ECONOMICS OF COAL FINES UTILIZATION

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ABSTRACT

In the twentieth century, coal has become the major fuel for electric power generation in the U.S. and most of the nonpetroleum-producing countries of the world. In 1988, the world coal-fired capacity for electric power generation was about 815 GW, consuming large quantities of coals of all ranks. Today, coal provides a third of the world's energy requirements. In fact, coal use for power generation has grown steadily since the oil embargo in 1973 and has seen an even faster rate of growth in recent years. It has been reported that the global demand for new coal will increase by more than 1500 million tons by the year 2000. However, this increased production of coal has its drawbacks, including the concomitant production of coal waste. Reported estimates indicate that billions of tons of coal waste have already been disposed of in waste impoundments throughout the U.S. Further, in the U.S. today, about 20-25% of each ton of mined coal is discarded by preparation plants as gob and plant tailings.

It appears that the most economical near-term approach to coal waste recovery is to utilize the waste coal fines currently discarded with the refuse stream, rather than attempt to recover coal from waste impoundments that require careful prior evaluation and site preparation. A hypothetical circuit was designed to examine the economics of recovery and utilization of waste coal fines. The circuit recovers products from 100 tons per hour (tph) of coal waste feed recovering 70 tph of fine coal that can be used in coal-fired boilers. The present analysis indicates that the coal waste recovery is feasible and economical. In addition, significant environmental benefits can be expected.

INTRODUCTION

Vast quantities of high-quality coal waste, estimated to range from 500 million tons to more than 2 billion tons, are stored in as many as 473 permitted coal slurry impoundments in the eastern U.S.[1,2]. Each year, operating coal preparation plants send 30-50 million more tons of fine material to these impoundments. An estimated one-fourth of the discarded fine coal could be made into usable fuel for pulverized coal-fired utility boilers, the predominant boiler-type for burning coal to produce electricity in the U.S. Although the quality of material is variable, some is of relatively high quality and would require little cleaning for recovery and utilization. Although each site would need to be evaluated on a case-by-case basis, the most likely candidates for reclamation projects are impoundment sites located near operating preparation plants, where existing infrastructure could be used to facilitate coal fines recovery.

The recovery and use of these coal fines either from impoundments or preparation plants could provide significant economic and environmental benefits. For example, coal producers could obtain a marketable product that is less susceptible to market fuel price fluctuations due to the inexpensive nature of the source material. The country's available supply of coal could be extended by replacing a portion of the coal supplied by operating coal mines with reclaimed coal from refuse ponds. The nation could also benefit from the reduced environmental impact of these refuse sites.

The utilization of the fine coal that is currently discarded with the refuse stream at preparation plant is feasible and economical. The technology for co-firing this fine coal waste with pulverized coal in a utility boiler is already being successfully demonstrated [3]. The purpose of this paper is to examine the economics of recovery and utilization of waste fine coal for combustion using this co-firing approach in a larger, 700-MWe boiler.

COAL FINES RECOVERY AND FUEL FORMULATION

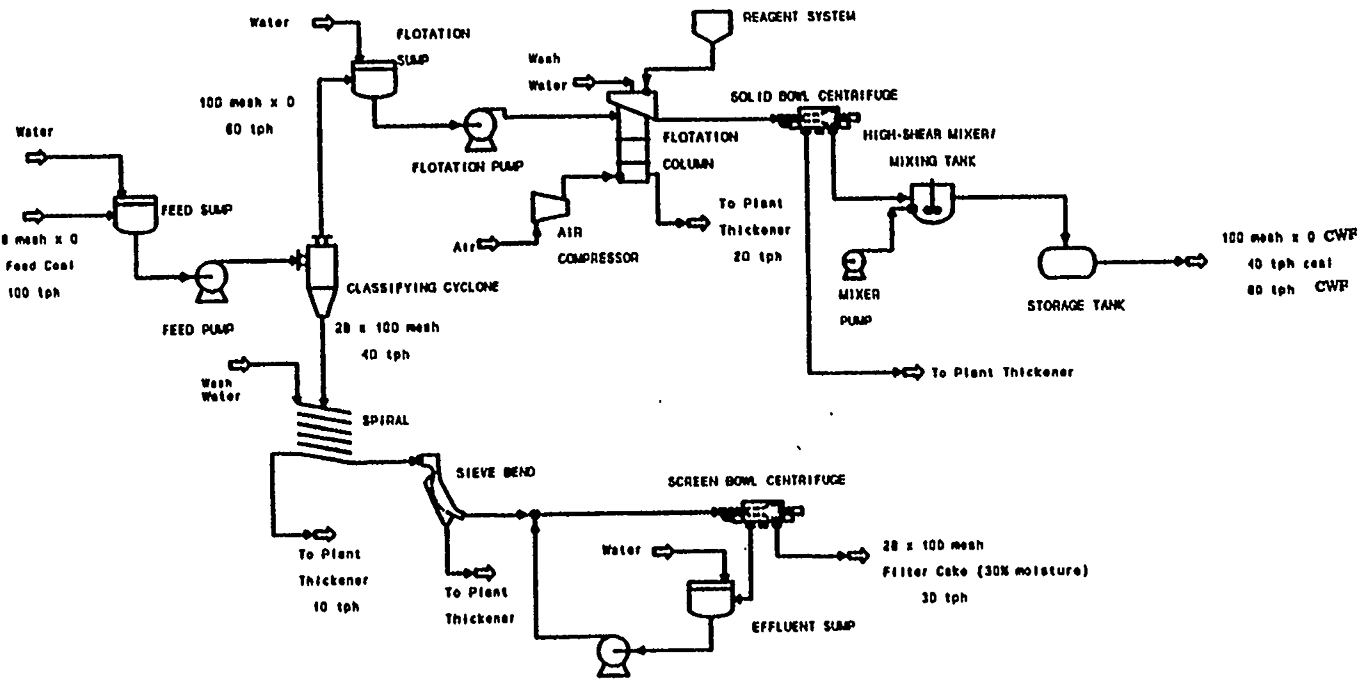
The hypothetical coal preparation plant under consideration recovers fine-coal refuse rather than discarding it. The plant processes 1,000 tons (907.2 metric tons) per hour (tph) of a high heating value, low-ash, and low-sulfur bituminous coal. Conventional coal cleaning equipment is used to process the plus 28-mesh material. The minus 28-mesh material is directed to a thickener for water clarification and subsequent disposal to a slurry pond. To reduce the loss of coal to the waste impoundment, a fine coal recovery circuit is added to the plant to recover fine coal from the minus 28-mesh material (100 tph). To present the least-cost scenario, the recovered coal fines material is utilized as a coal-water fuel (CWF). The use of this fuel form avoids the high costs of fine coal dewatering, and provides favorable handling and transportation characteristics.

Figure 1 is a conceptual flow diagram of the hypothetical coal fines recovery and CWF formulation circuit. Hourly, the circuit processes 100 tons or 90.72 metric tons (dry basis) of refuse stream with a heating value of 11,000 Btu/lb (25.59 MJ/kg) to produce two product streams of equal heating value (14,000 Btu/lb or 32.56 MJ/kg) : 40 tons (36.29 metric tons) of 100-mesh x 0 coal, which becomes 80 tph CWF, and 30 tons (27.21 metric tons) of 28-x 100-mesh coal filter cake. The remaining 30 tons (27.21 metric tons) is a 28-mesh x 0 refuse.

The minus 28-mesh stream is first sent to classifying cyclones which separate the feed material into a coarse underflow stream and a dilute fine overflow stream in the following manner:

(1) The 100-mesh x 0 overflow stream from the cyclones is directed to flotation columns. Frother and collector chemical reagents are added to the columns to facilitate flotation. The product from the flotation columns is sent to solid-bowl centrifuges for partial dewatering. Partial dewatering is sufficient to achieve the desired moisture content of 50% in the CWF. The product consists of 80 tph of 50 wt. % CWF containing 40 tph of dry coal fines. The CWF product is pumped to mixing and storage tanks for transportation to a utility site.

(2) The 28- x 100-mesh underflow stream from the classifying cyclones is treated in spiral concentrators and dewatered with sieve bends. The excess water in the clean coal product is removed in screen-bowl centrifuges. The coal product is a filter cake containing approximately 30% moisture, which is mixed with normal cleaned pulverized coal produced by the preparation plant. In this way, the 28- x 100-mesh coal that was not recovered in conventional coal cleaning is included. Refuse from the flotation columns and the sieve bends is sent to the existing plant thickener for disposal.



NOTE: Circuit contains the necessary pumps, sumps, and piping (not shown) to provide for water circulation from plant thickener. Makeup water is distributed where needed from a common head tank.

Figure 1. Schematic Flow Diagram of the Coal Fines Recovery and CWF Formulation Circuit.

It is assumed that the 100-mesh x 0 coal fines recovered from the preparation plant waste stream is formulated as CWF containing 50 wt. solids with no stabilizing chemicals added. The premise is that the fuel would be used within a few days of preparation and, therefore, not need chemical additives that are required for long-term storage. The CWF produced by this circuit has sufficient energy content to provide 7.84 x 10^{12} Btu/yr (8.27 x 10^6 GJ/yr), which represents approximately 20% of the thermal input to a 700-MWe coal-fired utility operating with a 65% capacity factor.

CAPITAL COSTS OF COAL FINES RECOVERY, FUEL FORMULATION, AND UTILIZATION

The costs of the individual pieces of equipment were obtained from several sources [4-6]. All costs were adjusted to 1994 dollars. For equipment where only purchased and delivered costs were available, a factor of 3.17 [7] was applied to obtain installation costs. The following discussion describes the major equipment items in various sections of the circuit:

- Coal Fines Recovery at the Cleaning Site: The circuit includes classifying cyclones, froth flotation columns, spiral concentrators, and clean-coal dewatering centrifuges. Various pumps and sumps are also required. The total capital required for this 100 tph coal fines recovery circuit is $4.84 million.
- CWF Formulation and Storage at the Cleaning Site: The coal fines CWF product from the solid bowl centrifuge is homogenized in high-energy mixing tanks to prepare coal-water fuel with 50 wt. % solids loading. Other equipment includes storage tanks and pumps. The total capital expenditure to formulate and store CWF at the coal fines cleaning and recovery site is $2.03 million.
- CWF Storage and Utilization at Utility Site: The major equipment needed at a utility site are tanks and pumps for CWF storage, a burner and atomizer system, and the distribution system to the burner. The total cost of these items is $3.0 million.

A 30% project contingency was added to the total capital cost. As shown in Figure 2, the capital requirements for coal fines recovery, formulation, storage and utilization, and contingency total $12.83 million, consisting of $8.93 million for fuel preparation and $3.90 for the user.

FIXED AND VARIABLE OPERATING & MAINTENANCE (O&M) COSTS

Table 1 summarizes the fixed and variable O&M costs for a 100 tph coal fines utilization system. All costs are in 1994 dollars.

Fixed Cost at Cleaning Site: Labor and CWF transport costs represent the fixed O&M costs for the cleaning plant. Eleven operators ($57,950/person/year) are required to operate the coal fines recovery and CWF formulation sections at the cleaning site. Two shift foremen ($69,750/person/yr) will supervise operators and maintain the quality of the coal fines recovery products at the cleaning site.

The CWF product is transported by truck to the point of utilization, assumed to be within a 25-mile radius of the formulation site. This item is actually a combination of fixed and

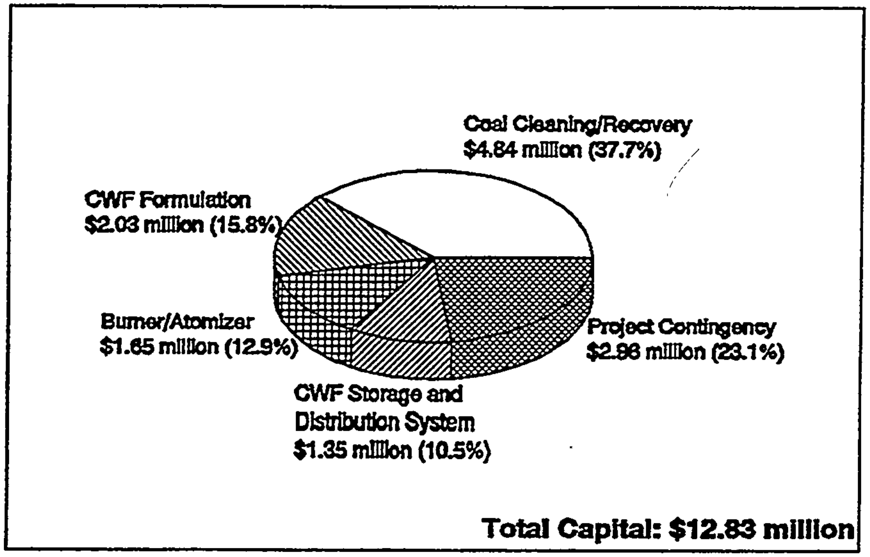


Figure 2 Capital Costs of the Coal Fines Cleaning/Recovery, CWF Formulation, and Utilization.

variable O&M costs but is conservatively carried here as a fixed cost. The transportation cost was obtained from reference 4 as $0.21/10^6$ Btu. For 7.84×10^{12} Btu/yr, the CWF transportation cost amounts to $1,646,400/yr. The cost of a vehicle at the preparation plant is based on the total tonnage of 28-mesh x 0 waste feed material processed by the cleaning circuit, as discussed in reference 5.

Fixed Labor at the Utility Site : Three persons work for the foreman responsible for maintaining boiler operation at the utility site. Using the same rate of wages for the operators at the cleaning site, the labor at the utility site is $173,850/yr.

Variable Costs at the Cleaning Site : The major components include electric energy, chemical additives, refuse disposal, and materials and supplies. The electricity cost for operating the coal fines recovery and CWF formulation units at the cleaning plant is $1,468,127, based on 6.0 cents/kWh. The cleaning plant frother and collector require reagents to separate coal fines; however, no chemical additives are used in the CWF formulation. Although a portion of the load is being removed from the existing thickener input, a cost was applied to dispose of the 30 tph refuse after coal fines recovery. The annual cost of materials and supplies is generally estimated as 5% of total capital cost [6].

Table 1. Summary of Annual Fixed and Variable O&M Cost for the Cleaning and Utility Sites.

Item	Fixed O&M, 1994 $		Variable O&M, 1994 $	
	Coal Fines Recovery & CWF Formulation Site	Utility Site	Coal Fines Recovery & CWF Formulation Site	Utility Site
Labor	776,950	173,850	–	–
CWF Transport	1,646,400	–	–	–
Material	–	–	446,900	195,200
Vehicle	21,000	–	–	–
Power	–	–	1,468,127	718,463
Chemicals	–	–	91,500	–
Heat Loss	–	–	–	739,900
Disposal	–	–	210,000	–
TOTAL	2,444,350	173,850	2,216,527	1,653,563

Variable Costs at the Utility Site: The cost to vaporize water contained in the CWF using 1,000 Btu/lb or 2.32 MJ/kg as the heat of vaporization and a rate of $\$1.00/10^6$ Btu [4] is \$560,000/yr. An additional cost of \$179,900/yr must be added to vaporize the moisture in the 28- x 100-mesh spiral concentrator product, which is in the form of a filter cake containing 30% moisture. For the utility, the costs for power and materials/supplies are \$718,463/yr and \$195,200/yr, respectively.

COST OF PRODUCTS FROM COAL FINES RECOVERY

The majority of coal preparation plants are independent from utilities. Thus, the capital and annual O&M cost components are shown separately in Table 2. There is no cost for the input waste stream to recover coal fines and formulate CWF since the waste stream is a portion of the original run-of-mine feed coal stream entering the preparation plant. Other financial assumptions are 20-year plant life, 80% plant operating factor, 4.1% inflation rate, and a 7-year declining balance depreciation method. The combined federal and state tax rate is 38% and the property tax rate is 2%. A nominal after-tax return of 15% or 25% on all equity investment was applied as applicable to the specific case. All capital investment funds are assumed to be expended one year before plant operation begins. Based on these assumptions and the cost data listed in Table 2 for the coal fines recovery site, product costs were calculated using a financial model developed in reference 8. Table 3 presents the results from the modeling effort in which identical financial assumptions were used for all three cases.

Table 2. Cost Components for Organizations Involved in Coal Fines Recovery and Utilization.

Category	Coal Fines Recovery Site Millions of 1994 $	Utility Site Millions of 1994 $
Capital cost without Contingency	6.87	3.00
Capital cost with Contingency (30%)	8.93	3.90
Fixed O&M Cost	2.44/yr	0.17/yr
Variable O&M Cost	2.22/yr	1.65/yr
Working Capital	0.38	0.16

Table 3. Product Cost Estimates.

Case	Waste Recovery (Percent)	Products[1]	Nominal Rate of Return after Tax (Percent)	Total Cost ($/$10^6$ Btu)
A	70	40% Coal Fines as CWF and 30% Dry Coal as Filter Cake	15	0.43
B	70	40% Coal Fines as CWF and 30% Dry Coal as Filter Cake	25	0.51
C	40	40% Coal Fines as CWF	15	0.75

For the best scenario, Case A, where both CWF and 28-x 100-mesh coal as filter cake products are sold to the utility, the overall product cost is $0.43/$10^6$ Btu. However, if only the CWF is produced (Case C), the product cost is $0.75/$10^6$ Btu. Case B is similar to Case A but includes a higher, 25% versus 15%, rate of return to the coal cleaning company. In this Case, both CWF and 28-x 100-mesh coal products are sold to the utility at a total price of $0.51/$10^6$ Btu.

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[1] All CWF products are 50 wt. % solids loading, 7,000 Btu/lb (16.28 MJ/kg), and are produced at a rate of 80 tph.

ECONOMIC ADVANTAGE TO A UTILITY

The three aforementioned cases are analyzed in terms of fuel cost savings to a utility and the payback periods to recover the utility's capital investment.

Savings in Fuel Cost: The major advantage to a utility is the availability of a low-priced fuel and the resultant savings in fuel costs. Economics have been estimated for the three product cases developed above, assuming a 700-MWe nameplate capacity unit, with a 65% capacity factor and a nominal heat rate of 10,000 Btu/kWh. This unit consumes approximately 7.0 x 10^9 Btu/hr (7.39 x 10^3 GJ/hr) of fuel. The contract price of the delivered feed coal is assumed to be \$1.65/10^6 Btu.

For Case C (recovery of coal fines as CWF only), the rate of the coal fines production [80 tph of 7,000 Btu/lb (16.28 MJ/kg) CWF with 50 wt. % coal solid loading] is sufficient to provide 20% of the power plant heat requirement. From Table 3, the price of CWF is \$0.75/10^6 Btu. The weighted average fuel cost is \$1.47/10^6 Btu, based on 80% pulverized coal fuel and 20% CWF prepared from coal fines. The fuel cost savings to the utility is \$1.65/10^6 Btu less \$1.47/10^6 Btu or \$0.18/10^6 Btu. At a fuel feed rate of 7.0 x 10^9 Btu/hr and a 65% capacity factor, the gross savings amount to \$7,174,000/yr.

Likewise, the gross fuel savings amount to \$17,140,000/yr for Case A (recovery of dry coal with CWF production) and \$ 15,940,000/yr for Case B (recovery of dry coal with CWF production at a higher rate of return).

Capital Investment Recovery Period: The effects of fuel cost savings on the utility are included in Table 4. From Table 2, the utility investment is \$3.90 million (including contingency). The annual incremental expense associated with using waste fuels on a boiler for all three cases is shown in Table 4. As a simplified method of expressing the financial benefit to a utility, a payback period to recover the capital investment is selected. Subtracting the incremental expense from the fuel savings gives the net savings before taxes, and applying a 38% tax rate gives the net savings after taxes. Payback period is calculated by dividing the incremental capital by the net annual savings. For Case A, the payback period is 0.40 yr; for Case B it is 0.44 yr; for Case C it is 1.12 yr.

SENSITIVITY OF THE COST OF CWF PRODUCT

The sensitivity of Case C to various technical and economic parameters was examined since this cost of producing only CWF was the highest among the three cases. The summary of the effect of change of each parameter is shown in Table 5. The base cost value in Table 5 are obtained from Table 1. Most parameters showed a change up to \$0.04/10^6 Btu for a 20% positive or negative variation. The parameters listed below exhibit noteworthy exceptions because they show an inverse trend in the CWF production cost.

Waste Recovery: As shown in Table 3, Case C is for 40% recovery of coal fines for use as CWF. As the total amount of coal fines for CWF production increases by 20%, the cost of

Table 4. Economics for Utilization of Recovered Coal Wastes at a 700-MWe Utility Operating with a 65% Capacity Factor (1994 $).

