

CONCEPTUAL DESIGN OF AN SO_x
SCRUBBER SYSTEM UTILIZING PLANT WASTE

by

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Thesis submitted to the Graduate Faculty of the
Virginia Polytechnic Institute and State University
in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

in

Mechanical Engineering

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October, 1975

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II. ACKNOWLEDGEMENTS

The author wishes to express sincere appreciation to the members of his advisory committee: Professors _____ and _____, Chairman. He is especially grateful to _____ for his assistance on this research.

Support of this investigation by the National Science Foundation is gratefully acknowledged.

Finally, the author is indebted to his wife _____, without her encouragement none of this would have been possible.

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VI. LIST OF SYMBOLS

a	interfacial area
B	atmospheric pressure
c	absorption rate
cfm	volumetric flow rate
C_p	specific heat
D	stack diameter
EC	energy cost
G	mass velocity
h	enthalpy
ha	volumetric heat transfer coefficient
HHV	higher heating value
H_p	shaft horsepower
Ka	volumetric mass transfer coefficient
L	length
M	molecular weight
\dot{m}	mass flow rate
Δp	pressure difference
Q	heat transfer rate
R	gas constant
R'	removal efficiency
SD	stack draft
SGE	steam generator efficiency
T	temperature

V gas velocity
v weight fraction of solute in solvent
ω humidity ratio

General Subscripts

a air
FG flue gas
ge flow loss
i interface
L liquid
RH reheater
SO₂ sulfur dioxide
mu makeup water
mue makeup water for evaporation

VII. INTRODUCTION

With the ever increasing demand for an economical and reliable energy source, environmentally available and acceptable to produce electrical energy, methods for efficient control of SO_x from combustion processes associated with power generation must be developed.

Sulfur dioxide (SO_2) is the predominate oxide of sulfur formed by the combustion of fuels containing sulfur compounds. Even when oxygen is present in large stoichiometric excess, sulfur trioxide (SO_3) is seldom found in amounts greater than a few percent of sulfur dioxide [1]. These constituents are usually denoted collectively by the term SO_x .

Atmospheric pollution caused by sulfur oxides and flyash emitted from coal-fired power plants is experienced almost universally and, with the present state of sulfur oxides emission control, seriously limits exploitation of coal energy resources. As a consequence, the demand for low sulfur gas and oil has increased leaving the high sulfur coal reserves unused. This represents a considerable amount of available energy. Table 1 shows the estimated remaining bituminous coal reserves by rank, sulfur content and state.

Several control methods have been proposed for reducing SO_x emission. Usually, these control methods are grouped:

- 1) fuel selection - use of low sulfur fuel
- 2) coal gasification - conversion of high-sulfur coal to low-sulfur oil and gas
- 3) fluidized-bed combustion - removing SO_2 within the boiler

Table 1. Estimated Remaining Bituminous Coal Reserves, Million Short Tons [2]

Coal rank and State	Sulfur content, percent									Total
	0.7 or less	0.8-1.0	1.1-1.5	1.6-2.0	2.1-2.5	2.6-3.0	3.1-3.5	3.6-4.0	Over 4.0	
Bituminous coal:										
Alabama.....	889.2	1,189.3	5,421.7	5,182.8	458.8	417.4	--	--	18.6	13,577.8
Alaska.....	20,287.4	1,100.0	--	--	--	--	--	--	--	21,387.4
Arkansas.....	--	--	1,128.4	293.1	154.0	--	40.3	--	--	1,615.8
Colorado.....	25,178.3	37,237.2	--	--	--	--	--	--	--	62,415.5
Georgia.....	--	76.0	--	--	--	--	--	--	--	76.0
Illinois ²	--	573.7	4,942.4	2,615.1	809.6	16,583.8	33,650.4	57,652.2	19,062.0	135,889.2
Indiana.....	197.5	173.0	3,645.2	4,248.8	3,543.4	4,110.5	10,872.8	5,105.9	2,944.0	34,841.1
Iowa.....	--	--	--	--	--	--	117.1	--	6,405.4	6,522.5
Kansas.....	--	--	519.9	519.7	1,038.7	2,070.6	4,148.0	8,287.3	4,153.8	20,738.0
Kentucky:										
West.....	--	--	1,119.6	162.0	336.3	3,793.6	12,759.3	13,643.3	5,081.3	36,895.4
East.....	13,639.9	8,491.9	2,286.8	1,658.8	1,158.3	2,154.4	24.7	--	--	29,414.8
Maryland.....	--	--	--	124.6	191.8	208.2	378.6	56.4	220.4	1,180.0
Michigan.....	--	--	--	--	--	--	--	205.0	--	205.0
Missouri.....	--	--	--	--	--	--	6,456.7	20,669.2	51,634.1	78,760.0
Montana.....	51.2	218.2	205.0	397.2	400.0	175.0	40.0	27.0	591.0	2,104.6
New Mexico.....	5,212.0	5,474.0	--	--	--	--	--	--	--	10,686.0
North Carolina.....	--	--	--	--	--	110.0	--	--	--	110.0
Ohio.....	--	611.0	369.0	2,110.2	2,750.4	7,810.5	9,785.3	10,148.2	8,439.4	42,024.0
Oklahoma.....	250.6	772.2	825.0	368.1	--	--	577.2	19.1	490.6	3,302.8
Oregon.....	--	14.0	--	--	--	--	--	--	--	14.0
Pennsylvania.....	44.0	1,154.4	7,624.4	12,424.9	19,689.5	9,995.6	5,287.6	1,150.5	580.6	57,951.5
Tennessee.....	3.3	160.9	715.9	258.7	178.2	190.5	219.7	43.8	68.5	1,839.5
Texas.....	--	--	--	--	7,978.0	--	--	--	--	7,978.0
Utah.....	8,551.4	13,584.0	--	1,524.9	--	--	--	--	3,997.7	27,658.0
Virginia.....	1,981.5	6,077.5	1,637.1	--	123.9	--	--	--	--	9,820.0
Washington.....	898.9	672.1	--	--	--	--	--	--	--	1,571.0
West Virginia.....	20,761.0	26,710.6	21,819.7	13,290.6	8,496.1	2,491.8	3,147.4	5,949.2	--	102,666.4
Wyoming.....	6,222.2	6,596.6	--	--	--	--	--	--	1.1	12,819.9
Other States ³	--	616.0	--	--	--	--	--	--	--	616.0
Total.....	104,168.4	111,502.6	52,260.1	45,179.5	47,307.0	50,111.9	87,505.1	122,957.1	103,688.5	724,680.2
Percent of total.....	14.4	15.4	7.2	6.2	6.5	6.9	12.1	17.0	14.3	100.0

immediately after combustion

- 4) flue gas scrubbing - removing SO_2 before the stack.

For this study, the gas scrubbing method was chosen.

A basic division in scrubbing processes is throwaway versus recovery which refers to the end product resulting from gas scrubbing. Throwaway processes produce a sludge which must either be stored at the plant or transported to an environmentally acceptable disposal site. Recovery processes produce a marketable product, usually sulfuric acid or elemental sulfur, in an effort to offset the gas treatment cost.

Of the several scrubbing methods available, lime-limestone wet scrubbing is considered to exhibit the most promise. However, some major problems associated with lime-limestone and other types of scrubbing devices have not been solved.

In order to meet emission control standards not only for air pollution but also water pollution (zero effluent), the pollution control system must be designed accordingly. Power plants presently operating are allowed to discharge cooling tower blowdown and diluted ashpond and ash transport water, but eventual control devices must be implemented which means, in some cases, treating large amounts of water. In an effort to approach a zero effluent plant with SO_x control, the use of cooling tower blowdown and flyash pond water were considered as possible scrubber liquors for SO_x removal. The advantage of such a system could be substantial since the SO_x sorbent is provided by the power plant waste and the waste water volume treated reduced.

Therefore, the object of this research was to design a scrubber process and system utilizing plant waste for SO_x control.

VIII. REVIEW OF THE LITERATURE

1. Introduction

The removal of sulfur oxides is not a new problem. Inhabitants of areas close to ore smelter operations have coped with unwanted fumes resulting from sulfur oxidation for more than a century. It was not until the disastrous incidents in Belgium, London and Donora, Pa. that the hazardous effects of air pollution were realized.

The physiological effects from exposure to SO_2 may be acute or chronic. Aggravation of systems in asthmatic and cardiopulmonary sufferers has been demonstrated repeatedly for daily SO_2 levels below the threshold for excessive mortality [3].

Effects of SO_2 on vegetation has been studied and the results with regard to the toxic effects on plant life accepted.

Presently, little disagreement occurs over the potential hazards which can result from exposure to SO_2 . However, there is controversy over realistic values for concentrations and times of exposure.

In an effort to accelerate implementation of air pollution control, Congress amended the Clear Air Act in 1970 to establish strict requirements and timetables for air pollution control. As a result, the US Environmental Protection Agency (EPA) has published primary and secondary ambient - air - quality standards as shown in Table 2. Limits for SO_2 emission from new stationary sources were set at 1.2 lb SO_2 /million Btu for coal firing and 0.8 lb SO_2 /million Btu for oil fired plants.

Table 2. National SO₂ Ambient Standards [4]

<u>Standard</u>	<u>Concentration</u>		<u>Description</u>
	μ g/m ³	ppm	
Primary	80	0.03	Annual arithmetic mean
	365	0.14	24-hr maximum not to be exceeded more than once per year
Secondary	60	0.02	Annual arithmetic mean
	260	0.1	24-hr maximum not to be exceeded more than once per year
	1300	0.5	3-hr maximum not to be exceeded more than once per year

2. Evolution of Alkaline Scrubbers

The earliest work on an alkaline scrubbing process was performed by London Power Co. around 1930 [5]. Flue gas scrubbers were installed in the Battersea and Bankside power stations using Thames River water on a once through basis. Analysis of the water revealed that the primary absorbents were dissolved calcium carbonate (CaCO_3) and magnesium carbonate (MgCO_3).

To avoid the water pollution problem encountered from using once through methods, Imperial Chemical Industries and James Howden and Co. combined and developed a closed loop process that produced only a solid waste end product (calcium sulfate and calcium sulfite). Since this earlier work, several new processes have been designed and tested in the US, USSR and Japan.

3. Operation of the Lime-Limestone Scrubber

Currently, lime-limestone scrubbing is being adopted as the basic method for controlling sulfur dioxide emission from power plants [6]. Typical lime-limestone scrubber systems are shown in Figs. 1 and 2.

The scrubber system is composed of three main components--the scrubber, pump tank and thickener. The purpose of the scrubber is to allow intimate mixing of the flue gas with a lime or limestone slurry that contains 5-15% solids by weight. The effluent from the scrubber is transported to the pump tank where chemical reactions go to completion producing a disposable precipitate. Makeup lime or limestone is added to the tank and the remaining solution returned to

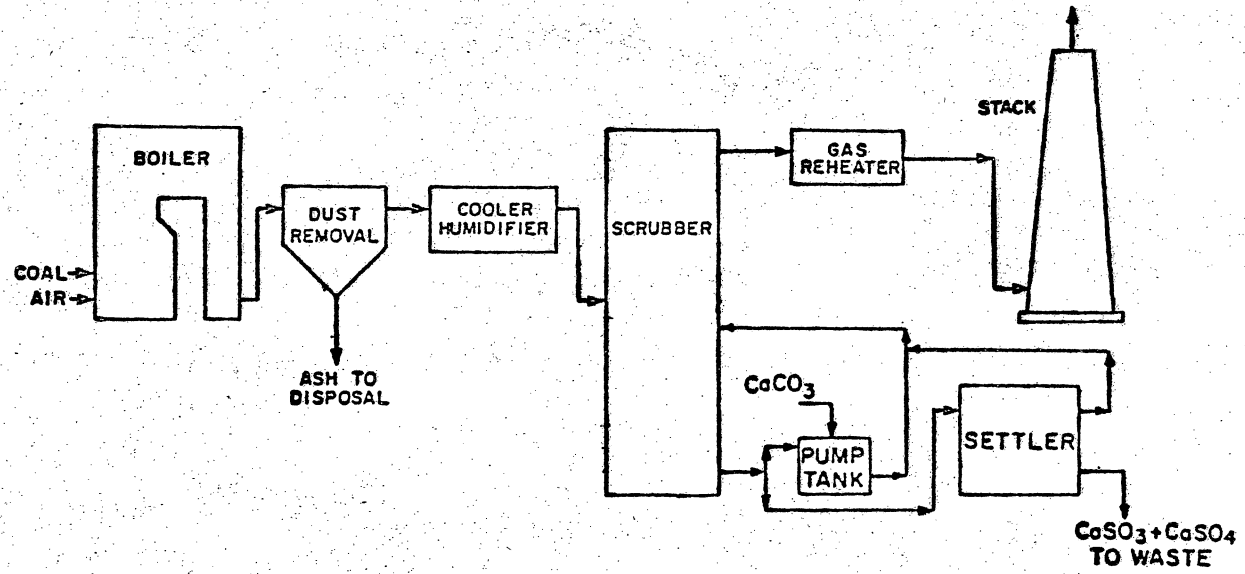


Figure 1. Flowsheet of Typical Limestone Scrubber

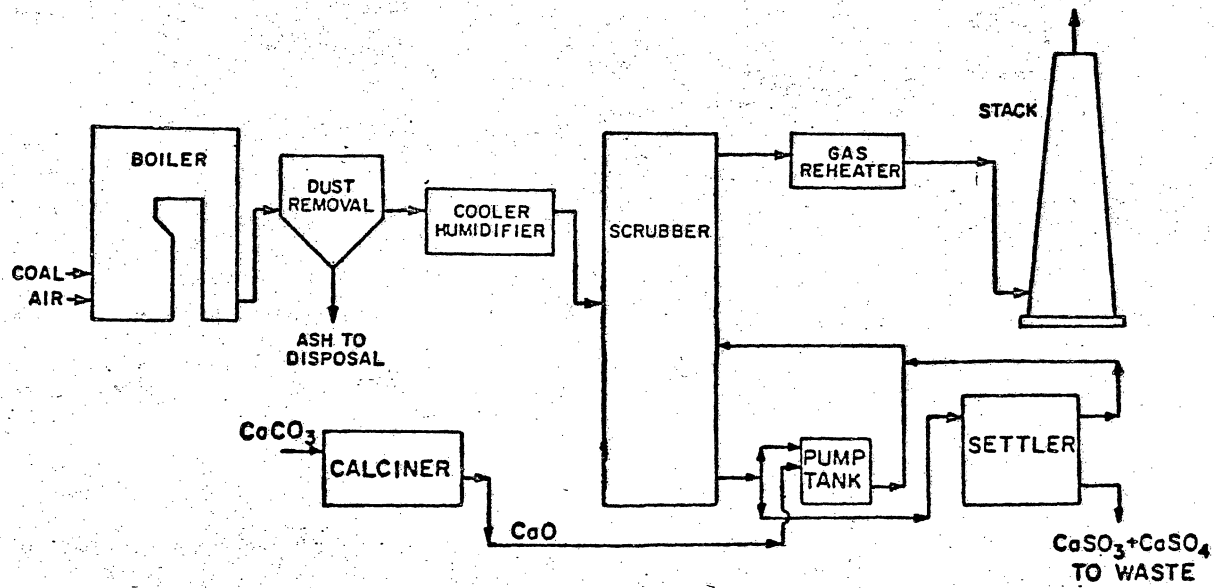


Figure 2. Flowsheet of Typical Lime Scrubber

the scrubber. The thickener or clarifier is used to concentrate the precipitate from the pump tank. The suspended solids are concentrated by the clarifier and then removed to a pond or disposal site.

4. Problems with Lime-Limestone Scrubbing

The major problems associated with lime-limestone scrubbing are numerous and are best described by summary:

- 1) Gas-liquid distribution: for treating large volumes of gas, obtaining the proper distribution of gas to liquid over the cross section of the scrubber is difficult.
- 2) Independent scrubber operation: provisions must be made for taking a scrubber(s) out of service without a plant shutdown or affecting other scrubbers in operation.
- 3) Corrosion: conclusions of the ICI - Howden work indicated that mild steel is a satisfactory construction material if the pH does not fall below 6.2 [7]. However, pH instability has been a problem.
- 4) Deposit buildup: at the wet-dry interface located at the scrubber inlet and exit, a deposit forms which can eventually restrict gas flow and lead to excessive pressure drop.
- 5) Mist elimination: in an attempt to prevent low pH and therefore to reduce scaling, a high rate of slurry recirculation is used. Combined with high turbulence, this promotes mist formation. Effective mist control is important since excessive carry over may cause deposition of solids on

the reheater and fan surfaces with subsequent component failure.

