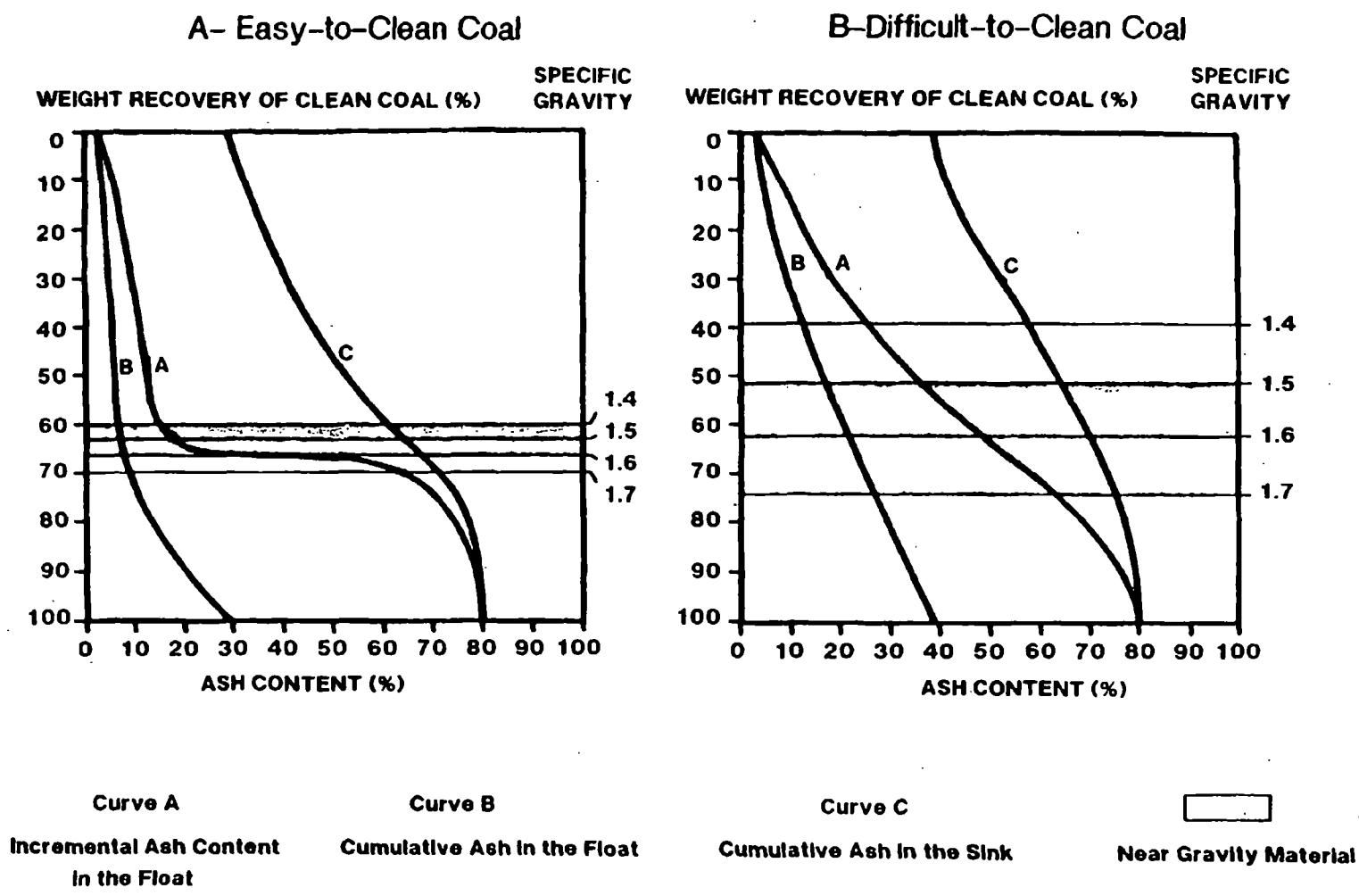


Figure 2.3-1. Coal washability curves. Reference 11.

Intermediate-sized particles are generally cleaned using concentrating tables and dense media cyclones. A concentrating table is an inclined, vibrating surface, equipped with diagonal riffles. Raw coal slurry is fed to an elevated corner of the table. The riffles act as miniature jigs to stratify the raw coal. Lighter clean coal is carried by water over the riffles and off the lower end of the table. Impurities sink between the riffles and are conveyed to the side of the table by the vibration. DM cyclones use centrifugal force to separate clean coal from refuse, much as gravity is used in dense media vessels. Because small particles settle slowly under gravity, the much stronger centrifugal forces in a cyclone permit larger tonnages of fine particles to be handled than in a DM vessel. Intermediate coal cleaning accounts for one-quarter to one-third of U.S. cleaned coal.

Fine coal (less than 28 mesh) is difficult to handle because of problems in separating fine particles from the cleaning fluid. The principal commercial methods for cleaning fine coal are froth flotation and fine coal cyclones, accounting for 5-10 percent of total production. Froth flotation takes advantage of the hydrophobic (i.e., water repelling) tendencies of most coals, versus ash which is hydrophilic (i.e., water adhering). When air is bubbled up through a flotation cell, the small coal particles attach to the air bubbles and rise to the surface.

Coal cleaning plants are frequently classified based on the following definitions (12):

<u>Level 1</u> - Crushing and particle sizing only. Little or no cleaning takes place.

<u>Level 2</u> - Coarse (+3/8 inch) coal cleaning only.

<u>Level 3</u> - Coarse (+3/8 inch) and intermediate (3/8 inch to 28 mesh) coal cleaning. Material less than 28 mesh is not cleaned and is either incorporated with the product or discarded depending on its quality.

<u>Level 4</u> - All size fractions are cleaned, but only a single product consisting of a blend of cleaned fractions is sold.

<u>Level 5</u> - This level involves the most rigorous coal cleaning, and produces two or more usable coal products of different ash and sulfur contents from one raw coal.

This numbering is generally compatible with existing coal cleaning literature. A schematic flow diagram for a Level 4 cleaning plant is shown in Figure 2.3-2. The quantity of material processed as coarse, intermediate, and fine coal will depend upon the characteristics of the coal and the desired product quality. The distribution of Level 2 through 4 cleaning plants by region and level is shown on Table 2.3-1. Level 1 plants are omitted because of their minimal effect on ash and sulfur levels of coal. The only Level 5 cleaning plant in the U. S. is the Homer City cleaning plant in Pennsylvania.

Applicability and Performance

Most coal cleaning plants in the U. S. were installed by coal companies to meet specific ash and sulfur specifications. Approximately two-thirds of the coal in the eastern U. S. is physically cleaned (12). For the nation as a whole in 1980, 241 million out of 687 million tons of steam coal were cleaned, or 35 percent.

Table 2.3-2 provides average sulfur dioxide emission rates for major seams in the two major high-sulfur coal regions in the U. S. Although not shown on the table, current coal cleaning plants (designed primarily to remove mineral matter) generally remove 20-30 percent of the sulfur in Illinois Basin coal and from 10-40 percent of the sulfur from Northern Appalachia coal. This level of sulfur removal is accomplished while recovering most of the heat content of the raw coal (>95 percent), thus minimizing the dollar value of coal lost during cleaning. Estimated emissions from state-of-the-art removal are assumed to be representative of cleaning in a Level 4 preparation plant. In general, state-of-the-art coal cleaning is most effective for bituminous coals from North Appalachia (especially those seams in the Allegheny Formation such as the Freeport and Kittaning seams) where sulfur reductions near 50 percent are possible and the Illinois Basin which contain significant amounts of macroscopic pyrite. Additional sulfur removal is possible, but a substantial portion of the heating value is lost in the reject material.

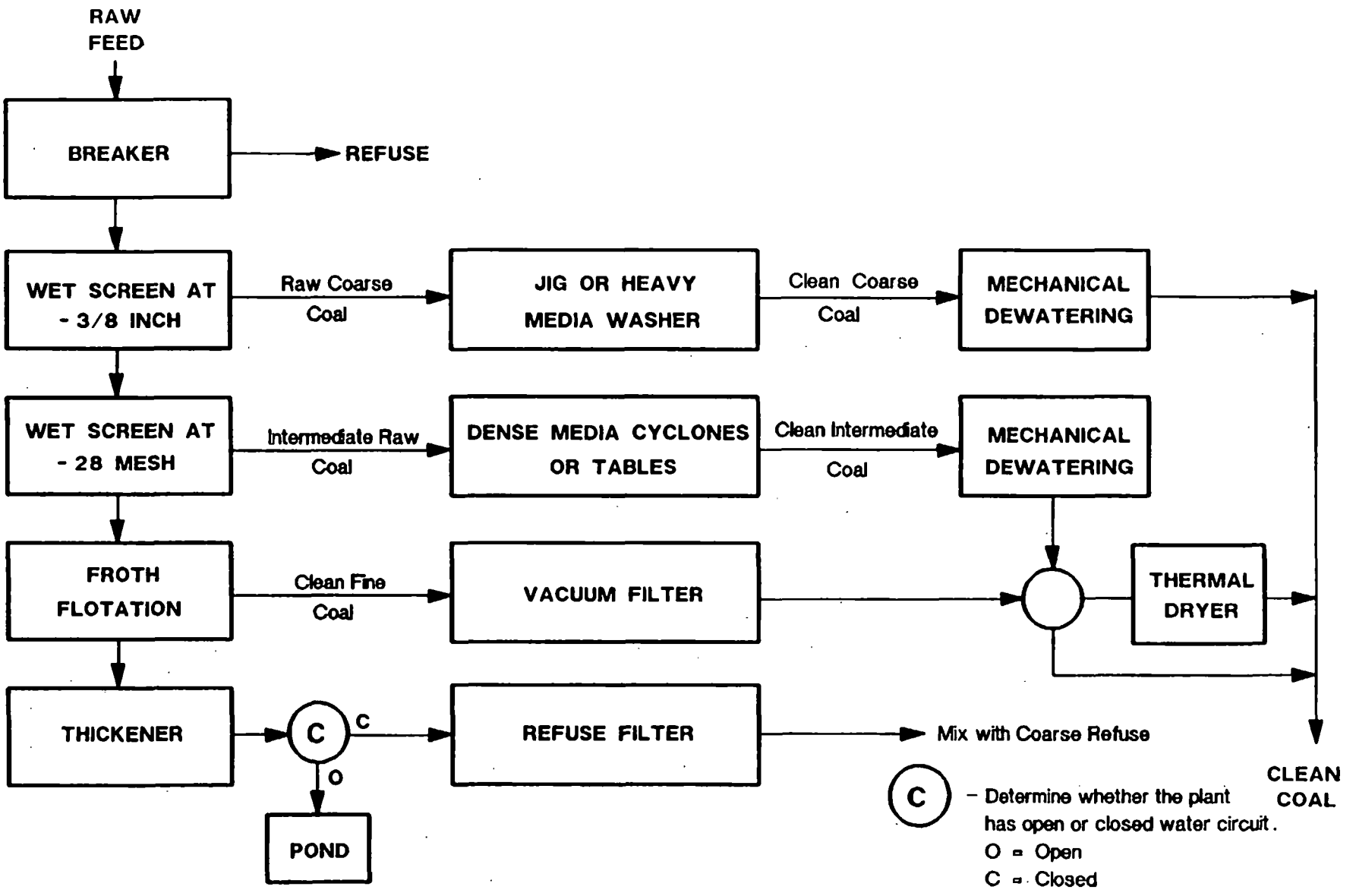


Figure 2.3-2. Level 4 Coal Preparation Plant
Source: Reference 1.

TABLE 2.3-1. BREAKDOWN OF COAL CLEANING PLANTS BY REGION AND LEVEL, 1978

Region	Total Plants[a]	Level 2	Level 3	Level 4
Northern Appalachia	128	35	52	41
Central Appalachia	239	47	71	121
Southern Appalachia	25	6	10	9
Midwest	53	23	22	8
Other	19	--	--	--
Total	464[a]	111[b]	155[b]	179[b]
Percent of Total		25	35	40

Source: 1979 Keystone Coal Industry Manual. "Directory of Mechanical Coal Cleaning Plants."
[a]Does not include 12 pneumatic preparation facilities.
[b]Only includes plants in the four major regions.

TABLE 2.3-2. POTENTIAL REDUCTIONS IN SULFUR DIOXIDE EMISSION RATES FROM PHYSICAL COAL CLEANING

	Raw Coal	State-of-the-Art[1]	Maximum Achievable[2]
Northern Appalachia			
Pittsburgh	5.7	4.2	3.3
Sewickley	6.2	4.8	3.8
Freeport	3.9	2.2	1.3
Kittaning	4.8	2.6	1.7
Illinois Basin			
Illinois 6/Kentucky 11	6.5	4.1	3.6
Illinois 5/Kentucky 9	6.5	4.4	3.9

[1]Based on 90% Btu recovery with 1 1/2" x 0 feed.
[2]Based on 14 mesh x 0 feed at s.g. = 1.4.
Source: Reference 14.

Cost

Physical coal cleaning has many economic benefits besides sulfur removal, although they can be difficult to quantify (15). These include reduction in coal transportation costs; reduced coal handling, pulverization, and storage requirements; generally improved boiler performance; smaller ash handling and waste disposal volumes; and reduced costs for post-combustion flue gas desulfurization (FGD) to meet state and federal emission limits. For utilities having trouble meeting particulate emission limits, this lower ash content can bring them into compliance and avoid expensive upgrading of ESPs. While these savings can be significant, they are hard to document quantitatively and vary from plant to plant. "Typical" cost benefits from coal cleaning have been estimated at $2.82/ton, but can range from $0.54 to over $9/ton (10). Cost savings associated with use of coal cleaning to reduce FGD requirements are discussed in Chapter 5. Other benefits, such as improved boiler performance and reduced coal handling costs, are very boiler- and site-specific and are not included in the costs discussed below.

The three major components of coal cleaning costs are the initial capital equipment, operation and maintenance, and the lost heating value of discarded coal. The most important variables in determining these costs are the characteristics of the raw coal and the desired level of cleaning (which in turn determines the selection of cleaning technology and the quantity of Btu rejected).

Of the total capital costs, roughly one-third are associated with the cleaning equipment per se. These costs (including equipment delivery and installation) range from around $2,200 per ton of raw coal per hour for a jig or DM vessel processing coarse coal to nearly $22,000 per ton per hour for a froth flotation cell for cleaning fine coal. Another third is for structural steel, electrical, engineering and construction fees, and other costs that are generally proportional to the size and complexity of the project design. The remaining third covers items such as refuse ponds, site preparation, conveyors, silos, and load out facilities for which costs are relatively independent of the cleaning level. On a total plant basis, total capital cost estimates range from around $11,000 per hourly ton of input capacity for a Level 2 plant to

near $43,000 per hourly ton for a Level 4 plant incorporating froth flotation for fine coal cleaning.

Operating costs increase substantially with higher levels of cleaning, as well as with coal washability. The operating cost for a coal circuit designed for cleaning intermediate coal (3/8" x 28 mesh) may be double on a per ton basis the cost of a circuit for coarse coal (+ 3/8"). More complex plants require more personnel and greater training in addition to power, supplies and maintenance. Magnetite costs can be very important in dense media plants. The heating value of the coal that is thrown away as refuse can be the highest cost of coal cleaning, and is both a function of the level of cleaning and the coal washability.

As a result, the cost of sulfur removal using coal cleaning varies widely depending on raw coal quality and user specifications. Table 2.3-3 shows ranges of capital costs, O&M cost, and Btu loss for several Northern Appalachian and Illinois Basin coals on a life-cycle basis (i.e., per ton of cleaned coal). Busbar costs and cost effectiveness are also shown. These values assume Level 2 and Level 4 cleaning of a raw coal with a sulfur content of 3.5 percent. Btu recoveries were 90-96.5 percent, weight recoveries 73-90 percent, sulfur reductions 17-62 percent, and raw coal cost $1.30/million Btu. Although Level 2 cleaning has lower capital and annualized costs than Level 4, costs per ton of SO_2 removed Level 4 are frequently lower because of higher sulfur removal rates. Cost per ton of SO_2 removed for lower sulfur coals generally increase due to the decreased amount of removable pyrite in the raw coal.

TABLE 2.3-3. ECONOMICS OF CONVENTIONAL PHYSICAL COAL CLEANING

$/Ton of Clean Coal	Current 1989 Dollars
Capital	0.54 - 2.43
O & M	3.79 - 7.84
Btu Loss	1.50 - 6.50
Total	5.83 - 16.77
Busbar Cost (mills/kwh)	2.7 - 6.2
$/ton of SO_2	325 - 595

2.4 NO_x COMBUSTION CONTROL TECHNOLOGY

The production of NO_x during combustion of coal (or other fuel) is dependent on boiler design, size, and operating parameters as well as fuel composition. By modifying flame geometry and the rate of air/fuel mixing, thermal oxidation of atmospheric and fuel nitrogen can be reduced. Specifically, "staging" of combustion to provide a fuel-rich (i.e., reducing) zone during the initial, high-temperature stage of combustion will favor formation of molecular N_2 instead of NO_x. Additional (i.e., secondary) air is added after the initial high-temperature reaction to complete combustion of unburned fuel. Secondary combustion air is added at a point where much of the fuel-bound nitrogen has been reduced to N_2 and flame temperatures are too low for thermal NO_x formation.

Wall-fired wet-bottom and cyclone units are large generators of NO_x due to their intense combustion conditions. Tangentially-fired units with their longer flames and cooler combustion temperature produce less NO_x. Because of the wide variations in boiler characteristics and NO_x levels, the applicability of combustion modification technologies for reducing NO_x is boiler specific. This section discusses the two major combustion modification technologies for NO_x control: low NO_x burners and overfire air.

Description

Low NO_x Burners--

Low NO_x burners are designed to reduce NO_x production (typically from wall fired boilers) by controlling air/fuel mixing. This changes combustion reactions in several ways. First, substoichiometric amounts of air are used during initial combustion. This results in fuel nitrogen forming molecular nitrogen (N_2) rather than NO_x. Second, due to controlled addition and mixing of secondary air, local zones of excess air that promote NO_x formation are minimized. Third, increasing combustion times and allowing for more heat transfer from the flame results in a cooler flame and reduced production of thermal NO_x.

The controlled air/fuel mixing is achieved by using separate air registers within the burner. Part of the air enters with the fuel and another part enters through an annulus surrounding the central fuel/air stream. The remaining air is injected through a second, outer annulus. Separating the air and fuel delays fuel/air mixing, thus "staging" the combustion process and creating the low NO_x conditions described above.

The same effect can be achieved on tangentially-fired boilers, but a different, somewhat more expensive burner design is required. However, the diagonal firing pattern and larger furnace dimensions of most tangentially-fired boilers result in lower flame temperatures and less NO_x formation than wall-fired boilers, thus the potential NO_x reduction by low NO_x burners is reduced. Because of this, overfire air is the predominant commercial NO_x control technique for tangentially-fired boilers. Therefore, the subsequent low NO_x burner discussion will focus on wall-fired units only.

Overfire Air--

Another fundamental method of controlling flame stoichiometry, in addition to (or in place of) the localized combustion staging of low NO_x burners, is combustion staging within the furnace volume using overfire air (OFA) ports With this approach, 15 to 20 percent of the required combustion air is diverted from active burners to OFA ports located above the top row of burners. The active burners operate fuel-rich, providing a reduction in both thermal and fuel NO_x generation. The unburned fuel escaping the fuel-rich flame zone burns higher in the furnace where the diverted combustion air is mixed. An overfire air system includes the OFA ports (involving penetrations in the furnace wall), additional duct work, potentially a separate fan, and dampers/air flow controls.

Applicability

The applicability and effectiveness of low NO_x burners and OFA, especially for retrofit cases, are dependent on a number of site-specific parameters, some of which are listed in Table 2.4-1. For example, NO_x emissions increase with increasing coal nitrogen and oxygen contents.

TABLE 2.4-1. SITE-SPECIFIC PARAMETERS AFFECTING NO_x EMISSIONS FROM COMBUSTION MODIFICATION CONTROLS

Furnace Characteristics	- Manufacturer - Size (width, depth, height) - Heat Input (Btu/hour) - Heat Release Rate (Btu/hour-ft^2) - Firing Type (tangential, wall, cyclone) - Bottom Ash Removal (slag, dry) - Number, Spacing, and Size of Burners - Nose/Top Burner Row Distance - Gas Velocity - Furnace Exit Temperature/Attemperation Margins - Windbox Size - Excess Fan Capacity
Coal Characteristics	- Heating Value - Composition (N, O, H_2O, S, Ash, wt. %) - Ash Fusion Temperature - Ash Slagging/Fouling Indices - Volatility - Variability

Furnace design, size, and dimensions also affect NO_x emissions, with low heat release rate furnaces generally exhibiting lower NO_x emissions. Specific reviews of the applicability of low NO_x burners and overfire air are presented below.

Low NO_x Burners--

Since the 1971 NSPS, low NO_x burners have been installed on more than 40,000 MW of boiler capacity. Most of these boilers have been new wall-fired units with circular burners. An additional 30,000 MW of pre-NSPS wall-fired boilers are believed technically capable of being retrofit with low NO_x burners. This 30,000 MW represents approximately 20 percent of nation-wide NO_x emissions from pre-NSPS utility boilers (16).

Retrofit of low NO_x burners to existing boilers is currently being examined in several full-scale utility boilers. Low NO_x burners are generally

larger than conventional burners and require more precise control of fuel/air distribution. Their performance depends partially on increasing the size of the combustion zone to accommodate longer flames while avoiding interaction with flames from other burners. Because of this, low NO_x burners are expected to be less effective when retrofit on relatively small furnaces.

Low NO_x burners are not currently available for dry-bottom cell-fired and roof-fired boilers, cyclone boilers, and wet-bottom boilers. Cell burners consist of two or three burners in a single cluster and were designed to produce an intense, high-temperature flame. Babcock & Wilcox is currently conducting a low NO_x burner development program for cell burner applications for the Electric Power Research Institute (17). Because of difficulties in applying low NO_x burners to cyclone boilers, primary emphasis for reducing NO_x emissions from these units is on reburning technology (see Section 3.7).

In order to retrofit low NO_x burners, the existing burners must be replaced. In some cases, some of the waterwall tubes may have to be bent in order to install the larger low NO_x burners. Also, low NO_x burners will generally produce a somewhat longer flame; thus, flame impingement on furnace walls and superheater tubes can be a problem for boilers with high heat release rates (typical of boilers built in the 1960's) (1). If flame impingement is a problem, potential solutions include adjusting burner tilt, adjusting coal/primary air versus secondary air velocities, biased firing of burner rows, and relocating some superheater tubes. Boilers with very small furnaces may have to be derated in order to prevent flame impingement at full load.

Other factors to consider in retrofitting low NO_x burners include fuel characteristics, furnace exit temperature, and fan capacity. Fuel-rich operating conditions in the lower furnace region associated with low NO_x burners can increase the slagging tendency of the coal. The generally longer flames of low NO_x burners will tend to increase furnace exit and superheat/reheat tube temperatures. Some low NO_x burners operate with a higher pressure drop or may require slightly higher excess air levels at full load to ensure good carbon burnout, thus increasing fan requirements.

Another consideration in retrofitting low NO_x burners is modifying the windbox. Modifications may include the addition of dampers and baffles for better control of combustion air flow to burner rows and combustion air distribution to burners within a row. Also, the windbox must be large enough to accommodate the low NO_x burners. If the existing windbox requires significant modifications to structural components, major re-piping, and/or windbox replacement, retrofitting low NO_x burners may not be feasible.

Overfire Air (OFA)--

Overfire air has been installed on approximately 30,000 MW of new tangentially-fired (T-fired) boiler capacity built since the 1971 NSPS. In addition an estimated 38,000 MW of pre-NSPS T-fired boilers are believed technically suitable for OFA retrofit; these units account for an estimated 15 percent of nation-wide pre-NSPS NO_x emissions (16). OFA can also be used with wall-fired boilers, but is less effective at controlling NO_x emissions than low-NO_x burners. As a result, use of OFA is primarily applied to T-fired boilers.

With retrofit applications, physical obstructions outside of the boiler may restrict extension of the windbox and make installations of ductwork needed to supply air to the OFA ports difficult. There must also be adequate distance between the top burner row, the OFA ports, and the furnace exit for good carbon burnout and maximum NO_x reduction performance. Extra fan capacity may also be necessary.

As in the case of low NO_x burners, increased furnace exit gas temperatures and flame impingement are concerns. To avoid these problems, the distribution of combustion air between the OFA ports and the burner registers must be carefully controlled. The injection velocity required for good OFA mixing with the combustion gases and the additional pressure drop associated with the OFA ductwork may require installation of a separate OFA fan.

Performance

In new boilers, low NO_x burners can reduce uncontrolled NO_x emissions by 40 to 60 percent. Performance in retrofit applications will depend on boiler-specific considerations, but is generally expected to be somewhat poorer than in a new boiler.

Overfire air is generally capable of a 15-30 percent NO_x reduction, with higher reductions possible in a few cases. OFA can achieve the 1971 NSPS NO_x emission limit for wall- and tangentially-fired utility boilers and 1978 NSPS for NO_x emission limit for tangentially-fired units.

Cost

Low NO_x burners and overfire air are relatively inexpensive control methods compared to post-combustion NO_x control techniques. NO_x reduction performance is the major uncertainty for a retrofit application.

The following table presents the IAPCS estimated capital costs, annualized costs and cost per ton of pollutant removed for OFA and LNB. The lower capital and annualized costs are for a large unit with a high capacity factor and the higher capital and annualized costs are for a small unit with a low capacity factor. Figures 2.4-1 through 2.4.3 show these costs as a function of boiler size for overfire air and low NO_x burners.

Basis for Range	Range
Overfire Air -	
Capital Cost ($/kW)	1.3 to 6.2
Annualized Cost (mills/kWh)	0.1 to 0.5
Cost Per ton of NO_x Removed ($/ton)	49.7 to 1,621.2
Low NO_x Burners -	
Capital Cost ($/kW)	5.5 to 25.5
Annualized Cost (mills/kWh)	0.2 to 2.1
Cost Per ton of NO_x Removed ($/ton)	79.9 to 2,388.8

Appendix B contains tables for estimating the cost of combustion controls as a function of boiler size, capacity factor and NO_x reduction.

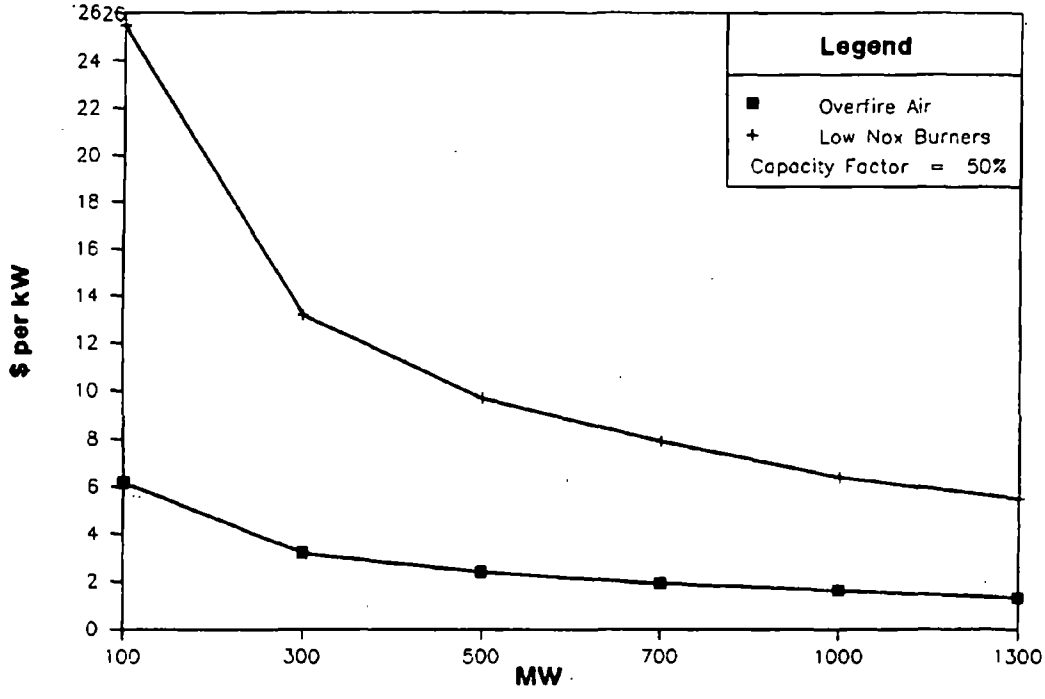


Figure 2.4-1. NOx Combustion Controls - Capital Cost, Current 1988 Dollars

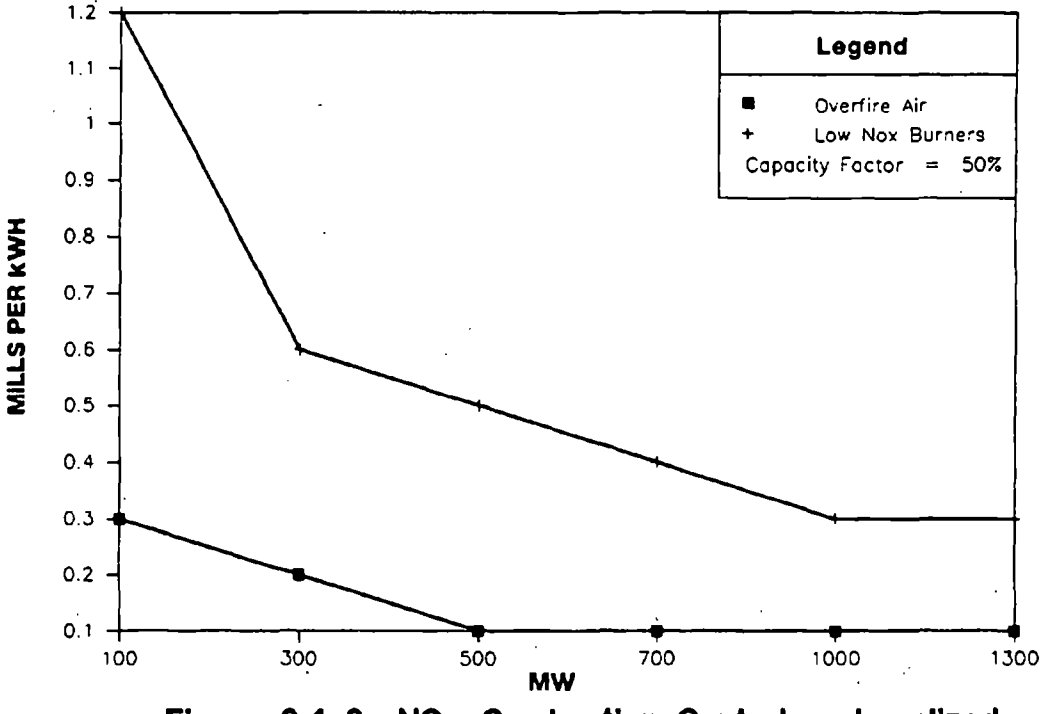


Figure 2.4-2. NOx Combustion Controls - Levelized Annual Cost, Current 1988 Dollars

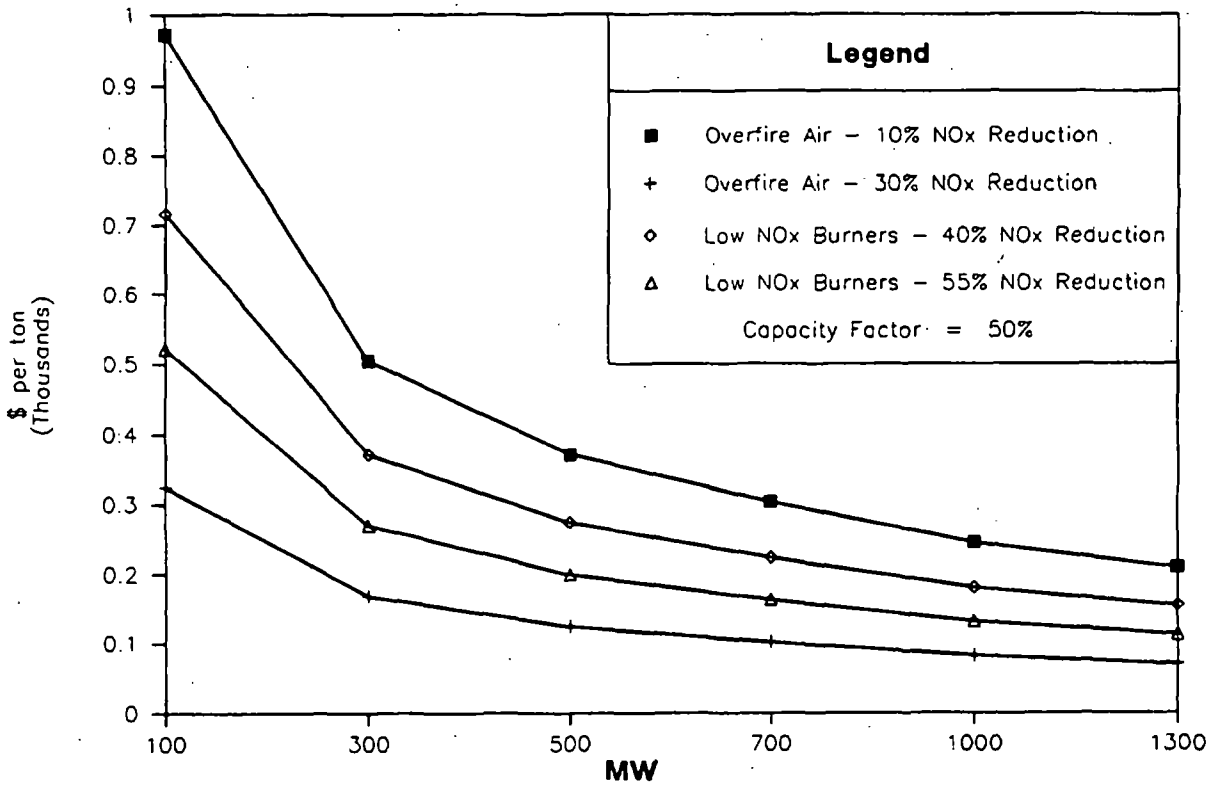


Figure 2.4-3. NOx Combustion Controls - Cost Per Ton of NOx Removed, Current 1988 Dollars

2.5 THROWAWAY WET FGD

Throwaway wet FGD technologies are the most widely used method of post-combustion SO_2 control in the U. S. This section discusses the two major processes used in the U. S.: lime/limestone scrubbing and dual alkali The use of additives to enhance lime/limestone performance is also discussed.

Additional processes are in use in Japan and Europe and have been pilot-scale tested in the U. S. These processes include Chiyoda Thoroughbred 121 and Dowa from Japan and Saarberg-Holter from West Germany. For information on these processes refer to Reference 18.

Description

Lime/Limestone Scrubbing--

Lime/limestone scrubbing is the most common post-combustion SO_2 control technique currently applied to utility boilers. A process flow diagram is shown in Figure 2.5-1. There are four fundamental steps to the process: preparation of the lime or limestone slurry, contacting the flue gas and slurry, reaction of the lime or limestone with SO_2, and removal of solid waste.

Reagent preparation involves crushing limestone or slaking lime to produce a slurry of fine hydrated lime or limestone particles. Although many existing systems use lime, most future systems are expected to use limestone due to its lower cost. Flyash from alkaline western coals has also been used in conjunction with lime and limestone. The slurry is fed to the scrubbing loop which consists of an absorber (usually a spray tower) and reaction tank; the scrubber system for most utility plants consists of multiple absorber/reaction tank modules. SO_2 is removed from the flue gas (which has already been cleaned of flyash by an ESP or fabric filter) in the absorber while the reaction tank provides time for limestone dissolution and reaction with the dissolved SO_2. The product of this reaction is a mixture of crystalline calcium sulfate and calcium sulfite. Small droplets of slurry which become entrained in the flue gas are captured by a mist eliminator. Reacted solids are removed and dewatered with the clear liquor being returned to the process. The solid waste is either ponded or landfilled.

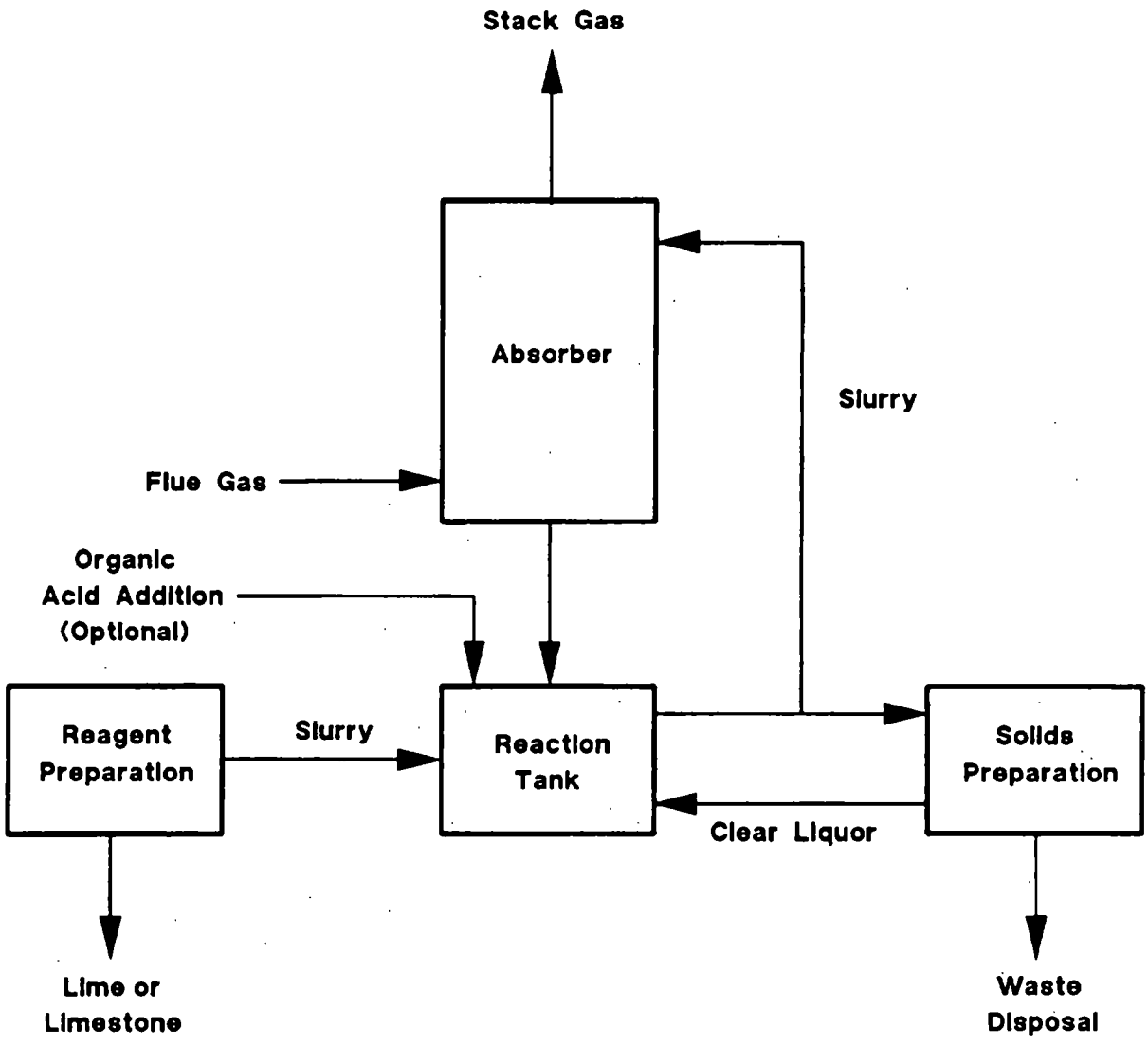


Figure 2.5-1. Lime / limestone FGD system flow diagram.

During the initial applications of lime/limestone FGD systems a variety of chemistry problems were encountered, often with the result being rock-like gypsum (calcium sulfate) scale formation on equipment. These problems have been overcome through experience and better understanding of the complex chemical interactions. Today, most lime/limestone FGD problems involve adjusting the chemistry to optimize raw material utilization and SO_2 removal. However, materials selection continues to be a problem. To reduce impacts on boiler operation, scrubber systems are generally designed with an extra absorber module which can be operated when any of the other modules are down for maintenance or repair.

One technique for improving the performance of limestone FGD systems is the use of organic acid additives, such as adipic and other dibasic acids. This improves SO_2 removal and enhances other aspects of the process, but does not greatly increase operating costs. As a result of EPA Air and Energy Engineering Research Laboratory (EPA-AEERL) sponsorship of development efforts, this technique has been successfully used in retrofit applications at several utility installations. Use of organic acids allows existing FGD systems to improve system reliability and SO_2 removal without affecting other aspects of the process or requiring significant additional SO_2 control equipment.

Dual Alkali Scrubbing--

Dual alkali scrubbing was developed to avoid the problems of erosion, scaling, and solids deposition found with lime/limestone FGD. As the name implies, two alkalis are used: a clear solution (generally sodium sulfite) in the absorber followed by addition of lime in the reaction tank to regenerate the spent solution for recycle to the absorber. The resulting calcium sulfite/sulfate sludge is then dewatered and landfilled. Makeup sodium (typically soda ash) is added to the regenerated solution to replace residual sodium lost in the filter cake.

Several variations of the dual alkali process have been investigated (e.g., using ammonia and potassium as the first-stage alkali). However, the use of sodium is the most advanced and is the only process in commercial use in the U. S. A simplified diagram of the sodium/lime process is shown in Figure 2.5-2.

