Investigation of High SO_2 Removal Design and Economics
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ABSTRACT

The Clean Air Act Amendments of 1977 require the Environmental Protection Agency to set new standards for control of sulfur dioxide emissions. SRI International estimated capital costs and incremental revenue requirements for flue gas desulfurization (FGD) to meet increasingly stringent SO_2 removal requirements. These costs were estimated by SRI for 14 conceptual designs based on process conditions and material balances determined by Radian Corporation. Analyses of these designs show the effects of various levels of SO_2 removal on the costs of burning eastern high-sulfur, high chlorine coal and western low-sulfur, low-chlorine ("compliance") coal, using limestone, lime, and regenerative magnesia scrubbing. Total capital requirement for FGD ranges from \$123 to \$213/kW (January 1979 dollars), if 25% spare capacity is provided for reliability. The sulfur "compliance" coal chosen in this study does not require FGD under the old standard (520 ng/J or 1.2 lb SO_2 emission/million Btu), but the incremental 30-year levelized busbar cost is 8 to 9 mill/kWh if a high percentage of SO_2 removal is required. For eastern coal, revenue requirements increase less than 3 mill/kWh as removal requirements are tightened.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

This publication is composed of two separate technical planning study reports that were undertaken to predict the effect of potential increasingly strict SO_2 emission limits on the economics of wet scrubbing. Both reports are part of the same EPRI effort and are published as separate volumes. In the first part of the effort, Radian Corporation performed process designs and material balances as input to the second half of the study, an economic evaluation performed by SRI International. The factors that significantly affect the process designs are summarized and the data for selected cases are presented.

PROJECT OBJECTIVE

The objective of Radian's work was to define representative cases and develop process designs and material balances that could be used to determine costs for each case. The process designs were developed using a process simulation computer program developed by the contractor. Cases were selected to span:

- Coal--eastern and western
- SO_2 removal--84%, 93% and 99%
- Alkali--magnesia, limestone and lime

The objective of SRI's work was to use the results of the Radian work to develop a cost estimate for each case and then analyze the results. In the cost estimates the latest vendor information was used to prepare the estimates.

PROJECT RESULTS

Process Designs. The major variables that were investigated in these designs were the liquid-to-gas ratio (L/G) in the scrubber and the volume of the process slurry holding tank. The former affects the SO_2 removal efficiency and the latter affects the scaling potential in the scrubber.

Under the study assumptions, higher SO_2 removals required moderate increases in L/G and were found to be dependent on the magnesium and chloride levels in the slurry liquors. This information is useful in gaining an understanding of the magnitude of the process changes required for high SO_2 removals.

Cost estimates. For low-sulfur coal systems, the design coal chosen meets the 1971 New Source Performance Standard (NSPS) for SO_2 without any further SO_2 removal. Increasing the design SO_2 removals to 93% and 99% results in a levelized cost increase of 8.5 and 8.9 mills/kWh, respectively. Magnesia scrubbing was about 7-8% more expensive than limestone scrubbing on a levelized basis for the low-sulfur western coal cases. For eastern higher-sulfur coal, increasing the removal requirements to 93% and 99% removal increases the levelized revenue requirement by 8% and 18%, respectively. Costs are significantly affected by chloride and magnesium levels in the coal. For high-sulfur coal, magnesia scrubbing is about 15% cheaper than limestone scrubbing on a levelized revenue basis.

The significance of the results of this study lies in the comparative numbers and not in their absolute magnitude. The increased costs are significant for higher SO_2 removals but they do not change by an order-of-magnitude as originally anticipated.

Probably the most significant unanticipated result of the study was the large effect that the Mg and Cl content of the scrubbing liquor has on system design and costs for lime and limestone systems. It is clear that this area should receive more attention in system design.

Finally, although the magnesia system appears less expensive than conventional lime/limestone systems for high-sulfur coals, it is still not well developed and its reliability remains uncertain.

Generalized cost estimates such as these are only an aid in planning either a research program or the selection of a flue gas desulfurization (FGD) process. It is not appropriate to generalize these comparisons or assume they represent manufacturers' current selling prices.

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SUMMARY

The Electric Power Research Institute (EPRI) commissioned Radian Corporation and SRI International (SRI) to study the technology and costs of meeting increasingly stringent emission standards. SRI's work is reported herein. The results of 14 flue gas desulfurization (FGD) case studies, given in the tabulation below, show that costs would increase substantially as SO_2 removal requirements are increased.

Power plants fired with the low-sulfur western coal of this study meet the old SO_2 standard of 1.2 lb/million Btu (520 ng/J) 2-hour average without FGD. The addition of FGD increases the plant levelized revenue requirement by about 8 to 9 mills/kWh, much more than the increase for eastern coal. The total cost of FGD is lower for western than for eastern coal, but the cost to remove a given amount of SO_2 is about four times higher for western coal.

For eastern coal, increasing the removal requirement to 93% removal and 99% removal increases the revenue requirement by about 8% and 18%, respectively. In this study designs capable of meeting the most stringent regulations use two scrubber loops in a single vessel. If separate vessels were used, the increase in revenue requirement would be greater. Lime scrubbing has about the same levelized busbar cost as limestone scrubbing, even though the cost of absorbent is much higher.

High chlorine content in coal results in increased costs, as the figures for the "High Cl" system show. In this case, the coal contains three times the chlorine content of base case coal. On the other hand, the presence of active magnesium in the limestone reduces costs, if the magnesium content is much more than chemically equivalent to the chlorine content in the coal. The "High Mg" case shows that magnesium reduced the revenue requirement by about 9% of the eastern coal base case cost, which assumed no active magnesium.

Magnesia scrubbing revenue requirements are about 6 to 8% more for 99% removal than for 93% removal. (These figures are consistent with limestone results.)

System	Percent SO_2 % Removal[2]	Type of Coal[1]	Levelized Revenue Requirement of FGD Mills/kWh[2]	Total Capital Requirement, $/kW
Limestone	84% removal	Eastern	13.0	165
Limestone[4]	93% removal	Eastern	14.1	194
Limestone	99% removal	Eastern	15.4	213
Limestone (High Cl)	93% removal	Eastern	14.6	204
Limestone (Low Mg)	93% removal	Eastern	13.8	189
Limestone (High Mg)	93% removal	Eastern	12.9	178
Lime	93% removal	Eastern	14.1	178
Limestone	93% removal	Western	8.5	123
Limestone	99% removal	Western	8.9	128
Magnesia[3]	93% removal	Eastern	12.1	193
Magnesia	99% removal	Eastern	13.1	207
Magnesia	93% removal	Western	9.1	155
Magnesia	99% removal	Western	9.6	163
Limestone	93% removal	Eastern	14.4	181

1. Eastern coal, 4.0% sulfur; western coal, 0.48% sulfur; uncontrolled emissions would be 7.5 and 1.1 lb/million Btu, respectively. Eastern coal 0.1% Cl in base case, 0.3% in High Cl case.

2. Assuming an inflation rate of 6.0% per year and a fuel cost increase of 6.2% per year; 30-year levelized revenue requirement at levelized capacity factor of 0.7. Methodology standardized by EPRI.

3. Base cases.

4. Variation of base case design.

By these estimates, magnesia scrubbing is cheaper than limestone scrubbing for eastern coal but more expensive for western coal. This difference arises because low-cost, single-loop scrubbers are used in western coal limestone scrubbing designs, whereas in all other cases (except the 84% SO_2 removal limestone case) a prescrubber or two loops are used. The competitive position of magnesia scrubbing for eastern coal is uncertain, however, because the process is not well developed, and its reliability has not been proven.

The U.S. Environmental Protection Agency is considering waste disposal regulations that could significantly increase the cost of disposal of some scrubber sludges. Such regulations could improve the competitive position of regenerative processes like magnesia scrubbing.

A number of aspects of FGD technology have not been thoroughly studied, and work in these areas would improve the reliability of economic comparisons. These include the most appropriate materials of construction for some services, performance of various towers (particularly for magnesia scrubbing), absorbent dissolution rates, and magnesia regeneration. Process innovations such as use of adipic acid and forced oxidation should be investigated, because they could have significant effects on process economics.

Section 1

INTRODUCTION

The Clean Air Act Amendments of 1977 require the Environmental Protection Agency (EPA) to reconsider the new source performance standards (NSPS) for SO_2 emissions from utility boilers. The new standards must establish percentage removal requirements for both high- and low-sulfur coals.

The Electric Power Research Institute (EPRI) requested Radian Corporation and SRI International (SRI) to study the technology and costs of meeting higher SO_2 removal requirements. The overall approach was to develop conceptual designs for various case studies. Radian was responsible for determining process conditions and material balances. SRI was responsible for designing equipment and determining costs. SRI's work is summarized in the balance of this report.

For the purposes of this study, three levels of control are considered. The first is the old standard of 1.2 lb SO_2 emission/million Btu of fuel burned. The second is 93% removal of SO_2 and the third level is 99% removal.

Efficiency of removal is affected by many variations in coal composition and flue gas desulfurization (FGD) process. The variables studied included:

- Coal--eastern high-sulfur and western low-sulfur
- Absorption promoters and inhibitors--magnesium and chloride
- Absorbent--limestone and lime (both in nonregenerative processes)
- Process--regenerative (magnesia) and nonregenerative.

The conceptual designs considered here and their designations are listed in the following tabulation.

System	Design Percent SO_2 Removal	Coal Type	Case Designation
Limestone (1.2 lb $SO_2/10^6$ Btu)	84	Eastern	LS84E
Limestone	93	Eastern	LS93E
Limestone	93	Eastern	LS93E SP
Limestone	99	Eastern	LS99E
Limestone (Cl sensitivity)	93	Eastern	LS93ECL
Limestone (low Mg sensitivity)	93	Eastern	LS93E0.5MG
Limestone (high Mg sensitivity)	93	Eastern	LS93E2.0MG
Lime	93	Eastern	LI93E
Limestone	93	Western	LS93W
Limestone	99	Western	LS99W
Magnesium oxide	93	Eastern	MG93E
Magnesium oxide	99	Eastern	MG99E
Magnesium oxide	93	Western	MG93W
Magnesium oxide	99	Western	MG99W

The base cases for nonregenerative and regenerative processes are fully described in Sections 2 and 3. The economic results of these and of derivative cases are given in Section 4. SRI's conclusions and recommendations are given in Section 5. The design and economic bases are given in Appendix A and design principles in Appendix B. The design of Case LS93E SP, which is a variation of the design of Case LS93E, is given in Appendix C.

Section 2

LIMESTONE SCRUBBING

Conceptual designs of nonregenerative limestone slurry processes are described in this section. A variation using a lime slurry is also described. All the processes were designed in detail. The base case, LS93E, is fully described. The other cases resemble the base case, and only distinquishing features are discussed.

BASE CASE LS93E

Flue gas from the electrostatic precipitator enters a plenum, and then goes to several FGD trains. Five trains are provided, any four of which can treat the full capacity gas load. Each train can be isolated from the rest of the system by an upstream and a downstream damper. The processed gas from these trains is collected in another plenum that connects to the stack. A bypass duct can carry the entire gas load between the two plenums in an emergency.

Figure 2-1 and Table 2-1 present a block flow diagram of the process and a material balance. Other streams have been added to the flow diagram and material balance produced by Radian to include sludge fixation, which was not considered by Radian. SRI designs show the use of quicklime for sludge fixation instead of slaked lime, because the use of quicklime is common practice.

A booster fan in each train forces the gas through the train to the stack. The gas at 149^{o}C (300^{o}F) is quenched and adiabatically saturated by a portion of the absorption slurry in the inlet duct to the first loop of the absorber. This first loop is a circular spray tower in which chlorides and a portion of the SO_2 in the flue gas are removed by recirculated slurry pumped from the first-loop hold tank through spray headers.

