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THE COAL FUTURE

APPENDIX G

Flue Gas Desulfurization  
and Low Btu  
Gasification--A Comparison

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Appendix G

CAC Document No. 163

Final Report

The Coal Future: Economic and Technological  
Analysis of Initiatives and Innovations to  
Secure Fuel Supply Independence

by

Michael Rieber  
Center for Advanced Computation  
University of Illinois at Urbana-Champaign

and

Shao Lee Soo  
James Stukel  
College of Engineering  
University of Illinois at Urbana-Champaign

Advisor

Jack Simon, Chief  
Illinois State Geological Survey

Center for Advanced Computation  
University of Illinois at Urbana-Champaign  
Urbana, Illinois 61801

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THE COAL FUTURE

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Flue Gas Desulfurization  
and Low Btu  
Gasification--A Comparison

by

James Stukel



## INTRODUCTION

Since the passing of the 1970 Clean Air Act, there has been a continuing debate regarding the use of high sulfur fuels. The most abundant high sulfur fuel available in the continental United States is coal. In the intervening five years a great deal has been written and said concerning ways to utilize high sulfur coal. During these five years the utilization of stack gas scrubbers for removing  $\text{SO}_2$ , the primary pollutant involved when high sulfur content coal is burned, has been considered the primary option open to potential users of high sulfur coal. Recently, low and medium BTU gasification systems have been proposed as another option for utilization of high sulfur fuels.

In an attempt to deal with the mideast oil embargo of 1974, a plan of action was proposed to make the U. S. self-sufficient in energy production by 1985. The implications of such a proposal on coal utilization are enormous. Even assuming that conservation measures will reduce the U. S. energy demand by as much as 10-15 percent, the production of coal would have to be doubled in the next ten years.<sup>1</sup> This requires the equivalent of opening one new strip and one new deep mine every month for the next ten years. This is an almost impossible task. To further complicate an already difficult situation, the majority of the coal available for immediate mining has a sulfur content which will not be able to be legally burned without adequate sulfur removal capabilities after May 30, 1975. This, then, defines the energy-environmental dilemma.

To gain some perspective regarding the number of power plants involved in such a dilemma consider the following facts. The report of the Hearing Panel on power plant  $\text{SO}_2$  compliance<sup>2</sup> indicated, that in 1972, there were 970 fossil fuel-steam power plants generating 302,000 mw of electricity.

Of these 55 percent (166,000 mw) were coal fired, 17 percent (51,000 mw) were oil fired and 28 percent (85,000 mw) were gas fired. The report estimated that there would be 209,000 mw being generated in 1975 by coal fired plants. Of these plants, 123,000 mw are not expected to need additional emission controls to meet primary air quality standards, 23,000 mw will require washing of high sulfur coal or blending with low sulfur coal and 63,000 mw will need elaborate sulfur control. After 1975, EPA estimates that 24,000 mw will be added each year, with 14,500 mw of these units being coal fired. Further, many units currently fired with oil or gas may be forced to switch to coal because of oil and gas shortages due to supply or to economics. One of the most obvious short term options for dealing with the sulfur problem is to switch all the units to coals having low sulfur contents. The SOCTAP<sup>3</sup> report states that a possible deficit on low sulfur coal in 1975 of as much as 250,000,000 tons may exist. Expressed as steam electric capacity, this is equivalent to 100,000 mw of capacity.

In the discussion which follows, two additional options for dealing with the utilization of high sulfur coal will be considered: stack gas scrubbers and low BTU gasification.

#### FLUE GAS DESULFURIZATION SYSTEMS

Sulfur dioxide removal processes are often described in terms of how the waste products are handled. One group of processes deals with the waste disposal problems by discharging absorbent to a sewer, by impounding or by removing the suspended solids from the slurry and discarding. These systems are called "throw-away" processes. The "throw-away" processes are inherently potential sources of water pollution and solid waste disposal problems. Another group of processes deals with the disposal problem by

regenerating the spent solvent so as to recover the  $\text{SO}_2$  absorber material for reuse in the scrubber and at the same time produce a useful by-product. These systems are called regenerative processes.

When evaluating  $\text{SO}_2$  removal efficiencies, it should be noted that removal efficiencies of the order of 75 percent are needed to meet the New Source Performance Standards with 3 percent sulfur bituminous coal. In general, efficiencies of 85 percent are sufficient to meet most state sulfur dioxide emission regulations.<sup>3</sup>

The recent SOCTAP report<sup>3</sup> stated that the most successful operation of  $\text{SO}_2$  stack gas clean up devices are the Chemico calcium hydroxide scrubber which has operated on a coal fired boiler at the Mitsui aluminum plant in Japan since March 29, 1972, the Babcock and Wilcox limestone scrubbing unit on Commonwealth Edison's Plant near Chicago, the Wellman-Lord regenerable sodium sulfite scrubbing process which has operated on the Japan Synthetic Rubber Chiba Plant since 1971, and Chemico's magnesium oxide system which was started up on the Boston Edison's Mystic Station in 1972.

The report went on to say that at the Mitsui aluminum plant the Chemico scrubbing plant has exhibited a removal efficiency between 80 and 90 percent, the Wellman-Lord Unit at the Japan Synthetic Rubber Plant has operated with a removal efficiency of 90 percent for over 9000 hours of operation, a short term test on the Babcock and Wilcox limestone scrubber at Commonwealth Edison's Will County Plant has exhibited efficiencies between 75 and 80 percent and finally the Chemico wet magnesium oxide scrubber

at Boston Edison's Mystic Station has demonstrated 90 percent SO<sub>2</sub> removal efficiencies.

Table 1 gives a listing of the present status of industrial commitments to stack gas scrubbers as of October 1974.<sup>9</sup> The table lists 99 commitments varying from actual operational units all the way to only considering FGD units at present. The remarkable point that can be made upon examination of the table is the almost total commitment of utilities to lime, limestone or lime/limestone processes. Table 2 shows that 81.4 percent or 29,439 mw of the generating capacity reported to be controlled are lime or limestone. Table 3 shows that 12,945 mw or 35.7 percent of the total are more or less committed to some form of stack gas cleaning scheme. The remainder still have options opened to them. This fact will become important in a later section.

Table 4 lists the distribution of FGD units according to the projected startup date. It is interesting to note that the number of retrofits are about equal to the number of new boilers. The average size of a retrofitted boiler is 243 mw whereas a new boiler averages 504 mw.

Table 5 gives the status of several tail-end gas cleaning systems. As can be seen from the table, the prime candidates for application in the near future are (1) wet (lime) limestone, (2) sodium hydroxide, (3) sodium carbonate, (4) magnesium oxide and (5) catalytic oxidation. A brief description of each of these systems is given in the next section.



Number and Company Name	Unit Name	Location	Start up Date	Status	New (N) Retrofit (R)	Boiler Size, mw	Proc. Cont. Type	Fuel	% S	Vendor
1 Alabama Electric Coop	Bigbee No 2	Jackson AL	1-79	6	N	225	A	Coal	0.8-1.5	NS
2 Alabama Electric Coop	Bigbee No 3	Jackson AL	1-79	6	N	225	A	Coal	0.8-1.5	NS
3 Arizona Electric Power	Apache No 2	Cochise AZ	6-79	4	N	200	A	Coal	0.5-0.8	Research Cottrell
4 Arizona Electric Power	Apache No 3	Cochise AZ	6-79	4	N	205	A	Coal	0.5-0.8	Research Cottrell
5 Arizona Public Service	Cholla No 1	Joseph City AZ	10-73	1	R	115	A	Coal	0.44	Research Cottrell
6 Arizona Public Service	Cholla No 2	Joseph City AZ	6-77	3	N	250	A	Coal	0.44	Research Cottrell
7 Arizona Public Service	Cholla No 3	Joseph City AZ	6-78	5	N	250	A	Coal	0.44	NS
8 Arizona Public Service	Cholla No 4	Joseph City AZ	6-79	6	N	372	A	Coal	-	NS
9 Arizona Public Service	Four Corners No 1	Farmington NM	10-76	6	R	175	B	Coal	0.8	Chemico
10 Arizona Public Service	Four Corners No 2	Farmington NM	12-76	6	R	175	NS	Coal	0.8	Chemico
11 Arizona Public Service	Four Corners No 3	Farmington NM	3-77	6	R	229	B	Coal	0.8	Chemico
12 Arizona Public Service	Four Corners No 4	Farmington NM	4-77	6	R	800	B	Coal	0.8	Chemico
13 Arizona Public Service	Four Corners No 5	Farmington NM	6-77	6	R	800	F	Coal	0.8	NS
14 Basin Electric	Missouri Basin No 1	Wheatland WY	6-79	6	N	550	A	Coal	-	NS
15 Basin Electric	Missouri Basin No 2	Wheatland WY	6-80	6	N	550	A	Coal	-	NS
16 Basin Electric	Missouri Basin No 3	Wheatland WY	6-83	6	N	550	A	Coal	-	NS
17 Boston Edison	Mystic No 6	Everett MA	4-72	1	R	150	C	Oil	2.5	Chemico
18 Central Illinois Light Co	Duck Creek No 1	Canton IL	3-76	2	N	100	A	Coal	2.5-3.0	Riley Stoker
19 Central Illinois Public Ser	Newton No 1	Newton IL	3-77	5	N	600	F	Coal	2.8-3.2	NS
20 Cincinnati Gas & Elec	East Bend No 1	Rabbit Hash KY	6-79	6	N	600	NS	Coal	0.5-0.8	NS
21 Cincinnati Gas & Elec	Miami Fort No 8	North Bend OH	1-77	5	N	500	F	Coal	3.2	NS
22 City of Key West	Key West Power Plant	Key West FL	10-72	1	N	37	A	Oil	2.4	Zurn Air Systems
23 Colorado UTE Elec	Craig Station No 1	Craig CO	6-78	6	N	450	A	Coal	0.45	NS
24 Colorado UTE Elec	Craig Station No 2	Craig CO	6-78	6	N	450	A	Coal	0.45	NS
25 Colorado UTE Elec	Hayden No 1	Hayden CO	6-78	6	R	202	A	Coal	0.45	NS
26 Colorado UTE Elec	Hayden No 2	Hayden CO	6-78	6	R	250	A	Coal	0.45	NS
27 Columbus & Southern Ohio	Conesville No 5	Conesville OH	1-76	3	N	375	B	Coal	-	UOP
28 Columbus & Southern Ohio	Conesville No 6	Conesville OH	1-78	3	N	375	B	Coal	-	UOP
29 Commonwealth Edison	Will County No 1	Romeoville IL	2-72	1	R	167	A	Coal	0.6-3.0	B & W
30 Dairyland Power Coop	Alma Station	Alma WI	6-71	1	R	80	B	Coal	3.0-3.5	Foster Wheeler
31 Detroit Edison	St. Clair No 6	Belle River MI	11-74	2	R	180	A	Coal	3.7-3.8	Peabody Eng.
32 Duquesne Light	Elrama	Elrama PA	5-75	2	R	510	A	Coal	1.0-2.8	Chemico
33 Duquesne Light	Phillips	So Height PA	7-73	1	R	100	A	Coal	2.0	Chemico

1. Operational Units
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- A. Limestone Scrubbing
- B. Lime Scrubbing
- C. Magnesium Oxide Scrubbing
- D. Sodium Carbonate Scrubbing
- E. Wellman-Lord
- F. Lime/Limestone Scrubbing
- G. Cut-Ox
- NS. Not Selected

Summary of FGD Systems (Ref. 9)

Number and Company Name	Unit Name	Location	Start Up Date	Status	New (N) Retrofit (R)	Boiler Size, mW	Proc. Cont. Type	Fuel	% S	Vendor
34 General Motors	Chev Pana 1 2 3 & 4	Parma OH	3-74	1	R	32	A	Coal	2.5	-
35 General Public Utilities	Homer City No 3	Homer City PA	4-77	5	N	650	NS	Coal	2.8	NS
36 Illinois Power	Wood River No 4	East Alton IL	10-72	1	R	110	G	Coal	2.9-3.2	Monsanto Env Chem
37 Indianapolis Power & Light	Petersburg No 3	Petersburg IN	4-77	5	N	515	D	Coal	3.0-3.5	NS
38 Kansas City Power & Light	Hawthorn No 3	Kansas City MO	11-72	1	R	140	A	Coal	.6-3.0	Combustion Eng.
39 Kansas City Power & Light	Hawthorn No 4	Kansas City MO	8-72	1	R	100	A	Coal	.6-3.0	Combustion Eng.
40 Kansas City Power & Light	La Cygne No 1	La Cygne KS	6-73	1	N	820	A	Coal	5.2	B & W
41 Kansas Power & Light	Jeffery No 1	St Mary KS	6-78	5	N	700	A	Coal	-	Combustion Eng.
42 Kansas Power & Light	Jeffery No 2	St Mary KS	6-79	5	N	700	A	Coal	-	Combustion Eng.
43 Kansas Power & Light	Lawrence No 4	Lawrence KS	12-68	1	R	125	A	Coal	3.5	Combustion Eng.
44 Kansas Power & Light	Lawrence No 5	Lawrence KS	11-71	1	N	400	A	Coal	3.5	Combustion Eng.
45 Kentucky Utilities	Green River 1 2 & 3	So Carlton KY	4-75	2	R	64	B	Coal	3.8	American Air Filter
46 Louisville Gas & Elec	Cane Run No 1	Louisville KY	6-80	6	R	110	B	Coal	3.5-4.0	NS
47 Louisville Gas & Elec	Cane Run No 2	Louisville KY	6-80	6	R	107	B	Coal	3.5-4.0	NS
48 Louisville Gas & Elec	Cane Run No 3	Louisville KY	6-80	6	R	137	B	Coal	3.5-4.0	NS
49 Louisville Gas & Elec	Cane Run No 4	Louisville KY	6-75	3	R	178	B	Coal	3.5-4.05	American Air Filter
50 Louisville Gas & Elec	Cane Run No 5	Louisville KY	6-76	6	R	183	B	Coal	3.5-4.05	NS
51 Louisville Gas & Elec	Cane Run No 6	Louisville KY	6-77	6	R	277	B	Coal	3.5-4.0	NS
52 Louisville Gas & Elec	Mill Creek No 1	Louisville KY	6-79	6	R	330	B	Coal	3.5-4.0	NS
53 Louisville Gas & Elec	Mill Creek No 2	Louisville KY	6-78	6	R	330	B	Coal	3.5-4.0	NS
54 Louisville Gas & Elec	Mill Creek No 3	Louisville KY	6-77	3	R	425	B	Coal	3.5-4.0	American Air Filter.
55 Louisville Gas & Elec	Mill Creek No 4	Louisville KY	6-79	6	R	425	B	Coal	3.5-4.0	NS
56 Louisville Gas & Elec	Paddys Run No 6	Louisville KY	4-73	1	R	65	B	Coal	3.5-4.0	Combustion Eng.
57 Montana Power	Colstrip No 1	Colstrip MT	5-75	2	N	360	B	Coal	0.8	Combustion Equip Assoc
58 Montana Power	Colstrip No 2	Colstrip MT	5-76	2	N	360	B	Coal	0.8	Combustion Equip Assoc
59 Montana Power	Colstrip No 3	Colstrip MT	7-78	6	N	700	NS	Coal	0.7	NS
60 Montana Power	Colstrip No 4	Colstrip MT	-79	6	N	700	NS	Coal	0.7	NS
61 Nevada Power	Reid Gardner No 1	Maopa NV	12-73	1	R	125	D	Coal	.5-1.0	Combustion Equip Assoc
62 Nevada Power	Reid Gardner No 2	Maopa NV	12-73	1	R	125	D	Coal	.5-1.0	Combustion Equip Assoc
63 Nevada Power	Reid Gardner No 3	Maopa NV	6-75	2	R	125	D	Coal	.5-1.0	Combustion Equip Assoc
64 Nevada Power	Reid Gardner No 4	Maopa NV	6-78	4	R	125	D	Coal	.5-1.0	Combustion Equip Assoc
65 New England Elec System	Brayton Point No 3	Somerset MA	0-0	6	R	650	NS	Coal	0.3	NS
66 Northern Indiana Pub Serv	Baily No 7	Chesteron IN	0-0	6	R	190	NS	Coal	0.3	NS

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- D. Sodium Carbonate Scrubbing
- E. Wellman-Lord
- F. Lime/Limestone Scrubbing
- G. Cut-Ox
- NS. Not Selected

Number and Company Name	Unit Name	Location	Start up Date	Status	New (N) Retrofit (R)	Boiler Size, mw	Type	Fuel	% S	Vendor
67 Northern Indiana Pub Serv.	Baily No 8	Chesterton IN	0-0	6	R	400	NS	Coal	0.3	NS
68 Northern Indiana Pub Serv.	Mitchell No 11	Gary IN	6-75	2	R	115	E	Coal	3.2-3.5	Davy Powergas
69 Northern States Power	Sherburne No 1	Becker MN	5-76	2	N	680	A	Coal	1	Combustion Eng.
70 Northern States Power	Sherburne No 2	Becker MN	3-77	2	N	680	A	Coal	1	Combustion Eng.
71 Pennsylvania Power Co.	Bruce Mansfield No 1	Shippingport PA	6-75	2	N	880	B	Coal	4.3	Chemico
72 Pennsylvania Power Co.	Bruce Mansfield No 2	Shippingport PA	8-75	2	N	880	B	Coal	4.3	Chemico
73 Pennsylvania Power Co.	Bruce Mansfield No 3	Shippingport PA	8-78	6	N	800	B	Coal	4.3	NS
74 Philadelphia Electric	Eddystone No 1	Eddystone PA	12-74	2	R	125	C	Coal	2.5	United Engineers
75 Potomac Electric & Power	Dickerson No 3	Dickerson MD	9-73	1	R	100	C	Coal	2.0	Chemico
76 Public Serv of Indiana	Gibson No 2	Princeton IN	5-75	2	R	650	F	Coal	1.5	NS
77 Public Serv of New Mexico	San Juan No 1	Waterflow NM	10-77	3	R	375	E	Coal	0.8	Davy Pow-Wellman Lord
78 Public Serv of New Mexico	San Juan No 1	Waterflow NM	1-77	3	R	340	E	Coal	0.8	Davy Pow-Wellman Lord
79 Public Serv of New Mexico	San Juan No 3	Waterflow NM	5-78	6	N	500	NS	Coal	0.8	NS
80 Public Serv of New Mexico	San Juan No 4	Waterflow NM	5-80	6	R	500	NS	Coal	0.8	NS
81 S. Carolina Pub Serv Auth	Georgetown No 2	Georgetown SC	5-77	5	N	140	F	Coal	1.0	NS
82 Salt River Project	Navajo No 1	Page AZ	3-76	6	N	750	F	Coal	.5-.8	NS
83 Salt River Project	Navajo No 2	Page AZ	10-76	6	N	750	F	Coal	.5-.8	NS
84 Salt River Project	Navajo No 3	Page AZ	3-77	6	N	750	F	Coal	.5-.8	NS
85 Southern California Ed	Mohave No 1	Laughlin NV	6-77	6	R	640	B	Coal	.5-.8	NS
86 Southern California Ed	Mohave No 1	Laughlin NV	12-74	2	R	160	A	Coal	.5-.8	VOP
87 Southern California Ed	Mohave No 2	Laughlin NV	11-73	1	R	160	B	Coal	.5-.8	Stern - Rogers
88 Southern California Ed	Mohave No 2	Laughlin NV	6-77	6	R	640	B	Coal	.5-.8	NS
89 So Mississippi Pwr Coop	Hattiesburg No 1	Hattiesburg MS	6-79	5	N	223	A	Coal	-	NS
90 So Mississippi Pwr Coop	Hattiesburg No 2	Hattiesburg MS	6-79	5	N	223	A	Coal	-	NS
91 Southwestern Public Serv	Harrington No 1	Amarillo TX	8-76	2	N	345	F	Coal	0.5	Combustion Eng.
92 Springfield Utility Board	Southwest No 1	Springfield MO	6-76	2	N	200	A	Coal	-	VOP
93 Tennessee Valley Authority	Shawnee No 10	Paducah KY	4-72	1	R	30	F	Coal	-	TVA
94 Tennessee Valley Authority	Widows Creek No 8	Bridgeport AL	1-77	2	N	550	A	Coal	3.7	TVA
95 Texas Utilities	Martin Lake No 1	Tatum TX	9-76	3	N	793	A	Coal	0.40	Research Cottrell
96 Texas Utilities	Martin Lake No 2	Tatum TX	6-77	3	N	793	A	Coal	0.44	Research Cottrell
97 Texas Utilities	Martin Lake No 3	Tatum TX	12-78	5	N	793	A	Coal	0.44	Research Cottrell
98 Texas Utilities	Martin Lake No 4	Tatum TX	12-79	5	N	793	A	Coal	0.44	Research Cottrell
99 Texas Utilities	Monticello No 3	Mt Pleasant TX	12-78	5	N	800	A	Coal	.8-1.0	NS