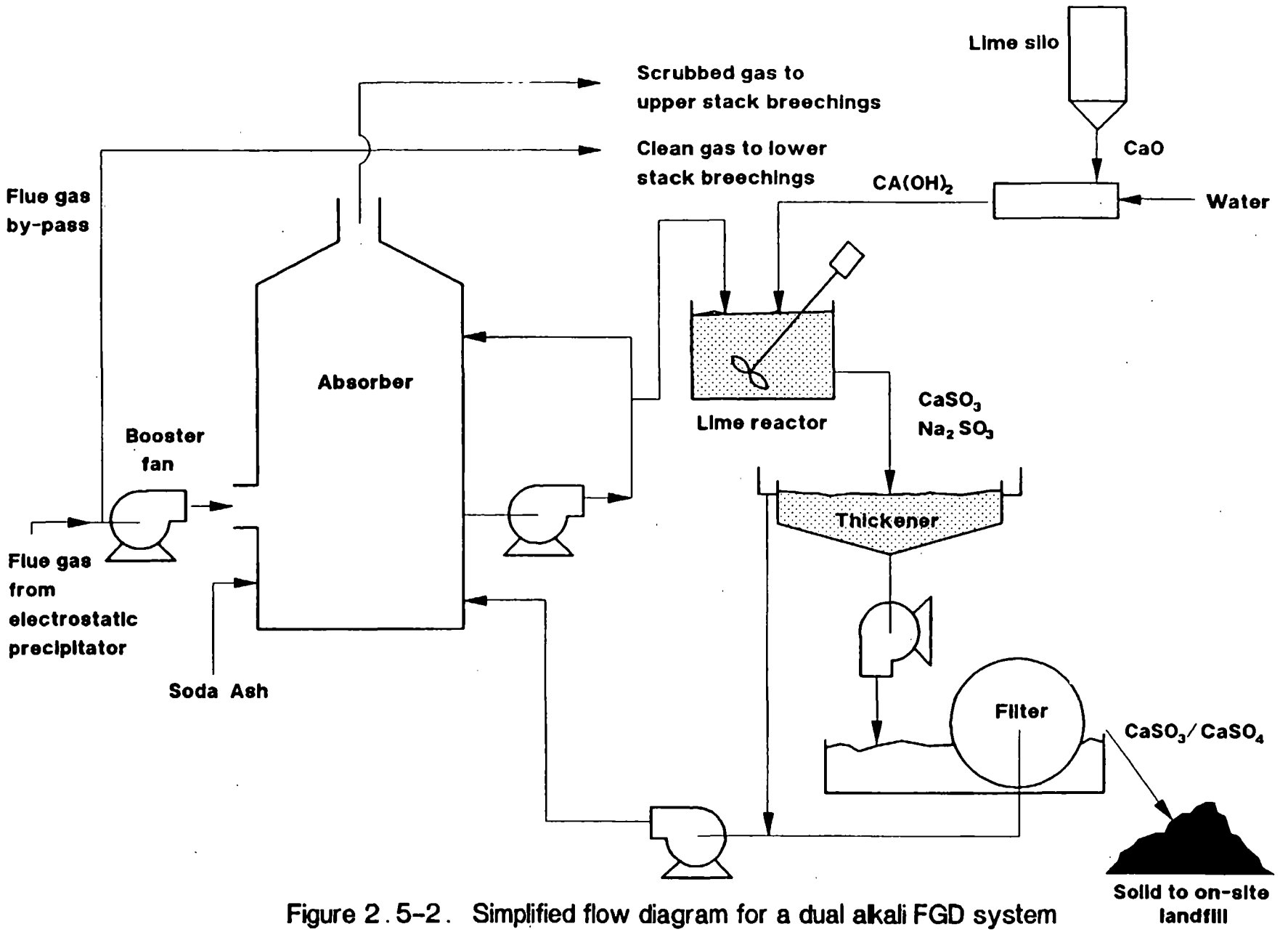


Figure 2.5-2. Simplified flow diagram for a dual alkali FGD system
Source: Reference. 1.

Sodium-based, dual alkali systems are more suitable for high-sulfur than for low-sulfur (less than 1.5 percent) coal applications. With low-sulfur coals, a greater percentage of the sulfite is oxidized to sulfate. This lowers the available alkalinity in the scrubber and, thus, lowers SO_2 removal capacity.

Additive-Enhanced FGD--

Several additives have been added to lime/limestone FGD systems to improve SO_2 removal, sorbent utilization, and/or system operation. These include: adipic acid, mixtures of dibasic organic acids (DBA), magnesium, and thiosulfate. The decision to use a specific additive depends upon system chemistry and performance. In general, except for the additive feed system, there are no major hardware differences compared to a conventional lime/limestone system.

Organic acids such as adipic or DBA are currently the most widely used additives. These acids buffer the gas/liquid contacting interface at a pH of 3 to 5, thus improving the driving force for SO_2 removal. For example, addition of only 2000 ppm of adipic acid can more than double the normal dissolved alkalinity of the absorbing limestone slurry, and thus can compensate for inadequate gas/liquid contact in the absorber. Organic acids buffering also allows operation at lower pH than limestone-only systems, without loss of SO_2 removal capability. This results in higher limestone utilization which can, in some cases, reduce mist eliminator scaling problems.

Magnesium-promoted FGD involves addition of magnesium-containing (i.e. dolomitic) lime or magnesium sulfate ($MgSO_4$) to either a lime or limestone system. Magnesium salts are more soluble than the corresponding calcium salts, providing higher liquid-phase alkalinity and increased absorption of SO_2.

Sodium thiosulfate ($Na_2S_2O_3$) addition has been used to reduce scaling in both lime and limestone systems and to improve sludge dewatering. Sodium thiosulfate is also a very active inhibitor of sulfite oxidation; as little as 100 ppm S_2O_3 can virtually halt oxidation of calcium sulfite to calcium sulfate. However, consumption of this additive can be high due to degradation

of thiosulfate to other species, such as trithionate, which do not inhibit oxidation. This approach was commercially demonstrated in 1985 at Seminole Electric Cooperative's plant near Palatka, Florida.

Applicability

Lime, limestone, sodium carbonate, dual alkali, and additive-enhanced FGD are commercially available and have been widely applied to utility boilers. The total capacity of operating and contracted throwaway FGD systems as of December 1985 exceeds 68,000 MW as shown in Table 2.5-1. There is currently an estimated 4,000 MW of scrubber capacity using additives (19). From a process viewpoint, lime/limestone FGD is applicable to all boilers. Dual alkali systems are best suited to boilers burning high-sulfur and high-chlorine coals.

Depending upon the arrangement of the boiler and associated equipment, a scrubber retrofit may be simple or difficult. Difficult retrofits often require either long runs or complex arrangements of ductwork, which can be very expensive and time consuming to install. In addition to the absorber and reaction tank in which SO_2 removal actually occurs, additional space is also required for sorbent receiving and storage, slurry preparation, dewatering of reacted sorbent, and solid waste disposal. Replacement or relining of the stack may also be required to eliminate corrosion problems caused by the wet flue gas. Modification of the boiler may also be required to balance and control pressure changes caused by retrofitting a scrubber and additional fans. The high capital cost of the equipment combined with retrofit difficulties makes it very expensive to apply lime/limestone FGD on plants operating at low capacity factors or with little remaining plant life.

In the U. S., additive enhancement is currently used solely as a retrofit method for upgrading the performance of existing FGD systems. The minimal changes in system design required for addition of additives make retrofit relatively easy. Future systems may be designed with additives to reduce capital and operating costs and to boost FGD removal efficencies well above 90%.

TABLE 2.5-1. TOTAL OPERATING AND PLANNED THROWAWAY FGD CAPACITY BY U. S. ELECTRIC UTILITIES (as of December 1985)

	Units	Capacity Factor
Limestone		
In operation	60	26,008
Under construction	9	17,113
Contracts awarded	8	1,505
	77	36,977
Lime		
In operation	39	17,113
Under construction	0	0
Contracts awarded	2	2,036
	41	19,149
Sodium Carbonate		
In operation	6	1,505
Under construction	1	550
Contracts awarded	2	1,100
	9	3,155
Dual Alkali		
In operation	5	1,963
Under construction	1	265
Contracts awarded	0	0
	6	2,228
Total	150	68,072

Source: Reference 19.

Performance

Performance of FGD systems is measured with respect to both SO_2 removal efficiency and system availability. In general, both lime/limestone and dual alkali are capable of up to 95 percent SO_2 removal with careful design and operation. Removals of 90 percent are more common. As discussed earlier, additives such as adipic acid frequently double the dissolved alkalinity of limestone slurries. This improvement in alkalinity can reduce SO_2 emissions from existing scrubbers by a factor of two or more while at the same time increasing unit availability.

In general, lime/limestone scrubbers tend to be more chemically complex in plants using high-sulfur coals versus low-sulfur coals, and thus availability is lower. Addition of spare absorber modules can improve overall system availability by providing backup units and allowing more time for maintenance and repair of each module. However, new scrubber design and the use of additives reduce the need for module sparing in the future.

Cost

The range of estimated costs using the IAPCS cost model for both new and retrofit lime/limestone FGD systems applied to coal-fired utility boilers are presented below. Depending on local sorbent availability, waste disposal options, and other site-specific factors, these costs may vary somewhat, but will generally be within the range shown. The lower capital and annualized costs are for large units with high capacity factors, low retrofit factors and low sulfur coal content, while the higher capital and annualized costs are for small units with low capacity factors, high retrofit factors and high sulfur coal content. Appendix C contains tables of costs as a function of boiler size, capacity factor, coal sulfur and retrofit difficulty.

Basis for Range	Range
Capital Cost ($/kW)	125 to 628.5
Annualized Cost (mills/kWh)	11 to 88.6
Cost Per ton of SO_2 Removed ($/ton)	508.3 to 10,618.6

These costs represent conventional design practices for new installations relative to the size of absorbers and installation of backup systems to assure

compliance with New Source Performance Standard (NSPS) reliability requirements. If these design practices are altered to comply with less stringent reliability requirements, larger--but fewer--absorbers might be installed. Analysis of cost savings at several plants using EPA's IAPCS computerized FGD costing model found this altered design philosophy could reduce capital cost by 16-32 percent and total annual costs by 9-12 percent relative to conventional designs. Under the Department of Energy Clean Coal program, the wet limestone Mitsubishi FGD process was selected for commerical demonstration. The program will demonstrate the technical and economic feasibility of an advanced, single, 600 MW absorber system to obtain 90-95% SO_2 removal on high sulfur (>3 % S) Indiana/Illinois coals.

The costs and benefits from additive enhancement depends in large part on the performance of the existing FGD system. For a limestone FGD unit operating at 90 percent SO_2 removal, but at high liquid-to-gas absorber ratios, use of organic acids can significantly reduce operating (i.e., slurry pumping) costs while maintaining the same SO_2 removal; EPA's IAPCS model estimates a total cost savings of roughly six percent at these conditions. For a unit needing to increase its SO_2 removal efficiency, the economic advantages of adding organic acids may be significantly higher. In general, the capital costs of converting an existing limestone FGD system to use these additives are minimal and quickly pay for themselves if improved system reliability is achieved.

Engineering estimates of capital and fixed O&M costs for dual alkali systems are generally lower than for lime/limestone systems (1). However, total costs (including variable O&M) are similar due to the high cost of lime. Enhancement of the dual alkali process to allow use of limestone rather than lime for regenerating the scrubber solution could significantly reduce the cost of the dual alkali process. Detailed calculations of dual alkali costs are not included in this report.

The costs for a specific FGD system are a function of many variables including unit size, coal sulfur level, percent SO_2 removal, and load factor. As previously mentioned, the remaining lifespan of a plant is also important. An operator does not want to make large capital investments in a plant that has a short remaining lifetime. Also, smaller plants (less than 100 MW) are much more expensive per kW to retrofit with scrubbers than larger plants. Figure 2.5-3 through 2.5-5 present costs as a function of plant size and coal sulfur content.

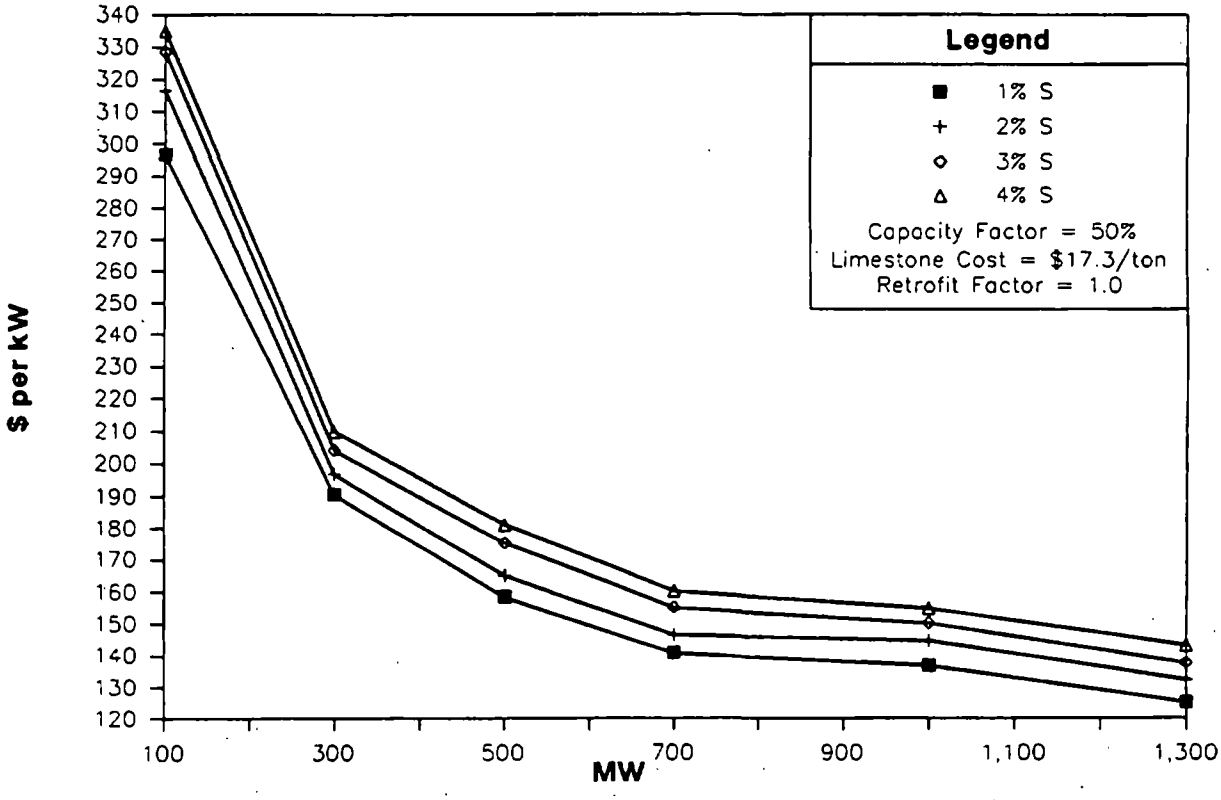


Figure 2.5-3. L/LS FGD - Capital Cost, Current 1988 Dollars

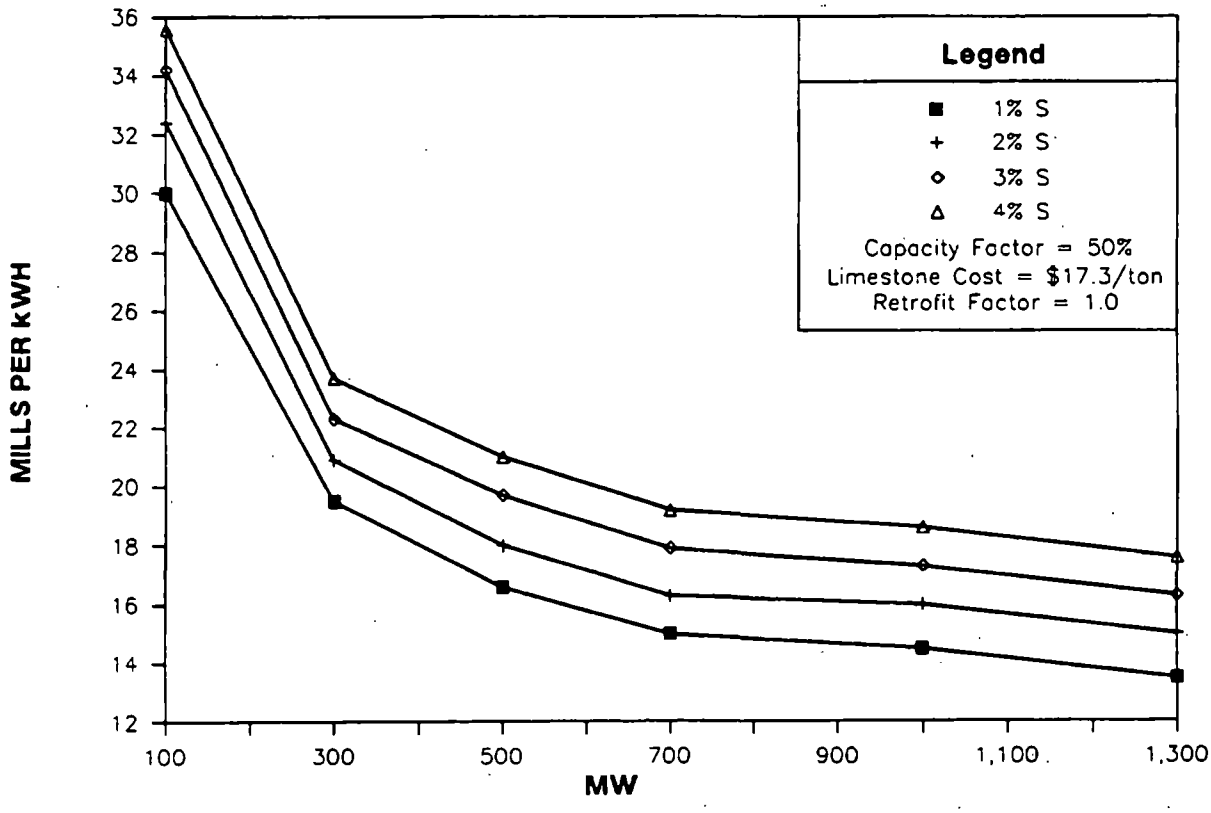


Figure 2.5-4. L/LS FGD - Levelized Annual Cost, Current 1988 Dollars

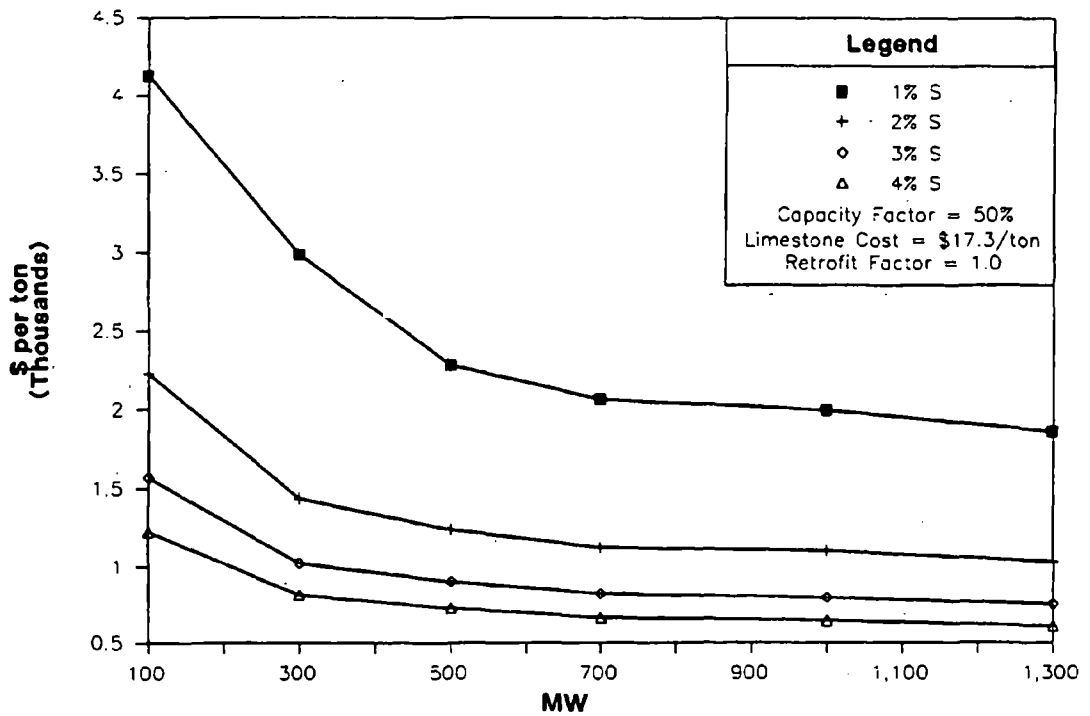


Figure 2.5-5. L/LS FGD - Cost Per Ton of SO2 Removed, Current 1988 Dollars

2.6 BY-PRODUCT RECOVERY FGD TECHNOLOGIES

Description

Recovery of sulfur from SO_2 as a useful product rather than discarding it is beneficial with respect to resource conservation, and also because of the potential improvement in process economics. Numerous processes have been proposed in an effort to achieve these goals. These processes vary widely in sorbent used, product made, and stage of development. Currently, three types of processes have attained commercial status.

- Gypsum recovery - Conventional FGD system with forced oxidation and additional process features to produce salable gypsum.
- Wellman-Lord - Reaction of sodium sulfite with SO_2 to form sodium bisulfite followed by steam stripping to recover sulfuric acid or elemental sulfur.
- Magnesia slurry - Reaction of $MgO/Mg(OH)_2$ with SO_2 to form magnesia sulfite which is then separated, dewatered, dried, and calcined to recover sulfuric acid.

Gypsum Recovery--

In general, wet lime/limestone FGD requires several process enhancements to supply consistently pure gypsum. These enhancements include forced oxidation to produce the required gypsum content in the slurry solids, particle agglomeration, and washing of gypsum filter cake. The gypsum slurry is then dewatered using conventional equipment, resulting in a cake that is greater than 80 weight percent solids (20, 21). Approximately 2.7 tons of gypsum is produced per ton of SO_2 removed from the flue gas (22). Washing of product gypsum, especially with high-chloride coals, generates a liquid waste stream.

Wellman-Lord--

The main features of the Wellman-Lord process (Figure 2.6-1) are the use of a sodium-based sorbent, regeneration of the sorbent solution, and conversion of SO_2 to sulfuric acid or elemental sulfur. The scrubbing portion

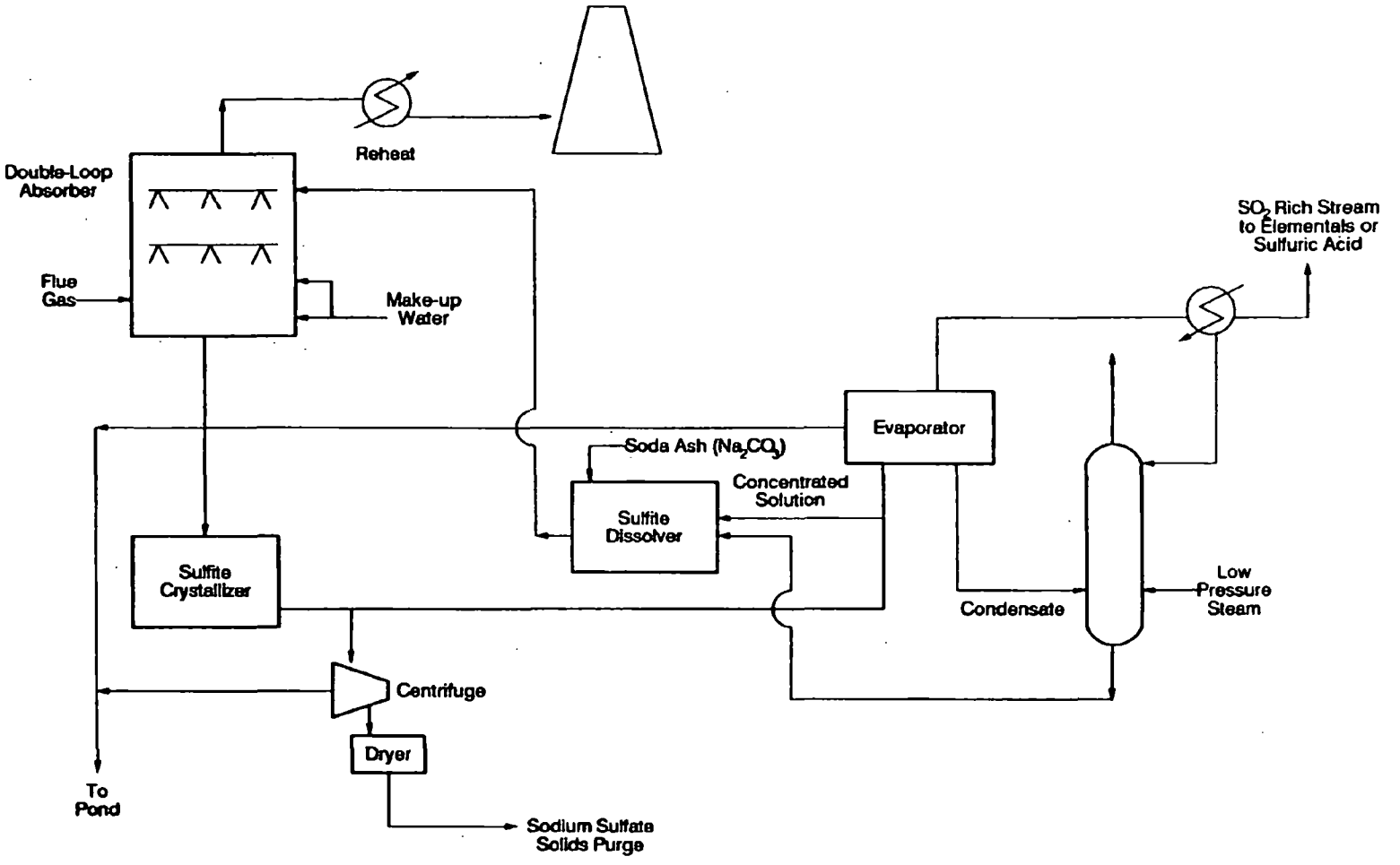


Figure 2.6-1. Wellman-Lord Process Schematic

of the process involves circulating sodium sulfite solution, typically in a plate-type absorber, to absorb SO_2. Sodium bisulfite is the primary reaction product, though some oxidation to sodium sulfate will occur depending on the flue gas oxygen content. This sulfate is purged as a solid, using a crystallizer followed by drying. The sodium bisulfite solution from the absorber is thermally decomposed in a steam-heated evaporator to regenerate sodium sulfite. Solids formed due to evaporation are redissolved in a separate vessel; at this point make-up soda ash (Na_2CO_3) is added as necessary. Overhead vapor from the evaporator is an SO_2-rich stream, suitable for conversion to elemental sulfur or sulfuric acid using commercial technology. A purge stream requiring treatment and disposal is produced in order to keep thiosulfate levels (i.e., dissolved solids) below a threshold.

Magnesia Slurry--

In the magnesia slurry process (Figure 2.6-2), an aqueous slurry of regenerable magnesium hydroxide and magnesium sulfite is used to absorb SO_2 from flue gas. Reaction products formed in the absorber include several magnesium sulfite hydrates and magnesium bisulfite. A portion of the reacted slurry from the main scrubbing loop is sent to a thickener for partial dewatering. A centrifuge is then used to recover a wet cake of magnesium sulfite/sulfate crystals. The crystals are sent to a rotary or fluidized bed dryer while the liquor is returned to the circulation loop. Following the solids drying step, decomposition of the magnesium sulfite/sulfate compounds is carried out in a high-temperature calciner. This liberates a concentrated SO_2 gas stream that can be routed to the sulfur (or sulfuric acid) production unit. The calcined solids are regenerated as MgO for recycle to the absorber loop.

Applicability

Gypsum Recovery--

Gypsum recovery technology can be used almost anywhere ordinary lime/limestone scrubbing is appropriate. It is only necessary to add forced oxidation, product solids washing (to assure adequate gypsum purity), and

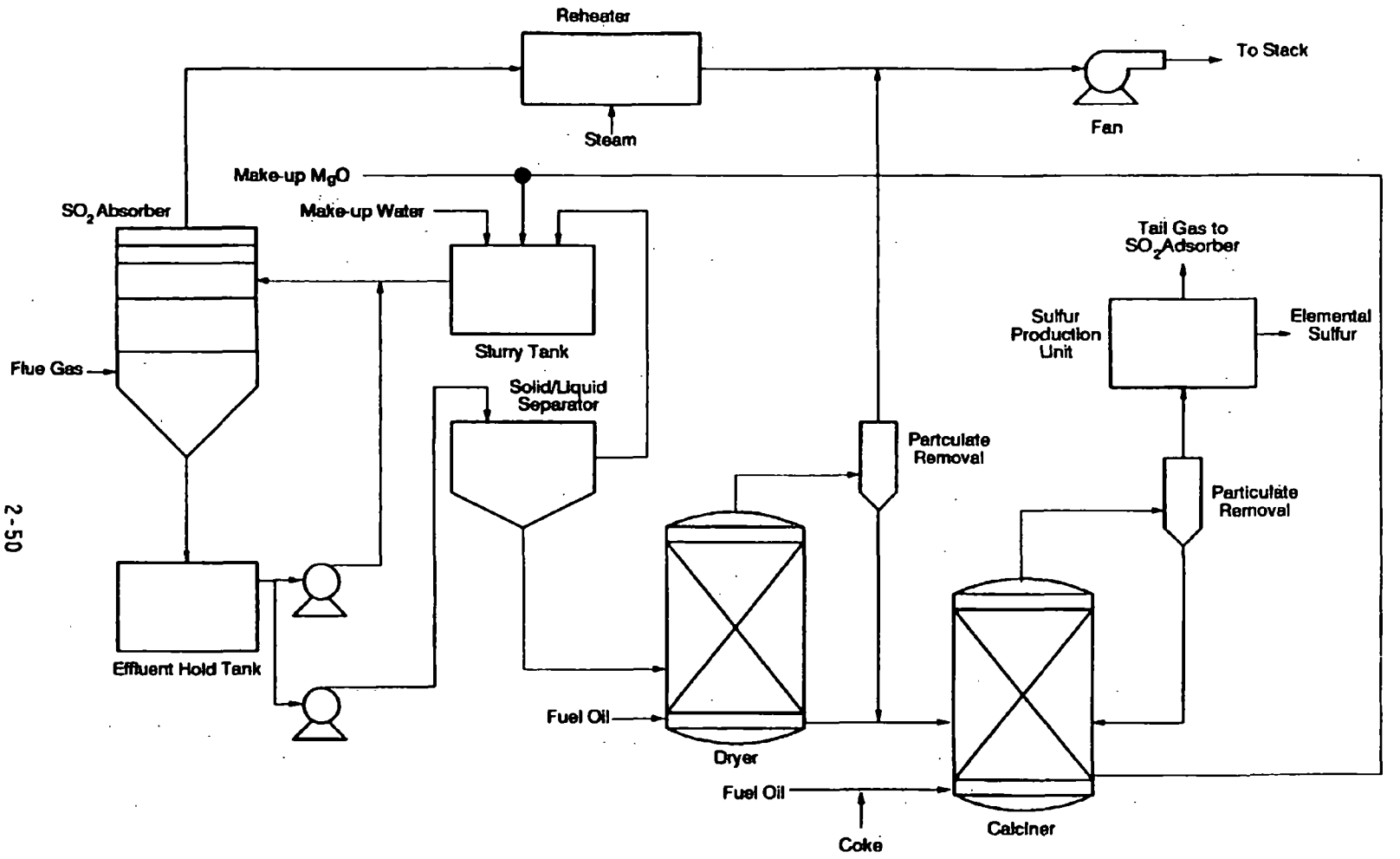


Figure 2. 6-2. Magnesia Slurry Process Schematic (23)

shipping facilities. The primary commercial use of this technology is to produce gypsum for wallboard or as a cement retarder. Approximately 20 forced oxidation units are operating in the U. S. However, only two plants have plans to recover gypsum for sale. In Japan, FGD gypsum is routinely substituted for natural gypsum, and practically all the lime/limestone FGD installations in Europe successfully market recovered gypsum. The main motivation for by-product utilization overseas is lack of space and high cost for waste disposal.

Annual consumption of raw gypsum in the U. S. was approximately 20 million tons as of 1984. The wallboard industry consumes about 70 percent of this amount, with the balance used in the manufacture of cement and plaster, and in agriculture (22). Wallboard manufacturers historically considered only purity and grain-size; however, for recovered FGD gypsum other factors must be considered. For example, as soluble salt concentration increases the gypsum calcination temperature decreases, which may result in weakening of the bond between the paper coating and gypsum core.

Wellman-Lord--

The Wellman-Lord process has been retrofit on coal-fired utility boilers in both the United States and Japan. Important features of the technology include SO_2 removal levels in excess of 90 percent over a wide range of inlet SO_2 concentrations (24) and minimal scaling and plugging. However, as with other regenerable FGD processes, the Wellman-Lord system is mechanically and chemically complex. Also, handling the concentrated SO_2 product stream as elemental sulfur or sulfuric acid entails additional complex processing, and advantageous economics require an available market for the end product. Another limiting factor for utility applications may be the disposal of the sodium sulfate solids, although the waste volumes are considerably less than for throwaway FGD systems.

The Wellman-Lord system is commercially available and as of 1985 has been applied to the four U. S. utility plants listed in Table 2.6-1 (25).

TABLE 2.6-1. WELLMAN-LORD UTILITY INSTALLATIONS IN THE U.S. (24)

Utility	Plant Name (MW)	%S Fuel	SO_2 End Product	Start-up
N. Indiana Public Ser.	Mitchell (115)	3.0	Elemental S	1976
Public Ser. Co. of N. M.	San Juan (700)	0.8	Elemental S	1978
Delmarva Power & Lt.	Delaware City (180)	7.0 (coke)	Sulfuric acid	1979
Public Ser. Co. of N. M.	San Juan (1100)	0.8	Sulfuric acid	1982

These applications represent both low sulfur (0.8 percent) and high sulfur (7 percent) coals. In early 1986, it was reported that one German and one Italian utility had ordered construction of Wellman-Lord units; the latter being a 10 MW demonstration plant (26). In Japan there are two utility installations, one at Chubu Electric's Nishinagoya station (620 MW) and the other at the Niigata station (380 MW) of Tohoku Electric (1).

Retrofits are possible, but there must be adequate land available near the boiler and it must be possible to modify the existing ductwork to route the flue gas to and from the FGD system. Many of the industrial sites where the Wellman-Lord system is employed are existing sulfuric acid or sulfur plants.

Magnesia Slurry--

The magnesia slurry process has been demonstrated on both oil- and coal-fired utility systems. Retrofit is possible and the process has good turn-down capability. The technology is well developed; in the mid-1970's two demonstration units having capacities of 90 and 150 MW operated for 2 years (27). Three commercial units totaling 724 MW were brought into service by Philadelphia Electric Company in 1983 (1). $MgSO_3$ is shipped off-site to existing sulfuric acid plants and the regenerated MgO is returned to the power plant site.

Performance

SO_2 removal performance of gypsum by-product recovery plants is about the same as conventional lime/limestone FGD systems. However, system operation must be more carefully controlled in order to avoid excessive unreacted reagent in the product solids. The 2250 MW Martin Lake plant has operated since late 1984 at better than 90 percent SO_2 removal, with greater than 95 percent oxidation and 90 percent utilization of limestone reagent. The plant burns a lignite with a sulfur content of 1-4 percent. The highly oxidized sludge clarifies quickly, but careful solids handling is needed to avoid line pluggage. Also, close pH control is necessary to maintain a sufficiently high oxidation level.

The 160 MW Muscatine, Iowa plant has successfully operated a forced oxidation system at 94 percent sulfur removal with 3.5 percent sulfur coal (20). After initial start-up in early 1984, the operability of the system has averaged better than 95 percent, and reliability above 99 percent. There have been some problems with scaling of gypsum on the scrubber internal components due to the high oxidation level. The largest problem to date with the produced gypsum has been excess unreacted limestone remaining in the solids.

Commercial Wellman-Lord systems have demonstrated removal of SO_2 in excess of 90 percent. High SO_2 removal is achievable due to the high alkalinity of the sodium-based sorbent solution. In addition, the process can handle widely varying inlet SO_2 concentrations, and scaling or solids-plugging do not present a problem. However, process complexity has resulted in greater maintenance requirements than for conventional lime/limestone FGD operation.

Cost

The costs for producing salable gypsum include many components, but are partially offset by disposal cost savings. In general, the additional operating costs for gypsum production roughly balance those for disposal The key variable then becomes whether the market and price for the product can justify the additional capital costs of increased washing capacity, forced oxidation, or other processing. Another factor affecting this option in the U.S., was a tax depletion allowance of 14 percent (as of 1984) granted to gypsum mining companies (22). This tax credit reduces the actual cost of mined gypsum, making it more difficult for FGD gypsum to be competitive.

Table 2.6-2 presents estimated capital costs, operating costs, and cost effectiveness for a Wellman-Lord FGD system as applied to a coal-fired utility boiler using EPRI guidelines (28). The Wellman-Lord process is more costly than conventional lime/ limestone FGD systems. However, the capital cost differential of the two systems decreases with lower sulfur coals. Operating costs are considerably higher for the Wellman-Lord process, as might be

TABLE 2.6-2. COST ESTIMATES FOR WELLMAN-LORD FGD
(Current 1988 $)

Technology	% SO_2 Removed	3.5% S	2% S
New 500-MW Baseload Power Plant:			
Capital Cost, $/kW	90	259 - 290	204 - 229
Mills/kWh, Constant $	90	11.7 - 12.1	9.4 - 9.7
Current $	90	20.0 - 21.0	16.1 - 16.7
$/Ton, Constant $	90	443 - 461	624 - 647
Current $	90	763 - 795	1,070 - 1,214
Retrofit 500-MW Baseload Power Plant[1]:			
Capital Cost, $/kW	90	335 - 375	265 - 296
Mills/kWh, Constant $	90	12.8 - 13.3	10.2 - 10.6
Current $	90	22.1 - 23.2	17.7 - 18.6
$/Ton, Constant $	90	484 - 506	680 - 709
Current $	90	832 - 870	1,169 - 1,270
Retrofit 250-MW Intermediate Load Power Plant[2]:			
Capital Cost, $/kW	90	538 - 602	424 - 475
Mills/kWh, Constant $	90	31.5 - 33.8	25.7 - 27.6
Current $	90	44.1 - 47.5	35.9 - 38.7
$/Ton, Constant $	90	1,197 - 1,289	1,712 - 1,839
Current $	90	1,679 - 1,811	2,398 - 2,579

[1]Retrofit difficulty of 1.3 and capacity factor of 65.

[2]Retrofit difficulty of 1.6 and capacity factor of 35.

expected in view of the greater mechanical and chemical complexity. Retrofit costs are also substantially higher relative to new installation of equivalent or greater capacity lime/limestone systems. In these cost comparisons, the market potential for recovered by-product at a particular location is not explicity included.

Extremely limited experience, to date, with the magnesia slurry process on a commercial scale would make specific cost comparisons unreliable. However, it is generally recognized that capital and O&M costs of the magnesia process are higher than conventional lime/limestone FGD.

2.7 SPRAY DRYING

Description

Spray drying is an established technology for desulfurization of flue gas from high- and low-sulfur coals. The first installation of utility spray dryer systems occurred in the late 1970's. Thirteen systems, representing nearly 4,300 MW of generating capacity, are currently in service, and an additional five systems representing over 2,800 MW of generating capacity are in design or construction phases (29). All of these full-scale systems are based on coals containing less than two percent sulfur. Evaluation of spray dryer systems for utility boilers burning greater than three percent sulfur coal was initiated in 1985 using two 1 MW, pilot-scale spray dryers at TVA's Shawnee Test Facility.

A typical spray dryer system is shown in Figure 2.7-1. Lime is the reagent in 17 of the utility systems mentioned above, while one (the first spray dryer system purchased) uses a sodium-based reagent. Only lime-based systems are described here.

In a lime-based system, quicklime is slaked to form calcium hydroxide, and the resulting slurry is combined with additional process makeup water and (in most systems) recycled solids. This combined slurry is then atomized and mixed with the flue gas at air heater exit conditions in the spray dryer vessel. Slurry atomization is generally accomplished with a rotary atomizer. Two-fluid air atomizing nozzles are used in some systems. Several simultaneous reactions occur immediately following atomization:

- Water in the slurry droplet evaporates;
- The evaporating water cools and humidifies the flue gas; and
- Flue gas SO_2 reacts with the lime, producing calcium sulfite and calcium sulfate solids.

The amount of water fed to the spray dryer is carefully controlled to avoid complete saturation of the flue gas. However, a close approach to

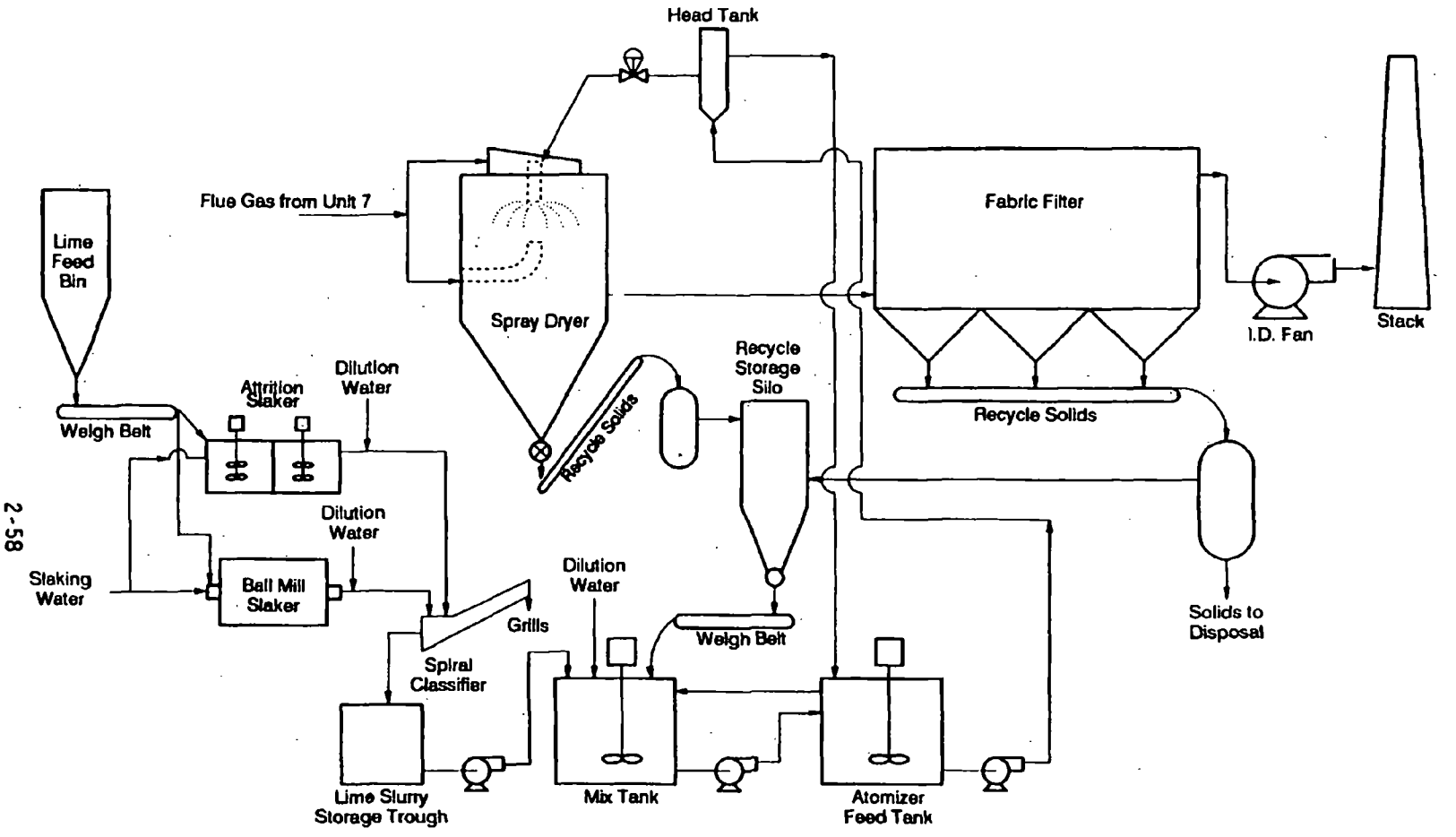


Figure 2.7-1. Lime Spray Drying Process Flow Diagram (30).

saturation temperature (20^0-35^0F) is needed to achieve high SO_2 removal and lime utilization. Sorbent feed rate, degree of slurry atomization (i.e., droplet size), flue gas residence time in the spray dryer vessel, and the approach to flue gas saturation temperature are carefully controlled to achieve acceptable SO_2 removal performance while avoiding problems with drop out or caking of incompletely dried solids in the spray dryer vessel. The calcium content in the slurry is limited to a maximum of about 35 percent solids. Typical flue gas residence time in the spray dryer vessel is 8-12 seconds.