	Base	Case A	Case B	Case C
Fuel Consumption, 10^{13} Btu/yr	3.98	3.98	3.98	3.98
Fuel Composition, Percent				
Coal	100	65	65	80
Coal-Water Fuel	–	20	20	20
Filter Cake	–	15	15	–
TOTAL	100	100	100	100
Fuel Price, $/10^6$ Btu				
Raw Coal	1.65	1.65	1.65	1.65
Coal-Water Fuel	–	0.43	0.51	0.75
Filter Cake	–	0.43	0.51	–
Weighted Average	1.65	1.22	1.25	1.47
Savings Relative to Base Case	–	0.43	0.40	0.18
Fuel Cost Savings, 10^6/yr	–	17.14	15.94	7.17
Incremental Capital Cost at Utility (Including Contingency), 10^6	–	3.90	3.90	3.90
Incremental Expense, 10^6/yr				
Fixed O&M + Property Tax at 2% of Capital	–	0.25	0.25	0.25
Variable O&M	–	1.65	1.65	1.65
Net Savings, Before Taxes, 10^6/yr	–	15.24	14.04	5.27
Depreciation (7-year Life)	–	-0.56	-0.56	-0.56
Net Savings, After Taxes, 10^6/yr	–	9.68	8.95	3.51
Payback on Incremental Capital, After Taxes, Yr.	–	0.40	0.44	1.12

Table 5. Sensitivity of the Technical and Financial Variables of Fine Coal Recovery and CWF Formulation (Production of CWF of 7,000 Btu/lb)

			Change from Product Cost Estimate of $\$0.75/10^6$ Btu.	
Item	Variation	Base Case	+ Change (+ or - $)	- Change (+ or - $)
Capital	± 20%	$8,936,000	+0.04	-0.04
Power	±20%	$1,468,127/yr	+0.03	-0.04
Labor	±20%	$776,950/yr	+0.02	-0.02
Transport	±20%	$1,646,400/yr	+0.04	-0.05
Materials & Supplies	±20%	$446,900/yr	+0.00	-0.01
Waste Recovery	±20%	40%	-0.13	+0.18
Disposal	±20%	$210,000/yr	0.00	-0.01
Plant Life	±50%	20 yr	-0.02	+0.06
Income Tax Rate	±20%	38%	+0.01	-0.01
Rate of Return (Nominal after Tax)	±33%	15%	+0.07	-0.07
Capacity Factor/(Labor Time)	59.4% (3 Shifts)[1]	80%		+0.14
	41.6% (2 Shifts)[2]			+0.38

[1] For three shifts, the capacity factor (59.4%) of the cleaning and coal fines recovery plant is based on 120 hr/week, 52 week/yr. The three-shift operation assumes 100 hr/week for production and 20 hr/week for maintenance.

[2] For two shifts, the capacity factor (41.6%) of the cleaning and coal fines recovery plant is based on 80 hr/week, 52 week/yr. The two-shift operation assumes 70 hr/week for production and 10 hr/week for maintenance.

the product decreases by \$0.13/10^6 Btu. However, a 20% decrease in the quantity of coal fines recovered for CWF production increases the product cost by a greater margin (\$0.18/10^6 Btu).

Plant Life: A 50% increase in plant life decreases the product cost by \$0.02/10^6 Btu because the costs are spread over a longer time period. A 50% decrease in plant life, however, increases the product cost by a much greater degree (\$0.06/10^6 Btu).

Capacity Factor/(Labor Time): A decrease in plant operating time significantly increased the product cost. Operating two shifts instead of three results in a CWF-only product cost of \$1.22/10^6 Btu. It should be noted here that reduced operating time impacts both labor costs and CWF production which changes the product cost.

The rate of return desired by the cleaning site has a significant change in the cost of CWF-only production. A 33% change from the base case alters the product cost by \$0.07/10^6 Btu.

Table 6 shows the breakdown for the Case C product cost for the coal fines recovery site based on 7.84 x 10^{12} Btu/yr of CWF. The CWF transport cost and power account for more than half of the total product cost; that the capital cost amounts to less than 22% of the total product cost of \$0.75/10^6 Btu.

Table 6. Parameters Contributing to the Cost of CWF-only Product.

Item	First Year Cost or Return, \$	\$/10^6 Btu	Percent of Total
Capital (Return on Investment)	1,223,040	0.16	21
Power	1,468,127	0.19	25
Labor	776,950	0.10	13
Transport	1,646,400	0.21	28
Materials & Supplies	446,900	0.06	8
Other Misc.	322,500	0.03	5
TOTAL	--	0.75	100

SUMMARY

This study shows favorable economics for coal fines recovery and utilization as CWF co-fired with pulverized coal. The economic benefits of coal fines recovery increase with the quantity of useful products generated by such processing. Recovery of 40% of the coal fines as CWF from currently discarded cleaning plant wastes can alone result in significant cost savings, requiring 1.12 years to recover the utility's capital investment. However, the recovery of additional combustible fuels for utilization can further enhance the economic benefit to a utility. The magnitude of this benefit is affected somewhat by the economic incentive provided to the cleaning plant owner. For example, if the owner's rate of return is increased from 15%

to 25% (Case B), the fuel cost to the utility increases by 2.5% and the time to recover its investment increases by 10%.

A sensitivity analysis of the CWF product cost was performed. This analysis identified three parameters that have the most effect on the CWF production cost: percent waste recovery, CWF transportation cost, and the capacity factor.

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CURRENT TECHNIQUES
IN THE RECLAMATION AND TREATMENT OF
FINE COAL REJECT RESERVES

by

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INTRODUCTION

The disposal of fine reject slurry from coal mines led to the requirement for impoundment structures and subsequent safety regulations for their construction and maintenance. Now, the changing technology of power generation, combustion technology and fine coal treatment has given these impoundments a new economic importance. In addition, the public awareness of environmental and safety issues has focused attention on these ponds creating a public demand to mitigate the problems associated with their existence. The position has been reached where it is feasible and economic to remove slurry from impoundments and to recover the energy resource in the form of fine coal.

Reclamation of fine coal slurry from impoundments has traditionally been approached with adaptations of standard coal mining and processing methods. Over a twelve year period, Mineral Development Corporation (MDC) has developed many processing, operating and evaluation techniques specifically for the recovery and treatment of coal slurry from active and abandoned slurry impoundments.

MDC is a U.S. company through which these technical developments have been introduced to United States.

This paper discusses the techniques and products unique to the MDC operations that provide the economic advantages to operate successfully on a long term basis in the "carbon recovery" industry.

HISTORY

MDC now leads the world in the production of fine coal product from the reclamation and treatment of reject slurry material. As illustrated in Fig. 1, over the last twelve years MDC has produced and sold in excess of 3.1 million tons of high quality, handleable fine coal product from seven projects in Australia and USA. During that period, MDC's unique dredging techniques have been employed to reclaim in excess of 11.0 million tons of material

MDC will commission its eighth coal refuse recovery project in the second quarter of 1995 and has obtained formal commitment for its ninth project expected to start in the fourth quarter of 1995.

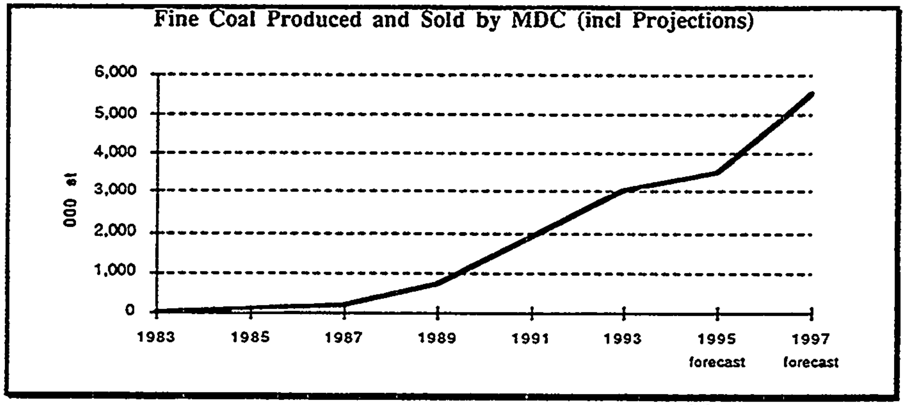

Fig. 1

The first MDC operation in the USA was constructed and commissioned in February 1991, at the Streamline mine in southern Illinois. The plant operated successfully, producing 400,000 tons of consistent, high quality fine coal from a site that manifests many of the problems that have for years plagued traditional "carbon recovery" operations from slurry impoundments.

MDC's recovery techniques are specifically designed to reduce reclamation costs and ensure timely release of liabilities. At the Streamline project, reclamation is complete and as shown in Fig. 2, the expected costs of reclamation were significantly reduced as a result of the MDC operating and reclamation techniques.

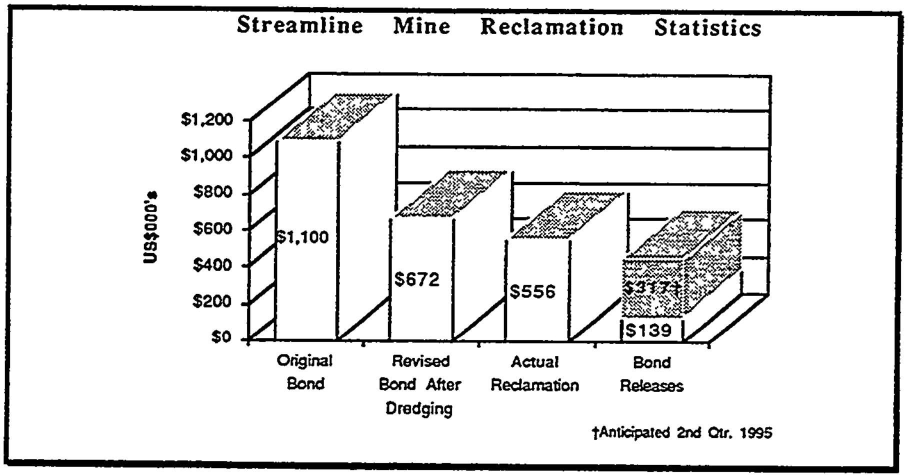

Fig. 2

As a result of this success, Mineral Development Corporation is embarking on three additional projects in the USA and continuing its exploration and evaluation of various slurry pond sites throughout North America.

The production of fine coal from slurry has been attempted over a number of years. Traditionally, this production consisted of smaller operators recovering the insitu slurry with "truck and shovel" operations with little or no secondary treatment. These operations suffered from poor quality product with poor handling characteristics. The operations usually fail early due to the inability to maintain production rates once the water table is encountered by the operation.

Those operations that use dredging as a means of slurry recovery and employ a secondary processing operation often find that the production costs in the current depressed market make the operation uneconomic. In many cases, the secondary processing does not produce a consistent quality product with good handling characteristics. The poor environmental condition remaining after such operations is typical in the "carbon recovery" industry and presents an entrenched "image problem" to be addressed by all future MDC operations.

The MDC approach and technology have addressed and overcome these traditional problems by introducing techniques to four basic areas of the business.

1. Detailed exploration and sampling of the in situ slurry to evaluate the resource and predict the yield, quality and size distribution of the product.

2. Dredging of the slurry in accordance with a mining plan so that feed rate and quality fluctuations can be foreseen and accommodated.

3. The efficient and effective use of equipment (some developed and manufactured by MDC) and processes for recovery of a high quality, low moisture product material with excellent handling characteristics, leaving the non combustible portion of the slurry (sandstone, shale, clay - including pyrite) as reject.

4. The effective handling of the reject streams from the process for other products or reclamation benefits.

EXPLORATION AND EVALUATION

The specific circuit design for any MDC project is formulated on the characteristics of the slurry. These characteristics are determined from laboratory tests of core samples. The samples are obtained by the "vibracore" method of drilling which recovers a complete, undisturbed core over a depth of 28 feet. The "vibracore" method utilizes a 30' aluminum tube of 3" diameter. An adapted concrete vibrating mechanism is attached to the tube to provide the motive force for penetration of the tube vertically into the slurry. Once the tube has reached full extension (or depth of slurry), the tube is capped (airtight) and then removed from the slurry. Atmospheric pressure allows the recovery of a core (approximately 28') of sample within the tube.

This method has also been successfully used for sample collection from slurry covered by water. MDC has constructed a light, mobile floating work platform from which this drilling operation can be performed.

In all cases, the "vibracore" drilling technique allows for effective and safe sample collection from the surface of the impoundments with little or no environmental impact on the slurry surface. The total drilling equipment is able to be operated and transported manually by two operators.

The slurry is usually deposited in a predictable manner. The coarser, heavier particles are deposited near the discharge point and the lighter, finer fractions migrating to the furthest point from the discharge. There is a variation in yield and quality in relation to this deposition. These patterns must be taken into account when planning the evaluation drilling and when preparing the operational mining plan.

As far as practical, the drilling is completed on a regular grid pattern over the entire area of the pond. The information from this exploration is used to develop the mining plan; predict the life of the operation; predict the quality of the product; and correlate the actual production results with predictions.

In very deep ponds, a number of deep holes are drilled to determine or confirm the floor profile and obtain samples of the deeply buried slurry. More recently seismic methods have been used to determine the profile of the pond floor and ascertain if there are hidden mining problems within the pond. The deep drilling (or seismic work) may not be required if the circuit of the original washery did not change over the life of the impoundment and the bottom profile of the impoundment is known through earlier surveys. Typically, as dredging progresses and slurry levels are lowered, additional drilling will be conducted ahead of the dredge to update and confirm the original exploration information.

The analytical test procedure of the drill core samples that has been developed by MDC, is extensive in order to provide suitable data for use in the in-house computer simulation model. The simulation model is used to predict the yield of coal from the impoundment and the quality of the material produced, with regard to ash, sulphur content, thermal energy and moisture. The simulation model has been developed over many years of test work and correlation with actual production.

In order to evaluate the economics of a slurry pond, four successively conditional, investigative stages are undertaken.

1. Visual inspection of the pond

   An initial inspection will result in a preliminary assessment of the quantity of slurry, a visual assessment of the quality of the slurry (e.g. size distribution) and the physical characteristics of the site that may have an impact on any potential operation (e.g. water and power availability, plant location, reject disposal potential, etc.).

2. Initial drilling and analyses

   In this stage, three or four drill core samples are taken at appropriate positions in the pond as an approximate representation of the total resource. These samples are subjected to a short size distribution analysis.

3. Short gravimetric float/sink analyses

   If the sizing analyses indicate the potential viability for slurry recovery and treatment, then a short (3 densities) gravimetric float/sink analysis is carried out which will indicate an approximate yield and quality of product.

4. Full evaluation/feasibility

   If the above test procedures are favorable, then a major evaluation is undertaken.

   This evaluation includes extensive drilling (both deep and shallow); detailed sizing analyses on each drill hole sample; extensive (10 densities) gravimetric float/sink analyses of each size fraction; mining plans; permit investigation/application and other engineering/feasibility studies.

DREDGE AND DREDGING TECHNIQUES

MDC recovers slurry from an impoundment developed by the use of a dredge that is designed specifically for this operation. The dredge is a fully contained unit (except for the power supply) and is remotely controlled. The dredge floats in a minimum of four feet of water. The slurry is recovered from beneath the water by means of a suction pipe positioned behind a revolving cutter head on the end of a boom or ladder. Depending of the dredge size, the cutter can operate up to 28' below the dredge. Also dependent on design, the MDC dredges can deliver from 100 to 150 dry tons of slurry per hour. The recovered slurry is pumped by floating and land pipelines direct to the plant. Control cables allow the dredge to be operated remotely from the control room in the plant. In some of the MDC operations, the dredge has operated automatically up to two miles from the plant.

The dredge is maneuvered by three cables each anchored to small tracked mobile equipment on the shoreline. The cables attach to three winches located on the dredge at port, starboard and aft positions. The cables are positioned so that the dredge is always directed toward the slurry face. With PLC control the dredge location can be determined accurately by reference to the status of the three winches. This allows automatic mining of the slurry face. With the integration of a flow meter and a density gage on the delivery line the PLC can also control the traverse speed of the dredge to maintain a constant feed rate to the plant.

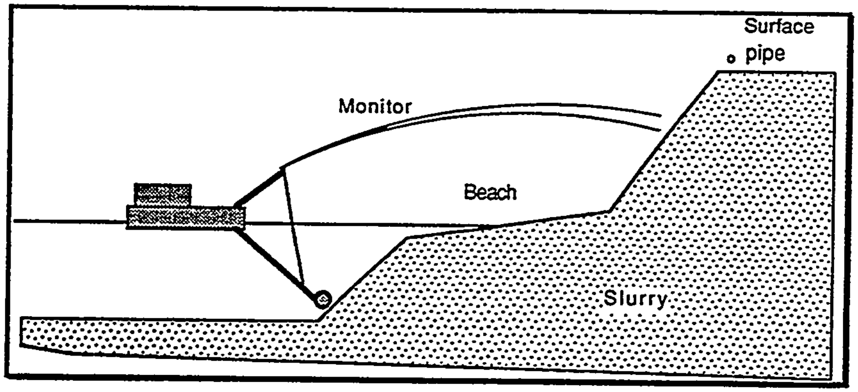

Fig. 3

Each impoundment is unique in its reaction to dredging but the usual procedure is to lower the water level so as to expose a 20 ft face and to wash the slurry above the water with high pressure monitors. This technique permits the slurry to slump in a controlled manner and to form a beach in front of the dredge in a condition partly prepared for suction (see Fig. 3). It also enables the dredge to operate a distance from the face in the unlikely event of a sudden uncontrolled fall. The fact that the dredge does not require on-board operators also increases the safety aspect in this situation. As an additional assistance to this technique, pipelines can also be placed on the slurry surface at the top of the face to wash the slurry through channels to the dredge.

The progress of dredging in a horizontal plane will be dependent on the distribution of the slurry - the extraction in a vertical plane is governed by the safe height of the wall in front of the dredge. In deep ponds this means that a number of passes are required to effect a complete extraction. Fig. 4 depicts a typical upstream wall impoundment and the sequence in which it will be dredged.

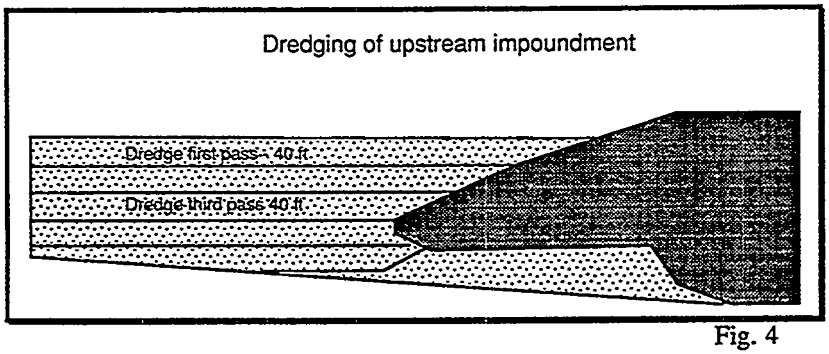

Fig. 4

EQUIPMENT AND PROCESSES

The MDC fine coal treatment plants have incorporated within their circuits a number of equipment and technical advancements developed specifically for the treatment of fine reject slurry material. The plant consists of the following equipment:

MDC PLANT EQUIPMENT UNITS	
1. **MDC Spirals (2 stages)**	- separate the coarse reject (low energy/high ash material) from the coal material to produce a high quality product. These units are 8 start columns that were developed and manufactured in house.
2. **MDC Rapped Sieve Bends**	- remove additional ultra-fine materials that cannot be concentrated by the spirals from the product. These units are manufactured by MDC in Australia.

PROPRIETARY EQUIPMENT	
1. **Sieve Bend/Shaker Screen**	- screen out the oversize (+ $\frac{1}{8}$ ") and pond debris and delivers a minus 9# material to the plant circuits.
2. **Deslime Cyclones**	- remove major portion of minus 200# material.
3. **Dewater Cyclones**	- remove excess water and ultra fine clays from coal product and,
4. **Screenbowl Centrifuges**	- remove water from the coal product prior to conveying the product to the bin or stockpile.

The arrangement of the various components and equipment contribute to the plant's ability to produce a coal product of high quality and excellent handling characteristics. A typical flowsheet is shown in Fig. 5.

Four of the patented components in the plant are:

1. Feed distribution system on rapped sieve bends
2. Rapped sieve bends
3. Eight start spirals
4. Feed distribution system for spirals

WATER REQUIREMENTS

The plant has a closed water circuit and requires minimal make-up water for the system. The majority of the water that accompanies the reject material is returned to the system through the use of pumps and holding/settling ponds. In general, the dredge requires about 280 tons of water per hour and the plant requires about 570 tons of recirculated water per hour, with a make-up water requirement of about 90 tons of water per hour. The water loss is attributed to that water that is bound within the coal product and evaporation.