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B. Lime Scrubbing  
C. Magnesium Oxide Scrubbing  
D. Sodium Carbonate Scrubbing  
E. Wellman-Lord  
F. Lime/Limestone Scrubbing  
G. Cut-Ox  
NS. Not Selected

Table 2  
Systems Classified By Scrubber Type

	mm.	%
Limestone	14,821	
Lime	9,305	
Lime/Limestone	5,313	
Subtotal	29,439	81.4
Sodium Carbonate	1,015	2.8
Wellman-Lord	830	2.3
Cat-Ox	110	.3
Not Selected	<u>4,415</u>	<u>13.2</u>
	36,184	100

Table 3

Status of Systems By Scrubber Type

	mw.	%
Operational Units (non-lime or limestone 610 mw)	2,849	7.8
Units Under Construction	6,192	17.1
Planned-Contract Awarded	3,904	10.8
Planned-Letter of Intent Signed	530	1.5
Planned-Requesting/Evaluating Bids	6,237	17.4
Considering Only FED Systems	<u>16,472</u>	<u>45.4</u>
	36,184	100

Table 4  
FDS Startup Schedule

<u>Projected Start Up Date</u>	<u>Number of New Units</u>	<u>New Units mw</u>	<u>Number of Retrofit Units</u>	<u>Retrofit Units mw</u>
71	1	400	1	80
72	1	37	6	697
73	1	820	6	690
74	0	0	2	285
75	3	2,040	5	1,132
76	9	4,351	3	533
77	10	5,153	8	4,151
78	10	5,818	5	1,032
79	12	5,016	2	755
80	1	550	4	854
81	0	0	0	0
82	0	0	0	0
83	<u>1</u>	<u>550</u>	<u>1</u>	<u>1</u>
	49	24,735	42	10,209
		Ave: 504		Ave: 243
No Date Selected			3	<u>1,240</u>
				11,449



Table 5 (Continued)  
Development Status of Scrubber Process Systems

Process Types	Concept		Bench		Pilot		Demo.		Commercial		Fuel Type		Efficiency or Projected Efficiency	Remarks
	1	2	3	4	5	6	7	8	9	10	Coal	Oil		
Ignite Ash Sorption	X	X	X	X	X	X								
Fixed Metal Oxides Sorp.	X	X	X	X	X	X								
Wolsten Salts	X	X	X	X										
Wolsten Carbonate	X	X	X	X										
Fluidized Bed	X	X	X	X	X	X								

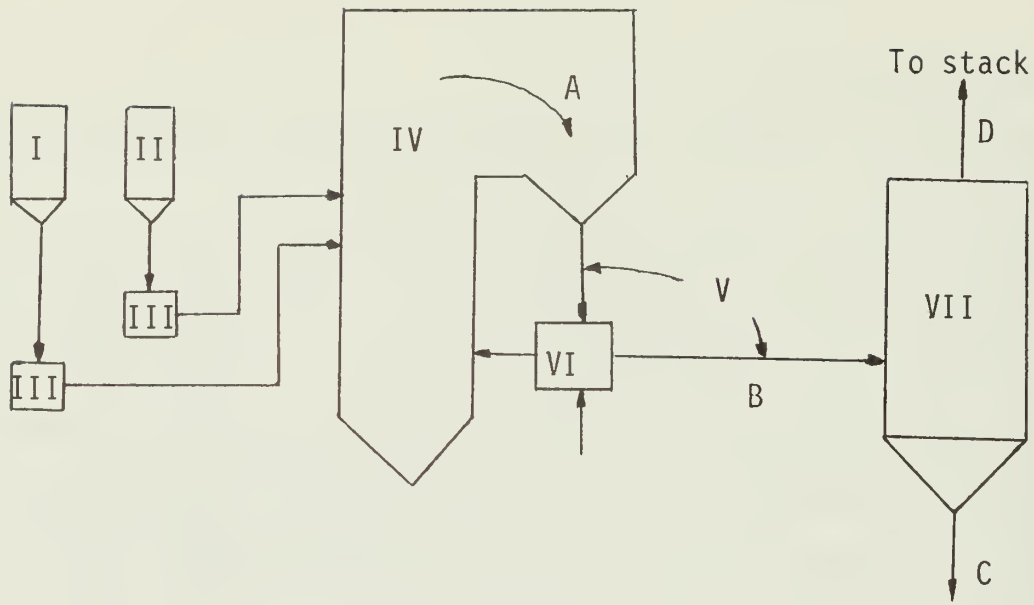


## Lime/Limestone Processes

Limestone-wet scrubbing can be effected either by injecting limestone into the power plant boiler and collecting the resulting calcium-sulfur compounds in a wet (water only) scrubber or by introducing limestone (or lime) directly into the scrubber system with no furnace injection.

In the first configuration, shown in Figure 1, pulverized limestone or dolomite may be introduced directly into the furnace through nozzles located above the coal burners. It may also be premixed with coal prior to pulverization and fed with the pulverized coal through common burners (I - IV). Partial reaction with sulfur compounds is effected in the furnace and in the flue gas circuit (IV, V) with completion of reactions occurring in packed marble-bed scrubbers (VIII) which are located between the air heaters (VI) and the stack. The advantages of this configuration are: (1) better utilization of the adsorbent due to the calcination of the limestone, which makes it more reactive, (2) reduced boiler corrosion, (3) sulfur dioxide sorption in the boiler, (4) adaptable to existing units, (5) uses normal chemical process components, (6) low capital investment, and (7) no potential marketing problems. The disadvantages of this configuration are: (1) a water pollution problem can be created, (2) a waste disposal problem is aggravated; the storage volume for the waste produced (excluding fly ash) is 200-300 percent greater than that required for the fly ash alone, (3) the possibility of serious scrubber scaling and corrosion problems, (4) a loss of plume buoyancy thus requiring reheat, (5) high maintenance and operating costs, (6) serious bed plugging problems, (7) no useful by-product is produced, (8) boiler fouling, (9) possible wasting of absorbent by "dead-burning" it in the furnace and losses which occur in the flue gas circuit.

In the second configuration shown in Figure 2, the limestone (lime) is



I. Coal Supply

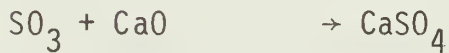
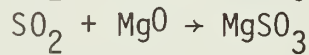
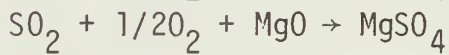
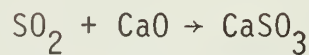
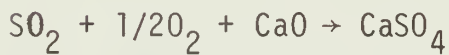
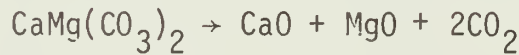
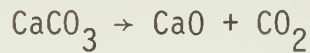
II. Limestone Supply

III. Pulverizers

IV. Furnace & V. Flue Gas Circuit (Ducts)

Coal + Air + Heat → Products (CO<sub>2</sub>, SO<sub>2</sub>, SO<sub>3</sub>, H<sub>2</sub>O, O<sub>2</sub>, NO<sub>x</sub>, Ash)

Limestone + Heat → Products



VI. Air Heater

All SO<sub>3</sub> is removed by the time the flue gas reaches this point

VII. Wet Scrubber

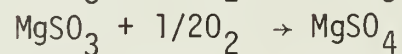
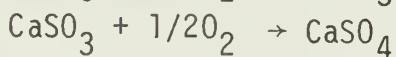
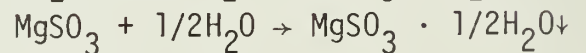
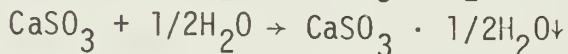
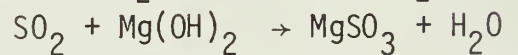
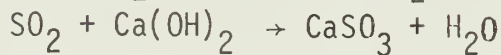
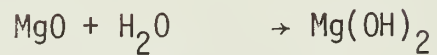
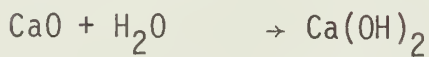
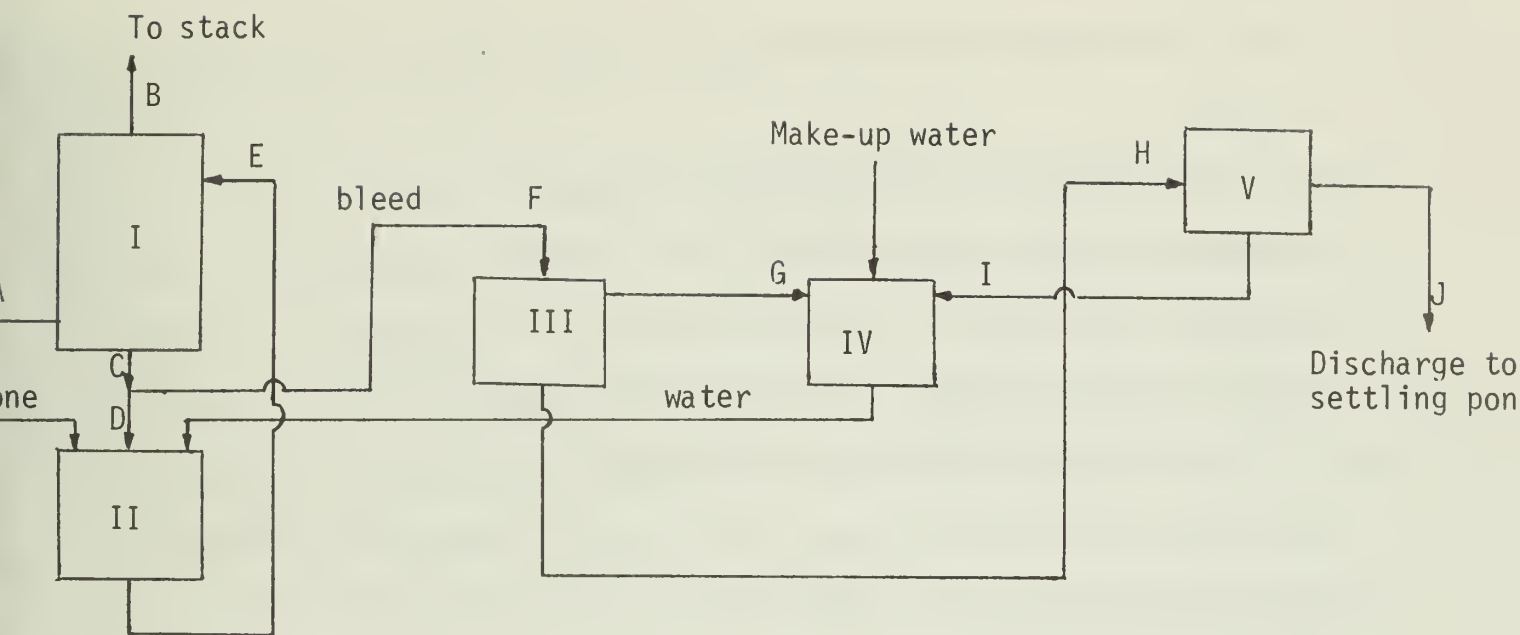
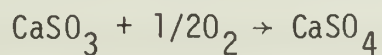
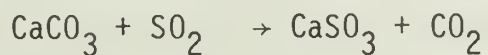


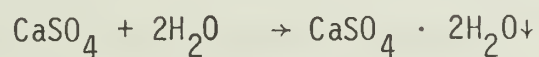
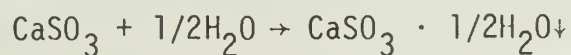
Figure 1 Lime/Limestone Injection Process Description



I. Scrubber - Removes  $\text{SO}_2$  and fly ash



II. Scrubber effluent hold tank



III. Clarifier - Solids separation

IV. Process water hold tank

V. Centrifuge or filter - Solids separation

Figure 2 Lime/Limestone Absorption Process Description

introduced directly into the scrubber system instead of the boiler ("tailend" system).

The flue gas containing  $\text{SO}_2$  is scrubbed (I.) with a slurry containing limestone and some undissolved solids (combined limestone-sulfur reaction products and fly ash). The level of solids is optimized to maintain the stability (with respect to scaling) of the slurry pH and  $\text{SO}_2$  removal efficiency.

The flue gas then leaves the scrubber to be exhausted to the atmosphere at a safe concentration of  $\text{SO}_2$ . The scrubber effluent moves from the scrubber to a hold tank (II.).

Between the scrubber and the hold tank, a side stream of spent scrubbing slurry is continuously bled and subjected to a dewatering operation (III., IV., V.). The amount of scrubber effluent removed in the side stream is small (on the order of a few percent), as it is necessary to purge only that amount of reaction products which are formed in the scrubber (under steady-state conditions).<sup>5</sup> The solids removed by the dewatering operation (reaction products, fly ash and unreacted limestone) are discharged as waste to either a disposal pond or landfill site.<sup>2</sup>

In the hold tank, fresh limestone is added to make up for limestone lost in the formation of the reaction products, and for amounts lost in the side stream to the dewatering operation.<sup>5</sup> Water is also added at this point to make up for various system losses due to evaporation in the scrubber, evaporation in the settling pond, and water of hydration associated with the discarded waste solids. The water added can be fresh and/or recycled pond water. The addition of water in the hold tank provides beneficial slurry dilution to assure a desupersaturated state before passage into the scrubber. The slurry is held in the hold tank for a sufficient period of time (residence time) to allow for completion of the chemical reactions and adequate desupersaturation.

The slurry is then pumped from the hold tank back to the scrubber, where the cycle begins again.

The advantages of this configuration (with limestone as the absorbent) are: (1) due to the amount of work on this process (pilot, demonstration and commercial plants), it is approaching a refined state, i.e., there are now methods of controlling the once ever-present problems of very severe scaling, plugging, and erosion., (2) adaptability to existing units, (3) and a relatively low capital investment.

The disadvantages of this configuration are: (1) a water pollution problem can arise if the waste products are not properly handled, (2) a waste disposal problem is aggravated since the storage volume for the waste products is 200-300 percent greater than that required for the fly ash alone, (3) a loss of plume buoyancy thus requiring reheat (characteristic of many other scrubbing processes), (4) no useful by-product is produced, (5) scaling, plugging, and erosion problems can arise if the process is not carefully designed and operated, (6) use of a less reactive absorbent material as compared to lime requires a higher stoichiometric ratio of limestone to  $\text{SO}_2$  absorbed with the result that more slurry is needed to be recirculated, and (7) the removal process consumes energy which reduces the overall efficiency of the power plant.

If lime is used in the configuration instead of limestone, the scrubbing efficiency is increased. This advantage is partially offset by the higher cost of lime over limestone and the greater potential for scaling under certain process configurations.<sup>6</sup>

## Mass Balance

There are two general assumptions which apply to both mass balances which are discussed below: (1) the  $\text{SO}_2$  removal system is in an operating equilibrium, i.e., the removal of  $\text{SO}_2$  from the flue gas does not disturb this equilibrium as the input rate of sulfur is equal to the output rate of sulfur, and (2) the reaction equations presented are overall reaction equations of the most important products and represent a hypothetical chain of events.

The chemistry associated with the limestone injection into the boiler is given on page 14.

For the purpose of calculation assume the following conditions: 100 percent removal of  $\text{SO}_3$ , 25 percent removal of  $\text{SO}_2$  in the flue gas circuit, 87 percent removal of the  $\text{SO}_2$  entering the scrubber, an overall removal efficiency of 90 percent, the limestone is 100 percent  $\text{CaCO}_3$  (no  $\text{MgO}$  present), and that limestone is introduced at a rate of 1.3 moles per mole of  $\text{SO}_2$  adsorbed.

The results of the calculations are given in Table 6. The calculations also show that this process yields 0.79 lbs of  $\text{CaSO}_3 \cdot 1/2 \text{H}_2\text{O}$ /lb of  $\text{CaCO}_3$  and 0.27 lbs of  $\text{CaSO}_4 \cdot 2 \text{H}_2\text{O}$ /lb of  $\text{CaCO}_3$ . 14.43 lbs of limestone are required for 100 lbs of coal burned.

The chemistry associated with the second configuration with limestone as the absorbent material is given on page 15.

The calculations for this configuration are based on the following assumptions: 90 percent removal of  $\text{SO}_2$ , 10 percent oxidation of the  $\text{CaSO}_3$  formed, and a 1.65  $\text{CaCO}_3/\text{SO}_2$  mole ratio.

The results of the calculations are given in Table 7. The calculations

Table 6

Material Balance - Boiler Injection of Limestone, followed by wet scrubbing  
 (lb. per 100 lb. of coal burned)

Component Stream	N <sub>2</sub>	CO <sub>2</sub>	H <sub>2</sub> O	O <sub>2</sub>	SO <sub>2</sub>	SO <sub>3</sub>	NO <sub>x</sub>	CaO <sup>+</sup>	CaSO <sub>3</sub>	CaSO <sub>4</sub>	Ash	CaSO <sub>3</sub> · 1/2H <sub>2</sub> O	CaSO <sub>4</sub> · 2H <sub>2</sub> O
A	773.146	229.638	19.35	39.136	7.232	0.72	0.54	+8.08	-	-	11.2	-	-
B	773.146	229.638	19.35	39.096	5.422	-0.72	0.54	5.996	+3.05	+0.38	-	-	-
C	-	-	-1.616	-0.18	-4.72	-	-	-5.996	-3.05	-0.38	11.2	+11.36	+3.94
D	773.146	229.638		38.916	0.702	-	0.54	-	-	-	-	-	-

14.43 lbs. of Limestone

Table 7

Material Balance - Alkaline earth (calcium\*) absorbent  
(lb. per 100 lb. of coal burned)

component Stream	N <sub>2</sub>	CO <sub>2</sub>	H <sub>2</sub> O	O <sub>2</sub>	SO <sub>2</sub>	SO <sub>3</sub>	NO <sub>x</sub>	Ash	CaSO <sub>4</sub> ·2H <sub>2</sub> O	CaSO <sub>3</sub> ·1/2H <sub>2</sub> O	CaCO <sub>3</sub>	Slurry
A	773.146	224.752	19.350	39.136	7.232	0.72	0.54	11.2				
I.		+4.47	-1.19	-0.16	-6.51			-11.1	+1.75	+11.81	-10.17	
B	773.146	229.222	18.16	38.976	0.722	0.72	0.54	0.1				
C									1.75	11.81	+16.78	
D									1.75	11.81	6.61	
E												
F									1.75	11.81	6.61	
G												
H									1.75	11.81	6.61	
I												
J									-1.75	11.81	-6.61	

-20.17



also show that this process yields 0.7 lbs of  $\text{CaSO}_3 \cdot 1/2 \text{H}_2\text{O}$ /lb of  $\text{CaCO}_3$  and 0.1 lbs of  $\text{CaSO}_4 \cdot 2 \text{H}_2\text{O}$ /lb of  $\text{CaCO}_3$ . 16.78 lbs of limestone are required for 100 lbs of coal burned.

### Calcium Hydroxide

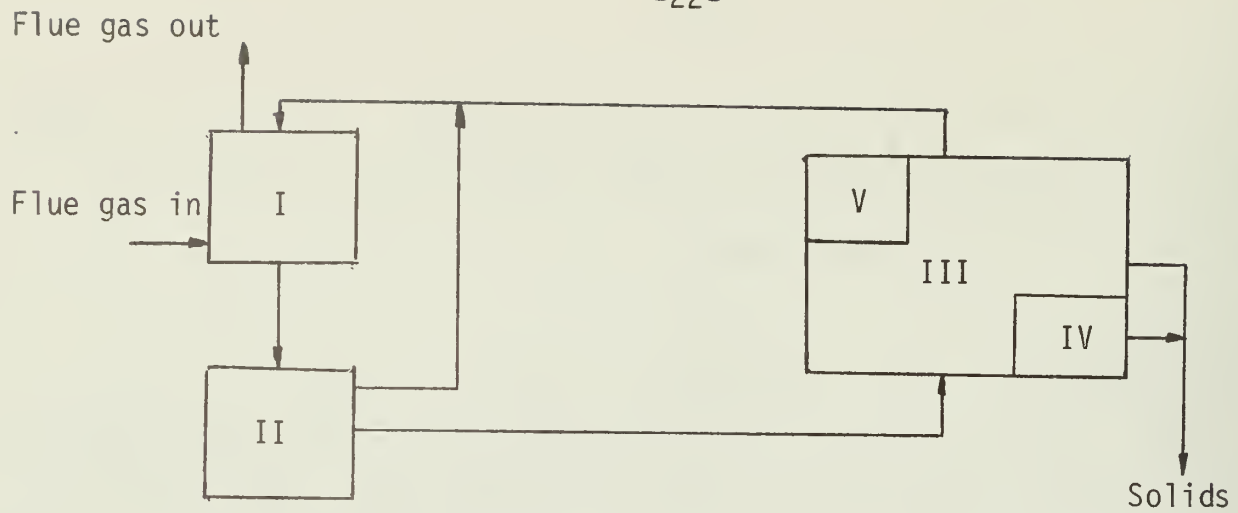
Since the chemistry and system characteristics of a calcium hydroxide system are sufficiently close to those associated with a lime/limestone system, the calcium hydroxide system will not be discussed separately in this report.