The nearly-saturated flue gas leaves the spray dryer vessel and enters the particulate collection device which is either a fabric filter or an electrostatic precipitator (ESP). Most utility experience to date has been with fabric filtration. In the particulate collection device, the original coal flyash and the dried sorbent are removed from the flue gas. Additional SO_2 removal occurs across the particulate control device. Fabric filters contribute as much as 20-30 percent to overall system SO_2 removal (30). Less information is available on SO_2 removal across an ESP, but it may be as high as 25 percent at low approach-to-saturation temperatures (20^0F) (31).

In most systems, a portion of the collected solids are recycled to the slurry mix tank. Solids recycle offers several advantages over once-through lime operation, including improved lime utilization, improved rotary atomizer operation, faster droplet drying, and reduced solid waste (32).

Lime spray drying has several perceived advantages over wet lime/limestone scrubbing. First, the need to recirculate large quantities of slurry are avoided. Liquid-to-gas volume ratios for spray dryer systems tend to be 10 to 50 times smaller than what is typical for lime/limestone scrubbing. Second, wastes from the spray drying process are produced as dry solids, rather than as a wet sludge. However, moving and storing relatively large quantities of dry solids can also be a troublesome operation. Third, fewer corrosion problems occur with lime spray drying since the system is dry rather than wet.

Applicability

Although spray drying is generally considered a commercially established technology for low-sulfur coal utility applications, data with high-sulfur coals are limited. In high-sulfur coal applications, heat balance considerations will limit SO_2 removal by limiting the amount of lime and water that can be added as well as the amount of sorbent recycle. Specifically, the differential between the spray dryer inlet temperature and the approach to flue gas saturation temperature at the spray dryer outlet will determine the amount of water that can be evaporated by the flue gas. If too much water is used, the flue gas will cool to below saturation temperature and result in scaling and corrosion problems. If too little water is used, poor slurry atomization and low lime utilization will occur. Additives (specifically chlorides) have been successfully used in short-term tests to increase sorbent utilization at high SO_2 removal rates.

Spray dryers can be used in retrofit applications if there is sufficient land near the boiler and if ductwork can be cost-effectively run from the air heater outlet to the spray dryer inlet and from the spray dryer outlet to the particulate control device.

The type of existing particulate control device may influence the applicability of spray drying as a retrofit FGD technique, particularly for high-sulfur coal applications. Fabric filters are well suited for spray dryer retrofits, but few existing utility boilers are so equipped (particularly in high-sulfur coal service). Existing ESPs have several limitations, including:

- Less contribution to system SO_2 removal than a fabric filter;
- Greater sensitivity to increased particulate loading caused by the sorbent (addition of a spray dryer will require greater particulate removal efficiencies to maintain emission rates at pre-spray drying levels); and
- Potential corrosion problems.

However, these impacts are poorly quantified and more testing is needed.

For low-sulfur coal applications where high recycle rates are achievable (30), adequate SO_2 removal performance may result even with relatively low SO_2 removal in the particulate control device. Also, most existing ESPs designed for low-sulfur coal have high specific collection area (SCA) values of 400 ft^2/thousand acfm and greater. With high initial SCA values, the beneficial effects of gas volume shrinkage and gas conditioning which results from cooling and humidification in the spray dryer may more than offset the effects of increased mass loadings.

Many existing ESPs in high-sulfur coal applications have SCA values of 200 or less. At these SCA values, it is unlikely that existing emission rates will be maintained after a spray dryer is installed upstream. Pilot studies conducted by TVA (31) indicate that an ESP with an SCA of 300 which currently meets existing particulate emission regulations by a comfortable margin should be able to maintain compliance after a spray dryer is added. If the ESP cannot be adequately upgraded, installation of a replacement baghouse would be required.

Other issues for spray dryer retrofits include increased corrosion of particulate control devices due to operation at lower gas temperatures and higher moisture content, and the need to upgrade existing particulate collector solids handling equipment to deal with increased loadings and alkalinity caused by the spray dryer. Replacement of wet sluice ash handling systems with a dry system would also be required to preheat plugging of the sluice lines.

Performance

All existing and currently planned lime spray dryer FGD systems are designed for coals with less than two percent sulfur; design SO_2 removals are 61 to 90 percent (29). Most of the 13 operating units readily meet their design SO_2 removal levels. Less experience is available for high-sulfur coal applications. Long- term utility test results are not yet available. Short-term spray dryer tests combined with fabric filtration have been conducted on high-sulfur coals at Argonne National Laboratory, General Motors' Buick Steam Plant, and Northern States Power Company's Riverside Station. These tests demonstrated over 90 percent SO_2 removal on coals up to 4.2 and 3.2 percent sulfur with and without the use of additives, respectively (33). In

pilot-scale testing using a lime spray dryer/baghouse in high-sulfur coal applications, TVA demonstrated that greater than 80 percent removal is feasible. Lime utilization at these SO_2 removal rates was 60-65 percent.

Pilot-scale tests for a high-sulfur spray dryer/ESP system conducted by Wheelabrator Air Pollution Control and the U.S. Department of Energy achieved SO_2 removal levels as high as 90 percent with a 2.7 percent sulfur coal (34). Lime utilization was only about 30 percent; however, these tests did not employ recycle and the approach to flue gas saturation was not as close as is possible to maximize lime utilization. The spray dryer inlet temperature for these tests was rather high at 400°F.

Tests conducted by TVA on a 10 MW spray dryer/ESP pilot unit at Shawnee with a 3.5 to 4 percent sulfur coal and a 300°F inlet flue gas temperature achieved overall SO_2 removal levels of 90 percent (35). Most of these tests were conducted at a 18°F approach to adiabatic saturation at the spray dryer outlet.

The data discussed above indicate that high levels of SO_2 removal (90 percent) are possible for new units that are designed to operate at these levels. However, for retrofit applications where the existing ESP will be reused, SO_2 removal may be limited by ESP performance. EPRI also is sponsoring research on lime spray drying at its High Sulfur Test Center. Pilot-scale tests on a 4 MW spray dryer with fabric filter have achieved 90% sulfur dioxide removal on coals with 3.5-4% sulfur and a calcium-to-sulfur ratio of 1.3 to 1.6.

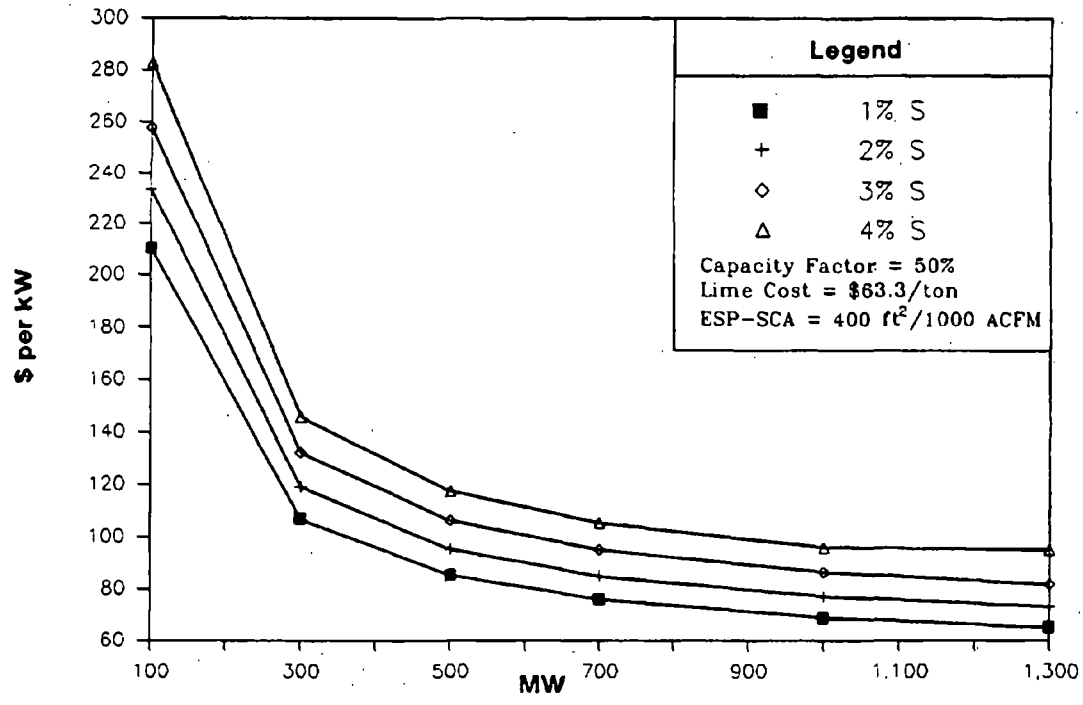
The potential for combined SO_2/NO_x reduction also exists by addition of roughly 10 weight percent sodium hydroxide to the lime. The NaOH both reacts with SO_2 and extends particle reactivity as well as serves as a catalyst to reduce NO_x emissions. In pilot-scale tests with a three percent sulfur coal and a lime-to-sulfur ratio of 1.3, SO_2 and NO_x reductions of 95 and 55 percent, respectively, were achieved. However, the process oxidizes a portion of the NO to NO_2, resulting in increased plume opacity and discoloration (36). Ammonia injection has been demonstrated to reduce the NO_2 plume.

Cost

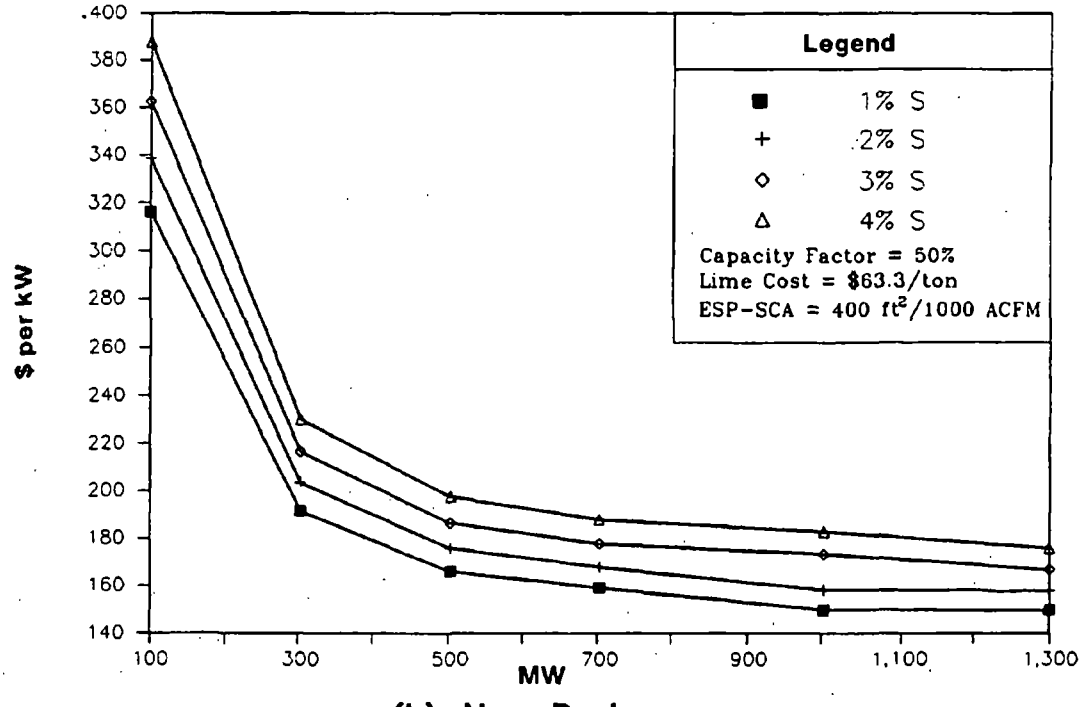
Due to the sensitivity of spray dryer performance to plant design and operating variables, capital and operating costs for retrofit spray dryer applications are expected to be more site specific than wet lime/limestone FGD. Key factors in evaluating spray dryer costs at a given site include air heater outlet temperature, length of ducting between the air heater outlet and spray dryer inlet, percent sulfur in the coal, and whether the existing particulate control device (typically an ESP) can be upgraded to handle the increased particulate loading resulting from spray dryer operation. The expected SO_2 reduction across a baghouse versus an ESP is also important.

A critical consideration in spray dryer economics will be whether to use the existing ESP or install a new baghouse. Installation of a new baghouse can increase retrofit capital costs by 30-50 percent and annualized costs by 25-40 percent above the cost of a spray dryer system in which the existing ESP can be used. Because of the uncertainty in spray dryer performance and cost, two cost cases are examined. One case assumes that the existing ESP can be upgraded and continues in service. The other assumes that a new baghouse will be installed.

The following table presents the range of values estimated using IAPCS cost model for capital cost, annualized cost, and cost per ton of pollutant removed. The lower costs are for a large unit with a low retrofit factor burning low sulfur coal. The higher costs are for a small unit with a high retrofit factor burning high sulfur coal. Figures 2.7-2 through 2.7-4 show these costs as a function of boiler size, coal sulfur content, and particulate control option.

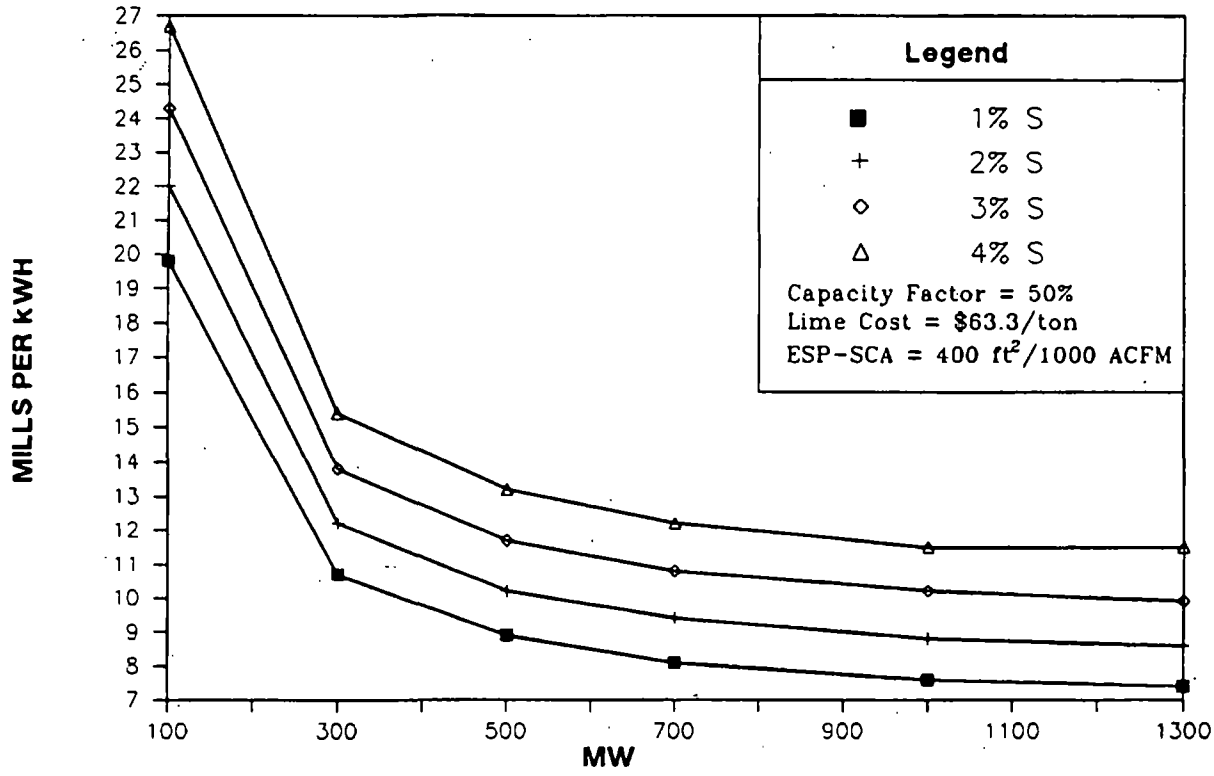


(a) Existing ESP

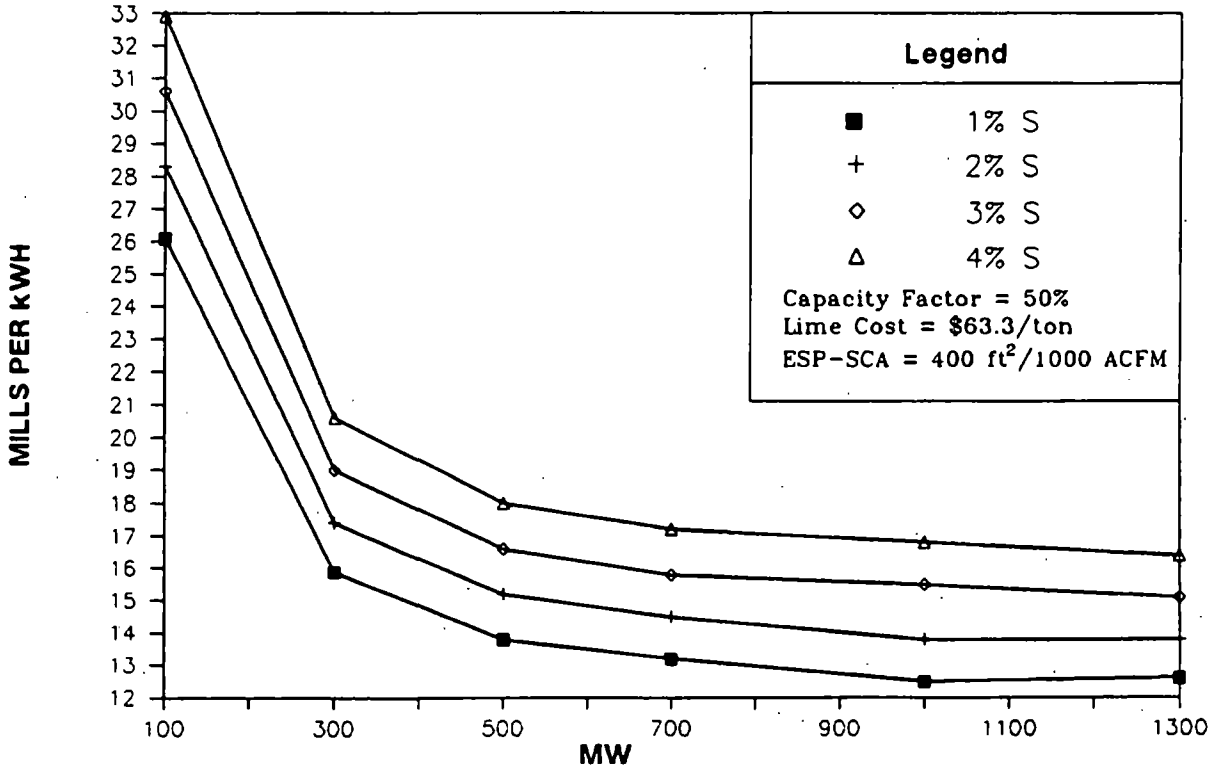


(b) New Baghouse

Figure 2.7-2. Lime Spray Drying - Capital Cost, Current 1988 Dollars

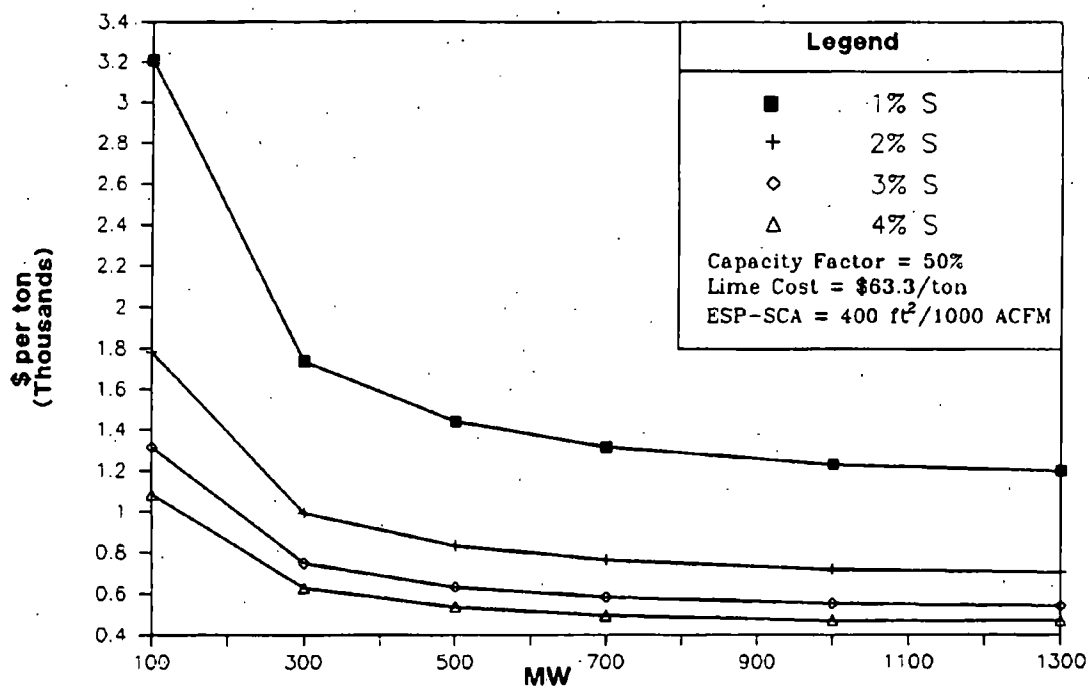


(a) Existing ESP

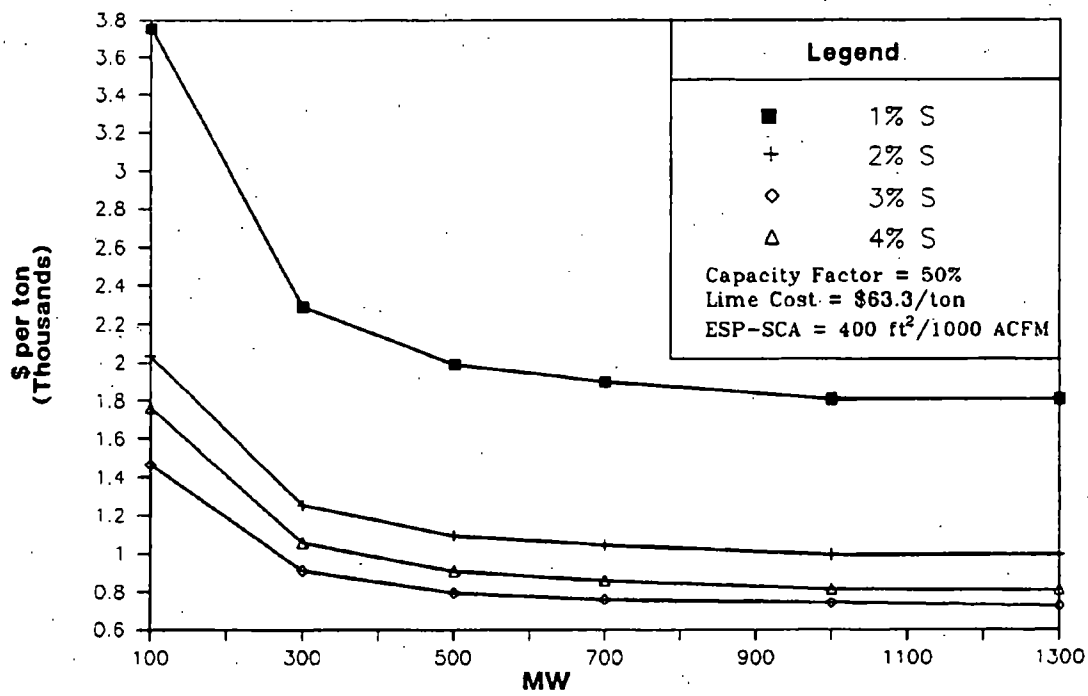


(b) New Baghouse

Figure 2.7-3. Lime Spray Drying - Annualized Cost, Current 1988 Dollars



(a) Existing ESP



(b) New Baghouse

Figure 2.7-4. Lime Spray Drying - Cost Per Ton of SO2 Removed, Current 1988 Dollars

	Range
Existing ESP -	
Capital Cost ($/kW)	64.7 to 535.4
Annualized Cost (mills/kWh)	6.3 to 65.7
Cost Per ton of SO_2 Removed ($/ton)	441.5 to 8,084.6
New Baghouse -	
Capital Cost ($/kW)	149.2 to 639.2
Annualized Cost (mills/kWh)	10.2 to 75.7
Cost Per ton of SO_2 Removed ($/ton)	522.2 to 8,632.4

Appendix F contains tables of costs as a function of boiler size, coal sulfur, capacity factor, and retrofit factor.

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SECTION 3

NEAR-COMMERCIAL TECHNOLOGIES

INTRODUCTION

This section discusses six technologies which are in the process of being demonstrated by U. S. utilities or which have been commercially applied to power plants overseas. The first two technologies, integrated gasification combined-cycle (Section 3.1) and fluidized bed combustion (Section 3.2), are integrated processes that can increase the electric generating capacity of a plant while minimizing SO_2 and NO_x emissions. Because of their design and high capital cost, these two repowering technologies will be used by utilities that have a need to increase or maintain base load electric generating capacity as well as to reduce emissions. When compared with the cost of a new plant (including emission control systems), addition of base load generating capacity through repowering of existing capacity can be very economical.

Post-combustion NO_x control technology (Section 3.3), has been commercially applied to boilers burning low-sulfur fuels in Europe and Japan, but has little or no long-term operating experience with medium- and high-sulfur coals. Because of specific process concerns, it is categorized in this report as a near-commercial technology until its use with higher sulfur fuels has been demonstrated.

The final three technologies, furnace and duct (also known as hot-side and cool-side) sorbent injection (Sections 3.4 and 3.5) and reburning (Section 3.6) have limited commercial application. These technologies have limited or planned large-scale demonstration experience in the U. S. and should be commercially available within the next few years.

3.1 INTEGRATED GASIFICATION COMBINED CYCLE

Description

Integrated gasification combined cycle (IGCC) is an alternative to conventional coal-fired electric power generation with post-combustion emission controls. Because of its overall design, emissions of sulfur and nitrogen oxides and particulates from IGCC facilities are projected to be significantly lower than from existing technologies. Figure 3.1-1 is a generalized block flow diagram of an IGCC facility designed to 1) convert coal (via partial oxidation and gasification) into a fuel gas, 2) remove reduced sulfur species (primarily hydrogen sulfide) from the fuel gas, 3) use the clean fuel gas to produce electricity in a combined-cycle application, and 4) treat and/or recover byproducts from the waste streams generated. Existing technology for removal of sulfur species from fuel gas requires gas cooling followed by scrubbing with any of several proprietary chemical processes. Cooling the gas imposes both additional capital cost and a significant penalty on process thermal efficiency. Advanced processes for removal of H_2S from hot fuel gas (1,000 to 1,200^{o}F) using adsorption by metal oxides are under development with funding from DOE. Development of hot gas cleanup technology will simplify IGCC process complexity and reduce costs.

A number of coal gasification processes have been examined for use in combined-cycle applications. The most successful demonstration of IGCC technology in the U. S. is the Cool Water Coal Gasification Demonstration Project. The Cool Water facility, located in Daggett, California, is based on use of Texaco gasification technology with cold gas cleanup. The facility is designed to gasify 1000 tons of coal per day and produce approximately 100 MW of net electrical power. The facility underwent successful startup in 1984 to become the first successful domestic operation combining coal gasification and combined cycle equipment. The description of Cool Water presented below is representative of the general concept of IGCC facilities.

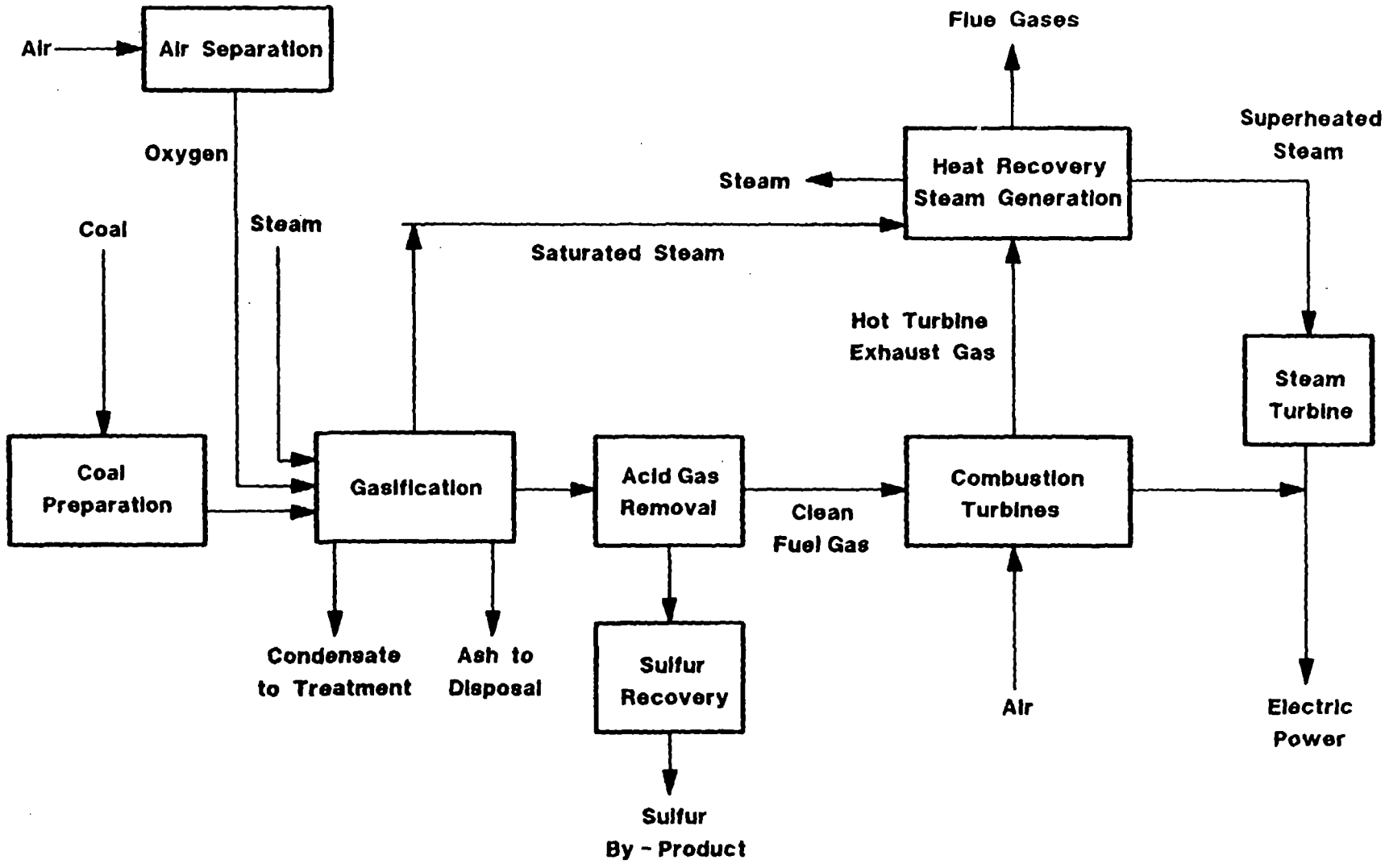


Figure 3.1-1. Generalized block flow diagram of combined cycle coal gasification power generation

The Texaco gasifier is a pressurized, downflow, entrained reactor in which pulverized coal is fed as a coal/water slurry. High purity oxygen (95+ percent) is also introduced into the top of the gasifier. The coal undergoes partial oxidation at high temperature and pressure to produce a raw fuel gas composed primarily of hydrogen, carbon monoxide, and carbon dioxide. The raw fuel gas also contains reduced sulfur and nitrogen species, water vapor, entrained molten slag, and soot particles; no condensible hydrocarbons are found in the gas.

The raw fuel gas is cooled in radiant and convective coolers, producing saturated steam for use in the combined-cycle power generation system. The fuel gas is initially treated in a particulate scrubber and a series of heat exchangers and coolers. Cooled fuel gas is then sent to the sulfur removal unit (the Selexol process is used at Cool Water) with a design sulfur removal efficiency of 97 percent. Removed sulfur compounds are subsequently converted into elemental sulfur in a Claus sulfur recovery unit equipped with a SCOT tail gas treating unit.

The clean fuel gas is then reheated and saturated with moisture before being combusted in a gas turbine to produce electricity. Moisture addition is used to reduce NO_x formation during combustion. Exhaust gas from the gas turbine is sent through a heat recovery steam generator (HRSG) to produce steam. This steam plus that from the fuel gas heat exchangers and coolers is sent to the steam turbine to produce additional electricity. Cooled flue gas from the HRSG is discharged through a stack to the atmosphere.

Auxiliary processes included in an IGCC facility include 1) coal storage and preparation, 2) oxygen production (if pure or nearly pure oxygen is used as the gasifier reactant) and compression, 3) wastewater treatment, and 4) other processes normally present in a coal-fired electric generating plant (e.g., cooling towers and ash handling/disposal).

Applicability

A number of coal gasification technologies are commercially available, and oil- and gas-fired combined-cycle power plants are currently in commercial use.(1)

With the recent success of the Cool Water Demonstration Program and continued research/demonstration activities, IGCC has become a rapidly emerging alternative for electricity generation at new plants and for repowering existing plants. Potential economic advantages include high modularity which will allow IGCC plants to be constructed in relatively small increments and ability to use lower priced (higher sulfur) coals. Retrofit to an existing coal-fired power plant will require construction of new gasification and combustion turbine facilities. However, depending on site-specific factors, it may be possible to incorporate some existing equipment into the IGCC plant (e.g., steam turbines, boiler feed-water and cooling systems, coal handling, and electric generator and auxiliaries).

Performance

The electric power industry has a growing interest in IGCC technology because of a number of environmental advantages versus conventional coal-fired power plants, including 1) lower SO_2, NO_x, and particulate emissions, 2) lower water consumption, and 3) lower land requirements. Sulfur dioxide emissions from an IGCC facility primarily arise from two sources: combustion flue gas and tail gases from the sulfur recovery unit. Very stringent control of sulfur emissions from an IGCC plant is attainable even for gasification of high-sulfur coals.

The sulfur removal unit in an IGCC is designed to remove sulfur species (primarily hydrogen sulfide) from the fuel gas prior to combustion, rather than removal of SO_2 following combustion. Sulfur control efficiencies greater than 95 percent are achievable with a properly designed sulfur recovery unit. Sulfur species remaining in the fuel gas are converted to sulfur oxides in the gas turbine combustor.

Sulfur emission test results from Cool Water are summarized in Table 3.1-1. These results indicate that SO_2 emissions from IGCC can be reduced to levels significantly below current utility NSPS requirements.

TABLE 3.1-1. AVAILABLE SO_2 EMISSIONS DATA FOR COOL WATER DEMONSTRATION PROGRAM

Coal Gasified	Sulfur Removal	Combined-Cycle Combustion Flue Gases			Incinerator Stack		(Reference)
		(SO_2 ppmvd[a])	(lb SO_2/hr)	(lb SO_2/10^6 Btu[b])	(SO_2 ppmvd[a])	(lb SO_2/hr)	
Low sulfur[c]	NR	NR	NR	0.032	NR	NR	2
Low sulfur[c]	97%	NR	NR	0.034	NR	NR	4
Low sulfur[c]	NR	8.59[d]	33.1[e]	NR	54	3.2	7

NR = Not reported.

[a]Parts per million by volume, dry basis.

[b]Pounds SO_2 per million Btu of coal.

[c]Sulfur content not reported.

[d]0.46 ppmvd of H_2SO_4 mist also reported.

[e]2.7 lb/hr of H_2SO_4 mist also reported.

Although the Cool Water test data shown are for gasification of a low-sulfur coal, similar results are expected with medium- and high-sulfur coals.

The Electric Power Research Institute (EPRI) has funded a number of studies relating to IGCC technologies (3,5,6,8,10). As part of these studies, estimates of the performance and cost of IGCC facilities gasifying high-sulfur Illinois #6 coal have been developed. Table 3.1-2 summarizes the sulfur control levels incorporated into these designs. Even higher levels of sulfur removal (97-99 percent) are achievable at small increases in overall capital cost.

Nitrogen oxide emissions from IGCC facilities are primarily associated with the gas turbine combustion gases. Test data from Cool Water, summarized in Table 3.1-3, indicate NO_x emissions equal to one-tenth the current utility NSPS for coal-fired boilers are achievable. As discussed previously, the Cool Water facility injects steam into the fuel gas prior to combustion in the gas turbine to reduce NO_x emissions.

Cost

Cost and emission data for several IGCC power plants are summarized in Table 3.1-4. Additional cost data compiled by EPRI (9) provide capital cost estimates of $1,204-1,576 per kW (in December 1984 $) for a 500 MW plant using bituminous coal. Key cost variables were the gasification technology (Texaco, Shell, KRW, and British Gas/Lurgi), method of gas cooling (none of these include hot-gas cleanup), and turbine design (current or advanced).

The following table presents the range of IAPCS values for capital and annualized costs. The lower capital and annualized costs are for a large unit with a high capacity factor and high heat rate, while the higher capital and annualized costs are for a small unit with a low capacity factor and low heat rate. Figures 3.1-2 and 3.1-3 show these costs as a function of boiler size and heat rate (cost bases differ from the costs presented in Table 3.1-4). Reported heat rate values (9) range from 7,130 Btu/kWh for Texaco fuel cell combined cycle to 10,250 Btu/kWh for Texaco gasification combined cycle burning bituminous coal. For this study two cases were considered for cost comparison, 8,000 and 10,000 Btu/kWh heat

TABLE 3.1-2. DESIGN SULFUR EMISSIONS INFORMATION FOR EPRI IGCC STUDIES

Gasification Technology	Feed Coal Type	Feed Coal Sulfur Content	Overall Sulfur Removed	Sulfur Dioxide Emissions, $lb/10^6$ Btu[a]
Texaco[b]	Ill. #6	3.4%	95%	0.28
British Gas Corp/ Lurgi (slagging)[c]	Ill. #6	3.4%	94.4%	0.34
Shell[d]	Ill. #6	3.4%	90%	0.56
Kellogg Rust Westinghouse[f]	Ill. #6	3.4%	95%	0.28

[a]Btu of coal feed.

[b]Reference 3.

[c]Reference 5.

[d]Reference 6.

[e]Plant was designed to meet NSPS (i.e., 90% control).

[f]Reference 8.

TABLE 3.1-3. AVAILABLE NO_x EMISSIONS DATA FOR COOL WATER DEMONSTRATION PROGRAM

Reference	$lb/10^6$ Btu[a]	ppmvd[b]	lb/hr[c]
2	0.058	23	61
4	0.059	NR	NR
7[d]	NR	22.8/20.8	61.2/57.2

NR = Not reported.

[a]Per million Btu of coal.

[b]Parts per million by volume, dry basis; normalized to ISO humidity (60%) and 15% O_2.

[c]Calculated as NO_2.

[d]Results shown are for two test periods.

TABLE 3.1-4. SUMMARY OF REFERENCE INFORMATION ON EMISSION AND COST DATA FOR IGCC FACILITIES

Facility	Plant Size, MW	SO_2 Removal	SO_2 Emissions lb/MBtu	NO_x Emissions lb/MBtu	Capital Requirement[a] Million $	Capital Requirement[a] $/kW	Annual O&M Cost,[b] $1,000	Cost Basis
Texaco (3)								
Radiant	600	95%	0.28	0.04	820	1,360	24,800	1/83 $
Radiant & Convective	589	95%	0.28	01.3	860	1,460	26,100	1/83 $
Total Quench	571	95%	0.28	0.04	740	1,300	22,700	1/83 $
Shell (6)	1,122	90%	0.56	NR	1,420	1,260	43,300	1981 $
Kellogg Rust Westinghouse (8)	547	95%	0.28	<75[c]	805	1,470	26,800	1/83 $

[a] Reflects cost of entire plant.

[b] Based on 100% capacity and combination of Illinois #6 coal.

[c] Units = ppmvd, parts per million volume (dry basis).