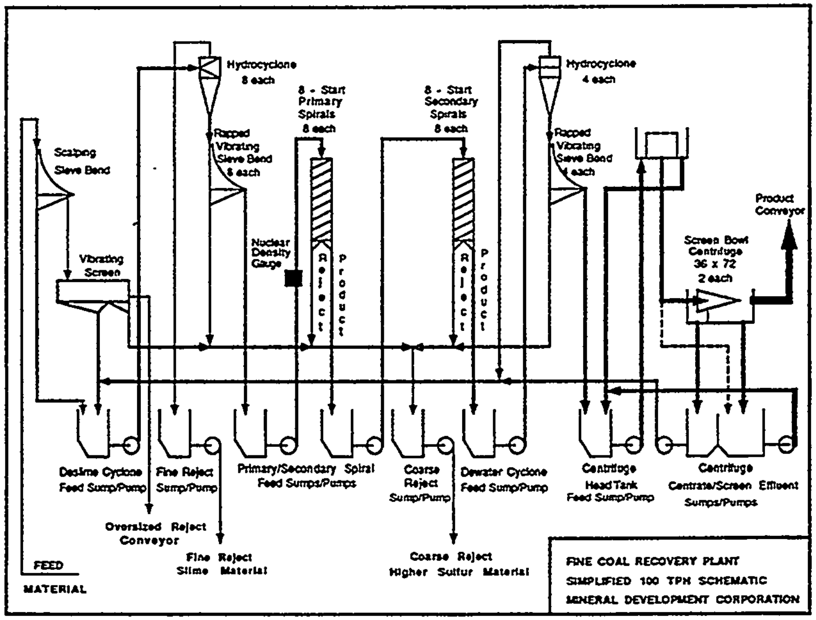

Fig. 5

POWER REQUIREMENTS

An MDC plant rated 130 ton per hour feed capacity with 1500 kilowatts installed power, has a power demand of 1,200 kilowatts. On average, the plant will use 170,000 kilowatt hours per week, based on the plant operating 24 hours per day, 6 days per week. This usage includes all power for the plant, dredge and associated pumps.

The plant is scheduled to operate 24 hours per day, 6 days per week; with the seventh day being available for minor maintenance or modifications to the plant equipment.

OPERATIONS

The plant is operated by a crew of six operators and one plant manager. The crew is on a rotation of three shifts per day, with two operators per shift. The plant manager is typically present at the site during the last hour of the night shift and the first hour of the afternoon shift. The plant can be operated with the small crew, due to the highly sophisticated PLC systems and computer monitoring capability.

Production from any project is dependent on yield achieved with the resource available. Typically, an MDC plant produces 250,000 tons to 400,000 tons on an annual basis.

RECLAMATION AND OTHER PROCESSES

Once material is dredged from the pond, there are unlimited options available to relocate and treat the various components of the material in the pond. These options allow the relocation of material to provide a better configuration in both the safety and environmental areas.

RECLAMATION AND SELECTIVE REDEPOSITION

Typically, the environmental concerns related to disposal of fine coal refuse center around the acid production from the pyrite (sulphur bearing) material. This acid production is responsible for the lack of plant growth on the slurry ponds. The MDC process separates the pyritic material from the finer clay bearing material. These two streams of material can then be selectively redeposited to provide the best environmental result (see Fig. 6).

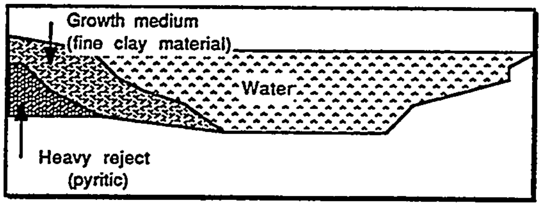

Fig. 6

If the pyritic material is located at the bottom of the reject area it can be isolated from oxygen that is required for acid production. In addition, the finer clay bearing material can be used to cap the pyritic material as a seal. The finer clay bearing material also provides the necessary nutrient base for the establishment of a plant growth medium. This aspect of the reclamation results from the dredging and re-deposition alone and does not require the expensive reclamation methods now in use.

This process allows MDC to reclaim the pond areas at a cost far lower than required by the current methods of reclamation. MDC is able to reduce the current liabilities of coal companies as well as produce a high quality energy resource from these environmental problem areas.

SOIL SUBSTITUTE

MDC is currently involved in a research and development program in conjunction with the Southern Illinois University to develop a method for production of a "soil substitute" from pond slurry. The materials would be produced from the reject circuits of the recovery plant. This product would be acceptable to the regulatory agencies as a topping medium for use in site reclamation. This material would also be transportable as a discrete product in addition to the coal product.

The current plan is to have this process to a prototype plant in operation within twelve months.

ULTRA-FINE PRODUCTS

The MDC fine coal product exhibits the high quality, low moisture and good handling characteristics due to the removal of the minus 200# material. In the minus 200# material rejected from the MDC process there can be significant quantities of energy product. With advances in new techniques such as column flotation and new product mediums such as coal/water mix fuels, the MDC process is well placed to provide a feed stock for such technologies with the benefit of no additional costs of dredging and feed materials handling.

SUMMARY

The recovery and treatment of fine coal reject from slurry ponds can be economic if the operation is approached with a logical and reasoned evaluation of the resource and treatment methods. Historically the "carbon recovery" industry has operated on limited tonnage basis and in a climate of "operate when prices allow". When supported by appropriate technologies, the business of fine coal recovery and treatment can compete with traditional coal mining operations in the range of economic conditions experienced by the coal industry. The combination of unique technologies and evaluations allows MDC not only to compete on a value for energy basis but also provides an added reclamation benefit to the coal industry.

Compacting Coal and Coal Fines into Coal Logs: A Discussion of the Process

by

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Capsule Pipeline Research Center
University of Missouri-Columbia

INTRODUCTION

The process of compacting coal, a heterogenous, porous, friable, viscoelastic material, into a cylindrical log shape, with dimensional stability and strength to travel many miles transported by water in a pipeline, has been under intensive R&D at the Capsule Pipeline Research Center, University of Missouri-Columbia, Missouri.

The desire for coal log pipeline is to take the coal at the mine and, instead of loading into a railroad hopper car or truck, to convert it into coal logs and inject them into a pipeline. Considering the fact that no two coal mines produce the "same coal" and that coal properties vary, it became obvious that a coal log fabrication method would have to be tailored to the coal. Also considering the research was being done at a university, some fundamental studies of an academic nature of the what and why of coal log fabrication were undertaken.

This paper highlights the results of the research effort to date on the fabrication of the coal logs for entry into the pipeline system. The fabrication process consists of three basic steps, a) coal particle preparation, b) mixing the coal particles with additives, and c) compacting the mixture into coal logs.

COAL PREPARATION

Coal at the mine is crushed to give a uniform size distribution with minimum fines generation, with a top size of 2" to 3". Used directly, uniform coal particle size distribution at half inch top size led to weak coal logs unless excessive (uneconomical) amounts of binder were used. By reducing the top size to 0.6 millimeter and using a grinder which gives a wide particle size distribution, the requirement of binder and/or process condition became economical. Ideally, the particle distribution giving the maximum bulk density is optimum. The effect of grinding with a roll mill and a hammer mill on particle size distribution is shown in Fig. 1. The equation indicating maximum density is simply weight percent passing equals the ratio of the diameter of the next smaller (D_P) particle to the diameter of the largest particle (D_L) raised to the 0.5 power. We have found experimentally this relationship gives the maximum bulk density for several different coals. We have also found that under compaction, coal particles break and the distribution shifts towards the maximum distribution. To illustrate this, coal particles screened to give a 4.75 to 3.5mm diameter cut were compacted at various pressures, and the resulting distribution determined.

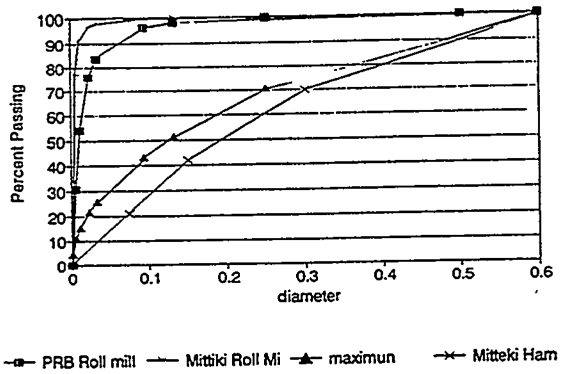

Fig. 1. Effect of Grinder Type
(30 mesh .6mm Topsize)

These data are shown in Fig. 2. This breakage produces uncoated surface (when binder is used) and weakens the log, enforcing the need for initial maximum bulk density coal particle distribution. For binderless logs breakage is not expected to be a problem.

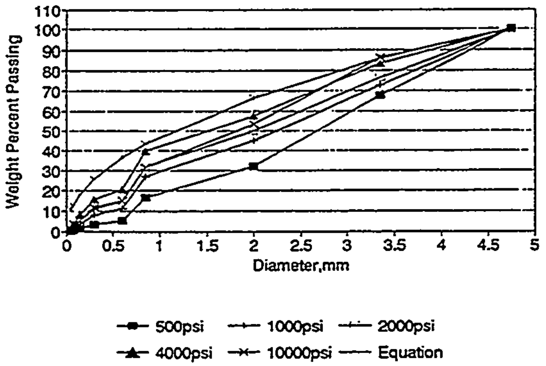

Fig. 2. Particle Breakage Effect of Pressure
(1.75"Die -4.75+3.5mm Antelope)

COMPACTION

Compacting coal is not a new endeavor. Our requirement of strength under water, wear-resistant in pipe, and low-cost fabrication led us to the use of asphalt as one path and the use of high temperature (150-200°C) and high pressure (20,000 psi) as a non-binder path.

The use of road asphalt required the need to dry the coal and mix with the hot asphalt. We were not able to make a log with an economical amount of asphalt. This work did show the difficulty of mixing the coal-asphalt to obtain uniform mixture.

The shift to emulsified asphalt allowed mixing and compacting undried coal. Work using this type binder again proved the need of good mixing. The mixing requires the coating of a large amount of surface with a very small amount of emulsion. Diluting the emulsion with water to give a higher liquid-to-solid ratio has proven to be satisfactory when used with suitable mixing. The combination of emulsion, temperature, and compaction pressure has also been shown to be a satisfactory system.

The key to successful coal log fabrication is the compaction process. Given a high bulk density and a uniform mix, the process compaction conditions set the quality of the log and the equipment cost sets the economics.

An eight inch pipeline requires injection of seven logs a second to maintain capacity. These logs are 7.0" in diameter, 14" in length and weigh 25 lbs. The weight, dimensions, and strength of each log must be maintained at the 7 logs a second rate, with equipment economical to do the job.

A single-screw extruder would produce logs with uneconomical amounts of road-asphalt. When attempted to be used with the emulsion-water-coal mixture, two process properties were found. An extruder is a forming device, and requires a material which will flow into the form desired. A coal-water-binder mixture will flow only so long as sufficient liquid is present to fill the voids and maintained being the continuous phase of the mixture. Coal-water-mixture when placed under pressure wants to separate. As the coal particles becomes closer, filling the void space, the water is pushed out of the mixture. The extent of this reaction is a function of the pressure and the amount of excess water available. The work with a single-screw extruder showed that before a satisfactory log could be made, the coal-water mixture became non-flowing and the die is plugged.

The reaction of a water-coal mixture as pressure is applied is shown in Fig. 3 for a bituminous coal. As indicated, at pressures below 1000 psi the water-coal begins to separate. In an extruder, as the screw creates pressure on the mixture the water flows backward and the coal gets drier as it advances until the coal will not flow into the smaller section of the die and plugs the machine. The same phenominon was demonstrated on a RAM extruder.

This separation reaction takes place with movable piston compactors. Here, however, as the mixture gets drier and the water escapes around the pistons, the coal doesn't have to "flow" through a die. The resistance to volume reduction is the escape of water and the drag on the side of the mold by the particles moving toward the center of the mold. The resulting log does not have to "flow" out of the mold but can be pushed out with only the need to overcome the friction of movement on the side of the mold, no further compaction is needed.

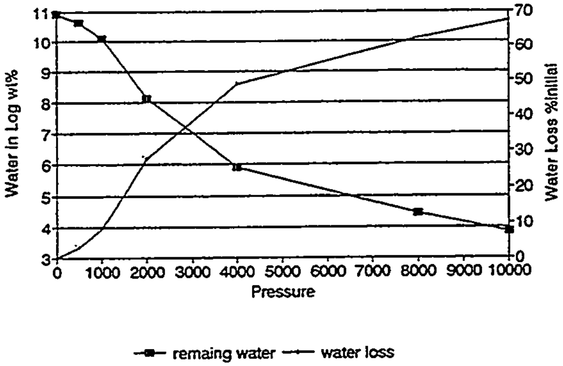

Fig. 3. Water During Compaction
(1.75"dia. Mitteki 20oc)

The need for water to escape during compaction generates a density profile along the compacted log. The water concentration varies both longitudinally and radially within the log. This density profile under two compacting conditions is shown in Fig. 4.

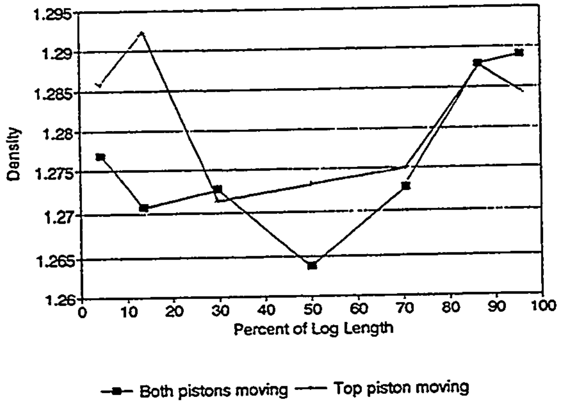

Fig. 4. Density Profile L/D = 2.0

These data show the resulting density effect of having both pistons moving during the compaction and only allowing one piston to move. Both methods show a lower density, thus a higher water concentration, in the center of the log.

Typical piston displacement-pressure curves are shown in Fig. 5a for a bituminous coal and 5b for a subbituminous coal. These curves were generated with the bituminous coal containing 20 wt %, water, the subbituminous containing 38 wt% water and both coals with 2 wt% asphalt binder. The pressure was increased at a constant rate to the 20,000 psi maximum, held for five minutes and then reduced at the same rate back to atmospheric pressure. This time is shown in Fig. 6(a&b). The log is then ejected using one of the compacting pistons. An increase on the displacement axis (y-axis) indicates a reduction in the volume of the log. Almost as soon as pressure is reduced the log begins to expand longitudinally and accelerates beginning at about 1000 psi. Expansion in radial direction happens as the log is ejected from the mold. Dimension change is about 5% in both directions. These profiles indicate that the softer subbituminous coal reacts similar to the harder bituminous coal during the compaction step but continues to compact during the constant pressure step. It then expands more during the pressure reducing step. The softer coal requires less stringent compaction conditions to make good coal logs.

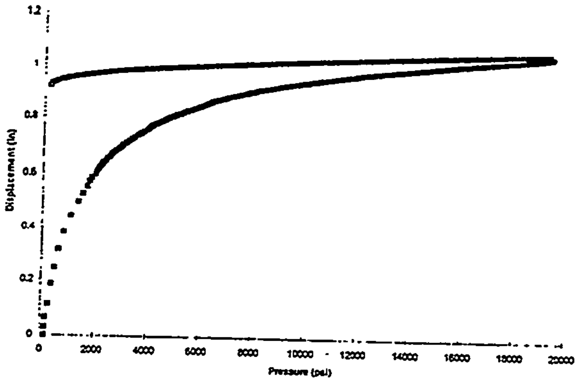

Fig. 5a. Compaction Profile
(Mettiki Coal, 1.75" Diameter, 97oC)

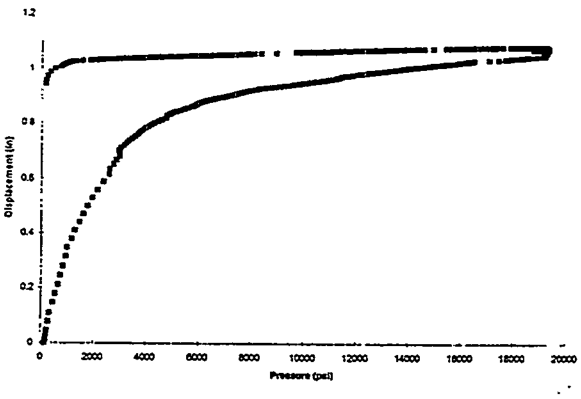

Fig. 5b. Compaction Profile
(PRB Coal, 1.75" Diameter, 97oC)

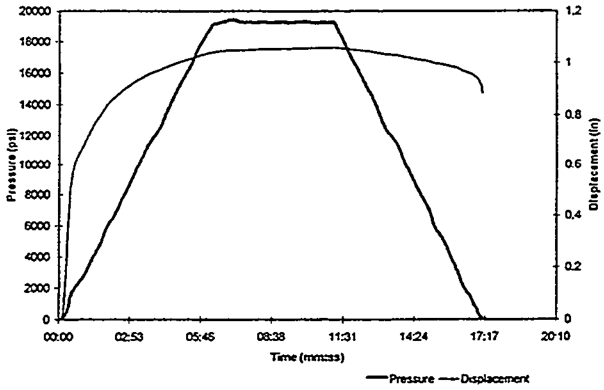

Fig. 6a. Compaction Profile
(Bituminous Coal)

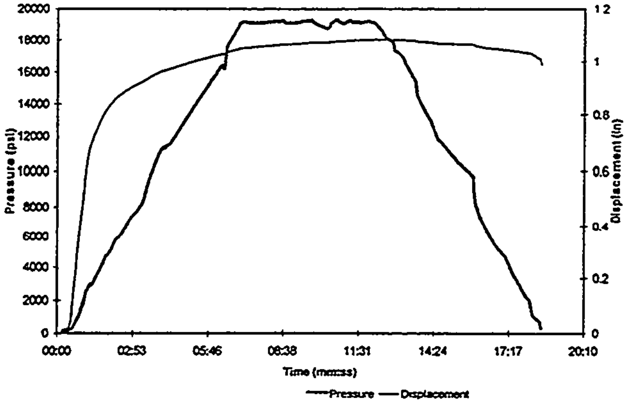

Fig. 6b. Compaction Profile
(Subbituminous Coal)

The application of temperature has allowed satisfactory logs to be made without binder. The need for loading and ejection of the logs while at these temperatures require doing so under pressure and until recently the machinery to do this has been uneconomical. A recent machine design may solve this problem.

SUMMARY

The strength of the coal log can be controlled by the amount of binder, the compacting temperature and pressure. Each of these have an effect on the cost of producing a coal log. Based on economics, the process conditions are tailored to the coal and to the shipping distance. The current research effort is to optimize the process conditions and to scale up the laboratory process to large (6"-8") coal log fabrication. Transportation in a 5 mile-long 6" pipeline has been demonstrated.

This paper has dealt with only the process considerations of fabricating the coal log. The entire pipeline system (injection, pumping, retrieval of the coal logs) has also been demonstrated in our laboratory.

Low-Cost Coal Fines Reconstitution Using the BioBinder Process, *R. Woodworth, J. Kelly and G. Miller, Altex Technologies Corporation, USA*

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This paper was unavailable at the time of publication and may be obtained directly from the author.

Developing the Solution for Accurate, Metered Feeding of Coal Fines and Resulting Boiler Performance Improvements, *Robert Black, Pennsylvania Power & Light; Jurgen Brat, CQ Inc., and Timothy W. Saunders, Stamet, Inc., USA*

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This paper was unavailable at the time of publication and may be obtained directly from the author.

Progress Report on the Proof-of-Concept Demonstration of Mulled Coal Technology for the Use of Wet Coal Fines, *Russell C. Maxwell, Energy International Corporation, USA*

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This paper was unavailable at the time of publication and may be obtained directly from the author.

DEVELOPMENT OF LIDS™ AS A PART OF THE BABCOCK & WILCOX LOW EMISSION BOILER SYSTEM

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ABSTRACT

Babcock & Wilcox (B&W) is currently developing an emissions control system capable of dramatically reducing SO_x and particulate emissions while addressing the concerns of solid waste generation and air toxics regulation. The work is being performed as an integral part of B&W's development of an advanced low-emission boiler system in a project entitled "Engineering Development of Advanced Coal-Fired Low Emission Boiler Systems (LEBS)." The program is sponsored by the U.S. Department of Energy's Pittsburgh Energy Technology Center (DOE - PETC). The overall goal of the DOE's program is to dramatically improve environmental performance and thermal efficiency of conventional, Rankine cycle, coal-fired power plants.

The initial phase of the LEBS multi-phase project required a thorough review and assessment of potential advanced technologies and techniques for control of power plant emissions. For the control of SO_x and particulate emissions, numerous near-term advanced flue gas cleanup options were assessed and evaluated. On the basis of these results and other qualitative considerations, the B&W Limestone Injection With Dry Scrubbing (LIDS™) process was selected for further development and evaluation in B&W's LEBS project. The B&W LIDS process is a limestone-based furnace injection/dry scrubbing SO_2 removal process. The process comprises the cost-effective integration of three commercially-proven flue gas cleanup technologies: furnace limestone injection, dry scrubbing, and pulse-jet fabric filtration.