### Sodium/Calcium Double Alkali Scrubbing

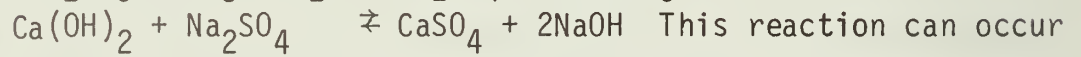
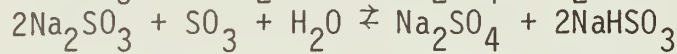
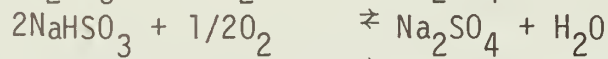
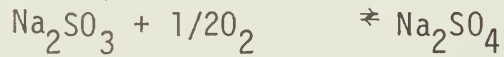
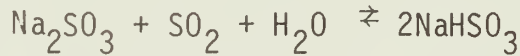
As noted earlier, there is an incompatibility in the lime/limestone scrubber between the system parameters which maximize  $\text{SO}_2$  removal and minimize scaling and deposition problems. In an attempt to alleviate this incompatibility, the double alkali scrubber has been proposed. In this configuration, the  $\text{SO}_2$  is removed using a sodium solution which is optimal for  $\text{SO}_2$  removal in one tank and uses calcium carbonate in another tank for regeneration of the sodium solution and the formation of calcium sulfate for ultimate disposal. This system is a hybrid regenerative and throwaway system.

### Sodium/Calcium Double Alkali Scrubbing (Sodium Hydroxide Process)

The sodium/calcium double alkali scrubbing process as shown in Figure 3 removes flue gas  $\text{SO}_2$  by contacting it with a solution of a soluble alkali salt. In this case sodium is used, but potassium and ammonia are also possibilities. The spent scrubbing solution is then treated with another alkali element, in this case Ca, to regenerate the alkali scrubbing solution and to remove the absorbed sulfur as calcium-sulfur solids.<sup>7</sup>



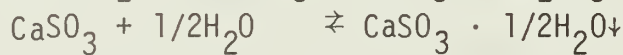
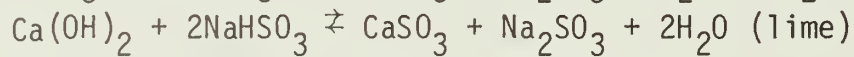
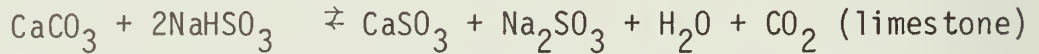
I. Scrubber



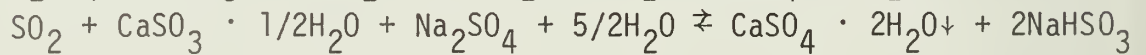
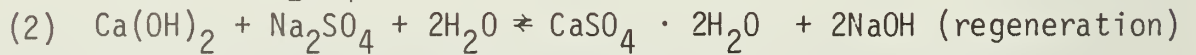
This reaction can occur if the system is not properly operated. Formation of excess  $\text{CaSO}_4$  can lead to scaling ( $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ ).

II. Recirculation tank

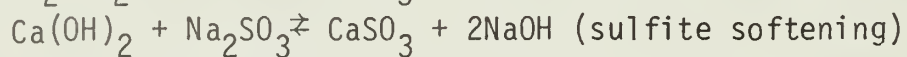
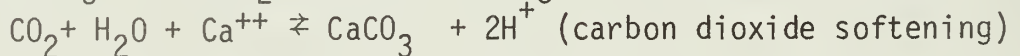
III. Regeneration Unit



IV. Sulfate removal



V. Excess calcium removal



Additional formation of active scrubbing agent  $\text{Na}_2\text{SO}_3$

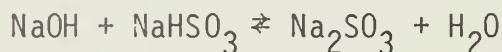


Figure 3 Double Alkali Process Description 1

A clear liquor solution containing sodium sulfite, sodium bisulfite and sodium sulfate is fed to the scrubber (I.) to absorb flue gas  $\text{SO}_2$ . Initially the scrubbing solution is at a high pH, indicating a high sulfite to bisulfate ratio.  $\text{SO}_2$  reacts with the sulfite in solution to form bisulfite, thereby lowering the sulfite to bisulfite ratio (as indicated by a lower solution pH). It should be noted that lowering the sulfite concentration can lead to scaling if there is insufficient sulfite to totally react with any calcium ions which enters the scrubber from a different process step. The calcium ions can react with the sulfate in solution to form calcium sulfate, which if present in sufficient quantities will precipitate. For this reason, the concentrations of calcium ions and sulfate entering the scrubber are carefully controlled. In other parts of the system, the formation of calcium-sulfur solids is promoted to rid the system of absorbed  $\text{SO}_2$ . But this is not the case in the scrubber.

Sodium sulfate is also formed in the scrubber by oxidation of bisulfite and sulfite in solution, and by the removal of  $\text{SO}_3$ .

After the scrubbing step, the effluent liquor moves from the scrubber to a recirculation tank (II.). From here, some of the effluent is recycled back to the scrubber and the remainder is processed for regeneration of the active scrubbing agent, sulfite, (III.), removal of the adsorbed  $\text{SO}_2$  (reaction with calcium) (III.), removal of excess calcium ions (softening) to prevent scaling (V.), and removal of sulfate (IV.).

During regeneration (III.), the bisulfite rich spent scrubbing solution is treated with lime or limestone to yield sulfite, the active scrubbing agent, and the adsorbed  $\text{SO}_2$  as calcium sulfur solids. The waste solids are purged from the system and the regenerated scrubbing liquor is subjected to further processing.

Calcium ion concentrations in the scrubbing solutions can be controlled by carbon dioxide, carbonate, or sulfite softening (V.). The net effect of these techniques is to reduce calcium ion concentrations to prevent scaling in the scrubber. Addition of sodium carbonate (for carbonate softening) has the advantage of both softening and replacing sodium losses from the system. Addition of sodium carbonate has the disadvantage of requiring a possible sodium purge if the amount required for softening exceeds the amount required to replace sodium losses.

The sulfate formed by oxidation must be removed from the system to prevent its build-up in the system and its combination with calcium (scaling) in the scrubber. It may be removed from the system (IV.) either as a solid waste or as a soluble sodium salt by a purge. From the standpoint of environmental acceptability, the removal of sulfate as a solid product is highly desirable as the soluble salt represents a serious water pollution potential. Sulfate removal is generally carried out by treating a side stream from the scrubber liquor loop.

After the spent scrubbing liquor is subjected to these regeneration and purification steps, it is sent back to the scrubber, where the cycle begins again.

Five different modes of operation have been identified for sodium/calcium double alkali scrubbing systems.<sup>7</sup> These include:

- (1) Limestone and lime regeneration, dilute active alkali, with sulfite softening.
- (2) Lime regeneration, dilute active alkali, with sulfite softening.
- (3) Lime regeneration, concentrated active alkali, sulfite softening, with side stream sulfate treatment (removal).
- (4) Limestone regeneration, concentrated active alkali, side stream sulfate treatment (removal), with sulfite softening.

(5) Lime regeneration, dilute active alkali, with carbonate softening.

Systems using lime have shorter hold times (10 minutes as compared to one hour) and greater utilizations (90 percent as compared to 80 percent) than systems using limestone. Also, systems using dilute active alkali must circulate large volumes of relatively dilute scrubbing liquor, whereas smaller volumes are associated with the concentrated alkali scrubbing liquors.

The advantages of the sodium/calcium double alkali scrubbing process are (1) adaptable to existing units, (2) low capital investment, (3) scale free operation if operated properly, and (4) no erosion or corrosion problems.

The disadvantages of this process are (1) a water pollution problem can arise if the waste products (both solid and liquid) are not properly handled, (2) a solid waste disposal problem is aggravated, (3) a loss of plume buoyancy thus requiring reheat (characteristic of many other scrubbing processes), (4) no useful by-product is produced, (5) scaling can occur if the system is not properly operated, and (6) the removal process consumes energy which reduces the overall efficiency of the power plant.

The absorption step has been tested on a 250,000 lb/hr industrial steam plant for a period of 1 year with a 90 percent  $\text{SO}_2$  removal efficiency being realized.<sup>17</sup> The regeneration step has been carried out only at the pilot plant stage. In a 2000 cfm pilot plant having an absorption/regeneration cycle, it was found that scaling was negligible, there was no sulfate built-up in the recirculating liquor, calcium levels in the recycle water were less than 70 ppm, the  $\text{SO}_2$  removal efficiency varied from 86 to 95 percent, and the make up requirements for the sodium hydroxide was 1-4 percent. The lime utilization factor ranged from 95-98 percent.

### Mass Balance

The flow chart of a double alkali  $\text{SO}_2$  scrubbing process is shown in Figure 4. Characteristics of this system are (1) it uses a concentrated active alkali to scrub  $\text{SO}_2$ , (2) it makes use of sulfite softening, (3) it purges sulfate in a side stream, and (4) lime is used in the regeneration step.

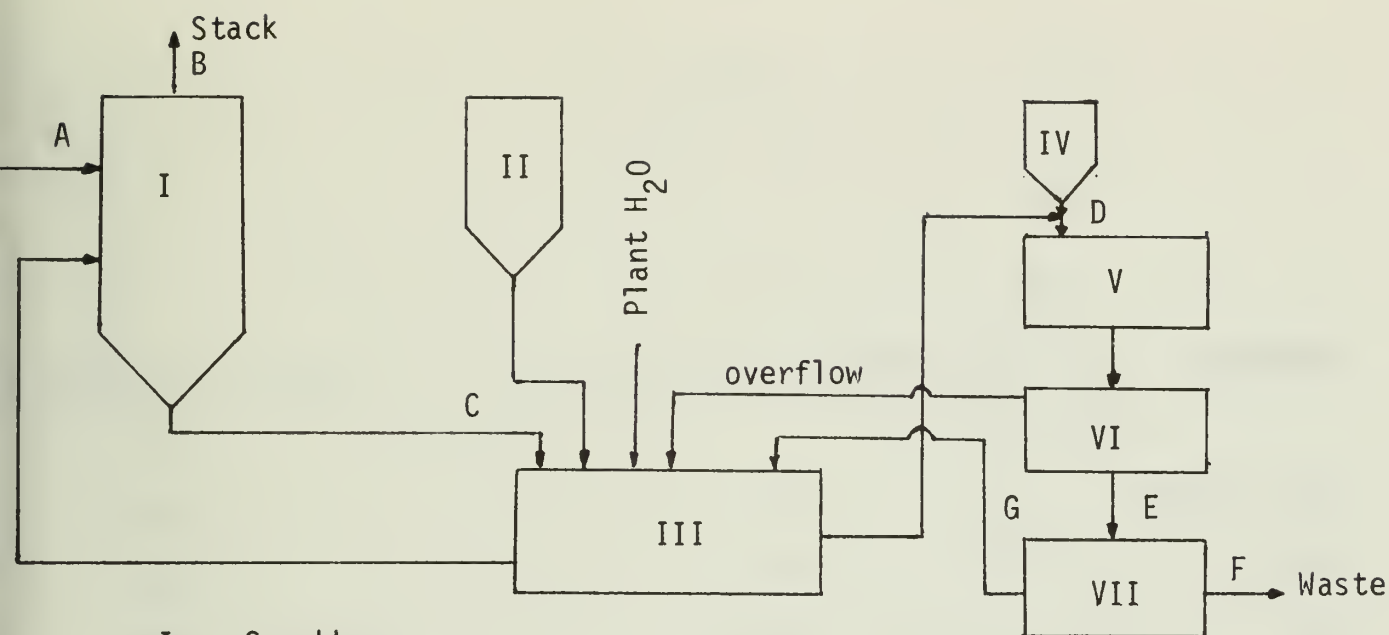
Two general assumptions used in the calculations are (1) the  $\text{SO}_2$  removal system is in an operating equilibrium, the removal of  $\text{SO}_2$  from the flue gas stream does not disturb this equilibrium as the input rate of sulfur is equal to the output rate of sulfur, and (2) the reaction equations presented are overall reaction equations of the most important components and represent a hypothetical chain of events.

For the purpose of calculation it is also assumed that there is 90 percent removal of  $\text{SO}_2$  and a 90 percent removal of  $\text{SO}_3$  from the flue gas. It is also assumed that the sodium sulfate is formed in the system by the removal of  $\text{SO}_3$ , by the oxidation of the bisulfite (15 percent of the absorbed  $\text{SO}_2$  is assumed to go to sulfate by oxidation of the bisulfite), and the oxidation of the sulfite present in the scrubbing solution. It should be noted that the amount of sulfate produced is equal to the amount formed by the oxidation of the bisulfite; the oxidation of sulfite does not involve absorbed  $\text{SO}_2$ , it involves the sulfite in the scrubbing solution.

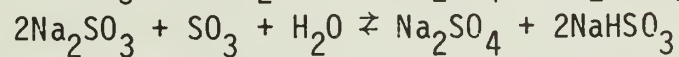
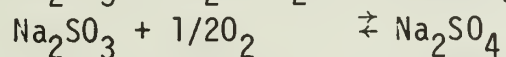
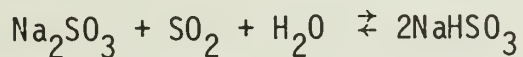
The results of the calculations are given in Tables 8 and 9.

In examining alternatives to lime or limestone based scrubbing processes, Shah and Quigley suggested the following criteria should be examined:<sup>8</sup>

- (1) it should be flexible enough for application to a variety of stack flue gases containing varying concentrations of sulfur dioxide and fly ash;
- (2) it should be capable of reducing the stack gas  $\text{SO}_2$  concentration to meet not



I. Scrubber

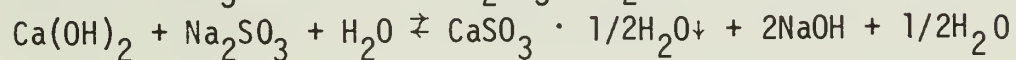
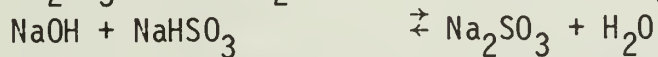
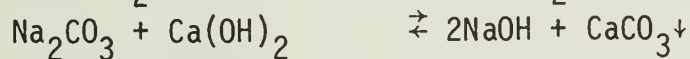
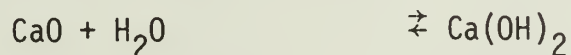


II. Soda ash supply

III. Recirculation tank

IV. Lime supply

V. Lime reactor & VI. Thickener



VII. Rotary filter

Figure 4 Double Alkali Process Description 2

Table 8

<u>Component</u>	<u>lbs./100 lbs. of coal burned</u>	<u>lbs./lb. of Na<sub>2</sub>CO<sub>3</sub> imp</u>
CaSO <sub>3</sub> · 1/2H <sub>2</sub> O	8.582	1.869
CaCO <sub>3</sub>	4.332	0.943
CaO	0.324	0.071
Na <sub>2</sub> SO <sub>4</sub>	5.488	1.195
Na <sub>2</sub> SO <sub>3</sub>	0.589	0.128



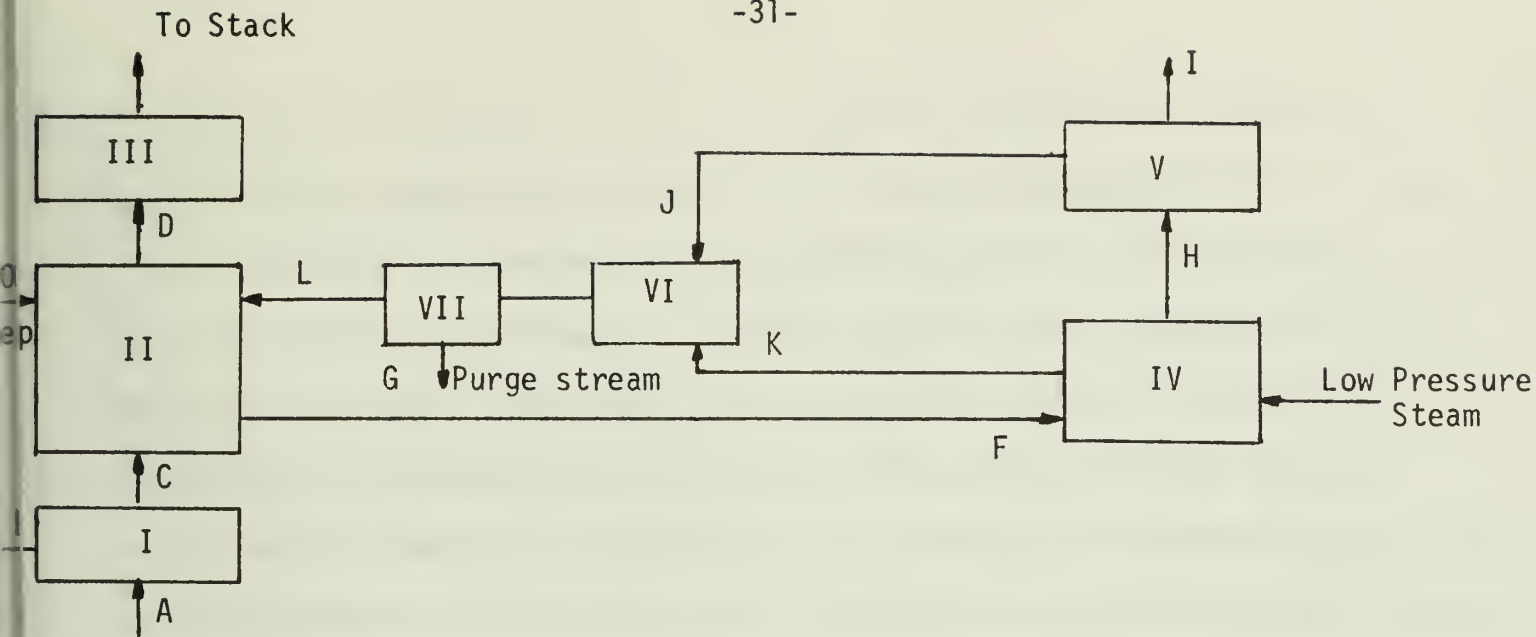


only today's pollution codes, but also tomorrow's code; (3) alkali used for adsorption of sulfur dioxide should be regenerated for reuse, the loss should be minimum; (4) the process should be flexible enough to produce a variety of saleable products, e.g., elemental sulfur, sulfuric acid, liquid sulfur dioxide, (5) the utility industry should be spared the burden to capitalize and operate chemical recovery processing plants. This means that only minimum equipment necessary to eliminate air pollution problems should be at the power plant sites; (6) the process should be adaptable to the central process recovery concept which reduces the economic burden and operating involvement of individual plants and makes possible the manufacture of saleable products of lower unit cost and in the proximity of the market; (7) elimination of air pollution should not result in land or water pollution; (8) the process economics should be such that to the power industry, the cost in terms of dollars per ton of coal should be less than the alternate for low sulfur oil or coal. To the central process operator, the unit cost of saleable products must be less than or equal to the cost of that product if manufactured conventionally.

#### Sodium (Wellman-Lord) Scrubbing

The Wellman-Lord sodium (potassium was tried without success) scrubbing process shown in Figure 5 absorbs flue gas  $\text{SO}_2$  into a sodium sulfite/bisulfite solution. Removal is effected by counter-currently contacting the  $\text{SO}_2$  gas with the scrubbing solution. The spent scrubbing solution is then processed to regenerate the active scrubbing solution and the absorbed  $\text{SO}_2$ . The  $\text{SO}_2$  can then be processed into a final saleable product.

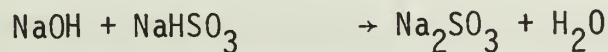
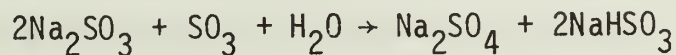
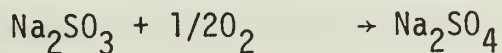
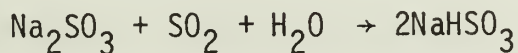
Before the absorption step takes place, the amounts of fly ash and  $\text{SO}_3$  (the  $\text{SO}_3$  forms sodium sulfite) contained in the flue gas are reduced (I.). This is necessary as both of these components can lead to contamination and



I. Prescrubber

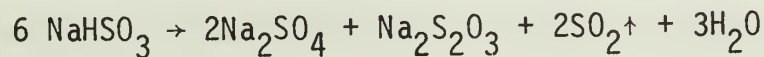
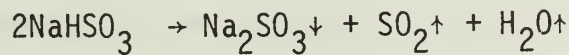
Removes fly ash and some SO<sub>3</sub>

II. Absorber



III. Reheater and Blower

IV. Evaporative Crystallizer



V. Condenser

Separates H<sub>2</sub>O and SO<sub>2</sub>

VI. Dissolver

Na<sub>2</sub>SO<sub>3</sub> crystals back into solution

VII. Removal of fly ash and sodium sulfate

Figure 5 Sodium Scrubbing Process Description

plugging of the removal system.