The partially scrubbed gas rises to the second loop, which is above the first loop in the same vessel. Although the gas goes directly to the second loop, the slurry circuits are kept separate by collecting the second-loop slurry in an inverted cone through which it drains to the second-loop hold tank. Pumps

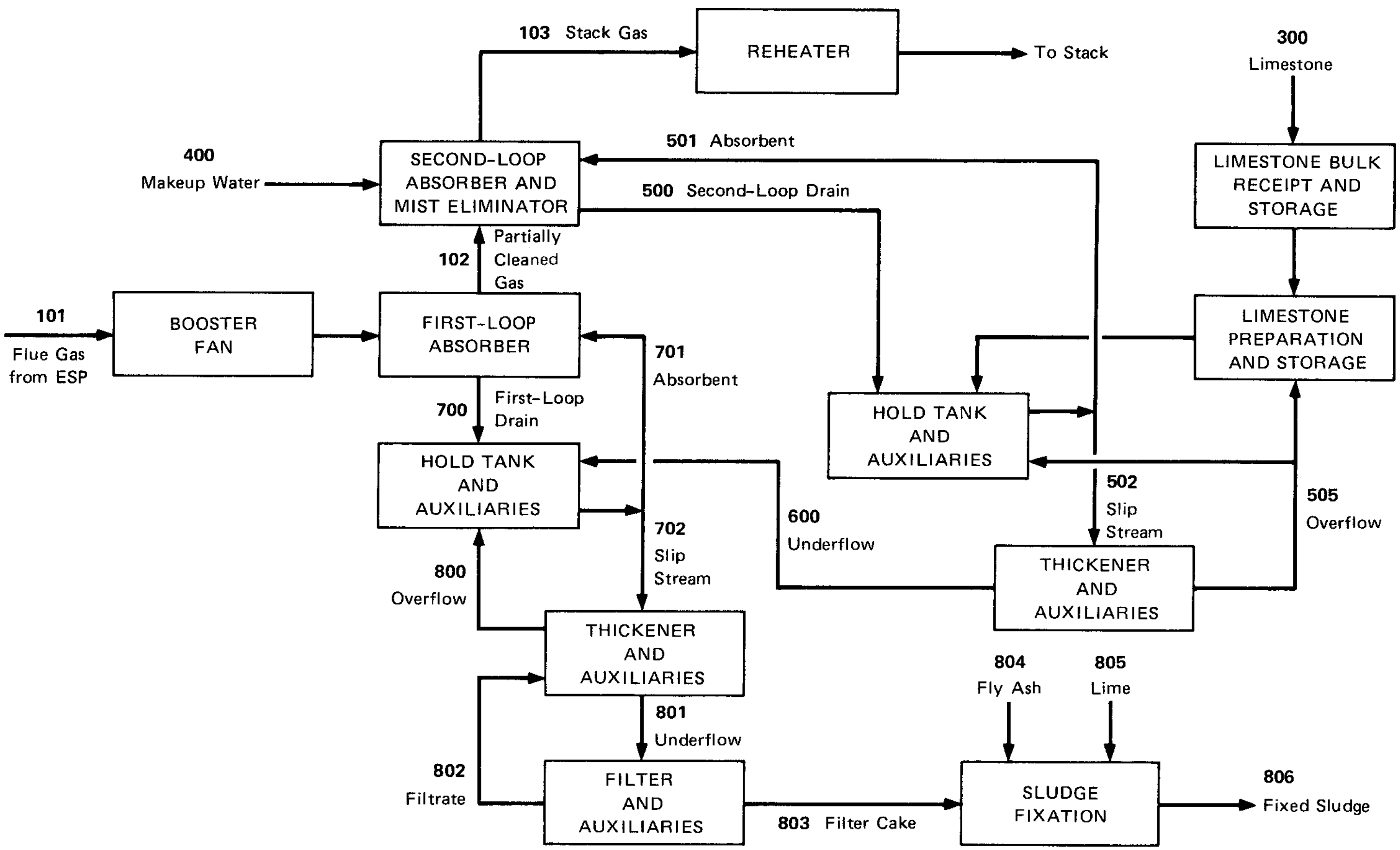


Figure 2-1. Block Flow Diagram Case LS 93 E

recirculate this slurry to spray headers that irrigate the open-grid packing to complete the absorption of SO_2 from the flue gas. Above this packing, a two-stage mist eliminator removes entrained droplets from the scrubbed gas. Makeup water not required for pump seals is used to wash deposits off the mist eliminator.

The cleaned gas leaves the top of the second loop and goes to an inline reheater. Here steam-heated tubes reheat the gas. The reheated gas goes through a flue to the plenum and then to the stack.

Limestone is received by rail, stored, and reclaimed for use as absorbent. It is ground in a wet ball mill, with clarified liquor to form a slurry. This slurry is added to the second-loop hold tank.

A portion of the second-loop slurry from the hold tank is diverted to a thickener, which aids in controlling the double-loop water balance. Overflow returns to the second-loop hold tank, and underflow goes to the first-loop hold tank. This underflow is the makeup absorbent for the first loop.

A portion of the first-loop slurry is dewatered to form the plant solid waste. The slurry is dewatered first in a thickener, from which the overflow returns to the first-loop hold tank. The underflow from the thickner is filtered in rotary drum vacuum filters. The filtrate returns to the first-loop thickener. The filter cake, containing 60% solids, is conveyed to the fixation system.

Besides filter cake, the fixation system receives fly ash and lime. These three components are blended in a pug mill to make the waste material suitable for landfill. This waste is pushed by a bulldozer into a chute for loading into trucks that carry the waste to the disposal area.

The principal operating conditions for this case are summarized in Table 2-2. Table 2-3 lists the major equipment used, and Table 2-4 presents the utility and operating requirements. For this and all other cases, the economic analysis is presented in section 4 of the report.

LS 84E

This case is for a power plant meeting an emission standard of 1.2 lb SO_2/ million Btu (520 ng/J) and burning eastern coal. Because of the less

Table 2-1

MATERIAL BALANCE FOR CASE LS 93E

	Stream Number									
	101	102	103	300	400	500	501	502	503	600
Total flow, lb/h	5,794,700	6,030,500	6,022,200	63,820	306,500	63,160,000	63,150,000	622,500	244,600	377,900
Solid flow, lb/h	-----	-----	-----	63,820	-0-	7,563,000	7,574,000	74,660	-0-	74,670
Liquid flow, lb/h	-----	-----	-----	-----	306,500	55,597,000	55,580,000	547,840	244,600	303,300
Slurry flow, gpm	-----	-----	-----	-----	612.5	118,100	118,100	1,164	493.9	670.4
Gas flow, acfm @ 1 atm	1.811×10^6	1.496×10^6	1.496×10^6	-----	-----	-----	-----	-----	-----	-----
SO_2, mol%	0.2959	0.1939	0.0194	-----	-----	-----	-----	-----	-----	-----
Temperature, °F	300	128	128	60	60	128	128	128	128	128
Liquid pH	-----	-----	-----	-----	7.30	4.94	6.08	6.08	6.21	6.21

	Stream Number									
	700	701	702	800	801	802	803	804	805	806
Total flow, lb/h	52,520,000	52,760,000	691,800	553,500	276,600	138,000	138,300	8,300	3,650	150,250
Solid flow, lb/h	6,315,000	6,320,000	82,870	-0-	82,960	-0-	82,960	8,300	3,650	94,910
Liquid flow, lb/h	46,210,000	46,440,000	608,900	553,500	193,600	138,300	55,290	-----	-----	-----
Slurry flow, gpm	97,090	97,570	1,279	1,103	430.7	275.4	276.8	-----	-----	-----
Gas flow, acfm @ atm	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
SO_2, mol%	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
Temperature, °F	128	128	128	128	128	128	128			
Liquid pH	3.01	5.69	5.69	5.87	5.87	5.87	5.87	-----	-----	-----

Table 2-2

PRINCIPLE DESIGN OPERATING CONDITIONS FOR CASE LS 93E

	Metric Units	English Units
Booster Fans		
Flow @ 149°C (300°F)	850,000 m^3/h	500,000 a ft^3/min
Pressure rise	3,500 N/m^2	14 in. H_2O
Double-loop absorber diameter x height (T/T)	9.5 I.D. x 26.5 m	31 I.D. x 87 ft
First-loop NTU (L/G)	0.46 (8.7 ℓ/m^3)	0.46 (65.2 gal/1000 a ft^3)
Second-loop NTU (L/G)	2.40 (10.6 ℓ/m^3)	2.40 (79.0 gal/1000 a ft^3)
Superficial gas velocity, saturated	2.6 m/s	8.4 ft/s
Packing height	1.2 m	4 ft
Mol alkali/mol SO^2 absorbed, overall	1.10	1.10
Reheater temperature rise	28°C	50°F
Heating steam pressure	3.55 x 10^6 N/m^2	515 psia
Tank residence time, minutes		
First-loop hold tank residence	25.6 min	25.6 min
Second-loop hold tank residence	9.8 min	9.8 min
Slurry solids content, wt %	12.0	12.0
Dewatered sludge solids, wt %	60.0	60.0
Limestone grind	95% minus 44 μm	95% minus 325 mesh
Thickener design rate	1.8 $m^2/(t/d)$	17.6 ft^2/(short ton/d) dry solids
Vacuum filter design rate	342 $kg/h/m^2$	70 $lb/h/ft^2$ dry solids

Table 2-3

MAJOR EQUIPMENT LIST FOR CASE LS 93E
(4 scrubber trains plus 1 spare)

Equipment	Number	Material[1]	Metric Units	English Units	Major Item Equipment Cost, $k
Flue gas treating system					
Double-loop absorbers					9,100
Diameter (I.D.) × height	5	Rubber-lined	9.5 × 26.5 m	31 × 87 ft	
Packing	5	PP	1.2 m	4 ft	
Mist eliminators	5	316 L & PP	2-stage	2-stage	
Bypass	1	CS	100%	100%	
Pumps					1,370
First-loop	15	Rubber-lined	2,700 m^3/h	12,000 gal/min	
Second-loop	15	Rubber-lined	3,400 m^3/h	15,000 gal/min	
Hold tanks					2,706
First-loop	5	Rubber-lined	2,650 m^3	700,000 gal	
Second-loop	5	Flake-lined	1,250 m^3	330,000 gal	
Flue gas reheaters (with soot blowers)	5	316 L	560 m^2	6,000 ft^2	1,520
Raw material preparation and storage					1,400
Ball mills	2		29 t/h	32 short ton/h	
Pumps					
Recycle	2	Rubber-lined	450 m^3/h	2,000 gal/min	
Feed	2	Rubber-lined	150 m^3/h	650 gal/min	
Makeup water treater	1		4 m^3/h	1,000 gal/min	
Feed surge (with agitators)	2	Rubber-lined	5,300 m^3	1,400,000 gal	
Makeup water tanks	1	CS	3,800 m^3	1,000,000 gal	
Raw material bulk receipt and storage					525
Unloading system	2		72 t/h	80 short ton/h	
Reclaim dozer and conveyor	2		29 t/h	32 short ton/h	
Silos	2	CS	1,100 t	1,200 short ton/h	
Waste separation and storage					2,730
First-loop thickener	1	Rubber-lined	56 D × 4.6 m	183 D × 15 ft	
Second-loop thickener	1	Flake-lined	53 D × 4.6 m	173 D × 15 ft	
Filters	4	317 L	37 m^2	400 ft^2	
Fly ash silos	2	CS	450 t	500 short ton	
Lime silos	2	CS	180 t	200 short ton	
Fly ash conveyors	1		9 t/h	10 short ton/h	
Lime conveyors	1		4 t/h	4 short ton/h	
Pug mill	2		90 t/h	100 short ton/h	
Disposal building storage	1		3 days	3 days	
Sludge conveyors	4		32 t/h	35 short ton/h	
Flue gas supply and discharge (fans)	5	CS	1,200 kW	1,400 bhp	1,000

[1]CS - Carbon steel
PP - Polypropylene

Table 2-4

UTILITY AND OPERATING REQUIREMENTS FOR CASE LS 93E

Requirement	Amount	
Operating labor (men/shift)	3	
Electric power (MW), average	18.0	
	Metric Units	English Units
Gas reheat thermal energy, maximum	79.4×10^9 J/h	75.3×10^6 Btu/h
Raw water makeup, average	139 m^3/h	613 gal/min
Limestone	26.8 t/h	29.5 short ton/h
Fly ash for fixation	3.8 t/h	4.2 short ton/h
Lime	1.66 t/h	1.83 short ton/h

Note: 2% handling loss is assumed for limestone, lime, and raw water.

stringent requirements, a single-loop absorption system is used, eliminating the need for a spray tower, a hold tank, and a thickener. The flue gas is quenched in the inlet duct to an open-grid packed tower, where both SO_2 and HCl are removed simultaneously. Otherwise, the process is similar to the base case process. Because there is less absorption, the alkali and sludge rates are lower than those in the base case.