This paper describes the results of the B&W Phase I LIDS process development and evaluation. Highlights include:

- Results of pilot-scale testing performed to demonstrate the feasibility of exceeding the B&W LEBS project SO_2 removal goal of 97% with a fully-integrated LIDS process under cost-effective operating conditions.
- Mercury removal data. Mercury emissions from coal-fired power plants are a matter of intense debate. It seems likely that mercury emissions will be regulated in some manner in the latter half of the decade.
- Results of a LIDS solid by-product engineering study on disposal characteristics and possible utilization specifications. B&W is aggressively pursuing ways to reduce or utilize the solid waste produced by its various FGD processes — including the LIDS process — through a variety of development programs.

COMBUSTION 2000 PROGRAM BACKGROUND

In order to improve environmental performance and thermal efficiency of future coal-fired power plants, the U.S. Department of Energy's Pittsburgh Energy Technology Center (DOE - PETC) has initiated a program called Combustion 2000 to address the clean and efficient use of coal for power generation for the first decade of the 21st century.[1] The Combustion 2000 program has two parallel engineering development activities that are expected to improve environmental performance and thermal efficiency of future coal-fired power plants: the Low-Emission Boiler System (LEBS) and the High-Performance Power System (HIPPS). The work presented in this paper was conducted under the LEBS portion of the Combustion 2000 program.

LEBS workscope calls for the development of a new boiler design equipped with improved combustion and heat transfer subsystems and advanced environmental control technologies capable of achieving emissions of SO_x, NO_x, and particulates far below current New Source Performance Standards (NSPS). The LEBS will use relatively near-term developing technology to dramatically lower SO_2, NO_x, air toxics, and particulate emissions, while keeping the cost of electricity comparable to that of conventional technology.

Major system development efforts are required for LEBS. In order to be successful, all major plant subsystems must be integrated. The new plant needs to be designed from the ground up with the new technologies in mind. Therefore, to ensure expertise in a broad range of technical areas, subsystem evaluation teams were formed. These Phase I subsystems are illustrated in Figure 1 and include: 1) NO_x control, 2) SO_x/particulates/air toxics/solid waste control, 3) boiler island, and 4) balance-of-plant.

Phase I of the LEBS program was completed over a 24-month period and included a LEBS concept evaluation and selection, a research and development test (R&DT) plan formulation and execution, and the creation of a preliminary commercial generating unit design. This paper covers highlights of the work completed in the SO_x/particulates/air toxics/solid waste control subsystem in Phase I.

SO_x/PARTICULATES CONTROL SUBSYSTEM

The SO_x/Particulates Control Subsystem consists of the SO_2 and particulate matter emissions control systems for the LEBS plant. Candidate SO_2 control processes considered included pre-combustion coal cleaning processes, various types of sorbent injection processes, and various types of dry and wet scrubbing processes. Other considerations to be addressed were control of hazardous air pollutant emissions (commonly called air toxics) and solid by-product disposal and/or utilization.

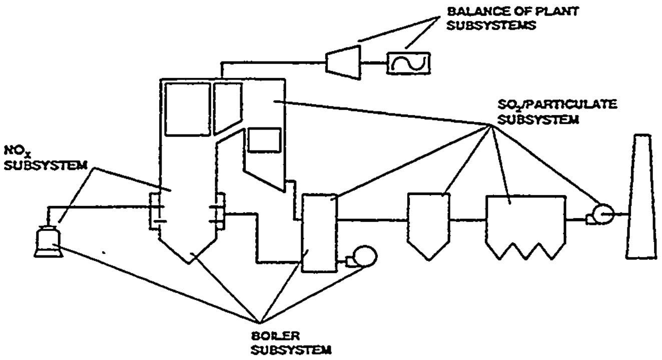

Figure 1. Phase I Subsystem Evaluation Teams

The specific goals for the SO_2/Particulates Control Subsystem were:

- SO_x — Greater than 97% SO_2 removal.
- Particulate — Less than 0.015 lb/10^6 Btu of fuel input.
- Any processes selected for application in the LEBS plant have the potential to comply with possible emissions control regulations for targeted air toxics.
- At a minimum, the selected processes produce an environmentally benign solid by-product. Recycle in the process (regeneration) or elsewhere (by-product utilization) is desirable.
- The selected processes integrate with other plant subsystems to yield optimal overall performance of the LEBS plant.

EVALUATION OF SUBSYSTEM CONCEPTS AND RESEARCH, DEVELOPMENT, AND TEST PLAN FORMULATION

The candidate screening process (Figure 2) began with the development of a "long-list" of candidate technologies for the LEBS plant subsystems. The technologies selected included promising processes that have been developed and installed on full-scale utility systems, pilot plants, or have received extensive bench-scale testing. The applicability of the different control technologies was then evaluated using a method developed by Charles Kepner and Benjamin Tregoe (K-T analysis).[2] The K-T analysis uses a systematic methodology to make qualitative decisions by establishing the criteria to be achieved, measuring each of a number of alternatives against these criteria, and then summarizing the results. This method of screening produces a numerical performance total for each of the technologies. The technology achieving the highest performance total is thus predicted to be the most suitable for achieving the objectives.

The final list of technologies selected in the K-T analysis was subsequently evaluated in detail — technically and economically — incorporating parameters specific to LEBS as well as the potential impact of the candidate concepts on all other plant subsystems. The analysis resulted in final recommendations for each B&W LEBS plant subsystem. The primary power plant concept developed by the B&W LEBS team is shown in Figure 3. The B&W LEBS system analysis and selection process has been documented in several technical papers.[3]

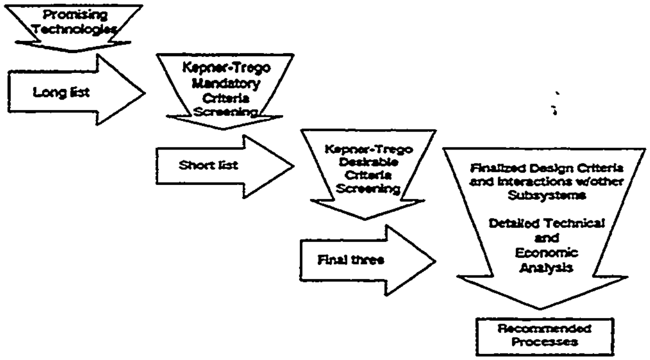

Figure 2. Methods of Process Selection

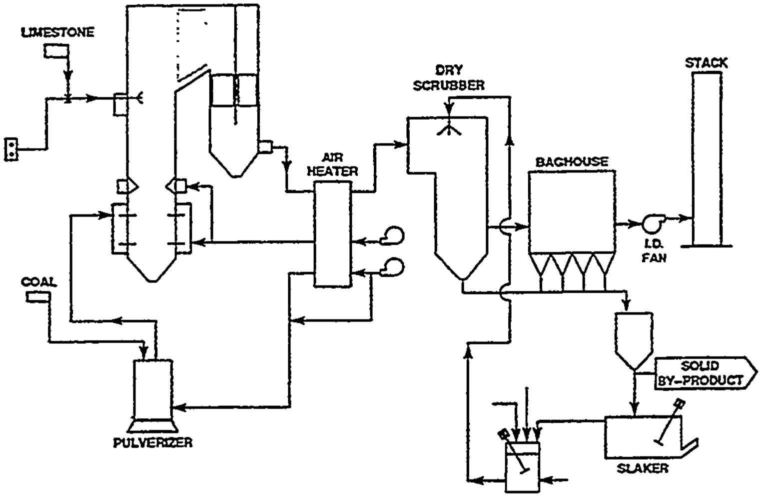

Figure 3. Phase I B&W LEBS Team Primary Power Plant Concept

A preliminary engineering design and a design uncertainties analysis were then completed to identify areas of uncertainty in the selected LEBS concept. For each major uncertainty, a rank according to the importance of the deficiency and the recommended approach to generating the required information was provided. Based on the areas of uncertainty identified, an R&DT plan was created to eliminate and/or reduce the deficiencies.

On the basis of results compiled during the concept selection process described above, the B&W Limestone Injection With Dry Scrubbing (LIDS) process was selected for further development and evaluation in B&W's LEBS project.

Limestone Injection With Dry Scrubbing (LIDS) Process

The B&W LIDS process — shown in Figure 4 — is a limestone-based furnace injection/dry scrubbing SO_2 removal process. The process actually comprises the cost-effective integration of three commercially-proven flue gas cleanup technologies: furnace limestone injection, dry scrubbing, and pulse-jet fabric filtration. Sulfur dioxide removal occurs in the boiler furnace and convection pass, in the dry scrubber, and in the fabric filter. Limestone is pulverized and injected as a dry powder into the flue gases in the upper furnace cavity of the boiler. Upon injection, the limestone undergoes calcination to form lime, a portion of which reacts with SO_2 in the flue gases forming calcium sulfate. The flue gases exiting the boiler then pass through a dry scrubber reactor where they are contacted by a slurry containing calcium hydroxide. In the dry scrubber, the flue gases are cooled and humidified to conditions near the water saturation temperature — commonly referred to as operation at a "low approach to saturation temperature". Under these conditions, a portion of the SO_2 in the gases reacts with the calcium hydroxide. Water contained in the slurry droplets evaporates as the droplets pass through the reactor vessel, and the products leave the dry scrubber as a dry powder still suspended in the flue gases.

Finally, the flue gases enter the pulse-jet fabric filter (baghouse) where coal fly ash, spent sorbent, and unreacted sorbent particles are collected. The use of a baghouse is a key feature of the LIDS process because of the additional SO_2 removal it yields as the flue gases pass through the sorbent-containing filter cake on the filter bags. The majority of the solids from the particulate collection device are recycled to a reagent preparation system to produce calcium hydroxide slurry for the dry scrubber. The remainder of the solids collected are conveyed to disposal and/or utilized.

The furnace limestone injection process facilitates the cost-effective use of a dry scrubber for downstream SO_2 removal by: 1) permitting the use of limestone as the sorbent (as opposed to the more expensive lime used in most dry scrubbing processes), and 2) by reducing the inlet SO_2 concentration to the dry scrubber through the in-furnace removal of SO_2. This latter fact permits the LIDS process to be applied to units firing high-sulfur coals by lessening the amount of calcium needed in the dry scrubber, thereby reducing the heat needed to evaporate the water contained in the sorbent slurry fed to the dry scrubber.

The remainder of this technical paper covers highlights of the LIDS development work performed in support of the B&W LEBS during Phase I of the project.

LIDS PILOT-SCALE FEASIBILITY TESTING

The feasibility of achieving the project SO_2 removal goal with a fully-integrated LIDS process under cost-effective operating conditions needed to be demonstrated. While previous data from a variety of sources were used to predict success, it was nevertheless essential to demonstrate this capability early in the project. The primary objective of the LIDS feasibility testing was to show the fully-integrated LIDS process capability of exceeding 97% SO_2 removal.

In the furnace limestone injection process, calcination of the limestone must occur under ideal conditions — injection temperature, time/temperature history, etc. — in order to yield a highly reactive sorbent for optimum SO_2 capture in the boiler furnace and convection pass, and for optimal performance of the downstream dry scrubbing and fabric filter SO_2 removal processes. In the dry scrubber, the SO_2 removal efficiency is maximized by increasing the amount of sorbent in the slurry

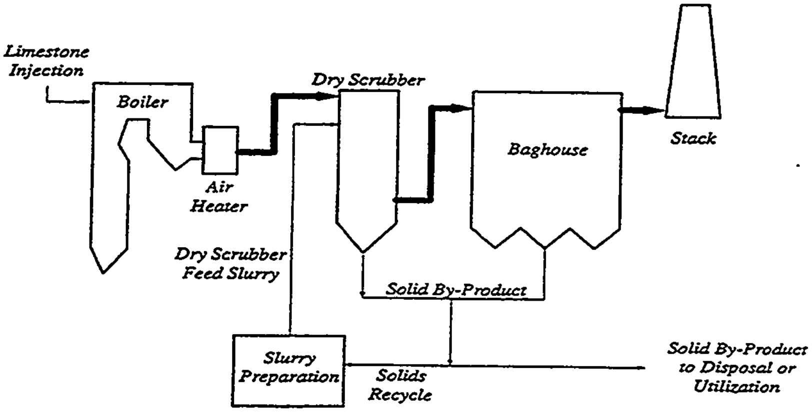

Figure 4. B&W LIDS Process

and by operation near the flue gas saturation temperature. The baghouse represents the final SO_2 removal stage for the LIDS system. Therefore, baghouse performance is critical to the overall SO_2 removal in LIDS. Key factors for the baghouse performance include the approach to saturation temperature, calcium stoichiometry, sorbent characteristics, and baghouse differential pressure (thickness of the filter cake). All of these things are examples of the process integration that needed to be accomplished during the fully-integrated LIDS feasibility demonstration.

Small Boiler Simulator (SBS) and LIDS Pilot Facility

The facility used for the LIDS pilot-scale testing is shown in Figure 5. It is made up of a combustion furnace called the small boiler simulator (SBS) and the LIDS pilot facility. The test unit is located at B&W's Research and Development Division in Alliance, Ohio. The main components of the SBS facility consist of the coal preparation and feeding system, SBS furnace, and gas monitoring system. Equipment specific to the LIDS facility includes the LIDS dry scrubber, slaker, baghouse, limestone feeder and injectors, a slurry and atomization system, and additional gas analyzers.

The 5-million Btu/hr SBS is a wall-fired, vertical furnace that simulates the characteristic geometry of B&W's commercial pulverized-coal-fired boilers. For LIDS testing, limestone is injected into the SBS furnace at the inlet to the convection pass. Injection temperatures are maintained between 2100 - 2300 F to maximize calcination and SO_2 removal. A heat exchanger is installed between the exit of the SBS and the dry scrubber. Its purpose is to cool the flue gas leaving the SBS to temperatures representative of utility boiler air heater exit temperatures.

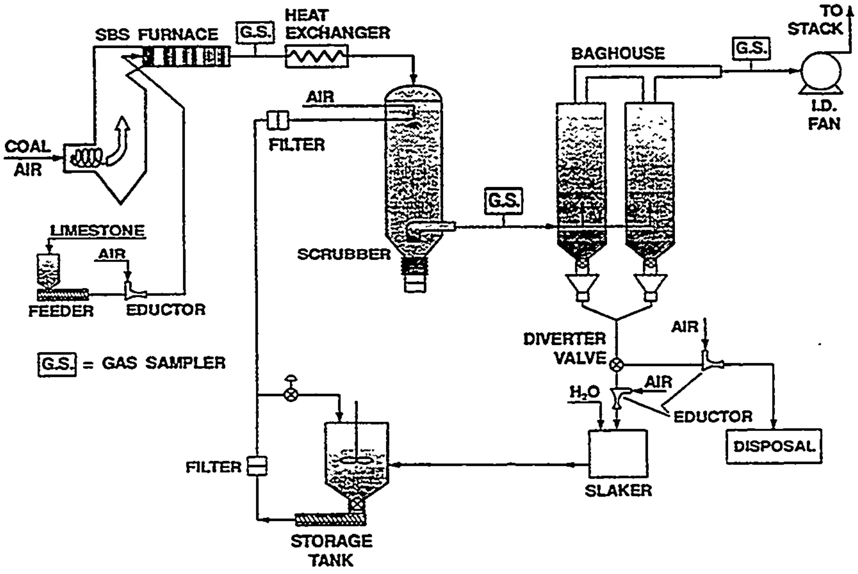

Figure 5. LIDS Pilot-Scale Test Facility

The dry scrubber is a cylindrical, vertical downflow reactor measuring 5 feet in diameter and 17 feet long from the B&W DuraJet™ atomizer to the exit. Flue gas enters the top through an expansion containing several flow straightening devices. The flue gas exits near the bottom of the chamber and flows to the baghouse. The pulse-jet fabric filter (baghouse) consists of two modules each containing twenty-three 4-5/8 inch diameter by 10-foot long bags. The gas analysis system installed on the SBS/LIDS facility consists of analyzers that continuously monitor and record the emissions at the inlet to the dry scrubber and the exits of the dry scrubber and baghouse.

Approach

During the LIDS feasibility demonstration, an Ohio #6 coal with 3.0 lb sulfur/10^6 Btu was burned in the SBS. The furnace convective pass, the dry scrubber, and the pulse-jet fabric filter were fully-integrated to achieve the SO_2 removal goal. The demonstration conditions were as follows:

Stoichiometric Ratio (Ca/S)	1.4 mole Ca/mole S
Approach to Adiabatic Saturation Temperature	10 F
Slurry Percent Solids	44%

These conditions were held relatively constant throughout the demonstration to obtain data above the 97% SO_2 removal goal over an extended period of time.

Results

The fully-integrated LIDS feasibility demonstration surpassed the goal of 97% SO_2 removal. The LIDS system removed 98% of the SO_2 based on an average of the data measured during the test series. Exceeding the SO_2 emissions goal using the B&W LIDS process required optimal SO_2 removal performance of the entire integrated system. Average values from the individual unit operations included 23% removal in the furnace, 58% in the dry scrubber, and 95% removal in the baghouse. These values are based on the inlet and outlet SO_2 concentrations of each operation. Figure 6 shows the total SO_2 removal accomplished at the outlet of each piece of the LIDS process.

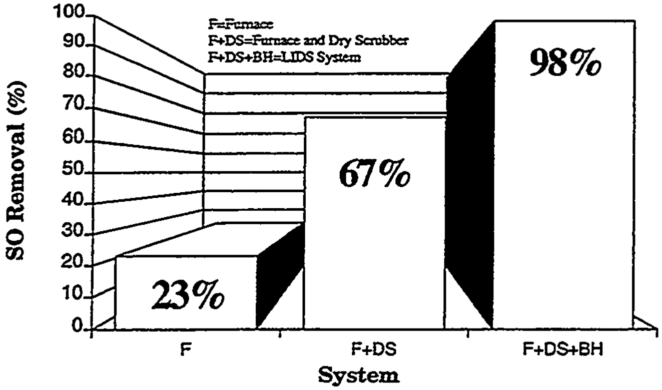

Figure 6. LIDS Demonstration SO_2 Removal by System Component

The furnace injection process performed well as an integral part of the LIDS system. The expected furnace injection SO_2 removal was achieved based on past applications. The SO_2 removal in the furnace was not optimized during the feasibility demonstration. There are factors that may influence the amount of SO_2 capture that were not addressed in this phase of the project, including limestone injection temperature (injector placement) and limestone particle size distribution. These issues will be addressed in future phases of LIDS testing.

The dry scrubbing and baghouse processes also performed well as an integral part of the LIDS system. The modest SO_2 capture achieved in the dry scrubber is to be expected because, as the total SO_2 removal of the integrated system improves (including significant baghouse SO_2 capture), the solids returned to the dry scrubber in the slurry become less reactive (the useful Ca/S entering the dry scrubber decreases). Based on qualitative observations, the dry scrubber and baghouse can be operated at an approach to saturation temperature in the range of 10 F without operability problems. The dry scrubber was operated without any excessive buildup and the hopper ash was easily removed. The baghouse solids felt dry, flowed freely, and were easily transported to the slurry preparation system.

AIR TOXICS

Title III of the Clean Air Act of 1990 mandated that the Environmental Protection Agency (EPA) evaluate emissions of 189 air toxics from the electric power plants. Since then, researchers have studied the trace emissions of heavy metals (including mercury), acid gases, and organics from power plants; their transport and fate in the environment; and health impacts. Currently, how these results will impact potential regulations is uncertain. Because air toxics regulation is somewhat of a wildcard, it was important to evaluate the B&W LEBS against potential air toxics areas of concern. The following is a summary of where potential air toxics regulations may be headed and how the B&W LIDS system stacks up against them.

Mercury

Mercury emissions from coal-fired power plants are a matter of intense debate. It seems likely that mercury emissions will be regulated in some manner in the future due to its potential to bioaccumulate in the food chain. Since mercury is a likely target for regulation, screening tests of mercury emissions from the LIDS system were performed by Frontier Geosciences, Inc., to characterize the process in terms of its mercury capture capabilities. The Frontier Geosciences measurement method has been documented in many technical papers; details of the method can be referenced in these papers.[4] Measurements were made at the dry scrubber inlet, dry scrubber outlet, and baghouse outlet during the LIDS feasibility demonstration discussed above. LIDS overall mercury removal results are presented in the Table 1.

Values recorded for total mercury (Hg) clearly indicate a repeated pattern of mercury reduction across both the dry scrubber and baghouse. Total mercury removal averaged 97% across the LIDS system (Figure 7).