In the absorption step (II.), flue gas  $\text{SO}_2$  combines with the sodium sulfite in the scrubbing solution to form more sodium bisulfite. Some sodium sulfate is also formed during this step through the absorption of  $\text{SO}_3$  and the oxidation of  $\text{Na}_2\text{SO}_3$ . The formation of sulfate in the system and the fly ash removed during  $\text{SO}_2$  scrubbing necessitates a purge bleed stream to prevent these contaminants from building up and interfering with the systems chemistry and operation. It is known that the sodium sulfate can affect the active level of sulfite in solution.<sup>9</sup> The formation of sulfate can be controlled somewhat by introducing oxidation inhibitors into the system.

The scrubbing solution, rich in bisulfite, is transferred to the regeneration section where it is treated to reform the active absorbing agent, sulfite, to release the absorbed  $\text{SO}_2$  for further treatment and to reduce the amounts of fly ash and sulfate to safe operating levels. All of the scrubbing solution moves to the regeneration step after the absorption step without any part being recycled back through the scrubber for additional reaction (as in some  $\text{SO}_2$  removal processes).

In the evaporative crystallizer of the regeneration section, the absorbed  $\text{SO}_2$  is released from the bisulfite solution and the sodium sulfite is reformed. The sodium sulfite precipitates as it forms and builds a dense slurry of crystals. The  $\text{SO}_2$  is now available for further processing.

The solids (sodium sulfite crystals, fly ash, and sodium sulfate) and sulfite/bisulfite solution from the evaporative crystallizer are sent to a tank (VI.) where water [from the water vapor released in the evaporative crystallizer (IV and V)] is supplied to dissolve the sodium sulfite crystals to reform the scrubbing solution, now rich in sulfite.

The amounts of fly ash and sodium sulfate in the solution are adjusted for safe operation (VII) before the solution reenters the scrubber to start

the process over again.

The removed sulfate can be disposed of, or treated with lime to remove the sulfur in the form of calcium-sulfur compounds (as in the sodium/calcium double alkali removal process) and to recover sodium (in the form of sodium hydroxide) to be reused in the process. If calcium ions remain in the solution, they may eventually produce scaling problems in the system.

Make up sodium, in the form of sodium hydroxide, is added to the scrubber to replenish sodium losses due to the formation of sulfate and other sodium-based solids, and any sulfite losses.

The advantages of this sodium scrubbing process are: (1) adaptable to existing units, (2) scale free operation due to favorable solubilities, (3) no erosion or corrosion problems, and (4) a useful by-product is produced.

The disadvantages of this process are: (1) the need to dispose of purge solids (sodium sulfate and other sodium salts), (2) a water pollution problem can arise if the waste products are not properly handled, (3) a loss of plume buoyancy thus requiring reheat (characteristic of many other scrubbing processes), and (4) the removal process consumes energy which reduces the overall efficiency of the power plant.

### Mass Balance

The flow chart and the chemical reactions associated with the Wellman-Lord sodium scrubbing process are shown in Figure 5.

There are two general assumptions which apply to the mass balance. They are (1) the  $\text{SO}_2$  removal system is in an operating equilibrium i.e., the removal of  $\text{SO}_2$  from the flue gas does not disturb this equilibrium as the input rate of sulfur is equal to the output rate of sulfur, and (2) the reaction equations presented are overall reaction equations of the most important products and represent a hypothetical chain of events.

For the purpose of calculation the additional assumptions made include a 90 percent removal of  $\text{SO}_2$ , a 90 percent removal of  $\text{SO}_3$  from the flue gas stream, a 1 percent oxidation rate of  $\text{SO}_2$  in the scrubber, and no regeneration of the sulfate formed to recover sodium hydroxide.

The results of the calculations are given in Table 10 and Table 11.

### Magnesia Scrubbing-Regeneration Process

There are several process variations employing magnesia chemistry to remove flue gas  $\text{SO}_2$ . In these magnesia scrubbing regeneration processes,  $\text{SO}_2$  can be removed by reaction with a magnesia base slurry, solution, or solid. Magnesium-sulfur compounds are formed during the scrubbing step. These compounds are then processed to regenerate the active scrubbing agent, magnesium oxide, and to recover the absorbed  $\text{SO}_2$ . The regenerated magnesium oxide is recycled for further scrubbing. The recovered  $\text{SO}_2$  is further processed to form a final product: liquid  $\text{SO}_2$ , sulfuric acid, or elemental sulfur.

There are three major process variations for magnesia base  $\text{SO}_2$  scrubbing systems:<sup>10</sup> (1) magnesia slurry, (2)  $\text{MgO} - \text{MnO}_2$  slurry, and (3) clear liquor.

#### Variation #1 - Magnesia Slurry<sup>10</sup>

The flue gas is first treated to remove fly ash and some of the  $\text{SO}_3$  (I.). It is then routed to the scrubber for  $\text{SO}_2$  removal.

In the scrubber,  $\text{SO}_2$  is contacted by a slurry of unreacted magnesium oxide, undissolved magnesium sulfite-sulfate, and dissolved sulfite-sulfate. The  $\text{SO}_2$  leaves the flue gas stream to form magnesium-sulfur compounds in the slurry. The formation of undissolved sulfite and sulfate during this removal step increases the amounts of these species in the scrubbing slurry above the steady state operating levels. These excess solids must be removed to keep the system operating.

Table 10

<u>Component</u>	<u>lbs /100 lbs of coal burned</u>	<u>lbs /lb of NaOH input</u>
NaSO <sub>4</sub>	0.886	0.968
NaSO <sub>3</sub>	0.553	0.604
Na <sub>2</sub> S <sub>2</sub> O <sub>3</sub>	0.128	0.140
Fly ash	11.16	-
SO <sub>2</sub>	5.892	6.439
NaOH (input)	0.915	1

Table 11

Material Balance - Alkalie (sodium) absorption process  
 (1b. per 100 lbs. of coal burned)

Component Stream	N <sub>2</sub>	CO <sub>2</sub>	H <sub>2</sub> O	O <sub>2</sub>	SO <sub>2</sub>	SO <sub>3</sub>	NO <sub>x</sub>	Ash	NaOH	NaSO <sub>3</sub>	NaHSO <sub>3</sub>	Na <sub>2</sub> SO <sub>4</sub>	Na <sub>2</sub> S <sub>2</sub> O <sub>3</sub>
A	773.146	224.752	19.136	39.136	7.232	0.72	0.54	11.2					
B						-0.36		-10.1					
C	773.146	224.752	19.350	39.136	7.232	0.36	0.54	1.1					
D	773.146	224.752	19.350	39.136	0.722	0.072	0.54	.04					
E									+ .915				
F			-1.465	-.016	-6.51	-.288		1.06	-.915	-10.842	+19.317	+0.655	
IV.			+1.672		+5.892			1.06	+11.395	-19.317	+0.231	+0.128	
G								-1.06		-0.553		-0.886	-0.128
H					5.892								
I					-5.892								
J													
K													
L													

Crystals Solution  
 In  
 Solution Solution



Upon leaving the scrubber, a portion of the slurry is recycled back to the scrubber for additional reaction, and the remainder (in a bleed stream) is treated to remove the solids formed during  $\text{SO}_2$  removal.

The slurry in the bleed stream is dewatered to separate the solids from the slurry (III.). The "cleaned" slurry (at a safe level of solids content) is then recycled back to the scrubber.

The solids from the dewatering step are dried (IV.), then calcined (V.) with coke to regenerate magnesium oxide and drive off the absorbed  $\text{SO}_2$ . The regenerated magnesium oxide is recycled back to the scrubber with make-up magnesium oxide and water (VI.). The  $\text{SO}_2$  is processed to form a final product.

#### Variation #2 - $\text{MgO-MnO}_2$ Slurry<sup>10</sup>

This concept is a variation of the basic magnesia slurry variation discussed above.  $\text{SO}_2$  is removed by scrubbing the waste gas stream with a magnesium oxide-sulfite slurry containing a scrubbing reaction activator, manganese dioxide ( $\text{MnO}_2$ ). The sulfites, sulfates, and unreacted manganese dioxide are dried and calcined to regenerate the absorbent (magnesium oxide) and the activator with the driven off  $\text{SO}_2$  being processed to a final product.

#### Variation #3 - Clean Liquor<sup>10</sup>

This variation is also similar to variation #1.

$\text{SO}_2$  and fly ash are simultaneously scrubbed from the flue gas stream by an acidic solution of magnesium sulfites.

The magnesium-sulfur compounds (magnesium sulfite-sulfate formed during the scrubbing step) remain in solution rather than precipitating as in variation #1. The only solid material in the solution after scrubbing is fly ash, which is eventually removed by filtration.

After leaving the scrubber, a portion of the scrubbing liquor is recycled back to the scrubber, with the remainder (in a bleed stream) being treated to remove the compounds formed during the removal of  $\text{SO}_2$ .

The bleed stream is sent to a reactor where it is reacted with magnesium oxide to precipitate magnesium sulfite hexahydrate and sulfate, which in variation #1 is accomplished in the scrubber. The hexahydrate is converted to trihydrate, then dried and calcined with the sulfate to regenerate the absorbent and produce  $\text{SO}_2$  for further processing (as in variation #1).

With the magnesia aqueous scrubbing methods, the regeneration of  $\text{MgO}$  and liberation of absorbed  $\text{SO}_2$  can take place on the power plant site, or off-site at a central processing site.

A central processing site<sup>10,11</sup> would receive the waste solids (from the scrubbing step) from a number of surrounding plants and would perform the regeneration step for them. The regenerated  $\text{MgO}$  would be shipped back to the plants and the liberated  $\text{SO}_2$  would be processed to the final product(s) at the central site.

Advantages of magnesia scrubbing<sup>10,11</sup> are (1) a useful by-product can be produced, (2) the ability to regenerate and recycle the absorbent, magnesium oxide, (3) the ease of separation of the sulfite salts formed from the scrubber liquor, (4) the avoidance of a solids disposal problem, (5) the capability of separating, both financially and operationally, the power unit scrubbing system from the chemical manufacturing and marketing function, (6) adaptable to existing units, and (7) minimum effect on the power plant.

The disadvantages are (1) a loss of plume buoyancy thus requiring reheat (characteristic of many other aqueous scrubbing processes, (2) possibility of scaling, (3) a water pollution problem can be created if the effluent from the particulate scrubber is not treated, and (4) the

removal process consumes energy which reduces the overall efficiency of the power plant.

A 150 mw demonstration plant for the MgO process is currently being tested at Boston's Edison Mystic Station on an oil-fired boiler.<sup>18</sup> The sulfur content of the oil ranges from 2-2.5 percent. To date, the plant has run intermittently for a total period of approximately 55 days. The tests have shown SO<sub>2</sub> removal efficiencies up to 90 percent for 2.1 percent sulfur oil. Over 2,000 tons of commercial grade sulfuric acid has been produced and marketed from the regeneration plant. Many of the early problems dealing with the formation of crystalline magnesium sulfite, recycled magnesia, pH control and unreacted MgO have been solved. During the last series of tests, the MgO was recycled approximately four times with only a slight reduction in the SO<sub>2</sub> removal efficiency.

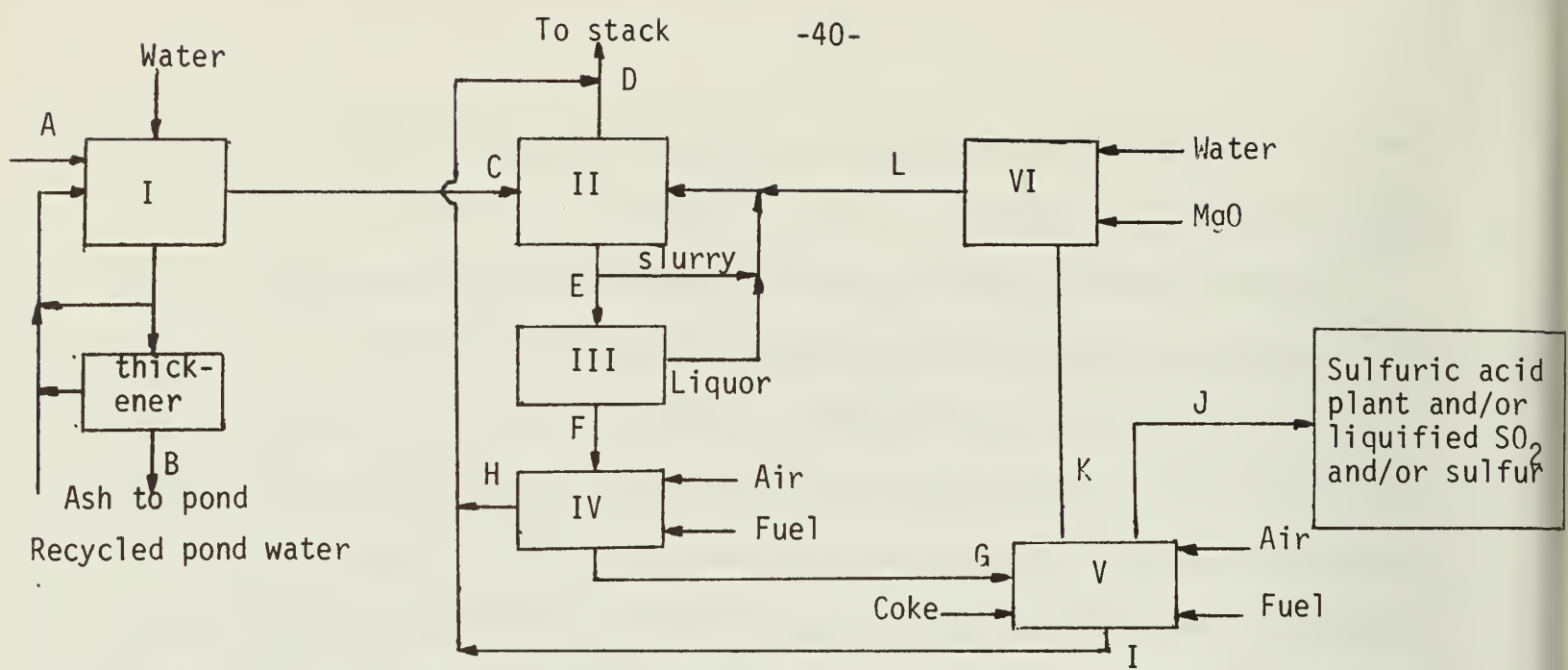
Because of the high efficiency fly ash removal capabilities of many electrostatic precipitators on Venturi scrubbers, it is felt that the demonstration of the MgO process on the Boston Edison Mystic Station oil-fired plant has direct application to coal burning installations.

### Mass Balance

The flow chart and the chemical reactions associated with the magnesium oxide scrubbing process are shown in Figure 6.

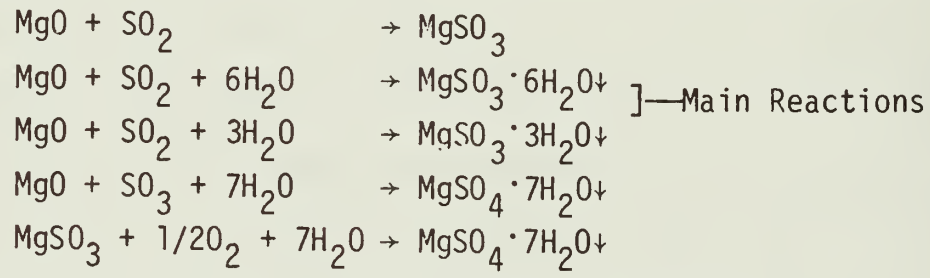
There are two general assumptions which apply to the mass balance. They are (1) the SO<sub>2</sub> removal system is in an operating equilibrium i.e., the removal of SO<sub>2</sub> from the flue gas does not disturb this equilibrium as the input rate of sulfur is equal to the output rate of sulfur, and (2) the reaction equations presented are overall reaction equations of the most important products and represent a hypothetical chain of events.

For the purpose of calculation the additional assumptions made include a 90 percent removal of SO<sub>2</sub> on 80 percent removal of SO<sub>3</sub> and a 2 percent oxidation rate in the scrubber.



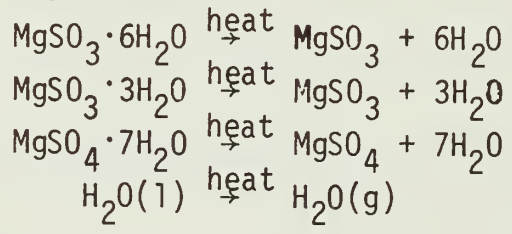
I. Particulate removal system

II. SO<sub>2</sub> absorber - venturi scrubber

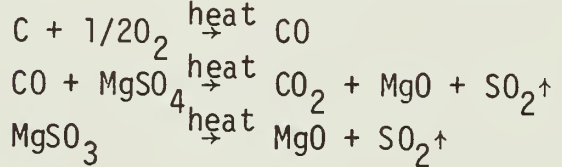


III. Dewatering system - Solids separated from the slurry

IV. Dryer system



V. Calciner system



VI. MgO tank

MgO brought into system

Figure 6 Magnesia Scrubbing Process Description

The results of the calculations are given in Table 12 and Table 13.

### Catalytic Oxidation

One process alternative to stack gas scrubbing which has been developed to the demonstration stage is the catalytic oxidation of the  $\text{SO}_2$  in the flue gas to product sulfuric acid or ammonium phosphate.<sup>12,13,14</sup>

In this process, flue gas sulfur dioxide is catalytically oxidized to sulfur trioxide. The sulfur trioxide reacts with water vapor present in the gas stream to form sulfuric acid, which is removed from the gas stream by condensation.

The converter unit of the process, in which the oxidation of sulfur dioxide occurs, requires a gas temperature which is higher than temperatures normally encountered in flue gas streams existing a boiler. This temperature requirement can be satisfied by one of two process variations: the integrated system or the reheat system.

The integrated system receives the flue gas for treatment directly from the boiler; this initial high temperature ensures a correct temperature for the oxidation reaction in the converter. At installations where the flue gas is not obtainable at a sufficiently high temperature, it must be reheated before it enters the converter; this is the reheat system.

With the integrated system, particulates are first removed (1.I.), then the sulfur dioxide is oxidized in the converter (1.II.), heat is recovered (1.III., 1.IV.), and the sulfur trioxide and water vapor are condensed in an absorbing tower (1.V.). This system is intended for use on new installations or retrofit installations which are modified to obtain the flue gas directly from the boiler.

Table 12

<u>Component</u>	<u>lbs /100 lbs of coal burned</u>	<u>lbs /lb of MgO input</u>
MgO (input)	0.09	1.0
C (input)	0.063	0.7
SO <sub>2</sub> (output)	6.683	74.26
Flyash	11.1	-

Material Balance - MgO Removal System  
(lb. per 100 lb of coal burned)

Component Stream	N <sub>2</sub>	CO <sub>2</sub>	H <sub>2</sub> O	O <sub>2</sub>	SO <sub>2</sub>	SO <sub>3</sub>	NO <sub>x</sub>	Ash	MgSO <sub>3</sub> ·6H <sub>2</sub> O	MgSO <sub>3</sub> ·3H <sub>2</sub> O	MgSO <sub>4</sub> ·7H <sub>2</sub> O	MgSO <sub>3</sub>	MgSO <sub>4</sub>	MgO	C
A	773.146	224.752	19.350	39.136	7.232	0.720	0.54	11.2							
B						-0.36		-11.1							
C	773.146	224.752	19.350	39.136	7.232	0.36	0.54	0.1							
D	773.146	224.752	19.350	39.136	0.722	0.144	0.54	0.1							
E			-8.288	-0.036	-6.51	-0.216			+9.004	+9.004	+1.218			-4.176	
F									9.004	9.004	1.218				
G									-9.004	-9.004	-1.218	+10.344	+0.594		
V.		+0.218		-0.079	+6.683							-10.344	-0.594	+4.086	+0.06
H & I	X	X													
J					-6.683										
K															4.086
L															+0.09
															4.176

-43-

X indicates combustion products at dryer and calciner

The reheat system is similar to the integrated system except for the reheat burners (2.II.) required to provide the correct temperature for the converter, and the position of the economizer. This system is intended for use on retrofit installations where the economizer follows the boiler. The heat recovery in the economizer leads to the requirement of heat addition with reheat burners to raise the temperature for the reaction in the converter.