- 6) Gas reheat: reheating of the gas after scrubbing is presently the accepted method for restoring gas buoyancy and to avoid condensation of corrosive liquids in the breeching and stack.
- 7) Scrubber type: the types and problems associated with the type of scrubbers are varied and have been summarized by Slack [8] in Table 3 for the scrubbers shown in Figs. 3 through 8.
- 8) Scaling: scale formation in flue gas scrubbers has been the main problem throughout lime-limestone scrubbing studies. The problem occurs because the liquid phase returned to the scrubber is saturated with calcium sulfite and calcium sulfate which tends to crystallize in the scrubber after SO_2 is absorbed. The best resistance to scaling has been obtained with low pressure drop type scrubbers with a high solids content (12-15%) in the slurry. The scrubbers that have had the most successful (no-scale) operation are the spray and mobile bed ("ping-pong ball packing") type scrubbers.
- 9) Waste disposal: the end product from lime-limestone scrubbing is calcium sulfate (gypsum) and calcium sulfite. This adds to the already considerable ash disposal problem. Calcium sulfate occurs in abundance in nature and calcium sulfite is known to occur in high pH areas. Because both are relatively insoluble, disposal by dumping appears to be the

Table 3. Comparison of Scrubber Types [8]

<u>Type</u>	<u>Holdup</u>	<u>Counter-current</u>	<u>Plugging Resistance</u>	<u>Turndown Ratio¹</u>	<u>Pressure Drop</u>	<u>Dust Removal</u>
Fixed Packing	Fair	Yes	Fair	Good	Medium	Poor
Tray ²	Good	No	Good	Poor	Medium	Good
Mobile Bed ³	Fair	No	Good	Poor	Medium	Good
Venturi	Poor	No	Good	Good ⁴	High	Good
Spray	Poor	Yes	Good	Good	Low	Poor
Doyle	Good	No	Good	Poor	Medium	Good

¹For operability rather than scrubbing efficiency.

²With glass spheres on the tray.

³Hollow plastic balls.

⁴With variable throat.

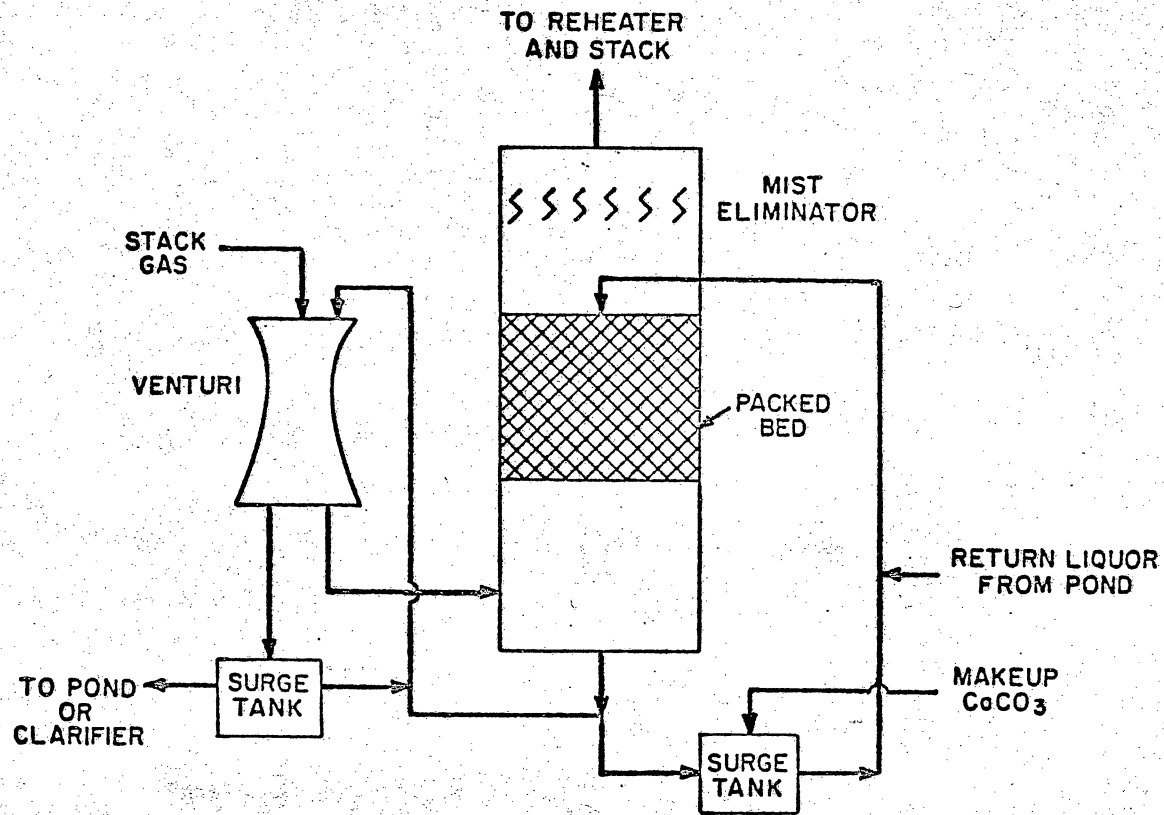


Figure 3. Packed Bed Scrubber with Upstream Venturi for Particle Collection [8]

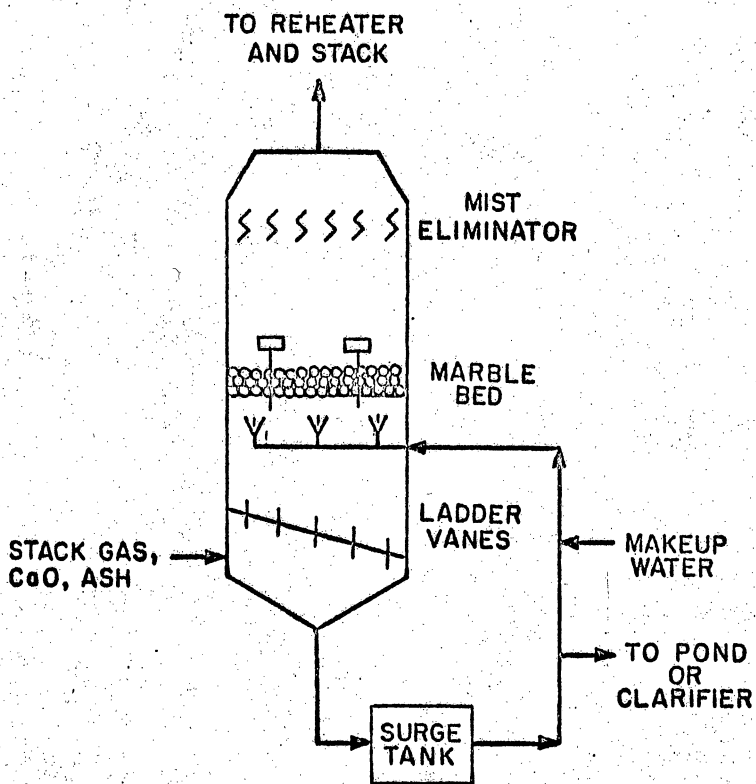


Figure 4. Tray Scrubber with Marble Bed Packing [8]

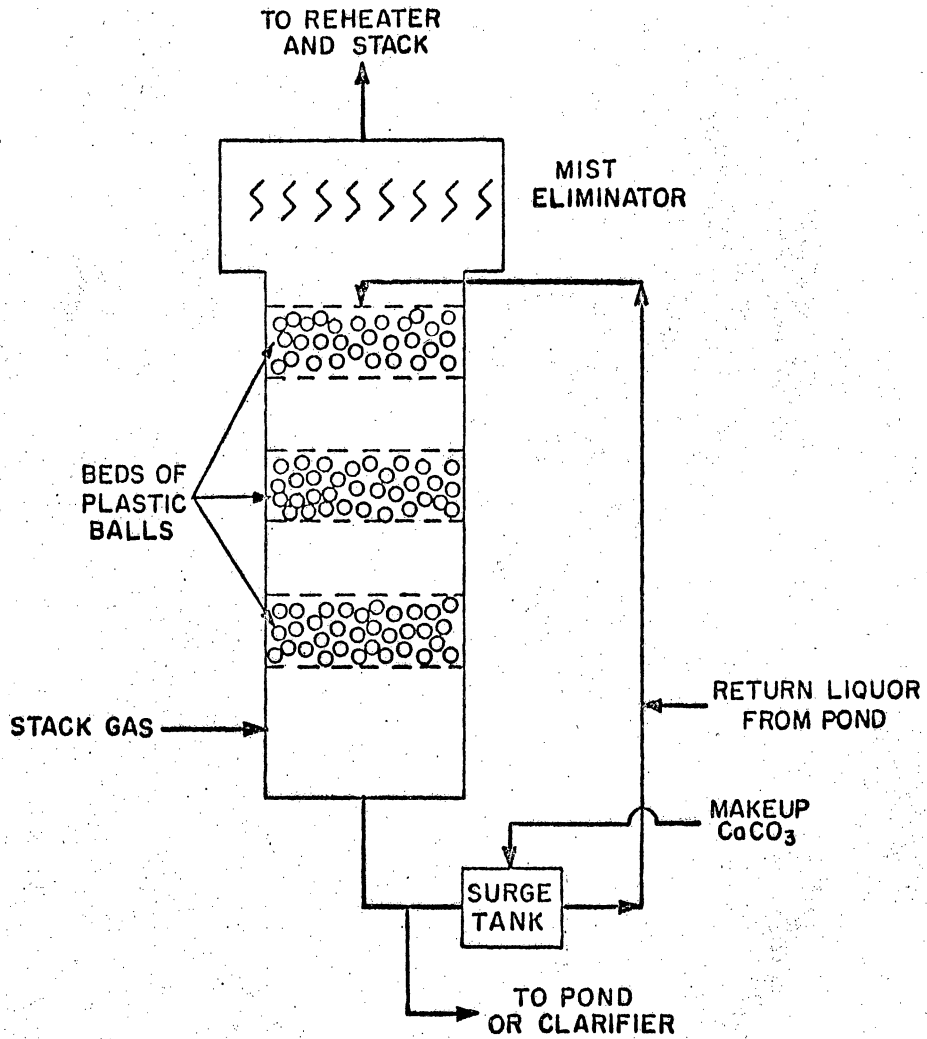


Figure 5. Mobile Bed Scrubber with "Ping Pong Ball" Packing [8]

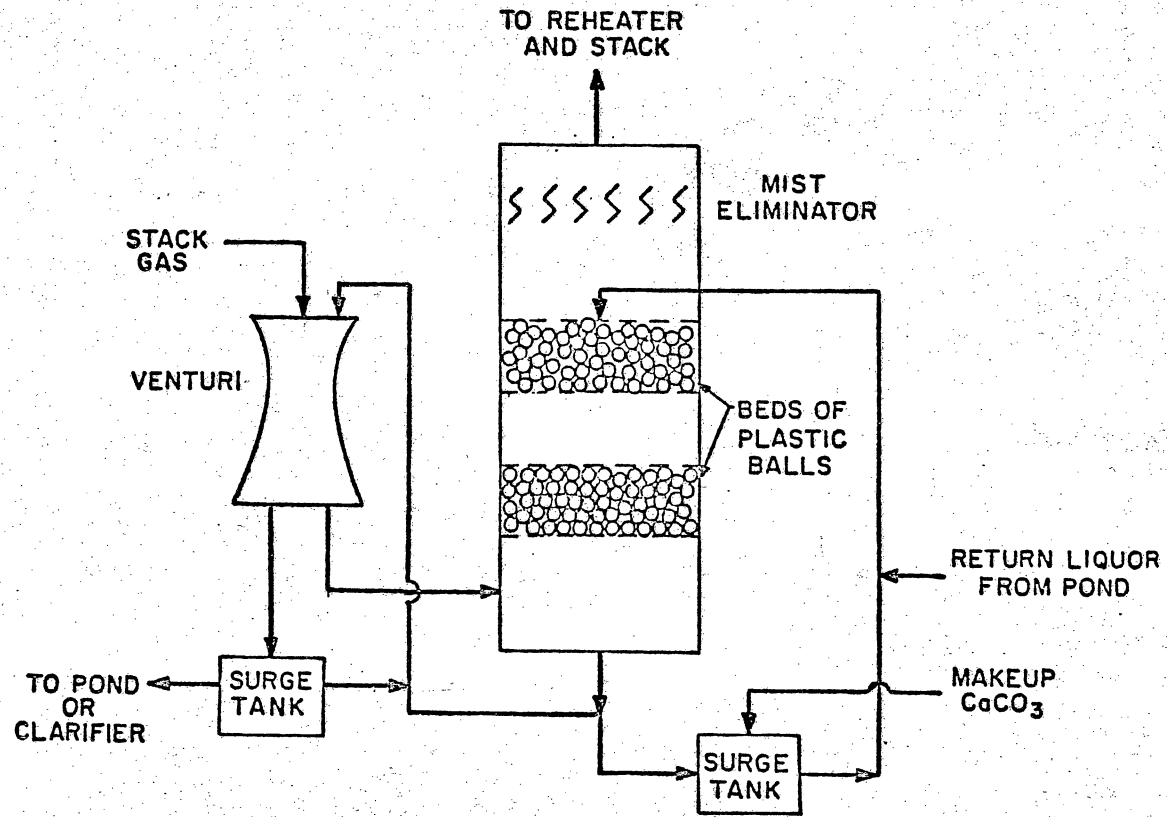


Figure 6. Venturi Scrubber in Series with Mobile Bed Scrubber [8]

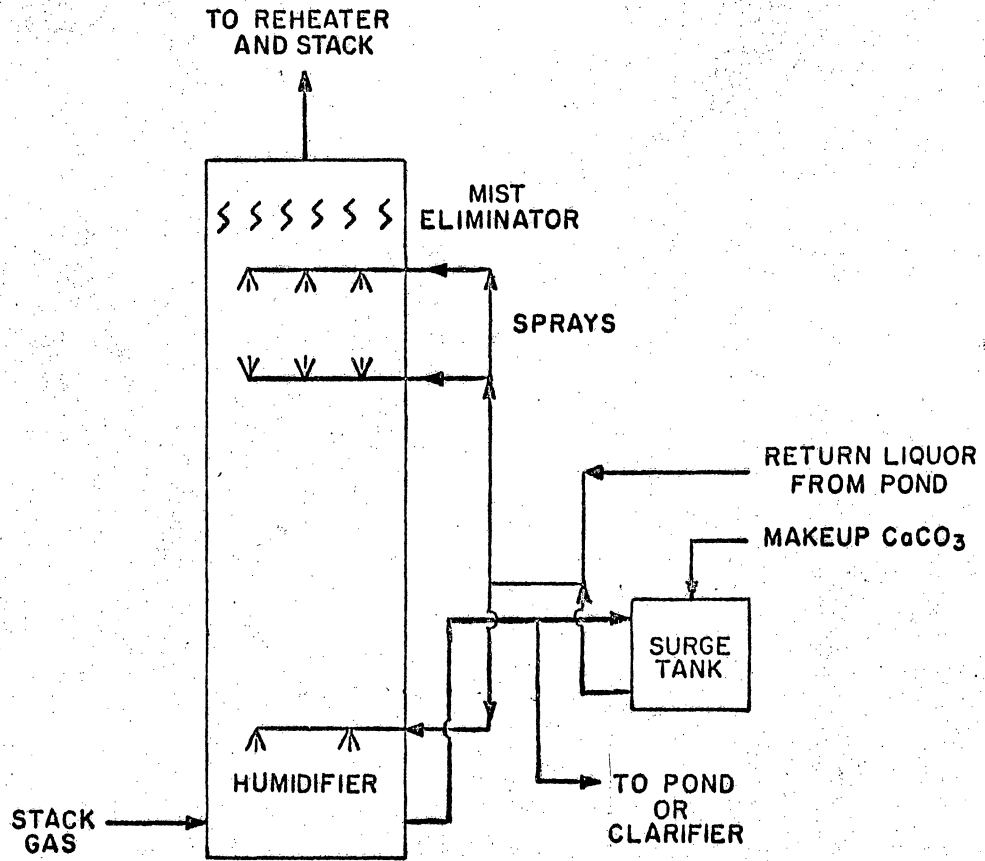


Figure 7. Spray Chamber Scrubber [8]

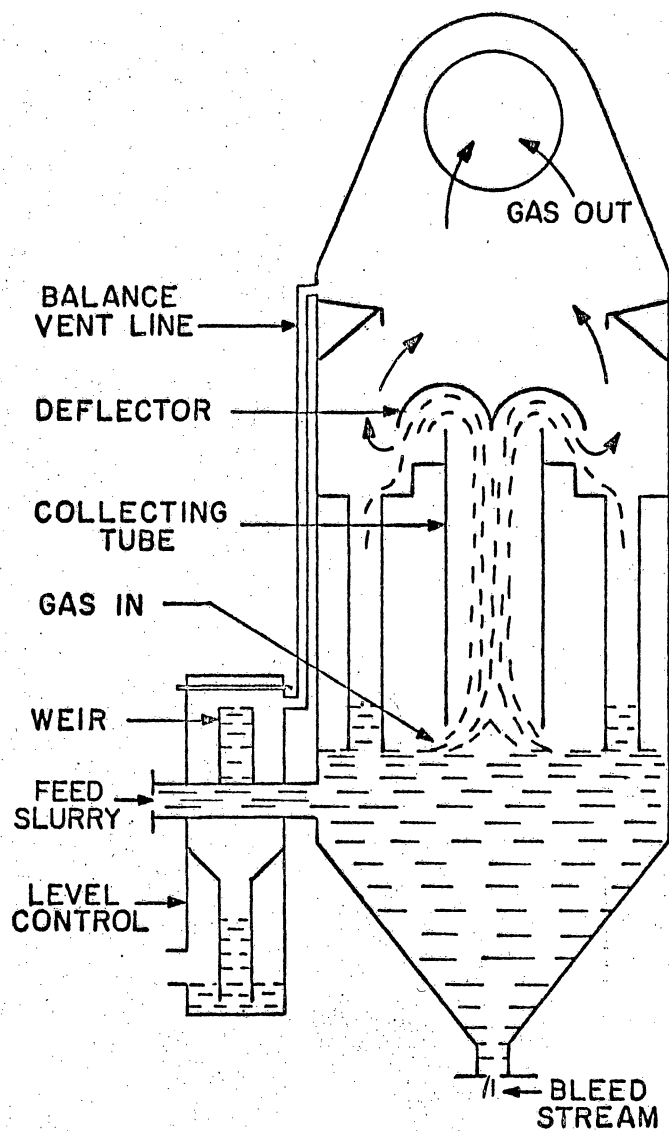


Figure 8. Doyle Impingement-Type Scrubber [8]

acceptable solution. The main problem for most utilities is obtaining the real estate for the disposal site.