Trace Metals, Acid Gases, and Organics

Particulate regulations, although particulate itself is not an air toxic *per se*, may well become more stringent. A reason for this is that many of the trace elements of interest condense onto the particulate as the flue gas cools through backend emissions control equipment. Therefore, particulate removal systems have the potential to remove large amounts of air toxic substances associated with the particulate matter. A pulse-jet fabric filter operating at low temperatures is an integral part of the B&W LIDS system. This gives LIDS two advantages in terms of air toxics removal: 1) fabric filtration has been shown to be extremely effective in removing particulate and therefore many trace metals, and 2) the LIDS dry scrubber and baghouse are operating at low temperatures creating a higher potential for trace metals to condense out of the flue gas.

Acid gases — primarily HCl and HF — have come under scrutiny primarily because of the large quantity of these substances emitted from electric power utilities. If acid gas regulations are required, coal-fired plants equipped with wet or dry FGD can achieve over 90% removal of HCl, as confirmed in field sampling.[5] Another form of HCl and HF control may be furnace sorbent injection.[6] LIDS has both furnace limestone injection and a dry scrubber to effectively control acid gases.

Table 6
LIDS TOTAL MERCURY REMOVAL

Sample ID	Stream	Total Hg ug/Nm^3	Hg Removal Across System %
A1	Inlet Scrubber	5.0	99.06%
A3	Inlet Baghouse	2.49	
A2	Outlet Baghouse	0.047	
A4	Inlet Scrubber	4.91	98.64%
A6	Inlet Baghouse	2.33	
A5	Outlet Baghouse	0.067	
A7	Inlet Scrubber	6.19	93.41%
A9	Inlet Baghouse	2.17	
A8	Outlet Baghouse	0.408	

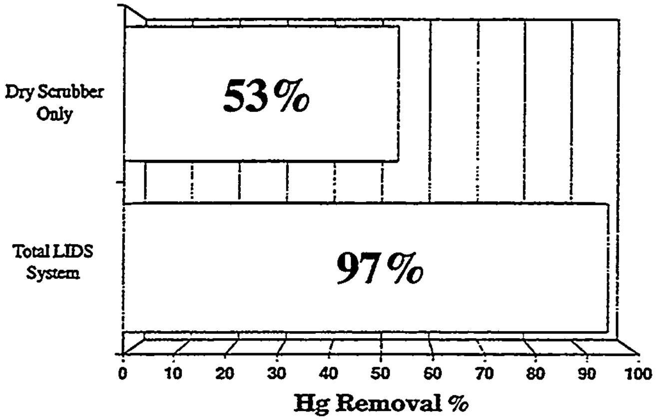

Figure 7. LIDS Total Mercury Removal

Preliminary measurements appear to indicate that, compared with trace metals, organic compounds in utility power plant emissions pose only a very small risk to human health or to the environment due to their typical presence at or below the detection limit of current EPA-recommended measurement methods. However, if organics such as furans and dioxins are regulated, LIDS has the potential for their control because of the use of a calcium-based sorbent injection process in the furnace/convection pass. The amount of HCl present in flue gas at lower temperatures — around 570 F — has been shown to directly affect the amount of dioxins and furans formed.[7] Therefore, injecting a compound that will react with HCl — such as limestone in the B&W LIDS process — may reduce the quantity of dioxins and furans formed.

Overall, the B&W LIDS process should be able to respond effectively to any of the potential air toxics regulations that have been indicated to date.

LIDS SOLID BY-PRODUCT UTILIZATION

Coal-fired utilities currently generate a large amount of solid by-products due to standards adopted to control the emission of SO_2 from electric power stations. In the past, solid by-product generation has been a low priority issue. However, recently it has become evident that it is important to assess the characteristics of the by-products produced by coal-fired utilities in terms of their readiness for disposal as well as utilization prospects. This is in large part due to increasing public and governmental pressure. An important objective of the B&W LEBS development work is to ensure that at a minimum, the system produces a solid by-product that is considered benign for environmentally acceptable disposal, with the more desirable option being to develop a system that utilizes or reuses the solid by-product.

To deem the LIDS solid by-product as environmentally benign for disposal, results of hazardous waste tests and water quality standards tests were obtained. All characteristics were below hazardous waste definitions. The results for all organic, volatile organic, and metal analytes were well below the EPA toxicity characteristics leaching procedure (TCLP) regulatory limits. The metals in the LIDS water leachate were well under the RCRA water quality criteria.

Utilization is an attractive economic and environmental alternative for management of the high-volume wastes frequently generated by energy production. Clean coal technology by-products, such as the LIDS residue, have been utilized as engineering and construction materials in numerous applications, and uses continue to be developed. The approach to the utilization of the LIDS by-product in Phase I was to follow the developments of independent solid waste management programs and to take advantage of knowledge gained. The Energy and Environmental Research Center (EERC) acted as a consultant on the solid by-product issues.

Based on high potential utilization applications, LIDS characteristics and laboratory- and field-scale characterization and demonstration projects for similar clean coal technology residues, the following utilization applications warrant further scientific and engineering investigation:

- Abatement of acid mine drainage and mine reclamation
- Structural fill and other fills, including controlled low-strength material
- Soil amendment (road base, subbase, and agriculture)

Demonstration projects will be required to evaluate performance and general interest in the utilization of LIDS materials. Testing will occur in Phase II of the B&W LEBS project and will focus on confirmation of strength parameters.

PHASE I LIDS CONCLUSIONS

Highlights of the Phase I LIDS R&DT results obtained in support of the B&W LEBS were as follows:

- The feasibility of achieving the project SO_2 removal goal of 97% with the fully-integrated LIDS system was demonstrated.
- LIDS total mercury capture was shown to be 97%.
- The LIDS solid by-product was found to be environmentally benign for disposal and potential utilization strategies were identified.

PHASE II LIDS ACTIVITIES

In Phase II, LIDS development and demonstration will continue. This will allow further definition of the LIDS process as well as progress in exceeding the original project goals. Key activities planned for B&W's LIDS process in LEBS Phase II include:

- LIDS process definition testing at 5 MBtu/hr—January 1995. This test series is aimed at further defining LIDS in terms of its optimum configuration to cost-effectively achieve the LEBS SO_2 removal goal. Mercury measurements are included for comparison to results obtained in the Phase I LIDS feasibility demonstration.
- LIDS subsystem testing in B&W's 100 MBtu/hr Clean Environment Development Facility—early 1996. The tests will primarily concern the optimization, operability, and sensitivity of the LIDS process as defined in the earlier pilot-scale tests. The tests will also provide an opportunity for the characterization of air toxics emissions and solid by-product streams.
- A preliminary engineering design of the proof-of-concept (POC) facility. The POC facility is to be constructed and operated in Phase IV of the LEBS program.

The presentation made regarding this technical paper will provide an update on results obtained during the LIDS process definition testing.

ACKNOWLEDGMENT

The authors express their thanks to the U.S. Department of Energy (PETC) for supporting the B&W Low Emissions Boiler System (LEBS) development (Contract Number DE-AC22-92PC92160). The assistance of Tony Mayne as technical monitor is appreciated.

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High Performance Power System (HIPPS) Plant Design and Economics

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ABSTRACT

The U.S. Department of Energy is sponsoring two multi-phase programs to evaluate and demonstrate High Performance Power System (HIPPS) fueled with coal. The major goal of the HIPPS program is to develop a high efficiency, low emissions, coal-fired power system for central power plant application early in the twenty-first century.

This paper describes the Phase I efforts for the HIPPS program carried out by a team lead by United Technology Research Center (UTRC). Phase I included research and tests to define the power generation and emissions performance capabilities of a HIPPS plant design using a High Temperature Advanced Furnace (HITAF) to heat air for a gas turbine. The Phase I near-term commercial HIPPS plant design allows natural gas to supply up to 35% of the total energy input, supplementing the coal combustion in the HITAF, and boosting the air temperature to the level required for modern gas turbines. The ultimate HIPPS goal calls for an all coal design.

The major objectives for the HIPPS program set by DoE are:

- Plant efficiency ≥ 47% (HHV)
- Nominal plant capacity suited to baseload plant, 300 MWe
- Air emissions at 25% of New Source Performance Standards (Phase I); more stringent requirements of 10% NSPS are set for subsequent Phases
- Cost of electricity at least 10% lower than modern coal plants with controls to meet NSPS

Phase I results, reported in this paper, include a commercial plant conceptual deign, capital and operating cost estimates, and economic evaluations. Phase II involves further analysis of commercial plant designs, component and sub-system demonstration, and engineering for a demonstration of technology on the order of 50 to 100 MWe. Phase III is the final engineering, installation, and operation of the HIPPS demonstration plant.

INTRODUCTION

Within the next decade, there will be a need for new generating capacity to meet increased demand and to replace existing units that are reaching the end of their service lives. The desires to use low-cost indigenous fuels, i.e., coal, to meet increasingly stringent environmental regulations, and to reduce energy consumption represent a challenge to the technical community. A major opportunity to meet this challenge was initiated by the Pittsburgh Energy Technology Center through its Coal-Fired High Performance Power Systems (HIPPS) program. During Phase I of this three phase program, a multi-disciplined Combustion 2000 Team (Table 1) led by the United Technologies Research Center has developed a conceptual commercial generating plant design of a nominal 300 MW HIPPS that meets, or exceeds, the major DoE goals for performance. This design is based on current state-of-the-art technology, or technology which could be commercialized by the year 2000.

Table 1- Combustion 2000 Team

Team Member	Major Responsibility
UTRC	Program Management, Heat Exchangers, Cycle Analysis
Turbo Power & Marine	Gas Turbine Technology
Power Technology, Inc.	
Physical Sciences, Inc.	Ash Depostion, Slag Screen, Control
University of North Dakota Energy & Environmental Center	Slag Interaction, Post Combustion SOx and NOx
Reaction Engineering, Inc.	Combustion Modeling, Burner Furnace Design
Oak Ridge National Laboratory	Ceramic Material
Bechtel Group	Overall Plant Design

The HIPPS conceptual design developed by the Combustion 2000 Team is shown in a simplified schematic in Fig. 1. The three major elements of the system are the High Temperature Air Furnace (HITAF), the gas turbine, and the steam system. The HITAF supplies 65% of the heat to a duct burner where natural gas boosts the temperature to that required by the turbine. The turbine exhaust stream, along with that from the HITAF, furnishes waste heat to a heat recovery steam generator (HRSG) and steam turbine. The overall efficiency of this system exceeds 47%, approximately 35% better than typical PC plants. Detailed descriptions of the equipment are given in the following sections.

In Fig. 1., it can be seen that the compressor discharge air is sent to a convective air heater in the HITAF. From there, it goes to a radiant heater located in the highest temperature portion of the HITAF, and then to the duct heater where the temperature is raised to the required combustor exit level. The exhaust from the gas turbine is split; one portion is sent to a "clean" HRSG while the remainder is used as preheated combustion air for the coal (Illinois No. 6) in the HITAF. The exhaust from the HITAF is sent to a "dirty" HRSG, baghouse, and FGD. To maintain the temperature required for a selective non-catalytic converter in the HITAF, as well as assure that the temperature for the convective section does not exceed 1800 F, a portion of the HITAF exhaust is recirculated. The steam bottoming system is atypical of currently installed combined cycles, with higher operating conditions of 2400 psi/1000 F/1000 F.

POWER PLANT DESIGN CRITERIA

The plant design scope is a non-site specific, greenfield power generation plant and includes all facilities required for power production; a mid-western location has been assumed. Technical and economic criteria for the study use the EPRI TAG^{TM} Volume 1: Rev. 7; June, 1993, as a basis. The HIPPS plant boundaries for

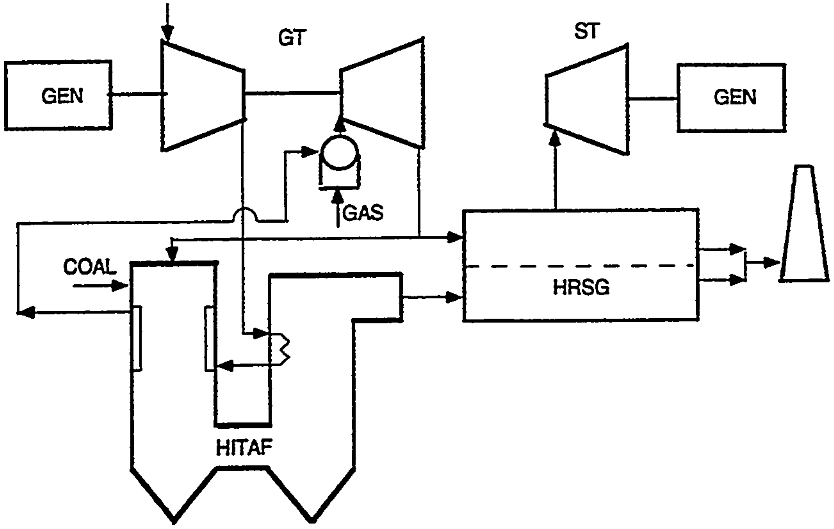

Fig. 1 - Simplified Schematic of HIPPS/HITAF

design and cost estimates include all the major operating systems such as the HITAF unit, a single modern frame-type gas turbine and a Rankine cycle steam bottoming turbine, heat recovery steam generators, environmental control equipment, auxiliary equipment and all support facilities needed to operate the plant (shops, offices, cafeteria, fuel handling and storage equipment, water intake structures, and waste treatment facilities). The plant includes the high voltage bushing of the generation step-up transformer, but not the switchyard and transmission lines. The switchyard and lines are generally influenced by transmission system specific conditions and hence are not included with the cost or design.

The nominal generating capacity of the HIPPS plant is 300 MWe. The plant is planned for a base loaded duty cycle with limited cycling capability. However, in recent years an important market aspect has surfaced in the power generation business: Load profiles are increasingly indicating a need for flexible plant operations, and new plant designs that account for daily output requirement swings. Thus, a minimum turndown of 50% is very desirable and is included. The nominal design life is 30 years. The net efficiency of the reference commercial plant is to be 47 % or higher (maximum heat rate 7,260 Btu/kWh, HHV) at ISO conditions. The plant burns Illinois No. 6 bituminous coal (EPRI TAG^{TM}) as the primary fuel. Coal is burned to supply 65 % or more of the total heat input to the HIPPS.

POWER PLANT DESCRIPTION

Figure 2 shows a detailed plan view of the HITAF and associated major power island equipment for the conceptual power plant. The facilities were designed and sized based on experience with similar pulverized coal-fired plants. The solid materials handling equipment and areas are designed for 60 days storage and an annual consumption of 350,000 tons. Conveyors, bins, storage piles and other items are sized for operations to efficiently receive and store materials, and to feed the HITAF units. The estimated land requirement is 49 acres inside the fence line. Solid waste disposal is offsite. The land requirement will vary over a wide range depending on the site and disposal method. Capital and operating costs for solids disposal are excluded from the plant estimate.

The major components shown in Fig. 2 are described briefly below. Except for the HITAF, and the way it is incorporated into the power plant, all the power generation and auxiliary equipment and processes are commercially available and used by the power industry.

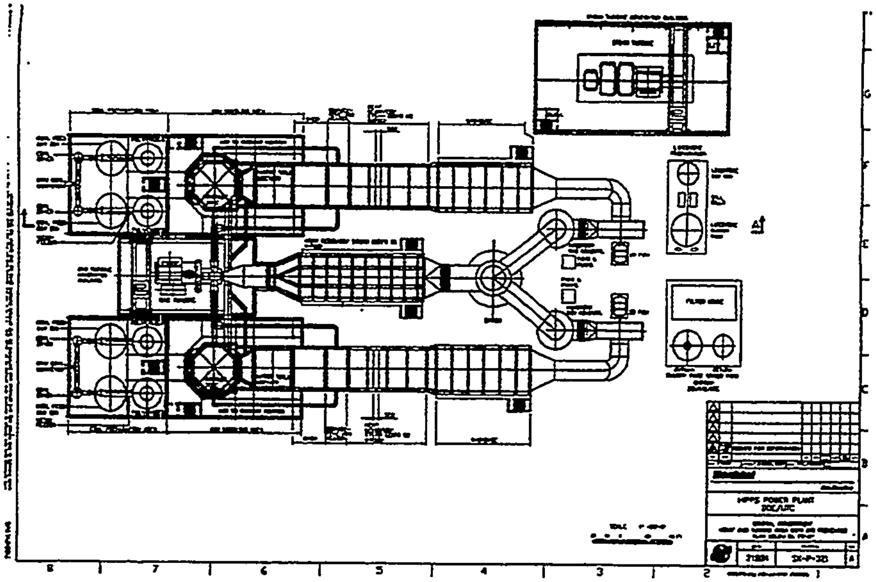

Fig. 2 Power Island Layout

Steam Generation Island

<u>High Temperature Air Furnace</u> The HITAF represents the major unproved technology in the commercial plant design. Two HITAF units (Fig. 3) of approximately 685 million Btu/hr each are used; the gas turbine

located between them. The octagonal-shaped combustor units are down fired with the coal injectors located around the top. A short, refractory-lined adiabatic section allows flame stabilization prior to the radiant section. Temperatures are high enough to keep slag running down the refractory walls of the radiant exchanger to a wet bottom collector. Finely pulverized coal and reburn air are introduced in the radiant section for NOx control.

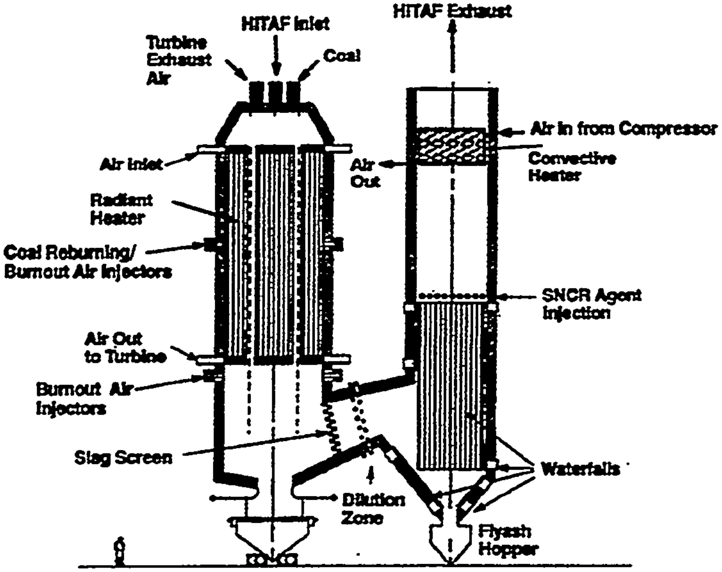

Fig. 3 HITAF Unit

The flow is turned 90° and passes through water-cooled slag screens and a waterwall section to freeze slag and reduce particulate carryover. A portion of the HITAF exhaust is recirculated to further reduce temperature and again turned 90° upwards before entering the SNCR and the convective heat exchanger. The exhaust exits the HITAF into a "dirty" heat recovery steam generator.

Considerable effort has been directed at identifying the refractory and heat exchanger materials. Currently, a castable alumina refractory is favored while a castable nickel-based alloy possibly lined with a newly developed FeCrAl alloy is proposed for the radiant exchanger. While use of an alloy limits the outlet temperature of the compressor discharge air to 1700 F, the fabricability and joining capability of the alloy lowers the risk of HITAF development. At the preliminary commercial plant level of conceptual design, the radiant and convective heat exchangers are included with the total HITAF cost estimate. Heat transfer from the coal combustion products to the gas turbine air is done in two steps: a radiant air heater in the combustion zone and a convective air heater in the furnace exhaust zone.

Radiant Air Heater The radiant heater will operate at a temperature high enough to ensure gravity-driven flow of molten slag from heater surfaces. This heater consists basically of many long hollow structural panels which will almost completely line the inside of the combustion zone (Fig. 4). The gas turbine compressor discharge air will be distributed to these panels by an arrangement of headers, manifolds, and ducts which will be staged to avoid excessive thermal stress. A ceramic refractory coating or tiles will be applied to the fire side of the panels to prevent slag-induced corrosion. Support for the radiant section will be provided by a massive structural shelf at the bottom of the furnace.