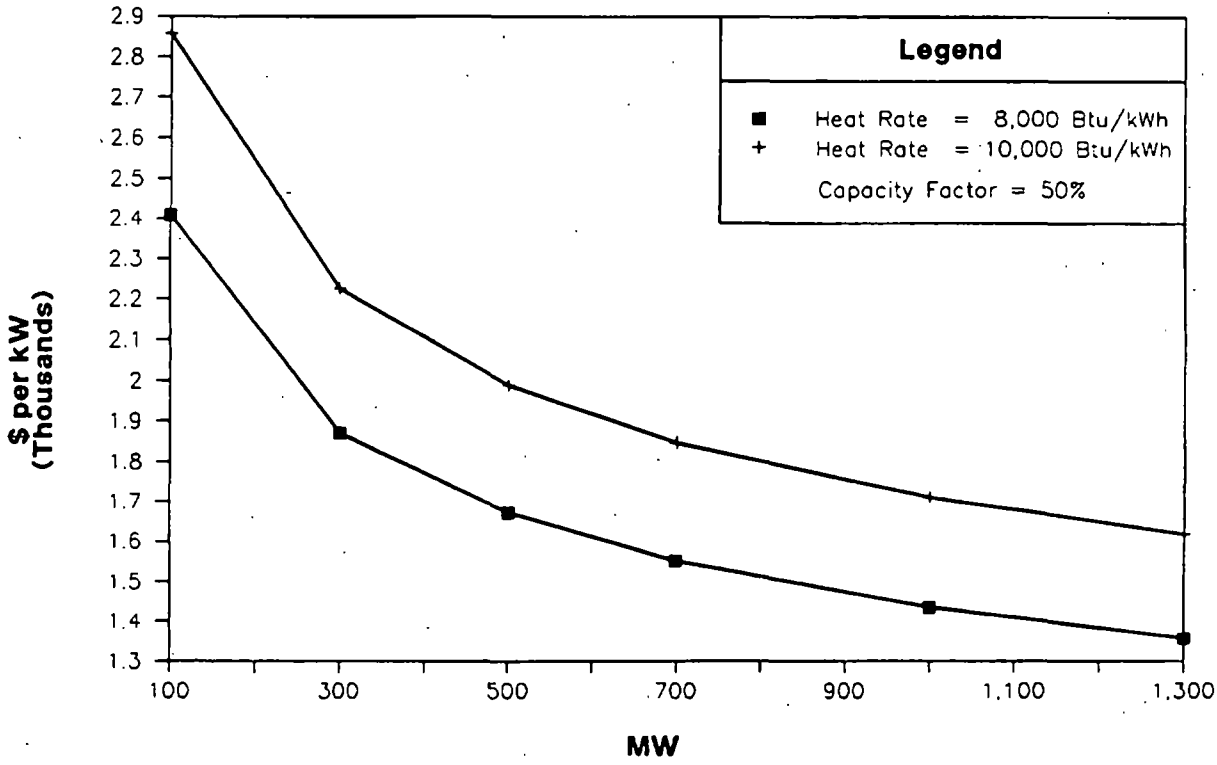


Figure 3.1-2. IGCC - Capital Cost, Current 1988 Dollars

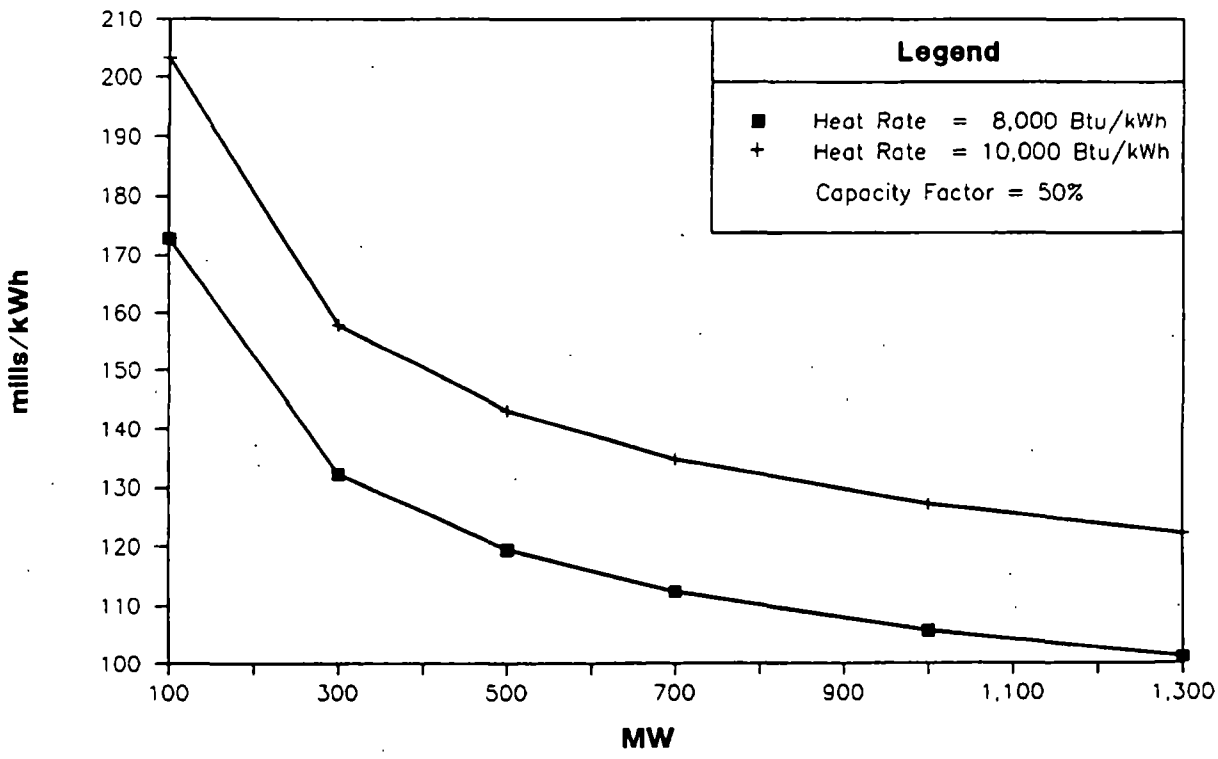


Figure 3.1-3. IGCC - Levelized Annual Cost, Current 1988 Dollars

rate, with direct capital costs for both cases assumed to be the same. This is not to be confused with the more advanced gasification process but is simply to present costs for a range of heat rates reported. The total capital cost for the low heat rate value is considerably lower due to the lower preproduction and inventory costs.

	Range
Capital Cost ($/kW)	1,344 to 2,919
Annualized Cost (mills/kWh)	80.3 to 206.0

Appendix D contains tables of costs as a function of boiler size, capacity factor, and boiler heat rate.

Because IGCC is an integrated process producing electricity as well as reducing SO_2 and NO_x emissions, it is difficult to separate costs associated with emission reduction processes. However, for one facility the capital cost for acid gas removal and sulfur recovery were calculated to be approximately 6 percent of the total plant capital investment. Additional cost information may be found in references 3, 6, and 8.

3.2 FLUIDIZED BED COMBUSTION

Description

Fluidized bed combustion (FBC) is an integrated technology for reducing SO_2 and NO_x emissions during combustion of coal. As with IGCC, FBC can be installed at new plants or used to repower existing plants. FBC technologies based on operation at atmospheric and pressurized conditions have been developed. Atmospheric FBC (AFBC) systems are similar to a conventional boiler in that the furnace operates at near atmospheric pressure and depends upon heat transfer of a working fluid (i.e., water) to recover the heat of combustion. Pressurized FBC operates at pressures greater than atmospheric pressure and recovers energy through both heat transfer to a working fluid and use of the pressurized gas to power a gas turbine. The pressurized systems offer the potential for smaller equipment sizes (and thus simpler retrofits), lower capital cost, and higher efficiency.

Atmospheric FBC --

Figure 3.2-1 presents a simplified AFBC process flow diagram. Coal and limestone are fed into a bed of hot particles (1400-1600°F) fluidized by upflowing air. SO_2 formed during combustion reacts with the calcined limestone to form calcium sulfate, thus reducing SO_2 emissions and avoiding the need for post-combustion controls. The relatively low combustion temperature limits NO_x formation, reduces ash fusion problems, and optimizes sulfur capture. One of the major advantages of AFBC technology is its ability to burn a wide range of fuels, including those with low heat and high sulfur contents.

There are two major AFBC types: the bubbling or dense bed, and the circulating or dilute bed. In the bubbling bed system, coal and limestone are continuously fed into the boiler from over or under the bed. The bed materials consisting of unreacted, calcined, and sulfated limestone, coal, and ash are suspended by the combustion air blowing upwards through the air distributor plate. Bed material is drained from the bed to maintain the

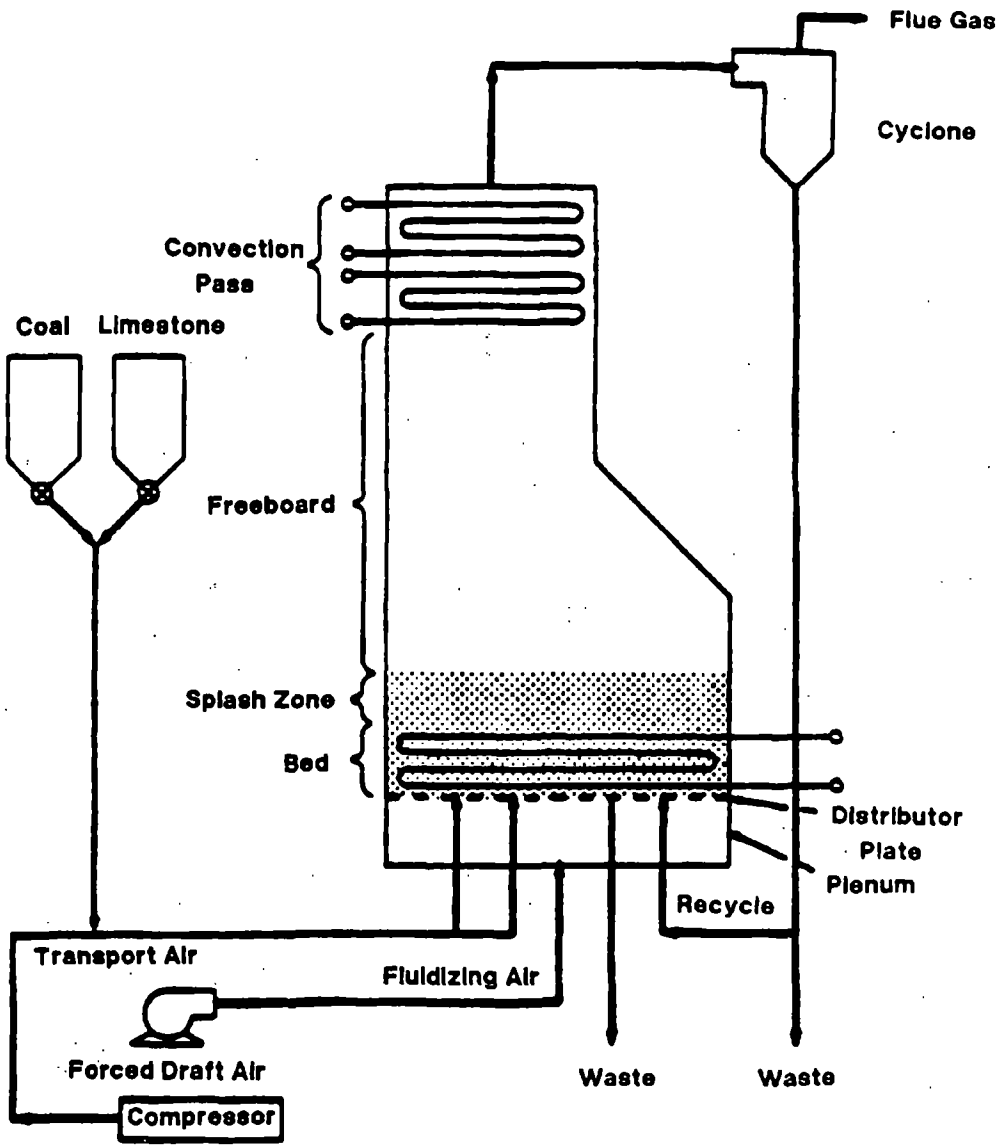


Figure 3.2-1. Simplified AFBC process flow diagram.

desired bed depth. Some bed material is entrained in the upflowing flue gas and escapes the combustion chamber. Approximately 80 to 90 percent of this fly ash is collected in the cyclone and is then either discarded or reinjected into the bed. Reinjection of ash is useful in improving combustion efficiency and limestone utilization. In general, combustion efficiency increases with longer freeboard residence times and greater ash recycle rates. Fly ash not collected in the cyclone is removed from the flue gas by an ESP or fabric filter.

Circulating fluidized bed is a more recent development in AFBC technology. The two major differences between circulating and bubbling AFBC's are: the size of the limestone particles fed to the system, and the velocity of the fluidizing air stream. Limestone feed to a bubbling bed is generally less than 0.1 inches in size, whereas circulating beds use much finer limestone particles, generally less than 0.01 inches. The bubbling bed also incorporates relatively low superficial air velocities, ranging from 4 to 12 ft/sec. This creates a relatively stable fluidized bed of solid particles with a well-defined upper surface. Circulating beds have superficial velocities ranging from 20 to 40 ft/sec. As a result, a physically well-defined bed is not formed; instead, solid particles (coal, limestone, ash, sulfated limestone, etc.) are entrained in the transport air/combustion gases. These solids are then separated from the combustion gases by a cyclone or other separating device and circulated back into the combustion region, along with fresh coal and limestone. A portion of the collected solids are continuously removed from the system to maintain material balances. Circulating beds are characterized by very high recirculated solids flow rates, up to three orders of magnitude higher than the combined coal/limestone feed rate (11).

Several advantages of circulating bed over bubbling bed design have been identified by FBC unit vendors (due to limited test and commercial operating data, however, not all of the advantages have been fully demonstrated):

- Higher combustion efficiency, exceeding 99 percent;
- Greater limestone utilization, due to high recycle of unreacted sorbent and small limestone feed size (greater than 85 percent SO_2

removal efficiency is projected with a Ca/S ratio of about 1.5, with the potential for greater than 95 percent SO_2 removal efficiency);

- Lower NO_x emissions because of staged combustion (less than 100 ppm NO_x are projected);
- Potentially fewer corrosion and erosion problems, since the heat transfer surface is less likely to be located in the primary combustion zone;
- Minimal excess air requirements, since the high velocities promote good mixing and combustion efficiency;
- Less dependence on limestone type, since reactivity is improved with the fine particle sizes; and
- Reduced solid waste generation rates, because of lower limestone requirements.

Potential disadvantages are:

- Increased capital costs;
- Greater energy losses due to high pressure drops;
- Increased combustor height compared to a bubbling bed;
- Uncertainty regarding the hot cyclone's ability to adequately separate solids and gases and to resist erosion and corrosion;
- Difficulty in scaling up for large units (over approximately 200 MW), limiting application generally to smaller boiler sizes;
- Less ability to use existing boiler equipment in retrofit applications; and
- Erosion of components subjected to impingement of high velocity particles.

Table 3.2-1 presents the estimated operating data for a 200 MW AFBC. Table 3.2-2 summarizes full-scale utility AFBC demonstration facilities

planned for startup between 1986 and the early 1990's. During retrofit of Northern States Power's Black Dog Unit No. 2, generating capacity was increased from 85 MW up to 125 MW by increasing operating pressures and modifying the turbine-generator as part of the overall retrofit program. Estimated cost was $58 million ($460/kW). Retrofitting AFBC on the four units at Wisconsin Electric's Oak Creek Station is expected to cost an estimated $380 million ($760/kW). The projected output includes recovering 130 MW of generating capacity lost due to coal changes and mechanical limitations on the existing boilers (12). The recovery of generating capacity lost during previous plant derates or expansion in capacity due to overall plant upgrading is one of the key economic benefits from repowering an existing plant with FBC.

TABLE 3.2-1. ESTIMATED OPERATING DATA FOR 200 MW AFBC

Performance at Full Load	
Steam Cycle Heat Rate, Btu/kWh	7724
Boiler Efficiency, percent	87.5
Gross Plant Heat Rate, Btu/kWh	8823
Auxiliary Power, MW	14.9
Net Plant Heat Rate, Btu/kWh	9500
Turbine Generator Gross Output, MW	209.6
Turbine Generator Net Output, MW	194.7
Consumption/Production @ 65% CF	
Coal, 10^3 TPY	521
Limestone @ Ca/S = 2.5, 10^3 TPY	173
Waste Solid, 10^3 TPY	229

Source: Reference 13.

TABLE 3.2-2. FULL-SCALE UTILITY AFBC DEMONSTRATIONS

Utility	Station	New/ Retrofit	Bed Type	Size (MW)	Start-Up
Colorado Ute	Nucla	New	Circulating	110	1987
Montana-Dakota Utilities	Heskett	Retrofit	Bubbling	85	1987
Northern States Power	Black Dog	Retrofit	Bubbling	125	1986
Tennessee Valley	Shawnee	New	Bubbling	160	1988

Pressurized FBC --

Pressurized fluidized bed combustion (PFBC) is similar to AFBC with the exception that combustion occurs under pressure. By operating at pressure, it is possible to reduce the size of the combustion chamber and to develop a combined-cycle or turbocharged boiler capable of operation at higher efficiencies than atmospheric systems. The turbocharged boiler approach recovers most of the heat from the boiler through a conventional steam cycle, leaving only sufficient energy in the gas to drive a gas turbine to pressurize the combustion air. The combined cycle system extracts most of the system's energy through a gas turbine followed by a heat recovery steam generator and steam turbine. As with IGCC (Section 3.1), hot gas cleanup technology is a critical need in combined cycle PFBC development.

The smaller size of a PFBC boiler may be especially attractive in retrofit applications where space is limited. However, PFBC is not as well developed as

AFBC. To date, work on PFBC has been limited mainly to component and pilot-scale testing at facilities in the U. S., England, and Sweden. Commercial PFBC systems have been ordered by two European utilities. There is currently one utility boiler size PFBC unit (75 MW) under construction in the United States (Tidd) (14) with another one planned for 1995 (Philip Sporn). Initial utility PFBC demonstration projects are not expected to be in operation until the early 1990's with earliest commerical availability projected in the 1995-2000 time frame (15).

Applicability

For new plants, AFBC can be applied to nearly any plant site because of the inherently wide design ranges for fuel and operating conditions. Single unit AFBC design is presently limited in size to about 300 MW. At these sizes, unit design can be modularized to reduce construction cost. Construction times of 2-3 years are possible, thus allowing the utility to schedule construction in response to load growth or other requirements.

Retrofitting AFBC at an existing plant requires either 1) installation of a totally new boiler to supply steam to an existing turbine or 2) replacement of the lower furnace section of the boiler while continuing to use portions of the convective heat transfer sections of the existing boiler. Major design questions in making this decision are the layout, condition, and operating limits (e.g., pressure/temperature) of the existing boiler and particulate control system and the availability of additional space if an entire new boiler is installed.

Current AFBC designs are limited to heat release rates of 1.0 million Btu/ft^2. Therefore, an existing boiler with a heat release rate of 1.5 million Btu/ft^2 would either require an AFBC approximately 50 percent larger than the existing boiler plan area or have to be derated to two-thirds of its present capacity (16). Based on this, retrofit of AFBC will likely be limited to boilers with heat release rates below 1.5 million Btu/ft^2. Because

most cyclone-fired and wet-bottom pulverized coal boilers have higher heat releases, these units are not likely candidates for AFBC retrofit. Bubbling beds are preferred for retrofit applications because they require less space and can use more of the existing plant equipment.

Performance

Extensive testing of AFBC has been performed in numerous vendor pilot plants, full-scale industrial AFBC's, the EPRI/Babcock & Wilcox test facility in Alliance, Ohio (rated at 20,000 lb/hr of steam), the 20 MW Tennessee Valley Authority pilot plant at Shawnee Station, and the 15 MW Northern States Power French Island Pilot Plant. The EPRI/B&W facility has unique scaling features for commercial units, such as a high free board zone, to simulate utility boiler residence times and temperatures. Test results have shown overall combustion efficiency of 98 percent and sulfur capture of 90 percent with a calcium-to-sulfur ratio of 2 to 1. NO_x emission levels achievable with AFBC are estimated at less than 0.4 lb/10^6 Btu.

The 20 MW TVA pilot plant has been used to evaluate a number of areas requiring technological developments related to size scale-up and long-term operation of commercial AFBC systems. Objectives of the 20 MW pilot test, cosponsored by TVA and EPRI, included:

- Demonstrate acceptable, long-term operation and performance (efficiency, reliability, emissions control, etc.);
- Develop and demonstrate automatic control and load following capability;
- Develop design correlations for confident scaling to commercial size boilers;
- Demonstrate the environmental control capability of the unit as a basis for assessing the environmental acceptability of AFBC on a commercial scale;
- Develop the capability to specify and design critical auxiliary equipment, especially the coal, limestone, and recycle feed systems;

- Test and evaluate materials of construction for the boiler and its auxiliaries; and
- Train engineers and operators for future units.

Performance results were favorable. Combustion efficiency, sulfur capture, and NO_x emissions have been well within expectations. The coal feed systems were the largest problem. The underbed feed system demonstrated high combustion and sulfur control efficiencies, but required a more complex feed system design and operated less reliably. The overbed feed system demonstrated high equipment reliability but less acceptable process efficiencies, especially when the coal contained significant (15-20 percent) fines. Remedies for these problems were investigated prior to the final design of the 160 MW TVA demonstration project.

Flyash recycle is another key design factor for sulfur capture and combustion performance optimization. As previously mentioned, combustion efficiency and limestone utilization increase with greater recycle. Other methods for improving limestone utilization include reinjection of sulfated limestone after pulverizing and hydration.

Cost

Analysis of system economies for new plants have generally found FBC to be cost competitive with pulverized coal firing combined with limestone FGD. Cost estimates by EPRI for a conventional boiler, AFBC, and PFBC are presented in Table 3.2-3 for a new plant.

The following table presents the range of values estimated using IAPCS for capital and annualized costs of new AFBC plants. The lower capital cost is for a large unit with a high capacity factor, while the higher capital cost is for a small unit with a low capacity factor. Figures 3.2-2 and 3.2-3 show these costs as a function of boiler size and heat rate. The reason for lower costs in the lower heat rate value cases is due to the lower preproduction and inventory costs.

	Range
Capital Cost ($/kW)	1,132 to 2,382
Annualized Cost (mills/kWh)	77.0 to 182.4

TABLE 3.2-3. COMPARATIVE ECONOMICS OF NEW 500 MW CONVENTIONAL AND FBC POWER PLANTS[a] (1988 dollars)

	Conventional Boiler w/ LS FGD	AFBC	Turbocharged PFBC[b]	Combined-Cycle PFBC
Capital Cost, $/kW	1,233	1,184	1,247	1,394
Heat Rate, Btu/kWh	10,060	10,000	9,703	8,980
Total Cost, Mills/kWh				
Constant $	37.3	23.0	36.9	39.4
Current $	64.2	62.4	63.5	67.8

[a]Based on a 500 MW boiler with 3.5 percent sulfur coal at $1.55/million Btu and capacity factor of 65 percent.

[b]Scaled from 250 MW based on a scaling factor of 0.8; circulating bed design.

Source: Reference 9.

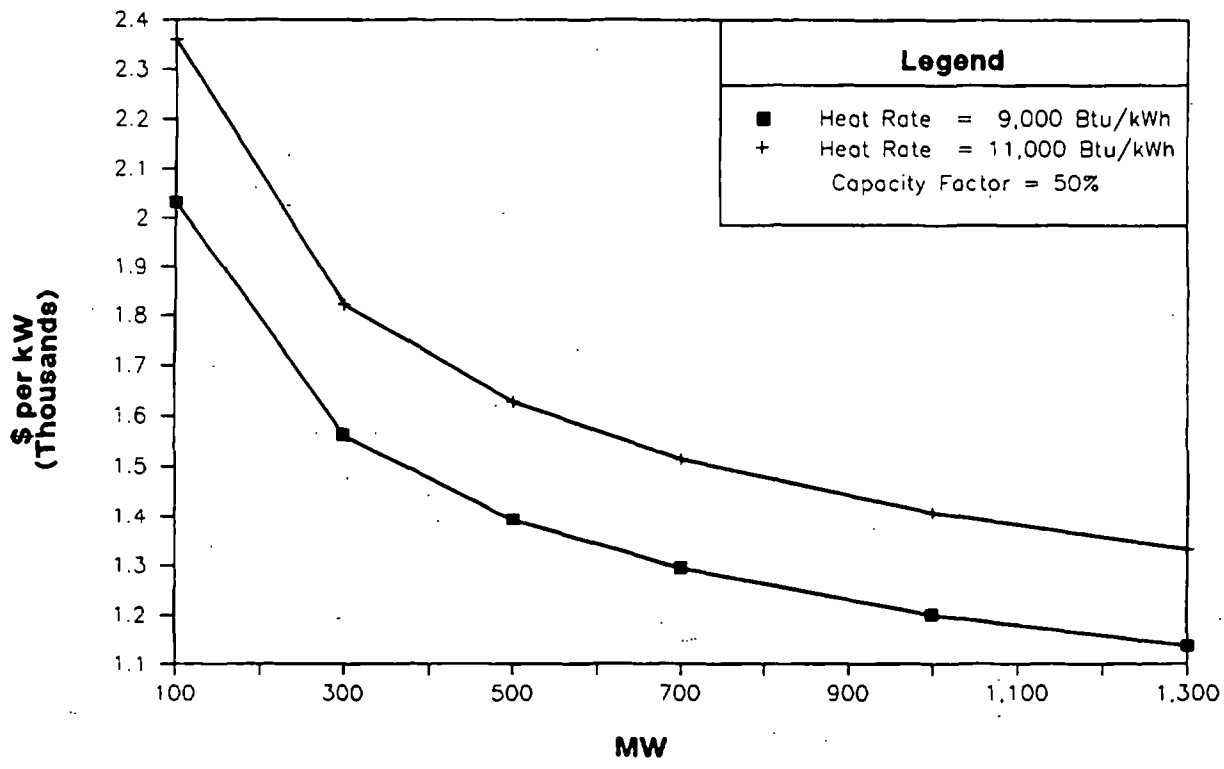


Figure 3.2-2. AFBC - Capital Cost, Current 1988 Dollars

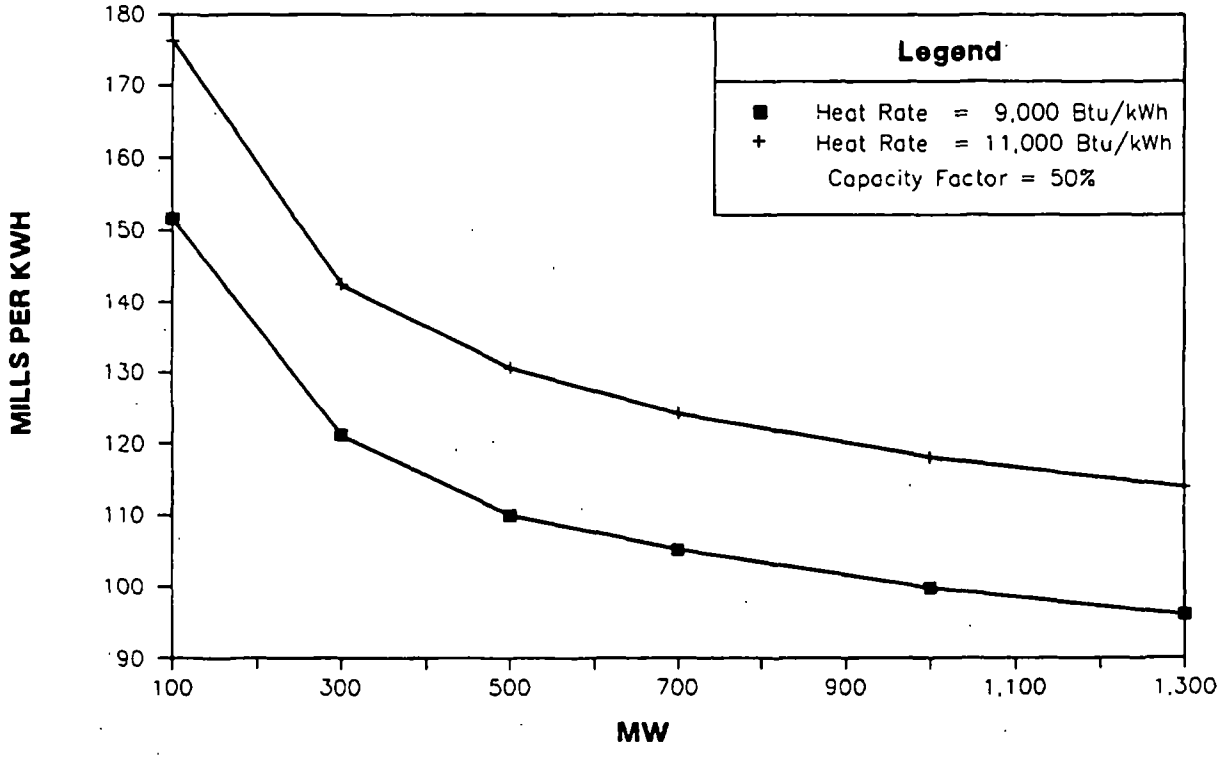


Figure 3.2-3. AFBC - Levelized Annual Cost, Current 1988 Dollars

Appendix E contains tables of costs as a function of boiler size, capacity factor, and boiler heat rate. Detailed costs are not available at this time for PFBC, and therefore these costs are not presented.

3.3 POST-COMBUSTION NOx CONTROL

Description

Selective Catalytic Reduction--

Selective catalytic reduction (SCR) is a flue gas treatment process for removal of nitrogen oxides (NO_x). In this process, gaseous ammonia diluted with either air or steam is injected into the flue gas upstream of the SCR catalyst. The ammonia/flue gas mixture then enters the catalyst where the ammonia reduces NO_x to elemental nitrogen and water (17).

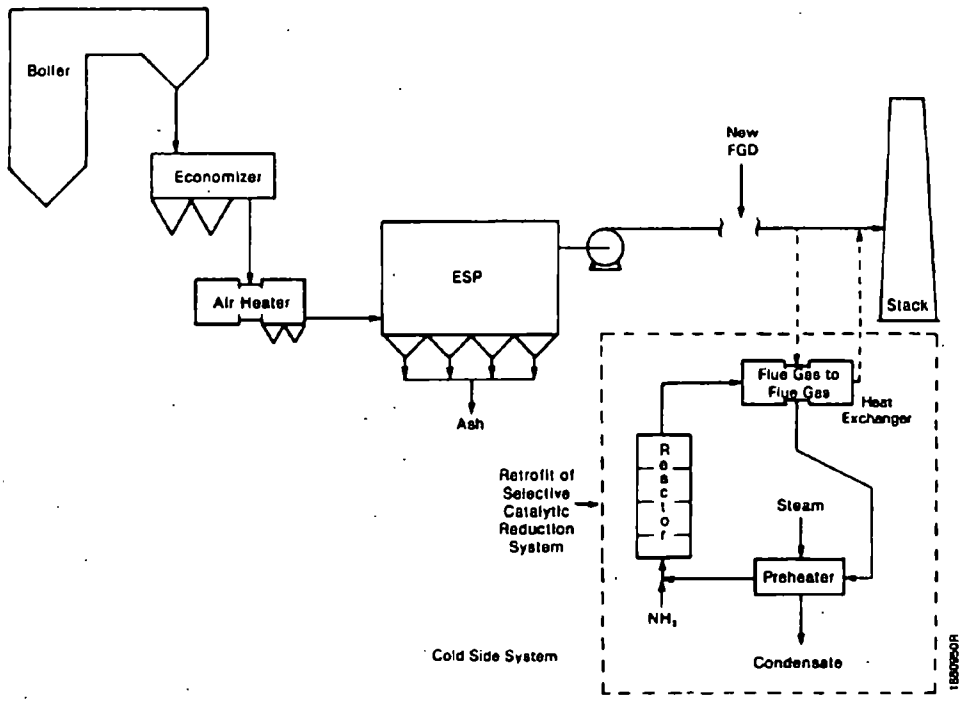
$$4NO + 4NH_3 + O_2 \longrightarrow 4N_2 + 6H_2O$$

$$2NO_2 + 4NH_3 + O_2 \longrightarrow 3N_2 + 6H_2O$$

Because the NO_x composition of the flue gas from combustion sources is primarily NO, the stoichiometric NH_3/NO_x mole ratio is approximately 1:1. An operating temperature of 570-750°F is required to catalyze NO_x reduction. Below this temperature range, the reaction between ammonia and NO_x slows significantly. Above this temperature range, ammonia is oxidized to NO_x and the catalyst can be damaged thermally.

There are two possible SCR configurations. The first design is a hot-side or high-dust SCR where the SCR system is located between the economizer and air preheater. The second design is a cold-side or low-dust SCR where the SCR is located typically downstream of the particulate control device and possibly downstream of the FGD system. Figure 3.3-1 presents both SCR configurations. An economizer bypass can be incorporated into the system to control low catalyst temperature excursions.

SCR catalyst formulations are typically based on oxides of titanium and vanadium. Other active metals such as platinum, palladium, and copper can also be used. In addition to reducing NO_x to N_2, the catalyst also promotes the oxidation of SO_2 to SO_3 which in turn reacts with NH_3 and H_2O to form ammonium bisulfate (NH_4HSO_4) as the flue gas cools. Ammonium bisulfate deactivates the SCR catalyst by coating active catalyst sites and the air



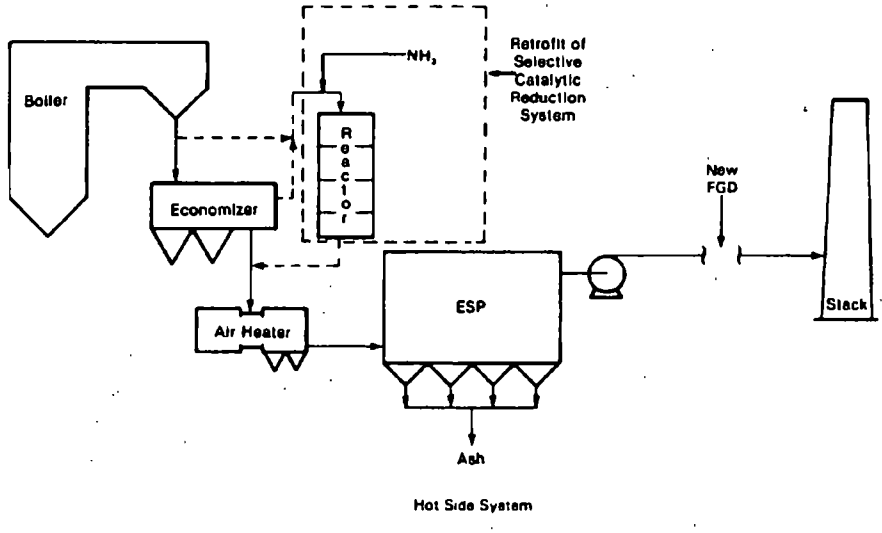


Figure 3.3-1. Possible SCR Configurations.

preheater elements reducing heat transfer efficiency and increasing metal corrosion.

Catalyst elements can be produced in two general forms: composed solely of catalyst material (homogeneous) or composed of either a ceramic or metallic substrate that is coated with catalyst material (heterogeneous). Each type of catalyst element has distinct advantages. For example, the homogeneous element will not lose activity even if the outer layer is lost due to erosion by flyash; however, eventually the physical strength can be affected. To overcome catalyst degradation, the inlet face of the catalyst can be protected with a hardened inert material. Coated catalyst elements also have the potential for erosion.

The coated element will not be structurally weakened, but catalytic activity can be lost. Metallic substrates are more resistant to breakage and cause less pressure drop than a ceramic substrate due to thinner walls. Alternatively, the ceramic material has better catalyst bonding properties, is acid resistant, and lightweight.

Honeycomb, plate, and pipe catalyst supports are preferred for use with coal-fired applications. A number of tests were performed on honeycomb and plate catalysts (18). The honeycomb shape is the most common for reasons of strength and ease of handling. The honeycomb shape consists of a square block with parallel channels passing through it. The block is typically 6-20 inches on a side and up to 40 inches in length. The channels can be either square, hexagonal, or triangular in shape. Plugging by particulates is typically not a major problem when a vertical downflow reactor arrangement and regular soot blowing are used.

Selective Non-Catalytic Reduction--

Selective non-catalytic reduction (SNR) is an alternate approach for post-combustion reduction of NO_x to N_2. The primary SNR technologies involve injection of ammonia or urea (decomposes to ammonia) into the convective section of the boiler at 1600-2000°F; no catalyst is used. At temperatures above and below this range, ammonia is oxidized to NO_x or exits as unreacted NH_3, respectively. In general, because of the short residence time at these temperatures in the convective section and the difficulty of thorough ammonia/flue gas mixing, NO_x reductions are lower (40-80 percent)

and ammonia emissions are higher than with SCR. Because a catalyst is not required, SNR costs are significantly lower than for SCR. However, because of the lower performance level and limited experience with SNR on coal-fired boilers, the rest of this section is limited to SCR technology.

Applicability

Since the SCR process requires flue gas at a specific temperature, it is simpler to apply it to new boilers than to existing boilers. In new boilers, the SCR system is normally placed between the economizer and air preheater as economizer exit temperatures of utility boilers are typically in the range of optimum SCR operation. Retrofit installations will require flow modifications and additional ductwork to divert and return flue gas to the existing flue gas handling system. Alternate arrangements include installation of the reactor vessel downstream of a hot-side particulate control system or, by combining with flue gas reheat, downstream of the existing emission controls for SO_2 and particulates. This last approach may be the most suitable for many retrofits due to boiler limitations.

SCR is considered a commercial technology; however, it has not been widely applied in the U. S. More than 30,000 megawatts of coal-fired utility boiler SCR systems have been operated in Japan and West Germany. Three coal-fired pilot SCR demonstration projects have been conducted in the U. S. The only full-scale U.S. application to a utility boiler is a gas- and oil-fired demonstration unit at Southern California Edison's Huntingdon Station. There are also three industrial boilers and a number of gas turbines and process heaters using SCR for NO_x control, all in California.

Most of these applications involve very low-sulfur fuels; this applicability of SCR technology to high-sulfur fuels has not been commercially demonstrated to date. A primary concern is the impact of high SO_3 levels on catalyst and air preheater plugging and corrosion due to formation of ammonium bisulfate. Air preheater plugging is most severe in units which fire high-sulfur fuels and remove particulates upstream of the NO_x reactor. When particulates are removed downstream, plugging problems are significantly reduced. It is believed that the particulates produce a sandblasting effect that cleans the air preheater elements and that some of

the NH_4HSO_4 condenses on the flyash particles rather than the air preheater. Air preheater plugging can also be reduced by limiting NH_3 slip to about 3 to 5 ppm (NH_3 slip is the reacted NH_3 exiting the SCR reactor) and redesigning the air heater. However, reducing NH_3 slip will require increasing catalyst volume if a specific NO_x removal rate is to be maintained. Off-line water washing of the air preheater may also be required to minimize pressure drop; soot blowers are generally ineffective at removing NH_4HSO_4 deposits from the air heater on-line.

NH_3 from an SCR reactor does not impair FGD SO_2 removal performance. However, in some cases, the wastewater may require treatment to remove nitrogen (nitrate) compounds. An unresolved concern is the potential for NH_3 slippage or NH_4HSO_4 to affect the performance of a downstream particulate control device (e.g., by increasing pressure drop across a downstream fabric filter or ESP).

Performance

SCR is capable of 80 to 90 percent NO_x reduction. The primary variables which determine the amount of NO_x removed are the amounts of ammonia and catalyst used; increasing ammonia and catalyst yields higher removals. However, increasing the amount of ammonia injected also increases ammonia slip. For this reason, the NH_3 slip limit set for the initial NO_x concentration is the key catalyst sizing parameter, rather than the required NO_x reduction.

Catalyst bed design is usually specified in terms of a space velocity or area velocity. The space velocity is the flue gas volumetric flow rate divided by the catalyst bulk volume, while area velocity is flue gas volumetric flow rate divided by the active catalyst area. High NO_x reductions require low space (or area) velocities and, hence, large quantities of catalyst. Low velocities are also required to control the amount of NH_3 slip. To reduce NH_3 slip to 3-5 ppm levels, space velocities in the range of 2,000-2,500 hr^{-1} are generally required. Furthermore, NH_3 slip increases dramatically at NH_3:NO_x injection ratios over 0.8 for high space velocities and 0.9 for low space velocities. This constraint contributes in holding SCR to a practical NO_x removal limit of about 80

percent. The exact relationships between velocity and NO_x reduction and NH_3 emissions, however, are specific to the proprietary catalysts used by the various SCR manufacturers.

Cost

The capital cost of an SCR system is a function of several factors including NO_x reduction level, flue gas flow rate, fuel type and sulfur content, and retrofit/new installation (19). Operating costs are primarily a function of catalyst life.

The following table presents the range of values for capital cost, annualized cost, and cost per ton of pollutant removed for hot side SCR systems using IAPCS. The lower costs are for a large unit with a low retrofit factor and the higher costs are for a small unit with a high retrofit factor. Figures 3.3-2 through 3.3-4 show these costs as a function of boiler size and catalyst life.

	Range
Capital Cost ($/kW)	85.4 to 240.6
Annualized Cost (mills/kWh)	6.1 to 39.4
Cost Per ton of NO_x Removed ($/ton)	1,856 to 12,334.1

Estimates of catalyst life vary from 1-5 years. For the purpose of this analysis, 1 and 3 year catalyst life is assumed. Appendix G contains tables of costs as a function of boiler size, retrofit difficulty, capacity factor, and catalyst life.