LS 99E

This case meets a more stringent SO_2 removal requirement (99% removal) than the base case. The flow scheme is identical, but limestone and sludge rates are greater.

LS 93E CL, LS 93 E 0.4 MG, AND LS 93E 2.0 MG

These cases meet the same requirements as the base case but explore the deleterious effects of chlorides and the beneficial effects of magnesia. The flow schemes for all these are identical.

The coal burned in Case LS 93E CL has three times the chlorine content of the coal in the base case (0.3 vs 0.1%). Cases LS 93E 0.5 MG and LS 93E 2.0 MG use the same coal as the base case, but the limestones have 0.5 and 2.0% active $MgCO_3$, respectively, compared with 0.0% in the base case.

LI 93E

This case uses lime in place of limestone as an absorbent. Lime is received by rail and conveyed to silos. Lime from the silos is fed by weigh feeders to slakers. Slaked lime is stored in agitated surge tanks and pumped as needed to the second-loop hold tank. Otherwise the flow scheme is identical to that of LS 93E.

LS 93W AND LS 99W

Because western coal is low in sulfur, single-loop spray towers are used. The flow schemes are identical to that of LS 84E, except that spray towers are used.

Section 3

MAGNESIA SCRUBBING

Conceptual designs of regenerative magnesia scrubbing processes are described in this section. All the processes were designed in detail. The base case, MG 93E, is fully described. The other cases resemble the base case, and only distinquishing features are discussed.

BASE CASE MG 93E

This base case uses the same general approach as that used for the throwaway base case LS 93E; i.e., the same gas rate, SO_2 removal, spare equipment policy, and so on, are used. Figure 3-1 and Table 3-1 present a block flow diagram and a material balance, both expanded from the absorption work of Radian to include regeneration and acid production.

A booster fan forces flue gas through each train. This gas is quenched in a prescrubber from 149^{o}C (300^{o}F) to adiabatic saturation temperature by recirculated makeup water. Water drains from the prescrubber to a hold tank, where a small amount of lime is added to prevent excessively low pH. The prescrubber removes HCl, but little SO_2, from the gas. Water is sprayed into the gas through spray headers in both the inlet duct and the spray tower prescrubber. In contrast with the throwaway designs, rectangular towers are used, as has been common practice for mobile bed absorbers.

The prescrubbed gas rises to the absorber, which is in the same vessel above a trapout device that collects the absorption slurry for diversion to its hold tanks. The absorber (which removes SO_2 from the gas) has three beds of mobile nitrile foam balls that improve gas-liquid contacting and thus help remove SO_2. Pumps recirculate slurry from the hold tank to spray headers above these beds. The gas rises through a two-stage mist eliminator that removes entrained droplets. Part of the makeup water can be used to wash the mist eliminator if necessary.

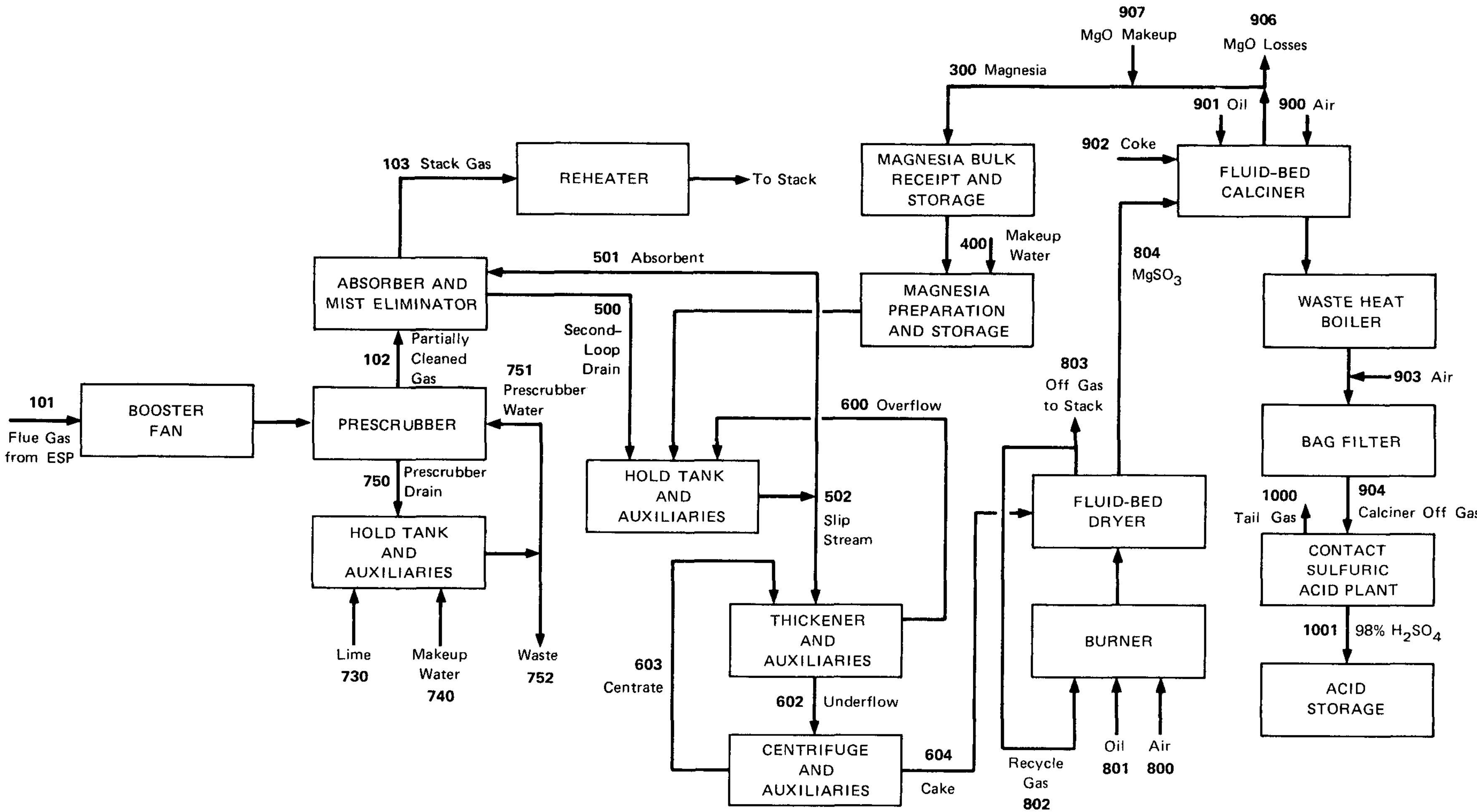


Figure 3-1. Block Flow Diagram for Case MG 93 E

The cleaned gas leaves the top of the second loop and goes to an inline reheater. Here steam-heated tubes reheat the gas. The reheated gas goes through a flue to the plenum and then to the stack.

Makeup magnesia is received by rail, stored, and reclaimed for use as absorbent. Recycled and makeup magnesia are slaked with makeup water and added to the absorber hold tank as a slurry.

The prescrubber waste is a small amount of acidic slurry. Because fly ash is collected dry in this plant, the prescrubber waste can be used for dust suppression. Lime is used, if necessary, to control pH.

A slip stream of absorbent is dewatered in a thickener. The overflow is returned to the absorber hold tank. The underflow is further dewatered in a centrifuge. The centrate is returned to the thickener. The centrifuge cake is composed of hydrated magnesium sulfite and sulfate ready for regeneration.

The centrifuge cake is dried in a fluid-bed dryer by hot gas from an oil-fired burner. This gas is tempered (cooled to a controlled temperature) by recycled off-gas. The net off-gas goes to the stack, providing about $3^{o}C$ ($5^{o}F$) reheat to the flue gas. The dried solids go to a fluid-bed calciner. Rotary kiln dryers and calciners could have been used instead; factors affecting the decision are discussed in Appendix B.

The fluid-bed calciner, also heated by burning oil, liberates SO_2 from $MgSO_3$ both directly and from $MgSO_4$ by reduction with added coke. The calciner off-gas containing the SO_2 generates steam in a waste heat boiler for use in reheating flue gas. Air is then added to this calciner off-gas to provide the oxygen needed for conversion to surfuric acid and to cool the gas before it enters a bag filter. The filtered gas goes to a standard, single-contact sulfuric acid plant, which makes 98% acid (97% yield). This acid goes to storage.

The acid plant tail-gas is returned to the absorber to recover the SO_2 content. The absorbers were sized for 4% additional gas from this source (the calculated actual amount is 2.8%). However, the material balance was not iterated to account for this small incremental flow.

Table 3-1

MATERIAL BALANCE FOR CASE MG 93E

	Stream Number									
	101	102	103	300*	400*	500	501	502	600	602
Total flow, lb/h	5,794,700	6,030,500	6,022,200	22,160	59,280	15,490,000	15,540,000	581,600	466,100	250,500
Solid flow, lb/h	-----	-----	-----	22,160	-----	1,856,000	1,857,000	69,900	-----	75,150
Liquid flow, lb/h	-----	-----	-----	-----	59,280	13,634,000	13,600,000	511,700	466,100	175,400
Slurry flow, gpm	-----	-----	-----	-----	119	23,900	23,760	894.3	759.5	355.6
Gas flow, acfm @ 1 atm	1.811×10^6	1.497×10^6	1.495×10^6	-----	-----	-----	-----	-----	-----	-----
SO_2, mol%	0.2959	0.2765	0.0194	-----	-----	-----	-----	-----	-----	-----
Temperature, °F	300	128	128	60	60	128	128	128	128	128
Liquid pH	-----	-----	-----	-----	7.30	5.45	7.94	7.94	7.94	7.94

	603	604	730†	740†	750	751	752	800	801	803
Total flow, lb/h	135,000	115,500	403	261,022	7,517,000	7,761,000	17,730	88,700	4,314	132,100
Solid flow, lb/h	-----	75,150	-----	-----	48,320	48,320	110.4	-----	-----	-----
Liquid flow, lb/h	135,000	40,320	-----	261,022	7,469,000	7,713,000	17,620	-----	4,314	-----
Slurry flow, gpm	220.0	135.6	-----	522	14,470	14,960	34.19	-----	-----	-----
Gas flow, acfm @ 1 atm	-----	-----	-----	-----	-----	-----	-----	19,350	-----	56,260
SO_2, mol%	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
Temperature °F	128	128	60	60	128	128	128	60	60	400
Liquid pH	7.94	7.94	-----	7.30	1.23	1.27	1.27	-----	-----	-----

	804	900	901	902	903	904	906	907	1000	1001
Total flow, lb/h	57,750	60,710	4,222	400	82,070	182,990	900	900	135,060	52,230
Solid flow, lb/h	57,750	-----	-----	-----	-----	-----	900	900	-----	-----
Liquid flow, lb/h	-----	-----	4,222	400	-----	-----	-----	-----	-----	52,230
Slurry flow, gpm	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
Gas flow, acfm @ 1 atm	-----	13,240	-----	-----	18,090	60,800	-----	-----	34,390	-----
SO_2, mol%	-----	-----	-----	-----	-----	9.1	-----	-----	0.3535	-----
Temperature, °F	400	60	60	60	60	400	200	60	160	100
Liquid pH	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----

*Radian figures converted to use of MgO instead of $Mg(OH)_2$.

†Radian figures converted to use of CaO instead of $Ca(OH)_2$.

Regenerated magnesia from the fluid-bed calciner is recycled. The material balance assumes that 4% makeup takes care of losses, maintenance of activity, and purge of impurities.

The principal operating conditions for this case are summarized in Table 3-2. Table 3-3 lists the major equipment and Table 3-4 the utility and operating requirements.

For this and all other cases, the economic analysis is given in a later section of the report.

MG 99E

This case meets a more stringent SO_2 removal requirement, 99% removal, than the base case. The flow scheme is identical to that for the base case, MG 93E, but five mobile beds are used instead of three to provide the greater number of transfer units needed for the higher removal.

MG 93W

There are no variations in the flow scheme from that of base case MG 93E.

MG 99W

As in case MG 99E, five mobile beds are used for the high removal requirement.