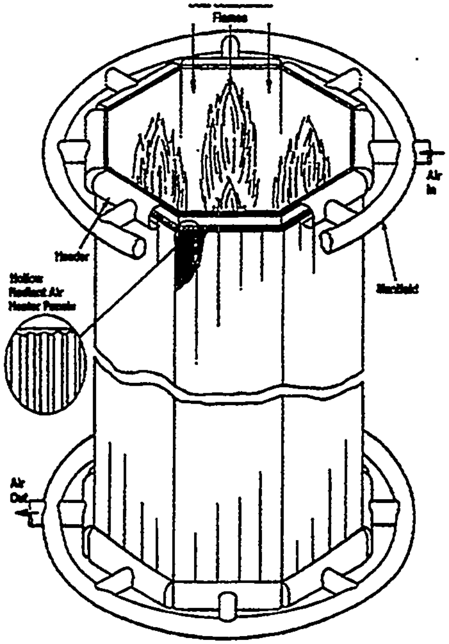

Fig. 4 Conceptual Design of Radiant Air Heater

Convective Air Heater The configuration for the convective heater is a modification of the shell and tube type where the air flows through banks of tubes and the hot combustion gases flows over and perpendicular to the tubes (Fig. 5). This modification, called the finned-tube-sheet, enhances the hot gas side conductance relative to the hot air side, provides additional structural rigidity, reduces circumferential thermal stresses in the tubes, and provides an aerodynamic shape which will reduce ash deposition on heater surfaces. A two-pass, cross-counter flow arrangement, with two air passes and one hot gas pass, provides the desired effectiveness without excessive pressure drop and with reasonable dimensions.

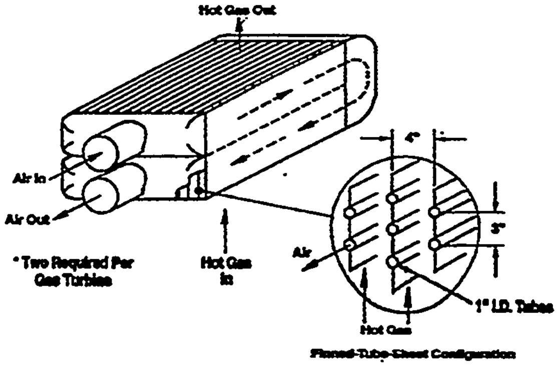

Fig. 5 Conceptual Design of Convective Air Heater

Heat Recovery Steam Generators Process Description Two HRSG's are used in the system to generate steam for electricity production. The "clean" HRSG (No. 1) operates with the gas turbine exhaust flow, and the two "dirty" HRSG's (Nos. 2A and 2B) use the HITAF flue gas as the source of thermal energy. Steam conditions are 1,000 F and 2,400 psia, with reheat to 1,000 F and 480 psia. The HRSG's supply a single steam turbine and share a common feedwater and condensate system.

Gas Turbine Generator

The frame-type gas turbine selected for the combined cycle system is based on the V. 84 design of Siemens. The Pratt & Whitney Division of UTC is assisting Siemens in the high temperature technology for this engine. The out board combustor configuration associated with early V. 84 engines has been retained for the this application to ease integration with the HITAF. Selected performance parameters (projected) are given in Table 2.

Table 2 - Heavy Frame Gas Turbine Characteristics

Inlet Flow - lb/sec	918
Rotor Speed - rpm	3600
Pressure Ratio	~16
Rotor Inlet Temperature - F	>2350
Exhaust Temperature - F	1005
Power - MW(nominal)	150

The duct burners which bring the hot turbine supply air to the desired turbine inlet temperature are located at the piping/engine interface and are arranged around the engine much like current can-annular combustor practice. The specially designed rapid mixing in these burners minimizes NOx from gas combustion as well as providing minimum size for ease of installation.

Steam Turbine

Two heat recovery steam generator sections are thermally integrated to generate 2,400 psia/1,000 F superheat and 480 psia/1,000 F reheat steam. The steam is used in a multiple pressure turbine to generate electric power. Steam from the low pressure turbine exhausts at 1.8 inches Hg absolute and is condensed by a shell and tube deaerating surface condenser.

Emission Control Systems

The systems selected for the commercial plant design are briefly discussed. Additionally, the Team is aware of parallel DoE and industry programs, such as the low emissions boiler system (LEBS). The programs' progress and data will be monitored to determine if equipment or processes can be used with the HIPPS commercial design or the demonstration plant.

Particulate Control Process Description The HITAF design uses a baghouse located between the HRSG and the FGD system for particulate control. This scheme provides adequate particulate emissions control, and is probably best available control technology (BACT).

Flue Gas Desulfurization Process Description The limestone forced oxidation process with a limestone-water slurry and throw-away product system is used for the commercial plant design. The process selection is based on engineering experience and the estimated process economics, auxiliary power requirements (pressure drop) and potential for sulfur reduction. The throw-away product assumes there is not a market for gypsum, and if that assumption were to change, the process design selection may be reviewed.

The scrubbed flue gas from the two absorbers is sent to a 350 foot tall stack and released to the atmosphere. Exhaust gas from the gas turbine heat recovery steam generator is sent to the same stack where it mixes with the flue gas and provides some reheat energy. The estimated temperature of the mixed stack gas is 157 F at the stack exit.

NO_x Control Process Description The HITAF design includes low NO_x burners in the coal combustor and both selective non catalytic (SNCR) downstream of the radiant heating section where the temperatures range from 1,600 to 1,900 F, and selective catalytic reduction (SCR) in a lower temperature zone of the HRSG. By using the SNCR as the primary NO_x control, the SCR can be relatively small and more cost effective. These controls are designed to reduce the NO_x to 0.15 lb. per million BTU of total energy input to the plant (coal/ natural gas).

The SCR process injects ammonia into the flue gas and passes the gas mixture through a catalytic bed with temperatures from 650 to 850 F. The ammonia reduces the NO_x to N_2 and H_2O. Commercial SCR technology in combination with low NO_x burners and overfire air systems can achieve 0.1 lb of NO_x per million Btu. The HITAF combination of SNCR and SCR is estimated to limit NO_x to below 0.1 lb/MMBtu.

Balance of Plant Facilities

The balance of plant facilities include the coal receiving, storage and handling processes; the limestone receiving, storage and handling processes; the water systems such as heat rejection, raw water supply and treatment, process and plant effluent treatment; support systems; and civil structures.

Design Point Performance

Table 3 shows the design point performance projections for the HITAF power generation plant. For reference, similar projections are shown for a commercial pulverized coal-fired power plant with FGD and SCR. The PC plant is more fully discussed below.

Table 3 - HITAF Generation and Emissions Performance

	300 MW HIPPS	300 MW PC Plant
Energy Inputs (MMBtu/hr)		
Natural Gas	740.4	None
Coal	1,370	2,960
Total	2,110.4	2,960
Generation Performance (MWe)		
Gas Turbine	161.0	None
Steam Turbine	150.0	324
Total Gross Power	311.0	324
In-house Power Consumption	17.6	24
Net Power Production	293.4	300
Plant Efficiency, HHV Input/Output %	47.4	34.6%
Heat Rate (HHV) Btu/kWh	7,195	9,870
Environmental Emissions (lb/MWh)		
SO_2	0.42	2.95
NO_x	0.43	1.18
CO_2	1,260	2,040
Particulate	0.02	1.00
Solid Wastes	130	261
Liquid Effluents (Boiler blowdown)	25	60

300 MW PC Steam Power Plant

A proprietary software program and data were used by Bechtel to conceptually define the performance of a 300 MW PC plant. Table 3 summarizes the generation and emission performance. As a brief process description, superheated steam for the turbine generator is supplied by a balanced draft steam generator. The steam generator is a subcritical unit with steam conditions of 2,400 psia/1,000 F with a single reheat to 1,000 F. The plant is designed for base load operations with limited cycling capability designed for PC operation, with startup on light oil. The steam generator furnace is designed to burn the coal completely and to cool the products of combustion so that the convection passes of the unit remain clean with an acceptable soot blowing arrangement. The furnace enclosure is designed with water cooled walls.

An electrostatic precipitator is used for particulate removal. A wet limestone FGD system is included, and a SO_2 removal of 90% is specified in the design. For NO_x reduction, a SCR unit is included. The design removal rate is 80% while restricting NH_3 slip to 5 ppm and SO_3 conversion to a maximum of 1% SO_2.

COMMERCIAL PLANT CAPITAL COST ESTIMATE

The HIPPS plant boundaries for design and cost estimates are consistent with the June, 1993 EPRI TAG™. The expected accuracy for the estimate is in the range of 30 % based on the level of conceptual engineering performed.

Cost Basis

The estimate of total conceptual plant investment and capital requirement is shown in Table 4. These costs reflect the plant scope noted above and the power generation process as defined by the flow sheets and other engineering documents. Costs for the commercially available power generation and balance of plant items (Account No. items 1, and 4 through 13) were estimated from informal budget quotes by potential suppliers and experience from similar power generation projects.

Items No. 2 and 3, the HITAF furnace system were estimated as a unit by comparison with commercial steam generation systems and air heat exchangers, adjusted for the unique HITAF requirements. The adjustments reflect HITAF material flow rates, temperatures, and pressures shown on the engineering documents. Because of the advanced nature of the HITAF, engineering judgments are also used to estimate the costs. Thus, the HITAF has a greater degree of cost uncertainty than other items, and has a special process contingency of 30% applied to the installed cost of the HITAF. This contingency is to account for the uncertainty of design at this stage of development. However, the HITAF is roughly 1/6 of the total direct field cost, and even by adding, for example, 50% more to the HITAF, the total direct field cost is only increased by approximately 7%. The factors used to estimate costs below the direct field level are consistent with the EPRI 1993 TAG.

Operating Cost Estimate

The O&M cost estimates (Table 5) were estimated from experience with pulverized coal and gas turbine combined cycle power plants, the literature and EPRI TAG information. Engineering judgment was used to adjust the commercial technology cost to HITAF conditions.

Economics

All costs are expressed in mid-1994 dollars, and the 1993 EPRI TAG is used to calculate a levelized revenue requirement using the current dollar methodology. The EPRI coal closest to the Illinois No. 6 fuel used in the process design is the Perry County, Illinois Basin coal delivered to the East Central U.S. The 1990 cost is given as \$1.44, and the year 2000 forecast delivered price is \$1.36 per million Btu. The price of coal in 1994 dollars is taken as \$1.40 per million Btu. The TAG has similar fuel forecasts for natural gas. The price of \$2.84 per million Btu (in 1994 dollars) delivered to the North Central U.S. on a long term contract is estimated from the TAG. Table 6 shows the results from calculating the levelized revenue requirements for a 10 year investment time frame. The HIPPS results are compared to a 300 MW PC power plant (Exhibit 2 in the TAG - a 300 MW subcritical PC plant with flue gas desulfurization). The costs for that plant were taken from the 1993 EPRI TAG. Escalation of 7% was added to bring the TAG costs to a mid-1994 level, and a selective catalytic reactor was added for NOx control and consistency with the DoE emissions requirements. While the HIPPS plant is nominally 310 MW, the actual net production, some 293 MWe are, of course, used to compute economics. The HIPPS revenue requirement is some 5% less than the conventional PC plant..

Table 4 - Capital Requirement for 300 MW HIPPS

Account No	Code of Account Item	Cost in $ Millions
	Power Generation	
1	Solid Feeding and Removal	7.8
2	Steam Generation Island/HITAF	47.7
3	High Temperature Heat Exchangers	with HITAF
4	High Temperature Piping and Ducting	8.0
5	Process Systems/Duct Burners	0.6
6	Gas Turbine - Generator	42.9
7	Steam Turbine and Boiler Feedwater	30.1
8	Emission Control Systems	36.9
9	Blank	
	Balance of Plant	
10	Solid Material Handling	11.6
11	Water Systems	7.8
12	Support Systems	58.3
13	Civil Structures	53.8
	Total Direct Field Cost	305.5
	Indirect Costs (12%)	36.7
	Process Facilities Capital Cost	342.2
	General Facilities and engineering Fee (10%)	34.2
	Subtotal	**376.4**
	Project Contingency (15%)	56.5
	Process Contingency (30% of HITAF)	13.5
	Total Plant Cost	446.4
	Total Cash Expended (Mixed Year Dollars)	428.8
	AFDC	48
	Total Plant Investment	476.8
	Owner Costs (Prepaid Royalties;Startup;Inventory; Land)	21
	Total Capital Requirement	**497.8**

Table 5 - 300 MW HITAF Operating and Maintenance Cost Estimate

FIXED COST $/KW-YR	39.0
INCREMENTAL COST MILLS/KWH	
• Variable Costs	1.4
• Consumables	1.4
• Byproducts	0.0
TOTAL INCREMENTAL COST	2.8

COMPARISON OF RESULTS

The following major results result from the comparison of the 300 MW HIPPS commercial plant and a PC plant.

- HIPPS 37% more efficient than the PC plant.

Table 6 - Current Dollar Summary of Levelized Revenue Requirements
300 MW HIPPS and Subcritical PC Power Plants
(Mills per kWh —10 Year Investment Time Frame)

	HIPPS 300 MW	PC PLANT 300 MW	HIPPS/PC AS %
Investment and Expenses	64.4	69.7	
Fuel Costs	18.2	16.9	
Total Levelized Revenue Requirement	82.6	86.6	95%

Note: The O&M cost estimates for the HIPPS and PC plants are 2.0 and 4.0 mills/kWh respectively.

- The PC emissions (NSPS) are nearly four times greater than HIPPS. While the PC plant emissions could be reduced further by specifying higher removal rates, the capital and operating costs would rise.

- The levelized cost of the HIPPS is some 5 % less than the PC plant; even with the HIPPS contingencies 55% higher. This is a major penalty for the advanced technology, especially as the costs following the addition of contingency are percentage factors that compound the larger HITAF contingency.

- The estimated fixed and incremental O&M costs are less for the HIPPS than the PC plant.

CONCLUSIONS

While the HIPPS plant is marginally less expensive than a PC plant, the advantage would still go to the HIPPS by virtue of the lower emissions and higher efficiency. Also, and most importantly, the preliminary HIPPS commercial design is the first step in a process to develop a HITAF system operating entirely (or nearly so) on coal. The HIPPS and PC comparison clearly shows the HIPPS to be more efficient, environmentally cleaner, and economically more attractive than current PC units.

ACKNOWLEDGMENT

The effort described here is supported by DoE Pittsburgh Energy Technology Center and Contract DE-AC22-92PC91155. In addition to the other Combustion 2000 Team members identified, the authors wish to recognize the contributions of Messrs. W. A. Blecher and S.J. Lehman in providing computer modeling and technical review.

ACHIEVING COMPLIANCE WITH
ADVANCED COAL-FIRED LOW-EMISSION BOILER SYSTEMS

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ABSTRACT

The paper reports on ABB's work in Phase I of the U.S. Department of Energy (DOE) project "Engineering Development of Advanced Coal-Fired Low-Emissions Boiler Systems" (LEBS) which is part of the DOE's Combustion 2000 Program. Work planned for future Phases is also described.

The overall objective of the LEBS Project is to dramatically improve environmental performance of future coal-fired power plants without adversely impacting efficiency or the cost of electricity. Near-term technologies, i.e., advanced technologies that are partially developed, will be used to reduce NO_X and SO_2 emissions to one-third current NSPS limits and particulates to one-half current NSPS limits. Air toxics must be in compliance. Waste must be reduced and made more disposable.

The work in Phase I included concept development and evaluation of several subsystems for controlling the emission of SO_2, NO_X, and particulates. Candidate technologies were then evaluated in various combinations as part of complete advanced supercritical power generation systems. One system was selected for preliminary design of a commercial generating plant.

INTRODUCTION

Combustion Engineering, Inc. (ABB CE) is one of three contractors for the US Department of Energy (DOE) Project titled "Engineering Development of Advanced Coal-Fired Low-Emission Boiler Systems". The overall objective of the Project is the expedited commercialization of advanced coal-fired low-emission boiler systems. The specified primary objectives are emissions of NO_X and SO_2 less than one-third New Source Performance Standards (NSPS), particulates less than one-half NSPS, and air toxics in compliance. Secondary objectives are improved ash disposability, reduced waste generation, and increased generating efficiency. All primary, and all or some secondary objectives must be met without increasing the cost of electricity from a current

"NSPS plant". Because emission requirements vary from site to site, ABB elected to have two NO_X targets (the contract target and approximately one quarter of that) and to favor SO_2 and particulate control technologies which can be designed for emission levels lower than the contract targets with minimum impact on costs. The final Project deliverables are a design data base that will allow future coal-fired power plants to meet the stated objectives and a preliminary design of a commercial generation unit (CGU).

In addition to the DOE - Pittsburgh Energy Technology Center, the project is being managed by ABB Power Plant Laboratories Division of Combustion Engineering, Inc. (ABB CE) as the contractor and the work is being accomplished and/or guided by this contractor and the following team members:

- ABB CE (NO_X, efficiency, waste disposability, cost of electricity)
- ABB Environmental Systems (SO_2, particulates, waste reduction)
- Raytheon Engineers and Constructors (Plant-wide evaluations, cost of electricity, A/E)
- Technical Consultants and Industry Advisors

SELECTION OF TECHNOLOGIES

The first major technical effort was concept development. The process began with the selection of coals and identification of candidate technologies through literature search and in-house sources. The team selected those candidates judged to have the potential of meeting the Project's primary objectives and one or more of the secondary objectives and to become commercially feasible within the Project's timeframe. Commercial feasibility in this sense encompasses not only technical and economic feasibility but also acceptance by the utility industry.

Near-term technologies were screened to identify those that best fit the criteria described above. From the remaining candidates, rough economic comparisons were made to compare system installation costs and annualized operation and maintenance costs. Five SO_2/particulate and five NO_X candidates were selected and subjected to comprehensive technical assessment and systems analysis and a "short list" was developed. Finally, six combinations of the short-listed subsystems were evaluated for technical, economic, and commercial feasibility as integral parts of an advanced supercritical power generation system.

Three test coals were selected and one was identified as the design coal to serve as the baseline test coal. The remaining test coals will be used in R&D and testing to determine process sensitivities to variations in coal characteristics. The test coals were selected so that the results of the engineering development work will be broadly representative of large classes of US coals whose current production is extensive, with significant remaining minable and uncommitted reserves. The sulfur content of the design coal, on entering the boiler, is at least three pounds of sulfur per million Btu. Illinois No. 6, Pittsburgh No. 8, and Upper Freeport were selected to be the project test coals and Illinois No. 6 will be the design coal.

The next step in the process was to select the plant steam cycle. A comprehensive evaluation was carried out which compared a conventional subcritical cycle with throttle conditions of 2400 psig/1000°/1000° to an advanced supercritical cycle with throttle conditions of 4500 psig/1100°/1100/1100°. The supercritical cycle was selected because of its higher plant efficiency. This not only provides performance improvements but it will reduce the amount of pollutant produced per kilowatt of electricity generated, regardless of the emissions technologies involved. The steam generator and turbine generator performance and cost values support this selection.

TECHNOLOGIES

Using the methodology described above, the team developed the following short list of subsystem technologies:

NO$_x$ Control:

*1. Advanced Tangential Firing.
*2. Coal Reburn.
*3. High-temperature SNCR
*4. Catalytic Filter (fabric or ceramic).
5. SCR for cost comparison.

* Selected for systems analysis.

SO$_2$/Particulate Control:

*6. Advanced Wet Limestone Scrubber with EP or FF.
*7. Thioclear Scrubber with EP or FF.
*8. SNO$_x$^{TM} Hot Process.
9. Catalytic Baghouse with WSA Tower.
10. Catalytic Baghouse with Wet Scrubber.

These subsystem technologies were evaluated as integral parts of these six advanced supercritical power generation systems.

1. Advanced Burners, Advanced Wet Limestone FGD, Advanced Electrostatic Precipitator.
2. Advanced Burners with Coal Reburn, Advanced Wet Limestone FGD, Advanced Electrostatic Precipitator.
3. Advanced Burners with High-Temperature SNCR, Advanced Wet Limestone FGD, Advanced Electrostatic Precipitator.
4. Advanced Burners, Thioclear FGD, Advanced Electrostatic Precipitator.
5. Advanced Burners, SNO$_x$^{TM} Process.
6. Advanced Burners, SNO$_x$^{TM} Hot Process.

Contractors were required to select one system as the basis for the commercial generating unit (CGU) design. Since all the primary and secondary objectives can be met by each of the six systems, the main evaluation criteria came down to the potential for commercial success and the cost of electricity. Based on evaluation of these concepts, the Team selected the SNO$_x$^{TM} Hot Process (System 6). This system is described schematically in Figure 1.