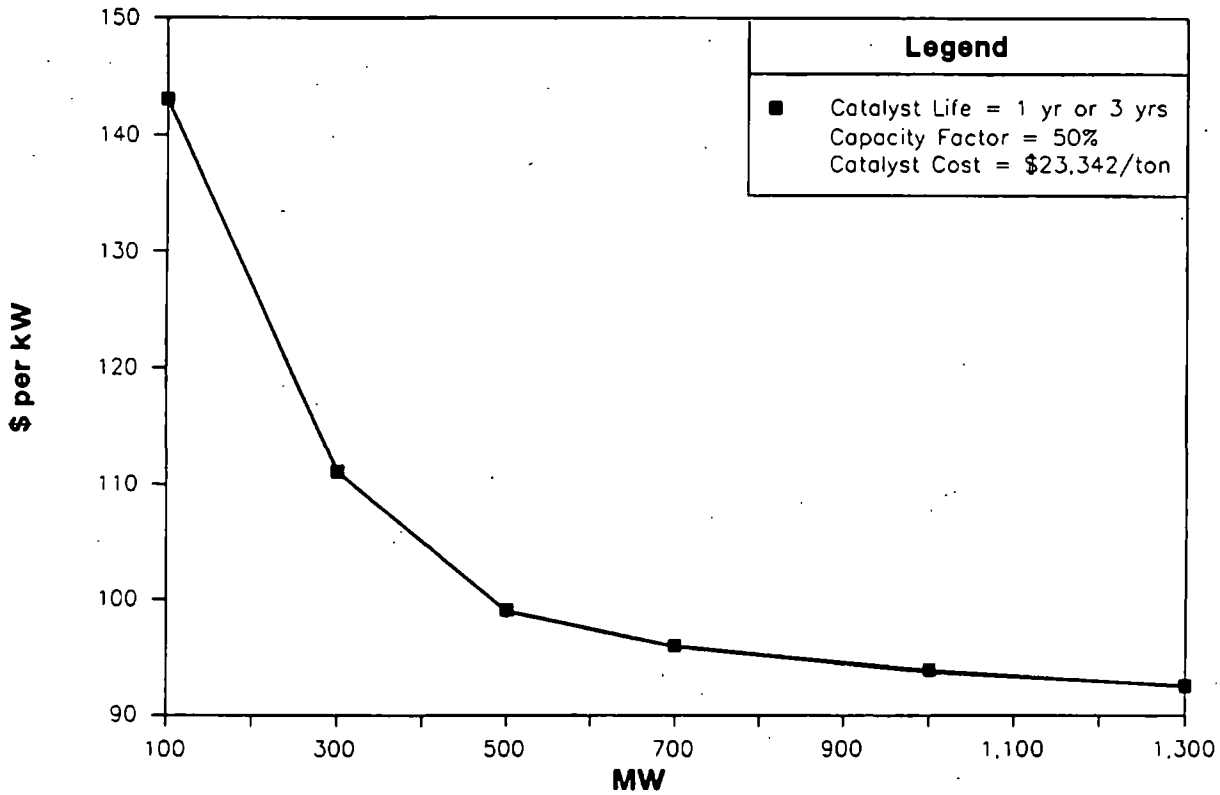


Figure 3.3-2. Selective Catalytic Reduction - Capital Cost Current 1988 Dollars

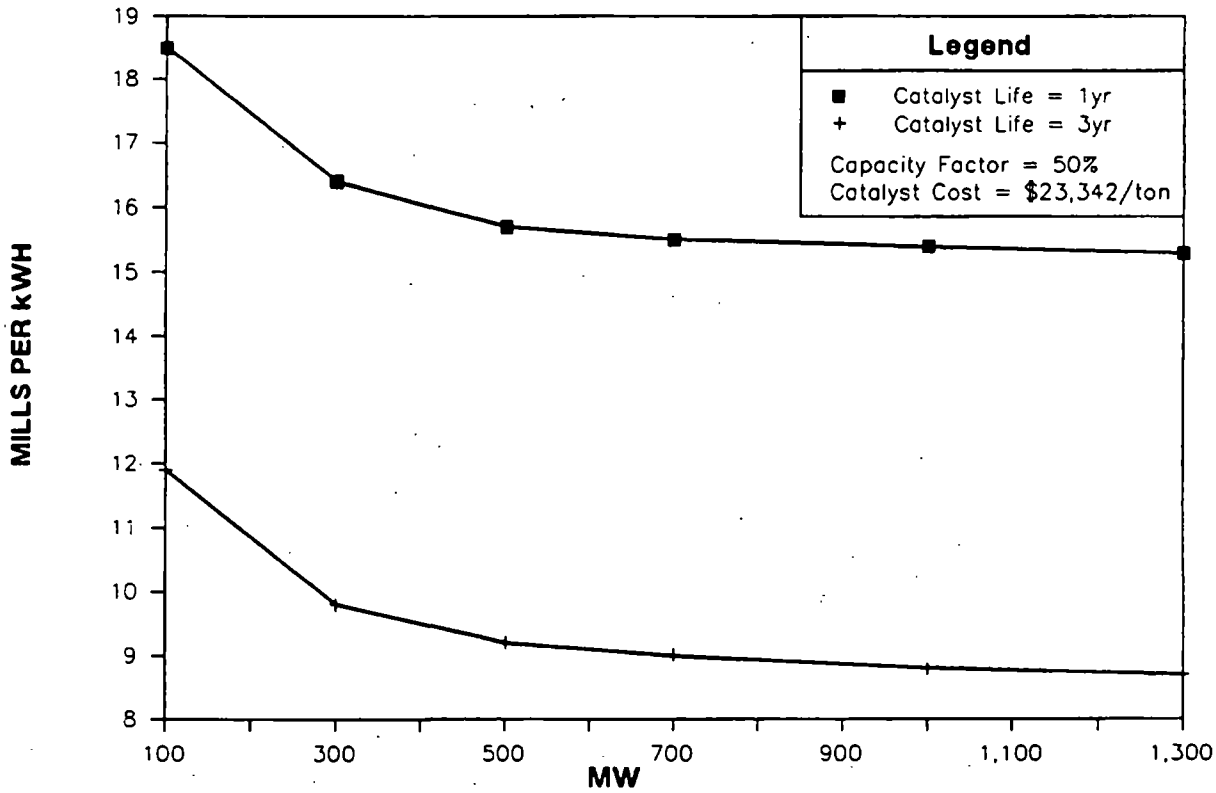


Figure 3.3-3. Selective Catalytic Reduction - Levelized Annual Cost, Current 1988 Dollars

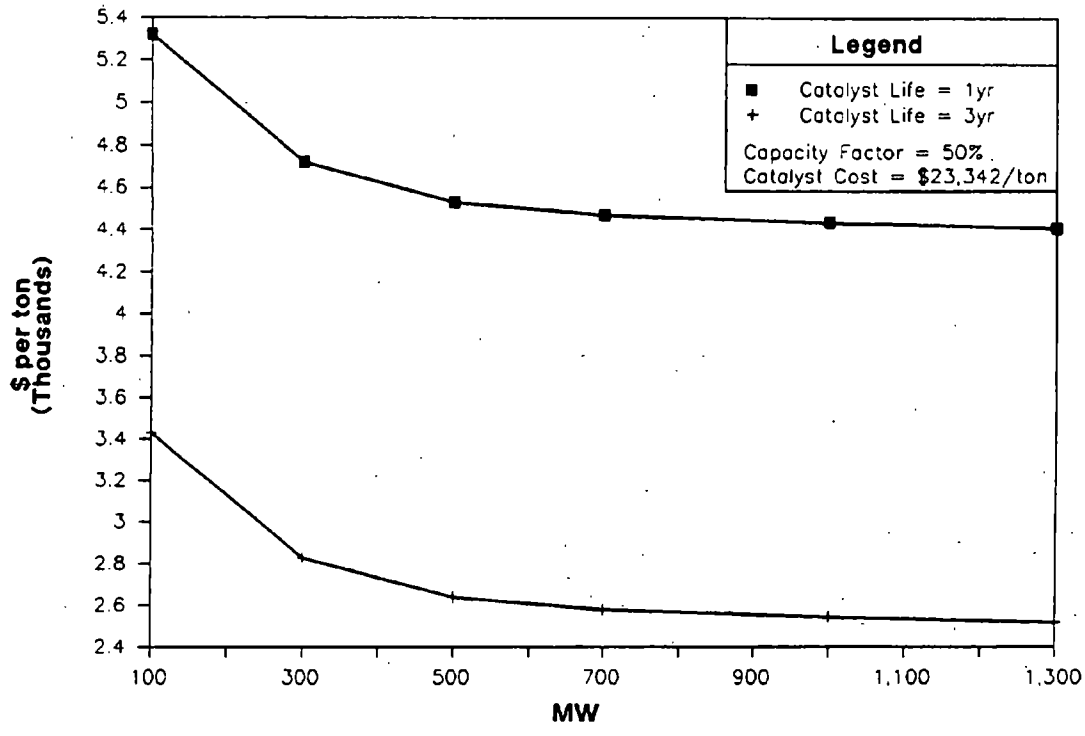


Figure 3.3-4. Selective Catalytic Reduction - Cost Per Ton of NOx Removed, Current 1988 Dollars

3.4 FURNACE SORBENT INJECTION

Description

Furnace sorbent injection (FSI, or LIMB) reduces SO_2 emissions by injecting powdered limestone or hydrated lime into the upper furnace of a coal-fired boiler. These sorbents decompose at high temperature to form calcium oxide which reacts with SO_2 to form calcium sulfate ($CaSO_3$) (20). Limestones ranging from calcitic (mainly $CaCO_3$, very low in $MgCO_3$) to dolomitic (containing roughly equal amounts of Ca and Mg) have been tested. Atmospheric and pressure-hydrated limes have also been used. The extent of SO_2 removal that can be achieved depends on the flue gas composition, temperature, and quench rate at the point of sorbent injection; sorbent composition and surface area; and the calcium-to-sulfur ratio (Ca/S) and degree of mixing between the sorbent and SO_2 in the flue gas (21).

The resulting calcium sulfate and unreacted sorbent are collected with the flyash in a baghouse or electrostatic precipitator (ESP). Collected sorbent and flyash can be either recycled to the furnace to increase sorbent utilization or discharged to the solid waste disposal facilities. Figure 3.4-1 is a generalized diagram of the FSI process.

The advanced silicate (ADVACATE) process (22) can be added to FSI (or stand alone process) for better SO_2 removal efficiency. Preliminary pilot tests on ADVACATE system have resulted in 90+ percent SO_2 removal with a calcium to sulfur ratio near 1.0. In this process cyclones may be used to knock out coarse dust upstream of the ESPs. Part of the high calcium flyash in the front section of the ESPs, silicate, is slurried under temperature and pressure. Silica in the fly ash dissolves and reacts with calcium to form high surface area silicates. The slurry is mixed with the remaining high calcium fly ash to form a damp powder, which is then injected into the ductwork.

Performance

The most important variable affecting the calcination and sulfation reactions is the gas temperature at the point of sorbent injection. As shown in Figure 3.4-2, a peak in SO_2 removal occurs for most sorbents in the range

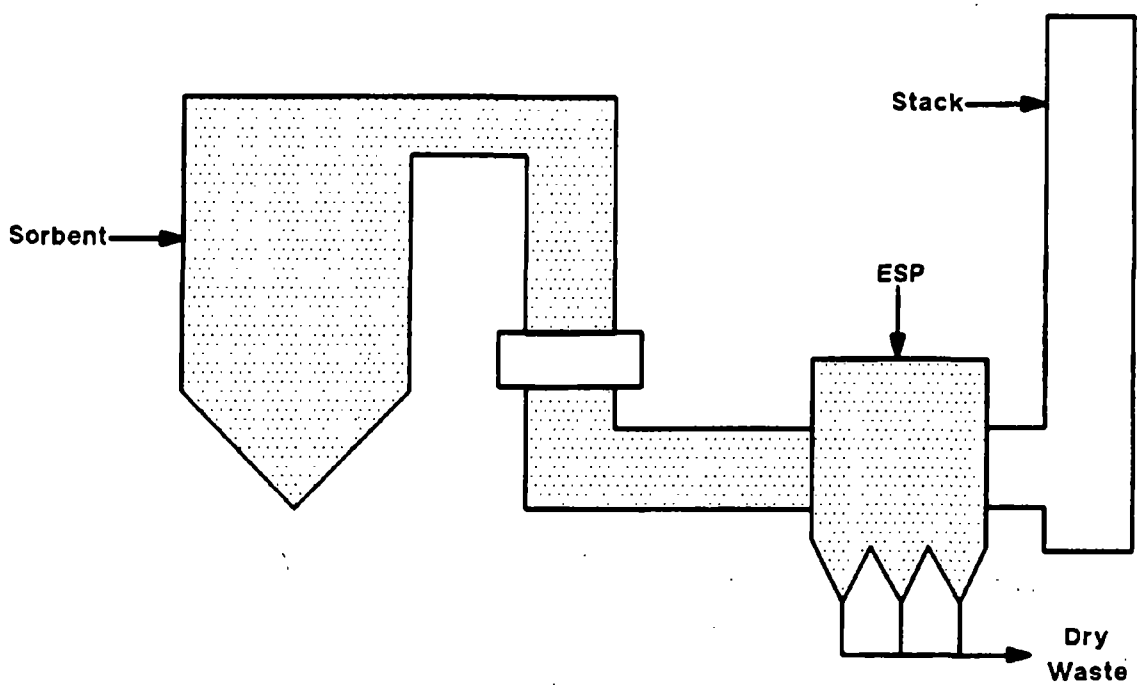


Figure 3.4-1. Simplified schematic of furnace sorbent injection.

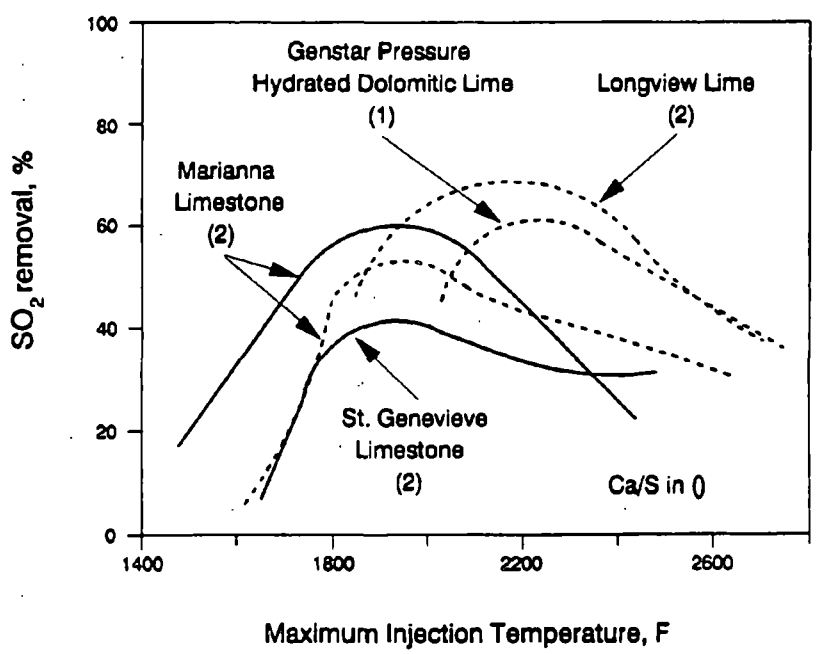


Figure 3.4-2. Peak sorbent reactivity as a function of temperature (23).

between 1800°F and 2200°F. In most boilers, these temperatures are found in the upper furnace. Because limited residence time is available in this zone to allow thorough mixing of the sorbent and combustion gases, rapid distribution of the sorbent over the full boiler cross-section is essential. At temperatures higher than 2200°F, sorbent deactivation (due to sintering of the sorbent surface and instability of the reaction products) lowers SO_2 removal. At temperatures below 1800°F, reaction rates for both calcination and sulfation are significantly reduced.

The sorbent quench rate (i.e., the rate of change in the flue gas temperature after the point of sorbent injection) also affects SO_2 removal, with SO_2 removal decreasing as quench rate increases (24). As quench rates increase, sorbent residence time in the 1800-2200°F window is correspondingly reduced. Expected SO_2 capture as a function of quench rate, Ca/S ratio, and sorbent are shown in Table 3.4-1. This factor is especially significant for cyclone-fired boilers for which quench rates in the upper furnace can approach 1500°F/sec.

Sorbent characteristics such as chemical composition and surface area also affect the sorbent injection rate needed to achieve good SO_2 removal. The surface area of several sorbents evaluated in a pilot-scale study are shown in Figure 3.4-3. Pressure-hydrated dolomitic lime was the most effective for removing SO_2; limestone was the least effective. Techniques that have been studied for improving sorbent surface area include thermal pretreatment and use of chemical promoters. Although these techniques have increased measured surface area, pilot-scale combustion tests have not conclusively demonstrated their effectiveness in improving SO_2 removal.

Sorbent residence time and distribution in the flue gas also impact SO_2 removal. As shown in Figure 3.4-4, longer residence times give the sorbent and SO_2 more time to react and increase SO_2 removal. However, for retrofit applications, the residence time of the boiler is already established and cannot be readily varied. Optimal location, number, and type of injection nozzles can maximize mixing of the flue gas and sorbent.

TABLE 3.4-1. SO_2 CAPTURE AS A FUNCTION OF Ca/S RATIO, QUENCH RATE, AND SORBENT

	Ca/S			
	1	2	3	4
Quench rate = 900[b]				
limestone	15-16	26-29	35-43	42-56
hydrate	19-20	35-40	53-56	69-70
CPH[c]	27-33	48-54	68-71	88-94
Quench rate = 700				
limestone	17-19	29-31	38-44	45-57
hydrate	24-26	40-46	57-62	73-76
CPH	33-38	54-62	75-79	92-95
Quench rate = 500				
limestone	18-22	31-33	41-45	48-58
hydrate	28-32	45-52	61-68	77-82
CPH	40-43	60-70	81-89	95+
Quench rate = 300				
limestone	19-25	32-36	44-46	51-59
hydrate	33-38	49-58	66-74	82-88
CPH	46-48	67-78	88-95	95+

[a] SO_2 capture is expressed as a percentage.

[b] Quench rate is expressed as ^{o}F per second for the sulfation "window" of $2200^{o}F$ to $1600^{o}F$.

[c] CPH = calcitic pressure hydrate.

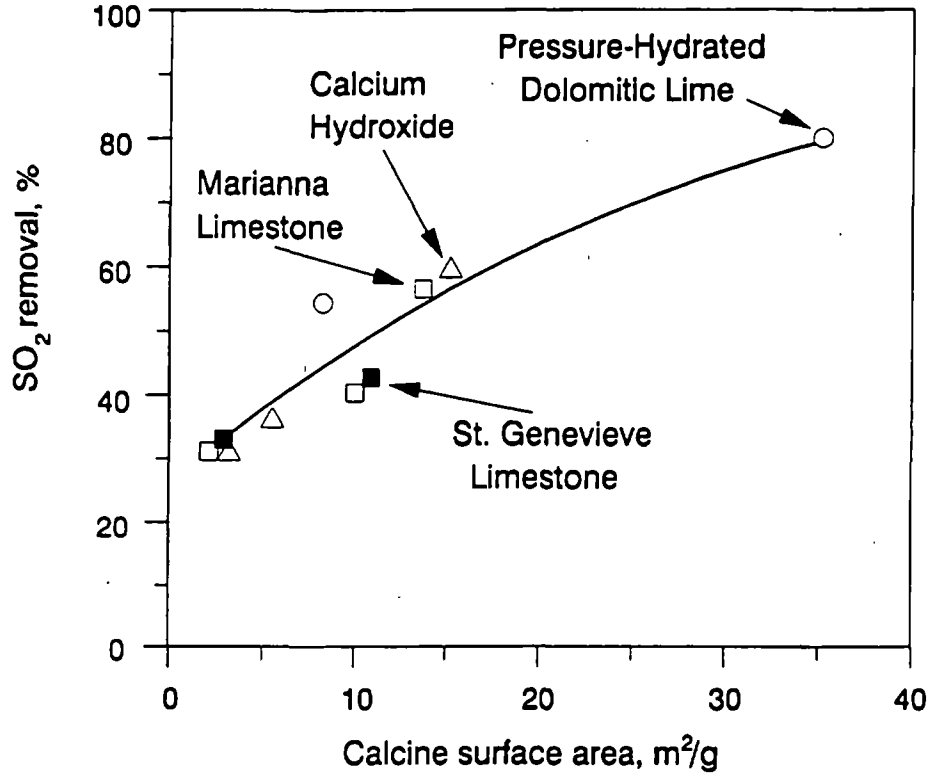


FIGURE 3.4-3. SO_2 removal as a function of calcine surface area (23).

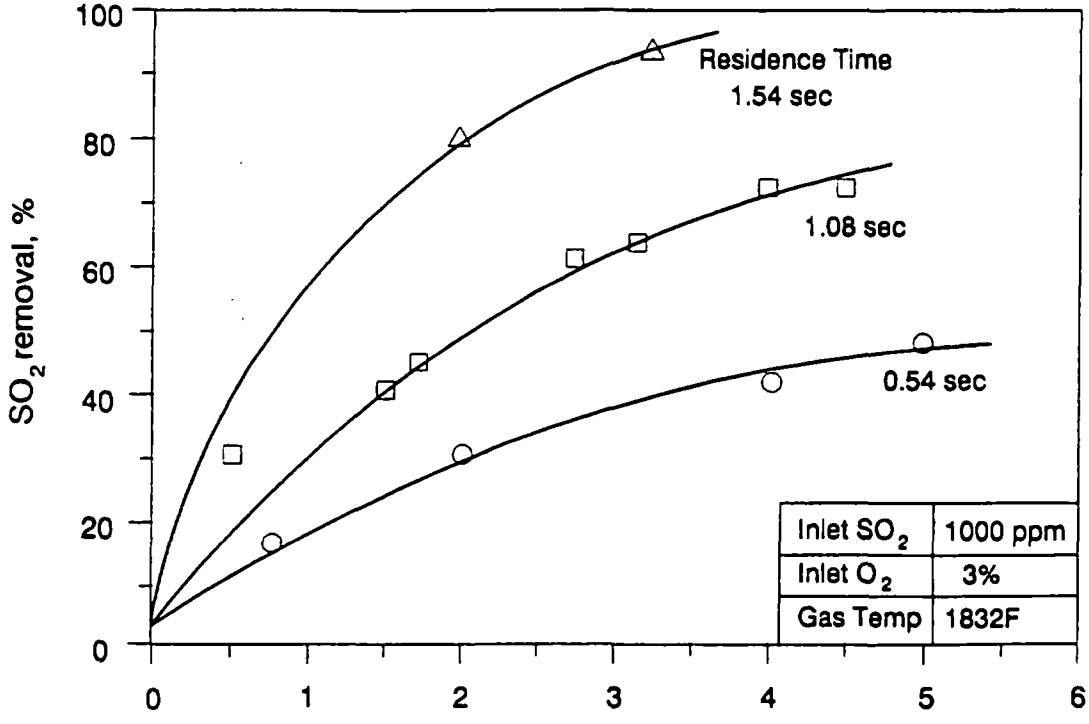


FIGURE 3.4-4. SO_2 removal as a function of Ca/S ratio and residence time using limestone (23).

Three commercial-scale utility projects in the U. S. are demonstrating furnace sorbent injection on eastern bituminous coal-fired boilers. The first project, sponsored by EPRI, U. S. EPA, and Richmond Power and Light, involves a 61 MW tangentially-fired boiler, Whitewater Valley No. 2 at Richmond Power and Light. The objective of this project is to evaluate different combinations of sorbents and coals (with varying sulfur contents), to assess process performance variables and determine retrofit control costs. Testing began in 1987 with sorbents including hydrated lime, limestone, dolomite hydrated lime, and "promoted" lime sorbent (25). Preliminary results of short-term tests indicate that sulfur dioxide emissions can be reduced by 25-40% at calcium to sulfur ration of 2. A flue gas humidification system was designed and installed to counteract degradation of ESP performance resulting from use of sorbent injection.

The second demonstration is a 105 MW wall-fired unit, Edgewater No. 4, owned and operated by Ohio Edison. This has been funded by the U. S. EPA, DOE, and the Ohio coal Development Office. Testing that began in 1987 showed that the technology can exceed its sulfur dioxide control goal of 50% at a calcium to sulfur ratio of 2 (26). The demonstration did encounter problems with downstream particulate control because of the high resistivity ash produced during sorbent injection. A full-scale flue gas humidification system was added in 1988 to reduce ash resistivity and increase overall sulfur dioxide removal.

The U. S. EPA and Combustion Engineering have begun the third demonstration--a program to demonstrate furnace sorbent injection on a tangential, coal-fired utility boiler at Virginia Power's Yorktown Unit No. 2. The overall objective is to achieve a significant reduction (approximately 50%) in sulfur dioxide while minimizing any negative effects on boiler performance during long-term operation of an integrated sorbent injection system and a low-NO_x firing system (27). For moderate calcium to sulfur ratios (2-3:1), SO_2 reductions of 40-53 percent were obtained using calcium hydroxide and of 25-42 percent were obtained using limestone (28).

Two Canadian utilities (Saskatchewan Power and Ontario Hydro) also are conducting full-scale furnace sorbent injection tests on lignite and medium-sulfur bituminous coals, respectively. Using a limestone slurry, the

Ontario Hydro demonstration system was able to remove up to 70% of the sulfur dioxide at a calcium to sulfur ratio of 2.5 to 3.

Baltimore Gas and Electric, Atlantic Electric, the Electric Power Research Institute, and Babcock and Wilcox have co-funded a program to evaluate the potential of dry sorbent injection technology on cyclone-equipped units (29). Upper-furnace sorbent injection for sulfur dioxide capture has been examined using a 6 million Btu/hr cyclone-equipped boiler simulator. The major operating problems which resulted was a significant increase in opacity from 10 to 60 percent.

Applicability

FSI can be applied to both new and retrofit applications. However, under existing utility NSPS, application to new boilers will be limited to units burning low-sulfur coals where 70 percent SO_2 removal is acceptable. Retrofit of FSI technology into an existing boiler will require installation of additional equipment as well as modifications to the boiler and particulate control device. Additional equipment includes sorbent preparation, storage, handling, and injection systems; particulate control modifications; and a new or expanded ash handling system.

FSI can increase the particulate loading in the boiler and particulate control device by two- to three-fold, depending on the Ca/S ratio and the coal sulfur and ash content. The increase in particulate loading as a function of Ca/S ratio and coal sulfur content for a typical 10 percent ash coal is shown in Figure 3.4-5.

Fouling and slagging of boiler tubes may become excessive due to increased solids loading and, particularly for acidic coal ashes, reduced ash fusion temperatures. These deposits reduce water wall heat transfer, and as a result, the superheater and reheater steam temperature. Increased dust loads may also increase erosion. Sootblower upgrading by increasing the number of blowers, blowing pressure, and frequency should reduce the severity of heat transfer problems. Finned-tube economizers probably will need to be replaced with larger bare-tube economizers with an in-line arrangement. In some cases, derating of the boiler may be necessary even after these modifications.

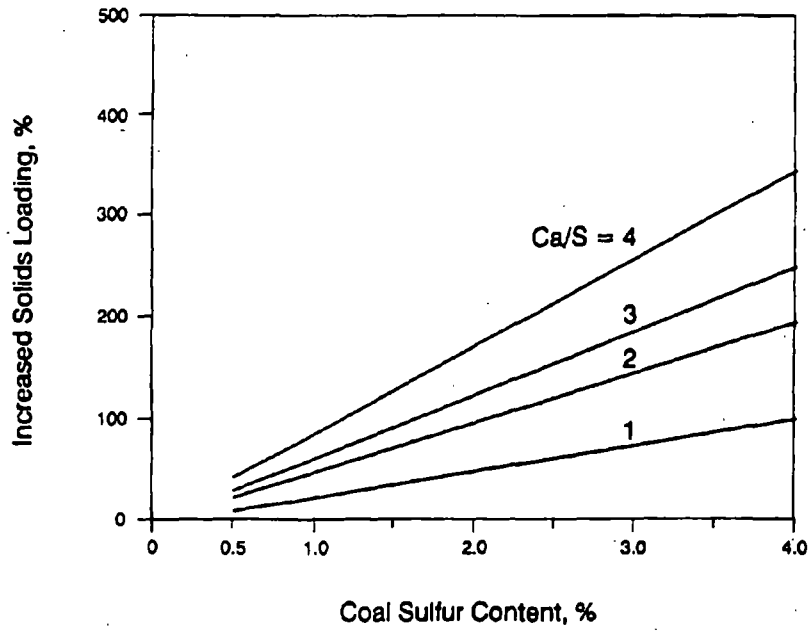


FIGURE 3.4-5. Increase in solids loading as a function of coal sulfur content and Ca/S ratio for a typical 10% ash coal (17).

FSI may also affect performance of the existing particulate control device (generally an ESP) due to the increased solids loading, the high alkalinity of the sorbent, and low SO_3 level in the flue gas (30). Pilot-scale tests indicate that flue gas humidification can return ESP performance to near pre-sorbent injection levels for most ESP's (26). For baghouses, additional bag area, more frequent cleaning, and additional induced draft fan capacity may be needed to handle the increased particulate loading and pressure drop. Also, the capacity of the ash handling system may need to be increased to handle the extra solids. Because of the large amount of unreacted lime present in the solids, conversion of wet ash handling systems to dry handling will also be necessary.

On the positive side, the lower SO_2/SO_3 content of the flue gas and increased alkali in the flue gas particulate may improve boiler efficiency by allowing lower boiler exit temperatures without cold-end acid corrosion.

Cost

Estimated total capital costs for retrofitting FSI on an existing power plant are much lower than for conventional FGD technologies (31). Depending on the design of the boiler and existing particulate control system, total capital costs range from $35-50/kW if the existing particulate control system can be used without significant upgrading. If the existing ESP must be replaced by a new fabric filter (e.g., due to space constraints that limit installation of additional ESP plate area), total capital costs can approach $150/kW.

Because of FSI's relatively low captial cost, total busbar costs are very sensitive to operating costs associated with sorbent purchase and disposal. Major factors affecting these costs are the Ca/S ratio required to achieve a given SO_2 reduction, coal sulfur content, and unit costs for sorbent and waste disposal. Also, because of the relatively low capital cost of FSI, the cost per ton of SO_2 removed is relatively uniform over a wide range of percent removals and plant sizes.

The following table presents the range of values for capital cost, annualized cost, and cost per ton of pollutant removed using IAPCS for FSI with humidification and upgrading of the existing ESPs (SCA = 300). A sulfur dioxide removal efficiency of 70% was assumed based on the use of humidification to improve both ESP performance as well as SO_2 capture. The lower costs are for a large unit burning low sulfur coal and the higher costs are for a small unit burning high sulfur coal. Figures 3.4-6 through 3.4-8 show these costs as a function of boiler size and coal sulfur content.

	Range
Capital Cost ($/kW)	23.6 to 81.4
Annualized Cost (mills/kWh)	3.6 to 21.8
Cost Per ton of SO_2 Removed ($/ton)	516.1 to 4,995

Appendix H contains tables of costs as a function of boiler size, capacity factor, coal sulfur and SO_2 reduction.

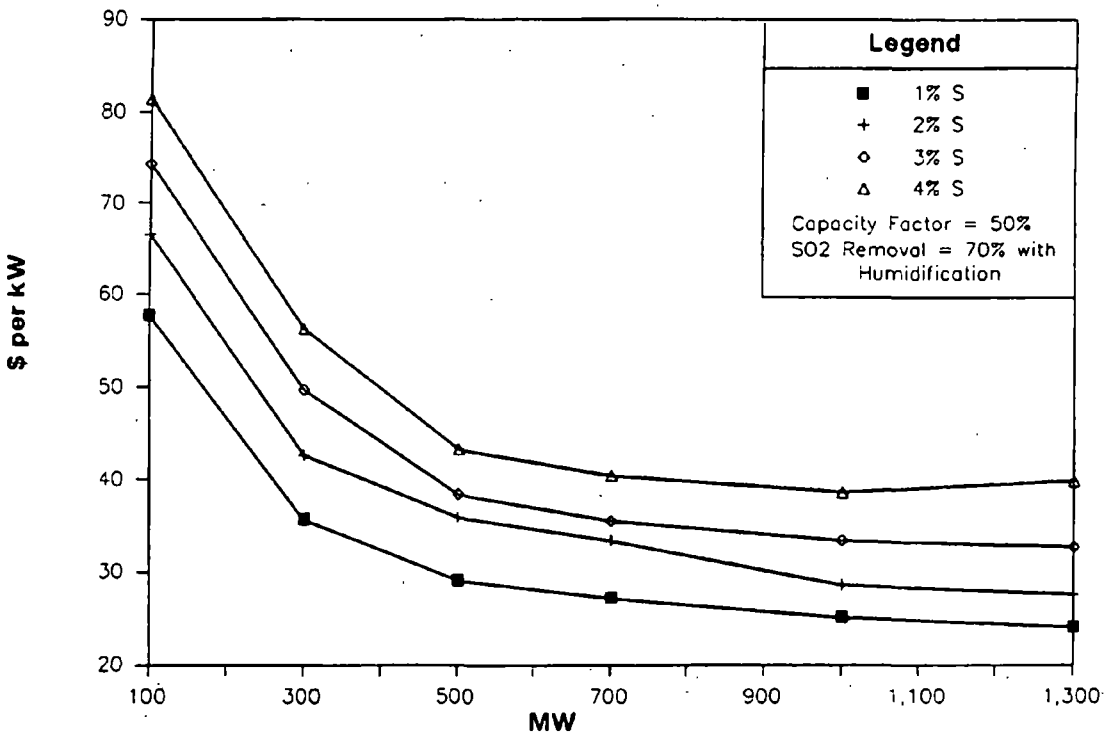


Figure 3.4-6. Furnace Sorbent Injection - Capital Cost Current 1988 Dollars

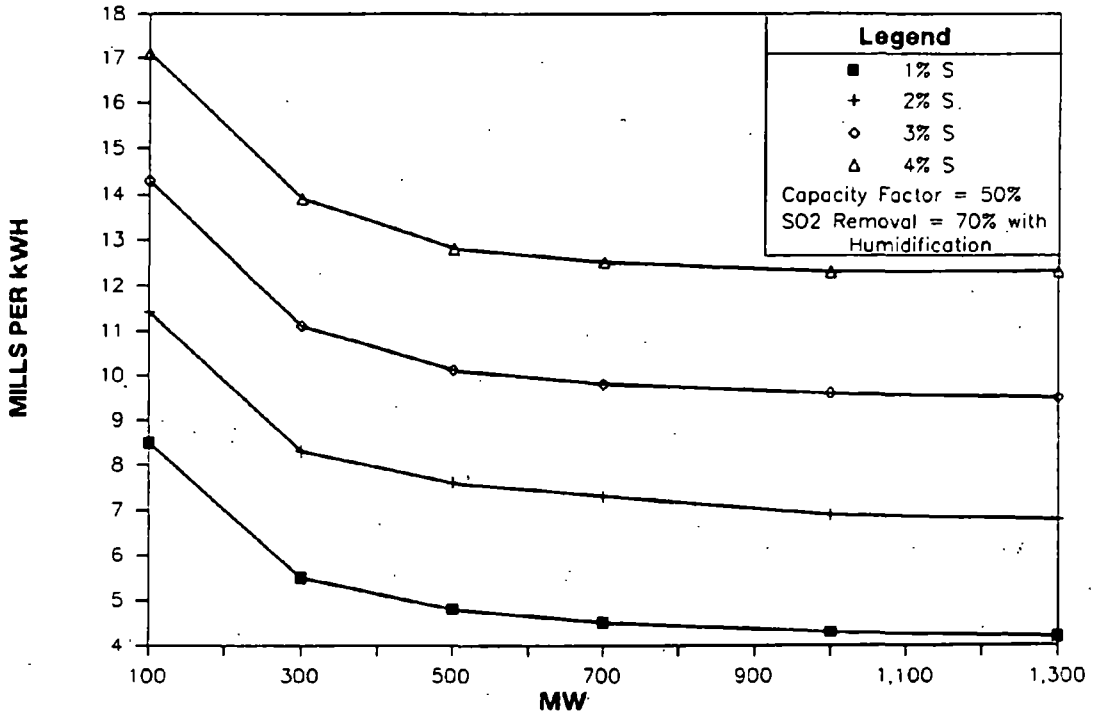


Figure 3.4-7. Furnace Sorbent Injection - Levelized Annual Cost Current 1988 Dollars

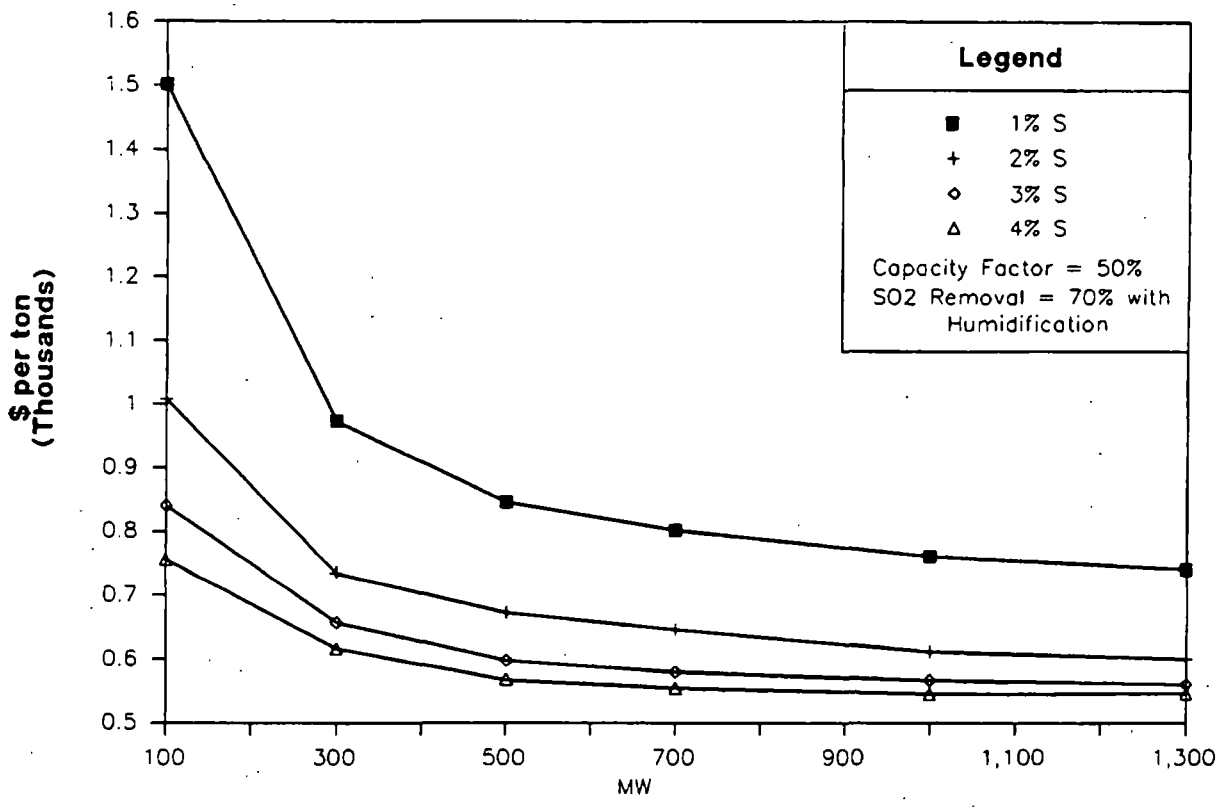


Figure 3.4-8. Furnace Sorbent Injection - Cost Per Ton of SO2 Removed, Current 1988 Dollars

3.5 LOW-TEMPERATURE SORBENT INJECTION

Description

Low-temperature sorbent injection (LTSI) involves injection of dry or slurried sorbents into the duct area between the air preheater and particulate control device. Several different process concepts have been proposed. These include injecting dry sodium compounds, powdered lime hydrates in conjunction with water or steam, and lime slurries. Injection locations range from immediately following the air preheater to the inlet of the particulate control device. Potential advantages of these processes include simplicity and low capital cost of sorbent preparation and injection equipment, and adaptability to retrofit applications.

Sodium-Based Sorbents--

In sodium-based systems, SO_2 is captured in a gas-solid reaction. Key to this reaction is the thermal decomposition of the sorbent to form a porous, high surface area matrix of reactive sodium carbonate (Na_2CO_3). The optimum injection temperature is 275-350°F, physically just downstream of the air heater. Because of its ability to "popcorn" during heating, nahcolite (a naturally occurring mineral containing 70-90 percent sodium bicarbonate--$NaHCO_3$) has very good reactivity and sulfur reduction performance. Trona (a naturally occurring form of sodium sesquicarbonate--$Na_2CO_3 \cdot NaHCO_3 \cdot 2H_2O$) is more densely packed and provides less surface area for reaction with SO_2, but is easier to handle. In general, 20-30 percent of SO_2 removal occurs in the duct while 70-80 percent of SO_2 removal occurs in the particulate control system. For this reason, a baghouse which allows considerable contact between the flue gas and sorbent is preferred.

Sulfur dioxide removal performance with nahcolite and trona during full-scale tests by the Colorado Springs (Colorado) Department of Utilities is shown in Figure 3.5-1. At SO_2 removals of up to 80 percent, nahcolite utilization approached 100 percent. NO_x reductions of up to 23 percent were also obtained during these tests (32). Earlier tests with soda ash (NaOH) were relatively ineffective.

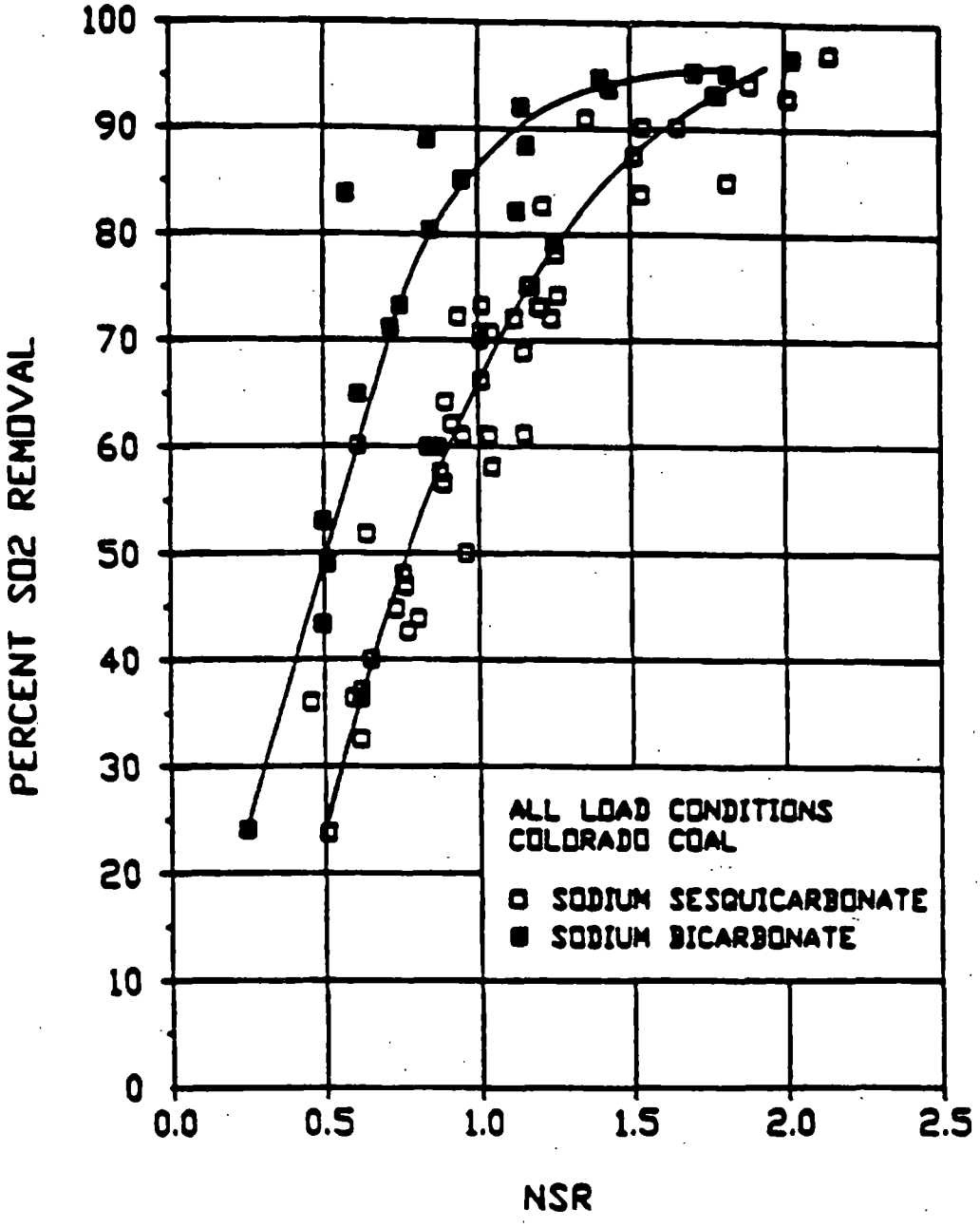


Figure 3.5-1. SO_2 removal as a function of normalized stoichiometric ratio (NSR). (32)

Calcium-Based Sorbents--

Both dry and slurried calcium-based sorbents are being investigated. For both sorbents, flue gas cooling to within 20-30°F of adiabatic saturation is critical (approximately 125°F in a typical coal-fired boiler system). To achieve these temperatures with dry sorbents, water and/or steam is injected into the flue gas downstream of the air preheater. In pilot-scale tests at these temperatures, SO_2 removal rates exceeding 75 percent have been achieved at Ca/S ratios of 2 using conventionally hydrated lime (33,34). Depending on the sorbent and flue gas temperature, 70-90 percent of SO_2 removal occurs during the initial few seconds while the sorbent is suspended in the flue gas. The remaining 10-30 percent occurs in the particulate control device. The quantitative advantages of a baghouse versus an ESP for SO_2 control are not well defined.