Table 3-2

PRINCIPAL DESIGN OPERATING CONDITIONS FOR CASE MG 93E

	Metric Units	English Units
Booster fans		
Flow @ 149° (300°F)	850,000 m^3/h	500,000 a ft^3/min
Pressure rise	5,200 N/m^2	21 in. H_2O
Prescrubber-absorber		
Prescrubber NTU (L/G)	0.0 (1.34 ℓ/m^3)	0.0 (10 gal/1,000 a ft^3)
Absorber NTU (L/G)	2.76 (2.13 ℓ/m^3)	2.76 (15.9 gas/1,000 a ft^3)
Superficial gas velocity, saturated	3.8 m/s	12.5 ft/s
Reheater temperature rise	28°C	50°F
Heating steam pressure	3.55×10^6 N/m^2	515 psia
Tank residence time, minutes		
Prescrubber hold tank residence	5.0 min	5.0 min
Absorber hold tank residence	72.6 min	72.6 min
Slurry solids content, wt %	12.0	12.0
Dewatered sludge solids, wt %	60.0	60.0
Thickener design rate	1.8 $m^2/(t/d)$	17.6 $ft^2/(short\ ton/d)$ dry solids

Table 3-3

MAJOR EQUIPMENT LIST FOR CASE MG 93E
(4 scrubber trains plus 1 spare)

Equipment	Number	Material[1]	Metric Units	English Units	Major Item Equipment Cost, $k
Flue gas treating system					
Prescrubber-absorber					6,150
Cross-section × height	5	Rubber-lined	49 m^2 × 26 m	530 ft^2 × 86 ft	
Mobile beds (3 level)	5	Nitrile foam	0.43 m	1.4 ft	
Mist eliminators	5		2-stage	2-stage	
Bypass	1	CS	100%	100%	
Pumps					460
Prescrubber	15	Rubber-lined	450 m^3/h	2,000 gal/min	
Absorber	15	Rubber-lined	680 m^3/h	3,000 gal/min	
Hold tanks					1,360
Prescrubber	5		8 m^3	2,000 gal	
Absorber	5		1,900 m^3	500,000 gal	
Flue gas reheater (with soot blowers)	5	316 L	560 m^2	6,000 ft^2	1,520
Raw material preparation and storage					600
Slakers	2		11 t/h	12 short ton/h	
Feed surge (with agitators)	2		2,050 m^3	540,000 gal	
Makeup water treater	1		35 m^3/h	150 gal/min	
Makeup water tank	1		570 m^3	150,000 gal	
Raw material bulk receipt and storage					190
Conveyors	2		27 t/h	30 short ton/h	
Silos	2		360 t	400 short ton	
Product separation and storage					7,200
Thickener	1	Flake-lined	53 D × 4.6	174 D × 15 ft	
Centrifuge	3		40 m^3/h	180 gal/min	
Sludge conveyor	3		55 t/h	60 short ton/h	
Fluid-bed dryer	1		84 × 10^9 J/h	80 × 10^6 Btu/h	
Fluid-bed calcining system	1		2.0 t/h	2.2 short ton/h	
Waste heat boiler	1		9.5 t/h	21,000 lb/h	
Bag filter	1		60,000 m^3/h	35,000 a ft^3/min	
Contact sulfuric acid plant	1		640 t/d	700 short ton/d	
Acid tank	1		170 m^3	45,000 gal	
Fuel oil tank	1		26 m^3	7,000 gal	
Flue gas supply and discharge (fans)	5	CS	1,550 kW	1,800 bhp	1,190

[1] CS - Carbon steel

Table 3-4

UTILITY AND OPERATING REQUIREMENTS FOR CASE MG 93E

Requirement	Amount	
Operating labor (men/shift)	3	
Electric power (MW), average	12.4	
	Metric Units	English Units
Fuel oil	163.9×10^9 J/h	155.4×10^6 Btu/h
Gas reheat thermal energy, maximum	79.4×10^9 J/h	75.3×10^6 Btu/h
Raw water makeup, average	37 m^3/h	163 gal/min
Coke	0.2 t/h	0.2 short ton/h
Lime	0.2 t/h	0.2 short ton/h

Note: A 2% handling loss is assumed for lime and raw water.

Section 4

ECONOMICS

In this section the results of the process economics are presented and interpreted and sensitivities to some important variables are defined. The evaluations of all 14 cases are summarized in Table 4-1, which gives the direct capital cost break- down by subsystem; Table 4-2, which gives the capital investment summary; and Table 4-3, which gives the levelized revenue requirements. The economic basis for the assessment is given in Appendix A.

The FGD levelized revenue requirement for eastern coal is increased 9% by the change from the 1.2 lb/million Btu standard (520 ng/J) to a requirement for 93% removal (LS 93 E versus LS 84E). The total capital requirement increases 18%. A 99% removal requirement for eastern coal increases the levelized revenue requirement by 18% and the total capital requirement by 30% over that for the 1.2 lb/million Btu standard (LS 99E versus LS 84E).

These cost increases depend on the design approach, and this approach could be constrained by site-specific factors. If the double-loop system had to be installed in two vessels, the increased foundation and support, duct, and mist eliminator costs might increase the total capital requirement by 30 to 40%, judging from the individual unit costs of single-loop systems.

The design western coal in this report meets the old federal standard of 1.2 lb/million Btu without FGD. Designing for 93 or 99% SO_2 removal increases levelized revenue requirements about 9 mill/kWh (LS 93W and LS 99W).

Although the costs are less for western coal than they are for eastern coal at the same level of removal, the amount of SO_2 removed is much smaller because of the low concentration of SO_2 in the flue gas from western coal. The levelized revenue requirement for removing a given amount of SO_2 is about four times as much for western coal as it is for eastern coal.

Table 4-1

PROCESS CAPITAL COST BREAKDOWN BY SUBSYSTEM
(Millions January 1979 Dollars*)

Case	LS 84E	LS 93E	LS 99E	LS 93E Cl	LS 93E 0.5 MG	LS 93E 2.0 MG	LI 93E	LS 93W	LS 99W	MG 93E	MG 99E	MG 93W	MG 99W	LS 93E SP
Net MW	501	499	497	498	500	502	501	507	506	505	502	507	506	494
Flue gas treating system (absorbers, ducts, dampers, slurry pumps, hold tanks, mist eliminators)	38.24	43.77	48.87	46.51	42.16	39.17	40.33	30.61	32.06	31.62	34.44	33.06	34.90	41.93
Flue gas reheat (heat exchangers, separate fans and ducts, soot blowers)	3.32	3.74	3.74	3.74	3.74	3.74	3.74	3.50	3.50	3.74	3.74	4.01	4.01	3.32
Raw material preparation and storage	2.99	3.08	3.17	3.08	3.08	3.04	2.17	0.97	1.00	1.32	1.36	0.43	0.44	3.15
Raw material bulk receipt and storage	0.87	0.90	0.93	0.90	0.90	0.89	0.51	0.63	0.65	0.41	0.42	0.26	0.27	0.92
Product or waste separation and storage (thickening, filtering, fixation, loading, regeneration, acid plant)	3.78	6.01	6.20	6.01	6.01	6.01	5.80	1.20	1.24	15.84	16.58	4.26	4.51	4.06
Flue gas supply and discharge (booster fans, chimney revisions)	1.41	2.20	2.29	2.20	2.20	2.20	2.11	1.52	1.52	2.62	2.97	2.82	3.15	1.41
Total process capital (excluding engineering and fees)	50.61	59.70	65.20	62.44	58.09	55.05	54.66	38.43	39.97	55.55	59.51	49.89	47.28	54.79
Total, $/kW	101.0	119.6	131.2	125.4	116.2	109.7	109.1	75.8	79.0	110.0	118.1	88.8	93.4	110.9

*550 MW gross plants; figures include sales tax.

Table 4-2

CAPITAL INVESTMENT SUMMARY*

	Case													
	LS 84E	LS 93E	LS 99E	LS 93E C1	LS 93E 0.5 MG	LS 93E 2.0 MG	LI 93E	LS 93W	LS 99W	MG 93E	MG 99E	MG 93W	MG 99W	LS 93E SP
Net MW	501	499	497	498	500	502	501	507	506	505	502	507	506	494
Capital Investment, January 1979 $/kW														
Process capital	101.0	119.6	131.2	125.4	116.2	109.7	109.1	75.8	79.0	110.0	118.1	88.8	93.4	110.9
11% engineering and home office fee	11.1	13.2	14.4	13.8	12.8	12.1	12.0	8.3	8.7	12.1	13.0	9.8	10.3	12.2
Process capital including engineering and fee	112.1	132.8	145.6	139.2	129.0	121.8	121.1	84.1	87.7	122.1	131.1	98.6	103.7	123.1
General facilities	5.6	6.6	7.3	7.0	6.4	6.1	6.1	4.2	4.4	6.1	6.6	4.9	5.2	6.2
Project contingency	17.7	20.9	22.9	21.9	20.3	19.2	19.1	13.3	13.8	19.2	20.6	15.5	16.3	19.4
Process contingency	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	12.2	13.1	9.9	10.4	0.0
Total plant investment	135.4	160.3	175.8	168.1	155.7	147.0	146.3	101.6	105.9	159.6	171.4	128.9	135.6	148.7
Royalty allowance	0.6	0.7	0.7	0.7	0.6	0.6	0.6	0.4	0.4	0.6	0.7	0.5	0.5	0.6
Reproduction costs	5.6	6.2	6.7	6.4	6.0	5.7	6.1	3.7	3.9	5.0	5.4	4.0	4.2	6.2
Inventory capital	0.5	0.5	0.5	0.5	0.5	0.5	0.9	0.1	0.1	0.6	0.7	0.1	0.1	0.5
Initial catalyst and chemicals (in process capital)	--	--	--	--	--	--	--	--	--	--	--	--	--	--
AFDC†	22.5	26.7	29.3	28.0	25.9	24.5	24.3	16.9	17.6	26.6	28.5	21.4	22.6	24.7
Land	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	
Total capital requirement	164.6	194.4	213.2	203.7	188.9	178.3	178.2	122.8	128.0	192.6	206.7	155.0	163.1	180.8

*550-MW gross coal-fired power plants; figures for FGD system excluding fly ash costs.

†Allowance for funds used during construction.

Table 4-3

REVENUE REQUIREMENTS FOR FGD

	Case													
	LS 84E	LS 93E	LS 99E	LS 93E C1	LS 93E 0.5 MG	LS 93E 2.0 MG	LI 93E	LS 93W	LS 99W	MG 93E	MG 99E	MG 93W	MG 99W	LS 93E SP
Fixed operating costs, first year ($/kW-yr)														
Operating labor	0.68	0.68	0.68	0.68	0.68	0.68	0.68	0.67	0.67	1.34	1.35	1.34	1.34	0.69
Maintenance labor	2.17	2.56	2.81	2.69	2.49	2.35	2.34	1.63	1.69	2.55	2.74	2.06	2.17	2.38
Maintenance materials	3.25	3.85	4.22	4.03	3.74	3.53	3.51	2.44	2.54	3.83	4.11	3.09	3.25	3.57
Administrative and support labor	0.85	0.97	1.05	1.01	0.95	0.91	0.90	0.69	0.71	1.17	1.23	1.02	1.05	0.92
Total fixed O&M first year	6.94	8.06	8.76	8.41	7.86	7.46	7.43	5.42	5.61	8.90	9.43	7.51	7.81	7.55
Variable operating cost excluding fuel, first year (mills/kWh)														
Water	.03	.03	.03	.03	.03	.03	.03	.03	.03	.01	.01	.00	.00	.03
Chemicals and catalysts	.63	.67	.71	.67	.66	.64	1.21	.10	.11	.22	.22	.04	.04	.70
Other consumables	1.72	1.59	1.74	1.68	1.55	1.38	1.46	1.41	1.47	1.09	1.24	1.03	1.13	1.94
Waste disposal	.80	.85	.91	.84	.85	.82	.79	.17	.18	.00	00	.00	.00	.90
Total variable (excluding fuel, first year)	3.18	3.14	3.39	3.22	3.09	2.87	3.49	1.71	1.79	1.32	1.47	1.07	1.17	3.57
By-Product credits, first year (mills/kWh)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Fuel cost, first year (mills/kWh)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.63	0.68	0.09	0.10	0.00
30-year levelized O&M costs														
30-year levelized fixed O&M, $/kW-yr	13.10	15.21	16.53	15.87	14.82	14.08	14.01	10.22	10.59	16.78	17.79	14.17	14.74	14.24
30-year levelized variable O&M (excluding fuel) mills/kWh	6.00	5.92	6.39	6.07	5.83	5.41	6.58	3.23	3.38	2.49	2.77	2.02	2.21	6.73
30-year levelized by-product credit, mills/kWh	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
30-year levelized fuel, mills/kWh	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.22	1.31	0.17	0.19	
30-year levelized fixed charges (capital) $/kW-yr	29.63	34.99	38.37	36.67	34.00	32.10	32.08	22.11	23.04	34.66	37.21	27.90	29.35	32.55
30-year levelized revenue requirement of FGD at levelized capacity factor 0.7 (mills/kWh)	12.97	14.11	15.35	14.64	13.79	12.94	14.10	8.50	8.86	12.10	13.06	9.05	9.59	14.36

Chlorine in coal increases FGD costs, but the presence of magnesium in the absorbent at concentrations higher than equivalent to the concentrations of chlorides and other strong acid ions reduces costs. Other strong acid ions include sulfate and nitrate ions. The lowest levelized revenue requirement at 93% SO_2 removal (LS 93.2.0 MG) is 8% less than that for the base case, assuming that active magnesium in limestone does not increase its price, but the highest (LS 93 E CL) is 4% higher than that for the base case.