- 10) Scrubber chemistry and kinetics: chemistry of lime-limestone scrubber systems is most complicated due to the large number of species present in the system. The main reaction in the scrubber are assumed to be:

- 1) absorption of SO_2
- 2) hydrolysis to form sulfurous acid (H_2SO_3)
- 3) reaction of sulfite ion from the H_2SO_3 with calcium ion from calcium carbonate or calcium hydroxide.

A most thorough study of the equilibrium chemistry has been performed by the Radian Corporation [9]. In their work, a computer program was developed to calculate partial pressures of SO_2 and CO_2 above aqueous solutions containing Ca^{++} , Mg^{++} , Na^+ , NO_3^- , CO_2 , SO_2 , $\text{SO}_4^{=}$ and Cl^- . Results were then compared to experimental data. In addition, thermodynamic data for the dissociation constants of CaCO_3 and MgSO_3 and the solubility product constant for $\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$ were determined experimentally.

The kinetics of lime-limestone scrubbing are not fully understood. When the system is assumed to be composed of H_2SO_3 formation with subsequent reaction with lime or limestone, then the following are resistances to the overall reaction rate:

- 1) SO_2 diffusion from the bulk gas through the gas film at the liquid surface

- 2) Dissolution of SO_2
- 3) Hydration of SO_2 to H_2SO_3
- 4) Dissociation of HSO_3^- to form $\text{SO}_3^{=}$
- 5) Diffusion of H_2SO_3 from the liquid film to the interior of the droplet
- 6) Hydration of CaO to Ca(OH)_2 (when lime is used)
- 7) Dissolution of Ca(OH)_2 or CaCO_3
- 8) Formation of Ca^{++} by reaction of Ca(OH)_2 or CaCO_3 with H^+
- 9) Formation of CaSO_3 by reaction of Ca^{++} with $\text{SO}_3^{=}$.

The reactions of steps 3, 4, 8 and 9 are rapid [10]. Therefore, depending on the type scrubber, the controlling mechanisms are gas diffusion, liquid diffusion, hydration of CaO (lime) or dissolution rate of CaCO_3 (limestone). In order to determine which of the above is controlling, it is necessary to know if CaCO_3 , CaO , or Ca(OH)_2 absorbent is added in the scrubber.

5. Use of Flyash Pond Water as a Scrubbing Liquor

No evidence of the use of flyash as an absorbent of SO_2 appears in the literature. The only investigations have primarily been to examine the uses of flyash as a construction material.

The areas of flyash study fall into two main categories:

- 1) extraction of valuable substances
- 2) development of uses of ash itself.

A considerable amount of research has been performed on the extraction

of valuable substances [11]. However, the practical importance has not been demonstrated. Success has been achieved in making use of the self-hardening characteristics (stabilization with lime and cement), and the ability to produce a lightweight aggregate.

6. Use of Cooling Tower Blowdown as a Scrubber Liquor for SO₂

No evidence of the use of cooling tower blowdown as a scrubber liquor could be found. In several scrubber systems, the cooling tower blowdown is used for makeup to replace the water lost due to evaporation in the scrubber. Since closed loop operation is desired, a means of utilizing blowdown and ash pond water were considered.

IX. UTILIZATION OF FLYASH AND COOLING TOWER BLOWDOWN

1. Utilization of Flyash as an SO_x Absorbent

The use of flyash as an absorbent has three main advantages:

- 1) available at plant site - no cost incurred for absorbent or transportation to site
- 2) small specific surface - enhances reactivity
- 3) no need for grinding - limestone must be ground when used in scrubbers.

To evaluate the possible species that would be available for reaction with SO₂, a study was made of a typical coal ash.

The analysis of ash varies with the type and location of the coal. A typical ash analysis for several US coals is shown in Table 4.

Silica (SiO₂) usually occurs in the greatest proportion of flyash. Because the flame temperature of a power boiler is in the range of 1500°C ± 200°C, most of the silica present in the coal ash is in the ceramic phase [13]. The only substance that would react with silica is a strong mineral acid which forms orthosilicic acid.

Alumina (Al₂O₃) is also found in abundance in flyash. Because of its relatively unreactive properties, alumina was not considered as a possible absorbent for SO₂.

Iron oxide (Fe₂O₃) is the next most abundant element usually found in flyash. In studying the possible use of iron it was found that research by Newman and Goebel [14] showed that oxides of iron are

Table 4. Ash Analysis of Seven U.S. Coals [12]

Rank:	Low-Volatile Bituminous	High Volatile Bituminous				Sub-bituminous	Lignite
Seam	Pocahontas No. 3	No. 9	Pittsburgh	No. 6			
Location	West Virginia	Ohio	West Virginia	Illinois	Utah	Wyoming	Texas
Ash, dry basis, %	12.3	14.10	10.87	17.36	6.6	6.6	12.8
Sulfur, dry basis, %	0.7	3.30	3.53	4.17	0.5	1.0	1.1
Analysis of ash, % by wt.							
SiO ₂	60.0	47.27	37.64	47.52	48.0	24.0	41.8
Al ₂ O ₃	30.0	22.96	20.11	17.87	11.5	20.0	13.6
TiO ₂	1.6	1.00	0.81	0.78	0.6	0.7	1.5
Fe ₂ O ₃	4.0	22.81	29.28	20.13	7.0	11.0	6.6
CaO	0.6	1.30	4.25	5.75	25.0	26.0	17.6
MgO	0.6	0.85	1.25	1.02	4.0	4.0	2.5
Na ₂ O	0.5	0.28	0.80	0.36	1.2	0.2	0.6
K ₂ O	1.5	1.97	1.60	1.77	0.2	0.5	0.1
Total	98.8	98.44	95.74	95.20	97.5	86.4	84.3
Ash fusibility							
Initial deformation temperature, F							
Reducing	2900+	2030	2030	2000	2060	1990	1975
Oxidizing	2900+	2420	2265	2300	2120	2190	2070
Softening temperature, F							
Reducing		2450	2175	2160		2180	2130
Oxidizing		2605	2385	2430		2220	2190
Hemispherical temperature, F							
Reducing		2480	2225	2180	2140	2250	2150
Oxidizing		2620	2450	2450	2220	2240	2210
Fluid temperature, F							
Reducing		2620	2370	2320	2250	2290	2240
Oxidizing		2670	2540	2610	2460	2300	2290

catalyst for the oxidation of gaseous sulfur dioxide. Oxidation of SO_2 to sulfur trioxide (SO_3) would be most desirable because the SO_3 could be removed in a simple H_2SO_4 spray tower. However, the work of Johnstone [15] found the maximum conversion of SO_2 to SO_3 to have been 1.8% of the SO_2 .

In further studies, the strong catalytic properties of solutions containing iron were noted. Ferrous sulphate is a known catalyst in the oxidation of dissolved sulfur dioxide. In experiments performed by Johnstone [16] 95% of the SO_2 in a gas containing 0.325% SO_2 was removed with only 1 gallon of water per 4000 cubic feet of gas. The liquid film resistance was reduced to such an extent that absorption of oxygen, not SO_2 was controlling.

The process appeared promising until phenolic compounds, resulting from combustion, were found to inhibit the catalytic action. Attempts were made by Johnstone to offset the inhibitor but little success was obtained. The remaining constituents of the flyash are CaO , Na_2O and K_2O . Of these, CaO usually occurs in the greatest proportion as 5 to 25% of the ash. When dissolved in water, these compounds will react with sulfurous acid which is formed by absorption of SO_2 .

The amount of CaO , Na_2O , and K_2O that is soluble in water for a given ash sample is not known [17]. Some of these will appear in the free state while the remainder is in combination with other constituents of the ash (silica, alumina, iron, sulfates, etc.).

The only source of information obtained was of ash pond water. The analysis for a power boiler is shown in Table 5. To estimate the

Table 5. Ash Pond Analysis

Sample	Determined ppm		
Calcium (Ca)	807.2	160.6	443.6
Magnesium (Mg)	5.8	3.6	5.8
Sodium (Na)	23.5	22.0	26.0
Potassium (K)	45.0	49.0	125.0
Nitrate (NO ₃)	4.0	2.0	0.1
pH @ 25 C	12.5	11.7	12.2

sulfur dioxide removal capabilities of the ash pond, the concentration of calcium only was used. For a plant burning 531,000 lbs/hr of coal with a 4% sulfur content, the sulfur dioxide produced would be 42,480 lb SO₂/hr. On a stoichiometric basis, 26,550 lb Ca/hr or 65,780 gpm of ash pond water with a Ca⁺⁺ concentration of 807.2 ppm would be required to scrub the gas. If the sulfur content of the coal was 1%, then 16,450 gpm of ash pond water would be required.

If the pond was saturated with Ca⁺⁺ from the ash, then the concentration would be 1314 ppm. Then, for the same fuel feed rate and 4% sulfur in the coal, the required flow of ash pond water to the scrubber would be 40,400 gpm compared to 65,780 gpm for a concentration of 807.2 ppm.

The use of ash pond water will be feasible if:

- 1) there is enough CaO in the ash to react with SO₂
- 2) the solubility of CaO is such that the amount of CaO required for reaction with SO₂ will be dissolved.

To determine the amount of calcium oxide (lime) required from the ash to reduce the SO₂ emission, the following analysis was made: Coals having 4%, 3%, 2% and 1% sulfur content were considered. The CaO required from the ash was calculated in Section XII and the results are shown in Fig. 9. The results of Fig. 9 are based on the CaO that is available for reaction with SO₂ in the water and not the total CaO content of the ash. The advantage of using ash pond effluent may be realized by noting that a 1000 mega-watt plant burning 531,000 lb/hr of 4%

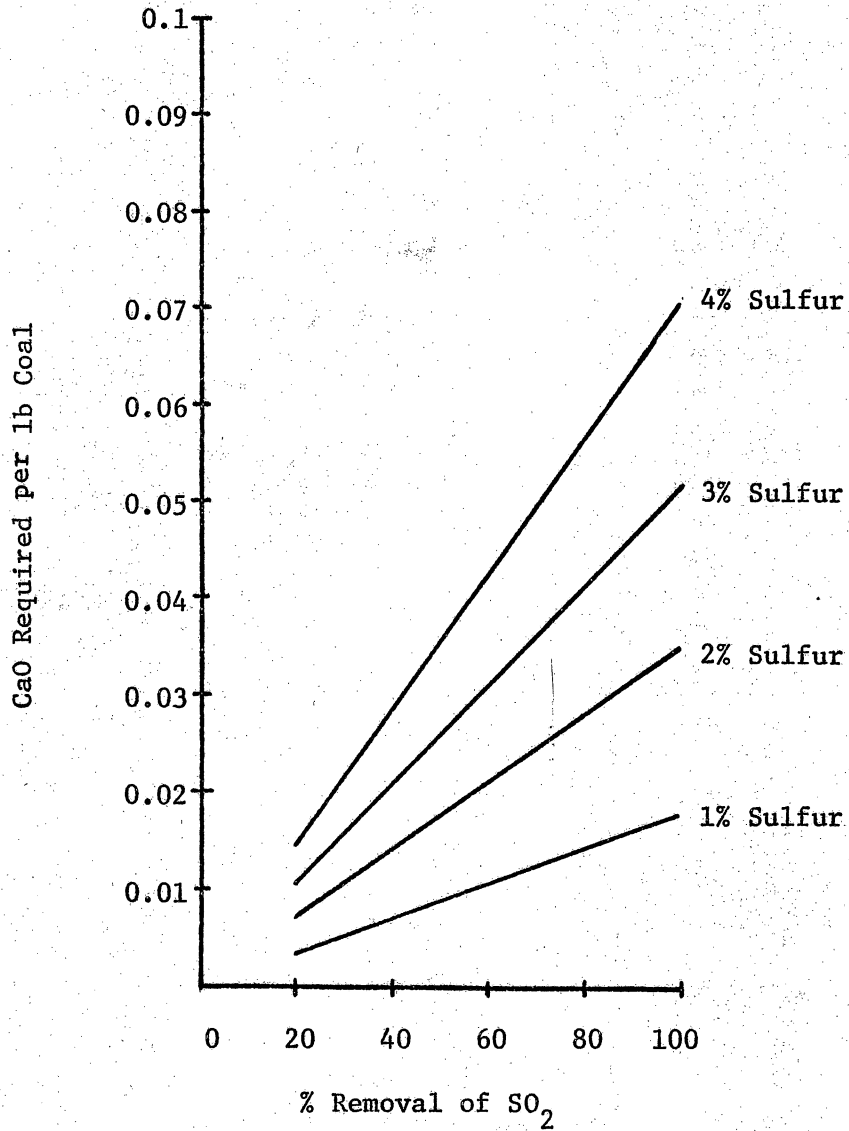


Figure 9. CaO Required from Ash to Produce a Given % Reduction in SO₂

sulfur coal produces 509 tons of SO_2 per day and requires 446 tons of CaO (lime) or 797 tons of limestone in the scrubber.

For some coals, the CaO content is low and the ash is almost totally insoluble in water. One method to produce an ash pond effluent with sufficient CaO to scrub the SO_2 would be to burn two types of coal--one high sulfur coal with low CaO content in the ash and one low sulfur coal with high CaO content in the ash. There would be three main advantages to this method:

- 1) SO_2 from the high sulfur coal would be removed by ash pond water
- 2) flyash from the low sulfur coal that usually could not be removed in an electrostatic precipitator due to its low sulfur content is removed
- 3) utilization of both high and low sulfur coals.

This method would not present any problem to the boiler proper since pulverizers are designed to handle a large variety of coals. However, implementation of the "two coal system" would depend on the economics of burning another coal as opposed to purchasing lime or limestone for the scrubber.

Carbon is not a constituent of coal ash but is present in flyash in amounts that vary with combustion efficiency. Ash from most modern power boilers usually contains less than 3% carbon. Carbon is of interest because of its absorption properties. Experimental work has been performed using activated carbon as an absorbent of SO_2 . In studies

with coal as an absorbent, the absorptive capacity was reported to have been 2.5 to 3.0 g SO₂/1000 g coal compared to 140 g SO₂/1000 g for activated carbon. No mention of the carbon content of the coal was given. The absorptive properties of carbon in the ash may be enhanced by the catalytic action of FeSO₄, MgO, etc. that would be present.

2. Use of Cooling Tower Blowdown in the Scrubber

Cooling tower blowdown utilization is of interest because of the impending EPA Water Pollution Standards. In 1972, the Federal Water Pollution Control Act was passed. After the amendments of 1974, rules and regulations were issued through the Federal Register describing ways for utilities to comply with regulations for waste water discharge. Reference to Table 6 shows the increased need for waste water recycle especially to meet the 1985 zero effluent goal.