Significant similarities in process chemistry exist between conventional spray drying and low-temperature sorbent injection. For slurried sorbents, the major differences are the available residence time and the requirement for rapid sorbent drying to prevent caking of wet sorbent on duct walls. Conventional spray dryers have a reactor vessel residence time of 5-12 seconds, whereas duct spray drying must accomplish particle drying and SO_2 reaction in two seconds or less. As a result, proper choice of atomizer design, slurry concentration, and slurry injection rate is essential.

The concentration of sorbent in the slurry is a major variable in removing SO_2 from the flue gas. In conventional lime-based spray dryers, the solids content of the slurry ranges from 10-40 percent by weight (lime slurries cannot be readily pumped at solids concentrations above 35 percent). To avoid caking on duct walls, the moisture content of the dried droplet should be less than 20 percent. Although high solids concentration slurry will dry more rapidly and have less impingement on duct surfaces, by drying faster the slurry quickly loses its reactivity and has less time to remove SO_2 from the flue gas. Testing will be required to determine the optimum slurry concentration and injection rate for a given SO_2 concentration and duct geometry.

Two basic atomizer designs are available: two-fluid and rotary. The two-fluid atomizer can project a narrow cone of spray, thus reducing impingement of slurry on the duct walls. However, an air compressor is needed and will be a significant energy consumer. Also, early experience with two-fluid systems in conventional spray dryers encountered problems with atomizer erosion and plugging.

Rotary atomizers offer reliability, thorough mixing, and flexibility in varying slurry feed rates without altering the droplet size distribution. Rotary atomizers are the predominant atomization method used in conventional spray drying. However, rotary atomizers offer less control over the spray pattern and may be more prone to duct wall caking problems.

By using a dry sorbent, caking problems are reduced, but flue gas humidification is required to achieve reasonable sorbent utilization. Dry injection has two potential mechanical advantages over slurry injection. First, the sorbent material is handled in a dry form rather than as a slurry, thus avoiding the need for a slurry mixing and handling system. Second, by separately feeding water and sorbent, operation of the system can be better controlled than with a single feed system. This is especially significant with high-sulfur coals where the quantities of calcium hydroxide required for significant SO_2 removal (>50 percent) are large. In such a case, the quantity of water required to maintain slurry rheology could result in excessive cooling of the flue gas. This flexibility in sorbent/water injection also can facilitate system adjustments required by fluctuations in boiler load.

Performance

Several in-duct sorbent injection processes are being developed. The U. S. Department of Energy (DOE) Flue Gas Cleanup Program is sponsoring development of low-cost emissions control technology that can be installed on existing coal-fired power plants if acid rain legislation is enacted. The emphasis of this program is on in-duct sulfur capture. In 1985, three projects were undertaken to test duct injection technology on slipstreams from operating coal-fired power plants.

One of the processes tested in the DOE program was the Hydrate Addition at Low Temperature (HALT) process. In this process, calcium hydrate particles

are injected into the flue gas stream before the gas is cooled by humidification with a fine spray of water. Sulfur dioxide removal efficiencies ranged from 42 to 52% (35). Higher SO_2 reductions of 60-75% have been obtained with corresponding 23 to 28% sorbent utilization, although wall wetting and deposition problems were encountered (36). Another process being evaluated in the DOE program is the Bechtel Confined Zone Dispersion (CZD) process which captures sulfur dioxide by spraying a finely atomized slurry of hydrated lime into the duct of a utility boiler between the air heater and the ESP. Two reactive lime reagents were used for a pilot-scale test. Sulfur dioxide removal efficiencies greater than 50% were achieved (Ca to S ratio of 1.5:1) using either pressure hydrated dolomitic lime or calcitic lime. A five-month, large-scale test was carried out when the CZD process was retrofitted onto one of two parallel flue ducts on a 140 MW unit. Wall wetting and deposition were generally a serious problem.

The General Electric in-duct scrubbing (IDS) process also has been tested at the pilot plant scale under DOE sponsorship. This type of scrubbing is accomplished by injecting a slaked lime slurry with a rotary atomizer directly into existing ductwork. Sulfur dioxide removal takes place in three areas: the slurry injection zone, the evaporation zone, and the drift zone. Test results indicated a lime utilization of 35% calculated for a calcium to sulfur ratio of 1.1 at an SO_2 removal efficiency of 39%.

Another type of injection process being sponsored by EPA is E-SO_x. This process appears to be especially appealing for retrofit applications since it makes use of an existing ESP modified for injecting a lime slurry into the first field of the ESP. The front end of an existing ESP is converted to a spray chamber where the gaseous sulfur dioxide comes into contact with lime slurry droplets. The remaining ESP fields are modified with cold pipe prechargers to bring the ESP particulate emissions back to current levels.

The Ohio Coal Development Office, U. S. EPA, Babcock & Wilcox, and Ohio Edison are jointly sponsoring an evaluation of this process at a 5 MW pilot plant; a 10-month field pilot test program at Ohio Edison's R. E. Burger Station was scheduled to begin in early 1989 (37). Testing will focus on demonstrating a minimum 50% sulfur dioxide removal rate, maintaining acceptable ESP performance with a reduced collector area (to make room for the spray chamber) and increased particulate loading, and showing that solid waste can be disposed of safely and economically.

Another variation of sorbent injection is a combination of furnace and duct injection. Tampella's LIFAC process, combines in-furnace injection with a subsequent vertical humidification reactor for activating the unreacted calcium oxide with water, thus increasing limestone utilization. The process can achieve an SO_2 removal efficiency of 80-85% with commercial grade pulverized limestone at calcium to sulfur ratios of 2.0 or less. Two full-scale applications of this technology exist in Finland, and pilot-plant tests reveal that the process is effective with coals of sulfur contents from 0.5 to 4.3%.

The Advanced Silicate (ADVACATE) process is an in-duct dry injection process that uses a highly reactive calcium silicate hydrate sorbent. The large surface area and high moisture retention capability of the sorbent makes it very reactive to sulfur dioxide and more effective than hydrated lime for dry injection. The sorbent can be produced by mixing a calcium-containing material (e.g., lime) and a silica-containing material (e.g., fly ash) in water at elevated temperature and pressure. The calcium silicate hydrate remains a free flowing dry material with a moisture content up to 35% (37). Pilot and bench scale testing have achieved greater than 80% SO_2 removal at a calcium to sulfur ratio of 2.5 (38). The unreacted lime and fly ash collected in the first particulate control section is hydrated to produce the calcium silicates. The ADVACATE process can be used without furnace injection of limestone by adding lime directly to the hydrator. Preliminary pilot testing of the ADVACATE process has shown a combined furnace/duct SO_2 removal efficiency greater than 90% with a calcium to sulfur ratio near 1.0.

Applicability

Reflective of its development status, a number of unresolved issues remain regarding retrofit of low-temperature sorbent injection: flue gas temperature and velocity limitations and their significance in boiler operations (e.g., load fluctuation); performance on boilers firing high-sulfur coals; method of water injection and limitations on flue gas approach to saturation temperature; sorbent distribution and residence time; sorbent utilization rates at various stoichiometric ratios, injection temperatures, and residence times; choice of and impact on the exiting particulate control system; potential for sorbent fallout and caking in the duct and particulate control systems; and disposal of solid wastes containing high levels of unreacted lime. Because of

similarities, much of the data collected from earlier furnace sorbent injection testing should be directly transferable.

Equipment for LTSI can potentially be installed without major alterations. To achieve satisfactory contact between the sorbent and flue gas, the sorbent injection point should be located at a point where the gas flow is reasonably uniform. Assuming a typical flue gas velocity of 50 feet per second at full load, 50 to 100 feet of straight duct downstream from the point of injection will be required for drying slurried sorbents. An estimated 75 percent of existing coal-fired power plants burning medium- and high-sulfur coals have adequate duct lengths to meet this requirement (39).

Cost

Although both sodium and calcium compounds are can be used with LTSI, uncertainties regarding the availability and cost of sodium compounds in the eastern U. S. and the disposal of sodium wastes are expected to limit their use. Therefore, cost estimates in this section apply only to lime-based systems.

Preliminary estimates of direct capital cost for lime injection systems (assuming the existing ESP can be used without significant modification) ranged from $11-16/kW for dry injection (33,40) to $27-32/kW for slurry injection systems (based on comparison with spray dryer costs). Whether the capital cost differences between dry and slurry injection systems (as well as between furnace and low-temperature injection) are real or result from variations in costing procedures used by different estimators is uncertain. If significant modifications to the particulate control system are required, capital costs can increase several-fold.

Sorbent utilization rates achievable with LTSI are also uncertain due to the lack of clear understanding of process chemistry and limited test data. Estimates of Ca/S ratios required to achieve 50 percent SO_2 removal range from 0.75 to 2.5. As a result, significant uncertainty exists regarding operating costs for LTSI. Based on 1) the uncertainties in capital and operating costs for LTSI and 2) general similarities in hardware for LTSI and FSI, it is likely that the cost of the two basic approaches to SO_2 control are similar. Therefore, the costs presented in Section 3.4 for FSI are likely to be applicable to LTSI.

3.6 REBURNING

Description

Reburning involves bypassing a portion of the fuel around the main combustion zone and injecting it above the main burners to form a reducing zone in which NO_x is converted to reduced nitrogen compounds. The overall process is divided into three zones:

- Main Combustion Zone. Approximately 80-85 percent of the total fuel to the boiler is combusted in this zone. The burners operate at air-fuel ratios less than or equal to those typical of a normal boiler (i.e., 10 percent excess air). No boiler modifications are required in this zone.
- Reburning Zone. The remaining 15-20 percent of the fuel is injected downstream of the main combustion zone to create a fuel-rich reburning zone. Nitrogen oxides (NO_x) produced in the main combustion zone react with hydrocarbon radicals formed by partial oxidation of the reburning fuel to form reduced nitrogen species such as NH_3 and HCN. A portion of these compounds are converted to N_2. The degree of NO_x reduction depends on overall air-fuel ratio (typically 0.80 to 0.95), temperature, residence time, and concentration of NO_x in the reburning zone.
- Burnout Zone. Additional air is added in the cooler, upper furnace to create overall fuel-lean conditions, and ensure complete oxidation of the reburning fuel. Reduced nitrogen species formed in the reburning zone are converted to either NO or N_2.

Any fuel can be used as the reburning fuel. However, U. S. pilot studies indicate that fuels with little or no fuel bound nitrogen, such as natural gas, can achieve greater NO_x reductions. These advantages are particularly significant under operating conditions of lower NO_x levels exiting the main combustion zone (less than 200 ppm) and short residence times. The volatility

and low fuel nitrogen content of natural gas are the primary contributing factors. SO_2 reduction also results from use of sulfur-free natural gas for reburning. The potential exists to further reduce SO_2 emissions by combining reburning with furnace sorbent injection.

Applicability

Because the main combustion zone of the furnace operates at normal air-fuel ratios, gas reburning is applicable to a wide range of wall, tangential, and cyclone-fired boilers. Boiler modifications for reburning involve installation of additional fuel injection burners and air ports above the top or cyclone burner row. Since the main combustion zone operates essentially under original design conditions, problems such as flame impingement and poor carbon burnout are minimized.

For retrofit applications, adequate space (and residence time) between the top burner row and the furnace exit must be available for the additional levels of fuel and air injection. If adequate space is not available, a loss of NO_x reduction performance and/or boiler derating at full load would likely be incurred. Also, if natural gas is used for reburning, a natural gas supply should already be available at or near the plant; otherwise, installation of a natural gas supply pipeline may cause retrofit costs to be unreasonable.

Performance

The development of reburning technology has occurred primarily in the U.S. and Japan, with the largest demonstrations occurring in Japan. In the U. S., the Gas Research Institute (GRI) has sponsored bench-scale (25 KW) and pilot-scale (3 MW) furnace tests at Energy and Environmental Research Corporation (EER) (34). Various fuels including natural gas, high-nitrogen residual oil, and coal have been evaluated as reburning fuels. These tests showed a NO_x reduction capability of 50-75 percent when using 15-20 percent natural gas as the reburning fuel. Use of oil and coal as reburning fuels resulted in less NO_x reduction and increased carbon content in flyash.

In the U. S., three commercial demonstration projects are currently planned; two of the projects use natural gas as the reburn fuel and one will

use coal as the reburn fuel. These projects are planned for start-up in the early 1990's.

In Japan, boiler manufacturer Ishikawajima-Harima Heavy Industries (IHI) has conducted pilot-scale tests as well as tests on two full-scale utility boilers, a 600 MW and a 55 MW unit. These tests were conducted using fuel oil and pulverized coal as the reburning fuels, respectively. The IHI pilot tests showed reburning with pulverized coal could achieve 40-50 percent NO_x reduction. However, in the two full-scale field tests, NO_x reductions were less than 20-25 percent using oil or coal as the reburning fuel (41).

Another Japanese manufacturer, Mitsubishi Heavy Industries (MHI), has conducted tests on a 125 MW oil-fired utility boiler. MHI has also over the last 3 years conducted a full-scale demonstration on a 600 MW oil/coal dual fuel-fired unit. Low-nitrogen oil was used as the reburning fuel in this long-term demonstration, and achieved 50 percent NO_x reduction (42).

Cost

The range of cost estimates for natural gas reburning applied to coal-fired utility boilers using the IAPCS cost model is presented below. The lower costs are for a large unit with a low fuel price differential (FPD). The higher costs are for a small unit with a high FPD. Fuel price differential is the cost per million Btu difference between the current coal and the reburn fuel. Figures 3.6-1 through 3.6-3 present capital cost, annualized cost, and cost per ton of pollutant removed as functions of plant size and FPD. A NO_x reduction of 60 percent was assumed.

	Range
Capital Cost ($/kW)	6.8 to 34.2
Annualized Cost (mills/kWh)	1 to 16.7
Cost Per ton of NO_x Removed ($/ton)	384.7 to 6,418.8

Appendix I contains tables of cost as a function of boiler size, capacity factor, FPD and percent gas substitution.

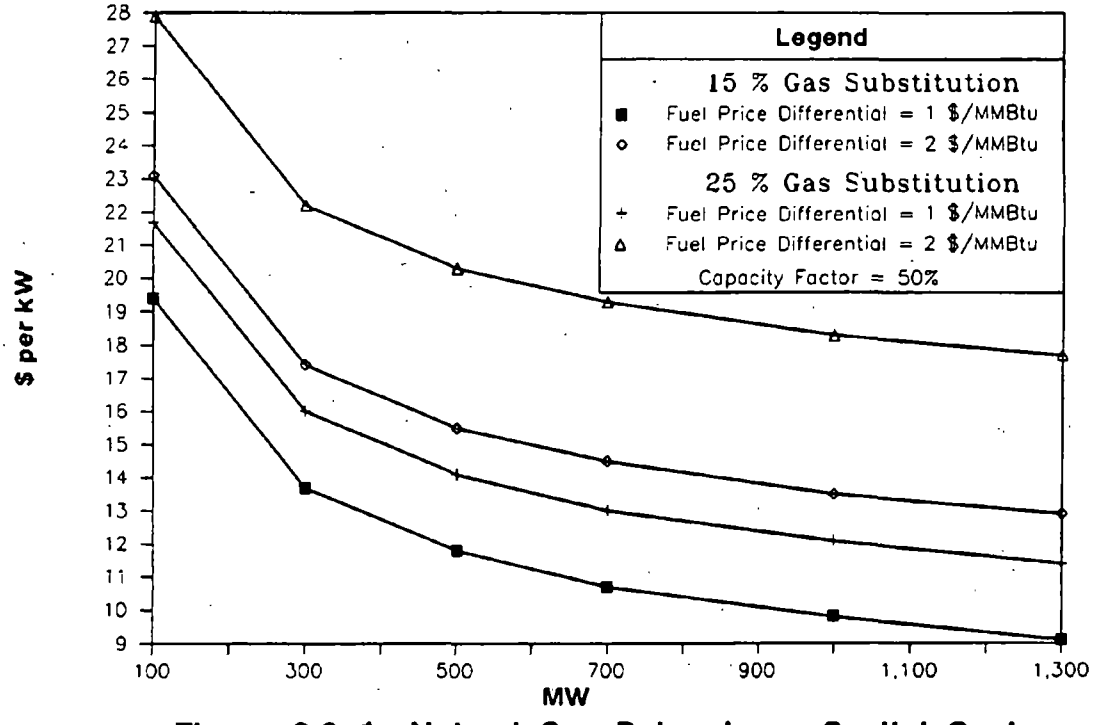


Figure 3.6-1. Natural Gas Reburning - Capital Cost Current 1988 Dollars

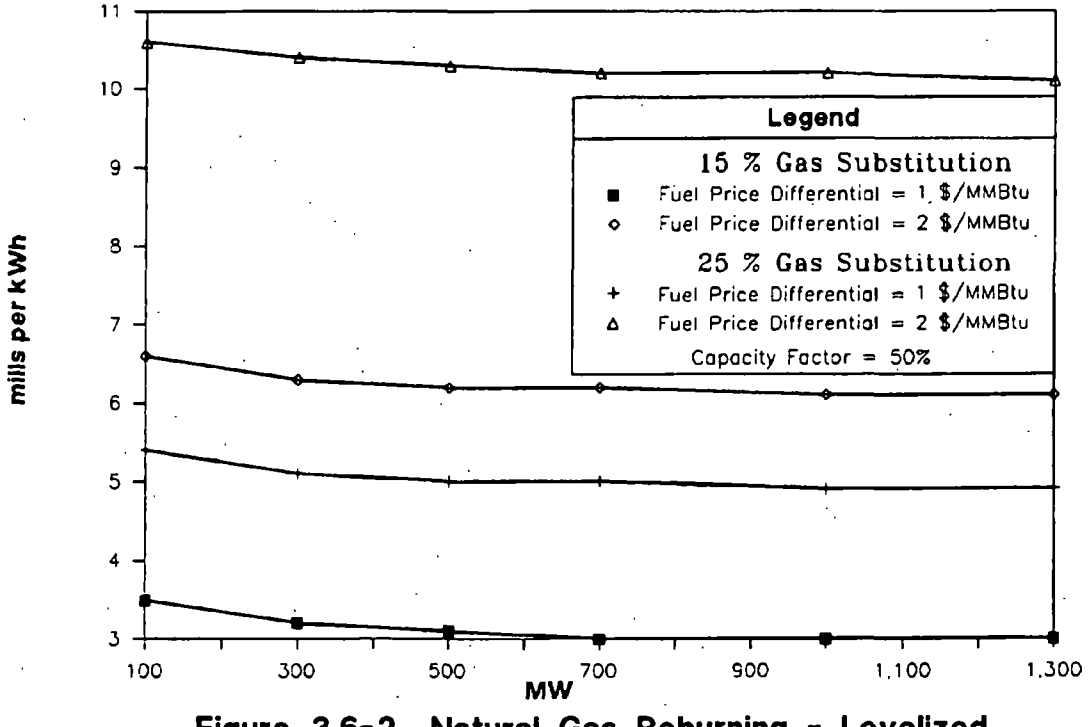


Figure 3.6-2. Natural Gas Reburning - Levelized Annual Cost, Current 1988 Dollars

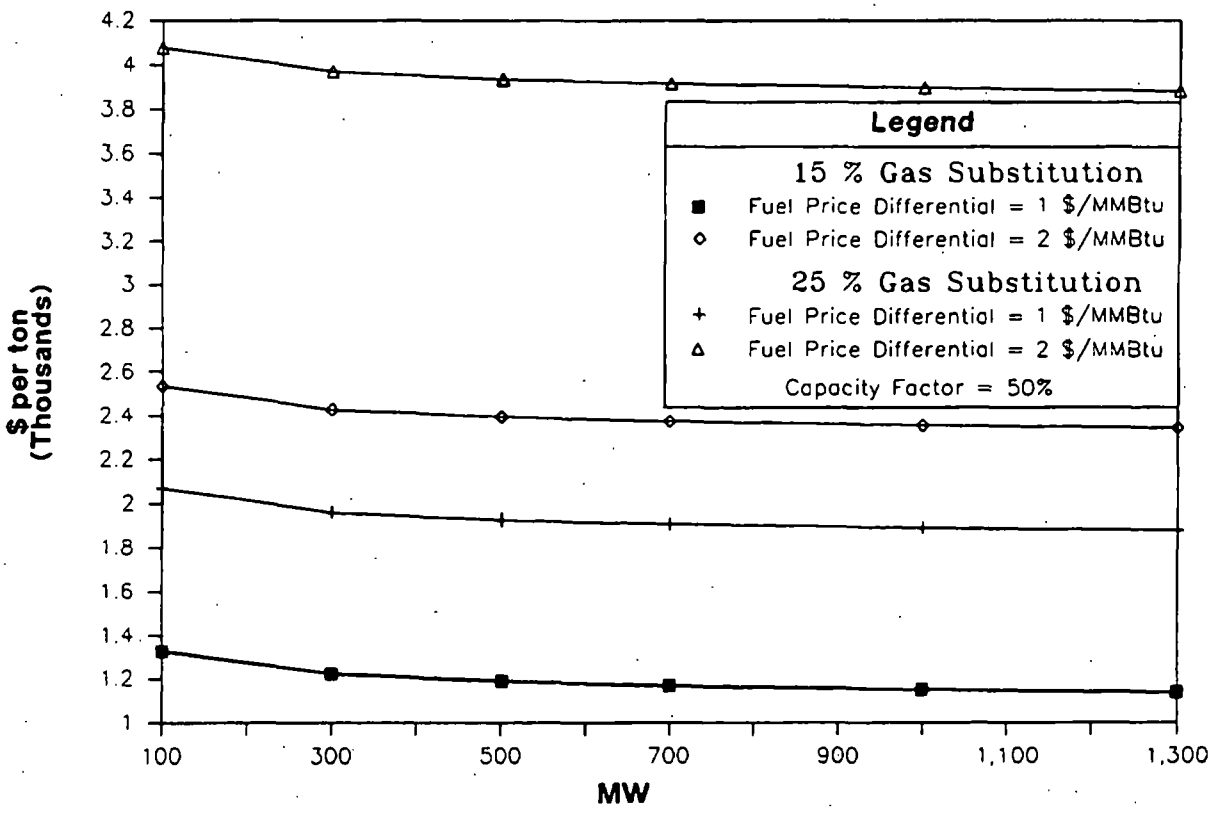


Figure 3.6-3. Natural Gas Reburning - Cost Per Ton of NOx Removed, Current 1988 Dollars

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SECTION 4

EMERGING TECHNOLOGIES

INTRODUCTION

This section reviews two areas of emerging technology which are currently being researched and developed, but have not been demonstrated using full-scale equipment. Depending on their economics and pollution control performance, however, it is possible that these technologies could be in commercial use by the mid- to late-1990's. These technologies are:

- Pre-Combustion Controls
  - Advanced Coal Cleaning (Section 4.1)

- Post-Combustion Controls
  - Advanced FGD (Section 4.2)

4.1 ADVANCED COAL CLEANING

Advanced coal cleaning is divided into two primary areas: 1) advanced physical cleaning and 2) chemical cleaning. Research is also underway in biological cleaning processes, but is currently limited to bench-scale experimentation. Evaluation of several advanced physical cleaning technologies is being conducted through a joint DOE/EPRI program at the EPRI Coal Cleaning Test Facility (CCTF) in Homer City, Pennsylvania.

Description

Advanced Physical Coal Cleaning--

Advanced physical coal cleaning processes are being developed to increase the removal of pyritic sulfur and mineral matter from fine coal (<28 mesh); organic sulfur is not affected. Although froth flotation and dense media cyclones have been used commercially on fines between 28 and 100 mesh, these techniques are currently ineffective at separating materials below 100 mesh (referred to a "ultrafines"). Additionally, flotation is more effective at removal of mineral matter than at removal of pyrite (1).

Research and development in advanced cleaning processes is concentrated in four areas: 1) dense media cyclones, 2) froth flotation, 3) electrostatic separation, and 4) selective coalescence (2). Most of these processes are being developed to clean coal fines produced during crushing at a conventional cleaning plant. Currently, these fines are generally either discarded or recombined with the coarser cleaned product without being cleaned. Briefly, these processes are:

- Dense media cyclones use heavy liquids such as fluorocarbon refrigerants to separate coal from mineral matter based on gravity differences.
- Froth flotation involves fine grinding and multistage flotation to liberate sulfur and mineral matter from coal based on differences in hydrophobic properties.

- Electrostatic separation involves feeding dry, electrically charged pulverized coal onto the surface of a rotating drum. Differences in the dielectric properties of clean coal versus pyrite and ash cause the impurities to separate.
- Small coal particles will selectively coalesce (i.e., agglomerate) into larger particles in the presence of an appropriate medium. The coalesced particles can then be separated from undesirable impurities that do not coalesce. Research is underway on both the basic physical mechanisms that are involved in this phenomenon and the use of novel, non-aqueous media such as liquid carbon dioxide.

Chemical Coal Cleaning--

Chemical cleaning processes are being developed to remove both pyritic and organic sulfur from coal. A variety of methods and chemical reagents have been used in chemical cleaning of coal, including alkali displacement, oxydesulfurization, and chlorinolysis. The major research and development efforts currently underway involve caustic leaching, microwave treatment, and partial hydrogenation.

- TRW's Gravimelt process is the primary caustic leach process under development. Gravimelt exposes coal to a mixture of molten potassium and sodium hydroxide at temperatures of 615-735°F. Sulfur reacts with these alkalis to form sulfides. Residence time is generally 2 to 3 hours. The cleaned coal floats to the top of the molten caustic and liberated mineral matter. The clean coal is then recovered by skimming and washed with water to remove soluble contaminants and residual caustic. Additional ash removal is achieved using an acid wash. Sulfur in the spent caustic is converted to sulfuric acid, calcium sulfate, or elemental sulfur in the regeneration section. The regenerated caustic is reconcentrated and recycled (3).
- Microwave treatment is based on the tendency of water, caustic, sulfur, and other ash components to absorb microwave radiation more readily than the other coal constituents. When exposed to microwave

radiation for less than a minute in the presence of potassium hydroxide, sodium hydroxide, and water, a portion of the sulfur is converted to alkali sulfides. The product coal is water washed to remove the sulfides and subsequently acid washed to dissolve the converted ash.

- Sulfur removal can also be achieved by hydrogenation at mild liquefaction conditions. Organic sulfur is converted to hydrogen sulfide which can in turn be converted to elemental sulfur or sulfuric acid. After separation of pyrite and insoluble mineral matter, the final product is cooled to form a clean solid.

Biological Coal Cleaning--

Several microorganisms have been shown in laboratory and bench-scale tests to promote removal of organic sulfur and pyrite occurring on the surface of coal particles (4,5). Biological desulfurization is a low energy process, has low operating costs, and does not reduce the heating value of the coal product. Because the microorganisms attack the coal particle surface, crushing enhances removal rates. Organic sulfur removals of 25-35 percent have been reported in bench-scale tests with -60 mesh coals. Biological treatment could be especially useful for coals containing very finely disseminated pyrite and organic sulfur that is generally not removable by mechanical techniques.

Potential problems are relatively long bioprocessing times (days to weeks) and the production of acidic, corrosive leaching solutions. However, bioprocessing of stored coal may reduce the time-factor problem, if corrosion problems can be overcome. Additionally, thermophilic microbial action (occurring at $>50^{0}C$) can accelerate processing with some bacteria.

Recent evidence shows that microorganisms can render pyrite hydrophilic in minutes, thereby making it amenable to separation from coal by techniques such as oil agglomeration. Development of such a treatment process would be valuable in that pyrite separation could be accomplished in minutes compared to several days required to remove comparable amounts of sulfur through bacterial oxidation of pyrite to sulfate. Still to be determined are process parameters and the biological mechanism.

Because of the limited data available on the performance and economics of biological treatment processes, no additional consideration of their potential is presented in this report.

Applicability

Advanced coal cleaning processes can be applied to both new and retrofit installations. However, since advanced physical coal cleaning cannot remove organic sulfur, its primary benefit is on coals with a high percentage of fine-grained pyritic sulfur and ash. Chemical cleaning can remove both pyritic and organic sulfur as well as mineral matter and is therefore applicable to a wider range of coals.

Performance

Pyrite in most coals generally accounts for 50-70 percent of the total sulfur. Assuming advanced physical cleaning can remove 90 percent of the pyritic sulfur, it would result in a 45-60 percent reduction in total sulfur. Chemical cleaning can remove over 95 percent of both pyritic and organic sulfur, and up to 99 percent of the mineral matter. Neither approach reduces the nitrogen content of coal and, thus, they have no effect on NO_x emissions.

Costs

Advanced physical and chemical cleaning processes are both more expensive per ton of product than conventional cleaning methods. They are therefore of primary use in cleaning coal fines which cannot be handled by conventional technologies or when additional sulfur removal is needed to meet more stringent regulatory limits.

Depending on the specific process, capital cost estimates for advanced physical coal cleaning equipment range from $22,000 to $82,000 per ton/hr of input capacity; O&M cost estimates range from $2-11 per ton of coal processed (6). The total capital cost of an integrated coal cleaning plant incorporating

an advanced physical cleaning process can exceed $120,000 per ton/hr of input capacity. Typical O&M expenses for the total plant range from $4-8 per ton of raw coal.

Table 4.1-1 reflects costs for a 450 ton per hour (input) integrated plant in which the raw coal is crushed to -3/4 inch (7). Conventional cleaning is used on 28 mesh x 3/4-inch material and advanced processes are used on material <28 mesh and on coarse middling products. Input to the advanced cleaning system equals 40-50 percent of the total plant feed (i.e., 180-225 tons per hour). Total capital cost ranged from $103,000-130,000 per ton/hr input capacity. O&M ranged from $5.20-6.82 per input ton. Coal sulfur content was assumed to be 3.5 percent. Btu and weight recoveries ranged from 83.5-94 percent and 68-85 percent, respectively. Sulfur levels (in lb SO_2/million Btu) decreased by 34-78 percent. Because the least capital cost system may have higher O&M or fuel loss costs (and vice versa), the lowest cost for the overall system is greater than summation of individual cost components.

The capital and busbar costs from Table 4.1-1 are significantly higher than for conventional physical cleaning (Section 2.3). The cost effectiveness values are fairly similar, however, reflecting the higher level of sulfur removal achievable. The cost for cleaning a 2 percent sulfur coal may be similar, but due to the lower level of pyritic sulfur, the cost per ton of SO_2 removed may more than double.

The cost of chemical coal cleaning technologies is uncertain because of the limited research results available (8). Current dollar cost estimates for the Gravimelt process presented in Table 4.1-2 range from $54-87 per ton of cleaned coal (at 13,400 Btu/lb and 10 percent of the original sulfur). This estimate is based on a 10,000 ton/day plant costing $270-380 million and operating 330 days/year. As shown in Table 4.1-2, capital costs account for 20-25 percent of the annualized product price, O&M expenses 70-75 percent, and fuel losses the remaining 5 percent or less.

These cost estimates do not include potential cost savings resulting from improved boiler operations due to reduced solids loadings to the boiler and emission control systems. These savings can significantly improve the economics of coal cleaning, but are frequently specific to individual boilers and coals.

TABLE 4.1-1. ECONOMICS OF ADVANCED PHYSICAL COAL CLEANING
(3.5% Sulfur Coal)

$/Ton of Clean Coal	Constant	Current
Capital	2.70- 3.80	5.20- 8.10
O & M	6.30- 9.00	10.80-15.70
Btu Loss	2.10- 7.00	3.60-12.30
Total	11.10-19.80	19.60-36.10
Busbar Cost (mills/kWh)	4.9 - 7.15	8.40-12.30
$/ton of SO_2	220-490	410-870

TABLE 4.1-2. ECONOMICS OF CHEMICAL COAL CLEANING

$/Ton of Clean Coal	Constant	Current
Capital	6.00- 8.40	12.50-17.60
O & M	21.70-37.90	37.90-66.60
Btu Loss	1.60- 2.20	2.70- 3.20
Total	29.30-48.50	53.10-87.40
Busbar Cost (mills/kWh)	11 -18	19 - 32
$/ton of SO_2	430-730	785-1300

4.2 ADVANCED POST-COMBUSTION SO_2/NO_x PROCESSES

Work is currently underway to develop improved methods for simultaneously controlling SO_2 and NO_x emissions. Common elements in most of these processes is either reagent regeneration or production of a salable by-product to reduce solid waste disposal volumes. This section focuses primarily on two processes which are in pilot-scale and proof-of-concept testing: electron-beam irradiation (E-Beam) and copper oxide. Other advanced processes for simultaneous SO_2/NO_x control which are less advanced include CONOSOX (based on a potassium salt reagent) and Flakt Boliden (based on a sodium citrate reagent).

Description

E-Beam--

Electron-beam processes involve the irradiation of flue gas treated with a reactant such as ammonia or lime to remove both SO_2 and NO_x. Factors affecting SO_2 and NO_x removal by electron-beam irradiation include gas moisture content, gas temperature, oxygen content, reagent ratio, and electron dosage. In addition, efficient electron penetration of the gas stream requires a unique discharge pattern and other special design considerations.

A simplified flow diagram of the E-beam/ammonia process is shown in Figure 4.2-1. Incoming flue gas is cooled and humidified to about 10 percent moisture content in a water quench tower. The cooled flue gas is injected with ammonia and then passed through an electron beam reactor. Oxygen and water in the flue gas are ionized by the electrons to form the radicals [HO], [O], and [HO_2] which react with SO_2 and NO_x to form sulfuric acid (H_2SO_4) and nitric acid (HNO_3). These acids are neutralized by the injected ammonia to form solid ammonium sulfate ($(NH_4)_2SO_4$) and ammonium sulfate nitrate ($(NH_4)_2SO_4 \cdot 2NH_4NO_3$). Reaction time for formation of the sulfate and nitrate salts is less than one second. Product solids are collected in the hopper located below the reactor or in a downstream particulate collector.

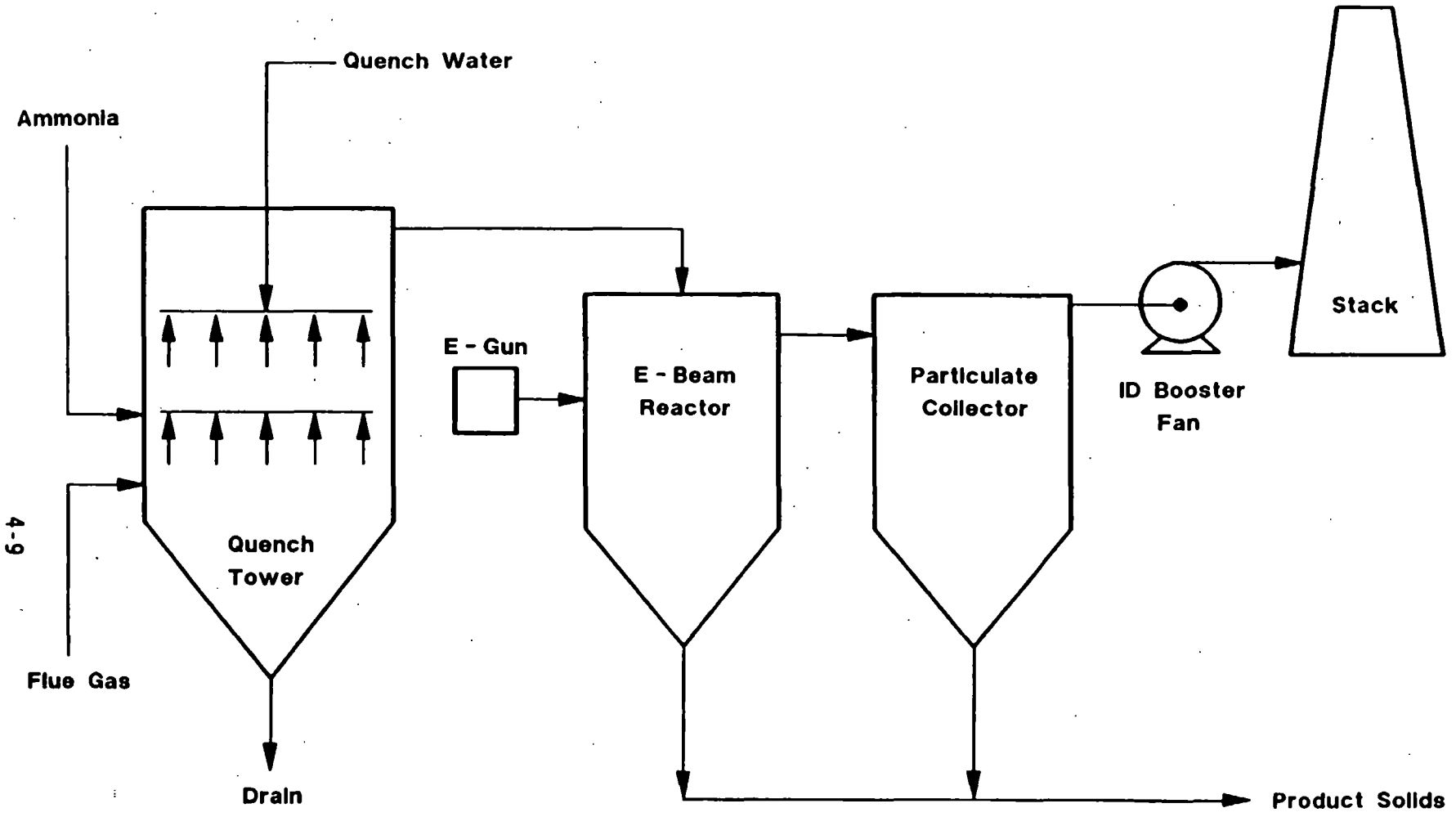

Figure 4.2-1. E-beam / ammonia process flow diagram (9).

In another version of the E-beam process, the water quench tower and ammonia injection system are replaced with a lime-based spray dryer. Reactions are the same manner as above except that the products formed are calcium salts ($CaSO_4$, $Ca(NO_3)_2$, and $CaSO_3$) instead of ammonium salts.

Copper Oxide--

Copper oxide (CuO) FGD is based on the reaction of copper oxide and SO_2 to form copper sulfate ($CuSO_4$). The $CuSO_4$, and to a lesser extent CuO, catalyzes the selective reduction of NO_x to N_2 in the presence of ammonia. Optimum reaction temperature is 650-850^oF. Spent $CuSO_4$ is sent to a second vessel for regeneration by a reducing gas. The resulting concentrated SO_2 stream can be economically recovered as salable sulfur or sulfuric acid.

Two copper oxide processes have been developed. Shell Oil's version of the process uses a set of specially designed, parallel-passage, fixed-bed reactors containing copper oxide bonded to an alumina substrate. The other approach, under development at DOE's Pittsburgh Energy Technology Center, uses a fluidized bed of copper-impregnated alumina spheres.

A simplified diagram of the fluidized-bed process is shown in Figure 4.2-2. In this configuration, the absorber is located between the plant economizer and air heater; ammonia is injected upstream of the absorber. The sulfated sorbent leaves the absorber through overflow pipes and weirs, and is pneumatically transported with pre-heated air either to a solids heater or directly to a moving-bed regenerator. The need for a solids heater depends on the regeneration gas being used. In the regeneration vessel, the sulfated sorbent contacts a reducing gas (CH_4, H_2, or a mixture of CO and H_2) flowing countercurrent to the sorbent. The regenerated sorbent leaves the bottom of the reactor and is pneumatically conveyed in heated air to the absorber. During transport, the copper formed during regeneration is oxidized to copper oxide.

Other Advanced Concepts--

During 1985, Argonne National Laboratory conducted laboratory and bench-scale tests of two advanced SO_2/NO_x processes. In laboratory tests

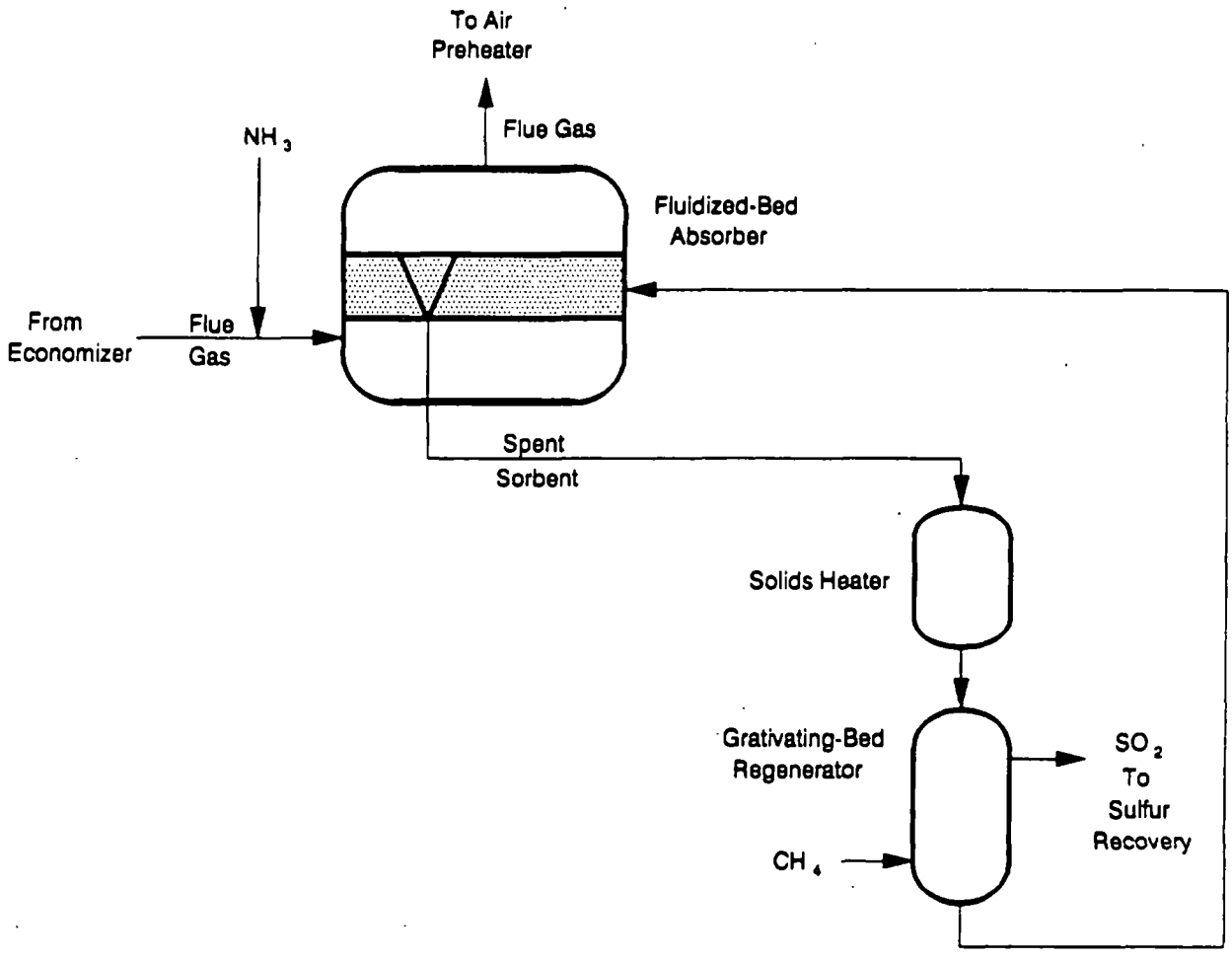

FIGURE 4.2-2. Simplified flow diagram for the Fluidized-Bed Copper Oxide Process (10).

using double-alkali chemistry with proprietary additives, Argonne achieved greater than 70 percent nitrogen oxides removal and 90 percent sulfur oxide removal. Other laboratory work at Argonne has shown 50-70 percent nitrogen oxide removal with greater than 90 percent SO_2 removal using lime spray drying combined with additives; because this work applies to current FGD processes, it could be scaled-up fairly quickly.