The levelized revenue requirement for 93% removal of SO_2 from eastern coal is essentially the same if the absorbent is limestone or lime.

For magnesia scrubbing increasing the required SO_2 removal from 93 to 99% increases the levelized revenue requirement 6 to 8%. This difference is similar to the effect for limestone scrubbing.

Magnesia scrubbing seems to compare favorably with limestone scrubbing for eastern coal, even with a higher process contingency to allow for the fact that limestone scrubbing is much further developed than magnesia scrubbing (see Table 4-2). Magnesia scrubbing has its principal advantage in the cost of the flue gas treating system (see Table 4-1). Part of the advantage may be the result of using different types of absorbers for the different processes. This occurred partly because of the limitations of available data. Further study is needed to confirm the comparison of processes.

On the other hand, for western coal this study shows limestone scrubbing is more economical than magnesia scrubbing. This is explained by the process requirements. Western coal permits the use of low-cost, single-loop scrubbers at moderate L/G with limestone, but magnesia scrubbing always requires a prescrubber to protect the regenerable magnesia absorbent. With high-chlorine eastern coal, both types of scrubbing use either a first loop or a prescrubber in addition to the main absorber.

The effects of large increases in waste disposal costs on the eastern coal base cases and their western coal counterparts are shown in Figure 4-1.

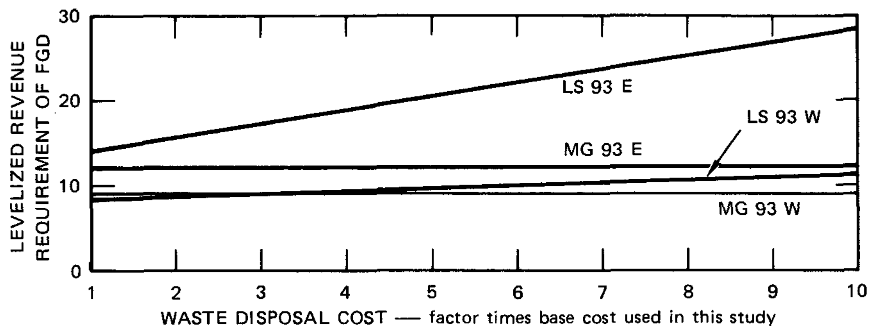


Figure 4-1. Effects of Waste Disposal Costs on Levelized Revenue Requirement of FGD

Figure 4-1 indicates a strong incentive to develop regenerative processes fully if waste disposal costs increase markedly. EPA is now considering future regulation of the disposal of fly ash and scrubber sludge. The possibility exists that some of these wastes will be considered hazardous and would thus require precautions that would enormously increase the cost of disposal, perhaps tenfold. However, neither the regulations nor their costs have yet been defined.

As noted in the design discussion, fly ash is removed before scrubbing, and its disposal was not considered in comparing the FGD processes. However, fly ash is used in fixing the sludge. The cost of sludge disposal could, therefore, be offset to a small degree by a credit for not having to dispose of this fly ash by itself. Such a credit would not be consistent with other EPRI evaluations and has not been taken in this report.

The last columns in Tables 4-1 through 4-3, Case LS 93E SP, may be compared with those for Case LS 93E to show the effects of using a single spray tower at high L/G instead of a two-loop scrubber. The capital costs are reduced, but the levelized revenue requirement is increased slightly. See Appendix C for details of this case.

Section 5

CONCLUSIONS AND RECOMMENDATIONS

The principal conclusions of this study are:

- Levelized revenue requirements of FGD increase about 8 to 9 mills/kWh as SO_2 removal requirements are tightened for plants burning western coal.
- The corresponding increases for plants burning eastern coal are between 1 and 3 mills/kWh.
- The greatest increases occur when scrubbing is required for part or all of the flue gas that otherwise does not require processing (i.e., flue gas from low-sulfur coal).
- Magnesia scrubbing may have economic advantages for high-chlorine, high-sulfur coals. Its reliability and economics have not been adequately demonstrated.
- Large increases in waste disposal costs would give regenerative processes a large advantage for high-sulfur coals.
- This study has taken account of effects of process requirements on L/G, construction materials, and other factors. New correlations of published data were made. These effects are not usually considered in this type of study, but their influence on costs is substantial. Further work of this kind is desirable to help to determine optimal design for specific processes.

Definitive public information should be developed to correct the following deficiencies:

- The most appropriate materials of construction for some environments, particularly materials that have minimum cost to provide acceptable reliability, have not been determined.
- Data on open-grid packed tower performance in terms of capacity, efficiency, and pressure drop for minimum cost design have not been collected.
- Comparisons of different types of absorbers for equivalent performance have not been made.
- Absorbent dissolution rates in different types of absorbers have not been determined.

- Data on magnesia slurry absorption in different types of absorbers and design requirements for countercurrent contacting at low L/G without bypassing are not available.

- Data on regeneration of magnesia, especially drying and calcining, and on fluid bed operations have not been collected.

- Several process innovations have not been studied by methods similar to those used in this report. These innovations include use of adipic acid as an absorbent promoter and forced oxidation to reduce sludge disposal cost. Forced oxidation may become important as a means of stabilizing sludge without fly ash.

Appendix A

DESIGN AND ECONOMIC BASES

DESIGN BASES

Design bases for SO_2 removal processes are summarized here. Included are assumptions on coal and ash compositions, coal combustion, boiler parameters, raw material compositions, and flue gas compositions. Eastern high-sulfur and western low-sulfur coals were chosen for the study. Both proximate and ultimate coal analyses, as well as ash analysis, are presented in Table A-1. Coal combustion assumptions and boiler parameters are shown in Tables A-2 and A-3.

In Table A-4, raw material compositions are given for the limestone, lime, magnesium oxide, and makeup water. Within certain limits, the quality of makeup water has little effect on FGD system operation. Therefore, the same makeup water composition, which is typical of Ohio River water, was used for all cases. Colorado River water was considered for the western cases, but an examination of data revealed that it had no "typical" composition. Compositions of the flue gas streams, shown in Table A-5, were calculated on the basis of assumed coal compositions and boiler parameters.

ECONOMIC BASES

Costs are based on process designs and equipment specifications for each of the 13 cases. Delivered equipment costs (including sales tax) were obtained from verbal vendor quotations and literature sources. Usually several sources were considered and a representative cost was chosen. Installation factors were applied according to the type of equipment. As an exception to this procedure, the cost of a complete sulfuric acid plant for regenerative cases was obtained by adjusting literature data. The direct installed costs were summed. Engineering and home office fees were assumed to be 11% of direct costs for process facilities. Estimates were built up on the following basis:

Table A-1

COAL COMPOSITIONS

Proximate Analysis	Eastern Coal (percent)	Western Coal (percent)
Moisture	12.0	30.4
Ash	16.0	6.4
Volatiles	33.0	31.2
Fixed carbon	39.0	32.2
Ultimate Analysis		
Carbon	57.5	47.85
Hydrogen	3.7	3.40
Nitrogen	0.9	0.62
Chlorine	0.1	0.03
Sulfur	4.0	0.48
Ash	16.0	6.40
Oxygen	5.8	10.83
Moisture	12.0	30.40
HHV (Btu/lb)	10,100	8,020
Ash Analysis		
Silica	45.0	31.59
Ferric oxide	20.0	4.55
Alumina	18.0	15.29
Titania	1.0	1.12
Calcium oxide	7.0	22.85
Magnesium oxide	1.0	4.74
Sulfur trioxide	3.5	16.55
Potassium oxide	1.9	0.44
Sodium oxide	0.6	1.27
Phosphorous pentoxide	0.2	0.75

Table A-2

COAL COMBUSTION ASSUMPTIONS

Carbon	All carbon completely oxidized to CO_2
NO_x	Assumed to be all NO, formed at rate corresponding to NSPS of 0.6 lb/10^6 Btu
Sulfur	SO_2 formed for 90% of western coal sulfur, 95% of eastern coal sulfur; no SO_3 formed
Ash	75% of ash forms fly ash
Unburned coal	None escapes with fly ash
Uncontrolled emissions	Emissions per million Btu without FGD: eastern coal 7.5 lb.; western coal 1.1 lb

Table A-3

BOILER PARAMETERS

Capacity	550 MW gross
Excess air	25%: 20% excess air to boiler plus 5% boiler leakage
Air leakage	Leakage in air preheater corresponds to 10% of entering flue gas
Plant thermal efficiency	Eastern coal, 38%; western coal, 36.5%
Water in air	0.013 lb H_2O/lb (80^oF, 60% relative humidity)

Table A-4

RAW MATERIAL COMPOSITIONS

Limestone Composition

$CaCO_3$	93.0%
Inerts	6.0%
H_2O	1.0%

Lime Composition

CaO	94.0%
Inerts	6.0%

Magnesia Composition

MgO	97.6%
CaO	1.5%
Inerts	0.9%

Makeup Water Composition

pH	7.3
T	60°F

Component	mg/ℓ
$CO_3^{=}$	84.4
$SO_4^{=}$	60.0
Ca^{++}	35.0
Mg^{++}	8.2
Na^{+}	12.0
Cl^{-}	15.0
NO_3^{-}	0.8

Table A-5

FLUE GAS COMPOSITIONS AND RATES

Component	Eastern Coal (mol %)	Western Coal (mol %)
SO_2	0.2959	0.0402
H_2O	8.190	11.92
CO_2	11.95	11.88
NO	0.0504	0.0478
N_2	73.68	70.46
O_2	5.824	5.656
HCl	0.00704	0.00252
	100.0	100.0
Flow rate (scfm)	1.172×10^6	1.287×10^6
Fly ash (lb/h)	58,690	30,780
Bottom ash (lb/h)	19,690	10,260
Coal requirements (lb/h)	489,100	641,200

Capital Investment

Process capital (as estimated)

General facilities:	5% of process capital
Project contingency:	15% of subtotal
Process contingency:	0% for limestone and lime; 10% for magnesia; based on process capital

Total Plant Investment (TPI)

Royalty allowance:	0.5% of process capital
Preproduction costs:	1 month fixed and variable operating costs, plus 1/4 month fuel cost, plus 2% of TPI
Inventory capital:	1 month of fuel plus 1 month other consumables other than water
Initial catalyst and chemicals:	included in process capital
AFDC*:	2 years at 8% of TPI compounded
Land:	$5,000 an acre

Total capital requirement (TCR): sum of above items

Fixed Operating Costs (first year)

Operating labor:	$12.90/h
Maintenance labor:	1.6% of TPI
Maintenance materials:	2.4% of TPI
Administrative and support labor:	30% of operating and maintenance labor

Total fixed O&M, first year: subtotal

Variable Operating Cost Excluding Fuel (first year)

Water:	$0.412/1,000 gal
Chemicals:	Limestone, $10.30/short ton; lime, $35/short ton; MgO, $200/short ton; coke, $32/short ton
Other consumables:	Steam $2.50/1000 lb; electricity 30 mills/kWh
Water disposal:	$8.75/short ton dry solids (includes capital charges)

Total variable (excluding fuel, first year): subtotal

*Allowance for funds used during construction.