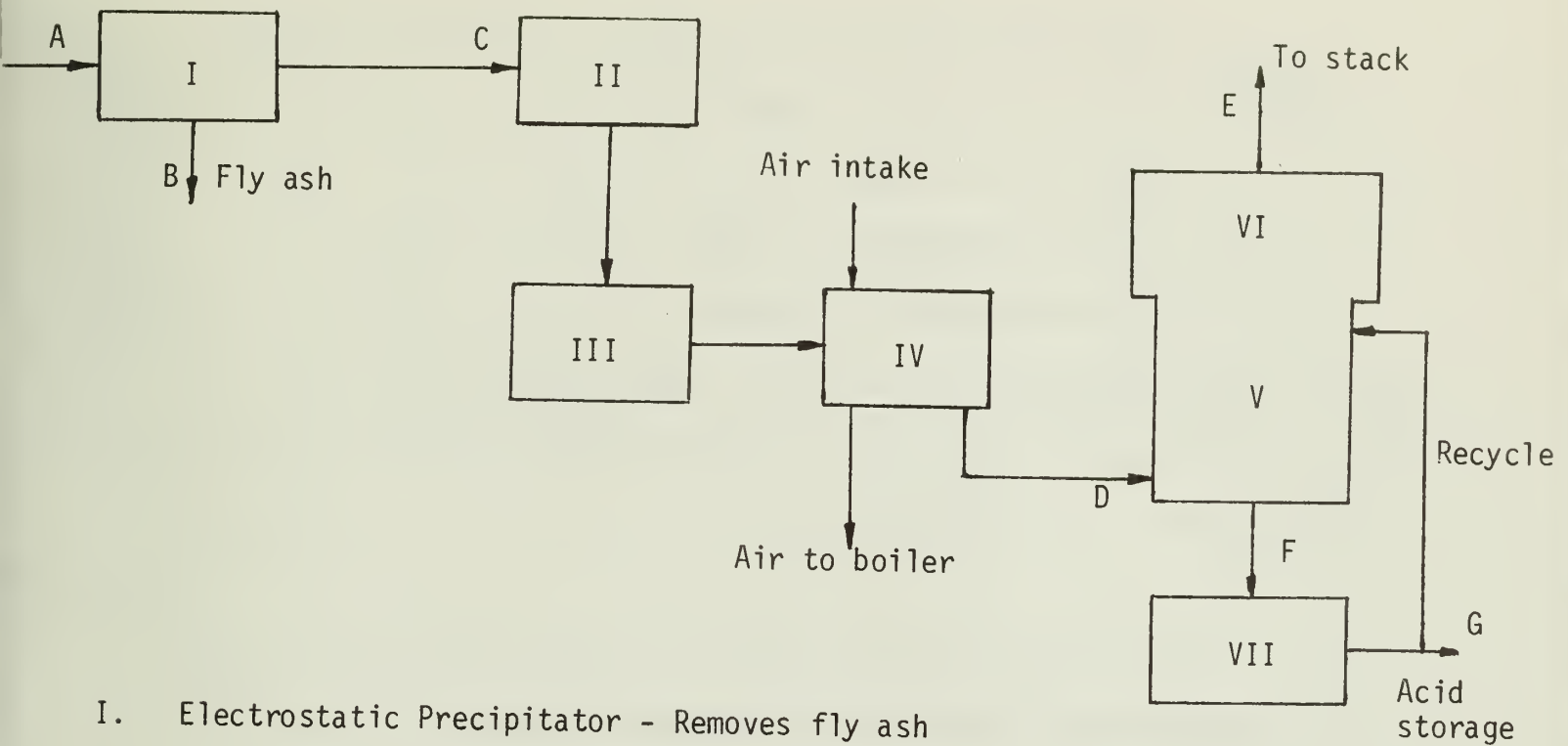
The advantages of this process are: (1) a useful by-product is realized directly from the abatement process without further processing and (2) it is attractive for high sulfur fuel. The disadvantages of this process are: (1) it is very difficult to retrofit on existing plants, (2) the catalyst can become fouled if the precipitator does not work efficiently, (3) if a sulfuric acid market is not available, the acid would have to be neutralized and disposed of.

The catalytic oxidation process has an impressive performance record to date. The Wood River installation of Illinois Power has a Cat-Ox process installed on a 100 mw plant. The initial tests have indicated that the system is capable of operating with a gas flow of  $1.12 \times 10^6$  lb/hr, producing 77.7 percent  $H_2SO_4$ , and achieving an 85 percent removal efficiency for  $SO_2$ . A company spokesman has said that "...it is evident that the system can remove at least 85 percent of the  $SO_2$  from the flue gas as 78 percent sulfuric acid..".<sup>19</sup> Over 1000 tons of  $H_2SO_4$  have been sold for fertilizer manufacturing to date and has been found acceptable.

### Mass Balance

The flow charts and chemical reactions associated with the catalytic oxidation  $SO_2$  removal processes are given for integrated and reheat configurations in Figure 7 and 8 respectively.



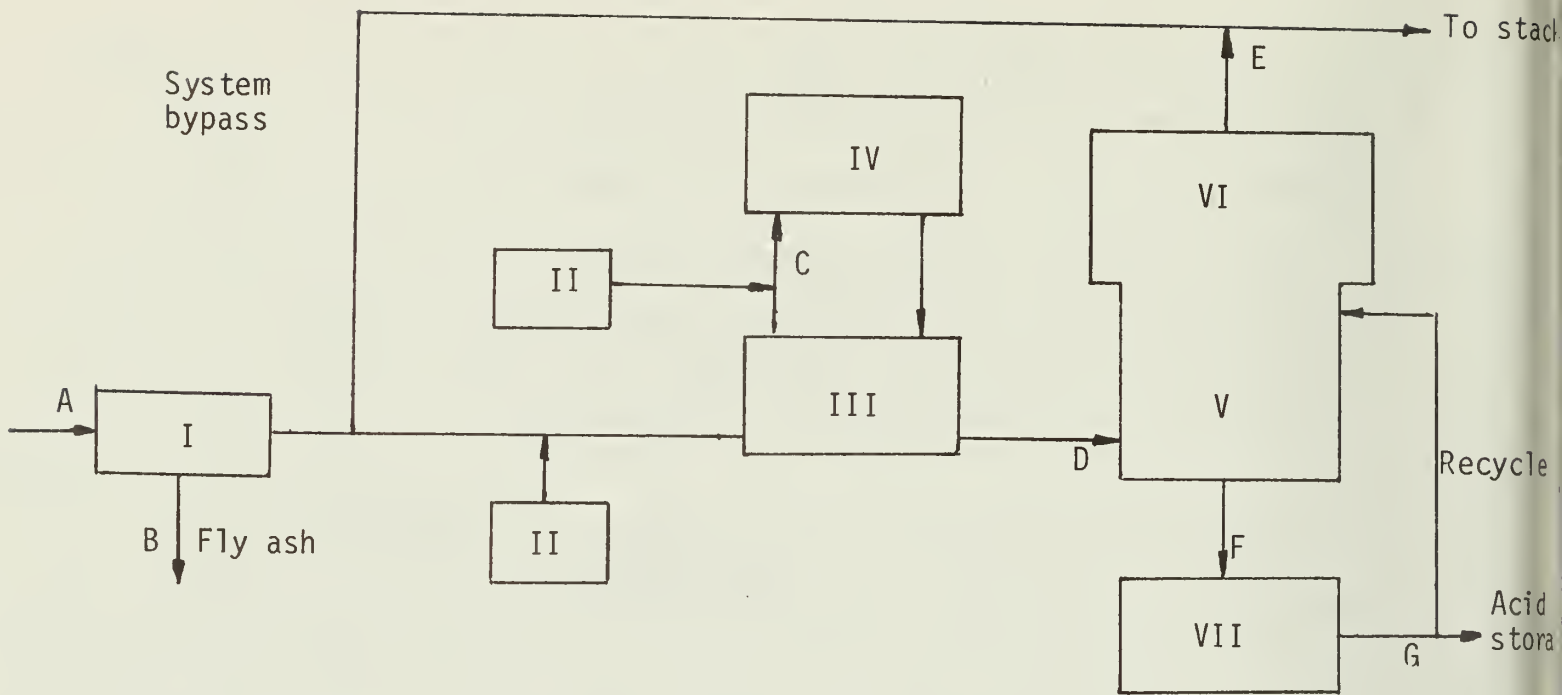


- I. Electrostatic Precipitator - Removes fly ash
- II. Converter - Temperature of the flue gas stream is about 850°F.  

$$\text{SO}_2(\text{g}) + 1/2\text{O}_2(\text{g}) \rightarrow \text{SO}_3(\text{g}),$$
 in the presence of a vanadium pentoxide catalyst.
- III. Economizer - cools the flue gas stream
- IV. Air heater - cools the flue gas stream
- V. Absorbing tower - Further reduces flue gas temperature, to about 250°F. During cooling, the  $\text{H}_2\text{O}$  and  $\text{SO}_3$  combine to form sulfuric acid which is subsequently condensed.  

$$\text{SO}_3(\text{g}) + \text{H}_2\text{O}(\text{g}) \rightarrow \text{H}_2\text{SO}_4(\text{g}) \rightarrow \text{H}_2\text{SO}_4(\text{l})$$
- VI. Acid mist eliminator - Removes acid mist particles and entrained droplets of circulating acid from the flue gas stream exiting V.
- VII. Acid cooler - Provides cool acid for storage and a recycle stream to V. to help cool the hot flue gas.

Figure 7 Integrated Cat-Ox System



- I. Electrostatic Precipitator - Removes fly ash
- II. Reheat burners - Raises flue gas temperature to 850°F. This is a requirement for the oxidation reaction in IV.
- III. Gas heat exchanger - Transfers heat from the oxidized flue gas stream (from IV.) to the incoming untreated flue gas stream (from I.).
- IV. Converter -
 
$$\text{SO}_2(\text{g}) + 1/2\text{O}_2(\text{g}) \rightarrow \text{SO}_3(\text{g}),$$
 in the presence of a vanadium pentoxide catalyst.
- V. Absorbing tower - Further reduces flue gas temperature, to about 250°F. During cooling, the  $\text{H}_2\text{O}$  and  $\text{SO}_3$  combine to form sulfuric acid which is subsequently condensed.
 
$$\text{SO}_3(\text{g}) + \text{H}_2\text{O}(\text{g}) \rightarrow \text{H}_2\text{SO}_4(\text{g}) \rightarrow \text{H}_2\text{SO}_4(\text{l})$$
- VI. Acid mist eliminator - Removes acid mist particles and entrained droplets of circulating acid from the flue gas stream existing V.
- VII. Acid cooler - Provides cool acid for storage and a recycle stream to V. to help cool the hot flue gas.

Figure 8 Reheat Cat-Ox System

There are two general assumptions which apply to the mass balance. They are (1) the  $\text{SO}_2$  removal of  $\text{SO}_2$  from the flue gas does not disturb this equilibrium as the input rate of sulfur is equal to the output rate of sulfur, and (2) the reaction equations presented are overall reaction equations of the most important products and represent a hypothetical chain of events.

For the purpose of calculation the additional assumptions made include a 90 percent conversion of  $\text{SO}_2$  to  $\text{SO}_3$  and a 100 percent conversion of  $\text{SO}_3$  to sulfuric acid.

The process yields 18.85 lbs of  $\text{H}_2\text{SO}_4$ /100 lbs of coal burned.

The results of additional calculations are shown in Table 14.

#### Economic Impact of $\text{SO}_2$ Regulations

It was estimated recently that price increases directly attributable to pollution control will be 2 percent by 1975 and 1-1.5 percent by 1978.<sup>16</sup> Of fourteen industries which were studied, it was estimated that 200 to 300 plant shutdowns will occur in the 1970-80 time frame. This is expected to affect 50-150 communities. The plants shut down, however, will be those which are currently marginal operations. The enforcement of the law will merely hasten the demise of these industries. Unemployment is projected to total between 50,000 and 125,000. This is approximately 1-4 percent of the employment in the affected industry. Only 0.05 percent of the total national labor force is expected to lose their jobs due to  $\text{SO}_2$  related legislation.<sup>16</sup>

During the 1974 national hearings<sup>2</sup> on  $\text{SO}_2$  scrubber systems it was estimated that the impact of scrubber costs could be as much as 15 to 20 percent where a company only had one plant. The nationwide average price increase was estimated in the same hearings to be 3 percent by 1980.<sup>2</sup> It is assumed

Table 14

Material Balance - Catalytic Oxidation Process  
(1b per 100 lb of coal burned)

Component Stream	N <sub>2</sub>	CO <sub>2</sub>	H <sub>2</sub> O	O <sub>2</sub>	SO <sub>2</sub>	SO <sub>3</sub>	NO <sub>x</sub>	ASH	H <sub>2</sub> SO <sub>4</sub>
A	773.146	224.752	19.350	39.136	7.232	0.72	0.54	11.2	
B								-11.1	
C	773.146	224.752	19.350	39.136	7.232	0.72	0.54	0.1	
D	773.146	224.752	19.350	37.506	0.722	8.86	0.54	-0.1	
E	773.146	224.752	17.360	37.506	0.722	-	0.54	-	
F									+10.85
G									-10.85

that the capital cost of a scrubbing system will be \$60/kw and the inflation rate will be 7.5%, the increase in industry's capital requirement for 1980 will be 4%.<sup>2</sup>

The average coal sulfur content of coal purchased in 1970 was 2.5 percent.<sup>16</sup> This percentage is expected to decrease to about 1.1 percent by 1975.<sup>16</sup> If state regulations are enforced in 1975, it is estimated that more than  $150 \times 10^6$  tons of coal for steam generation currently being produced could not be legally burned unless sulfur control technology is used.

One consequence of the sulfur regulations could be the continued importation of oil. If sufficient oil is imported to meet the regulations, this could lead to a dislocation in the coal industry which could displace 1/4 to 1/3 of current production, with a potential loss of up to 26,000 jobs in coal mining.<sup>16</sup>

In reviewing the economics of SO<sub>2</sub> tailend control devices, it is agreed at the outset that each process application is a special case and that the uniformity of analysis associated with the economic projections of each process developer cannot be guaranteed. In reviewing all the cost projections, however, certain unmistakable trends appear.

Table 15 gives the projected and actual (when available) capital and operating costs associated with the process classes of interest.

More recent projections for retrofit systems indicate that capital costs are in the range of \$80-\$90/kw, while new systems are costing about \$75/kw for 100 mw installations. One retrofit, however, reported a high of \$108/kw. If one counts both the capital and operating costs for scrubbers, it is found that the apparent cost of coal is increased \$5 to \$6 per ton of coal

Table 15

<u>Capacity (MW)</u>	<u>Capital Cost \$/kw</u>	<u>Annual Cost mills/kw-hr</u>	<u>Basis</u>	<u>Reference</u>
<u>Lime/Limestone systems-retrofit</u>				
80	84	-	Estimate	23
80	80	-	Estimate	23
100	80	4.1	Estimate	21
135	68	-	Estimate	3
135	64	-	Estimate	3
155	108	10	Experience	2
170	50	-	Estimate	2
200	52	2.4	Estimate	2
300	61	3.2	Estimate	21
380	70-75	2.1	Estimate	2
470	62	4	Estimate	2
500	49	-	Estimate	23
500	46	-	Estimate	23
500	51	2.9	Estimate	21
550	49	-	Estimate	3
550	76	-	Estimate	2
615	62	-	Estimate	2
615	47	4-6	Estimate	2
670	45-60	-	Estimate	2
800	43	2.6	Estimate	21
880	68	3.2	Estimate	2
880	66	3.7	Estimate	2

<u>Capacity (mw)</u>	<u>Capital Cost \$/kw</u>	<u>Annual Cost mills/kw-hr</u>	<u>Basis</u>	<u>Reference</u>
<u>Lime/Limestone systems-new plant</u>				
820	42	-	Experience	2
840	72-86		Estimate	2
1000	35	1.5	Estimate	2
1600	93	2.7	Estimate	2
<u>Magnesium Oxide systems</u>				
100	100	4.2	Estimate	21
155	39	-	Experience	2
300	73	3.1	Estimate	21
500	62	2.7	Estimate	21
800	53	2.4	Estimate	21
<u>Sodium system (Double Alkali)</u>				
80	88	-	Estimate	23
80	67	-	Estimate	23
100	85	4.0	Estimate	21
135	63	-	Estimate	23
135	51	-	Estimate	23
300	62	3.1	Estimate	21
500	52	2.7	Estimate	21
500	42	-	Estimate	23
500	35	-	Estimate	23
800	45	2.4	Estimate	21

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for new systems and \$7 per ton for retrofits. The expected price increase in electricity for utilities using scrubbers is 2 to 3.5 mils per kwh (1973 prices).

Another cost of interest is the sludge disposal costs for throwaway processes. For onsite disposal, the estimated costs are \$1 to \$2 per ton where as for offsite disposal the cost rise to \$12-15 per ton.<sup>24</sup>

The Monsanto Cat-0x process has reported actual capital costs of \$77/kw based on a 110 mw gross capacity. The estimated operating and maintenance costs of this process are \$2.18 per ton of coal after the initial de-bugging period.

A Radian<sup>20</sup> report suggests that the age and size distributions of coal fired generating units are very important in regard to the utilization. Another factor which is important is the load factor. The load factor is defined in this report as the annual heat input to the unit divided by its design annual input. The Radian<sup>20</sup> report states that the annual load factor for boilers less than 25 years old ranges from 50 to 85 percent. The total load factor for the entire group is 70 percent. Generally, load factors ranging from 10 to 55 percent can be assigned to boilers older than 25 years. An average load factor for this group is approximately 35 percent. The differential load factors between new and old boilers exists because the new boilers are significantly more efficient than those constructed before about 1950. For obvious economic reasons, the utility will use old units during periods of peak capacity demand. A comparison of relative cost of control of new or old units in the state of Ohio is given in Table 16. These results should be typical of most midwestern states.

Tables 17 and 18 give the distribution of costs associated with both capital and operating costs for some typical FDS. Since the cost estimates are for different size installations and for different years, they are not

Table 16

Typical Relative Costs of Retrofitting  
Existing Boilers in Ohio (Ref. 20)

<u>Boiler Age Group (Years)</u>	<u>0 - 10</u>	<u>11 - 20</u>	<u>21 - 30</u>	<u>31 - 40</u>
Retrofittable Capacity				
Relative to 0 - 10 yr. Unit	1.00	.64	.21	.045
Ave. Size (mw)	540	170	66	60
Useful Operating Life (yrs.)	35	25	15	5
Annual Operating Hrs.	6000	6000	4725	2900
Sulfur Content Coal, %	3.33	3.33	3.33	3.33
Ave. Capital Investment				
Relative to 0 - 10 yr. Unit	1.00	1.30	1.50	1.50
Annual Costs of Pollution				
Control Relative to 0 - 10 yr.				
Unit	1.00	1.67	3.69	12.10

Table 17

## Distribution of Estimated Capital Costs for Selected FDS

Process Category	Limestone Wet Scrubbing 500 mw (1972)	Double Alkali Scrubbing 500 mw (1974)	Magnesia Scrubbing 500 mw (1972)	Sodium (W-L) Scrubbing 500 mw (1975)	Catalytic Oxidation 100 mw (1974)
I. Gas handling area: fans, reheat system, scrubbers, duct work	3,401,000	3,072,000	4,483,000	7,695,000	
II. Chemical System	293,000	2,808,000	2,734,000	6,840,000	
III. Sulfuric acid plant	-	-	2,877,000	2,565,000	
IV. Bulk materials: steel, concrete, piping, electrical & instrumentation work, paint, optional bypass duct, etc.	2,777,000	2,521,000	3,411,000	7,600,000	
V. Construction Costs: labor, indirect field costs, construc- tion fees	5,513,000	5,800,000	3,796,000	8,740,000	
VI. Contractor's Costs: design engineering, fees, contingency, start-up etc.	3,800,000	3,300,000	6,361,000	4,560,000	
VII. Interest during construction	629,000	-	865,000	-	
VIII. Solids Disposal	934,000	-	-	-	
IX. Yard Improvements	375,000	-	350,000	-	
Totals:	17,722,000	17,501,000	24,877,000	38,000,000	8,500,000

Table 18

## Distribution of Estimated Operating Costs for Selected FDS

Cost Category	Process	Limestone Wet Scrubbing 500 mw (1972)	Double Alkali Scrubbing 500 mw (1974)	Magnesia Scrubbing 500 mw (1972)	Sodium (W-L) Scrubbing 500 mw (1975)	Catalytic Oxidation 100 mw (1974)
Direct Costs:						
I. Chemicals conversion costs -		\$ 394,600	\$2,193,378	\$ 137,000	\$ 956,000	\$ 30,000
II. Operating Labor & Supervision		139,700	638,074	235,200	216,000	36,000
III. Utilities		542,700	1,316,027	1,181,200	1,500,000	736,000
IV. Maintenance		1,149,100	438,676	1,478,800	842,000	-
V. Laboratory (Analyses)		38,000	-	85,000	72,000	-
VI. Misc. Supplies		-	-	-	45,000	-
Subtotal		2,264,100	4,586,155	3,117,200	3,631,000	962,000
Indirect Costs:						
VII. Capital Charges		2,140,200	1,236,268	3,721,500	2,712,000	-
VIII. Overhead		342,400	-	923,800	343,000	-
Subtotal		2,482,600	1,236,268	4,645,300	3,055,000	-
XI. On site solids disposal		629,600	Unknown	None	None	None
Total Annual Operating Costs		\$5,376,300	\$5,822,423	\$7,762,500	\$6,140,000 (no sulfur credit)	\$762,000 (no acid credit)

directly comparable. The tables do, however, give an indication of the relative expense of various components of FDS. Further, the rapid escalation of capital costs from 1972 to 1975 for comparable units is particularly evident in Table 17.