In the early stages of this design, it was proposed to use cooling tower blowdown as a scrubber liquor for SO₂. However, in the treatment of cooling water, sulfuric acid (H₂SO₄) is added to control and prevent precipitation of calcium carbonate scale. Calcium carbonate is the most common type scale found in cooling water systems [19]. Because it is in the bicarbonate form, it does not present a problem until breakdown to the carbonate occurs. This can occur with aeration, heat or increase pH. The chemical reaction for the formation of calcium carbonate is:

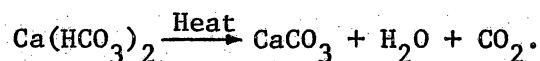


Table 6. Condenser Cooling Water Regulations - 500 Mega-Watt and Larger on Line After 1970 [18]

Effective Date 7/1/77

Discharge Limitations

Free Available Chlorine Limited to 0.5 mg/1 (Daily Max.) and 0.2 mg/1 (Daily Avg.)

No Discharge of Polychlorinated Biphenol Compounds is Permitted

Discharge of Either Free Available Chlorine or Total Residual Chlorine From Any One Unit is Limited to Two Hours Per Day With Not More Than One Unit in Any Plant Discharging at Any One Time.

Monitor Requirements

Daily - 24 Hr. Cont. - Chlorine

Daily - 24 Hr. - Total Flow

Effective Date 7/1/81

Discharge Limitations

No Discharge of Polychlorinated Biphenol Compounds is Permitted

No Discharge of Heat Except in Blowdown From Recirculated Cooling Water System

Pollutants	Maximum Concentration	Average Concentration
Free Available Chlorine	0.5 mg/1	0.2 mg/1
Zinc	1.0 mg/1 (Daily)	1.0 mg/1 (30 Day Avg.)
Chromium	0.2 mg/1 (Daily)	0.2 mg/1 (30 Day Avg.)
Phosphorous	5.0 mg/1 (Daily)	5.0 mg/1 (30 Day Avg.)
Other Corrosion Inhibitors	Limit to be Established on a Case by Case Basis	

Monitor Requirements

Daily - 24 Hr. Comp. - Chlorine

Daily - 24 Hr. Comp. - Phosphorous

Daily - 24 Hr. Comp. - Zinc

Daily - 24 Hr. Comp. - Other Corrosion Inhibitors

Daily - 24 Hr. Comp. - Chromium

Daily - 24 Hr. - Total Blowdown Flow

The addition of sulfuric acid causes formation of the more soluble calcium sulfate while the concentration is controlled by blowdown. Precipitation of calcium sulfate will not occur until a concentration of approximately 2000 ppm is attained. However, the level of concentration is usually kept at about 250 ppm because the concrete cooling tower basin is endangered. Above this level of concentration, voluminous calcium-alumina sulfate are formed which can cause the concrete to crack.

Another method used to slow the rate of scale formation is the use of polyphosphate. Polyphosphates will delay the rate of precipitation until cooling water has passed through the equipment. The use of blowdown containing polyphosphates was not considered.

Since closed loop operation is desired, it was proposed to use the SO_2 and SO_3 available from the flue gas to supply the acid required to form the more soluble calcium sulfate in the cooling water. Also because the flue gas must be cooled before entering the scrubber it was decided that the makeup water to the cooling tower could also be used for this purpose.

The design of the gas cooler absorber is presented in the next section.

X. SYSTEM DESIGN

1. Design Considerations

A 1000 mega-watt unit size was selected as the base case for the study. From the chart of power plant sizes, Fig. 10, it can be seen that this will be a midrange boiler size in the 1980's.

The steam generator was assumed to be a balanced draft pulverized coal-fired type having a capacity of 6.5×10^6 lbm/hr with 8 stages of feedwater heating. A schematic diagram of the plant is shown in Fig. 11. Table 7 is a summary of the design data used in the analysis.

Implementation of the gas scrubber system was assumed to be a retrofit operation. The plant was assumed to have an electrostatic precipitator in operation. Figure 12 is a diagram of the proposed system for SO_x control and reduction of plant waste water discharge.

2. Operation

Flue gas from the induced draft fan is passed through a cooler shown in Fig. 12. Makeup water for the cooling tower is sprayed in the cooler where some of the SO_2 and SO_3 is absorbed and the flue gas temperature reduced. The SO_2 and SO_3 absorbed by the water from the flue gas will react with the calcium bicarbonate in the cooling tower makeup water and form calcium sulfate ($CaSO_4$). If an excess of SO_2 is absorbed, provisions are made for adding lime for pH control. The dissolved solids concentration ($CaSO_3$ and $CaSO_4$) will be controlled

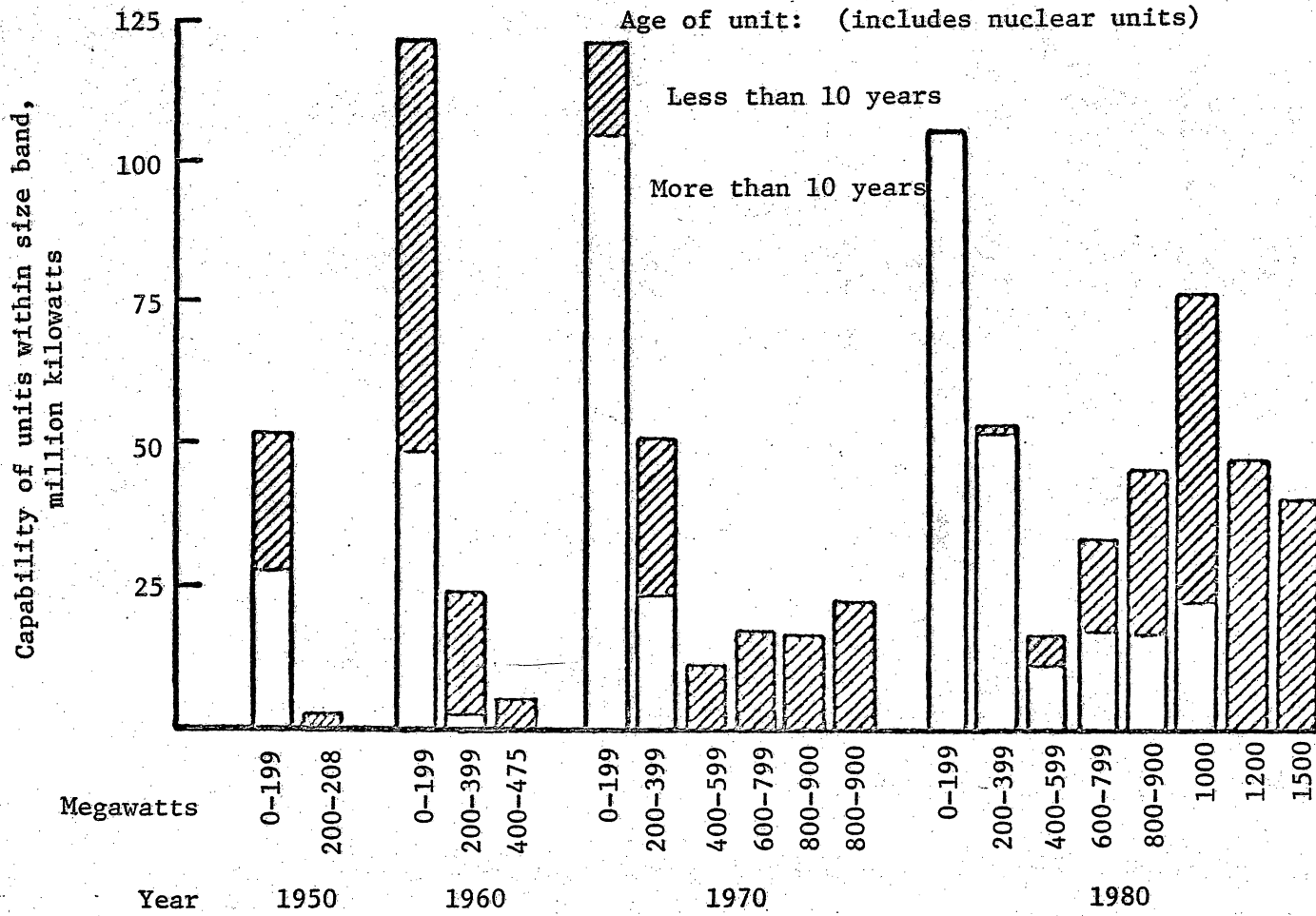


Figure 10. Chart of Power Plant Sizes [20]

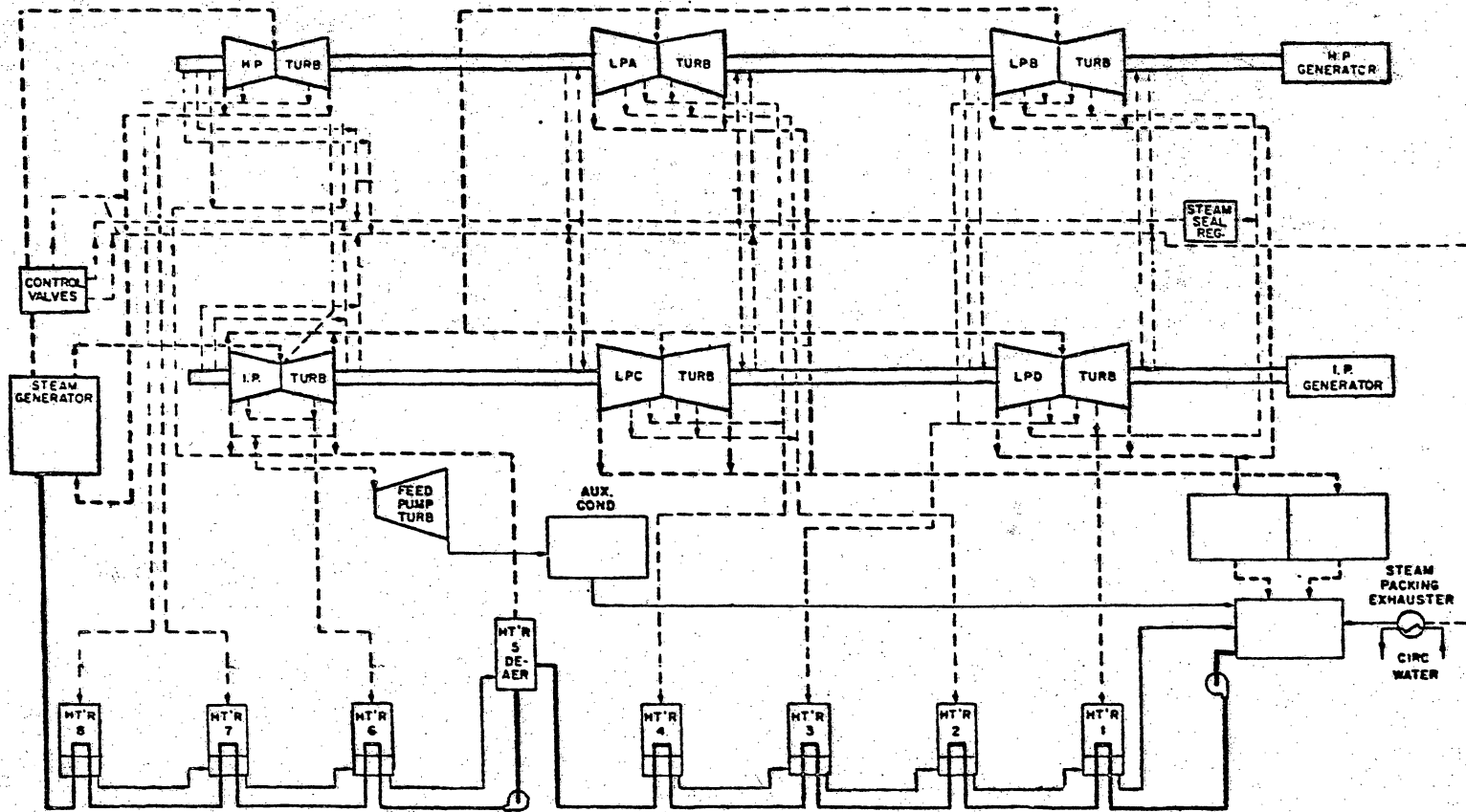


Figure 11. Schematic Diagram of 1000 Mega-Watt Power Boiler

Table 7. Summary of Design Data

Boiler

Steam flow rate	6.5×10^6 lbm/hr
Steam flow through condenser	3.9×10^6 lbm/hr
Flue gas temperature at gas cooler inlet	300F
Flue gas humidity ratio at cooler inlet	$0.059 \frac{\text{lb H}_2\text{O}}{\text{lb dry gas}}$
Flue gas flow rate	5,841,000 lbm/hr
Fuel consumption rate	531,000 lb coal/hr
Sulfur content	1 ↔ 4% S/lb coal
Higher heating value	12,000 B/lbm
Steam temperature at turbine exit	101F
Steam pressure at turbine exit	1 psia
Steam quality at turbine exit	10%
Cooling water temperature at condenser inlet	85F

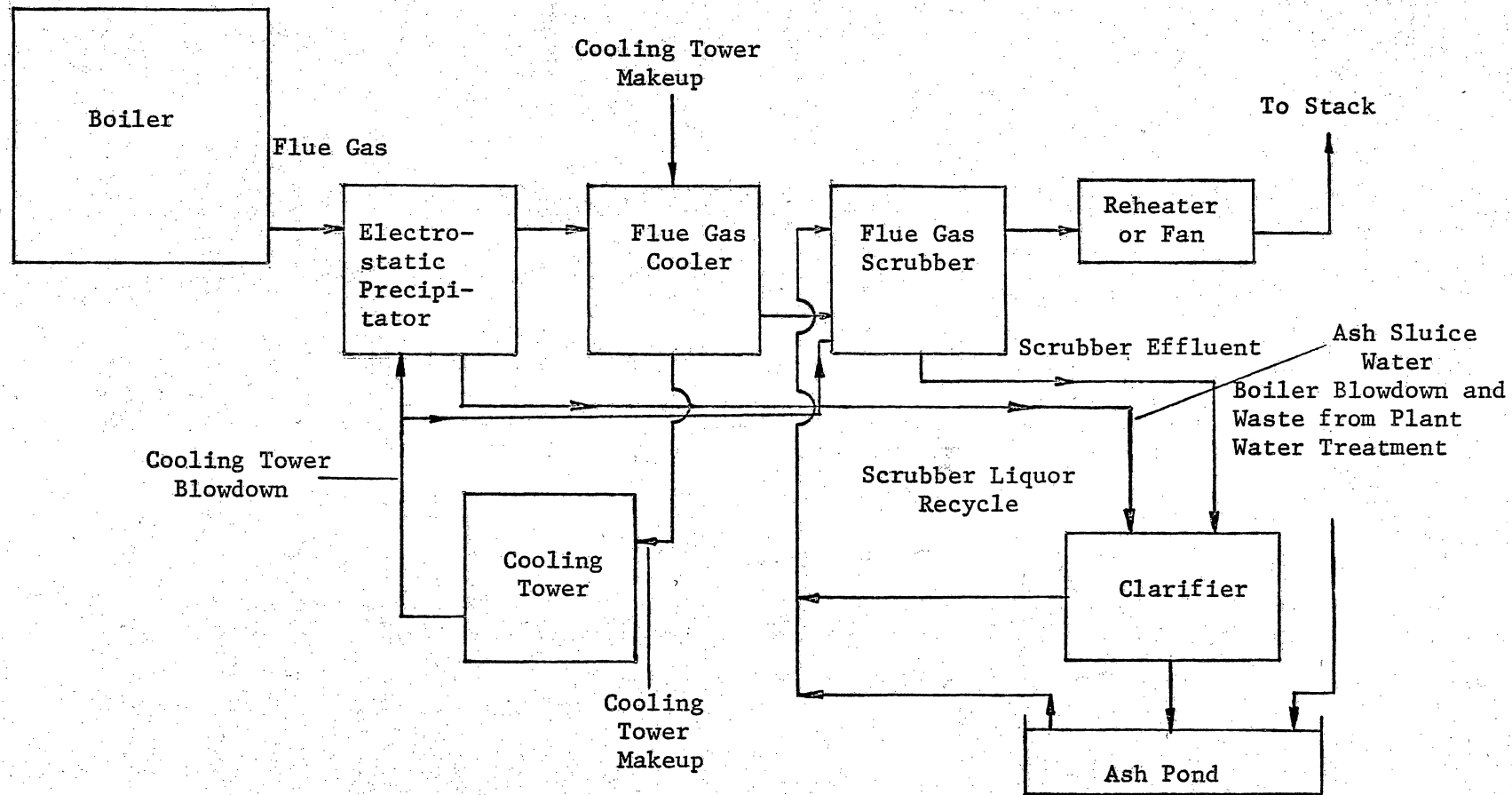


Figure 12. Flowsheet of Proposed System for SO_x Control

by cooling tower blowdown.