However, for the following reasons none of the short listed subsystem technologies were abandoned: (1) They may be more suitable for a particular project or customer. (2) They may become commercially available sooner. (3) They would be fallbacks if a selected technology

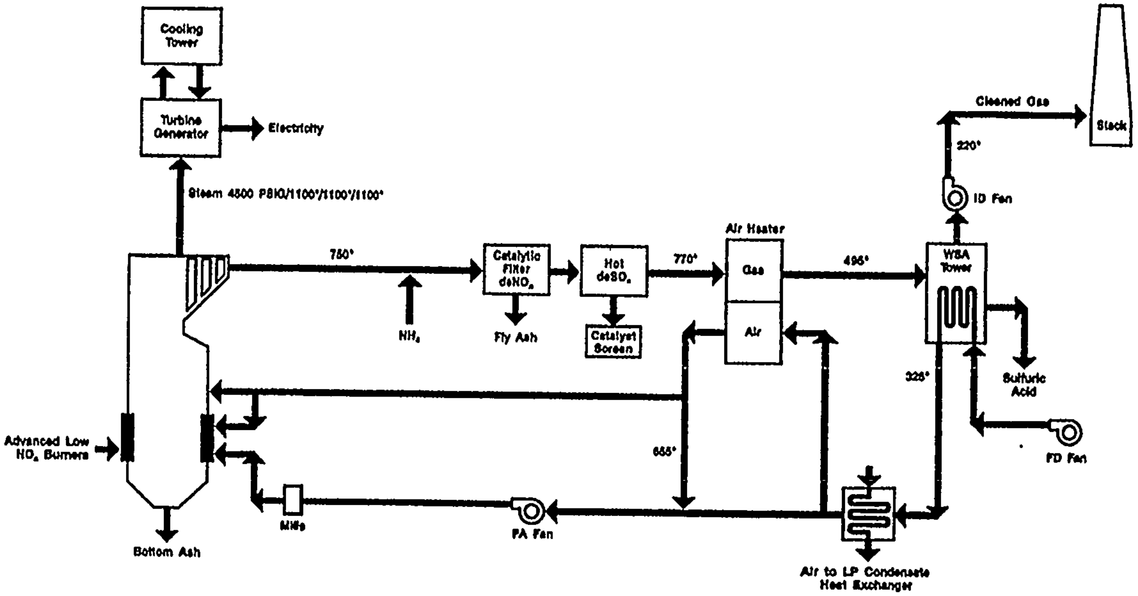


Figure 1 System Diagram

proves unsuccessful. (4) There is insufficient information to reject or select some of them at this time. (5) Their development will continue under other programs. (6) The requirement that air toxics be "in compliance" is undefinable at this time since regulations have not been established. Therefore, control technologies cannot be fixed. It is entirely possible that the final output of the Project is one system design with a menu of options for emission control subsystems.

A specific ash disposal technology was not selected. Because the choice is highly site-specific, the team only identified and assessed candidate technologies and reduced the list to three: (1) Chemical fixation with lime for stabilized landfill. (2) Production of lightweight aggregate from flyash. (3) Vitrification. The latter two are being commercialized under other projects and all three will remain as an option in future work on this Project.

Brief descriptions of each selected subsystem incorporated into the CGU design follow.

Advanced Tangential Firing. This combustion system is the basis of the low NO_X approach for the continuing evolution of tangential firing. Pilot-scale tests have demonstrated NO_X emissions below the Project's target with reasonable carbon loss. This technology will continue to be developed and advances will be applied to this Project as they evolve. The Project supports expanding the data base, component development, and system evaluations.

Catalytic Filter. The catalytic filter with SCR catalyst is an integral step in the SNO_X™ systems. (See below.) Also, post combustion NO_X reduction will facilitate an alternate, lower NO_X emission goal. The two candidate designs are the CeraMem™ catalytic ceramic filter and the University of North Dakota - Energy and Environmental Research Center catalytic fabric filter. Both are expected to be available prior to the LEBS commercial-readiness target date. The CeraMem™ filter was selected for the CGU design.

SNO_X™ Hot Process. The SNO_X™ (Sulfur and Nitrogen Oxide Reduction) technology utilizes two catalytic reactors to control NO_X and SO_2 in the flue gas stream. No sorbents are added and no sludge is formed. The process is capable of greater than 97% SO_2 removal, 80-90% NO_X removal and additional particulate removal while producing commercial grade sulfuric acid and useable heat. The process was developed in the early 1980's and has been successfully tested in Europe and Asia in pilot scale units as well as on a full scale 310 MWe coal-fired unit Denmark. In the United States, it was demonstrated under DOE's Clean Coal Program as a 35 MWe slipstream facility on a Ohio utility unit burning high-sulfur coal. Following successful demonstration, the host utility elected to continue operation on a commercial basis. The by-product sulfuric acid is sold locally.

The SNO_X™ Hot Process is an adaptation of the SNO_X Process that, by taking advantage of a high temperature (750°F) CeraMem™ catalytic filter with integral NO_X SCR catalyst, allows process simplification with reduction in capital equipment as well as an improvement in the thermal efficiency of the steam cycle. (See Figure 1.) The process is expected to achieve greater than 97% SO_2 removal and 80% NO_X removal. The CeraMem™ filter is the least "near-term" of all the technologies selected and the fallback is the high temperature fabric filter.

THE COMMERCIAL GENERATING UNIT DESIGN

The nominal 400 MWe Commercial Generating Unit (CGU) illustrated in Figure 2 is an adaptation of a conventional pulverized coal-fired steam-electric plant in which selected technologies have been introduced to achieve reduced levels of airborne emissions, increased thermal efficiency, reduced waste and improved costs. These technologies involve primarily three areas:

- An advanced low-NOx combustion system in the furnace that limits NO_x concentrations in the flue gas leaving the steam generator to 0.2 lb/MMBtu.
- The SNO_x™ Hot Process integrated with a ceramic-element gas filtration system. Here, particulates are separated, NO_x concentrations are further reduced, and SO_2 is removed from the flue gas in the form of a commercial grade sulfuric acid by-product.
- Advanced supercritical boiler and turbine with throttle conditions of 4500 psig, 1100°/1100°/1100°.

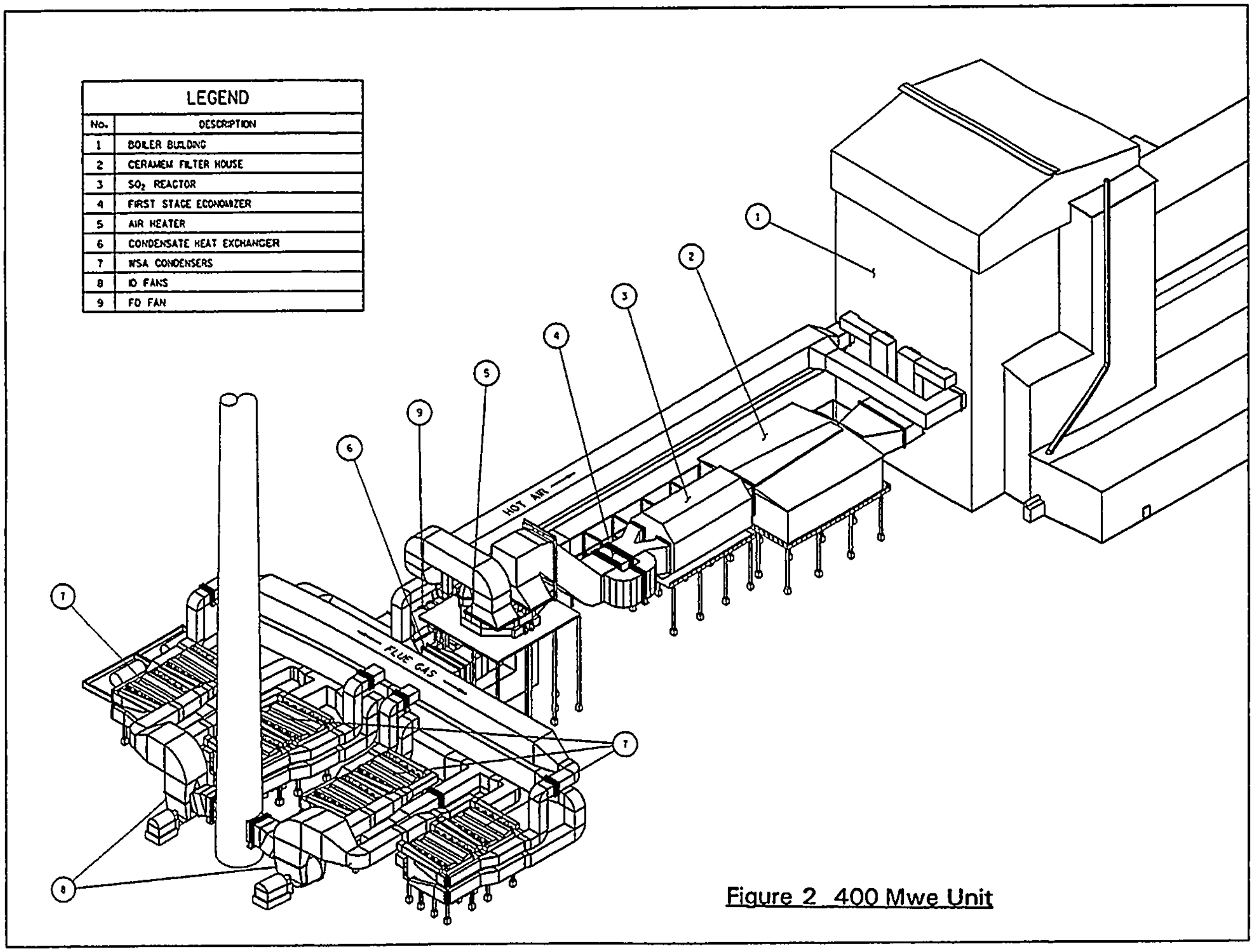
This combination of emission control processes meets or betters all of the target emission levels for the LEBS Project, while producing either benign or saleable by-products from the gas treatment. The advanced steam cycle and the SNO_x™ Hot Process enable the design to meet the efficiency objective and, indirectly, the cost of electricity objective.

The design and economic bases are summarized in Table 1. The design coal analysis is given in Table 2.

Table 2 - **Midwestern Bituminous Coal Analysis, as-received (Illinois No. 6)**

Proximate Analysis, %		Ultimate Analysis, %	
Moisture	12.0	Moisture	12.0
Volatile Matter	33.0	Carbon	57.5
Fixed Carbon	39.0	Hydrogen	3.7
Ash	16.0	Nitrogen	0.9
	100.0	Chlorine	0.1
		Sulfur	4.0
Higher Heating Value, Btu/lb	10,400	Oxygen	5.8
Grindability, Hardgrove	56.0	Ash	16.0
			100.0

The CGU design includes all of the structures, equipment and material for a complete power plant. This includes the steam generation, electricity generation, and pollution control systems. It also includes fuel storage and handling systems, cooling water and service water systems, and the switchyard.



No.	DESCRIPTION
1	BOILER BUILDING
2	CERAMEM FILTER HOUSE
3	SO_2 REACTOR
4	FIRST STAGE ECONOMIZER
5	AIR HEATER
6	CONDENSATE HEAT EXCHANGER
7	WSA CONDENSERS
8	ID FANS
9	FD FAN

Figure 2 400 Mwe Unit

Table 1 - CGU Design Bases

GENERAL

Site	Kenosha, WI
Number of Units	One
Net Plant Heat Rate (Btu/kWh)	8305 (2.5" Hg)
Net Plant Efficiency (%)	41.1
Net Power Output at Rated Load (MWe)	408.1 at Step-Up Transformer
Fuel	Illinois No. 6

CIVIL/STRUCTURAL

Boiler and Turbine Building	Enclosed
Foundations	Spread Footings

MECHANICAL

Forced Draft Fans	One, motor driven
Induced Draft	Two, motor driven
Primary Air	One, motor driven
Coal Delivery	100 Car Unit Train at 5 hour Nominal Turnaround
Coal Storage	90 Days at Rated Load, 8 hours in Silos
Coal Handling System	Track Hopper, Lowering Well, Crusher, Boiler House Transfer Tower, Trippers
Pulverizers (Total/Spares)	5/1
Stack	500 feet
Waste Disposal	Ash is Trucked Off-Site
Turbine-Generator	
a. Configuration	Tandem-Compound 2-Flow
b. Speed (RPM)	3600
c. Last Stage Blade Length (in)	33.5
Gross Turbine-Generator Output at 2.5 in-HgA (MWe)	428
Condensers	
a. Type	Single Pressure
b. Shell/Divisions per Shell	2/1
c. Arrangement	Longitudinal
d. Number of Passes	Two
e. Pressure (in-HgA)	2.5
Cooling Tower	
a. Type	Natural Draft Wet Evaporative
b. Number/Total Flow-Normal (gpm)	1/171,200

Cooling Tower Conditions (Design)	
a. Approach (°F)	16
b. Range (°F)	20
c. Wet Bulb (°F)	67
Feedwater Pumps	
a. High Pressure (number/Drive)	1/Main Turbine Shaft
b. Other (Number/Service/Drive)	1/Booster/Motor 1/Start-Up Motor
Feedwater Heaters	
a. Open Stages (Number)	One
b. High Pressure Closed Stages (Number/Number Trains)	2/1
c. Intermed. Pressure Closed Stages (Number/Number Trains)	2/1
d. Low Pressure Closed Stages (Number/Number Trains)	4/1
e. Air to Condensate Heat Exchanger	One

ELECTRICAL

Connection to Off-site Power (No./kV)	2/230
Generator	
a. Rating (MVA)	500
b. Voltage (kV)	24
c. Power Factor	0.9
d. H_2 Pressure (psig)	45
Generator Disconnect	Bolted Disconnect Links
Auxiliary Power System	
a. Medium Voltage System A (kV)	6.9
b. Low Voltage System (V)	480
c. Direct Current Systems (V)	250/125
Station Service Transformers Nameplate Rating (MVA)	20/24
Unit Auxiliary Station Transformers Nameplate Rating (MVA)	20/24
Natural Gas Engine-Generator Unit	
a. Type/Number	High Speed/1
b. Voltage (V)	480
c. Rating (kW/PF)	600/0.8
Control Room Wiring	Multiplexed to Logic Cabinets in Control Room
Multiplexing of Cables	Distributed Microprocessors, metal Conductor and Fiber Optics Data Highways
Instrumentation	Independent Sensors for Computer Input

Expected emissions performance is listed in Table 3 and shown graphically in Figure 3.

Table 3 - **Emissions Reduction Performance**

	NSPS	LEBS Target	CGU	Reduction
SO2, lb/mm Btu*	0.6	0.2	0.12	80%
NOX, lb/mm Btu	0.6	0.2	0.05	91.67%
Particulate, lb/mm Btu*	0.03	0.015	0.005	0.8333

* 3 lb S and 15.4 Lb ash per million Btu in the coal.

Most inorganic air toxics should be removed at near 99% with the particulate removal equipment. Volatile organic emissions, CO and ammonia slip will be oxidized in the SO_2 oxidizer.

The CGU will produce significantly less waste than the NSPS plant. Part of this is due to the lower amount of ash produced per kWh because of the higher efficiency boiler and steam cycle and the SNO_X^{TM} Hot Process. The major portion of this reduction results from the production of sulfuric acid as a commercially saleable by-product rather than the sludge normally generated by an FGD system. A summary of waste generation and a comparison to the NSPS plant is listed in Table 4 and shown graphically in Figure 4.

Table 4 - **Waste Reduction Performance**

	NSPS	CGU	Reduction
Ash, Lb/kWh	0.148	0.128	13.5%
lb/mm Btu	15.4	15.4	0.0%
FGD Waste, Lb/kWh	0.204	0.00	100%
lb/mm Btu	21.2	0.00	100%
Total Waste, Lb/kWh	0.352	0.128	63.6%
lb/mm Btu	36.6	15.4	57.9%

The plant uses a 4500 psig, 1100°F supercritical thermodynamic cycle with two reheat streams at 1100°F each. The gross output of the generator is 428 MWe. The net plant output is 408 MWe. Net plant heat rate is 8,305 Btu/kWh for a net efficiency of 41.1% based on fuel HHV. (Note: a detailed breakdown of energy losses should be reviewed when comparing efficiencies of different systems.) There is the potential to increase net plant efficiency beyond 41.1% and this will be pursued in ongoing work.

Figure 3 **Emissions Performance Comparison**

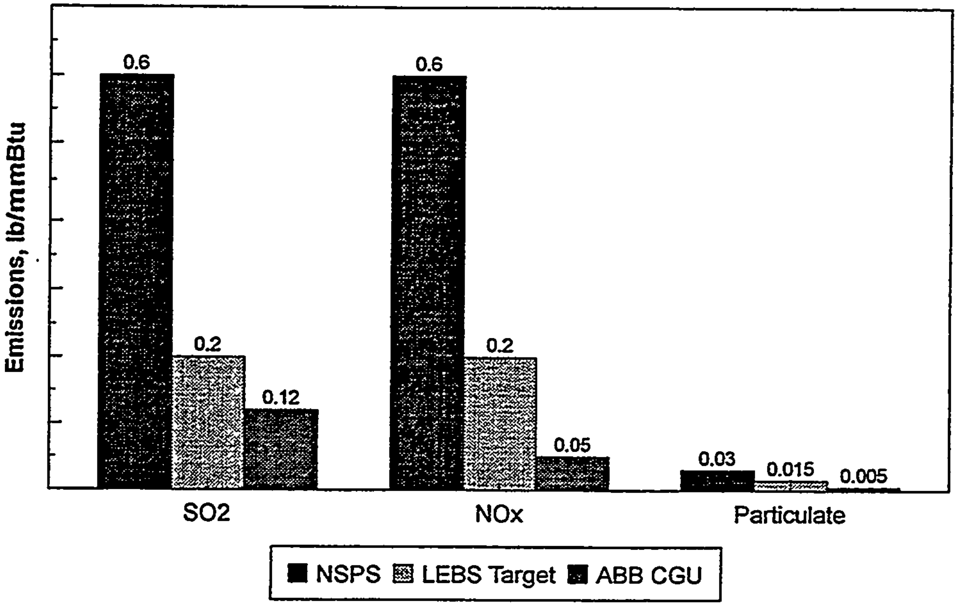
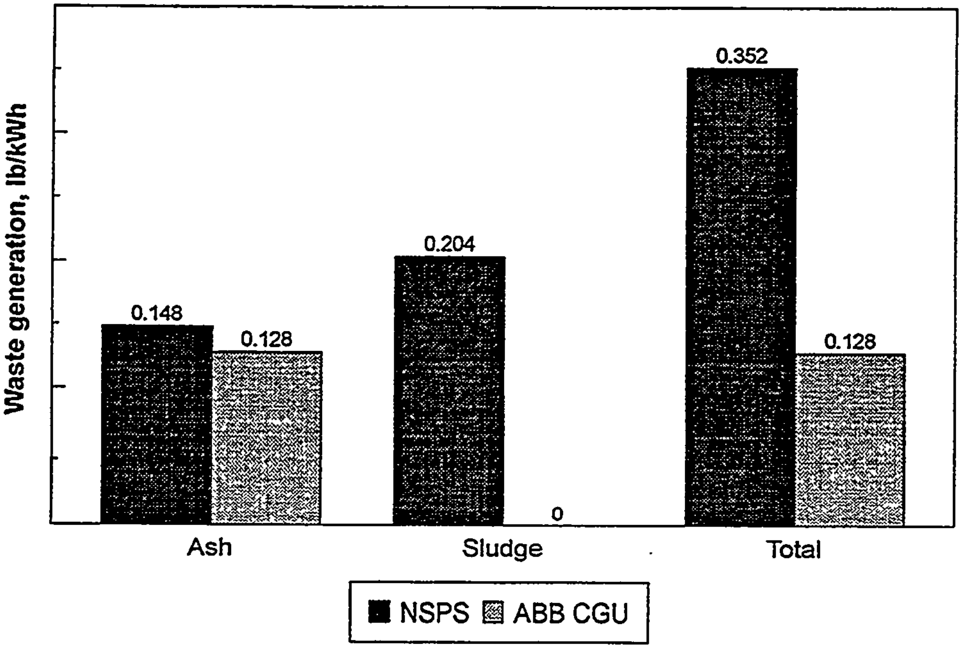


Figure 4 **Waste Generation Comparison**



The CGU has a total plant cost that is less than the cost of a current NSPS plant. The total capital requirements estimate includes the time related portions of the project estimate such as allowance for funds used during construction. The improvements come from the adoption of an aggressive commercialization plan that utilizes the concept of a "Consortium" formed to produce a number of these units on a replicated and modularized basis. These factors, coupled with ABB's commitment to a significantly reduced "cycle time" for the boiler and other key equipment results in significantly reduced schedule from award to start-up. This aggressive construction schedule improves the time-related costs.

The CGU will satisfy the objective of having a cost of electricity equal to or less than that for the NSPS plant. The calculated cost of electricity is reduced by the by-product credit received from the sale of the sulfuric acid (using a figure confirmed by an outside market study) and by an aggressive but achievable capacity factor. An independent reliability, availability and maintainability analysis was completed for the CGU. The study was based on performance data obtained from the NERC data base and utilized the industry accepted "Delphi" process to adapt the data for the CGU. Since DOE required that the cost of electricity analysis be completed assuming baseloaded operation, the equivalent availability factor estimated in this study was used for the CGU capacity factor. One reason enhanced reliability and equivalent availability are achieved is that the SNO_X™ Hot Process is "passive", i.e., it has far less mechanical equipment than is typically found in flue gas desulfurization processes that utilize lime or limestone. The simpler process, absence of mechanical equipment, and the passive character of the process results in higher reliability and availability.