By-Product Credits (first year): Sulfuric acid, no credit

Fuel Cost (first year): Low-sulfur oil, $2/million Btu

Levelizing Factors

O&M, excluding fuel: 1.886 x first year's cost

Fuel: 1.932 x first year's cost

Fixed Charges: 0.18 x TCR

Add levelized O&M and fixed charges, all in mills/kWh

Appendix B

EQUIPMENT DESIGN PRINCIPLES

PROCESS CONDITIONS

The process factors affecting equipment design include liquor-to-gas ratio (L/G), pH, liquor composition, relative saturation, and type of absorbent.

The L/G was determined for each case by the Radian RIPS program, assuming a dissolution rate for absorbent and an SO_2 vapor pressure 0.1 times the partial pressure in the gas leaving the absorber. Calculated L/G values were radically different for different absorber configurations and SO_2 removal requirements.

The L/G determines the pump requirements, which differ markedly for different cases. Because the dissolution rate affects L/G, Radian recommended that dissolution rates be studied further to increase the reliability of predictions of them.

Radian's program determines pH and liquor compositions, which are important factors in selecting construction materials. Similarly, the program determines the residence times and hold-tank volumes necessary to prevent scaling.

The following sections detail the equipment selection for the scrubber system.

Gas Saturation Provisions

In the designs in this report, quench sprays in the scrubber inlet ducts cool the flue gas entering the scrubber. These ducts must resist both hot and dry conditions and cool and wet conditions. Corrosion under the latter conditions is aggravated by solid deposits, low pH, and high chloride-ion concentrations. Organic coatings are not suitable because they fail when hot. Gunnited or castable linings may be suitable, but they are subject to cracking by differential thermal expansion. Accordingly, resistant alloys were chosen as construction materials for this corrosive service--a high-nickel molybdenum alloy such as Hastelloy C-276™ for prescrubbers and first-loop absorbers and a "20 family" alloy* for single-loop absorbers. These materials are used in the duct for a distance upstream from the absorber equal to three times the longest side of the duct perimeter.

Informal discussions with materials specialists indicate that no one is sure of the least expensive way to design the transition zone for some of the process conditions given in Table B-1 because the tightly closed water loops make conditions more severe than those in many existing FGD plants. The material selections given above are believed to be conservative, and to allow estimation of the trends of costs as a function of process conditions.

The design velocity for ducts in 915 m/min (3,000 ft/min). Soot blowers are provided to remove deposits from the wet-dry zone.

Absorbers

Absorbers remove SO_2 from flue gas and are therefore the heart of an FGD system. Table B-2 gives design data for the cases studied in this report.

All tower shells for the designs of this report are neoprene lined. Nozzles may be silicon carbide or Stellite™. Internal supports and piping are chosen according to the pH and chloride concentration previously shown in Table B-1; 316L is used if outlet pH $\geq$ 5.5 and Cl $\leq$ 10,000 mg/ℓ, 317L for pH $<$ 5.5 and Cl $\leq$ 10,000 mg/ℓ, a "20 family" alloy* for 3.0 $<$ pH $<$ 5.5 or Cl $>$ 10,000 mg/ℓ, and C-276 for pH $<$ 3.0. These criteria are generally conservative according to experience and internally consistent for the various cases.

Three types of absorbers are used: spray, grid, and mobile bed towers. The next three subsections explain how each was selected and designed for the requirements of a particular process.

*The "20 family" includes (in order of increasing molybdenum content) Incolloy 825™, Uddeholm 904L™, Jessops JS-700™, and Hastelloy G™.

Table B-1

CORROSION PARAMETERS

Case	First (or only) Loop or Prescrubber pH In	First (or only) Loop or Prescrubber pH Out	First (or only) Loop or Prescrubber Chloride (mg/1)	Second Loop pH In	Second Loop pH Out	Second Loop Chloride (mg/1)
LS93E	5.69	3.01	9,070	6.08	4.94	15
LS84E	5.35	4.11	9,600	N/A	N/A	N/A
LS99E	5.71	3.00	8,530	6.14	5.59	15
LS93ECL	5.26	2.71	28,300	6.08	4.94	15
LS93E0.5MG	5.65	3.13	9,120	6.15	4.80	15
LS93E2.0MG	5.90	3.43	9,470	6.27	5.02	15
LI93E	6.25	3.35	9,760	8.85	4.90	15
LS93W	5.51	4.85	20,100	N/A	N/A	N/A
LS99W	5.64	5.31	18,900	N/A	N/A	N/A
MG93E	1.27	1.23	28,700	7.94	5.45	324
MG99E	1.27	1.23	28,700	7.94	5.97	306
MG93W	1.29	1.26	28,200	7.83	6.03	823
MG99W	1.29	1.26	28,200	7.83	6.47	772

N/A = not applicable

Table B-2

ABSORBER DESIGN DATA

	First (or only) Loop or Prescrubber									Second Loop									
		L/G		Design Velocity			Packed Height or Ball Depth		Internal Supports & Piping‡		L/G		Design Velocity			Packed Height or Ball Depth		Internal Supports & Piping‡	No. of Beds
Case	Type	ℓ/m³	gal/kcf	m/s	fps	NTU§	m	ft		Type	ℓ/m³	gal/kcf	m/s	fps	NTU§	m	ft		
LS 93E	Spray	8.7	65.2	*	*	0.46	--	--	317L	Grid†	10.6	79.0	2.6	8.4	2.40	1.2	4	317L	--
LS 84E	Grid	19.4	145.0	2.1	7.0	1.99	1.2	4	317L	--	--	--	--	--	--	--	--		--
LS 99E	Spray	8.6	64.6	*	*	0.46	--	--	317L	Grid	14.2	106	2.4	7.9	4.35	2.4	8	316L	--
LS 93ECL	Spray	11.1	82.8	*	*	0.46	--	--	20	Grid	10.6	79.0	2.6	8.4	2.40	1.2	4	317L	--
LS 93E 0.5MG	Spray	8.7	65.0	*	*	0.46	--	--	317L	Grid	9.8	73.0	2.6	8.6	2.40	1.2	4	317L	--
LS 93E 2.0MG	Spray	8.9	43.8	*	*	0.46	--	--	317L	Grid	7.4	55.2	2.8	9.3	2.40	1.2	4	317L	--
LI 93E	Spray	7.8	58.7	*	*	0.46	--	--	317L	Grid	9.1	68.4	2.7	8.9	2.40	1.2	4	317L	--
LS 93W	Spray	8.6	64.0	2.6	8.5	2.82	--	--	20	--	--	--	--	--	--	--	--	--	--
LS 99W	Spray	10.0	75.1	2.6	8.5	4.77	--	--	20	--	--	--	--	--	--	--	--	--	--
MG 93E	Spray	1.3	10.0	*	*	--	--	--	C-276	Mobile bed	2.1	15.9	3.8	12.5	2.76	0.43	1.40	316L	3
MG 99E	Spray	1.3	10.0	*	*	--	--	--	C-276	Mobile bed	4.6	34.1	3.8	12.5	4.71	0.51	1.67	316L	5
MG 93W	Spray	1.3	10.0	*	*	--	--	--	C-276	Mobile bed	0.6	4.78	3.8	12.5	2.76	0.53	1.75	316L	3
MG 99W	Spray	1.3	10.0	*	*	--	--	--	C-276	Mobile bed	1.4	10.3	3.8	12.5	4.71	0.81	2.67	316L	5

*Set by size of second loop.

†Grid is open-packed honeycomb type.

‡316L and 317L are stainless steels; 20 is "20 family" (see text); C-276 is Hastelloy C-276 or an equivalent.

§Number of transfer units.

Spray Towers. Spray towers were chosen for all prescrubbers and for the first loop of the double-loop absorbers considered in this report. The single-loop absorbers for western coal cases are also spray towers.

Spray towers are often preferred because of:

- Low potential for fouling
- Freedom from flooding
- Low gas-side pressure drop.

However, spray towers may not provide reliably high absorption efficiency because of the potential for bypassing and because they provide only one theoretical stage of absorption. To obtain similar mass transfer, spray towers require higher L/G values than other absorbers, and they therefore lead to higher pumping costs.

Because prescrubbers need remove only HCl, highly soluble gas, and first-loop absorbers have moderate requirements for SO_2 absorption, spray towers are appropriate for these applications. They are also appropriate for western coal cases because the low SO_2 content of the gas leads to moderate L/G values.

Spray towers were not chosen for magnesia scrubbing because the low L/G required for magnesia leads to a significant potential for bypassing of the liquor and gas.

The design velocity for spray towers is consistent with commercial practice. However, the diameters of spray sections installed below other types of absorber are the same size as the absorbing section to avoid swedging the tower. These practices are reflected in the data presented in Table B-2.

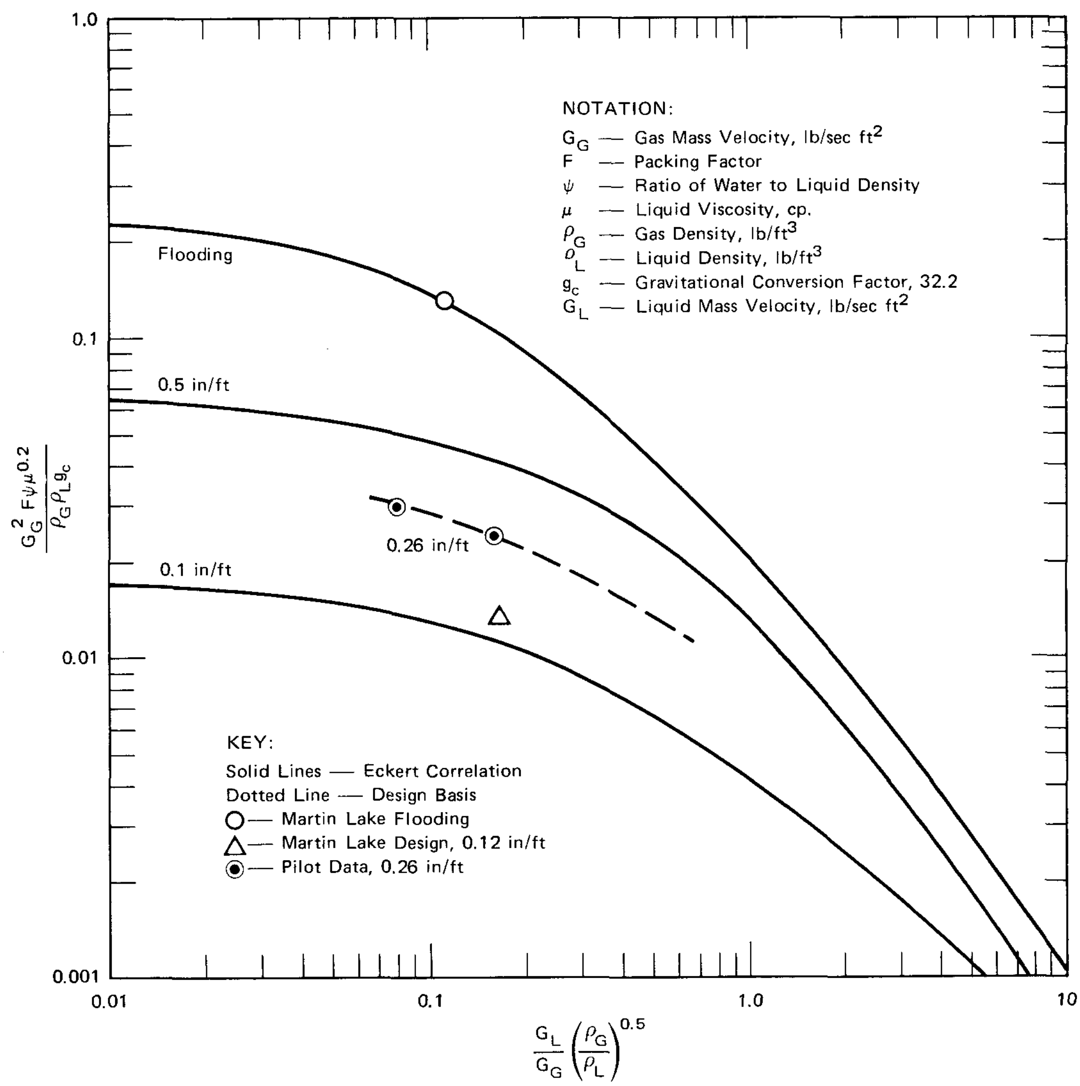
Grid Towers. Open-grid packed towers have been chosen for cases that require high SO_2 recovery with maintenance of a moderate L/G. These cases include the LS 84E and all second loops of nonrecovery processes. Several factors in this design are favorable for high recovery: countercurrent contact of gas and absorbent, little bypassing, and high alkali dissolution resulting from high alkali residence time.