The cooled gas then flows through the scrubber. In the scrubber, pressure nozzles are used to spray recycled ash pond water on the scrubber "packing." The objective of the packing is to provide intimate contact between the gas and liquid over a large interphase surface. However, packed beds, which provide high surface area, used in lime-limestone scrubbers do not offer good resistance to scale formation. The scrubber for the proposed system would consist of several "cells" or banks of tubes mounted perpendicular to the direction of gas flow. Due to scaling and plugging tendency in scrubbers, provisions would be made to facilitate removing the tube banks or cells from the scrubber while the scrubber is operating. The pressure drop across each scrubber cell should be monitored. During times of excessive pressure drop warning should be given in order that the cell may be removed for maintenance or additional manual cleaning.

A cell design consisting of removable tube banks was selected because:

- 1) Tubes are a standard construction material.
- 2) Simple fabrication.
- 3) Cells can be rapidly cleaned while the scrubber is in operation. Therefore, plant down time is reduced.
- 4) The particle collection will increase due to the venturi effect produced as the gas accelerates between the tubes.

The proposed removal techniques consist of three steps shown in Figs. 13a, 13b and 13c. In Step 1, high pressure drop signals cell should be removed. The access door is opened and the lift mechanism attached. Step 2 shows the cell removed from the scrubber and the access door closed. In Step 3, the clogged cell is pulled out for cleaning. Seals are located around the top of the cell to prevent excessive leakage of flue gas while the cell is being removed. By using several cell sections, the overall efficiency of the scrubber should not be effected greatly if only one cell has been removed.

As the recycled ash pond water strikes the tubes of the cell, the cell becomes wetted and gas passing over the tube surface is scrubbed. As the gas flows between the tubes, it is accelerated and scrubbing similar to scrubbing in a venturi scrubber should occur. In a venturi scrubber, relative velocity between the gas and the liquid on the walls of the venturi causes the liquid to be atomized. Subsequently, vigorous scrubbing action occurs. In addition, if a staggered tube arrangement is used, the particle removal efficiency will be increased due to impaction on tube surfaces in the tube bundle.

Since counter flow is desired in absorption devices, the recycled ash pond water is sprayed in the top of the scrubber and flows down through the cells as shown in Fig. 14. The ash pond water sprayed at the scrubber cell inlet is used not only for scrubbing but also to keep the cell tube surfaces clean. Blowdown from the cooling tower is sprayed on the walls of the scrubber inlet to help keep surfaces clean and to supply water lost due to evaporation in the scrubber. The

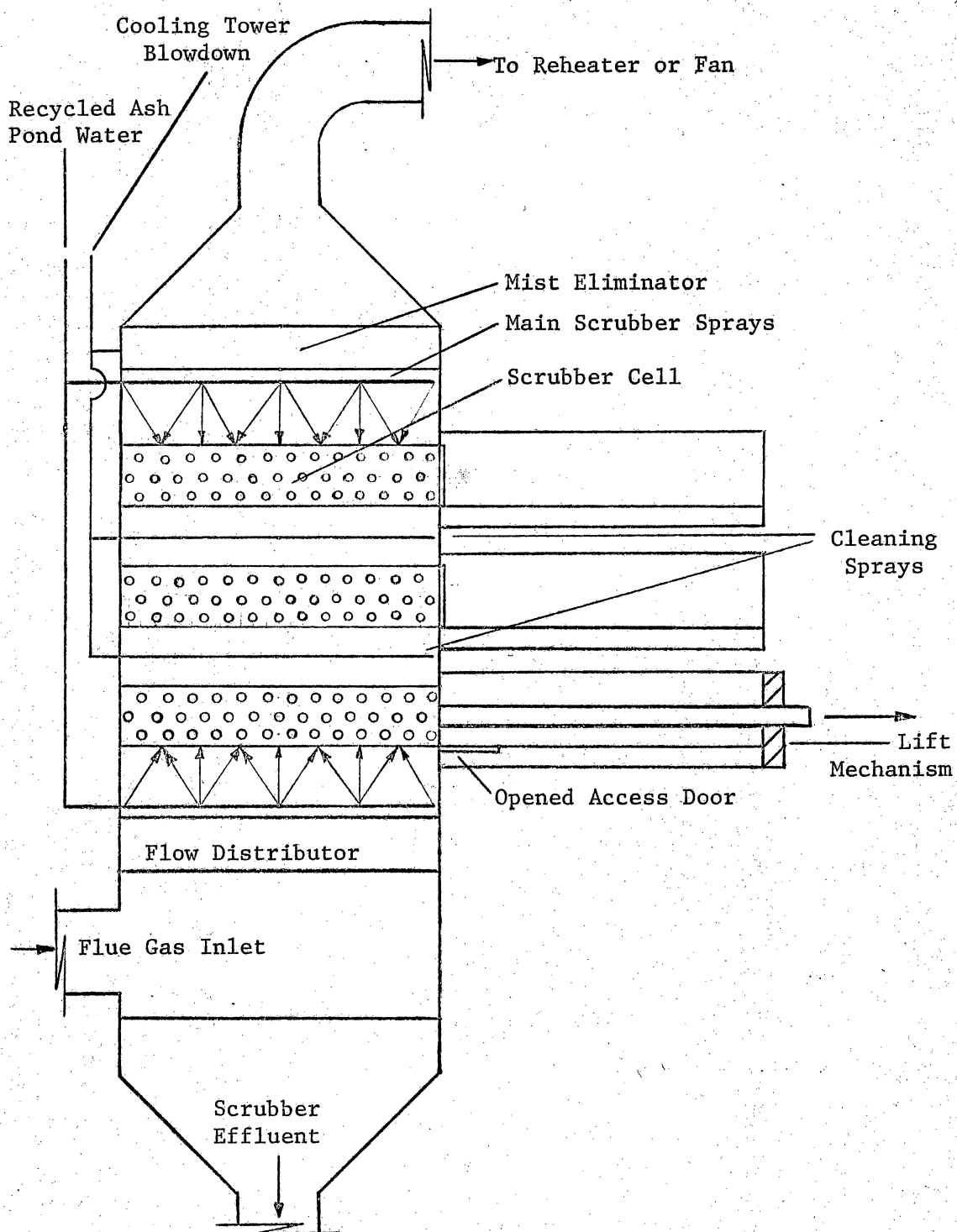


Figure 13a. Scrubber Cell Removal - Step 1

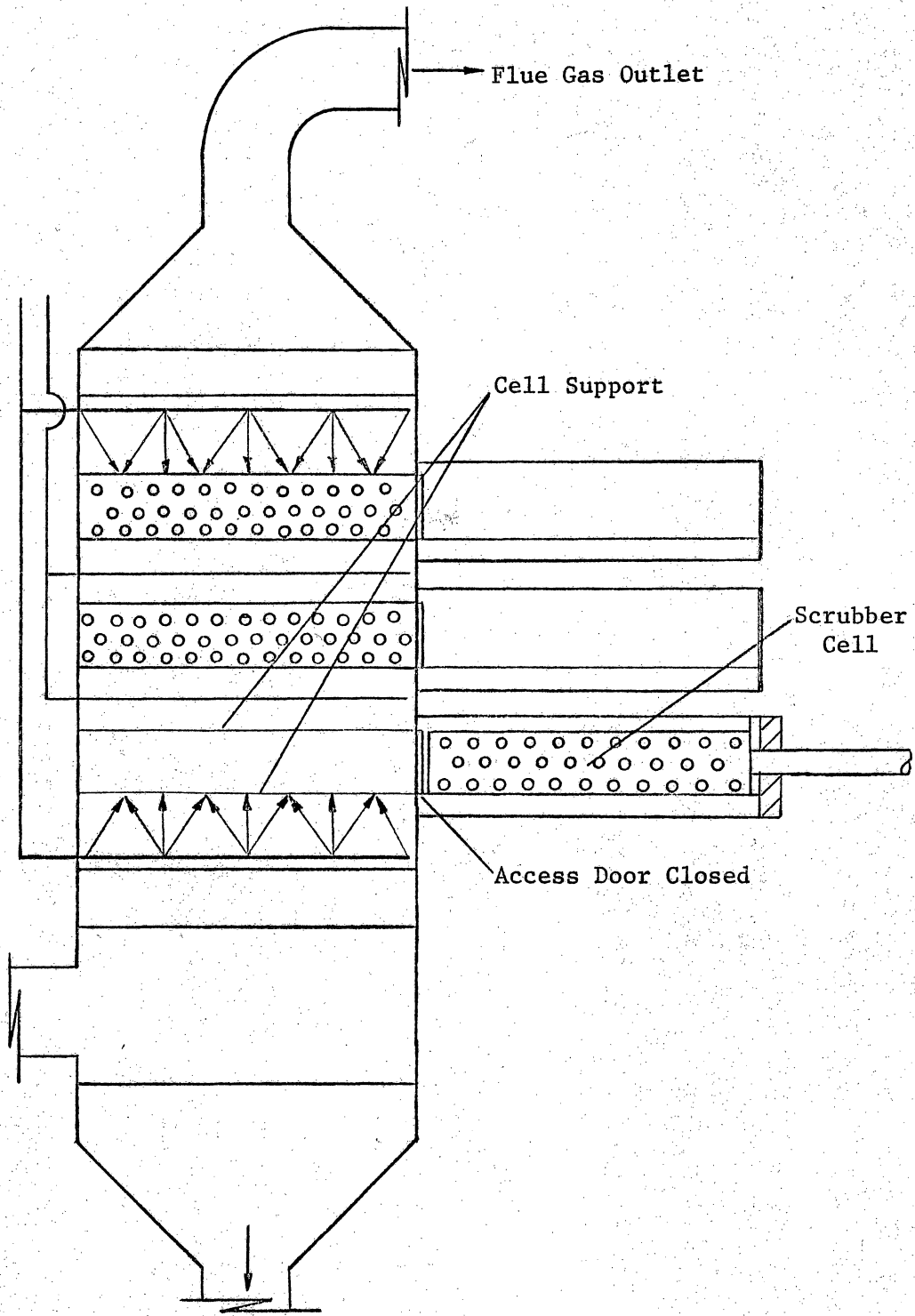


Figure 13b. Scrubber Cell Removal - Step 2

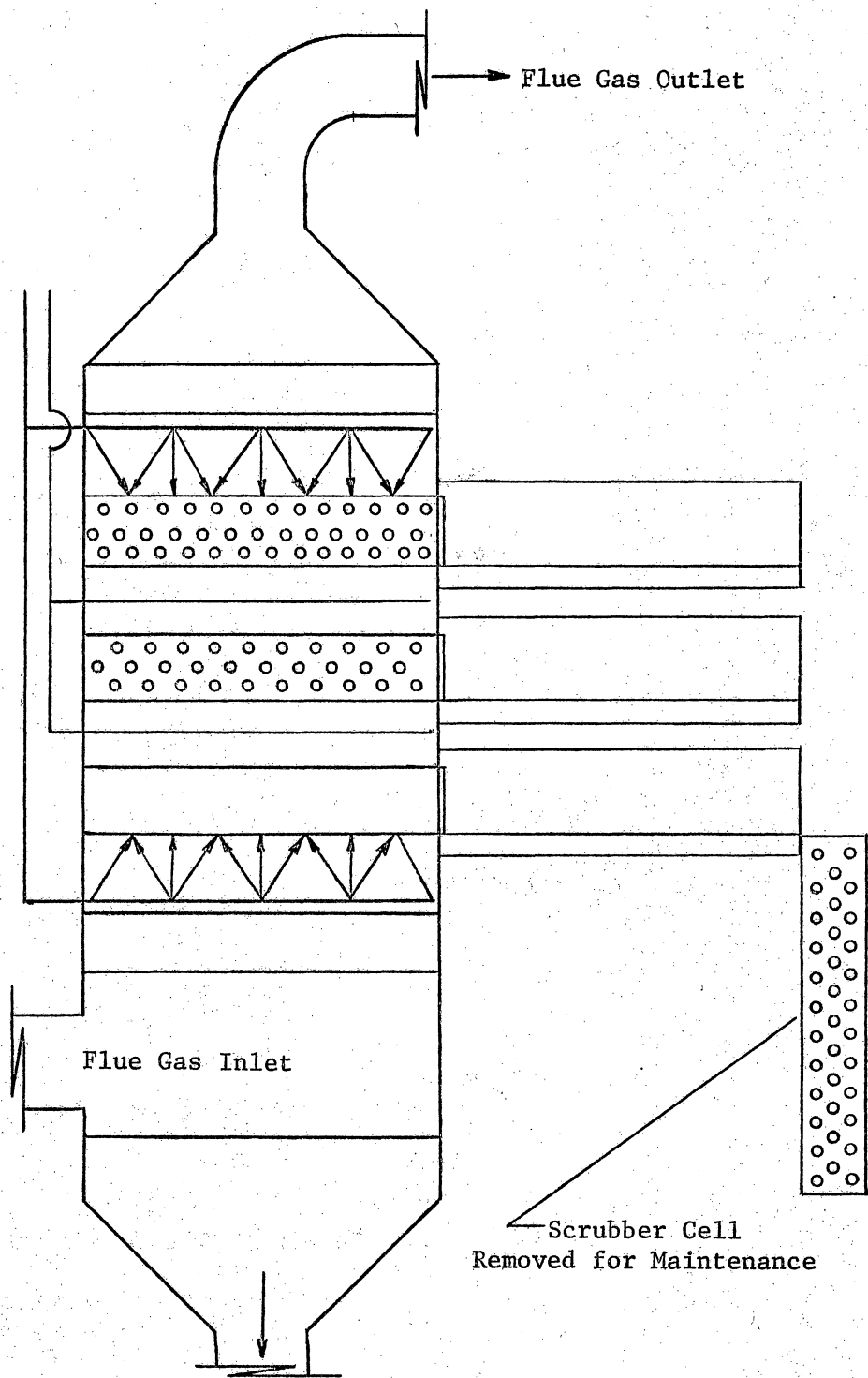


Figure 13c. Scrubber Cell Removal - Step 3

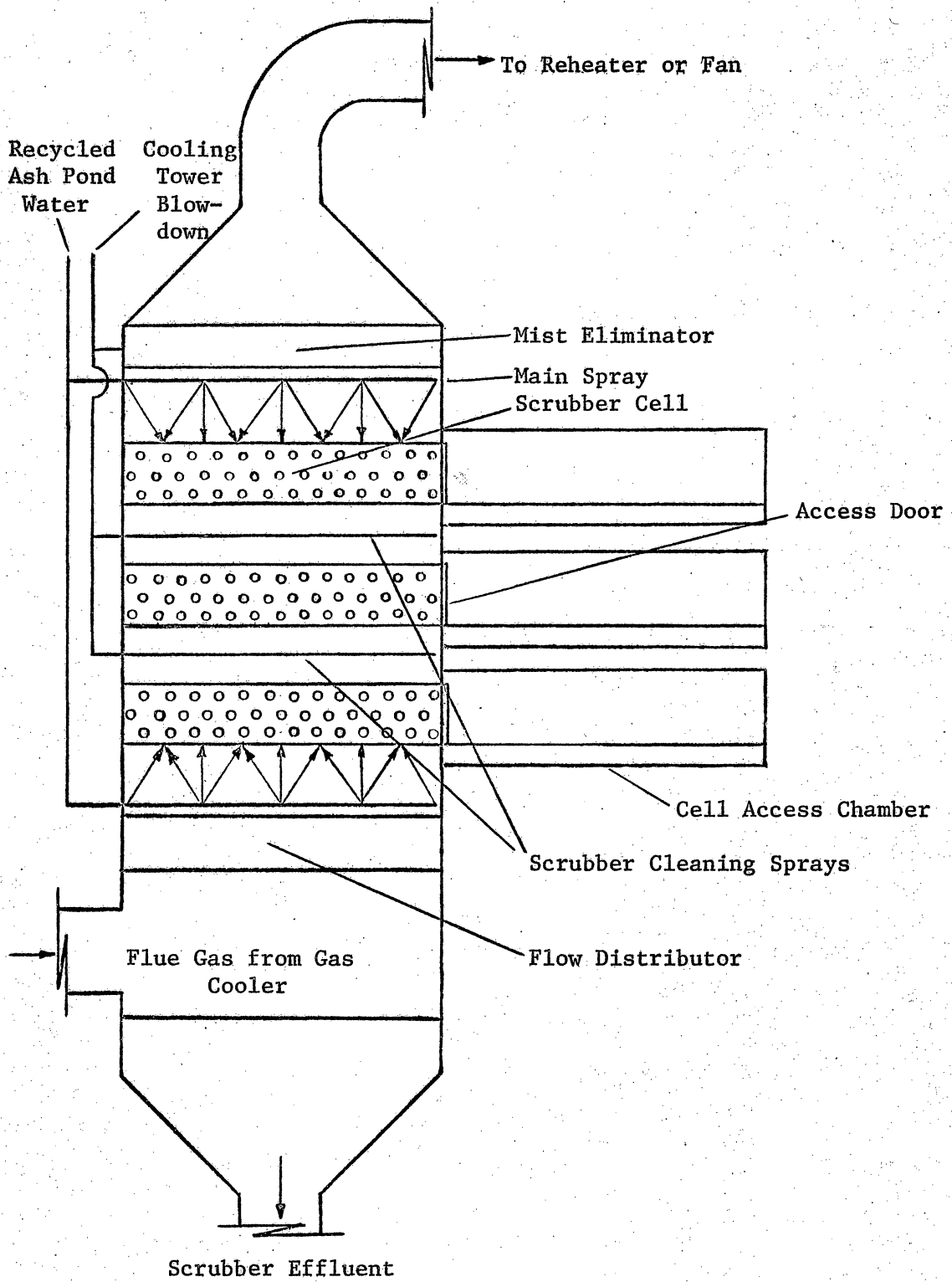


Figure 14. Proposed Flue Gas Scrubber Design

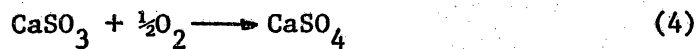
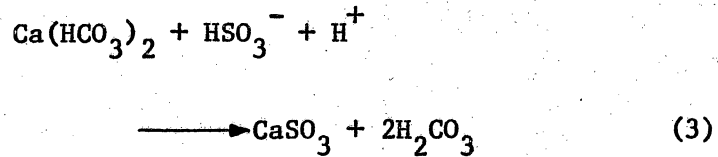
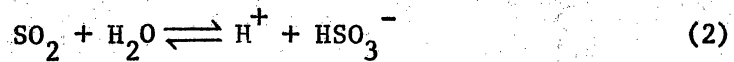
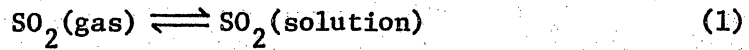
mist eliminator is kept clean also by washing with cooling tower blow-down. Guide vanes are used at the scrubber inlet to assure proper distribution of flue gas over the scrubber cross-section. A monitor is placed after the last cell section to signal control of the amount of ash pond water recirculation which will depend on the concentration of SO_2 allowed to leave the scrubber.