### Waste Disposal

One of the major environmental problems facing the throwaway stack gas cleaning processes is the ultimate disposal of the sludge. Both lime and limestone processes generate a large quantity of waste products. By way of example, the National Electric Reliability Council's Report<sup>24</sup> estimates that a single 825 megawatt generating unit will produce 3.63 tons of sludge per day for every megawatt of power which is produced. For this particular installation this would amount to more than 3000 tons of sludge each day which must be placed in a suitable repository. The most common disposal method currently being considered requires gravity settling in lined ponds followed by ultimate disposal in a landfill. Because the sludges are often thixotropic and may absorb water after a rainfall, any soluble salts or toxic elements in the sludge could be leached out and drained away from the disposal site. This, of course, could lead to serious environmental consequences. Based on the land requirements reported for fly ash disposal in the recent SOCTAP, the land requirement for sludge disposal is almost nine times that required for fly ash disposal. Given that 80 out of the 92 planned scrubber installations reported in the PEDCO study are lime or limestone based installations, this sludge disposal problem could be enormous.<sup>9</sup>

Perhaps the most effective way to alleviate this sludge disposal problem is to promote the development of regenerative processes and discourage the further sale of throwaway processes. The rationale behind such a strategy lies in the fact that elemental sulfur is the most desirable product for flue

gas desulfurization systems. Elemental sulfur is the preferred waste product because it can be economically stored for subsequent use at some future date, and is a relatively insoluble and inert material with no apparent major water pollution potential. The difficulty with this strategy is that of unfavorable economics.

### Space Requirements

While it is true that the ground space requirements for certain flue gas cleaning processes vary, the major flue gas scrubbing processes all require essentially the same equipment in the area immediately adjacent to the boiler stack. Since all of the components of the system used need not be located in this area, the overall space required for the installation is not significant for a retrofit installation.

A recent study by the Radian Corporation<sup>20</sup> examined the space requirements for lime/limestone, MgO and sodium based scrubbing processes. All of these processes are expected to have essentially the same space requirements in the area immediately adjacent to the boiler and stack.

The basic process equipment required for lime/limestone processes include the scrubber, mist eliminator, hold or delay tanks, solid separation devices, a reheat system, storage bins for the alkaline additive, slurry tanks and pumps and a solid disposal system. The main components of the waste disposal system include a clarifier or thickener, vacuum filter and a method of sludge fixation.

In order to size the area required to house the scrubber it was assumed that the gas velocity in the scrubber was 9.5 ft/sec, and that each scrubber handles 450,000 ACFM. These conditions define the scrubber area required which is 800 ft<sup>2</sup>. If a 4 sec gas residence time is assumed for the scrubber, the scrubber dimensions would be of the order of 20 ft in diameter and 40 ft

high. If a plant burning coal with a 3.5 percent sulfur content was limited to a sulfur emission rate of one (1) lb SO<sub>2</sub> per million BTU input, a 450,000 ACFM scrubber module would require a holding tank 50 ft in diameter and 55 ft high. In addition, a mist eliminator having a height twice that of the scrubber would be required on top of the scrubber module.

Following the example used in the Radian report<sup>20</sup> consider a 550 mw unit burning coal with a 3.5 percent sulfur content. If the holding tanks are placed below the scrubbers, there is a 15 ft space between the holding tanks, the pump houses are 7.5 ft from the holding tanks and the pump houses have dimensions of 30 ft by 60 ft, the total area required for the scrubber installation is approximately 25,000 ft<sup>2</sup> or 45 ft<sup>2</sup>/mw. If the scrubber size and pump house dimensions remain unchanged, and it is assumed that there is 15 ft between the scrubbers and 10 ft between the scrubbers and the pump houses, then the total area required is only 13,000 ft<sup>2</sup> providing the holding tanks are removed to some remote area or their size is reduced to fit beneath the scrubber units. This amounts to a space requirement of approximately 24 ft<sup>2</sup>/mw.

It should be noted that although the above space requirements were based on a 550 mw plant, the result is independent of plant size. The basis for this generalization lies in the fact that the scrubber area is proportional to flue gas flow rate. Since the flue gas flow rate and the size of the plant are also proportional, the ground space required for a scrubber is directly proportional to the size of the unit in mw. The space requirements would change, of course, if the coal sulfur content and heating value were changed.

The Radian Corporation<sup>20</sup> study assumed that the very minimum amount of space required adjacent to the plant was 20 ft<sup>2</sup>/mw. The impact of this type of load restriction for boilers in the State of Ohio was that about

74 percent of the total capacity surveyed in the State of Ohio had space equal to or greater than the 20 ft<sup>2</sup> mw minimum. Although this study was limited to the State of Ohio, there is no reason to believe it is not representative of the entire country.

In a similar study undertaken by the M. W. Kellogg Company,<sup>25</sup> a survey was conducted to determine the applicability of nine different SO<sub>2</sub> control processes to existing power plants based on space considerations. The study was limited to large (200 mw or greater) coal or oil-fired power plants. The nine processes included were limestone scrubbing (TVA), limestone injection (TVA), catalytic oxidation (Monsanto), molten carbonate scrubbing (Atomic International), sodium or potassium sulfite scrubbing (Wellman-Lord), magnesium oxide scrubbing (Chemico), formate scrubbing (Consolidation Coal), and ammonia scrubbing (TVA). The results indicated that, based on space requirements alone, none of the candidate processes could be installed in over 50 percent of the units studied. Further, the maximum theoretical space applicability ranged from approximately 60-70 percent for "throw-away" processes, down to 30-40 percent for the regenerable type processes which produce saleable by-products. The report noted that newer and larger plants could accommodate the processes better than older and smaller ones.

#### Time Requirements for Installation

Another important parameter in making economic predictions is the time requirement for installation of flue gas desulfurization. The recent hearing panel on sulfur dioxide controls<sup>2</sup> concluded that a reasonable time scale from the decision to control to compliance is broken down as follows: signed contracts in 6-9 months, construction begun 8-11 months, start up in 21-30 months, and compliance in 27 to 36 months. Industry representatives placed the overall period to be from 36 to 48 months.<sup>2</sup>



### Tie-In Requirements

The normal load cycle for electric power generation peaks in summer and winter due to the extreme temperatures encountered. The high winter peak load is due to space heating while the summer load peaks are due to space cooling. A typical generating plant is scheduled for routine maintenance at least once each year. The down time associated with the maintenance is of the order of one to three weeks. This time space would not be long enough to allow the installation of even the pre-assembled sulfur oxide removal processes. Once every four to five years a generating plant is down for a period of five to eight weeks for maintenance. This time span is considered sufficient<sup>3,20</sup> for the installation of most SO<sub>2</sub> scrubbing processes. Based on these figures, a recent report by SOCTAP<sup>3</sup> concluded that on the average a maximum 20 percent of the electrical generating capacity could be retrofitted in any one year. A report by Radian Corporation<sup>20</sup> seriously questioned whether this conversion percentage was realistic. The report stated that the percentage was too high because of the certainty of slippage in many retrofit installations. This is particularly true in the Midwest and East where brownouts have occurred during the past couple of years.

### Institutional Barriers

Institutional barriers can combine to delay the ordering, fabricating, assembling, and placing into operation an SO<sub>2</sub> scrubbing system. A list of the most important barriers are listed by SOCTAP:<sup>3</sup> 1) the adequacy of the market demand to encourage development of a supply industry; 2) the necessity to maintain adequate electrical reserve generation margin; 3) lack of process chemical expertise in the electrical utility industry; 4) fuel switching alternatives for higher costs of low sulfur fuels may be passed through to consumers by means of fuel adjustment clauses.

The things that are currently restricting the use of SO<sub>2</sub> systems include 1) lack of confidence in the ability of the vendors to perform as promised, 2) anticipation that regulations may be altered in the near future, 3) potential difficulties in raising capital and obtaining rate increases in covering expenses for pollution abatement, 4) the lack of suitably trained personnel in the industry to evaluate and operate these systems.

#### Power Requirements for Scrubbers

Recent reports<sup>2,24,20,3,22</sup> estimate that between 2 to 7 percent of the power output from boilers outfitted with SO<sub>2</sub> scrubbing systems are required to run the scrubbers. Energy requirements of 4 to 7 percent were reported in the recent EPA hearings on SO<sub>2</sub> scrubber technology held in Washington, D. C. in January of 1974.<sup>2</sup> In the SOCTAP<sup>3</sup> report assessing flue gas desulfurization systems, it was reported that the energy requirements to run TVA's Willow Creek No 8 plant, rated at 550 megawatts, was 24.5 mw. This amounts to roughly 4.5 percent of the total energy output of the plant. Gifford<sup>22</sup> in reporting on the Will County Unit No 1 of Commonwealth Edison estimated that the power requirement to run the limestone scrubber was 5.1 percent of the unit gross capacity. He noted that this is nearly equivalent to the auxiliary power consumed by the rest of the unit. If a National average of 5.5 percent energy penalty is used, EPA<sup>2</sup> estimates that the total electricity used by flue gas desulfurization systems in 1980 will be about 1 percent of the total electricity projected to be used in that year. Industry reports point out, however, that power companies do not have sufficient reserve capacities to supply this power. Since the Federal Power Commission requires that reserves of the order of 20 percent of the expected peak loads are necessary to avoid sporadic power curtailments, the installation

of FGD systems may be delayed.

### LOW BTU GASIFICATION

Another promising method for utilizing low sulfur fuels is the production of low BTU gas from coal. There are two advantages of this approach to the problem. First, the gas is produced under reducing conditions with the result that sulfur is converted into  $H_2S$ . This is an advantage because  $H_2S$  can be readily removed with existing technology at least at low temperatures. It should be noted, however, that the conventional processes for  $H_2S$  removal have the disadvantage of necessitating the gas to be cooled prior to treatment. This results in a considerable loss of heat and a lowering of the efficiency. This problem can be eased by the design of efficient heat transfer and recovery systems. The gas can also be generated at high pressures. This is advantageous because there is the possibility of using more advanced power cycles to generate power in new power installations.

It should be recognized that the production of low BTU gas is not a new technology. Long before the discovery and the ability to have long-range transmission of large volumes of natural gas, there were a number of local town gas facilities which produced a low BTU gas. In many parts of Europe this is still the case. So the technology to produce low BTU gas exists. But, its application to supplying a power plant with a varying load has not been demonstrated. Low BTU gases are generally acceptable fuels for gas, steam, and turbine power cycles and should be adequate for conventional power plants, although studies are needed to determine the lower acceptable limits in the heating value with regards to combustion characteristics and reduction of boiler ratings.<sup>31</sup>

Coal gasifiers can be divided into two principal configurations. These are processes in which the fuel is maintained in a fixed bed and where the

fuel is suspended in a gas. Suspension processes are categorized into fully entrained, fluidized bed systems and cyclone or vortex gasifiers.

The gas and solid flow may be countercurrent or con-current in the fixed bed processes. Countercurrent processes have found more industrial usage. In the suspension processes the particles may move with the gas, as in full entrainment; or they may move relative to it, which happens in vortex, cyclone and fluidized bed systems.

Counter-current gasifiers have a down flow of coal and an upflow of gas, and generally have high thermal efficiencies and good flexibility. They have, however, low gasification rates, small capacities, and excessive tar formation. Co-current gasifiers either up or down flow, have higher gasification rates and minimal tar formation because of higher temperatures but thermal efficiency is lower unless energy recovery is effected or the synthesis gas is cleaned and used at a high temperature. Fluidized bed properties are intermediate. Fluidized bed reactors can be made in large sized but the operating range is small. Process modifications such as recycle and waste heat recovery, can provide improvements. Examples of the fixed bed process and the three suspended processes are: The Lurgi, the Kopper-Totzek (entrained bed) Winkler (fluidized bed) and the Ruhrgas gasifiers (vortex bed), respectively.

It is of interest to give a brief description of three commercial low BTU gasification systems. Namely, the Lurgi, Kopper-Totzek and Winkler processes.

### Lurgi Process

The Lurgi process is the only fixed bed high pressure reactor which is commercially available. This is due to the difficulties faced in feeding reactors with coal under pressure.<sup>26</sup> There are two typical types of Lurgi processes, namely the Lurgi dry-ash and Lurgi slagging gasifiers. The latter is a modified Lurgi process which is designed to raise the process efficiency

and to make the system more suitable for utilizing caking coals.

The Lurgi dry-ash gasifier is a fixed bed pressurized generator which is made up of three main parts: the coal hopper, the reactor and the coal ash lock. Coal is fed from the top of the generator through the coal hopper and into the reactor. In the reactor the coal contacts a mixture of oxygen and steam (or air and steam) countercurrently. The oxygen and steam are injected into the lower region of the gasifier.<sup>26</sup> The mixture is proportioned to control the temperature in the reactor so that the formation of clinkers, due to the fusion of ash, can be avoided.<sup>27</sup> A mechanical stirrer keeps the coal and ash separated. The ash is continuously removed from the bottom through the ash lock after being separated by a revolving grate.<sup>26</sup> The fuel bed is heated by the rising gases and the volatile compositions of coal are vented as part of the raw gas from the top of the gasifier.<sup>27</sup> The range of the required operating conditions include a pressure range from 350 - 450 psi, a temperature range from 1,150 - 1,400°F and a residence time of almost one hour. The crude gas leaves the gasifier at a temperature of 700 - 1100°F and may contain tars, oil, naphta, phenols, and ammonia along with coal and ash dusts. A brief description of two commercial operated plants are summarized below:

a. Lurgi dry-ash commercial synthesis gas process (Scotland) - This autothermically and externally steam heated plant is supplying town gas, under 25 ATM pressure, at the rate of 310 BTU/scf. Highly volatile C bituminous coal is fed to the assembly at the rate of 49 lb/M scf of steam. The gasifier, which has a capacity ranging from 9,300 - 12,400 ft<sup>3</sup>, processes the gas at 11.9 M scf/ft<sup>2</sup> hr and uses the coal at the rate of 584 lb/ft<sup>2</sup> hr. The total heat input is 11,040 BTU/lb of fuel gasified and the total heat out is 6,230 BTU/lb. The gas analysis for this installation is given in Table 19.

Table 19

Gas Analysis for the Output from a Lurgi Dry-Ash Commercial Gas Process<sup>30</sup>

## Scotland Plant (25 atm.)

H <sub>2</sub>	37.4%
O <sub>2</sub>	negligible
CO	26%
CO <sub>2</sub>	26.1%
CH <sub>4</sub>	9.1%
other hydrocarbons	0.6%
N <sub>2</sub>	0.8%

## South Africa Plant (10-14 atm.)

H <sub>2</sub>	24.2%
CO	15.5%
CO <sub>2</sub>	15.0%
CH <sub>4</sub>	4.5%
N <sub>2</sub>	40.8%

## South Africa Plant (20 atm.)

H <sub>2</sub>	24%
CO	15%
CO <sub>2</sub>	15-17%
CH <sub>4</sub>	6%
N <sub>2</sub>	40-45%

b. Lurgi dry-ash commercial fuel gas process in South Africa is autothermally heated, and is pressurized to between 10 and 20 ATM. Bituminous coal along with steam and air are supplied to the generator and gas is produced having a HHV ranging from 140 - 170 BTU/scf. The generator has a capacity of 8,250 - 9,750 MM BTU/day and the gasification rate is 16.4 - 19.3 M scf/ft<sup>2</sup> hr. The gas composition analysis is given in Table 19 for operating pressures of 10 and 20 atmospheres. The heating value of the gas produced may be raised to 204 BTU/scf after being washed from CO<sub>2</sub> residual gases.<sup>28,29</sup>

Tables 20 and 21 give an output gas analysis, and a heat balance on a typical Lurgi dry-ash gasifier.<sup>30</sup>

Lurgi Slagging Gasifier: The thermal efficiency of the commercial type Lurgi process is reduced due to the presence of clinkering conditions in the gasifier. This reduced efficiency is due to the excess steam required to keep the ash below its fusion temperature. In addition, the extra non-decomposed steam causes the production of a considerable amount of dilute liquor which must be processed.

In order to raise the thermal efficiency, various investigators over the years have tried to operate at temperatures at which slagging occurs. This is a desirable condition for non-caking coals. Operation under these conditions not only raises the efficiency but also removes the necessity for grating the coal.

The heating value of this process is almost similar to the normal fixed bed gasifier. Methane formation is slightly reduced in the lower part of the gasifier and a higher partial pressure of hydrogen is realized due to an increased steam decomposition. It is estimated that the thermal efficiency of the slagging generator can be raised to 70 percent, which is approximately 10 percent higher than the commercial dry ash types.<sup>32</sup> Tables 22 and 23 give an

Table 20

Heat Balance for Lurgi Dry-Ash Gasifier<sup>30</sup>

<u>INPUT</u>	LBS.	C	H	O	S	N	ASH
Coal	1000.0	773.0	54.0	63.0	25.0	14.0	71.0
98% O <sub>2</sub>	502.7			492.7		10.0	
Steam	1815.0		201.7	1613.3			
Jacket Stm.	285.0		31.7	253.3			
Add'l Moist.	36.6		4.1	32.5			
<b>TOTAL IN</b>	<b>3639.3</b>	<b>733.0</b>	<b>291.5</b>	<b>2454.8</b>	<b>25.0</b>	<b>24.0</b>	<b>71.0</b>
<u>OUTPUT</u>							
Ash	74.9	3.1	0.4	0.4			71.0
CO <sub>2</sub>	1258.0	343.0		915.0			
CO	574.0	246.0		328.0			
H <sub>2</sub>	88.1		88.1				
CH <sub>4</sub>	168.0	126.0	42.0				
C <sub>N</sub> H <sub>M</sub>	8.8	7.5	1.3				
N <sub>2</sub>	14.2					14.2	
H <sub>2</sub> S	26.6		1.6		25.0		
Tar/Oil	51.3	43.2	4.7	3.4			
Phenols	5.5	4.2	0.4	0.9			
NH <sub>3</sub>	11.9		2.1			9.8	
H <sub>2</sub> O	1358.0		150.9	1207.1			
<b>TOTAL OUT</b>	<b>3639.3</b>	<b>773.0</b>	<b>291.5</b>	<b>2454.8</b>	<b>25.0</b>	<b>24.0</b>	<b>71.0</b>

Based on 1000 Pounds of Coal, Including 1.2% Moisture.



Table 21

Heat Balance for Lurgi Dry-Ash Gasifier<sup>30</sup>

<u>INPUT</u>	<u>BTU</u>
Coal Heating Value	13,990,000
Coal Sensible Heat	20,400
Coal Moisture Sensible Heat	2,500
Oxygen Sensible Heat	24,100
Steam Total Enthalpy	2,460,000
Jacket Water Sensible Heat	42,000
Solid Recycle Sensible Heat	-
Liquid Recycle Sensible Heat	-
Total In	<u>16,538,000</u>
<u>OUTPUT</u>	
Ash Sensible Heat	8,000
Ash Combustibles Heating Value	-
Hot Raw Gas Heating Value	13,263,000
Hot Raw Gas Sensible Heat	1,105,000
Water Vapor Total Enthalpy	2,140,000
Entrained Solids Sensible Heat	-
Jacket Steam Total Enthalpy	-
Heat Loss and Others	<u>22,000</u>
Total Out	16,538,000

Based on 1,000 Pounds of Coal, Including 1.2% Moisture.

Based Temperature, 32°F. (0°C.)

output gas analysis and a heat balance for this system.

A comparison of the process conditions for the two Lurgi configurations is given in Tables 24 and 25.

The basic advantage of Lurgi gasifiers are that they are commercially available. The disadvantages are (1) a low thermal efficiency, (2) limited to non caking coals for most applications, and (3) excessive tar formation during the gasification process.