Applicability and Performance

E-Beam--

E-beam processes are in the early stages of development and have not been tested at full-scale. Two major pilot projects are being conducted under the U. S. Department of Energy Flue Gas Clean-up Program: one with Research-Cottrell to evaluate the E-beam/lime spray dryer process, and the other with Ebara International Corporation to evaluate E-beam/ammonia injection.

The E-beam/lime spray dryer pilot-scale system was installed on a slipstream (4,000 ACFM) from a 150 MW coal-fired boiler at the TVA Shawnee Steam Plant in Paducah, Kentucky. The pilot work began in early 1983 and was completed in 1985. During the test program, five parameters were varied. The inlet SO_2 concentration was observed to have the greatest impact on SO_2 and NO_x removal. At a high inlet SO_2 concentration (2,500 ppm) and an absorbed electron dose of 1.5 Mrad (a Mrad equals 10 joules/g of flue gas), SO_2 removal was greater than 82 percent and NO_x removal was greater than 90 percent. For a low inlet SO_2 concentration (400 ppm), SO_2 removal was greater than 95 percent and NO_x removal was about 55 percent. The increased NO_x removal at high SO_2 concentrations probably results from reaction of NO and NO_2 with H_2SO_4 to form $HNOSO_4$. (11)

The E-beam/ammonia injection pilot-scale test facility is located at Indiana Power and Light Company's E. W. Stout plant. Construction of the pilot plant began in 1983. Testing completed in 1985 claims to have demonstrated 90 percent reduction in both SO_2 and NO_x. Detailed test results are not yet available.

Copper Oxide--

The copper oxide process has been tested at both bench- and pilot-scale levels. Bench-scale tests with a fluidized-bed absorber using flue gas from a natural gas combustor doped to 3,000 ppmv have shown 90 to 95 percent SO_2 removal and over 97 percent NO_x removal. Pilot-scale fluidized-bed tests using flue gas from a 500 lb/hour combustor burning a 3 percent sulfur bituminous coal achieved 90 percent removal of both SO_2 and NO_x (12). In 1985, DOE awarded a contract to scale this process up to the proof-of-concept level

(i.e., 5 MW equivalent). This pilot-scale facility will be constructed and operated at Commonwealth Edison's Kincaid Station (Kincaid, IL). The Shell process has also been tested at the bench- and pilot-scale levels and has been applied on a commercial 40 MW oil-fired boiler in Japan (13).

The integrated CuO process involves three major reactor systems (the absorber, the regenerator, and SO_2 recovery) and a pneumatic solids handling system between the absorber and regenerator. The relative complexity and space requirements of this system plus the need for 650-850°F flue gas will limit the potential for applying the CuO process to many existing boilers. One possible alternative for overcoming space constraints is to install the absorber downstream of the existing environmental control systems and to reheat the flue gas to the desired temperature. However, operation at the lower end of the desired temperature range (for economic reasons) will reduce SO_2/NO_x removal rates. Because the sulfation reaction is exothermic, part of the reheat energy may be recoverable from the cleaned flue gas.

Cost

Because of the relative complexity of these processes, retrofit onto existing plants will be more difficult than other post-combustion FGD technologies. The impact of this difficulty on retrofit costs has not been quantified, however. Cost estimates for lime-based electron beam technology are presented in Table 4.2-1. The costs shown for SO_2 control are the same as for lime spray drying. The incremental costs for the system are included under the costs for NO_x control. The combined costs for the total system are shown in the right-hand columns. Cost estimates for control of SO_2 and NO_x using copper oxide technology are presented in Table 4.2-2. In both cases, retrofit factors for lime/limestone wet FGD were assumed.

TABLE 4.2-1. COST ESTIMATES FOR ELECTRON BEAM

Technology	% SO_2 Removed	3.5% S	2% S	% NO_x Removed	NO_x	SO2 & NOx Combined 3.5% S	SO2 & NOx Combined 2% S
New 500-MW Baseload Power Plant:[1]							
Capital Cost, $/kW	70 - 90	174 - 207	153 - 182	90	93 - 109	284 - 315	248 - 275
Mills/kWH, Current	70 - 90	15.7 - 18.6	12.0 - 14.0	90	6.3 - 6.7	22.4 - 25.2	18.3 - 20.1
$/Ton, Current $	70 - 90	677 - 804	880 - 1,081	90	2,632 - 3,732	930 - 1,135	1,298 - 1,629
Retrofit 500-MW Baseload Power Plant:[2]							
Capital Cost, $/kW	70 - 90	226 - 270	198 - 235	90	122 - 143	369 - 408	323 - 357
Mills/kWh, Current $	70 - 90	17.1 - 20.4	13.2 - 15.4	90	7.0 - 7.6	24.7 - 27.8	20.3 - 22.4
$/Ton, Current $	70 - 90	730 - 883	961 - 1,202	90	2,944 - 4,182	1,017 -1,260	1,431 - 1,819
Retrofit 250-MW Intermediate Load Power Plant:[3]							
Capital Cost, $/kW	70 - 90	339 - 403	297 - 352	90	183 - 213	552 - 612	483 - 535
Mills/kWh, Current $	70 - 90	30.2 - 35.3	25.2 - 29.2	90	12.9 - 14.3	44.5 - 49.4	38.1 - 41.8
$/Ton, Current $	70 - 90	1,219 -1,609	1,753 - 2,365	90	5,396 - 7,854	1,760 - 2,329	2,608 - 3,500

[1] Retrofit factor of 1.0 and capacity factory of 65 percent.

[2] Retrofit factor of 1.3 and capacity factor of 65 percent.

TABLE 4.2-2. COST ESTIMATES FOR COPPER OXIDE

Technology	% SO_2 Removed	3.5% S	2% S	% NO_x Removed	NO_x	SO2 & NOx Combined 3.5% S	SO2 & NOx Combined 2% S
New 500-MW Baseload Power Plant:[1]							
Capital Cost, $/kW	70 - 90	187 - 203	155 - 167	90	(10)	187 - 203	155 - 167
Mills/kWH, Current	70 - 90	17.7 - 20.5	13.0 - 14.7	90	(12)	17.7 - 20.5	13.0 - 14.7
$/Ton, Current $	70 - 90	760 - 883	949 - 1,143	90	760 - 1,143	760 - 883	949 - 1,143
Retrofit 500-MW Baseload Power Plant:[2]							
Capital Cost, $/kW	70 - 90	235 - 255	192 - 207	90	(10)	235 - 255	192 - 207
Mills/kWh, Current $	70 - 90	18.8 - 21.9	14.0 - 15.8	90	(12)	18.8 - 21.9	14.0 - 15.8
$/Ton, Current $	70 - 90	813 - 951	1,028 - 1,236	90	813 - 1,236	813 - 951	1,028 - 1,236
Retrofit 250-MW Intermediate Load Power Plant:[3]							
Capital Cost, $/kW	70 - 90	360 - 391	290 - 315	90	(10)	360 - 391	290 - 315
Mills/kWh, Current $	70 - 90	31.3 - 35.0	24.6 - 27.1	90	(12)	31.3 - 35.0	24.6 - 27.1
$/Ton, Current $	70 - 90	1,269 -1,613	1,715 - 2,216	90	1,269 - 2,216	1,269 - 1,613	1,715 - 2,216

[1] Retrofit factor of 1.0 and capacity factory of 65 percent.

[2] Retrofit factor of 1.3 and capacity factor of 65 percent.

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APPENDIX A

Summary of Control Costs
Coal Switching and Blending

COAL SWITCHING AND BLENDING
CAPITAL COST ($/kW)

ESP SIZE	FUEL PRICE DIFFERENTIAL ($)	SULFUR (%)	CAPACITY FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
LARGE ESP		1.0			30.7	23.2	21.3	20.4	19.6	19.1
(400 SCA)	5	2.0	30.0	$/kW	30.7	23.2	21.3	20.4	19.6	19.1
		3.0			30.7	23.2	21.3	20.4	19.6	19.1
		4.0			30.7	23.2	21.3	20.4	19.6	19.1
LARGE ESP		1.0			30.7	23.2	21.3	20.3	19.6	19.1
(400 SCA)	5	2.0	50.0	$/kW	30.7	23.2	21.3	20.3	19.6	19.1
		3.0			30.7	23.2	21.3	20.3	19.6	19.1
		4.0			30.7	23.2	21.3	20.3	19.6	19.1
LARGE ESP		1.0			30.7	23.2	21.3	20.3	19.6	19.1
(400 SCA)	5	2.0	70.0	$/kW	30.7	23.2	21.3	20.3	19.6	19.1
		3.0			30.7	23.2	21.3	20.3	19.6	19.1
		4.0			30.7	23.2	21.3	20.3	19.6	19.1
SMALL ESP		1.0			31.8	24.3	22.4	21.5	20.7	20.2
(200 SCA)	5	2.0	30.0	$/kW	31.8	24.3	22.4	21.5	20.7	20.2
		3.0			31.8	24.3	22.4	21.5	20.7	20.2
		4.0			31.8	24.3	22.4	21.5	20.7	20.2
SMALL ESP		1.0			31.8	24.3	22.4	21.5	20.7	20.2
(200 SCA)	5	2.0	50.0	$/kW	31.8	24.3	22.4	21.5	20.7	20.2
		3.0			31.8	24.3	22.4	21.5	20.7	20.2
		4.0			31.8	24.3	22.4	21.5	20.7	20.2
SMALL ESP		1.0			31.8	24.3	22.4	21.5	20.7	43.1
(200 SCA)	5	2.0	70.0	$/kW	31.8	24.3	22.4	21.5	20.7	43.1
		3.0			31.8	24.3	22.4	21.5	20.7	43.1
		4.0			31.8	24.3	22.4	21.5	20.7	43.1

COAL SWITCHING AND BLENDING
CAPITAL COST ($/kW)

ESP SIZE	FUEL PRICE DIFFERENTIAL ($)	SULFUR (%)	CAPACITY FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
LARGE ESP		1.0			41.1	33.6	31.6	30.7	29.9	29.5
(400 SCA)	15	2.0	30.0	$/kW	41.1	33.6	31.6	30.7	29.9	29.5
		3.0			41.1	33.6	31.6	30.7	29.9	29.5
		4.0			41.1	33.6	31.6	30.7	29.9	29.5
LARGE ESP		1.0			41.1	33.5	31.6	30.7	29.9	29.5
(400 SCA)	15	2.0	50.0	$/kW	41.1	33.5	31.6	30.7	29.9	29.5
		3.0			41.1	33.5	31.6	30.7	29.9	29.5
		4.0			41.1	33.5	31.6	30.7	29.9	29.5
LARGE ESP		1.0			41.1	33.5	31.6	30.7	29.9	29.5
(400 SCA)	15	2.0	70.0	$/kW	41.1	33.5	31.6	30.7	29.9	29.5
		3.0			41.1	33.5	31.6	30.7	29.9	29.5
		4.0			41.1	33.5	31.6	30.7	29.9	29.5
SMALL ESP		1.0			42.2	34.7	32.8	31.8	31.1	30.6
(200 SCA)	15	2.0	30.0	$/kW	42.2	34.7	32.8	31.8	31.1	30.6
		3.0			42.2	34.7	32.8	31.8	31.1	30.6
		4.0			42.2	34.7	32.8	31.8	31.1	30.6
SMALL ESP		1.0			42.2	34.7	32.7	31.8	31.1	30.6
(200 SCA)	15	2.0	50.0	$/kW	42.2	34.7	32.7	31.8	31.1	30.6
		3.0			42.2	34.7	32.7	31.8	31.1	30.6
		4.0			42.2	34.7	32.7	31.8	31.1	30.6
SMALL ESP		1.0			42.2	34.7	32.7	31.8	31.1	30.6
(200 SCA)	15	2.0	70.0	$/kW	42.2	34.7	32.7	31.8	31.1	30.6
		3.0			42.2	34.7	32.7	31.8	31.1	30.6
		4.0			42.2	34.7	32.7	31.8	31.1	30.6

COAL SWITCHING AND BLENDING
ANNUAL COST (mills/kWh)

ESP SIZE	FUEL PRICE DIFFERENTIAL ($)	SULFUR (%)	CAPACITY FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
LARGE ESP		1.0			8.2	6.9	6.6	6.5	6.4	6.3
(400 SCA)	5	2.0	30.0	mills/kWh	8.2	6.9	6.6	6.5	6.4	6.3
		3.0			8.2	6.9	6.6	6.5	6.4	6.3
		4.0			8.2	6.9	6.6	6.5	6.4	6.3
LARGE ESP		1.0			7	6.1	5.9	5.8	5.7	5.6
(400 SCA)	5	2.0	50.0	mills/kWh	7	6.1	5.9	5.8	5.7	5.6
		3.0			7	6.1	5.9	5.8	5.7	5.6
		4.0			7	6.1	5.9	5.8	5.7	5.6
LARGE ESP		1.0			6.5	5.7	5.5	5.5	5.4	5.4
(400 SCA)	5	2.0	70.0	mills/kWh	6.5	5.7	5.5	5.5	5.4	5.4
		3.0			6.5	5.7	5.5	5.5	5.4	5.4
		4.0			6.5	5.7	5.5	5.5	5.4	5.4
SMALL ESP		1.0			8.3	7	6.7	6.6	6.5	6.4
(200 SCA)	5	2.0	30.0	mills/kWh	8.3	7	6.7	6.6	6.5	6.4
		3.0			8.3	7	6.7	6.6	6.5	6.4
		4.0			8.3	7	6.7	6.6	6.5	6.4
SMALL ESP		1.0			7.1	6.1	5.9	5.8	5.8	5.7
(200 SCA)	5	2.0	50.0	mills/kWh	7.1	6.1	5.9	5.8	5.8	5.7
		3.0			7.1	6.1	5.9	5.8	5.8	5.7
		4.0			7.1	6.1	5.9	5.8	5.8	5.7
SMALL ESP		1.0			6.6	5.8	5.6	5.5	5.4	5.4
(200 SCA)	5	2.0	70.0	mills/kWh	6.6	5.8	5.6	5.5	5.4	5.4
		3.0			6.6	5.8	5.6	5.5	5.4	5.4
		4.0			6.6	5.8	5.6	5.5	5.4	5.4

COAL SWITCHING AND BLENDING
ANNUAL COST (mills/kWh)

ESP SIZE	FUEL PRICE DIFFERENTIAL ($)	SULFUR (%)	CAPACITY FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
LARGE ESP		1.0			17.1	15.8	15.6	15.4	15.3	15.2
(400 SCA)	15	2.0	30.0	mills/kWh	17.1	15.8	15.6	15.4	15.3	15.2
		3.0			17.1	15.8	15.6	15.4	15.3	15.2
		4.0			17.1	15.8	15.6	15.4	15.3	15.2
LARGE ESP		1.0			15.7	14.7	14.5	14.4	14.4	14.3
(400 SCA)	15	2.0	50.0	mills/kWh	15.7	14.7	14.5	14.4	14.4	14.3
		3.0			15.7	14.7	14.5	14.4	14.4	14.3
		4.0			15.7	14.7	14.5	14.4	14.4	14.3
LARGE ESP		1.0			15.1	14.3	14.1	14	13.9	13.9
(400 SCA)	15	2.0	70.0	mills/kWh	15.1	14.3	14.1	14	13.9	13.9
		3.0			15.1	14.3	14.1	14	13.9	13.9
		4.0			15.1	14.3	14.1	14	13.9	13.9
SMALL ESP		1.0			17.2	16.3	15.7	15.5	15.4	15.4
(200 SCA)	15	2.0	30.0	mills/kWh	17.2	16.3	15.7	15.5	15.4	15.4
		3.0			17.2	16.3	15.7	15.5	15.4	15.4
		4.0			17.2	16.3	15.7	15.5	15.4	15.4
SMALL ESP		1.0			15.7	14.8	14.6	14.5	14.4	14.4
(200 SCA)	15	2.0	50.0	mills/kWh	15.7	14.8	14.6	14.5	14.4	14.4
		3.0			15.7	14.8	14.6	14.5	14.4	14.4
		4.0			15.7	14.8	14.6	14.5	14.4	14.4
SMALL ESP		1.0			15.1	14.3	14.1	14.1	14	14
(200 SCA)	15	2.0	70.0	mills/kWh	15.1	14.3	14.1	14.1	14	14
		3.0			15.1	14.3	14.1	14.1	14	14
		4.0			15.1	14.3	14.1	14.1	14	14

COAL SWITCHING AND BLENDING
SO2 COST EFFECTIVENESS ($/TON)

ESP SIZE	FUEL PRICE DIFFERENTIAL ($)	SULFUR (%)	CAPACITY FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
LARGE ESP		1.0			9650.8	8147.4	7797.3	7637.1	7505.2	7428.7
(400 SCA)	5	2.0	30.0	$/TON	916.8	774	740.7	725.5	713	705.7
		3.0			481.2	406.3	388.8	380.8	374.3	370.4
		4.0			326.3	275.4	263.6	258.2	253.7	251.1
LARGE ESP		1.0			8266.4	7159.9	6910	6797.9	6706.4	6653.3
(400 SCA)	5	2.0	50.0	$/TON	785.3	680.2	656.4	645.8	637.1	632
		3.0			412.2	357	344.6	339	334.4	331.8
		4.0			279.5	242	233.6	229.8	226.7	224.9
LARGE ESP		1.0			7674	6736.9	6529.8	6438.3	6364	6321.1
(400 SCA)	5	2.0	70.0	$/TON	729	640	620.3	611.6	604.6	600.5
		3.0			382.7	335.9	325.6	321.1	317.3	315.2
		4.0			259.4	227.7	220.7	217.7	215.1	213.7
SMALL ESP		1.0			9783.8	8281.9	7931.9	7771.7	7640	7563.3
(200 SCA)	5	2.0	30.0	$/TON	929.4	786.7	753.5	738.3	725.8	718.5
		3.0			487.9	413	395.5	387.5	381	377.2
		4.0			330.8	280	268.1	262.7	258.3	255.7
SMALL ESP		1.0			8349.9	7244.3	6994.2	6882.2	6790.8	6737.7
(200 SCA)	5	2.0	50.0	$/TON	793.2	688.2	664.4	653.8	645.1	640
		3.0			416.4	361.2	348.8	343.2	338.6	336
		4.0			282.3	244.9	236.4	232.7	229.6	227.8
SMALL ESP		1.0			7736.4	6799.8	6592.5	6501.1	6426.8	6383.9
(200 SCA)	5	2.0	70.0	$/TON	734.9	645.9	626.3	617.6	610.5	606.4
		3.0			385.8	339.1	328.7	324.2	320.5	318.3
		4.0			261.5	229.9	222.9	219.8	217.3	215.8

COAL SWITCHING AND BLENDING
SO2 COST EFFECTIVENESS ($/TON)

ESP SIZE	FUEL PRICE DIFFERENTIAL ($)	SULFUR (%)	CAPACITY FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
LARGE ESP		1.0			20220.6	18717.4	18367.4	18207.2	18075.4	17998.9
(400 SCA)	15	2.0	30.0	$/TON	1920.9	1778.1	1744.8	1729.6	1717.1	1709.8
		3.0			1008.3	933.4	915.9	907.9	901.3	897.5
		4.0			683.6	632.8	620.9	615.5	611.1	608.5
LARGE ESP		1.0			18510.4	17403.3	17153.5	17041.3	16949.8	16896.8
(400 SCA)	15	2.0	50.0	$/TON	1758.4	1653.2	1629.5	1618.8	1610.1	1605.1
		3.0			923	867.8	855.4	849.8	845.2	842.6
		4.0			625.8	588.3	579.9	576.1	573	571.2
LARGE ESP		1.0			17777.6	16840.3	16633.3	16541.8	16467.5	16424.5
(400 SCA)	15	2.0	70.0	$/TON	1688.8	1599.7	1580.1	1571.4	1564.3	1560.2
		3.0			886.5	839.8	829.4	824.9	821.2	819
		4.0			601	569.3	562.3	559.2	556.7	555.2
SMALL ESP		1.0			20354.1	18851.9	18501.9	18341.8	18210.1	18133.5
(200 SCA)	15	2.0	30.0	$/TON	1933.5	1790.8	1757.6	1742.4	1729.9	1722.6
		3.0			1015	940.1	922.6	914.6	908.1	904.2
		4.0			688.1	637.3	625.5	620.1	615.6	613
SMALL ESP		1.0			18593.9	17487.6	17237.7	17125.6	17034.2	16981.1
(200 SCA)	15	2.0	50.0	$/TON	1766.3	1661.2	1637.5	1626.8	1618.2	1613.1
		3.0			927.2	872	859.6	854	849.4	846.8
		4.0			628.6	591.2	582.7	578.9	575.9	574.1
SMALL ESP		1.0			17840	16903.2	16696	16604.6	16530.3	16487.4
(200 SCA)	15	2.0	70.0	$/TON	1694.7	1605.7	1586	1577.3	1570.3	1566.2
		3.0			889.6	842.9	832.6	828	824.3	822.2
		4.0			603.1	571.4	564.4	561.3	558.8	557.4

APPENDIX B

Low NO_x Combustion Control Technologies

NOx COMBUSTION CONTROL TECHNOLOGIES:
OVERFIRE AIR, LOW NOx BURNERS
CAPITAL COST ($/kW)

FIRING CONFIGURATION	CAPACITY FACTOR (%)	NOx REDUCTION (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
TANGENTIAL		10		6.2	3.2	2.4	1.9	1.6	1.3
FIRING	30	20	$/kW	6.2	3.2	2.4	1.9	1.6	1.3
(OVERFIRE AIR)		30		6.2	3.2	2.4	1.9	1.6	1.3
TANGENTIAL		10		6.2	3.2	2.4	1.9	1.6	1.3
FIRING	50	20	$/kW	6.2	3.2	2.4	1.9	1.6	1.3
(OVERFIRE AIR)		30		6.2	3.2	2.4	1.9	1.6	1.3
TANGENTIAL		10		6.2	3.2	2.4	1.9	1.6	1.3
FIRING	70	20	$/kW	6.2	3.2	2.4	1.9	1.6	1.3
(OVERFIRE AIR)		30		6.2	3.2	2.4	1.9	1.6	1.3
WALL FIRING		20		25.5	13.2	9.7	7.9	6.4	5.5
(LOW NOx	30	40	$/kW	25.5	13.2	9.7	7.9	6.4	5.5
BURNERS)		55		25.5	13.2	9.7	7.9	6.4	5.5
WALL FIRING		20		25.5	13.2	9.7	7.9	6.4	5.5
(LOW NOx	50	40	$/kW	25.5	13.2	9.7	7.9	6.4	5.5
BURNERS)		55		25.5	13.2	9.7	7.9	6.4	5.5
WALL FIRING		20		25.5	13.2	9.7	7.9	6.4	5.5
(LOW NOx	70	40	$/kW	25.5	13.2	9.7	7.9	6.4	5.5
BURNERS)		55		25.5	13.2	9.7	7.9	6.4	5.5

NOx COMBUSTION CONTROL TECHNOLOGIES:
OVERFIRE AIR, LOW NOx BURNERS
ANNUAL COST (mills/kWh)

FIRING CONFIGURATION (%)	CAPACITY FACTOR (%)	NOx REDUCTION (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
TANGENTIAL		10		0.5	0.3	0.2	0.2	0.1	0.1
FIRING	30	20	mills/kWh	0.5	0.3	0.2	0.2	0.1	0.1
(OVERFIRE AIR)		30		0.5	0.3	0.2	0.2	0.1	0.1
TANGENTIAL		10		0.3	0.2	0.1	0.1	0.1	0.1
FIRING	50	20	mills/kWh	0.3	0.2	0.1	0.1	0.1	0.1
(OVERFIRE AIR)		30		0.3	0.2	0.1	0.1	0.1	0.1
TANGENTIAL		10		0.2	0.1	0.1	0.1	0.1	0
FIRING	70	20	mills/kWh	0.2	0.1	0.1	0.1	0.1	0
(OVERFIRE AIR)		30		0.2	0.1	0.1	0.1	0.1	0
WALL FIRING		20		2.1	1.1	0.8	0.6	0.5	0.4
(LOW NOx	30	40	mills/kWh	2.1	1.1	0.8	0.6	0.5	0.4
BURNERS)		55		2.1	1.1	0.8	0.6	0.5	0.4
WALL FIRING		20		1.2	0.6	0.5	0.4	0.3	0.3
(LOW NOx	50	40	mills/kWh	1.2	0.6	0.5	0.4	0.3	0.3
BURNERS)		55		1.2	0.6	0.5	0.4	0.3	0.3
WALL FIRING		20		0.9	0.5	0.3	0.3	0.2	0.2
(LOW NOx	70	40	mills/kWh	0.9	0.5	0.3	0.3	0.2	0.2
BURNERS)		55		0.9	0.5	0.3	0.3	0.2	0.2

NOx COMBUSTION CONTROL TECHNOLOGIES:
OVERFIRE AIR, LOW NOx BURNERS
NOx COST EFFECTIVENESS ($/TON)

FIRING CONFIGURATION	CAPACITY FACTOR (%)	NOx REDUCTION (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
TANGENTIAL		10		1621.2	838.8	617.3	504.7	407.6	348
FIRING	30	20	$/TON	810.6	419.4	308.7	252.4	203.8	174
(OVERFIRE AIR)		30		540.4	279.6	205.8	168.2	135.9	116
TANGENTIAL		10		972.7	503.3	370.4	302.8	244.6	208.8
FIRING	50	20	$/TON	486.4	251.6	185.2	151.4	122.3	104.4
(OVERFIRE AIR)		30		324.2	167.8	123.5	100.9	81.5	69.6
TANGENTIAL		10		694.8	359.5	264.6	216.3	174.7	149.1
FIRING	70	20	$/TON	347.4	179.7	132.3	108.2	87.4	74.6
(OVERFIRE AIR)		30		231.6	119.8	88.2	72.1	58.2	49.7
WALL FIRING		20		2388.8	1235.5	909.4	743.1	600.1	512.6
(LOW NOx	30	40	$/TON	1194.4	617.8	454.7	371.6	300	256.3
BURNERS)		55		868.6	449.3	330.7	270.2	218.2	186.4
WALL FIRING		20		1433.3	741.3	545.7	445.9	360	307.6
(LOW NOx	50	40	$/TON	716.6	370.7	272.8	222.9	180	153.8
BURNERS)		55		521.2	269.6	198.4	162.1	130.9	111.8
WALL FIRING		20		1023.8	529.5	389.8	318.5	257.2	129.7
(LOW NOx	70	40	$/TON	511.9	264.8	194.9	159.2	128.6	109.9
BURNERS)		55		372.3	192.6	141.7	115.8	93.5	79.9

APPENDIX C

Lime/Limestone FGD

L/S-FGD
CAPITAL COST ($/kW)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		296.6	190.5	158.2	140.5	136.7	125
1.0	30	1.5	$/kW	426.3	271	223.6	198	192.2	175.3
		2.0		556	351.5	289.1	255.4	247.7	225.5
		1.0		296.8	190.6	158.3	140.6	136.7	125.1
1.0	50	1.5	$/kW	426.5	271.1	223.7	198.1	192.3	175.3
		2.0		556.2	351.6	289.2	255.5	247.8	225.6
		1.0		297	190.7	158.4	140.7	136.8	125.1
1.0	70	1.5	$/kW	426.7	271.2	223.8	198.1	192.3	175.4
		2.0		556.4	351.7	289.2	255.6	247.8	225.6
		1.0		316.4	196.8	164.8	146.3	144.5	132
2.0	30	1.5	$/kW	455.2	279.8	232.8	206	203.3	185.2
		2.0		594	362.7	300.9	265.7	262.1	238.3
		1.0		316.6	196.9	164.9	146.4	144.6	132.1
2.0	50	1.5	$/kW	455.4	279.9	232.9	206.1	203.4	185.2
		2.0		594.2	362.8	301	265.8	262.2	238.4
		1.0		316.7	197	164.9	146.4	144.7	132.2
2.0	70	1.5	$/kW	455.5	280	233	206.1	203.5	185.3
		2.0		594.4	362.9	301.1	265.8	262.3	238.4
		1.0		328.4	204	175.2	155.1	150.1	137.5
3.0	30	1.5	$/kW	472.5	289.9	247.8	218.4	211	192.7
		2.0		616.6	375.8	320.5	281.7	271.9	247.9
		1.0		328.6	204.1	175.3	155.1	150.2	137.6
3.0	50	1.5	$/kW	472.7	290	247.9	218.5	211.1	192.8
		2.0		616.8	376	320.6	281.8	272	248
		1.0		328.7	204.2	175.4	155.2	150.3	137.7
3.0	70	1.5	$/kW	472.8	290.1	248	218.5	211.2	192.9
		2.0		616.9	376.1	320.7	281.9	272	248.1
		1.0		334.8	209.8	180.9	160.1	154.8	143
4.0	30	1.5	$/kW	481.5	298	255.6	225.4	217.5	200.2
		2.0		628.1	386.2	330.3	290.6	280.1	257.5
		1.0		335	209.9	181	160.2	154.9	143.1
4.0	50	1.5	$/kW	481.7	298.1	255.7	225.4	217.6	200.3
		2.0		628.3	386.3	330.4	290.7	280.2	257.5
		1.0		335.2	210	181.1	160.3	155	143.2
4.0	70	1.5	$/kW	481.9	298.2	255.8	225.5	217.6	200.4
		2.0		628.5	386.4	330.5	290.7	280.3	257.6

L/S-FGD
ANNUAL COST (mills/kWh)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		45.6	29	24.3	21.7	20.9	19.4
1.0	30	1.5	mills/kWh	61.4	38.7	32.2	28.7	27.7	25.5
		2.0		77.1	48.5	40.1	35.7	34.4	31.6
		1.0		30	19.5	16.6	15	14.5	13.5
1.0	50	1.5	mills/kWh	39.4	25.4	21.3	19.2	18.5	17.2
		2.0		48.8	31.2	26.1	23.4	22.6	20.8
		1.0		23.2	15.4	13.2	12.1	11.7	11
1.0	70	1.5	mills/kWh	29.9	19.6	16.7	15.1	14.6	13.6
		2.0		36.7	23.8	20.1	18.1	17.5	16.2
		1.0		49	30.6	26	23.3	22.8	21.1
2.0	30	1.5	mills/kWh	65.8	40.7	34.2	30.6	30	27.6
		2.0		82.7	50.8	42.5	37.8	37.1	34
		1.0		32.4	20.9	18	16.3	16	15
2.0	50	1.5	mills/kWh	42.5	26.9	22.9	20.7	20.3	18.8
		2.0		52.6	33	27.9	25	24.6	22.7
		1.0		25.2	16.7	14.5	13.3	13.1	12.3
2.0	70	1.5	mills/kWh	32.4	21	18.1	16.4	16.1	15.1
		2.0		39.6	25.3	21.6	19.5	19.2	17.8
		1.0		51.3	32.4	28.1	25.3	24.4	22.7
3.0	30	1.5	mills/kWh	68.8	42.8	37	33	31.8	29.4
		2.0		86.3	53.3	45.8	40.7	39.2	36.1
		1.0		34.2	22.3	19.7	17.9	17.3	16.3
3.0	50	1.5	mills/kWh	44.7	28.6	25	22.5	21.8	20.3
		2.0		55.2	34.9	30.3	27.1	26.2	24.3
		1.0		26.8	18	16	14.7	14.3	13.5
3.0	70	1.5	mills/kWh	34.3	22.5	19.8	18	17.5	16.4
		2.0		41.8	26.9	23.6	21.3	20.6	19.3
		1.0		53	34	29.7	26.8	25.9	24.3
4.0	30	1.5	mills/kWh	70.8	44.7	38.8	34.8	33.5	31.2
		2.0		88.6	55.4	47.9	42.7	41.1	38.2
		1.0		35.6	23.7	21	19.2	18.6	17.6
4.0	50	1.5	mills/kWh	46.3	30.1	26.5	24	23.2	21.8
		2.0		57	36.5	31.9	28.7	27.7	25.9
		1.0		28.1	19.2	17.3	16	15.5	14.8
4.0	70	1.5	mills/kWh	33.7	23.8	21.2	19.3	18.7	17.7
		2.0		43.3	28.4	25.1	22.7	22	20.7

L/S-FGD
SO2 COST EFFECTIVENESS ($/TON)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		6282.2	3988.9	3340	2993.2	2885.3	2668.1
1.0	30	1.5	$/TON	8450.5	5334.1	4434	3953.7	3813.5	3508.2
		2.0		10618.6	6679.3	5528.1	4914.2	4761.6	4348.4
		1.0		4125.1	2984.6	2280.2	2064.2	1993.9	1859.6
1.0	50	1.5	$/TON	5426	3491.7	2936.7	2640.5	2550.8	2363.7
		2.0		6726.9	4298.8	3593.1	3216.9	3107.7	2867.8
		1.0		3189.5	2122.3	1824.5	1665.4	1611.9	1513.6
1.0	70	1.5	$/TON	4118.7	2698.8	2293.4	2077.1	2009.7	1873.6
		2.0		5047.9	3275.3	2762.3	2488.7	2407.5	2233.7
		1.0		3372.6	2109.2	1787.2	1607	1571.4	1455.6
2.0	30	1.5	$/TON	4533	2802.7	2356.4	2106	2063	1899.7
		2.0		5693.5	3496.2	2925.6	2605.1	2554.6	2343.9
		1.0		2230.1	1438	1237.3	1125.2	1101.3	1029.8
2.0	50	1.5	$/TON	2926.4	1854.1	1578.8	1424.6	1396.3	1296.3
		2.0		3622.7	2270.2	1920.3	1724	1691.2	1562.8
		1.0		1735	1148.6	1000.8	918.3	899.8	847.5
2.0	70	1.5	$/TON	2232.3	1445.9	1244.8	1132.2	1110.5	1037.9
		2.0		2729.7	1743.1	1488.7	1346.1	1321.2	1228.2
		1.0		2357.1	1487.7	1292	1162	1120	1042.5
3.0	30	1.5	$/TON	3160.2	1966.6	1696.9	1515	1459.4	1350.1
		2.0		3963.3	2445.5	2101.9	1867.9	1798.8	1657.7
		1.0		1570	1025.5	903.4	822.7	795.5	747.6
3.0	50	1.5	$/TON	2051.9	1312.8	1146.4	1034.5	999.2	932.2
		2.0		2533.7	1600.1	1389.4	1246.2	1202.8	1116.8
		1.0		1229.1	826.5	736.6	677	656.5	621.4
3.0	70	1.5	$/TON	1573.3	1031.5	910	828.3	801.9	753.3
		2.0		1917.5	1236.7	1083.5	979.6	947.4	885.1
		1.0		1825.9	1171.3	1023.6	923.8	890.9	835.7
4.0	30	1.5	$/TON	2438.8	1539.9	1333.9	1196.5	1152.7	1074.9
		2.0		3051.8	1908.6	1648.2	1469.5	1414.5	1314
		1.0		1225.8	815.8	723.7	661.8	640.6	606.5
4.0	50	1.5	$/TON	1593.6	1037	911.1	825.4	797.7	749.9
		2.0		1961.3	1258.2	1098.5	989	954.7	893.4
		1.0		965.9	662.6	594.8	549.3	533.3	508.3
4.0	70	1.5	$/TON	1228.6	820.6	728.7	666.2	645.5	610.8
		2.0		1491.3	978.6	862.5	783.1	757.7	713.3

APPENDIX D

Integrated Gasification Combined Cycle

IGCC-$/kW

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	HEAT RATE (Btu/kWh)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
1.0	50	8000	$/kW	2,373	1,849	1,652	1,536	1,423	1,345
		10000		2,816	2,199	1,967	1,829	1,695	1,603
1.0	70	8000	$/kW	2,372	1,848	1,652	1,535	1,422	1,344
		10000		2,815	2,198	1,966	1,828	1,694	1,602
2.0	50	8000	$/kW	2,409	1,872	1,672	1,553	1,437	1,358
		10000		2,857	2,226	1,989	1,848	1,712	1,618
2.0	70	8000	$/kW	2,408	1,871	1,671	1,552	1,436	1,358
		10000		2,856	2,225	1,988	1,847	1,711	1,617
3.0	50	8000	$/kW	2,438	1,891	1,687	1,566	1,449	1,369
		10000		2,890	2,248	2,007	1,864	1,725	1,631
3.0	70	8000	$/kW	2,437	1,890	1,686	1,566	1,448	1,368
		10000		2,889	2,247	2,006	1,863	1,724	1,630
4.0	50	8000	$/kW	2,463	1,908	1,701	1,578	1,459	1,379
		10000		2,919	2,267	2,022	1,878	1,737	1,642
4.0	70	8000	$/kW	2462	1,907	1,700	1,577	1,459	1,378
		10000		2,918	2,266	2,021	1,877	1,736	1,641

IGCC-mills/kWh

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	HEAT RATE (Btu/kWh)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
1.0	50	8000	mills/kWh	171.3	131.2	118.7	111.7	105.2	100.8
		10000		201.4	156.8	142.3	134.2	126.7	121.6
1.0	70	8000	mills/kWh	130.7	102.0	93.1	88.0	83.3	80.3
		10000		154.4	122.3	112.0	106.2	100.8	97.1
2.0	50	8000	mills/kWh	172.9	132.3	119.5	112.4	105.7	101.2
		10000		203.4	157.9	143.2	134.9	127.2	122.2
2.0	70	8000	mills/kWh	131.8	102.7	93.6	88.4	83.7	80.5
		10000		155.6	123.0	112.5	106.8	101.2	94.5
3.0	50	8000	mills/kWh	174.1	133.0	120.1	112.9	106.1	101.5
		10000		204.8	158.7	143.9	135.6	127.8	122.5
3.0	70	8000	mills/kWh	132.7	103.3	94.0	88.7	83.1	80.7
		10000		156.4	123.7	113.1	107.1	101.5	97.8
4.0	50	8000	mills/kWh	175.2	133.7	120.6	113.2	106.4	101.9
		10000		206.0	159.4	144.6	136.0	128.1	122.9
4.0	70	8000	mills/kWh	133.4	103.6	94.2	88.9	84.0	80.9
		10000		157.3	124.1	113.4	107.5	101.7	98.0

APPENDIX E

Atmospheric Fluidized Bed

AFBC-$/kW

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	HEAT RATE (Btu/kWh)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
1.0	50	9000	$/kW	2,018	1,554	1,386	1,289	1,195	1,132
		11000		2,346	1,814	1,621	1,509	1,401	1,328
1.0	70	9000	$/kW	2,018	1,554	1,386	1,289	1,195	1,132
		11000		2,346	1,814	1,621	1,509	1,401	1,328
2.0	50	9000	$/kW	2,031	1,562	1,393	1,295	1,200	1,137
		11000		2,361	1,823	1,628	1,515	1,407	1,333
2.0	70	9000	$/kW	2,031	1,562	1,393	1,295	1,200	1,137
		11000		2,361	1,823	1,628	1,515	1,407	1,333
3.0	50	9000	$/kW	2,042	1,568	1,398	1,299	1,205	1,141
		11000		2,372	1,830	1,634	1,521	1,412	1,338
3.0	70	9000	$/kW	2,042	1,568	1,398	1,299	1,205	1,141
		11000		2,372	1,830	1,634	1,521	1,412	1,338
4.0	50	9000	$/kW	2,050	1,574	1,403	1,304	1,209	1,145
		11000		2,382	1,836	1,640	1,526	1,416	1,342
4.0	70	9000	$/kW	2,050	1,574	1,403	1,304	1,209	1,145
		11000		2,382	1,836	1,640	1,526	1,416	1,342

AFBC-mills/kWh

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	HEAT RATE (Btu/kWh)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
1.0	50	9000	mills/kWh	148.9	118.8	108.7	102.9	97.3	93.8
		11000		172.9	139.5	127.9	121.5	115.3	111.1
1.0	70	9000	mills/kWh	116.4	95.4	87.7	83.5	79.6	77.0
		11000		135.8	111.8	103.6	98.9	94.5	91.7
2.0	50	9000	mills/kWh	151.7	121.3	110.0	105.2	99.8	96.1
		11000		176.2	142.5	130.7	124.3	118.1	113.9
2.0	70	9000	mills/kWh	119.0	97.3	90.0	85.8	81.9	79.3
		11000		138.8	114.6	106.4	101.7	97.3	94.3
3.0	50	9000	mills/kWh	154.2	123.6	113.2	107.5	102.0	98.4
		11000		179.4	145.3	133.7	127.0	120.8	116.7
3.0	70	9000	mills/kWh	121.5	99.6	92.2	88.0	84.0	81.6
		11000		141.8	117.4	109.2	104.5	100.0	97.0
4.0	50	9000	mills/kWh	156.8	126.0	115.5	109.7	104.3	100.6
		11000		182.4	148.1	136.5	129.9	123.6	119.4
4.0	70	9000	mills/kWh	123.7	101.9	94.3	90.1	86.3	83.7
		11000		144.6	120.2	111.8	107.1	102.7	99.8