The effluent from the scrubber is pumped to a clarifier and the overflow returned for reuse in the scrubber. Underflow from the clarifier is transported to the waste settling pond where reactions go to completion and precipitation of calcium sulfite (CaSO_3) and calcium sulfate (CaSO_4) occur. Cooling tower blowdown is used to sluice flyash from the electrostatic precipitator and bottom ash from the boiler. Ash sluice water is then mixed with the scrubber effluent. Water from the flyash-scrubber effluent settling pond is recycled to the scrubber.

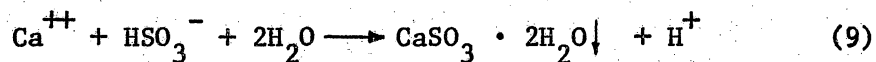
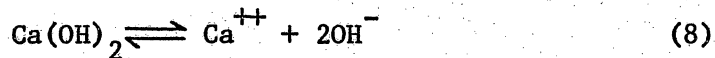
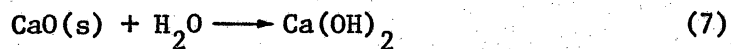
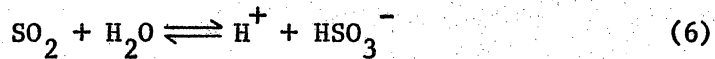
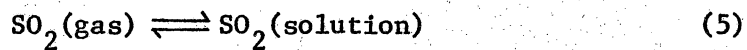
Boiler blowdown and water treatment waste are received in a reaction tank where lime (CaO) is added for pH control. Effluent from the reaction tank is transported to the flyash-scrubber effluent settling pond.

3. Chemistry of System

The basic system circuitry was shown in Fig. 12. The main reactions that would occur in the flue gas cooler if reaction with calcium bicarbonate from the cooling tower makeup water only is considered are:



The reactions that would occur in the scrubber are similar except calcium oxide (CaO) from the ash is supplied rather than the bicarbonate form. The reactions that would occur in the scrubber, assuming reaction with the CaO from the ash only, are:



The only unknown reaction rate would be reaction (7) because the hydration rate of CaO from coal ash could not be found in the literature.

When reusing the ash pond water, the water returned for use in the scrubber would be saturated with dissolved solids. Control of

scaling could be controlled by recycling a portion of the cooling tower blowdown with the scrubber recirculation water which would tend to dilute it. Aeration might be necessary to convert the calcium sulfite (CaSO_3) to calcium sulfate (CaSO_4) which will precipitate much more rapidly. Settling of calcium sulfite and calcium sulfate will be aided since flyash promotes coagulation and settling [21].

4. Design of Direct-Contact Cooler

4.1 Calculation of Cooling Tower Losses

To determine the amount of makeup water for the cooling tower and therefore the flow to the gas cooler, the evaporation loss from the tower was calculated. For a known plant heat rejection load Q_p , a heat balanced of the tower yields

$$Q_p = \dot{m}_{L1} h_{L1} - (\dot{m}_{L1} - \dot{m}_{mue}) h_{L2} \quad (10)$$

The energy transferred to the air is

$$Q_t = \dot{m}_a (h_{a2} - h_{a1}) \quad (11)$$

and the evaporation is

$$\dot{m}_{mue} = \dot{m}_a (\omega_2 - \omega_1). \quad (12)$$

The above equations were combined and two equations were obtained in terms of the unknown makeup water flow:

$$\dot{m}_{mue} = \dot{m}_a (\omega_2 - \omega_1) \quad (13)$$

and

$$\dot{m}_{mue} = \frac{Q_p}{\left(\frac{h_{a2} - h_{a1}}{\omega_2 - \omega_1} \right) - h_{mue}} \quad (14)$$

The only unknown of Eq. (13) and (14) are the humidity ratio and enthalpy of air leaving the tower. By assuming that the air leaving the cooling tower is saturated, the Eqs. (13) and (14) could be solved by trial and error to determine the makeup water flow rate. At design conditions the evaporation loss was calculated to have been 2.5% of the circulating water flow. A summary of design conditions for the cooling tower is given in Table 8.

The total makeup requirement was then calculated with the following addition losses considered:

- (a) drift loss was 0.1 percent of circulating water flow
- (b) blowdown was based on 10 levels of concentration.

Therefore, the total makeup requirement for the tower was calculated to be 7272 gpm at 75F. The next step in the analysis was to determine the losses in the flue gas cooler, SO₂ absorbed and an estimation of the size.

4.2 Direct-Contact Gas Cooler

The principle objective of the spray chamber is to reduce flue gas temperature before the gas enters the scrubber. Reducing the stack gas temperature is necessary because if the flue gas temperature

Table 8. Summary of Cooling Tower Data

Cooling tower: Mechanical draft type

Liquid to gas ratio	$1.6 \frac{\text{lb H}_2\text{O}}{\text{lb air}}$
Circulating water temperature rise through condenser	25F
Condenser terminal difference	10F
Inlet air temperature	80F
Inlet air humidity ratio	$0.013 \frac{\text{lb H}_2\text{O}}{\text{lb dry air}}$
Makeup water temperature from flue gas cooler	120F
Atmospheric pressure	14.7 psia

is much greater than the wet bulb temperature of the gas, the temperature of the liquid film will be raised and the solubility of SO_2 will be decreased.

4.3 Basic Equations for Simultaneous Heat and Mass Transfer

For the direct-contact spray chamber, Fig. 15, the basic equations for parallel flow as derived in reference [22] were:

Mass Transfer

$$-dG_L = G_{FG} d\omega = K_G a_M (\omega_i - \omega) d\ell \quad (15)$$

Heat Transfer

$$G_{FG} c_{pFG} dT_{FG} = h_{FG} a_H (T_i - T_{FG}) d\ell. \quad (16)$$

Total Energy Transfer to Flue Gas

$$G_{FG} dh = K_G a_M (h_i - h) d\ell \quad (17)$$

which assumes the interfacial area for heat transfer, a_H , is equal to the interfacial area for mass transfer, a_M , Lewis number is 1, and variations in h_{fg} are neglected.

Energy Balance

$$G_{FG} dh = -G_L c_L dT_L. \quad (18)$$

Heat transfer to water

$$G_L c_L dT_L = h_L a_H (T_L - T_i) d\ell. \quad (19)$$

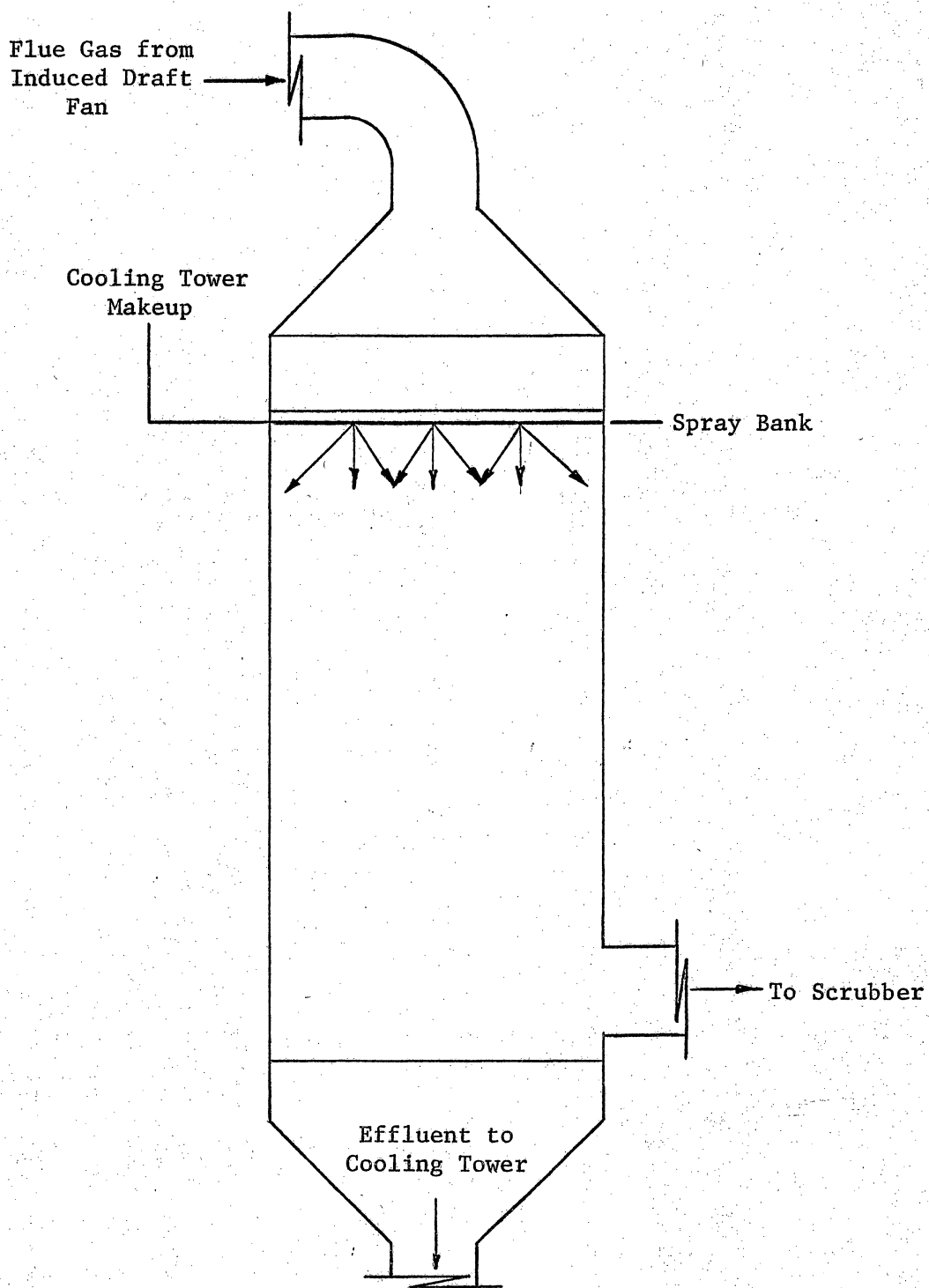


Figure 15. Flue Gas Cooler

Three equations may be extracted from the above equations [22].

The equations are

$$-\frac{h_L a}{K_G a} = \frac{h - h_i}{T_L - T_i} \quad (20)$$

and

$$\frac{dh}{dT_{FG}} = \frac{h - h_i}{T_{FG} - T_i} \quad (21)$$

and

$$\frac{d\omega}{dT_{FG}} = \frac{\omega - \omega_i}{T_{FG} - T_i} \quad (22)$$

and from Eq. (18)

$$\frac{dh}{dT_L} = -G_L/G_{FG} \quad (23)$$

where $h_L a$ = liquid volumetric heat transfer coefficient, B/hr F ft³
 $K_D a$ = flue gas volumetric mass transfer coefficient lbm/hr ft³
 h = flue gas enthalpy in differential length $d\ell$, B/lbm
 h_i = enthalpy at the liquid-gas interface, B/lbm
 T_{FG} = flue gas temperature in differential length $d\ell$,
 T_i = temperature at liquid-gas interface
 ω = humidity ratio, $\frac{\text{lb H}_2\text{O}}{\text{lb dry gas}}$
 ω_i = humidity ratio at liquid-gas interface, $\frac{\text{lb H}_2\text{O}}{\text{lb dry gas}}$

To facilitate the use of Eqs. (20) through (23), the equations were solved on a digital computer. For inlet conditions and knowing the

exit liquid temperature from the chamber, the exit gas temperature and humidity, can be calculated. Also using the equation

$$L = \frac{G_{FG}}{K_D a} \int_{\text{inlet}}^{\text{outlet}} \frac{dh}{(h_i - h)} \quad (24)$$

where L = length of spray chamber, ft

G_{FG} = flow of flue gas per unit area, $\text{lbm}/\text{ft}^2\text{hr}$

K_D = flue gas volumetric mass transfer coefficient,
 $\text{lbm}/\text{ft}^3\text{hr}$

h_i = enthalpy of the liquid-gas interface, B/lbm

h = enthalpy of flue gas, B/lbm

the length of the spray chamber could be determined. The major difficulty is the determination of the gas volumetric mass and heat transfer coefficient and the liquid volumetric heat transfer coefficient. Equations for the gas side coefficient do exist in the literature [23]. However, data on the liquid side coefficient is sparse [23]. For estimation, a typical value for $h_L a$ and $K_D a$ were used based on values given in reference [24] for a spray nozzle operating in air.

The cooling tower makeup was 7272 gpm at 75F and the flue gas mass flow was 5,841,000 lbm/hr . The inlet gas to the cooler was at 300F with a dewpoint temperature of 110F. Using these design conditions and the heat and mass transfer coefficients from reference [24] the exit gas temperature was calculated to have been 134F and

humidity ratio was 0.073 $\frac{1 \text{ lb H}_2\text{O}}{1 \text{ lb dry gas}}$. The water evaporated in the cooler was

$$\begin{aligned} \dot{m}_{FG}(\omega_2 - \omega_1) &= 5,841,000 (0.073 - 0.0594) \\ &= 79,400 \text{ lbm/hr.} \end{aligned}$$

The chamber length was calculated to have been 12 ft.

At the cooler exit the saturation temperature is 116.5F. Therefore, by using the makeup water from the cooling tower, the flue gas was lowered to within 17.5F of the saturation temperature of the gas. This is desirable because the lower exit temperature should increase scrubber efficiency as shown in Fig. 16 [25].

4.4 Absorption of SO₂ in the Flue Gas Cooler

The problem in calculating SO₂ absorption in the flue gas cooler is that no data on SO₂ absorption could be found in the literature for absorption of SO₂ with heat transfer in a parallel flow spray chamber. To estimate the maximum absorption the following analysis was made: Assuming the liquid temperature at the spray chamber inlet is constant and the tower height is infinite, then the SO₂ absorbed was calculated from

$$C_{\text{SO}_2} = \dot{m}_L v_{\text{SO}_2} \quad (25)$$

where C_{SO_2} = amount of SO₂ absorbed, lb/hr

\dot{m}_L = liquid mass flow rate, lb/hr

v_{SO_2} = weight fraction of solute in solvent, lb SO₂/100 lb H₂O.