### Kopper-Totzek Process

In the Kopper-Totzek process, a homogeneous mixture of coal and oxygen is injected from both sides of the reactor.<sup>30</sup> The opposite jet configuration creates a high degree of turbulence which increases the reaction rate significantly. The combustion reaction takes place very close to the injection point while the reduction reaction continues in the remaining space.<sup>30</sup> Due to the concurrent flow of the reactants, the hydrocarbons produced at low temperatures are passed into a very high temperature zone where they decompose rapidly. As a result, the coagulation of coal particles does not occur.<sup>33</sup> Some unreacted carbon and about 50 percent of the ash, entrained in the hot gas (1800-2300°F), leave the gasifier. This gas passes through the waste heat boiler, the dust collector, the wet scrubbing tower, and the disintegrator. The gas is then compressed, after it is put through a water separator, for future usage or further processes. The remaining ash, in the gasifier, is slagged and gets tapped off from the bottom side of the reactor. The water quenching is used to granulate the slag, and it is later removed through a water seal.<sup>30</sup>

Based on the resulting quality of the produced gases, this process uses almost 30 percent more oxygen than the Lurgi gasifier, for the equivalent amount of  $\text{CO} + \text{H}_2$ . The economic penalties associated with this usage are

Table 22

Material Balance for Lurgi Slagging Gasifier<sup>30</sup>

<u>INPUT</u>	LBS.	C	H	O	S	N	ASH
Coal	1000.0	773.0	54.0	63.0	25.0	14.0	71.0
98% O <sub>2</sub>	646.7			633.8		12.9	
Steam	463.0		51.4	411.6			
Lime	77.0						77.0
Ash Recycle	160.0						160.0
Add'l Moist.	36.6		41.1	32.5			
<b>Total In</b>	<b>2383.3</b>	<b>773.0</b>	<b>109.5</b>	<b>1140.9</b>	<b>25.0</b>	<b>26.9</b>	<b>308.0</b>
<u>OUTPUT</u>							
Ash	308.0						308.0
CO <sub>2</sub>	189.7	51.7		138.0			
CO	1393.7	597.3		796.4			
H <sub>2</sub>	50.2		50.2				
CH <sub>4</sub>	100.0	75.0	25.0				
C <sub>N</sub> H <sub>M</sub>	7.0	6.0	1.0				
N <sub>2</sub>	17.8					17.8	
H <sub>2</sub> S	26.6		1.6		25.0		
Tar/Oil	45.8	39.4	4.1	2.3			
Phenols	3.7	2.9	0.3	0.6			
NH <sub>3</sub>	11.0		1.9			9.1	
H <sub>2</sub> O	228.0		25.3	202.7			
Fatty Acids	1.8	0.7	0.2	0.9			
<b>Total Out</b>	<b>2383.3</b>	<b>773.0</b>	<b>109.5</b>	<b>1140.9</b>	<b>25.0</b>	<b>26.9</b>	<b>308.0</b>

Based on 1000 Lbs. of Coal, Including 1.2% Moisture

Table 23

Heat Balance for Lurgi Slagging Gasifier<sup>30</sup>

<u>INPUT</u>	<u>BTU</u>
Coal Heating Value	13,990,000
Coal Sensible Heat	20,400
Coal Moisture Sensible Heat	2,500
Oxygen Sensible Heat	31,000
Steam Total Enthalpy	627,000
Jacket Water Sensible Heat	-
Solid Recycle Sensible Heat	4,800
Liquid Recycle Sensible Heat	<u>-</u>
Total In	14,675,700
<u>OUTPUT</u>	
Ash Sensible Heat	252,500
Ash Combustibles Heating Value	-
Hot Raw Gas Heating Value	12,488,000
Hot Raw Gas Sensible Heat	775,000
Water Vapor Total Enthalpy	360,000
Entrained Solids Sensible Heat	-
Jacket Steam Total Enthalpy	-
Heat Loss and Others	<u>790,200</u>
Total Out	14,675,700

Based on 1000 Pounds of Coal, Including 1.2% Moisture.

Base Temperature, 32°F (0°C.)

Table 24

Estimated Performance for a Typical High - Volatile Weakly Caking Coal.<sup>30,31</sup>

	<u>Slagging</u>	<u>Dry ash</u>
Process steam (lb/therm of gas)	3.8 to 3.0	17.9
Oxygen (ft <sup>3</sup> therm of gas)	54 to 64	48
Steam: oxygen ratio (by vol)	1.5 to 1.0	8.0
Final reaction temperature (°C)	1250	750
Steam decomposition (per cent)	90 to 76	40
Gas yield (therms/ton d.a.f. coal)	262 to 257	267
Gas composition (per cent by vol)		
(i) Crude -		
CO <sub>2</sub>	4.7 to 7.0	28.5
CO	58.8 to 59.3	18.5
C <sub>2</sub> H <sub>4</sub>	0.2 to 0.2	0.2
H <sub>2</sub>	26.3 to 25.6	42.5
CH <sub>4</sub>	9.0 to 5.9	9.6
N <sub>2</sub>	<u>1.0 to 1.0</u>	<u>0.7</u>
	<u>100.0</u> <u>100.0</u>	<u>100.0</u>
(ii) Washed to 2 per cent of CO <sub>2</sub>		
CO <sub>2</sub>	2.0 to <b>2.0</b>	2.0
CO	60.4 to 62.5	25.3
C <sub>2</sub> H <sub>4</sub>	0.2 to 0.2	0.2
H <sub>2</sub>	27.1 to 27.0	58.4
CH <sub>4</sub>	9.3 to 7.3	13.1
N <sub>2</sub>	<u>1.0 to 1.0</u>	<u>1.0</u>
	<u>100.0</u> <u>100.0</u>	<u>100.0</u>
Gas calorific value (Btu/ft <sup>3</sup> )	375 to 360	400
By-products (therms/ton d.a.f. coal)	42 to 33	42
Liquor (lb/therm of gas)	0.9 to 2.1	12.0
Thermal efficiency of gas production		
(i) Excluding oxygen manufacture and carbon monoxide conversion	89.3 to 86.2	73.6
(ii) Including oxygen production but excluding carbon monoxide conversion	81.1 to 77.4	68.5
(iii) Including oxygen production and compression and carbon monoxide conversion	72.4 to 68.9	67.8

Table 25

The Process Data Summary for Normal and Slagging Lurgi Gasifiers<sup>30,31</sup>

<u>Item</u>	<u>Units</u>	<u>Fixed-bed Processes</u>	
		<u>Process 11</u>	<u>Processes 18,19,20</u>
		<u>Lurgi Dry-ash</u>	<u>Lurgi Slagging</u>
Coal (4.7% Moisture)	ton/day	12,860	13,410
Oxygen	ton/day	5,700	8,050
Cooling Water Circulated	gpm	263,000	356,000
Cooling Water Makeup	gpm	13,000	17,500
Boiler Feed Water Makeup	gpm	5,300	3,900
Fired Boiler Steam Production	M lb/hr	1,470	693
Total Steam Production	M lb/hr	4,765	3,103
Dry Gas to CO Shift	MM scfd	310	785
Dry Raw Synthesis Gas	MM scfd	883	802
Methane by Synthesis	MM scfd	133	166
CO <sub>2</sub> Removal	ton/day	16,050	18,500
Sulfur	ton/yr	49,000	59,000
Ammonium Sulfate	ton/yr	187,500	187,500
Raw Phenols	ton/yr	20,500	15,200
Char	ton/day	---	---
Overall Efficiency Btu in Gas/Btu in Coal	Percent	66.9	64.1

offset by the lower cost in the purification of the exhaust gases.<sup>34</sup>

This process is versatile in that it can gasify many types of fuel; char, petroleum coke, tars, heavy residuals, light to heavy oils, pumpable slurries of carbonaceous material in liquid hydrocarbon and all ranks of coal. The transferring from solid to liquid fuels require only the change of the burner heads. The turn down ratio can be varied easily due to access on multiple burners, and the reactor is flexible for closing down the operation instantaneously. Thirty minutes are required for start up. The mechanical and pressure sealing equipment are not complicated devices, and the screw feeders for solids or pumps for liquid feed stocks are the only moving parts designed. This simplicity results in a reduction in the amount of maintenance necessary between the periodic annual maintenances. The Kopper-Totzek gasifier has a high feed rate coal capacity of 850 tons per day, and has proven to be a safe operation for more than 20 years. The productive availability of the unit, which is also termed as the on-stream time, is above 95 percent.<sup>35</sup>

The reported overall thermal efficiencies for this process range from 67 to 75 percent.<sup>36,33,37</sup>

Another advantage is that the process is almost free from pollution. The resulting gases have no form of tars, condensable hydrocarbons or ammonia and phenol.

The gases produced are water quenched in order to remove the entrained slag droplets. When this type of gas is passed from the waste heat boiler, high pressure steam of up to 1500 psig can be produced. A high energy scrubbing system is used for cleaning the gas by removing the entrained solids to .002 - .005 grains/SCF, and lowering the temperature to almost 95°F.<sup>36</sup>

A commercial plant is currently being operated in Poland under atmospheric pressure utilizing pulverized Bituminuous coal. It produces a synthesis gas with a 640 BTU/lb heat content. The thermal efficiency is reported to be

72.4 percent.<sup>30,38,39</sup>

Tables 26 and 27 give an output gas analysis and a heat balance for the systems.

The process advantage of a Kopper-Totzek gasifier are (1) single construction, (2) operater on all coal types without pretreatment, (3) no oils or tars are produced, (4) commercially available and (5) good carbon utilization. The basic process disadvantage is large oxygen requirements.

The Winkler Process - The Winkler process is a fluidized bed process which was first proposed by Dr. Fritz Winkler of Germany, in 1922. The primary advantage of this gasifier in 1922 was it's ability to process brown coal.<sup>40</sup> It is the only fluidized bed commercially operated plant.<sup>41</sup> The first plant was erected at Leuna, Germany in 1926. There are currently about 36 plants operating at 16 different installations at this time. These are listed in Table 1.<sup>40</sup>

Fluidized beds are characterized by the upward flow of a fluid through a bed of small particles suspended in the fluid. Ideally, the temperature gradient across the bed will approach zero. The formation of agglomerates by the sintering of the ash, which may occur at temperatures as low as 1040°C, maintains the limiting point on the maximum rise of temperature in the fluidized bed processes.<sup>41</sup> Therefore the temperature in these processes ranges from 800° to 1000°C (or almost 1500 to 1850°F), relative to the reactivity of the fuel.<sup>42</sup> The reactor pressure in the Winkler process is almost atmospheric.

The process can be used for gasifying different types of the solid fuels. They may vary from lignite to coal, and of course some are more desirable than the others.<sup>40</sup> Therefore this process is mainly used for the gasification of sub-bituminous coal or its coke. Other types of the coal which may be



Table 26

Material Balance for the Kopper-Totzek Process<sup>30</sup>

<u>INPUT</u>	LBS	C	H	O	S	N	ASH
Coal	1000.0	773.0	54.0	63.0	25.0	14.0	71.0
98% O <sub>2</sub>	1020.0			1000.0		20.0	
Steam	500.0		55.5	444.5			
Total In	2520.0	773.0	109.5	1507.5	25.0	34.0	71.0
<u>OUTPUT</u>							
Ash	95.8	23.8				1.0	71.0
CO <sub>2</sub>	495.0	135.0		360.0			
CO	1431.0	613.0		818.0			
H <sub>2</sub>	66.4		66.4				
CH <sub>4</sub>	1.6	1.2	0.4				
C <sub>N</sub> H <sub>M</sub>							
N <sub>2</sub>	33.0					33.0	
H <sub>2</sub> S	26.6		1.6		25.0		
Tar/Oil							
Phenols							
NH <sub>3</sub>							
H <sub>2</sub> O	370.6		41.1	329.5			
Total Out	2520.0	773.0	109.5	1507.5	25.0	34.0	71.0

Based on 1000 Pounds of Coal, Including 1.2% Moisture.

Table 27

Heat Balance for the Kopper-Totzek Process<sup>30</sup>

<u>INPUT</u>	<u>BTU</u>
Coal Heating Value	13,990,000
Coal Sensible Heat	20,000
Coal Moisture Sensible Heat	-
Oxygen Sensible Heat	49,000
Steam Total Enthalpy	667,000
Jacket Water Sensible Heat	-
Solid Recycle Sensible Heat	-
Liquid Recycle Sensible Heat	-
Total In	14,736,000
<u>OUTPUT</u>	
Ash Sensible Heat	79,000
Ash Combustibles Heating Value	-
Hot Raw Gas Heating Value	10,530,000
Hot Raw Gas Sensible Heat	2,090,000
Water Vapor Total Enthalpy	895,000
Entrained Solids Sensible Heat	-
Jacket Steam Total Enthalpy	-
Heat Loss and Others	1,142,000
Total Out	14,736,000

Based on 1000 Pounds of Coal, Including 1.2% Moisture.

Base Temperature, 32°F. (0°C.)

used include brown coal, brown coal-char, semicoke, caking coal, and high rank brown coal. Grude which is a product char obtained from the low-temperature carbonization of lignite is also used. Examples of installations which use various coal types are the gasifiers which were built in Japan which use caking coal or semi-coke and the bituminous fed reactor in Spain.<sup>43</sup>

A list of Winkler plants is given in Table 28. The composition of the produced gas is tabulated in Table 29. Table 30 gives additional information.

The process advantages include 1) commercially available, 2) work on moderately caking coals, 3) can gasify high ash coal, and 4) the process does not produce tars or oils. The disadvantages are 1) the process is limited to atmospheric pressure operation and 2) the need to recycle entrained coal for good efficiency.

Table 31 gives the status of a number of low BTU gasification projects.

#### Advanced Power Systems - The Combined Cycle

An example of a more advanced power design that can be used for high pressure low BTU gasification is the combined cycle power plant.<sup>31,44,45</sup> In this configuration, the gas is expanded from a high pressure to a low pressure through a turbine-generator configuration which generates electricity. The gas is then burned to produce steam in a conventional boiler. Steam from the boiler is then used to operate steam turbines which when coupled to generators produce electrical output. The combination of the electricity generated from the gas turbine and steam turbine generator outputs gives rise to an increase overall plant efficiency. This configuration has the potential to increase the overall power plant efficiencies from 38 to 47%.

There are, however, some disadvantages in operating a low BTU gasification combined cycle installation. The disadvantages are

Table 28

Plant List - Winkler Generators.<sup>40</sup>

PLANT NO.	PLANT	YEAR	PRODUCT	CAPACITY PER GENERATOR			
				NORMAL		MAXIMUM	
				1000 NM <sup>3</sup> /HR	1000 SCFH	1000 NM <sup>3</sup> /HR	1000 SCFH
1	Leuna-Werk Leuna, Germany	1926-	Fuel Gas	60	2240	100	3730
		1930	Water Gas	30	1120	50	1870
2	Braunkohle-Benzin AG Bohlen, Germany	1936	Water Gas	27.6	1030	30	1120
3	Braunkohle-Benzin AG Magdeburg, Germany	1936	Water Gas	27.6	1030	33	1230
4	Yahagi Japan	1937	Water Gas	8.75	330	-	-
5	Braunkohle-Benzin AG Zeitz, Germany	1938	Water Gas	22.5	840	-	-
6	Dai-Nihonyinzo-Hiryo Japan	1938	Synthesis Gas	14	520	-	-
7	Nippon Tar Japan	1938	Water Gas	14	520	-	-
8	Toyo-Koatsu Japan	1939	Synthesis Gas	15	560	20	750
9	Sudetenlandische Treibstoffwerke Brux, Czechoslovakia	1943	Water Gas	27.6	1030	30	1120
*10	Fabrika Azotnih Jendinjenja Gorazde, Yugoslavia	1953	Synthesis Gas	5	190	-	-
11	Calvo Sotelo Puertollano, Spain	1954	Water Gas	9.5	350	-	-
12	Union Rheinische Braunkohlen Wesseling, Germany	1956	Synthesis Gas	12	450	17	630
13	Calvo Sotelo Puertollano, Spain	1957	Synthesis Gas	9.5	350	-	-
*14	Azot Sanyyii TAS Kutahya, Turkey	1959	Synthesis Gas	12	450	18	670
15	Neyveli Lignite Corporation Madras, India	1959	Synthesis Gas	41.6	1550	-	-
16	Union Rheinische Braunkohlen Wesseling, Germany	1960	Synthesis Gas	12	450	17	630

\* PRESENTLY OPERATING

Table 29

Raw Gas Composition for the Winkler Process<sup>35</sup>

RAW GAS COMPOSITIONS			
Component	High-BTU	Mo1%	Low-BTU
CO	25.7		19.0
CO <sub>2</sub>	15.8		6.2
H <sub>2</sub>	32.2		11.7
H <sub>2</sub> O	23.1		11.5
CH <sub>4</sub>	2.4		0.5
H <sub>2</sub>	0.8		51.1
H <sub>2</sub> S	2,500 ppm		1,300 ppm
COS	400 ppm		200 ppm
	<hr/>		<hr/>
Total	100.0		100.0
Higher heating value (dry basis) BTU/ccf	275		118

Table 30

Typical Winkler Product Gas Composition<sup>40</sup>

PLANT NO.	PRODUCT	COMPOSITION, VOLUME PERCENT DRY					HHV DRY BTU/SCF
		CO <sub>2</sub>	CO	H <sub>2</sub>	CH <sub>4</sub>	N <sub>2</sub>	
1*	Fuel Gas	10	22	12	0.7	55.3	116.6
1	Water Gas	19	38	40	2	1	271.8
1	Water Gas	20	38	39	1.5	0.5	263.8
6	Synthesis Gas	14	38	26	0.2	21.8	208.2
8	Synthesis Gas	12	34	31	0.5	22.5	214.6
10	Synthesis Gas	20	37	37	3	3	269.0
12	Synthesis Gas	13.8	48.2	35.3	1.8	0.9	287.3
13	Synthesis Gas	21.4	32.8	41.3	2.8	2	267.4
14	Synthesis Gas	19.9	35.5	40.0	2.8	1.8	271.8

\* REFERS TO PLANT NO. IN TABLE 28

Stage of Development of Typical Low BTU Gasification Processes

	Concept		Bench		Pilot		Demo.		Commercial		Coal Condition		HHV*	Type of Reactor				Operating Pressure									
	1	2	Const.	Test	3	4	Const.	Test	5	6	Const.	Test		7	8	Const.	Test	9	10	Caking	Non-Caking	Fluid Bed	Entrained Bed	Moving Bed	0-15 psig	100-500 psig	1000-1500 psig
Atgas	X	X	X	X	X	X													X					X			
Bi-Gas	X	X	X	X	X	X													X					X			
Hydrane	X	X	X	X	X	X													X					X			
CO <sub>2</sub> Acceptor	X	X	X	X	X	X													X					X			
Koppers Totzek	X	X	X	X	X	X													X					X			
Lurgi	X	X	X	X	X	X													X					X			
Molten Salt	X	X	X	X	X	X													X					X			
Synthane	X	X	X	X	X	X													X					X			
U-Gas	X	X	X	X	X	X													X					X			
Union Carbide	X	X	X	X	X	X													X					X			
Wellman Galusha	X	X	X	X	X	X													X					X			
Westinghouse	X	X	X	X	X	X													X					X			
Winkler	X	X	X	X	X	X													X					X			

\* Higher Heating Value (BTU/SCF)

1. Its operation is more like chemical plant than a power plant
2. It requires a high degree of control of flow of composition of most of the process streams.
3. It is more difficult to start up and shut down.
4. It requires more time to reach optimum conditions than an ordinary power plant.

As a result, special training of personnel and possible the addition of manpower with different skills, that is chemical plant experience, might be needed. This latter factor of personnel may make an installation too costly for small installations, but could possibly have great advantage in the case of large installations. Another problem is the need to start up and shut down the gasification facilities simultaneously with the power plant. This assumes, of course, that large scale storage of the low BTU gas is impractical. Because start ups and shut downs are known to cause difficulties in the chemical industry, it might not be desirable for utilizing low BTU gasifiers for peak operations. Rather, they should be used in plants that have a steady demand factor.

As noted above, it is necessary to operate the gasifier under pressure in a combined cycle configuration. Not only does this produce a more efficient power plant, but is also simplifies the design of the gasification process itself. It should be noted, however, that because of the problems in feeding coal through a system under pressure, only one high pressure process, the Lurgi process, has been commercialized. It currently has fourteen plants now operating around the world and is in the process of expanding its operation to a number of additional installations.

It is of interest to point out a few of the more important limitations of the Lurgi process. These limitations include<sup>31</sup>



1. The coal has to be carefully sized,
2. The process has difficulties in utilizing caking coals, and
3. The process is limited in terms of capacity.

The largest reactor vessel that has been built is only about 12 - 13 feet in diameter which makes it necessary to install multiple units for a plant designed to produce a large amount of gas. While it is recognized that a larger vessel would be desirable, one has not been built, at least not to date.

Although Lurgi is the only pressurized process available for immediate workable application, a number of new processes are in the development stage, with several more at the research and conceptual stage. Preliminary design and economic evaluation should make it possible for further development. It appears that given sufficient time for development, other processes will become economically superior to Lurgi.