High recovery is achieved at low pressure drop in open-grid packed towers. The available SO_2 absorption data were correlated (see Figure B-1) with earlier correlations, and SRI assumed that a constant pressure gradient would yield a constant height of a transfer unit. Although grid towers have been used for FGD, relatively little public information on this use is available. The correlation approach is believed to be the most reliable one for determining the effect of L/G on required tower size, but experimental confirmation would be desirable.

The solid lines in Figure B-1 give the correlation of Eckert (1,2), which includes various packings but does not include grid packings. Some grid packing data are available (3,4,5). Gleason applied this correlation to Munters wetted film contactor, apparently the CF 12060 type (3). The appropriate coordinates of the correlation are somewhat controversial (6). A single correlation is unlikely to apply to slurries and clear liquids in all packings, and the data for Martin Lake plotted on Figure B-1 show an unusual effect of velocity (7). However, the pilot plant data (5) are consistent with the correlation and give a reasonable basis for estimate of the effect of liquid loading on absorber performance. Therefore, the dotted line on Figure B-1 was used to calculate the design velocities given in Table B-2.

Mobile Bed Towers. Mobile bed towers were chosen for magnesia scrubbing to provide countercurrent contact (which is not achievable in the venturi used in some magnesia processes) at moderate pressure drop with little danger of bypassing. To facilitate designing the apparatus, SRI correlated pressure drop data anew and inferred mass transfer from magnesia-promoted lime scrubbing data. When necessary, the designs provide additional mobile beds to obtain the required number of transfer units.

Extensive pressure drop data from the EPA Shawnee pilot plant were empirically correlated (8), but the literature equation does not extrapolate to reasonable results outside the range of the data. Therefore, SRI developed a more reasonable correlation by postulating three terms for correlating pressure drop. The first term accounts for fluidizing the balls and is proportional to the ball depth. The second term accounts for pressure drop from gas flowing through the grids and is proportional to the velocity squared. The third term accounts for interaction between gas and liquor and is proportional to the product of the ball depth, the velocity squared, and an interaction



SOURCE: SRI

Figure B-1. Packed Tower Pressure Drop and Flooding function whose exponent is to be determined from the data. The following equation represents nonflooding data given by Epstein:

$$\Delta P = 0.134\ h_s + 0.0142\ V^2 + 1.53 \times 10^{-7}\ h_s\ (L/G \times V)^{1.4}\ V^2$$

where:

ΔP = pressure drop through 3 beds (in. H_2O)
h_s = total ball depth (in.)
V = gas velocity (ft/s)
L/G = liquid to gas ratio (gal/1000 ft^3).

This equation represents data for polypropylene and TPR balls, which have been superseded by heavier nitrile foam balls. The following equation represents more limited pressure drop data for absorption with magnesia-enhanced lime:

$$\Delta P = 0.174\ h_s + 0.0142\ V^2 + 2.0 \times 10^{-7}\ h_s\ (L/G \times V)^{1.4}\ V^2 \qquad (B.2)$$

Fortunately, designing near the area of operating experience was possible. The low L/G of magnesia systems would theoretically make high design velocities feasible, but the mist eliminator might be troublesome. Instead, a typical velocity of 3.8 m/s (12.5 ft/s) was chosen. A constant pressure drop of 15 cm H_2O (6 in. H_2O) was achieved for all the cases by varying the total height of balls. This pressure drop was approximately the same as the pressure drop when 90 to 99% removal was achieved with magnesia-enhanced lime in three beds (Shawnee run 608-2B, September 1976). The design basis assumed 93% removal with three beds at this pressure drop for Cases MG 93E and MG 93W. The greater NTU of Cases MG 99E and MG 99W requires five beds and a proportionately higher pressure drop.

Mist Eliminators

Mist eliminators are used to remove droplets of liquid or slurry from the gas leaving the absorber. Table B-3 summarizes designs for the various cases. Factors affecting the choices of eliminator include:

- Anticipated difficulty in cleaning
- Corrosive environment (pH, chloride, SO_2 concentration)

Table B-3

MIST ELIMINATOR DESIGN DATA

Case	Absorbent	SO_2 in Gas ppm	Absorber Loop Stoich., Mol Alkali per Mol SO_2 Sorbed	Absorber Loop Alkali Utilization, Percent	Chloride in Absorber Liquor, mg/ℓ	Inlet pH	First Stage					Second Stage				
							Passes	Material*	Configuration	Approx. Height, ft	Wash†	Passes	Material*	Configuration	Approx. Height, ft	Wash†
LS 93E	Limestone	194	1.62	61.5	15	6.08	2	316L	Slanted baffle	2	1,3	2	PP	Horizontal	1	2
LS 84E	Limestone	442	1.15	87.0	9,600	5.35	3	FRP	Horizontal	1	1,3	None				
LS 99E	Limestone	28	1.58	63.3	15	6.14	2	316L	Horizontal	1	1,3	2	PP	Horizontal	1	2
LS 93ECL	Limestone	194	1.62	61.5	15	6.08	2	316L	Slanted baffle	2	1,3	2	PP	Horizontal	1	2
LS 93E .5MG	Limestone	194	1.62	61.5	15	6.15	2	316L	Slanted baffle	2	1,3	2	PP	Horizontal	1	2
LS 93E 2MG	Limestone	194	1.62	61.5	15	6.27	2	316L	Slanted baffle	2	1,3	2	PP	Horizontal	1	2
LI 93E	Lime	194	1.21	82.6	15	8.85	3	316L	Horizontal	1	2,4	None				
LS 93W	Limestone	26	1.15	87.0	20,100	5.51	3	FRP	Horizontal	1	2,4	None				
LS 99W	Limestone	4	1.15	87.0	18,900	5.64	3	FRP	Horizontal	1	2,4	None				
MG 93E	MgO	194	<1.01	>99	324	7.94	2	FRP	Horizontal	1	2,4	2	PP	Horizontal	1	2
MG 99E	MgO	28	<1.01	>99	306	7.94	2	FRP	Horizontal	1	2,4	2	PP	Horizontal	1	2
MG 93W	MgO	26	<1.01	>99	823	7.83	2	FRP	Horizontal	1	2,4	2	PP	Horizontal	1	2
MG 99W	MgO	4	<1.01	>99	772	7.83	2	FRP	Horizontal	1	2,4	2	PP	Horizontal	1	2

*316L, stainless steel; FRP, fiberglass-reinforced plastic; PP, polypropylene.

†Wash arrangements: 1, continuous bottom; 2, intermittent bottom; 3, intermittent high-pressure top; 4, intermittent low-pressure top.

- Strength required
- Type of reheater downstream.

Deficiencies of mist eliminators have caused many operating problems that have spawned an extensive body of literature (9-14). Factors that determined the selections indicated in Table B-3 are described in the rest of this subsection.

The low utilization limestone cases are also the low chlorine cases; for them, Type 316L can be used in the first stage to make a strong structure capable of supporting loads of mud and resisting abrasion by high-pressure sprays. Shawnee experience shows that simple horizontal chevrons can be kept clean under similar conditions (13). Conservatively, slanted baffles were chosen for most cases, but horizontal baffles were chosen for use when SO_2 concentration is less than 50 ppm. Horizontal baffles were also chosen for lime scrubbing, which causes less fouling than limestone. The highest utilization cases are also high chloride cases. FRP systems were chosen for these mist eliminators to prevent chloride corrosion, because cleaning is easy at high utilization, and great strength is not needed.

The first stage of the mist eliminator removes slurry entrained in the gas, but some of the spray used to clean the first stage is entrained. In limestone cases using inline reheat downstream, a simple polypropylene second stage removes the entrained spray. In cases using hot air reheat, no second stage is provided because heater tubes are not in the path of the entrained droplets. In the lime scrubbing case, the second stage is omitted, as is usual practice (9). Only intermittent spraying is needed with lime scrubbing, and the amount of chloride carryover to the reheater is small.

Few data based on experience with mist eliminators in magnesia systems are available. The choice of a simple, horizontal, plastic construction is based on the nonfouling characteristics of magnesia, the recommendations of vendors, (14), and analogy with other cases. The chloride level is moderate in the magnesia cases. The mist eliminators chosen should be able to protect inline reheaters downstream.

Reheaters

Not all designs include reheaters (15), but the reheater selections summarized in Table B-4 for the various cases were intended to represent a conservative

Table B-4

REHEATER DESIGN DATA
(Each of 4 Exchangers Plus 1 Spare)

Cases	Tubes	Type	Surface* m^2	Surface* ft^2	Heat Used
LS 93E, ECL, E 0.5 MG, E 2.0 MG, LS 99E, MG 93E, MG 99E, LI 93E	316L bare	Inline	560	6,000	1.6
MG 93W, MG 99W	316L bare	Inline	600	6,500	1.6
LS 93W, LS 99W	CS finned	Hot air	170	1,800	2.4
LS 84E	CS finned	Hot air	160	1.700	2.3

*Based on bare outside tube area.

Percent of plant heat input transferred to flue gas.

approach to ensuring system reliability. All reheaters are heated 28°C (50°F) with 3.55 megapascal (500 psig) steam. The selections are:

- Inline reheaters with Type 316L bare tubes for double-loop scrubbers and magnesia systems
- Indirect hot air reheaters with finned carbon-steel tubes for high-chloride, single-loop systems.

The choice between the two methods may be difficult. In general, inline reheat is lower in capital and utility cost but higher in maintenance cost and lower in reliability. The reheaters discussed in this report are designed according to the chloride level of the system. Low-chloride environments (which are a result of both the process and the mist eliminator design) can use in-line tubes of 316L stainless steel. For high-cloride environments, indirect hot-air reheaters were chosen to prevent corrosion.

Reported steam pressures in use at commercial FGD plants range from 115 to 575 psig (15). The choice depends on plant-specific factors. SRI used 500 psig (3.55 megapascal) as a reasonable value for all cases.

FANS

Booster fans upstream of the scrubbers are used in these designs. The principal factors leading to this choice are the noncorrosive and nonfouling conditions in this location. Only if the system has no electrostatic precipitator are the fans likely to be downstream of the scrubbers. The fans are designed to handle a gas rate 10% higher than the full-load gas rate.

SOLIDS PROCESSING AND DISPOSAL

In nonrecovery processes in this report, sludges are thickened, filtered, fixed, and disposed as landfill. In magnesia recovery processes, magnesium sulfite is thickened, centrifuged, dried, calcined, and reused in the system.

Nonrecovery Processes

Design rates in the following tabulation were used. Actual rates vary considerably; these estimates are intented to be conservative.

	Eastern Coal	Western Coal
Thickener, $m^2/(t/d)$ [ft^2/short ton/d]	1.8 [27]	1.3 [19]
Filter, kg/h m^2 [(lb/h ft^2)]	342 (70)	489 (100)

In preparation for disposal, the sludge filter cake is blended with a small fraction of fly ash and lime, as in the standard EPRI approach. Western coal fly ash already contains the required lime. The blends used in this report are:

	Parts by Weight (dry)	
Component	Eastern Coal Sludge	Western Coal Sludge
Filter cake	100	100
Fly ash	10	30
Lime	4.4	0

A conventional system prepares the sludge for disposal. Belt conveyors deliver sludge to the fixation area. Pneumatic conveyors deliver fly ash and lime. These materials are blended in pug mills. The fixed sludge is loaded into trucks. Such systems have been studied by Michael Baker and by Bechtel (16, 17).

The fixation process converts the sludge from a thixotropic mass to a soil-like material with low permeability and increased load-bearing strength suitable for landfill. Fixation entails pozzolanic reactions. The blend can be varied to alter the properties of the landfill. Thus, the blends presented here are representative for generalized comparisons but not for specific sludge and landfill requirements.