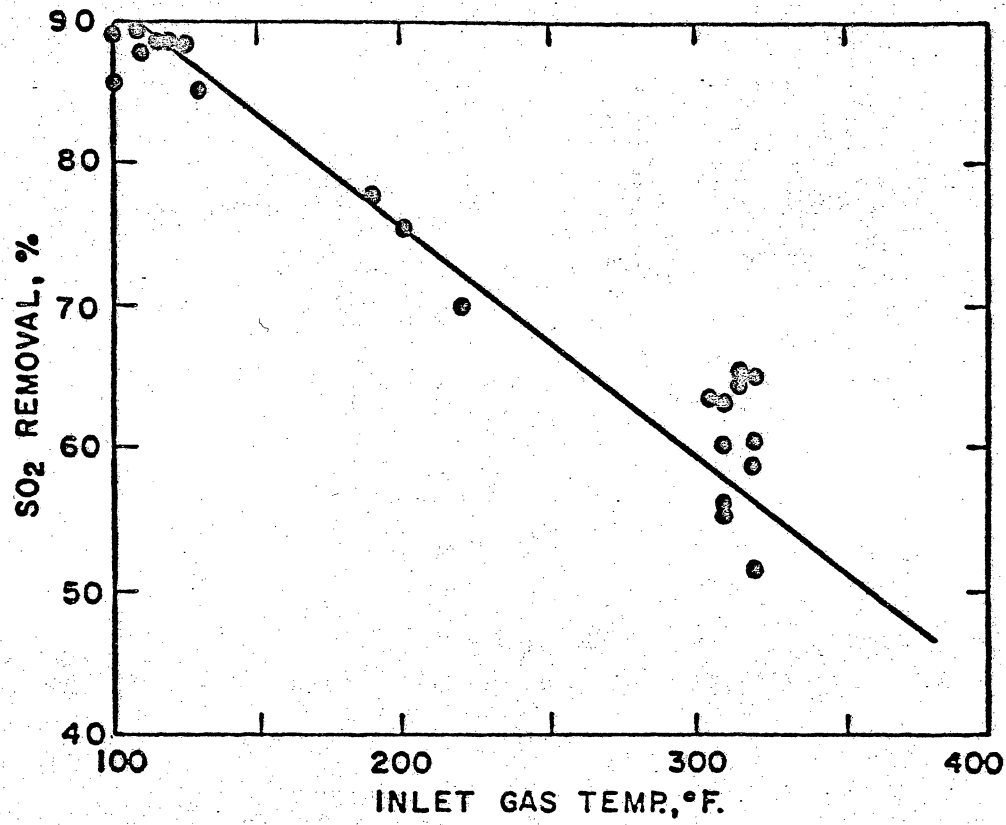


Figure 16. Scrubber Efficiency Variation with Inlet Flue Gas Temperature [25]

At design conditions, the SO_2 absorbed as a function of sulfur content of the coal was calculated. The results are shown in Table 9. From Table 9 it may be seen that the maximum SO_2 absorbed was for the highest SO_2 concentration.

This estimation of maximum absorption was conservative because:

- 1) the liquid temperature increases through the gas cooler and the solubility of SO_2 decreases, and
- 2) the tower has finite length.

By using a parallel flow cooler, heat transfer is accomplished but the mass transfer is limited. If the cooler did absorb the amounts shown in Table 9, this would cause scaling problems.

If 2574 lb SO_2 /hr were absorbed and lime (CaO) was added for pH control, then 5470 lb CaSO_4 /hr would be produced. If the makeup flow rate through the tower was 3.63×10^6 lbm/hr, then the concentration of CaSO_4 would be 1510 ppm. The solubility of CaSO_4 is approximately 2200 ppm at 70F and decreases as temperature increases therefore, after less than 2 levels of concentration in the cooling tower precipitation of CaSO_4 would occur. Therefore, provisions must be made for adding sodium hydroxide (NaOH) or soda ash (Na_2CO_3) during periods of high SO_2 absorption. The solubility of Na_2SO_3 and Na_2SO_4 are 139,000 ppm at 32F and 194,000 ppm at 68F respectively. Also both solubilities increase with temperature.

Summarizing, the use of cooling tower makeup water appears to be a feasible method for:

Table 9. Maximum SO₂ Absorbed in Flue Gas Cooler*

Sulfur content of coal	SO ₂ produced	SO ₂ absorbed
	<u>1b SO₂</u> hr	<u>1b SO₂</u> hr
4%	42,480	2574
3%	31,860	2132
2%	21,240	1570
1%	10,620	817

* Plant burning 531,000 lb coal/hr

- 1) reducing flue gas inlet temperature to the scrubber and possibly increasing scrubber efficiency, and
- 2) eliminating the need for sulfuric acid addition to the cooling tower makeup water.

The use of sulfur dioxide for control may cause changes in makeup water chemistry. If the SO_2 is absorbed in amounts greater than that required to react with the calcium bicarbonate, then lime (CaO) or soda ash (Na_2CO_3) must be added depending on the level of calcium sulfate (CaSO_4) concentration that is allowed in the cooling tower water circuit.

5. Zero Blowdown System

The proposed system of Fig. 12 could operate on a zero blowdown basis if

- 1) the evaporation from the ash pond was sufficient to offset cooling tower blowdown
- 2) the settled solids of the ash pond-scrubber effluent would constitute a water bleed from the system if these wastes were being transported away from the plant.

During times of adverse weather conditions, (rain, etc.) provision must be made for preventing overflow of the concentrated ash pond into the adjacent water course and adding an SO_x absorbent (lime, soda ash, etc.) when the pond becomes diluted.

When insufficient pond evaporation was occurring, some type of chemical or physical water treatment must be performed and the treated water recycled if zero blowdown is to be obtained. A diagram for zero blowdown operation is shown in Fig. 18.

6. Comparison of Flue Gas Reheating to Fan Power Energy Cost

One of the major problems of wet scrubbing processes is that the flue gas is cooled which results in loss of stack gas buoyancy. Under periods of low turbulence, the cooled plume will return to the ground without sufficient dilution and, thereby, increase the ground level concentration of objectional stack gas constituents. Currently, reheating of the gas is considered necessary to reduce possible plume formation and to restore buoyancy to the gas [26].

For purposes of comparison, the use of an indirect flue gas reheater was considered. In this type reheater, steam or hot water is passed through the reheater and the flue gas is reheated.

Applying the First Law to the reheater, the energy required was calculated from

$$Q_{RH} = \dot{m}_{FG} \Delta h_{FG} \quad (26)$$

or

$$Q_{RH} = \dot{m}_{FG} c_{pFG} \Delta T_{FG} \quad (27)$$

where \dot{m}_{FG} = flue gas mass flow rate, lbm/hr

c_{pFG} = specific heat of the flue gas, B/lbm F

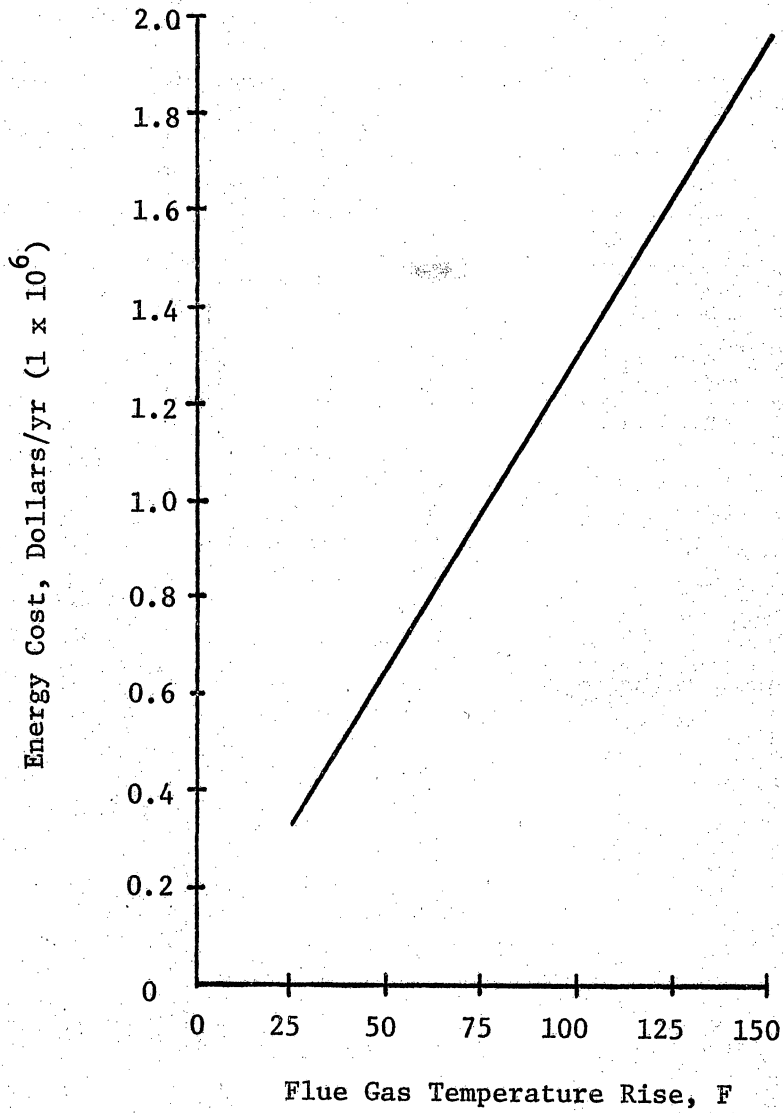


Figure 17. Energy Cost Per Year Versus Flue Gas Temperature Rise

ΔT_{FG} = temperature rise of flue gas across the reheater, F

Q_{RH} = energy required, B/hr.

To calculate the energy required as a function of reheat temperature, the temperature rise across the reheater was varied in 25F increments. The flue gas was assumed to be leaving the scrubber at 100F and saturated. The maximum temperature rise across the reheater was set at 150F since 250F is the temperature at which power boilers equipped with low level economizers are now exhausting flue gas. A higher temperature rise would not be practical.

To estimate the energy cost it was assumed that the reheater was supplied with steam from the steam generator.

For a known temperature rise across the reheater, the energy cost could be calculated from the expression

$$EC = (Q_{RH}/HHV)(PC/2000)(1/SGE) \quad (28)$$

where EC = energy cost, dollars/hr

Q_{RH} = energy supplied to reheater, B/hr

PC = price of coal per ton, dollars

HHV = higher heating value of coal, B/lbm

SGE = efficiency of steam generator, percent/100.

For a steam generator operating 8760 hr/yr with 90% efficiency and burning coal with a higher heating value of 12,000 B/lbm at the present

cost of \$23/ton [27], the energy cost per year as a function reheat temperature was calculated. The results are shown in Fig. 17.

In order to compare the energy cost of a fan to the reheater, the pressure rise required across the fan must be known. For this study it was assumed that the fan would supply the same pressure rise as the draft produced across the stack in addition to the flow losses of the stack.

For stack heights of 400, 600 and 800 feet the draft produced for a flue gas at 250F was calculated from the equation

$$SD = 7.84(L) \left(0.00179 - \frac{1}{T_g} \right) B/30 \quad (29)$$

where SD = stack draft, in. of H₂O

B = atmospheric pressure, in. of H₂O

T_g = average flue gas temperature, R

L = stack height, ft.

The exit velocity of the flue gas from the stack was selected to have been 80 ft/sec. At this velocity, the gas will penetrate the turbulent wake created at the stack exit in high wind conditions [28]. Since this was a retrofit operation, the stack diameter is too large to allow an exit velocity of 80 ft/sec. Therefore, it was proposed that a convergent nozzle could be placed at the stack exit to increase flue gas exit velocity.

For a 60° nozzle the flow loss was calculated to have been 0.1 in. of H₂O from the expression obtained from reference 29

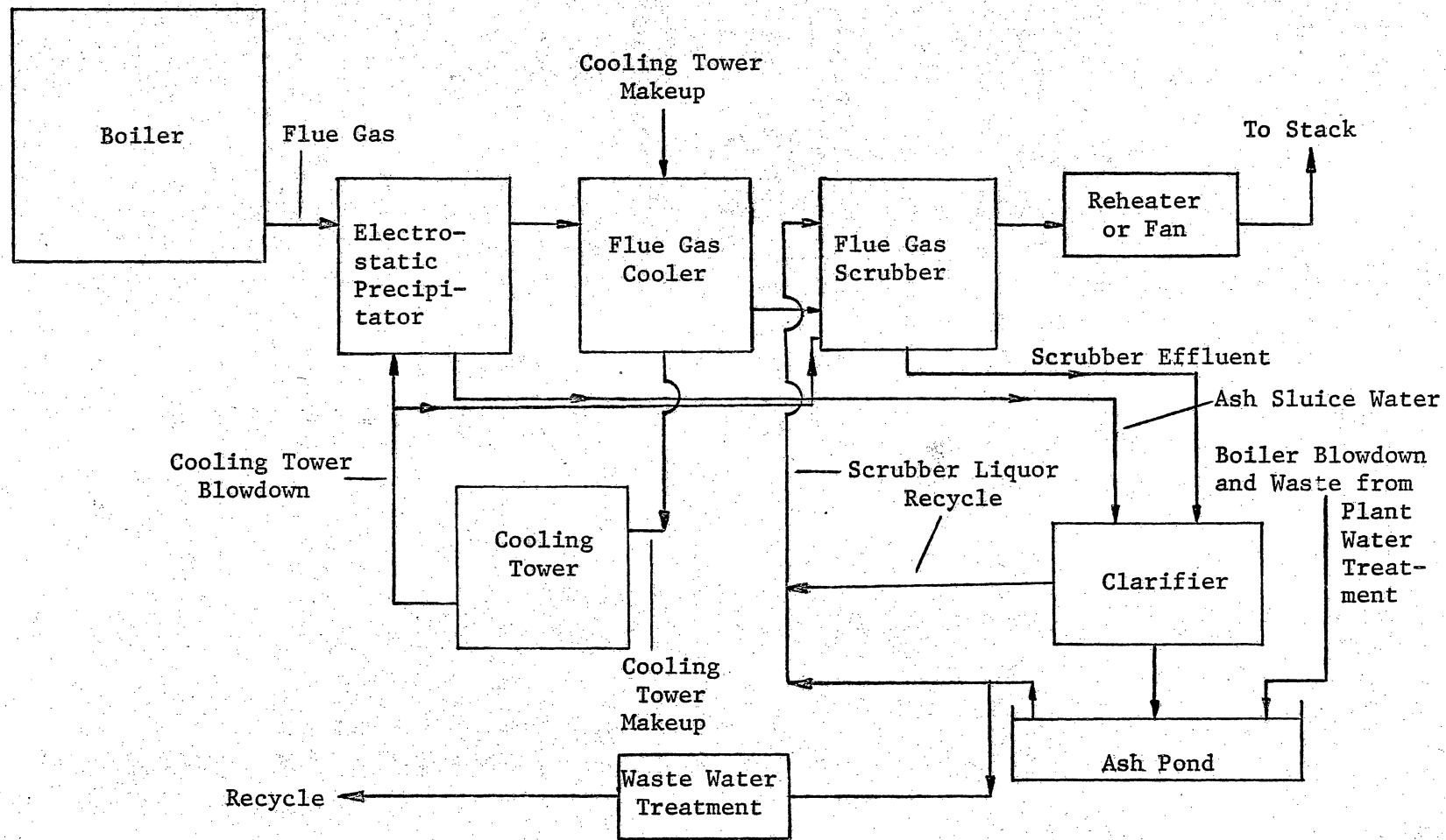


Figure 18. Flowsheet of Proposed System for SO_x Control and Zero Blowdown

$$H_{ge} = (0.07) \left(\frac{V(60)}{4005} \right)^2 \quad (30)$$

where H_{ge} = flow loss, in. of H_2O

V = exit velocity from nozzle, ft/sec.

The pressure rise requirements for the fan were calculated for stack heights of 400, 600 and 800 ft.

The shaft horsepower was then determined assuming a fan efficiency of 60%. At these design conditions the shaft horsepower was calculated from

$$H_{p_s} = (cfm/6350) (\Delta p / .60) \quad (31)$$

where H_{p_s} = shaft horsepower, Hp

cfm = flue gas volume flow rate, ft^3/min

Δp = pressure rise across fan, in. of H_2O .

If the energy conversion efficiency for the plant is 30% and the fan motor losses are neglected, the energy requirements may be calculated. Using the same design conditions that were used for the reheater, the energy requirements per year were calculated as shown in Section X. The comparison in energy cost per year for the fan and reheater are shown in Table 10.

Because of the environment that the fan must operate in, it must be kept clean to avoid solids deposition which will cause the fan rotor to become unbalanced. It is recommended that a centrifugal fan using radial blades be used and kept clean by water washing and soot-

blowing. Also, the blades must be covered with baked phenolic or plasma sprayed alumina coating to prevent erosion of the fan surfaces. Corrosion in downstream equipment should not be a problem since the current trend recommends using plastic liners in stacks.

XI. DISCUSSION OF RESULTS

The use of ash pond water as a scrubber liquor will be possible if the amount of Na_2O , CaO , and K_2O available from the ash is sufficient to achieve the desired removal of SO_2 in the scrubber. As shown in Section IX, the ash pond water did contain sufficient CaO to remove the SO_2 . However, ash compositions vary and some types of ash will not be suitable. Therefore, lime (CaO) or some other type of absorbent must be added to the ash pond water. In addition, the pond must be designed such that it will not release water to the adjacent water courses.

The advantages of using ash pond water will be

- 1) no cost incurred for absorbent transportation or preparation and,
- 2) utilization of a usual plant waste.