### Economics

In assessing the economics of low BTU gasification systems two facts seem clear. First, system costs are increasing tremendously and second, there is a large discrepancy between cost estimates made by various organizations. This fact is particularly evident in Table 32 where cost estimates for retrofit low BTU gasification systems range from \$110/kw to \$498/kw.<sup>31,44,46,47</sup>

Table 33, a table prepared by IGT, gives a comparison of the costs for several different retrofit gasification configurations. Table 34 gives a comparison of the total capital investment associated with various gasification systems. Table 34 was prepared by TVA for EPRI. A comparison of the results from these two studies point up that honest disagreements exist between two organizations with considerable expertise in this area. Tables 35 and 36

give detailed cost breakdown information for the two studies.

The only conclusion that can be drawn, at this time, from these two cost estimates is that a detailed study into where the major disagreements occur is necessary. The identification of the sources and validity of the differences will be examined in the next phase of this study.

For new installations, IGT<sup>44</sup> reports costs of \$280/kw for an integrated power system which includes a low BTU gasifier operating in a combined cycle configuration. Table 37 gives a cost breakdown for the combined cycle facility. This cost compares favorably with the \$300/kw which is often quoted for conventional coal burning installations without provisions for stack gas cleanup if high sulfur fuels are used.

#### Waste Disposal and Environmental Control

The waste disposal problems associated with low BTU gasification processes do not appear to be materially different from those of coal fired units. The ash which is formed is comparable to flyash with regard to its pollution potential. The hydrogen sulfide is converted to elemental sulfur which is the best by-product from an environmental point of view. A list of hydrogen sulfide abatement systems is given in Table 38. Table 39 gives a summary of desulfurization processes proposed for coal gasification system.

#### Space Requirements

The space requirements for low BTU installations are not trivial. The sole advantage of the gasification system is that it can be constructed a short distance away from the power plant. This allows a certain flexibility in siting the installation. This area should be the subject of a detailed investigation.

Summary of IGT U-Gas Process Applications with Air- and Oxygen-Blown Gasification  
(Coal Rate: 7345.7 tons/SD, 181,996.8 X 106 Btu/SD; Coal Cost, 30¢/106 Btu)<sup>44</sup>

	Retrofit to Existing Power Plant		New Separate Facility		Combined Power Cycle	
	Air-Blown	Oxygen-Blown	Air-Blown	Oxygen-Blown	Air-Blown	Oxygen-Blown
Products						
Gas, 10 <sup>6</sup> Btu/SD (dry HHV)	129,000 (155)	140,000 (225)	129,000 (155)	140,000 (255)	---	---
Electric Power, kW	20,400	9,700	146,000	102,000	942,000	967,000
Turbine Steam, lb/hr	721,000	522,000	---	---	---	---
Overall Efficiency, % of coal HHV						
Coal to Gas	70.9	76.9	70.9	76.9	---	---
Coal to Power	0.9	0.4	6.6	4.6	42.4	43.5
Coal to Steam	13.1	9.5	---	---	---	---
	84.9	86.8	77.5	81.5	---	---
Total Capital Investment, \$10 <sup>6</sup>	66.2	73.1	90.3	94.4	216.0	231.1
Gross Annual Required Revenue, \$10 <sup>6</sup>	32.27	34.74	37.55	38.47	61.77	64.77
Total By-product Value-Steam, Power, Sulfur*-\$10 <sup>6</sup>	-7.02	-4.96	-9.84	-7.05	-0.62	-0.64
Net Annual Required Revenue, \$10 <sup>6</sup>	26.25	29.78	27.71	31.42	61.15	64.13
Net Annual Manufacturing Cost, \$10 <sup>6</sup>	21.99	26.09	22.03	25.48	47.83	49.90
20-Year Average Product Pricet						
Gas						
¢/10 <sup>6</sup> Btu	61.9	64.8	65.3	68.3	---	---
¢ above coal cost	31.9	34.8	35.3	38.3	---	---
Power						
mills/kWhr	---	---	---	---	8.2	8.4
80% load factor	---	---	---	---	9.0	9.2
Gas Manufacturing Cost						
¢/106 Btu	51.9	56.7	51.9	55.4	---	---
¢ above coal cost	21.9	26.7	21.9	25.4	---	---

\* High-pressure turbine steam at 90¢/1000 lb; power at 8 mills/kWhr; sulfur at \$10/long ton.

+ 90% Stream factor, A.G.A.-OCR Accounting Procedure.

Table 34

Total Capital Investment for Six Gasification Systems<sup>47</sup>

<u>System</u>	<u>Total capital investment, \$</u>
Wellman-Galusha/iron oxide	160,982,000
Wellman-Galusha/iron oxide/fines gasification	201,767,000
Lurgi/Benfield	211,510,000
Wellman-Galusha/Stretford	221,068,000
Lurgi/Stretford	234,807,000
Lurgi/iron oxide	234,876,000

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<sup>a</sup>Includes both Wellman-Galusha and Koppers-Totzek gasification units.

Table 35

Total Plant Investment for Retrofit Application of the IGT U-Gas  
Process to a Coal-Fired Power Plant<sup>44</sup>

	Air-Blown Gasification (Plant Capacity, $129.1 \times 10^9$ Btu/SD)	Oxygen-Blown Gasification (Plant Capacity, $140 \times 10^9$ Btu/SD)
Coal Storage	Existing	Existing
Coal Preparation	Existing	Existing
Solids Feed and Discharge System	\$2,360,000	\$2,360,000
Pretreaters	1,300,000	1,300,000
Gasifiers	6,970,000	4,130,000
Recycle Fines Cyclones and Feeders	300,000	220,000
Oxygen Plant and/or Pretreatment Air Compressor*	6,400,000	3,100,000
Second-Stage Pretreatment Air Compressor*	---	1,100,000
Oxygen Compressor*	---	1,650,000
Sulfur Removal and Recovery	7,200,000	6,030,000
Gasifier Steam Superheater	260,000	260,000
Gas Expander and Generator	16,290,000	10,830,000
Waste-Heat Boiler and BFW Pump	5,120,000	3,710,000
Oxygen Plant	---	16,650,000
Gas Plant Power Distribution, \$115/kW	1,480,000	1,570,000
Cooling and Makeup Water	430,000	430,000
General Facilities	<u>Existing</u>	<u>Existing</u>
Subtotal	48,110,000	53,340,000
Sulfur Removal Contingency at 20%	1,440,000	1,210,000
Plant Contingency at 10% (Excluding Sulfur Removal)	<u>4,090,000</u>	<u>4,730,000</u>
Total Bare Cost	53,640,000	59,280,000
Contractor Overhead and Profit at 7.73%	<u>4,150,000</u>	<u>4,580,000</u>
Subtotal	57,790,000	63,860,000
Interest During Construction at 7.5%	<u>4,330,000</u>	<u>4,790,000</u>
Total Fixed Investment	62,120,000	68,650,000
Working Capital	<u>4,100,000</u>	<u>4,430,000</u>
Total Investment	\$66,220,000	\$73,080,000

\* Driven by expansion turbine.

Table 36

Lurgi-Stretford System<sup>a</sup> Summary of total Capital Investment Requirements<sup>48</sup>

	Investment, \$	Percent of direct inv
Coal conveying and screening (conveyors, bins, feeders, screens, dust collectors, chutes, galleries and supports, belt scales, and hoppers)	1,272,000	0.9
Lurgi gasification (gasifiers, bunkers, conveyors, chutes, slide gates, dust collectors, pumps, compressors, ash locks, ash hoppers, depressurizing vessels, and ductwork)	45,858,000	33.5
Power unit modifications (fan capacity increased, superheater and economizer modified, burners and windboxes replaced, coal pipes and related equipment removed, ductwork, and structural supports)	5,550,000	4.1
Stretford H <sub>2</sub> S removal and sulfur recovery	21,096,000	15.4
Gas cleanup <sup>2</sup> and heat exchange (scrubbers, heat exchangers, internal coils, tanks, pumps, separators, coolers, and ductwork)	9,086,000	6.6
Tar removal	8,326,000	6.1
Phenosolvan waste water treatment	5,778,000	4.2
Expander-compressor motor set (expander-compressor sets, motors, and ductwork)	18,717,000	13.7
Utilities (instrument air generating and supply system, plus distribution systems for process steam, water, and electricity)	3,740,000	2.7
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	5,911,000	4.3
Construction facilities	6,514,000	4.8
Cooling pond (land, excavation, clay liner, pumps, and piping)	3,783,000	2.8
Byproduct storage (storage facilities for 30 days' production of sulfur, aqueous ammonia, and crude phenols)	1,181,000	0.9
Subtotal direct investment	<u>136,802,000</u>	100.0
Engineering design and supervision	16,416,000	12.0
Construction field expense	17,784,000	13.0
Contractor fees	9,576,000	7.0
Contingency	15,048,000	11.0
Subtotal fixed investment	<u>195,626,000</u>	143.0
Allowance for startup and modification	19,563,000	14.3
Purchased power during additional outage for tie-in <sup>b</sup>	840,000	0.6
Interest during construction	<u>15,650,000</u>	11.4
Total capital investment excluding raw materials and royalties	231,670,000	169.3
Raw materials	2,573,000	1.9
Royalties - Stretford process	<u>555,000</u>	0.4
Total capital investment	234,807,000	171.6

a. Basis:

Existing 500 MW coal-fired unit retrofitted with coal gasification system and derated to 475 MW.

Remaining life of power plant, 25 yr.

Coal gasified, 1,840,300 tons/yr; 10,800 Btu/lb; 11,955 Btu/kWh.

Existing ash disposal facilities assumed adequate.

Midwest plant location, 1975 average project cost date.

b. Transmission charge, \$0.001/kWh.

Table 37

Total Plant Investment for the Combined-Cycle  
Facility Application of the IGT U-Gas Process<sup>44</sup>

	Air-Blown Gasification (Power Output, 942 MW)	Oxygen-Blown Gasification (Power Output, 967 MW)
Coal Storage	\$1,160,000	\$1,160,000
Coal Preparation	2,420,000	2,420,000
Solids Feed and Discharge	2,360,000	2,360,000
Pretreater	1,300,000	1,300,000
Recycle Fines Cyclone Separators and Feeders	300,000	220,000
Oxygen Plant and Pretreatment Air Compressor*	6,400,000	3,100,000
Second-Stage Pretreater Air Compressor*	---	1,100,000
Sulfur Removal and Recovery	7,200,000	6,030,000
Gasifier Steam Superheater	260,000	260,000
Gas Turbine	80,600,000	81,300,000
Gas Turbine Effluent Waste-Heat Steam Cycle II	27,100,000	29,300,000
Steam Cycle I Turbine Generator	13,530,000	10,440,000
Oxygen Plant	---	16,650,000
Oxygen Compressor*	---	1,650,000
Gas Plant Power Distribution at \$115/kW (21,740 kW)	2,370,000	2,500,000
Cooling and Makeup Water	6,120,000	5,970,000
General Facilities	<u>5,000,000</u>	<u>5,000,000</u>
Subtotal	163,090,000	174,890,000
Sulfur Removal Section Contingencies (20%)	1,440,000	1,210,000
Plant Contingencies at 10% (Excluding Sulfur Removal)	<u>15,590,000</u>	<u>16,890,000</u>
Subtotal Bare Cost	180,120,000	192,990,000
Contractor's Overheat and Profits (7.73%)	<u>13,920,000</u>	<u>14,920,000</u>
Subtotal	194,040,000	207,910,000
Interest During Construction (7.5%)	<u>14,550,000</u>	<u>15,590,000</u>
Fixed Investment	208,590,000	223,500,000
Working Capital	<u>7,360,000</u>	<u>7,640,000</u>
Total Capital Investment	\$215,950,000	\$231,140,000

\* Data not available.

Table 38

Summary of Commercial H<sub>2</sub>S and Acid Gas Absorption Processes<sup>31,48</sup>

<u>Process</u>	<u>Developer</u>	<u>Absorbent</u>	<u>Pressure psi*</u>
Adip	Shell	Alkaloamine	50-500+
Alkaxid	Badisch Anilin	Potassium Salt	50-500+
Benfield	Benfield	Potassium Carbonate	100-2000
Catacarb	Eickmeyer & Assoc.	Pottasium Salt Solution	200-1000
Econamine	Fluor	Alkaloamine	~ 1000
Fluor Solvent	Fluor	Propylene Carbonate	~ 1000
Purisol	Lurgi	N-methyl pyrrolidone	1070
Rectisol	Lurgi	Methanol	685
Seloxol	Allied Chemical	Dimethyl ether of polyethylene glycol	1000
SNPA-DEA	Ralph M. Parsons Co.	Diethanolamine	600-1100
Sulfinol	Shell	Tetrahydrothiophene	~ 1000

\*will vary depending upon process and sulfur removal requirements.



Table 39

Summary of Desulfurization Processes  
Proposed for Coal Gasification Systems<sup>31</sup>

<u>Process</u>	<u>Desulfurization Process</u>
Lurgi (high Btu) (28)	Rectisol (low temperature methanol)
Lurgi (low Btu) (28)	Hot Potassium Carbonate
HYGAS (84)	Monoethanol Amine or Diglycol Amine
CO <sub>2</sub> Acceptor (84)	Hot Potassium Carbonate
Synthane (45)	Hot Potassium Carbonate
Bi-Gas (85)	Rectisol
Atgas (69)	Limestone

### Time Requirements for Installation

The time requirements for installation of a low BTU gasification system are comparable to those for stack gas scrubber systems.

### Tie-In Requirements

The tie-in requirements for a low BTU gasification system are comparable to those for stack gas scrubber systems.

### Institutional Barriers

The institutional barriers which were listed for stack gas scrubbing are generally applicable to low BTU gasification. The only major difference is in the genesis of the control strategy.

Stack gas scrubber strategies for controlling sulfur oxide emissions were put forth by United States Environmental Protection Agency. The reaction of the industrial community was that the systems were not reliable, were costly and offered no hope of contributing to the overall economic strength of the industry. Low BTU gasification strategies were basically the brain child of industry. In contrast to the FDS, industry viewed low BTU gasification as a costly alternative but one which has the potential to contribute to the economic growth of the industry over the long haul. For this reason, it is felt that low BTU gasification systems offer the best alternative to utilize high sulfur coals while at the same time improve our overall energy supply picture.

### Energy Efficiencies for Low BTU and Low BTU-Combined Cycle Gasification Systems

The thermal efficiency for producing low BTU gas is estimated to range from 70 to 80 percent.<sup>31,35,44</sup> Electric power generation efficiencies for conventional power plants are approximately 38 to 40 percent.<sup>31</sup> Combined gas, steam, turbine systems have potential for overall plant efficiencies

of up to 47 percent when high temperature gas turbines are developed.<sup>44,45</sup> The realization of these higher overall plant efficiencies in the future is not only attractive economically but necessary if our coal reserves are to be utilized in a more efficient manner.

#### COMPARISON OF STACK GAS CLEANUP SYSTEMS WITH LOW BTU-COMBINED CYCLE GASIFICATION SYSTEMS

It was noted in an earlier section that over 99 utilities have committed themselves in varying degrees to the installation of SO<sub>2</sub> stack gas cleanup devices. Eighty-two percent of these installations, amounting to approximately 29,000 mw are currently committed to lime or limestone based processes. The remaining applications are distributed more or less equally among the other processes; i.e., magnesium oxide, sodium hydroxide and catalytic oxidation systems. This suggests that the most advanced process, at least from the view point of industry, is the lime or limestone scrubber system.

In reviewing the low BTU gasification installation, it was found that the major processes which have been operated commercially around the world are the Lurgi, Koppers-Totzek, and the Winkler processes. When low BTU gasification processes are considered for integration into a power generation as the sole source of gas, the Lurgi process is the only one which is considered commercial. If a compressor-turbine set is added to the low BTU gasifier to form a combined cycle configuration, then Lurgi is the only gasifier that is under test at this time at the pilot scale level.

In comparing the relative merits of stack gas scrubbing to low BTU gasification as a means of making high sulfur coal available in the near term it appears that lime or limestone scrubbing units are in competition with the Lurgi gasification systems.

In assessing the technological merits of stack gas scrubbing it was

found that the scrubbers use approximately 5.5% of the energy generated by the plant on which they are installed.<sup>2</sup> For lime or limestone systems, approximately 3 tons/mw day of sludge is generated on a dry basis or 8 ton/mw day on a wet basis which must be disposed of in some manner.<sup>5</sup> For regenerative systems, a by-product is produced, usually sulfur, which can be stored or sold. Based on current and any foreseeable possible technologies there is no possibility that a stack gas scrubber system can ever be anything but an energy drain on the system it controls. In this sense it is a deadend technology. Further, the waste disposal problems associated with throw-away processes are immense.

By way of contrast, the low BTU gasification combined-cycle system properly integrated into a conventional power generation system has tremendous possibilities for technological improvement. The primary technological advances that must be made in order to fully realize the potential of the low BTU gasification system is a higher allowable inlet gas temperature to the compressor, higher particulate removal efficiencies in collection devices upstream of the turbine and an  $H_2S$  removal system capable of operating at high temperatures. This configuration has the potential to increase the overall system efficiencies of conventional power plants from 38% to 47%.<sup>44,45</sup> In addition, the probability that these technological improvements will occur is high. For example, Pratt Whitney, a major manufacturer of turbines, has reported that aircraft turbines now cruise with inlet temperatures around 2000°F.<sup>44</sup> It is projected that attainable inlet temperatures could reach 2800°F during the next decade. Inlet temperatures of 2400°F could result in an overall plant efficiency of almost 44%.<sup>44</sup> In addition, the current  $H_2S$  cleanup processes for gasification require that the temperature of the process stream be lowered considerably in order to use present day abatement technologies. Subsequent technological development of processes which can remove  $H_2S$  at high temperature could result in a further increased efficiency of the

gasification system. Again, there is every reason to believe that these advances will occur. This would further increase the potential overall efficiencies of power generation configuration. In low BTU gasification systems, elemental sulfur is the major by-product. As with the stack gas cleanup systems, this by-product can be either stored or sold. The amount of ash generated from these processes is approximately the same as that generated by conventional coal burning installations.

In comparing the two technologies discussed above, it is seen that in terms of energy efficiencies, the stack gas cleanup method is a deadend technology whereas the low BTU gasification option provides tremendous potential for increased energy efficiencies. For this reason, the most attractive long term technological alternative is low BTU-combined cycle gasification.

Current estimates of capital costs for retrofit stack gas cleanup range from \$45 to \$108 per kw.<sup>2,3,24</sup> The reported capital costs for low BTU gasification retrofit configurations range from \$110/kw to \$498/kw.<sup>31,44,46,47</sup> These costs are for using these two technologies on conventional power generation systems. For new plant installations, however, IGT<sup>44</sup> reports costs of \$280/kw for an integrated power system which includes a low BTU gasifier operating in a combined cycle configuration. This compares with the \$300/kw which is often quoted for conventional coal burning installations without provisions for stack gas cleanup if high sulfur fuels are used.<sup>31</sup> These costs are summarized in Tables 40 and 41.

Because of the potential for technological improvement associated with low BTU gasification systems, it is recommended that a subsidy be created to encourage the development of gasification processes in the utility industry. The subsidy should be guaranteed for a fixed time period after which the utilities would assume full economic responsibility. The duration of the

Table 40

Comparison Between the Capital Cost of Retrofitting a Coal Gasifier to an Existing Boiler and Stack Gas Cleanup

<u>Process</u>	<u>Cost</u>
EPRI <sup>47</sup>	\$339-495/kw
IGT U-Gas <sup>44</sup>	\$110/kw
Federal Power Commission Report <sup>46</sup>	\$148/kw
BCR Air Blown Two-Stage Gasifier <sup>31</sup>	\$117/kw
Stack Gas Clean-up - Retrofit <sup>3</sup>	\$45-65/kw
Retrofit <sup>24</sup>	\$65-100/kw
Retrofit <sup>3</sup>	\$50-108/kw
Retrofit <sup>47</sup>	\$51-91/kw

Table 41

Comparison Between the Capital Cost of a New Low BTU - Combined Cycle to a New Coal Fired Conventional Boiler

<u>Process</u>	<u>Cost</u>
Low BTU-Combined Cycle <sup>44</sup>	\$216-268/kw
New Conventional Boiler <sup>31</sup>	\$300/kw

subsidy would be determined by reasonable estimations regarding the time required for realizing the technological improvements in turbine blading, high temperature H<sub>2</sub>S removal and dust removal capabilities necessary for the ultimate development of the gasification system.

Institutional considerations also play a large role in mapping out sulfur abatement strategies. The stack gas sulfur oxide removal method was the solution proposed by the regulatory agencies for utilizing high sulfur coal. The industries were less than enthusiastic regarding this solution because it represented a drain on their power generation systems and would never improve their economic position. Low BTU gasification with a combined cycle, however, was an alternative proposed by industry and which has the potential to improve their economic position over the long term. For this reason, it seems likely that with the proper encouragement; i.e., a subsidy, there is a much higher probability of industry providing real leadership to see that this technology reaches maturity in the shortest period of time.

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