APPENDIX F

Lime Spray Drying

LIME SPRAY DRYING WITH REUSE OF THE EXISTING ESPs
LSD+ESP (LARGE ESP:400 SCA)
CAPITAL COST ($/kW)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		209.8	106.1	84.8	75.5	68.5	64.7
1	30	1.5	$/kW	301.8	152.5	122.1	108.9	98.9	93.6
		2.0		393.7	198.9	159.4	142.3	129.4	122.5
		1.0		210.8	106.8	85.4	76.1	69.0	65.3
1	50	1.5	$/kW	302.8	153.2	122.7	109.5	99.5	94.2
		2.0		394.7	199.6	160.0	142.9	130.0	123.1
		1.0		211.8	107.5	86.0	76.6	69.6	65.8
1	70	1.5	$/kW	303.7	153.9	123.3	110.0	100.0	94.7
		2.0		395.6	200.3	160.6	143.4	130.5	123.6
		1.0		231.6	117.6	94.2	84.0	76.3	72.2
2	30	1.5	$/kW	333.7	169.2	135.7	121.2	110.3	104.4
		2.0		435.8	220.9	177.3	158.4	144.2	136.6
		1.0		233.5	118.9	95.3	85.1	77.3	73.2
2	50	1.5	$/kW	335.6	170.6	136.9	122.3	111.3	105.4
		2.0		437.7	222.2	178.4	159.5	145.3	137.7
		1.0		235.2	120.2	96.5	86.2	78.4	74.2
2	70	1.5	$/kW	337.3	171.8	138.0	123.4	112.4	106.4
		2.0		439.4	223.5	179.6	160.6	146.3	138.7
		1.0		255.2	130.2	104.5	93.4	85.0	80.5
3	30	1.5	$/kW	368.4	187.6	150.8	134.9	122.9	116.4
		2.0		481.6	245.1	197.1	176.4	160.8	152.4
		1.0		257.8	132.1	106.2	95.0	86.5	82.0
3	50	1.5	$/kW	371.0	189.5	152.5	136.5	124.4	117.9
		2.0		484.2	247.0	198.8	178.0	162.3	153.9
		1.0		260.3	133.9	107.9	96.6	88.0	87.5
3	70	1.5	$/kW	373.5	191.4	154.2	138.1	125.9	123.5
		2.0		486.7	248.8	200.5	179.6	163.8	159.5
		1.0		279.6	143.2	115.3	103.1	94.0	89.1
4	30	1.5	$/kW	404.3	206.7	166.5	149.1	136.0	129.0
		2.0		529.0	270.2	217.7	195.1	178.1	168.9
		1.0		282.9	145.7	117.5	105.3	96.0	95.1
4	50	1.5	$/kW	407.6	209.2	168.7	151.2	138.0	135.0
		2.0		532.3	272.7	220.0	197.2	180.1	175.0
		1.0		286.0	148.1	119.7	107.3	98.0	97.3
4	70	1.5	$/kW	410.7	211.5	170.9	153.3	140.0	137.3
		2.0		535.4	275.0	222.1	199.3	182.1	177.2

LIME SPRAY DRYING WITH REUSE OF THE EXISTING ESPs
LSD+ESP (LARGE ESP:400 SCA)
ANNUAL COST (mills/kWh)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		30	15.7	12.8	11.5	10.7	10.3
1	30	1.5	mills/kWh	39.9	20.7	16.8	15.1	13.9	13.4
		2.0		49.8	25.7	20.8	18.7	17.2	16.5
		1.0		19.8	10.7	8.9	8.1	7.6	7.4
1	50	1.5	mills/kWh	25.7	13.7	11.3	10.3	9.5	9.2
		2.0		31.6	16.7	13.7	12.4	11.5	11.1
		1.0		15.5	8.7	7.4	6.8	6.4	6.3
1	70	1.5	mills/kWh	19.8	10.9	9.1	8.3	7.8	7.7
		2.0		24	13	10.8	9.9	9.2	9
		1.0		32.8	17.4	14.2	12.9	11.9	11.5
2	30	1.5	mills/kWh	43.8	22.9	18.7	16.9	15.6	15
		2.0		54.8	28.5	23.2	20.9	19.2	18.4
		1.0		22	12.2	10.2	9.4	8.8	8.6
2	50	1.5	mills/kWh	28.6	15.6	12.9	11.8	11	10.7
		2.0		35.2	18.9	15.6	14.2	13.2	12.8
		1.0		17.7	10.4	8.9	8.3	7.9	7.7
2	70	1.5	mills/kWh	22.4	12.8	10.8	10	9.4	9.2
		2.0		27.1	15.1	12.7	11.7	11	10.7
		1.0		35.8	19.2	15.8	14.3	13.3	12.8
3	30	1.5	mills/kWh	48	25.4	20.8	18.8	17.4	16.7
		2.0		60.1	31.5	25.7	23.3	21.5	20.6
		1.0		24.3	13.8	11.7	10.8	10.2	9.9
3	50	1.5	mills/kWh	31.6	17.5	14.7	13.5	12.6	12.3
		2.0		38.9	21.2	17.7	16.2	15.1	14.6
		1.0		19.9	12	10.5	9.8	9.4	9.4
3	70	1.5	mills/kWh	25.1	14.7	12.6	11.7	11.1	11
		2.0		30.3	17.3	14.7	13.6	12.9	12.7
		1.0		38.8	21	17.4	15.8	14.7	14.2
4	30	1.5	mills/kWh	52.3	27.9	22.9	20.8	19.2	18.5
		2.0		65.7	34.7	28.4	25.7	23.8	22.8
		1.0		26.7	15.4	13.2	12.2	11.5	11.5
4	50	1.5	mills/kWh	34.7	19.5	16.5	15.2	14.2	14.1
		2.0		42.8	23.6	19.8	18.1	17	16.7
		1.0		22.1	13.7	12	11.3	10.9	10.9
4	70	1.5	mills/kWh	27.9	16.7	14.4	13.5	12.8	12.7
		2.0		33.7	19.6	16.8	15.6	14.7	14.6

LIME SPRAY DRYING WITH REUSE OF THE EXISTING ESPs
LSD+ESP (LARGE ESP:400 SCA)
SO2 COST EFFECTIVENESS ($/TON)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		4868.4	2549.9	2074.5	1872.7	1730.5	1667.8
1	30	1.5	$/TON	6476.4	3361.7	2727	2457	2263.5	2173.3
		2.0		8084.6	4173.5	3379.5	3041.3	2796.6	2678.8
		1.0		3208.9	1738.3	1439.4	1314.6	1230.5	1198.8
1	50	1.5	$/TON	4173.7	2225.3	1830.9	1665.2	1550.3	1502.1
		2.0		5138.6	2712.4	2222.4	2015.8	1870.2	1805.4
		1.0		2526.2	1419.4	1196.1	1104.3	1045	1026.7
1	70	1.5	$/TON	3215.4	1767.3	1475.8	1354.7	1273.5	1243.3
		2.0		3904.6	2115.2	1755.4	1605.1	1501.9	1460
		1.0		2661.6	1411.3	1155	1046.2	969.1	934.7
2	30	1.5	$/TON	3554.6	1863	1518.4	1371.7	1266.3	1216.7
		2.0		4447.6	2314.6	1881.8	1697.3	1563.5	1498.6
		1.0		1784.5	993.1	832	764.6	718.9	701.2
2	50	1.5	$/TON	2320.3	1264.1	1050	960	897.2	870.3
		2.0		2856.1	1535	1268.1	1155.3	1075.5	1039.5
		1.0		1436.6	842.2	721.9	672.3	639.9	629.4
2	70	1.5	$/TON	1819.3	1035.7	877.6	811.8	767.3	750.2
		2.0		2202	1229.3	1033.4	951.3	894.6	871
		1.0		1936.6	1038.1	854.1	775.8	720.3	695.2
3	30	1.5	$/TON	2596.7	1373	1123.9	1017.8	941.4	905
		2.0		3256.7	1707.8	1393.7	1259.8	1162.4	1114.8
		1.0		1315.9	748.3	632.8	584.3	551.2	538.1
3	50	1.5	$/TON	1711.9	949.3	794.7	729.5	683.9	664
		2.0		2107.9	1150.2	956.6	874.7	816.5	789.9
		1.0		1077.7	652.3	566.1	530.4	506.9	507.9
3	70	1.5	$/TON	1360.5	795.8	681.7	634.1	601.7	597.9
		2.0		1643.4	939.3	797.3	737.8	696.4	687.8
		1.0		1577.5	853.6	705.4	642.4	597.5	577
4	30	1.5	$/TON	2122.9	1131.1	929.4	843.5	781.4	751.6
		2.0		2668.3	1408.7	1153.4	1044.5	965.2	926.2
		1.0		1083.5	627.2	534.2	495.2	468.4	466.9
4	50	1.5	$/TON	1410.7	793.7	668.6	615.8	578.7	571.6
		2.0		1737.9	960.2	803	736.4	689	676.4
		1.0		899.5	558.1	488.8	460.1	441.1	441.5
4	70	1.5	$/TON	1133.2	677.1	584.8	546.3	519.9	516.4
		2.0		1366.9	796.1	680.8	632.4	598.6	591.2

LIME SPRAY DRYING WITH A NEW BAGHOUSE
LSD+FF
CAPITAL COST ($/kW)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		315.6	191.0	165.4	158.5	149.5	149.2
1	30	1.5	$/kW	407.2	237.3	202.6	191.7	179.8	178.0
		2.0		498.8	283.5	239.8	225.0	210.2	206.8
		1.0		316.4	191.7	166.0	159.1	150.0	149.8
1	50	1.5	$/kW	408.0	237.9	203.2	192.3	180.4	178.6
		2.0		499.6	284.2	240.4	225.6	210.7	207.4
		1.0		317.3	192.4	166.6	159.6	150.6	150.4
1	70	1.5	$/kW	408.9	238.6	203.8	192.9	180.9	179.2
		2.0		500.5	284.8	241.0	226.2	211.3	208.0
		1.0		337.1	202.4	174.7	166.9	157.2	156.7
2	30	1.5	$/kW	438.8	253.8	216.1	203.9	191.1	188.8
		2.0		540.6	305.3	257.5	241.0	224.9	220.9
		1.0		338.7	203.6	175.8	167.9	158.2	157.8
2	50	1.5	$/kW	440.5	255.0	217.2	205.0	192.1	189.9
		2.0		542.2	306.5	258.6	242.1	225.9	222.1
		1.0		340.3	204.8	176.9	169.0	165.4	159.0
2	70	1.5	$/kW	442.1	256.3	218.3	206.1	199.2	191.1
		2.0		543.8	307.7	259.7	243.1	233.0	223.2
		1.0		360.5	214.8	185.0	176.4	165.8	165.0
3	30	1.5	$/kW	473.2	272.0	231.1	217.7	203.6	200.9
		2.0		586.0	329.3	277.2	259.0	241.3	236.7
		1.0		362.8	216.6	186.6	177.9	173.4	166.7
3	50	1.5	$/kW	475.6	273.8	232.7	219.2	211.2	202.5
		2.0		588.4	331.0	278.8	260.6	249.0	238.4
		1.0		365.1	218.3	188.2	179.4	175.1	168.3
3	70	1.5	$/kW	477.9	275.6	234.3	220.8	212.9	204.2
		2.0		590.7	332.8	280.4	262.1	250.7	240.0
		1.0		384.6	227.7	195.6	186.0	174.7	173.7
4	30	1.5	$/kW	508.9	291.0	246.7	231.8	216.6	213.5
		2.0		633.1	354.2	297.7	277.6	258.5	253.2
		1.0		387.7	230.1	197.7	188.1	183.0	175.9
4	50	1.5	$/kW	511.9	293.3	248.8	233.9	224.9	215.6
		2.0		636.2	356.5	299.8	279.7	266.8	255.4
		1.0		390.6	232.3	199.8	190.0	185.3	178.0
4	70	1.5	$/kW	514.9	295.6	250.8	235.8	227.1	217.8
		2.0		639.2	358.8	301.9	281.6	269.0	257.5

LIME SPRAY DRYING WITH A NEW BAGHOUSE
LSD+FF
ANNUAL COST (mills/kWh)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		40.2	24.0	20.7	19.7	18.6	18.5
1	30	1.5	mills/kWh	50.1	19.0	24.7	23.3	21.9	21.6
		2.0		59.9	34.0	28.7	26.8	25.1	24.7
		1.0		26.1	15.9	13.8	13.2	12.5	12.6
1	50	1.5	mills/kWh	32.0	18.9	16.2	15.3	14.5	14.4
		2.0		37.9	21.9	18.6	17.5	16.5	16.3
		1.0		20.2	12.6	11.1	10.6	10.1	10.2
1	70	1.5	mills/kWh	24.5	14.7	12.8	12.1	11.5	11.5
		2.0		28.7	16.9	14.5	13.7	12.9	12.8
		1.0		43.0	25.7	22.1	21.0	19.9	19.8
2	30	1.5	mills/kWh	53.9	31.2	26.6	25.0	23.5	23.2
		2.0		64.9	36.8	31.0	29.0	27.2	26.7
		1.0		28.3	17.4	15.2	14.5	13.8	13.8
2	50	1.5	mills/kWh	34.8	20.7	17.8	16.9	16.0	15.9
		2.0		41.4	24.0	20.5	19.3	18.2	17.9
		1.0		22.3	14.2	12.6	12.0	11.8	11.6
2	70	1.5	mills/kWh	27.0	16.6	14.5	13.7	13.4	13.1
		2.0		31.7	19.0	16.4	15.5	14.9	14.6
		1.0		45.9	27.4	23.7	22.5	21.2	21.1
3	30	1.5	mills/kWh	58.1	33.6	28.6	26.9	25.3	25.0
		2.0		70.2	39.8	33.6	31.4	29.4	28.8
		1.0		30.6	19.0	16.6	15.8	15.5	15.1
3	50	1.5	mills/kWh	37.8	22.7	19.6	18.5	17.9	17.4
		2.0		45.1	26.4	22.5	21.2	20.3	19.7
		1.0		24.5	15.9	14.1	13.5	13.3	13.0
3	70	1.5	mills/kWh	29.7	18.5	16.2	15.4	15.0	14.7
		2.0		34.9	21.1	18.3	17.4	16.8	16.3
		1.0		49.0	29.3	25.2	23.9	22.6	22.5
4	30	1.5	mills/kWh	62.4	36.1	30.7	28.9	27.1	26.8
		2.0		75.7	42.9	36.2	33.8	31.6	31.0
		1.0		32.9	20.6	18.0	17.2	16.8	16.4
4	50	1.5	mills/kWh	40.9	24.6	21.3	20.2	19.5	19.0
		2.0		49.0	28.7	24.6	23.1	22.2	21.5
		1.0		26.7	17.5	15.6	15.0	14.8	14.5
4	70	1.5	mills/kWh	32.5	20.4	18.0	17.2	16.7	16.3
		2.0		38.2	23.4	20.4	19.3	18.6	18.2

LIME SPRAY DRYING WITH A NEW BAGHOUSE
LSD+FF
SO2 COST EFFECTIVENESS ($/TON)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		1.0		5790.9	3460.8	2981.6	2834.1	2681.0	2672.0
1	30	1.5	$/TON	7211.6	4177.9	3558.0	3350.2	3151.8	3118.5
		2.0		8632.4	4895.0	4134.3	3866.2	3622.6	3564.9
		1.0		3757.0	2289.8	1990.1	1898.2	1807.3	1807.9
1	50	1.5	$/TON	4609.5	2720.1	2335.9	2207.9	2089.8	2075.8
		2.0		5462.0	3150.3	2681.7	2517.5	2372.3	2343.7
		1.0		2913.5	1816.3	1593.5	1525.4	1461.2	1465.9
1	70	1.5	$/TON	3522.3	2123.7	1840.5	1746.6	1663.0	1657.3
		2.0		4131.3	2431.0	2087.5	1967.8	1864.8	1848.6
		1.0		3094.5	1849.4	1593.7	1513.0	1431.3	1425.0
2	30	1.5	$/TON	3883.5	2248.4	1914.6	1800.5	1693.7	1674.0
		2.0		4672.4	2647.4	2235.6	2088.1	1956.2	1922.9
		1.0		2035.5	1252.7	1092.7	1042.5	993.8	993.4
2	50	1.5	$/TON	2508.9	1492.1	1285.3	1215.0	1151.3	1142.8
		2.0		2982.3	1731.4	1477.9	1387.5	1308.8	1292.2
		1.0		1607.8	1023.2	904.3	867.1	850.4	834.6
2	70	1.5	$/TON	1945.9	1194.2	1041.8	990.3	962.9	941.3
		2.0		2284.1	1365.2	1179.4	1113.5	1075.4	1048.0
		1.0		2205.3	1318.1	1136.0	1078.1	1019.1	1013.6
3	30	1.5	$/TON	2788.5	1613.9	1374.3	1291.8	1214.4	1198.9
		2.0		3371.7	1909.7	1612.7	1505.6	1409.6	1384.2
		1.0		1467.4	910.4	796.5	760.5	742.0	724.4
3	50	1.5	$/TON	1817.3	1087.9	939.6	888.7	859.2	835.6
		2.0		2167.2	1265.4	1082.6	1017.0	976.3	946.8
		1.0		1176.9	761.2	676.6	649.9	637.4	626.0
3	70	1.5	$/TON	1426.9	888.0	778.8	741.5	721.0	705.4
		2.0		1676.8	1014.7	880.9	833.1	804.7	784.8
		1.0		1763.8	1054.2	908.8	861.5	814.5	809.3
4	30	1.5	$/TON	2245.7	1299.5	1106.7	1039.2	976.9	963.5
		2.0		2727.5	1544.7	1304.6	1216.8	1139.3	1117.7
		1.0		1185.3	740.3	649.4	620.0	604.9	590.7
4	50	1.5	$/TON	1474.5	887.4	768.2	726.6	702.3	683.2
		2.0		1763.6	1034.6	886.9	833.2	799.7	775.8
		1.0		962.8	630.9	563.5	541.7	531.4	522.2
4	70	1.5	$/TON	1169.3	736.0	648.3	617.8	601.0	588.3
		2.0		1375.8	841.1	733.1	693.9	670.6	654.4

APPENDIX G

Selective Catalytic Reduction

POST-COMBUSTION NOx CONTROLS:
SELECTIVE CATALYTIC REDUCTION
CAPITAL COST ($/kW)

CATALYST LIFE (YEARS)	NOx REDUCTION (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
			1.00		133	103.2	91.5	88.6	86.6	85.4
1	70	30	1.50	$/kW	176.7	133	118.5	114.4	111.5	109.8
			2.00		220.5	162.8	145.5	140.3	136.4	134.2
			1.00		143	111	98.9	95.9	93.7	92.4
1	80	30	1.50	$/kW	191.8	144.7	129.6	125.3	122.1	120.3
			2.00		240.5	178.4	160.3	154.7	150.6	148.2
			1.00		133	103.2	91.5	88.7	86.6	85.4
1	70	50	1.50	$/kW	176.8	133	118.5	114.5	111.6	109.9
			2.00		220.5	162.8	145.5	140.3	136.5	134.3
			1.00		143.1	111.1	99	95.9	93.8	92.5
1	80	50	1.50	$/kW	191.8	144.8	129.6	125.3	122.2	120.4
			2.00		240.6	178.5	160.3	154.7	150.6	148.3
			1.00		133	103.2	91.6	88.7	86.7	85.5
1	70	70	1.50	$/kW	176.8	133	118.6	114.5	111.6	109.9
			2.00		220.6	162.8	145.6	140.3	136.5	134.3
			1.00		143.1	111.1	99	96	93.8	92.5
1	80	70	1.50	$/kW	191.9	144.8	129.7	125.4	122.2	120.4
			2.00		240.6	178.5	160.4	154.8	150.7	148.3
			1.00		133	103.2	91.5	88.6	86.6	85.4
3	70	30	1.50	$/kW	176.7	133	118.5	114.4	111.5	109.8
			2.00		220.5	162.8	145.5	140.3	136.4	134.2
			1.00		143	111	98.9	95.9	93.7	92.4
3	80	30	1.50	$/kW	191.8	144.7	129.6	125.3	122.1	120.3
			2.00		240.5	178.4	160.3	154.7	150.6	148.2
			1.00		133	103.2	91.5	88.7	86.6	85.4
3	70	50	1.50	$/kW	176.8	133	118.5	114.5	111.6	109.8
			2.00		220.5	162.8	145.5	140.3	136.5	134.3
			1.00		143.1	111.1	98.9	95.9	93.7	92.5
3	80	50	1.50	$/kW	191.8	144.8	129.6	125.3	122.2	120.4
			2.00		240.6	178.5	160.3	154.7	150.6	148.3
			1.00		133	103.2	91.5	88.7	86.7	85.5
3	70	70	1.50	$/kW	176.8	133	118.6	114.5	111.6	109.9
			2.00		220.6	162.8	145.6	140.3	136.5	134.3
			1.00		143.1	111.1	99	96	93.8	92.5
3	80	70	1.50	$/kW	191.9	144.8	129.7	125.4	122.2	120.4
			2.00		240.6	178.5	160.4	154.8	150.7	148.3

POST-COMBUSTION NOx CONTROLS:
SELECTIVE CATALYTIC REDUCTION
ANNUAL COST (mills/kWh)

CATALYST LIFE (YEARS)	NOx REDUCTION (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
			1.00		29.4	26.1	25	24.7	24.5	24.3
1	70	30	1.50	mills/kWh	33.4	28.8	27.5	27.1	26.8	26.6
			2.00		37.5	31.6	30	29.5	29.1	28.9
			1.00		30.4	26.9	25.8	25.4	25.2	25.1
1	80	30	1.50	mills/kWh	34.9	30	28.6	28.2	27.9	27.6
			2.00		39.4	33.1	31.5	30.9	30.5	30.2
			1.00		17.9	15.9	15.3	15.1	14.9	14.9
1	70	50	1.50	mills/kWh	20.3	17.6	16.8	16.5	16.3	16.2
			2.00		22.8	19.2	18.3	17.9	17.7	17.6
			1.00		18.5	16.4	15.7	15.5	15.4	15.3
1	80	50	1.50	mills/kWh	21.2	18.3	17.4	17.2	17	16.9
			2.00		23.9	20.2	19.2	18.8	18.6	18.4
			1.00		13	11.6	11.1	11	10.9	10.8
1	70	70	1.50	mills/kWh	14.7	12.7	12.2	12	11.9	11.8
			2.00		16.5	13.9	13.3	13	12.9	12.8
			1.00		13.4	11.9	11.5	11.3	11.2	11.1
1	80	70	1.50	mills/kWh	15.4	13.3	12.7	12.5	12.3	12.3
			2.00		17.3	14.6	13.9	13.6	13.5	13.4
			1.00		18.4	15.1	14.1	13.7	13.6	13.4
3	70	30	1.50	mills/kWh	22.5	17.9	16.6	16.1	15.9	15.7
			2.00		26.6	20.7	19.1	18.5	18.2	17.9
			1.00		19.4	15.9	14.8	14.5	14.3	14.1
3	80	30	1.50	mills/kWh	23.9	19.1	17.7	17.2	16.9	16.7
			2.00		28.5	22.2	20.5	19.9	19.6	19.3
			1.00		11.3	9.3	8.7	8.5	8.4	8.3
3	70	50	1.50	mills/kWh	13.7	11	10.2	9.9	9.8	9.7
			2.00		16.2	12.7	11.7	11.4	11.2	11
			1.00		11.9	9.8	9.2	9	8.8	8.7
3	80	50	1.50	mills/kWh	14.6	11.7	10.9	10.6	10.4	10.3
			2.00		17.4	13.6	12.6	12.2	12	11.9
			1.00		8.3	6.9	6.4	6.3	6.2	6.1
3	70	70	1.50	mills/kWh	10	8	7.5	7.3	7.2	7.1
			2.00		11.8	9.2	8.6	8.3	8.2	8.1
			1.00		8.7	7.2	6.8	6.6	6.5	6.5
3	80	70	1.50	mills/kWh	10.7	8.6	8	7.8	7.7	7.6
			2.00		12.6	9.9	9.2	9	8.8	8.7

POST-COMBUSTION NOx CONTROLS:
SELECTIVE CATALYTIC REDUCTION
SO2 COST EFFECTIVENESS ($/TON)

CATALYST LIFE (YEARS)	NOx REDUCTION (%)	CAPACITY FACTOR (%)	RETROFIT FACTOR (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
			1.00		9657.6	8576	8230.3	8119.6	8056.6	8005.8
1	70	30	1.50	$/TON	10996.2	9487.1	9055.9	8908.7	8818.2	8752.7
			2.00		12334.1	10398.1	9881.6	9697.8	9579.8	9499.5
			1.00		8735.4	7730.2	7415.9	7314.1	7255.2	7208.8
1	80	30	1.50	$/TON	10039.9	8631.8	8236.9	8100.6	8015.7	7955.4
			2.00		11344.2	9533.3	9057.9	8887.1	8776.3	8702.1
			1.00		5877.9	5228	5020.4	4954	4916.1	4885.6
1	70	50	1.50	$/TON	6681	5774.8	5515.8	5427.4	5373.1	5333.8
			2.00		7483.7	6321.4	6011.2	5900.8	5830	5781.9
			1.00		5320.6	4716.7	4528	4466.8	4431.4	4403.6
1	80	50	1.50	$/TON	6103.1	5257.7	5020.6	4938.7	4887.8	4851.5
			2.00		6885.8	5798.6	5513.2	5410.6	5344.1	5299.5
			1.00		4263.8	3799	3650.6	3603.1	3576	3554.2
1	70	70	1.50	$/TON	4837.4	4189.5	4004.4	3941.2	3902.4	3874.3
			2.00		5410.8	4579.9	4358.3	4279.4	4228.8	4194.3
			1.00		3862.3	3430.3	3295.4	3251.6	3226.3	3206.4
1	80	70	1.50	$/TON	4421.4	3816.7	3647.2	3588.7	3552.3	3526.4
			2.00		4980.4	4203.1	3999.1	3925.7	3878.2	3846.3
			1.00		6058.5	4977.1	4631.3	4520.7	4457.7	4406.9
3	70	30	1.50	$/TON	7396.9	5888.3	5457	5309.8	5219.3	5153.8
			2.00		8734.9	6799.4	6282.6	6098.9	5980.9	5900.6
			1.00		5586.1	4581.2	4266.8	4165.1	4106.1	4059.7
3	80	30	1.50	$/TON	6890.5	5482.8	5087.8	4951.6	4866.7	4806.4
			2.00		8195	6384.3	5908.8	5738	5627.2	5553
			1.00		3718.5	3068.7	2861.1	2794.6	2756.8	2726.3
3	70	50	1.50	$/TON	4521.5	3615.5	3356.5	3268.1	3213.7	3174.4
			2.00		5324.2	4162.1	3851.9	3741.5	3670.7	3622.5
			1.00		3431	2827.3	2638.5	2577.4	2542	2514.1
3	80	50	1.50	$/TON	4213.6	3368.3	3131.1	3049.3	2998.3	2962.1
			2.00		4996.3	3909.2	3623.7	3521.2	3454.7	3410.1
			1.00		2721.4	2256.7	2108.2	2060.7	2033.6	2011.8
3	70	70	1.50	$/TON	3295	2647.2	2462	2398.9	2360	2331.9
			2.00		3868.4	3037.6	2815.9	2737	2686.4	2652
			1.00		2512.7	2080.8	1945.8	1902.1	1876.7	1856.8
3	80	70	1.50	$/TON	3071.7	2467.2	2297.6	2239.1	2202.7	2176.8
			2.00		3630.8	2853.6	2649.5	2576.2	2528.6	2496.8

APPENDIX H

Furnace Sorbent Injection

FURNACE SORBENT INJECTION AND HUMIDIFICATION
FSI+ESP (LARGE ESP:400 SCA)
CAPITAL COST ($/kW)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	SO2 REMOVAL (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		30		56	34.8	28.4	26.6	24.6	23.6
1	30	50	$/kW	56.9	35.2	28.7	26.9	24.9	23.8
		70		57.7	35.7	29.1	27.2	25.1	24.1
		30		56	34.8	28.4	26.6	24.6	23.6
1	50	50	$/kW	56.9	35.2	28.8	26.9	24.9	23.8
		70		57.7	35.7	29.1	27.2	25.1	24.1
		30		56	34.8	28.4	26.6	24.6	23.6
1	70	50	$/kW	56.9	35.3	28.8	26.9	24.9	23.8
		70		57.7	35.7	29.1	27.2	25.1	24.1
		30		64.3	41.3	34.9	32.4	27.8	26.8
2	30	50	$/kW	65.5	42	35.4	32.9	28.2	27.2
		70		66.5	42.6	35.9	33.4	28.6	27.6
		30		64.3	41.3	34.9	32.5	27.8	26.8
2	50	50	$/kW	65.5	42	35.4	32.9	28.2	27.2
		70		66.6	42.6	35.9	33.4	28.6	27.6
		30		64.3	41.3	34.9	32.5	27.8	26.9
2	70	50	$/kW	65.5	42	35.4	32.9	28.2	27.2
		70		66.6	42.6	35.9	33.4	28.6	27.6
		30		71.6	48	37	34.3	32.4	31.7
3	30	50	$/kW	73	48.9	37.7	34.9	32.9	32.3
		70		74.3	49.7	38.4	35.5	33.4	32.7
		30		71.6	48	37.1	34.3	32.4	31.8
3	50	50	$/kW	73	48.9	37.8	34.9	32.9	32.3
		70		74.3	49.7	38.4	35.5	33.4	32.7
		30		71.6	48.1	37.1	34.3	32.4	31.8
3	70	50	$/kW	73	48.9	37.8	34.9	32.9	32.3
		70		74.3	49.7	38.4	35.5	33.4	32.8
		30		78.2	54.4	41.7	38.9	37.3	38.6
4	30	50	$/kW	79.9	55.4	42.6	39.7	37.9	39.3
		70		81.4	56.3	43.3	40.4	38.6	39.9
		30		78.2	54.4	41.7	39	37.3	38.7
4	50	50	$/kW	79.9	55.4	42.6	39.7	38	39.3
		70		81.4	56.3	43.3	40.4	38.6	39.9
		30		78.3	54.4	41.8	39	37.3	38.7
4	70	50	$/kW	79.9	55.4	42.6	39.7	38	39.3
		70		81.4	56.3	43.4	40.4	38.6	39.9

FURNACE SORBENT INJECTION AND HUMIDIFICATION
FSI+ESP (LARGE ESP:400 SCA)
ANNUAL COST (mills/kWh)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	SO2 REMOVAL (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		30		12.1	7.2	6	5.6	5.3	5.1
1	30	50	mills/kWh	12.2	7.3	6.1	5.7	5.3	5.1
		70		12.4	7.4	6.2	5.8	5.4	5.2
		30		8.3	5.3	4.6	4.4	4.1	4
1	50	50	mills/kWh	8.4	5.4	4.7	4.4	4.2	4.1
		70		8.5	5.5	4.8	4.5	4.3	4.2
		30		6.6	4.5	4	3.8	3.7	3.6
1	70	50	mills/kWh	6.7	4.6	4.1	3.9	3.7	3.7
		70		6.8	4.7	4.2	4	3.8	3.7
		30		15.2	10.1	8.9	8.5	7.8	7.6
2	30	50	mills/kWh	15.4	10.3	9.1	8.6	8	7.8
		70		15.6	10.5	9.3	8.8	8.1	7.9
		30		11	8	7.3	7	6.6	6.5
2	50	50	mills/kWh	11.2	8.2	7.4	7.2	6.8	6.6
		70		11.4	8.3	7.6	7.3	6.9	6.8
		30		9.3	7.1	6.6	6.4	6.1	6
2	70	50	mills/kWh	9.4	7.2	6.7	6.5	6.2	6.2
		70		9.6	7.4	6.9	6.7	6.4	6.3
		30		18.1	13	11.6	10.9	10.5	10.4
3	30	50	mills/kWh	18.4	13.3	11.6	11.1	10.7	10.6
		70		18.7	13.5	11.9	11.3	11	10.8
		30		13.7	10.7	9.7	9.4	9.2	9.1
3	50	50	mills/kWh	14	10.9	9.9	9.6	9.4	9.3
		70		14.3	11.1	10.1	9.8	9.6	9.5
		30		11.9	9.7	9	8.8	8.6	8.5
3	70	50	mills/kWh	12.1	9.9	9.2	9	8.8	8.7
		70		12.3	10.1	9.4	9.2	9	8.9
		30		21	15.9	14.1	13.6	13.3	13.3
4	30	50	mills/kWh	21.4	16.2	14.4	13.9	13.5	13.6
		70		21.8	16.6	14.7	14.2	13.8	13.9
		30		16.4	13.3	12.3	12	11.7	11.8
4	50	50	mills/kWh	16.7	13.6	12.5	12.2	12	12.1
		70		17.1	13.9	12.8	12.5	12.3	12.3
		30		14.5	12.3	11.5	11.3	11.1	11.1
4	70	50	mills/kWh	14.8	12.5	11.8	11.5	11.4	11.4
		70		15.1	12.8	12	11.8	11.6	11.7

FURNACE SORBENT INJECTION AND HUMIDIFICATION
FSI+ESP (LARGE ESP:400 SCA)
SO2 COST EFFECTIVENESS ($/TON)

SULFUR CONTENT (%)	CAPACITY FACTOR (%)	SO2 REMOVAL (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		30		4995	2981.9	2499.3	2327.4	2172.8	2092
1	30	50	$/TON	3032.2	1814.8	1522.7	1418.3	1324.8	1275.6
		70		2189.7	1313.9	1103.6	1028.7	961.1	925.5
		30		3412.2	2200.2	1909.8	1806.4	1713.6	1665.2
1	50	50	$/TON	2074.3	1341.4	1165.7	1102.9	1046.9	1017.5
		70		1500.3	973.1	846.5	801.5	761	739.7
		30		2736.2	1867.6	1659.6	1585.4	1519.2	1484.7
1	70	50	$/TON	1665.3	1140.1	1014.2	969.2	929.3	908.2
		70		1205.8	828	737.3	705.2	676.3	661.1
		30		3129.9	2088.2	1847.7	1749.2	1614.5	1575.6
2	30	50	$/TON	1906.4	1275.3	1129.8	1069.7	988.2	964.4
		70		1381.1	926.5	821.7	778.2	719.5	702.2
		30		2278.7	1651.5	1507	1448	1367.1	1343.7
2	50	50	$/TON	1390.2	1010.3	923	887	838	823.6
		70		1008.9	735.2	672.4	646.4	611.1	600.6
		30		1914.9	1465.5	1362.2	1320.1	1262.3	1245.5
2	70	50	$/TON	1169.6	897.3	833.1	809.4	774.3	764
		70		849.8	653.7	609	590.4	565.1	557.6
		30		2494.9	1794.4	1569.6	1498.8	1448.6	1427.2
3	30	50	$/TON	1522.6	1098.1	961.8	918.6	888	874.8
		70		1105.2	799.3	701	669.8	647.5	637.8
		30		1892.7	1471.1	1336.4	1293.9	1263.6	1250.7
3	50	50	$/TON	1156.9	901.6	819.9	793.9	775.4	767.5
		70		841.2	657.3	598.4	579.6	566.1	560.3
		30		1635.4	1333.3	1237.2	1206.8	1185.2	1175.9
3	70	50	$/TON	1000.7	817.9	759.6	741	727.7	722
		70		728.6	596.9	554.7	541.2	531.6	527.4
		30		2171.4	1643.1	1456.4	1403.8	1368.7	1374.2
4	30	50	$/TON	1326.8	1006.4	893.3	861.1	839.6	842.6
		70		964.3	733.2	651.7	628.4	612.7	614.6
		30		1696	1378.5	1266.5	1234.8	1213.7	1216.8
4	50	50	$/TON	1038	845.5	777.6	758.2	745.2	746.9
		70		755.6	616.9	567.9	553.8	544.4	545.5
		30		1493	1265.7	1185.7	1163	1147.8	1150
4	70	50	$/TON	914.6	776.9	728.4	714.5	705.1	706.3
		70		666.4	567.3	532.2	522.1	515.3	516.1

APPENDIX I

Natural Gas Reburn

NATURAL GAS REBURN
CAPITAL COST ($/kW)

FUEL PRICE DIFFERENTIAL	CAPACITY FACTOR (%)	FUEL SUBSTITUTED (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		5		17	11.4	9.5	8.4	7.5	6.8
1 $/MMBTU	30	15	$/kW	19.4	13.7	11.8	10.7	9.8	9.1
		25		21.7	16	14.1	13	12.1	11.4
		5		17	11.4	9.5	8.4	7.5	6.8
1 $/MMBTU	50	15	$/kW	19.4	13.7	11.8	10.7	9.8	9.1
		25		21.7	16	14.1	13	12.1	11.4
		5		17	11.4	9.5	8.4	7.5	6.8
1 $/MMBTU	70	15	$/kW	19.4	13.7	11.8	10.7	9.8	9.1
		25		21.7	16	14.1	13	12.1	11.4
		5		18.3	12.6	10.7	9.7	8.7	8.1
2 $/MMBTU	30	15	$/kW	23.1	17.4	15.5	14.5	13.5	12.9
		25		27.9	22.2	20.3	19.3	18.3	17.7
		5		18.3	12.6	10.7	9.7	8.7	8.1
2 $/MMBTU	50	15	$/kW	23.1	17.4	15.5	14.5	13.5	12.9
		25		27.9	22.2	20.3	19.3	18.3	17.7
		5		18.3	12.6	10.7	9.7	8.7	8.1
2 $/MMBTU	70	15	$/kW	23.1	17.4	15.5	14.5	13.5	12.9
		25		27.9	22.2	20.3	19.3	18.3	17.7
		5		19.6	13.9	12	10.9	10	9.3
3 $/MMBTU	30	15	$/kW	26.9	21.2	19.3	18.2	17.3	16.6
		25		34.2	28.5	26.6	25.5	24.6	23.9
		5		19.5	13.9	12	10.9	10	9.3
3 $/MMBTU	50	15	$/kW	26.9	21.2	19.3	18.2	17.3	16.6
		25		34.2	28.5	26.6	25.5	24.6	23.9
		5		19.5	13.9	12	10.9	10	9.3
3 $/MMBTU	70	15	$/kW	26.9	21.2	19.3	18.2	17.3	16.6
		25		34.2	28.5	26.6	25.5	24.6	23.9

NATURAL GAS REBURN
ANNUAL COST (mills/kWh)

FUEL PRICE DIFFERENTIAL	CAPACITY FACTOR (%)	FUEL SUBSTITUTED (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		5		1.9	1.5	1.3	1.2	1.2	1.1
1 $/MMBTU	30	15	mills/kWh	3.9	3.5	3.3	3.2	3.2	3.1
		25		5.9	5.5	5.3	5.2	5.1	5.1
		5		1.5	1.3	1.2	1.1	1.1	1
1 $/MMBTU	50	15	mills/kWh	3.5	3.2	3.1	3	3	3
		25		5.4	5.1	5	5	4.9	4.9
		5		1.4	1.2	1.1	1.1	1	1
1 $/MMBTU	70	15	mills/kWh	3.3	3.1	3	3	2.9	2.9
		25		5.2	5	4.9	4.9	4.8	4.8
		5		3	2.6	2.4	2.3	2.2	2.2
2 $/MMBTU	30	15	mills/kWh	7.2	6.7	6.6	6.5	6.4	6.3
		25		11.3	10.9	10.7	10.6	10.5	10.5
		5		2.6	2.3	2.2	2.2	2.1	2.1
2 $/MMBTU	50	15	mills/kWh	6.6	6.3	6.2	6.2	6.1	6.1
		25		10.6	10.4	10.3	10.2	10.2	10.1
		5		2.4	2.2	2.1	2.1	2.1	2
2 $/MMBTU	70	15	mills/kWh	6.4	6.2	6.1	6.1	6	6
		25		10.3	10.1	10.1	10	10	10
		5		4.1	3.6	3.5	3.4	3.3	3.3
3 $/MMBTU	30	15	mills/kWh	10.4	10	9.8	9.7	9.6	9.6
		25		16.7	16.3	16.1	16	15.9	15.9
		5		3.6	3.4	3.3	3.2	3.2	3.1
3 $/MMBTU	50	15	mills/kWh	9.7	9.5	9.4	9.3	9.3	9.2
		25		15.9	15.6	15.5	15.4	15.4	15.4
		5		3.4	3.2	3.2	3.1	3.1	3.1
3 $/MMBTU	70	15	mills/kWh	9.5	9.3	9.2	9.2	9.1	9.1
		25		15.5	15.3	15.2	15.2	15.2	15.1

NATURAL GAS REBURN
NOx COST EFFECTIVENESS ($/TON)

FUEL PRICE DIFFERENTIAL	CAPACITY FACTOR (%)	FUEL SUBSTITUTED (%)		BOILER GENERATING CAPACITY (MW) 100	300	500	700	1000	1300
		5		747.6	570	510.7	477.8	447.4	427.6
1 $/MMBTU	30	15	$/TON	1510.9	1333.8	1274.5	1241.6	1211.2	1191.5
		25		2274.9	2097.6	2038.4	2005.4	1975.1	1955.3
		5		589.5	483	447.4	427.6	409.6	397.5
1 $/MMBTU	50	15	$/TON	1329.6	1223.3	1187.7	1167.9	1149.6	1137.8
		25		2069.9	1963.5	1927.9	1908.1	1890	1878.1
		5		521.8	445.7	420.2	406.2	393.1	384.7
1 $/MMBTU	70	15	$/TON	1251.8	1175.9	1150.4	1136.3	1123.3	1114.8
		25		1981.9	1906	1880.6	1866.5	1853.5	1845
		5		1161.8	984.6	925	892.1	861.7	842
2 $/MMBTU	30	15	$/TON	2754.2	2577.1	2517.7	2484.8	2454.4	2434.6
		25		4346.8	4169.5	4110.3	4077.4	4047	4027.3
		5		991.2	884.6	849	829.2	811	799.1
2 $/MMBTU	50	15	$/TON	2534.3	2428	2392.4	2372.6	2354.4	2342.5
		25		4077.8	3971.4	3935.9	3916.1	3897.9	3886
		5		917.9	841.8	816.3	802.2	789.2	780.7
2 $/MMBTU	70	15	$/TON	2440.1	2364.1	2338.7	2324.6	2311.6	2303.1
		25		3962.4	3886.5	3861	3846.9	3833.9	3825.5
		5		1576.4	1398.8	1339.4	1306.5	1276.1	1256.4
3 $/MMBTU	30	15	$/TON	3997.2	3820.2	3760.8	3727.9	3697.5	3677.8
		25		6418.8	6241.5	6182.3	6149.3	6119	6099.2
		5		1392.7	1286.2	1250.6	1230.8	1212.5	1200.7
3 $/MMBTU	50	15	$/TON	3739.1	3632.7	3597.1	3577.4	3559.2	3547.3
		25		6085.8	5979.3	5943.7	5924	5905.8	5893.9
		5		1314	1237.9	1212.4	1198.3	1185.3	1176.8
3 $/MMBTU	70	15	$/TON	3628.4	3552.4	3527	3512.9	3499.8	3491.4
		25		5942.7	5866.9	5841.5	5827.4	5814.4	5805.9