The proposed scrubber design should be advantageous from an operability standpoint. Since the scrubber may be cleaned while in operation, plant down time would be reduced.

Comparison of energy cost for operating a fan versus a reheater has shown the fan should be less expensive. However the fan blades must be protected to prevent solid deposition and erosion. It is proposed that this may be accomplished by water washing or soot blowing in addition to using a protective coating such as baked phenolic or plasma sprayed alumina. The disadvantage of not reheating is that if the downstream equipment is not protected, corrosion can occur. In addition, the

cooled gas exhausting from the stack may produce an objectional plume which when cooled, would settle on the surrounding areas.

XII. CONCLUSIONS

The possible use of ash pond water and cooling tower blowdown as a scrubber liquor in a sulfur dioxide scrubber were studied. The main conclusions that may be drawn were:

- 1) Blowdown from cooling towers using sulfuric acid for calcium carbonate scale control do not appear to be an acceptable scrubber liquor. However, makeup water to the tower might be used to cool the flue gas before the gas enters the scrubber. This could increase scrubber efficiency and also replace the sulfuric acid required for cooling tower water treatment.
- 2) The use of ash pond water is recommended when the calcium oxide (CaO) available from the ash is sufficient to scrub the gas.
- 3) The use of a fan to disperse the flue gas will require less energy therefore the operating cost will be less compared to a reheater.
- 4) This method should be feasible and could meet EPA air pollution standards if the CaO, Na₂O, K₂O available from the ash is sufficient.

XIII. RECOMMENDATIONS

The operation of the proposed scrubber system appears feasible. However, several areas must still be investigated. Further study is recommended to:

- 1) determine the acid (sulfurous and sulfuric) neutralizing capabilities of several types of coal ash,
- 2) determine possible catalytic reactions, of sulfur dioxide (SO_2) caused by the flyash,
- 3) in order to avoid operational problems, the chemistry of the cycle must be studied,
- 4) measure the mass transfer coefficient for the proposed scrubber and determine optimum tube diameter in cell to yield maximum SO_2 absorption with minimum pressure drop, and
- 5) study the effects of using a high exit velocity at the stack to ensure sufficient dilution of gas without excessive plume formation.

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XV. APPENDIX

Calculation of CaO Required for a Given SO₂ Removal

The mass of sulfur dioxide (SO₂) produced per lb of coal was calculated from:

$$\frac{M_{SO_2}}{MS} (UA) = m_{SO_2}$$

where M_{SO_2} = molecular weight of sulfur dioxide

MS = molecular weight of sulfur

UA = ultimate analysis value for sulfur, lb S/lb coal

m_{SO_2} = lb SO₂ produced/lb coal.

The calcium oxide (CaO) required as a function of percent sulfur dioxide removed

$$\frac{M_{CaO}}{M_{SO_2}} \frac{(R')}{100} m_{SO_2} = m_{CaO}$$

where

M_{CaO} = molecular weight of calcium oxide

M_{SO_2} = molecular weight of sulfur dioxide

R' = percentage of sulfur dioxide removed

m_{CaO} = lb CaO required from ash/lb coal.

A sample calculation is shown below for a coal containing 4% sulfur. The sulfur dioxide (SO₂) produced per lb of coal is

$$\frac{64}{32} (0.04) = 0.08 \text{ lb SO}_2/\text{lb coal}$$

and if 80% of SO_2 is to be removed, then

$$\left(\frac{56}{64}\right) \left(.08\right) \left(\frac{80}{100}\right) = 0.07 \text{ lb CaO/lb coal}$$

is required.

Derivation of Design Data

The amount of fuel required was calculated from the design information of Table 7. At the outlet from the steam generator

$$T = 1000\text{F}$$

$$P = 2500 \text{ psia}$$

$$\text{and } h = 1458.4 \text{ B/lbm.}$$

At the steam generator inlet

$$T = 550\text{F}$$

$$P = 2550 \text{ psia}$$

$$\text{and } h = 575.6 \text{ B/lbm.}$$

If the required steam flow rate was 6.5×10^6 lb/hr and the steam generator efficiency was 90%, then the energy required was

$$Q = \frac{6.5 \times 10^6 (1458.4 - 575.6)}{0.90}$$

or

$$Q = 6,376 \times 10^6 \text{ B/hr.}$$

Assuming the fuel has a higher heating value of 12,000 B/lbm, then the fuel feed rate was

$$\begin{aligned}\dot{m} &= \frac{6,376 \times 10^6}{12,000} \\ &= 531,000 \text{ lb coal/hr.}\end{aligned}$$

The theoretical air W_{ta} could be calculated using the equation given by Potter [30]

$$W_{ta} = (7.65)(12,000)/10,000$$

$$W_{ta} = 9.18 \text{ lbm Air/lbm Fuel}$$

and with 120% total air

$$W_a(1.2)(9.18) = 11 \frac{\text{lb Air}}{\text{lb Fuel}}$$

The mass flow of flue gas was assumed to be equal to the air supplied to the boiler

$$\dot{m}_{FG} = (11)(531,000) = 5,841,000 \frac{\text{lb Flue gas}}{\text{hr}}$$

Plant Heat Rejection Load

The heat rejection load of the plant was calculated from

$$Q_R = \dot{m}(h_{Ci} - h_{Co})$$

where Q_R = heat rejection load, B/hr

\dot{m} = steam flow rate, lbm/hr

h_{Ci} = enthalpy of steam at condenser inlet, B/lbm

h_{Co} = enthalpy of saturated water at condenser outlet, B/lbm.

The design conditions were:

- 1) Steam enters condenser at 1 psia and 10% moisture
- 2) Water leaves the condenser saturated
- 3) 60% of total steam flow passes through condenser.

Therefore,

$$Q_R = (0.60)(6.5 \times 10^6)(1002.4 - 69.7) = 3.637 \times 10^9 \text{ B/Hr.}$$

Cooling Tower Circulating Water Flow for Tower

For a 25F temperature rise for the cooling water

$$\dot{m}_{cir} = Q_R / (h_{fo} - h_{fi})$$

where

\dot{m}_{cir} = circulating water flow, lbm/hr

Q_R = plant heat rejection load, B/hr

h_{fo} = enthalpy of cooling water at condenser outlet, B/lbm

h_{fi} = enthalpy of cooling water at condenser inlet, B/lbm.

Therefore, for a cooling water inlet temperature of 85F and a 25F temperature rise across the condenser

$$\begin{aligned} \dot{m}_{cir} &= 3.637 \times 10^9 / (77.9 - 53.01) \\ &= 146.1 \times 10^6 \text{ lbm/hr} \end{aligned}$$

or

$$\dot{m}_{cir} = 292,300 \text{ gpm at 60F.}$$

Calculation of Cooling Tower Losses

A heat balance of the cooling tower yields

$$Q_t = \dot{m}_{L1} h_{L1} - \dot{m}_{L2} h_{L2}$$

or

$$Q_t = \dot{m}_a (h_{a2} - h_{a1}) = \dot{m}_{L1} h_{L1} - (\dot{m}_{L1} - \dot{m}_{mue}) h_{L2}$$

and

$$\dot{m}_a (h_{a2} - h_{a1}) = \dot{m}_{L1} (h_{L1} - h_{L2}) + \dot{m}_{mue} h_{L2}$$

The heat rejection by the plant is

$$Q_p = \dot{m}_{L1} h_{L1} - \dot{m}_{L2} h_{L2} - \dot{m}_{mue} h_{mu}$$

and

$$Q_p = \dot{m}_{L1} h_{L1} - (\dot{m}_{L1} - \dot{m}_{mue}) h_{L2} - \dot{m}_{mue} h_{mu}$$

Therefore,

$$Q_p + \dot{m}_{mu} h_{mu} = \dot{m}_{L1} (h_{L1} - h_{L2}) + \dot{m}_{mu} h_{L2}$$

This may be rewritten as:

$$\dot{m}_a (h_{a2} - h_{a1}) = \dot{m}_{mue} h_{mu} + Q_p$$

Dividing the above equation by the evaporation rate

$$\frac{\dot{m}_a (h_{a2} - h_a)}{\dot{m}_a (\omega_2 - \omega_1)} = \frac{\dot{m}_{mue} (h_{mu})}{\dot{m}_a (\omega_2 - \omega_1)} + \frac{Q_p}{\dot{m}_a (\omega_2 - \omega_1)}$$

and noting

$$\dot{m}_a (\omega_2 - \omega_1) = \dot{m}_{mue},$$

then,

$$\dot{m}_{mue} = \frac{Q_p}{\left[\left(\frac{h_{a2} - h_{a1}}{\omega_2 - \omega_1} \right) - h_{mu} \right]}.$$

Solution by trial and error yielded

$$\dot{m}_{mue} = \frac{3.637 \times 10^9}{\left[\left(\frac{77.4 - 33.8}{0.0476 - 0.013} \right) - 87.9 \right]} = 3,130,000 \text{ lbm/hr.}$$

The drift loss was 0.1 percent of the circulating water flow, therefore,

$$\dot{m}_D = 146.1 \times 10^6 \times .001 = 146,200 \text{ lbm/hr.}$$

The blowdown was based on 10 levels of concentration, therefore the blowdown was

$$\dot{m}_{BD} = \left(\frac{1}{10 - 1} \right) 3,130,000 = 348,000 \text{ lbm/hr.}$$

The total makeup was

$$\dot{m}_{mue} + \dot{m}_D + \dot{m}_{BD} = 3,624,000 \text{ lbm/hr}$$

or 7272 gpm at 75F.

Maximum Absorption of SO₂ in Flue Gas Cooler

For a fuel rate of 531,000 lb coal/hr and 4 percent sulfur

$$(531,000) (0.04) \left(\frac{64}{32}\right) = 42,480 \frac{\text{lb SO}_2}{\text{hr}} .$$

The percent SO₂ by volume was estimated from

$$\frac{\dot{m}_{\text{SO}_2}}{\dot{m}_{\text{FG}}} \frac{R_{\text{SO}_2}}{R_{\text{FG}}} (100) = \text{percent SO}_2 \text{ by volume}$$

where \dot{m}_{SO_2} = mass rate of SO₂ flow, $\frac{\text{lbm}}{\text{hr}}$

\dot{m}_{FG} = mass rate of flue gas flow, lbm/hr

R_{SO_2} = gas constant for SO₂, $\frac{\text{ft} - \text{lb}}{\text{lbm} - \text{R}}$

R_{FG} = gas constant for flue gas, $\frac{\text{ft} - \text{lb}}{\text{lbm} - \text{R}}$

then,

$$\left(\frac{42,480}{531,000}\right) \left(\frac{24}{53.3}\right) (100) = 0.327\% \text{ SO}_2 \text{ by volume.}$$

Assuming pressure in flue gas cooler is 14.7 psia, then the partial pressure was

$$(0.327) 760 = 2.48 \text{ mm of Hg.}$$

From Perry [31], the solubility of SO₂ was

$$0.071 \text{ lb SO}_2 / 100 \text{ lb H}_2\text{O}.$$

Therefore, for a makeup water flow rate of 3,625,000 lbm/hr, then

$$3,625,000 \left(\frac{0.071}{100} \right) = 2574 \frac{\text{lb SO}_2}{\text{hr}}$$

neglecting the increase of solubility due to reaction with alkaline elements in the makeup water.

Calculation of Stack Flow Losses and Fan Power

To calculate the fan power required, an estimate for the stack design is necessary to ascertain the flow losses.

Determination of Stack Diameter

Inlet gas temperature = 250F. Assume average gas temperature = 225F. The stack diameter is given by [32]

$$D = \frac{11.7}{12} \sqrt{\frac{G_{FG}(T_{FG})}{(B)V}}$$

where D = stack diameter, ft

G_{FG} = mass velocity, lbm/sec ft²

B = atmospheric pressure, in. of Hg

T_{FG} = average gas temperature, R

V = velocity, ft/sec.

V	D
20	41.9
30	34.9
40	29.7

The stack diameter is determined such that the flow loss will be 5% of the stack draft.

$$\text{Stack Flow Loss} = \left(\frac{2.76}{B}\right) \left(\frac{T_{FG}}{D^4}\right) \left(\frac{\dot{m}_{FG}}{10^5}\right) \left(\frac{fL}{D} + 1\right)$$

where B = atmospheric pressure, in. of Hg

T_{FG} = average gas temperature, R

\dot{m}_{FG} = flue gas mass flow rate, lbm/hr

L = stack height, ft

f = friction factor (estimated from [33])

D = stack diameter, ft.

Stack Diameter, ft	Flow Loss, in. of Hg		
	400	600	800
41.9	0.078	0.082	0.086
34.2	0.18	0.19	0.2
29.7	0.32	0.34	0.37
	400	600	800

Stack Height, ft

Stack Draft

$$\text{Stack Draft} = 7.84(L) \left(0.00179 - \frac{1}{T_{FG}}\right) \frac{B}{30} \text{ from [33]}$$

Stack Height, ft	Stack Draft, in. of H ₂ O
400	1.03
600	1.55
800	2.0

The stack diameter will be 41.9 ft if the flow loss through the stack is 5% of the stack draft.

Calculation of Nozzle Flow Loss

For a flue gas exit velocity of 80 ft/sec

$$H_{ge} = (0.07) \left(\frac{80(60)}{4005} \right)^2 = 0.1 \text{ in. of H}_2\text{O}$$

where the factor (0.07) is from [34]. Then, for stack heights of 400, 600, 800 ft, the flow losses and pressure rise required by the fan are shown below.

	Stack Height, ft		
	400	600	800
Stack Draft	1.03	1.55	2.0
Nozzle Loss	0.1	0.1	0.1
Stack Flow Loss	0.078	0.082	0.086
Unaccounted Loss (30% of above total)*	0.06	0.086	0.11

*Unaccounted losses include loss in breeching, stack entrance, and fan inlet.

Pressure rise required by fan

	Stack Height, ft		
	400	600	800
in. of H ₂ O	1.3	1.82	2.3

Fan Power Calculation

Flue gas conditions at fan inlet

$$T_{FG} = 100F$$

$$P = 14.7 \text{ psia}$$

$$A = 685.6 \text{ ft}^2$$

$$\dot{m}_{FG} = 5,841,000 \text{ lbm/hr}$$

assuming flue gas may be treated as an ideal gas with the physical properties of air

$$\rho = \frac{(14.7)(144)}{(53.3)(560)} = 0.0709 \text{ lbm/ft}^3$$

the flow rate (cfm) at design conditions is

$$\frac{5,841,000}{0.0709} \frac{1}{60} = 1.37 \times 10^6 \text{ cfm.}$$

The air horsepower is given by

$$H_{Pa} = \frac{\text{cfm} (\Delta p)}{6350}$$

and the shaft horsepower is

$$H_{Ps} = \frac{H_{Pa}}{\text{Eff}}$$

where H_{Ps} = shaft horsepower

Eff = fan efficiency.

Assuming the fan has a 60% Eff at design conditions, then the shaft horsepower could be calculated.

	Stack Height, ft		
	400	600	800
H_{Pa}	281	393	496
H_{Ps}	468	654	827

Calculation of Fan Energy Cost

A sample calculation is shown below for calculating the energy cost for operating a fan 8760 hr/yr with a plant energy conversion efficiency of 30% and stack height of 400 ft.

The energy cost is

$$EC = 468 \text{ Hp} \frac{550 \text{ ft-lb}}{\text{Hp sec}} \frac{3600 \text{ sec}}{\text{hr}} \frac{8760 \text{ hr}}{\text{yr}} \left(\frac{1 \text{ lbm}}{12000 \text{ B}} \right) \left(\frac{\text{ton}}{2000 \text{ lbm}} \right) \left(\frac{23 \text{ dollars}}{\text{ton}} \right)$$

$$EC = 33,330 \text{ dollars/yr.}$$

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CONCEPTUAL DESIGN OF AN SO_x
SCRUBBER SYSTEM UTILIZING PLANT WASTE

by

George Todd Wright

(ABSTRACT)

The use of ash pond water and cooling tower blowdown as possible SO_x scrubber liquors has been studied. Information on analysis of ash pond water indicates that this method may be feasible. It was found that ash from the furnace contained salts which could be used to absorb SO_x in the scrubber. This would significantly reduce the cost of lime or limestone addition to the scrubber water as presently proposed. Blowdown from cooling towers using sulfuric acid for calcium carbonate scale control does not appear to be an acceptable scrubber liquor. However, the makeup water may be passed through a flue gas cooler where a portion of the sulfur dioxide (SO₂) and sulfur trioxide (SO₃) are absorbed and the flue gas cooled before entering the scrubber.

A scrubber design has been proposed which allows cleaning a portion of the scrubber packing while the scrubber is operating.

A comparison of reheat versus fan power to assist in flue gas removal via the stack was made. A fan which would disperse the flue gas will be more economical if the maintenance costs of the fan and the saturated plume emitted from the stack at a high velocity are compatible.