Regenerative Processes

In the regenerative magnesia scrubbing processes analyzed in this report, the magnesium sulfite slurry from the absorber is thickened, dewatered by centrifuges, dried, and calcinated to regenerate magnesia for reuse. Occluded sulfate is reduced by coke added to the dried cake. The gas liberated by calcining contains SO_2, which is converted to sulfuric acid in a contact acid plant.

The thickeners are similar to those used for nonrecovery processes. Centrifuges dewater the thickener underflow as in demonstration plant practice. Centrifuges are chosen over filters because the high centrifugal force of centrifuges is helpful in dewatering and thorough dewatering is desirable to reduce the energy requirement for regeneration. Magnesium sulfite may form a trihydrate, which is difficult to dewater. The choice of centrifuges improves reliability for handling variations in crystal compositions.

Fluid-bed dryers and calciners were chosen over the alternative rotary kilns primarily because design information was available from the work of TVA (14). Neither type has been proven in practice. A dryer-calciner combination was chosen over single-stage calcining of the wet solid because the combination is characterized by a lower heat load and more concentrated and more uniform calciner gas, and it should ease control problems in the acid plant.

The fluid beds are fired with low-sulfur fuel oil. The dryer off-gas is injected into the flue gas to provide part of the reheat. The calciner gas is cooled in a waste heat boiler, generating steam to supply part of the requirement of the steam-heated reheater.

OTHER EQUIPMENT

Rubber-lined pumps are widely used in FGD slurry service and have been chosen for the designs in this report.

Ducts are sized for a gas flow rate of 915 m/min (3000 ft/min). All the trains and the bypass duct have an upstream damper of carbon steel and a downstream damper of stainless steel.

The absorbent for the process can be unloaded in a 40-hour week. Two unloading trains capable of handling two and one-half times maximum capacity are provided.

EPRI DESIGN STANDARDS AND ALLOWANCES

EPRI is attempting to improve the consistency of FGD cost estimates by setting design standards and allowances. Because this study is specifically intended to determine the effects of process variations on costs, the process condition standards (e.g., SO_2 removal and L/G) have not been applied. As much as possible, however, EPRI storage and spare equipment and material criteria and the like have been followed; these criteria are listed in Table B-5.

EPRI has also specified uniform groupings for capital costs. These groupings have been used in the economics section.

Table B-5

STANDARD AND SPECIFIC DESIGN ALLOWANCES

Flue Gas Treating

Four trains plus one spare

100% bypass capacity

No wash tray (differs from standard)

Prequench sprays

Process conditions as required (may differ from standard)

Pump redundancy 50%

Reheat

500 psig steam

Feed Preparation and Surge

100% spare wet ball mill

20% slurry, two vessels, 1.5 day surge each

150% lime-treating capacity for makeup water

Feed

200% reclaim capacity, two silos, 1.5 day surge each

Waste Separation and Storage

Thickener, 50% spare in one unit

Filter, 33% spare (four for three)

Blending, 100% spare

Storage silos, two silos, two days capacity each

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Appendix C

SINGLE-LOOP DESIGN FOR 93% SO_2 REMOVAL, EASTERN COAL

Figures C-1 and C-2 present a simplified FGD system flow diagram and FGD system design material balance for SO_2 removal supplied by Radian for single-loop, 93% SO_2 removal for eastern coal. SRI developed equipment specifications and economic analyses for this system that are comparable with those given for the other limestone scrubbing cases previously presented.

This design, designated LS 93E SP, was developed to confirm engineering judgment that two-loop scrubbers are preferable to unusually high L/G in a single-loop scrubber (total L/G is 144 in a two-loop process, 255 in a single-loop process). The Radian results for LS 93E SP are presented here because they were developed after the Radian report had been submitted in final form.

In comparison with a two-loop system, the higher capital cost of pumps for LS 93E SP is more than offset by the reduction in absorber cost and the elimination of one thickener. The absorber cost comparison depends on use of the same design velocity in spray towers regardless of L/G. The data needed to judge the validity of this design approach are not available. Radian uses two thickeners in two-loop systems to improve process control; however, a two-loop system can also be operated with a single thickener. Thus, the uncertainties in the design favor the two-loop system, reinforcing the conclusion that the levelized revenue requirement of FGD is slightly lower for the two-loop system because of its advantage in operating cost.

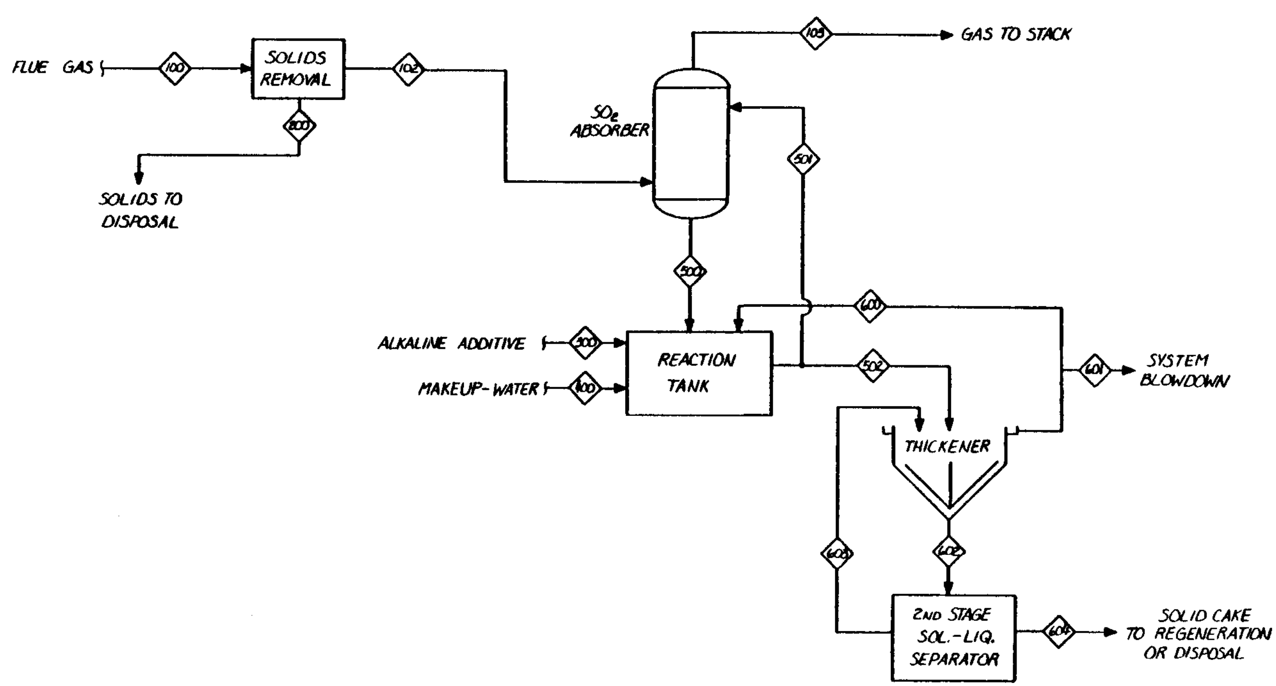

Figure C-1. FGD System Flow Diagram

FGD SYSTEM DESIGN

DESIGN CONDITIONS

1. COAL ANALYSIS

COMPONENT	WT. %
ASH	16.0
S	4.0
H	3.7
C	57.5
N	0.9
O	5.8
CL	0.1
H_2O	12.0
TOTAL	100.0

HIGHER HEATING VALUE = 10,100 BTU/LB.
FUEL CONSUMPTION = 245 TON/HR.
EXCESS AIR & BOILER LEAKAGE = 25%
PREHEATER AIR LEAKAGE = 10% OF FLUE GAS ENTERING

2. CAKE LIQUOR COMPOSITION

COMPONENT	MG/L	COMPONENT	MG/L
CA	5270	CO_3	297
MG	44.4	NO_3	4.33
NA	65.0	SO_3	38.7
CL	8650	SO_4	1060

3. RELATIVE SATURATION

SPECIES	500	501	750	751
$MgCO_3$				
$MgSO_3 \cdot 3H_2O$				
$Mg(OH)_2$				
$CaCO_3$	4.66×10^{-4}	0.200		
$CaSO_3 \cdot \frac{1}{2}H_2O$	0.376	2.27		
$CaSO_4 \cdot 2H_2O$	1.28	1.20		

4. REACTION TANK VOL. = 1.91×10^6 GAL.
HOLD TANK VOL. (PRESCRUBBER) = GAL.

5. L/G = 255 GAL./MACF OUTLET
L/G (PRESCRUBBER) = NONE

6. NTU = 2.82

CASE DESCRIPTION

CASE: LS 93 E SP

FGD SYSTEM: LIMESTONE

SO_2 REMOVAL: 93%

HCL REMOVAL (PRESCRUBBER):

COAL TYPE: EASTERN

TOWER TYPE: SPRAY

STREAM NUMBER	200	300	400	500	501	502	600	601	602	603	604	730	740	750	751	752	760	761
TOTAL FLOW, LB./HR.	58,690	66,720	308,500	2.060×10^8	2.062×10^8	725,400	580,700	0	289,300	144,600	144,700							
SOLID FLOW, LB/HR	58,690	66,720	0	2.465×10^7	2.467×10^7	86,770	0		86,790	0	86,790							
LIQUID FLOW, LB/HR			308,500	1.814×10^8	1.816×10^8	638,600	580,700		202,500	144,600	[illegible]							
SLURRY SOLIDS CONTENT, WT%			0	12.0	12.0	12.0	0		30.0	0	60.0							
LIQUID FLOW, GPM			616.6	361,500	361,900	1273	1157		403.7	288.3	115.5							
SLURRY FLOW, GPM			616.6	381,300	381,700	1342	1157		451.0	288.3	185.2							
TEMPERATURE, °F			60.0	128	128	128	128		128	128	128							
TDS, WPPM			217	15,750	15,540	15,540	15,200		15,200	15,200	15,200							
pH			7.30	4.43	5.52	5.52	5.88		5.88	5.88	5.88							
SOLIDS COMPOSITION, WT. %																		
$CaCO_3$		93.0		8.58	8.65	8.65			8.54		8.54							
$CaSO_4 \cdot 2H_2O$				18.64	18.62	18.62			18.81		18.81							
$[(CaSO_3)_{.85}(CaSO_4)_{.15}] \cdot \frac{1}{2}H_2O$				67.38	67.32	67.32			67.27		67.27							
INERTS		7.0		5.40	5.40	5.40			5.38		5.38							

STREAM NUMBER	100	101	102	103
ACFM	1.811×10^6		1.811×10^6	1.495×10^6
LB. MOLES/HR.	195,900		195,900	207,200
TEMPERATURE, °F	300		300	128
PRESSURE, PSIA	14.7		14.7	14.7
FLY ASH, LB./HR.	58,690		0	0
COMPOSITION, MOLE %				
SO_2	0.2959		0.2959	0.01939
H_2O	8.190		8.190	14.14
CO_2	11.95		11.95	11.45
NO	0.0504		0.0504	0.04715
N_2	73.68		73.68	68.94
O_2	5.824		5.824	5.410
HCL	0.00704		0.00704	0

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Figure C-2. FGD System Design

Appendix D

ABBREVIATIONS

a ft^3	actual cubic feet
AFDC	allowance for funds used during construction
atm	standard atmosphere
bhp	brake horsepower
Btu	British thermal unit
CS	carbon steel
d	day
D	diameter
FGD	flue gas desulfurization
FRP	fiberglass-reinforced plastic
ft	feet
gal	U.S. gallon
kcal	kilocalorie
kg	kilogram
h	hour
HTU	height of a transfer unit
I.D.	inside diameter
in.	inch
J	joule
kW	kilowatt
ℓ	liter
lb	pound
L/G	liquid to gas ratio
m	meter
mg	milligram
min	minute
MW	megawatt
N	newton
ng	nanogram
NSPS	new source performance standard
NTU	number of transfer units

O&M	operating and maintenance
PP	polypropylene
psia	pounds per square inch absolute
psig	pounds per square gauge
s	second
scfm	standard cubic feet per minute
t	metric ton
TCR	total capital requirement
TPI	total plant investment
T/T	tangent to tangent
m	micrometer