The design also incorporates advanced diagnostics concepts which provide early warning of impending failures in the plant equipment. This advanced knowledge has several benefits that result in improved reliability and availability. Advanced diagnostics should enable maintenance outages to be both more effective by providing maintenance information in areas which might not be readily amenable to inspection, and shorter because preparation should be better due to a reduced number of "surprise" repairs.

Finally, the CGU was designed for access and ease of maintenance because it was designed for access and ease of construction. The plant is laid out with the "ranch" concept. This means that the stacking of equipment is minimized. Rather it is spread out in the horizontal plane. In addition, the plant design incorporates a "backbone" utility rack for piping, cable, conduit and electrical wiring. The ground level portion of this rack is used as a maintenance access corridor that runs throughout the plant. Also, organizing piping and conduit on overhead racks provides more ground level access to equipment for maintenance. Incorporating these features in the conceptual design of the plant, and coupling them with the implementation of a "design for maintainability" approach during the detailed design stage, will result in a plant with superior availability and higher capacity factor.

Some of the features of the CGU, which indirectly should make it attractive to utilities, are as follows:

- The design incorporates a supercritical cycle but a subcritical version can be provided also.

- The CGU eliminates the FGD waste disposal problem, including future licensing, because it produces commercial grade sulfuric acid which is a marketable commodity in most parts of the country. The heat rate and efficiency are improved because the design recovers the heat generated from condensation of the sulfuric acid to heat the combustion air. In addition, the SNO_x™ Hot Process has, except for small acid pumps, no mechanically driven equipment. This results in a savings in auxiliary power consumed inside the plant and reduced O&M costs.
- The passive nature of the SNO_x™ Hot Process, coupled with the incorporation of advanced diagnostics systems and an aggressive maintenance program, will enable the CGU to have high availability and increase capacity factor in baseload service. This is particularly attractive to utilities under pressure from their PUC to increase availability and lower reserve margins.
- The use of the SNO_x™ Hot Process to make sulfuric acid eliminates future licensing risk to the utility about the regulatory requirements surrounding the disposal of FGD sludge.
- The base model for the CGU is designed for baseload operation with limited cycling capability but cycling capability can be added.
- The plant is significantly more cost effective than the NSPS plant, due to in part to replication and modularization, an aggressive construction schedule, and the "ranch" and "backbone" concepts in the arrangement.

All of the foregoing features are responsive to the technical, regulatory and economic needs of the power generation industry. The superior performance of the ABB CGU coupled with efforts to minimize investor risk should make it extremely attractive to the utilities and IPPs. The near term character of the LEBS technologies chosen, coupled with the attractive performance and cost features of the ABB CGU, support a confidence in the Project Team that the proposed CGU design will be acceptable and marketable.

DEVELOPMENT OF PRESSURIZED COAL PARTIAL COMBUSTOR

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INTRODUCTION

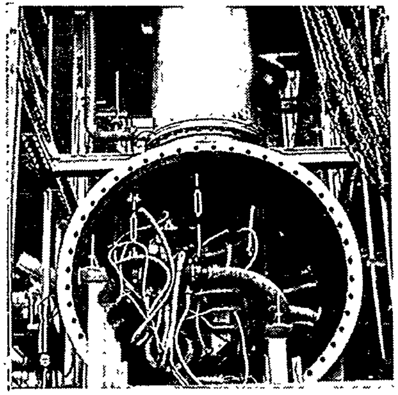

Fig.1 7t/d coal bench scale PCPC

Coal reserves are abundant and distributed worldwide. Since its supply is stable, coal is considered one of the important fuels for power generation in Japan. As environmental problems such as global-warming and acid rain are emerging recently, reduction of CO_2 emissions is becoming increasingly important along with efforts to reduce sulfer oxide and nitrogen oxide emissions. Under these circumstances, the need for development of highly efficient , coal-fueled power generation systems has been increasing.

The integrated gasification combined cycle (IGCC), an environment-friendly power generation system of high thermal efficiency, is being developed via various approaches around the world. The oxygen-blown entrained flow gasification process is a relatively simple method of producing medium calorie coal gas suitable for application to gas turbines. Various systems for this process have been developed to a demonstration level in Europe and America(1),(2).

The air-blown entrained flow gasification process does not require the power needed to produce oxygen, therefore it is more efficient than the oxygen-blown process. Japan has actively been developing the air-blown process(3),(4). However, Taking stable molten slag discharge into consideration, coal must be supplied at two stages to raise the combustor temperature in ash molten part. Only two reports have been presented regarding two-stage coal supply. One is the report on an experiment with the Hycol gasifier, in which air feed ratio is varied, with coal feed fixed(5). The other is a report on a simulation study with various gasifier coal feed ratios, conducted at Central Research

Institute of Electric Power Industry (6). It seems that the appropriate feed ratio has not yet been established.

Meanwhile, R & D on a Coal partial Combustor (CPC) under atmospheric pressure has been conducted in a project sponsored by the Ministry of International Trade and Industry (MITI) (7)-(9). This project has been conducted between Center for Coal Utilization, Japan and Kawasaki Heavy Industries Ltd since 1984. Since 1994, Chubu Electric Power Company Incorporated and Electric Power Development Co., Ltd have been cooperating.

Through this activity, a unique furnace construction has been established, and these influences of stoichiometric air ratio, of oxygen enrichment, of char recycling and of coal types on performance have been clarified. The purpose of the present study is to apply this developed CPC techniques to a Pressurized CPC (PCPC), thereby improving the IGCC technology.

For the present study, we conducted systematic experiments on the air-blown process with a two stage dry feed system, using a 7 t /d-coal **bench scale PCPC test facility** (**Fig. 1**), operated at the pressure of 0.4 MPa, and clarified the influence of coal feed ratio on coal gasification performance.

This report describes the above-mentioned bench scale test procedures and results, and also some informations about a plan of a 25 t/d-coal pilot test system .

BASIC CONCEPT AND STRUCTURE OF CPC

The basic concept of CPC is a two stage dry feed entrained flow system: Coal and air are supplied at high speed tangentially into the cylindrical combustor to cause partial combustion (gasification) in a highly reducing atmosphere under high temperature and high heat load, while most of the coal ash is melted, separated and removed. The resultant product gas is used effectively as fuel gas.

Fig. 2 shows the conceptual structure of the CPC, which is composed mainly of the **precombustor**, the **CPC**, and the **slag-free duct**. The precombustor has a pulverized coal burner at the top , in which coal is supplied downward together with air, so that the coal is partially burned in a fuel-rich atmosphere. Air is also fed from the side of the precombustor, to increase partial combustion gas temperature. High temperature partial combustion gas is then blown tangentially into the CPC installed under the precombustor.

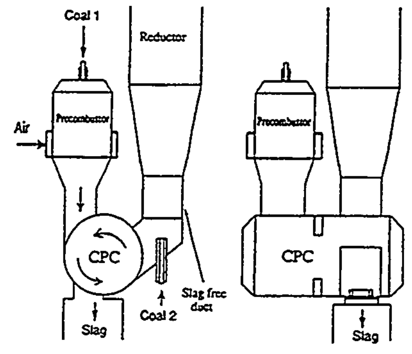

Fig.2 Conceptual Structure of CPC

Most of the coal ash is captured on the CPC wall due to centrifugal force, and discharged in the form of granulated slag through the slag hole.

The water-cooled self-coating wall protects the CPC from high-temperature gas. The slag-free duct connected to the CPC outlet is made of water-cooled flat panel. This duct damps the swirling motion of the product gas discharged through the CPC outlet, preventing captured molten slag from flowing downstream. Pulverized coal is supplied in the middle of the slag free duct, to lower product gas temperature quickly and raise the heating value of the product gas.

TEST FACILITY AND PROCEDURE

Test Facility

Fig. 3 shows the schematic flow diagram of the bench scale PCPC test facility, which comprises compressor, pulverized coal storage, pulverized coal feeder (lock hopper type), pressure vessel containing CPC and reductor, product gas cooler, char recycle feeder, ceramic filter, and operation control system. Various draft equipment, product gas incinerator, cooling equipment, exhaust gas treatment facility etc. are also installed as auxiliary.

The test system provides a coal processing capacity of 7 t/d, design pressure of 1.1 MPa, and normal operating pressure of 0.4 MPa. The normal operating pressure of the test facilities was set much lower than the estimated actual operating pressure (2 to 3 MPa) for the following reason. Under the restriction of the product gas incinerator capacity (7 t/d), if the operating pressure were set as high as that of the demonstration plant, this CPC size would have to be extremely small. To obtain molten slag discharge performance data as well, it was considered reasonable to conduct the test with a gasifier of the adequate size, under medium operating pressure.

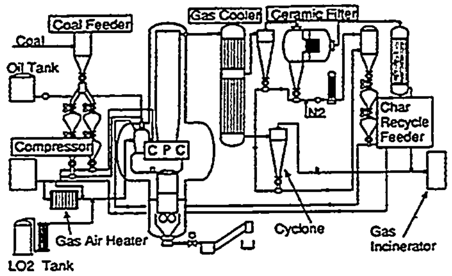

Fig.3 Schematic Flow Diagram of 7t/d coal bench scale PCPC

The ceramic filter to collect fly char, and the char recycle feeder, are located downstream of the test facility. To obtain basic data via the simplest method, these equipments were not used in this report.

Operation Procedure

Compressed air was used as gasifying agent. Air of ambient temmperature was used to transport pulverized coal. The other air was preheated to about 600 K. To conduct the test with systematically varied coal feed ratio, air flow rate to the CPC precombustor (**A1**) and to the CPC outlet (**A2**) were fixed at 810 kg/h and 230 kg/h, respectively. Air-transported pulverized coal was fed at constant volume through the dry lock hopper system and supplied at two stages to CPC precombustor (**coal1**) and CPC outlet(**coal 2**).

The system was operated by the following procedure:

The entire system was purged at the beginning. The LPG pilot burner was then ignited, and pressure was raised to the specified value (normally 0.4 MPa). The preheating burner with kerosene was then ignited under excess air condition. Upon completion of preheating, kerosene flow was increased, and the system was purged under a stoichiometric air ratio of about 1 (0.5% or lower O_2 concentration in product gas). After purging oxygen, kerosene flow was further increased to enter the gasification mode.

In the gasification mode, the downstream gas incinerator was checked for ignition and combustion of product gas, and the operating stoichiometric ratio was lowered to increase the heating value of the product gas. At this stage, coal supply was begun, and the conditions were set to start test operation.

At the end of the test, the stoichiometric air ratio was increased gradually to minimize carbon monoxide generation. To prevent combustion of deposited char in the system, the entire system was purged with nitrogen gas simultaneously with fuel feed stopping. Finally, the system was cooled and stopped while filled with nitrogen gas.

Measurement Method

Combustion air feed was measured via an orifice, and coal feed was measured with an impact flowmeter and a load cell in combination. To evaluate coal gasification performance, the composition of the product gas sampled at the cyclone outlet, was analyzed by gas-chromatography. Components for the analysis were CO, CO_2, H_2, O_2, N_2, and hydrocarbons of CH_4, C_2H_6, and C_3H_8. The result was cross-checked with a continuous analyzer (CO/H_2/calorimeter) to confirm correctness.

Properties of Test Coal

Taiheiyo coal, which is popular in Japan and easy to gasify, was used for the test. Since this coal ash melts at relatively low temperature at 1,590 K, melting point adjusting agent was not added. Element analysis revealed that the coal was composed of C, H, N, S, O and ash in the ratio of 63.9%, 5.6%, 1.0%, 0.2%, 14.9% and 14.4%, respectively.

TEST RESULTS

Test Conditions of Coal Feed Ratio

Figs. 4 and 5 show the test conditions of coal feed ratio for gasification characteristics with various feeds of coal1 and coal2 under the conditions shown in .

In Fig. 4, the vertical axis represents the total coal feed (coal1 + coal2), the horizontal axis the ratio (R2) of coal2 feed to total feed. For any point on the graph, it is possible to identify the coal1 and coal2 coal feeds. Total coal feed was set in the range between 200 and 350 kg/h, with coal2 feed varied from 0 to 200 kg/h at increments of 50 kg/h. **The solid lines plot a coal feed in which coal2 ratio is varied with total feed fixed, the broken lines plot a coal feed in which coal2 feed is varied with coal1 feed fixed** (this rule applies to all other charts appearing herein).

As shown in Fig. 5, total stoichiometric air ratio (λ_t) in the entire gasifier (CPC + reductor) was varied from 0.35 to 0.6. The value of λ_t can be calculated by Equation (1), and the value of stoichiometric air ratio for CPC (λ_1) by Equation (2).

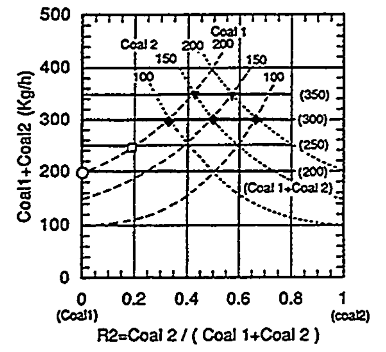

Fig.4 Total coal feed as a function of coal2 feed ratio R2

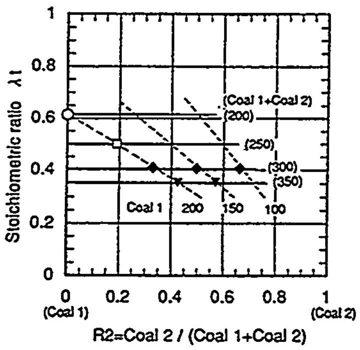

Fig.5 Stoichiometric ratio λ_t vs R2

$$\lambda_t = (A1 + A2) / (A_0 \times (Coal1 + Coal2)) \quad \text{——— (1)}$$

$$\lambda_1 = A1 / (A_0 \times (Coal1) \quad \text{——— (2)}$$

in which A_0 is the theoretical air requirement of Taiheiyo coal (=8.65kg/kg-coal).

Properties of Product Gas

It is possible to determine the important gasification performance factors of the gasifier by analyzing the composition of the product gas at the gasifier outlet ⁽⁷⁾. The heating value of the product gas is the basic indicator of gasification performance. For the present study, the heating value was obtained by gas chromatography.

The heating value of the product gas shows two characteristic behaviors, as shown in **Fig. 6**. First, it rises with monotonous increase in coal2 feed (broken lines). This implies that monotonous coal2 feed contributes to a increase in gas heating value. Second, the heating value rises with lower coal2 feed ratio(solid lines). In other words, the heating value increases with higher the coal1 feed ratio. With the coal1 feed ratio exceeding a certain level, CPC temperature drops, possibly hampering stable slag discharge. However, it is desirable to operate the system with higher coal1 feed ratio, so long as stable slag discharge is secured.

Fig. 7 shows the behavior of CO in the product gas. Carbon monoxide, the primary component of combustible gas, shows a behavior similar to that of the heating value. As coal2 feed ratio is increased (solid lines), CO concentration in the product gas decreases substantially due to decreased carbon conversion efficiency (Fig. 10) and increased concentrations of hydrocarbon (Fig. 8) and CO_2 (Fig. 9). CO concentration has maximal value as coal2 feed is varied with coal1 feed fixed (broken lines). H_2 concentration shows behavior similar to that of CO concentration. However, drop in H_2 concentration is smaller with higher coal2 feed ratio, because hydrogen in coal can be gasified easier than carbon .

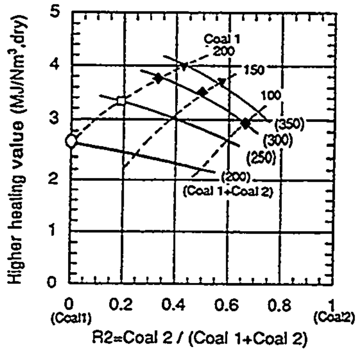

FIg.6 Higher heating value

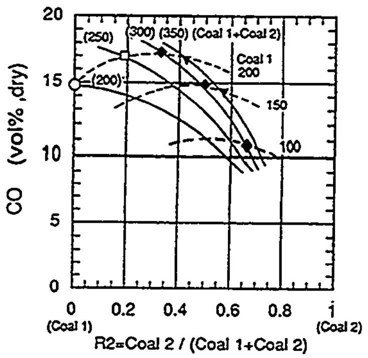

Fig.7 CO Concentration

Fig. 8 shows the hydrocarbon concentration behavior. The product gas contains CH_4 as primary hydrocarbon, and may also contain C_2H_6 and C_3H_8. The hydrocarbon concentration is therefore C_1-equivalent concentration (CH_4^*). This concentration increases sharply with higher coal2 feed, indicating that hydrocarbon is generated from coal2, not from coal1.

As shown in Fig. 9, CO_2 concentration increases with higher coal2 feed ratio (solid lines). In coal gasification, it is desirable that CO increase and CO_2 decrease. To meet this requirement, operation with lower coal2 feed ratio is preferable. Under the conditions set for our test, however, CO_2 concentration was higher than 8%, even with maximum coal.

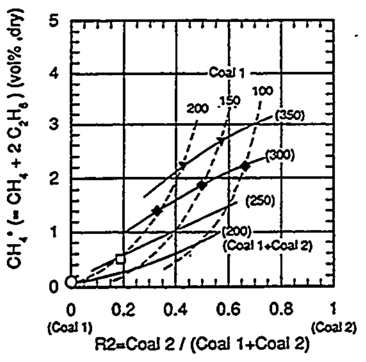

Fig.8 Hydrocarbon Concentration

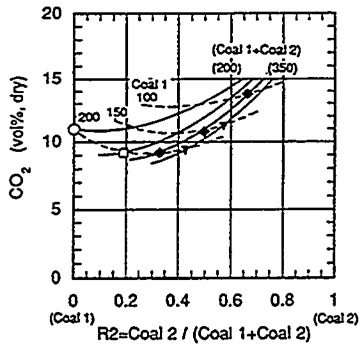

Fig.9 CO2 Concentration

Carbon Conversion Efficiency

Fig. 10 shows the carbon conversion efficiency (gasification efficiency) data. As

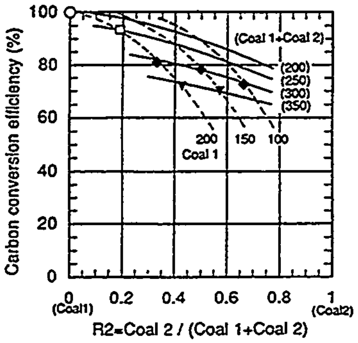

Fig.10 Carbon Conversion Efficiency

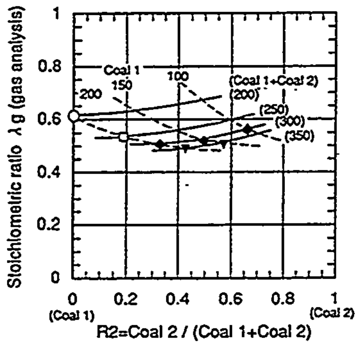

Fig.11 Stoichiometric ratio λ_g

shown, carbon conversion efficiency η_{C1} decreased with higher coal2 feed ratio, indicating that operation with higher coal1 feed ratio is desirable. **Fig. 11** shows the stoichiometric air ratio for gasification reaction (λ_g). λ_g can be calculated by Equation (3) from the composition of the product gas and element analysis values of coal (7).

$$\lambda_g = 1 - 0.5((CO + H_2 + CH_4 \times 4 + C_2H_6 \times 7 + C_3H_8 \times 10) / ((CO + CO_2 + CH_4 + C_2H_8) \times (1 + 3h/c - 3o/8c)) \qquad \text{———(3)}$$

The relation between λ_g and λ_t (total stoichiometric air ratio) can be approximately determined by Equation (4) (7).

$$\lambda_g = \lambda_t / \eta_{C1} \qquad \text{———(4)}$$

The experimental value of λ_g is distributed between 0.6 and 0.5.

If complete char recycling is attained, approximate cold gasification efficiency η_g can be calculated by Equation (5).

$$\eta_g = 1.32(1 - \lambda_g) \qquad \text{———(5)}$$

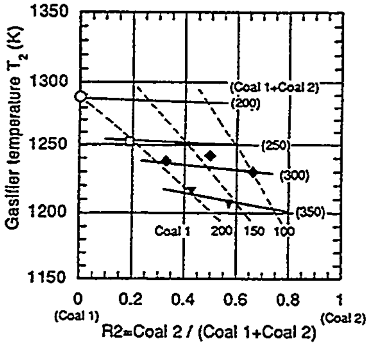

Fig.12 Gasifier Temperature

Gas Temperature in Reductor

Gas temperature (T_2) of the reductor inlet drops as coal2 feed is increased, as shown in **Fig. 12**. How coal feed ratio contributes to the T_2 decrease has not been clarified. Drop in reductor gas temperature with higher coal2 feed is attributable to the following three factors:

① Progress in endothermic reaction for gasification as expressed by the following:

$C + CO_2 \rightarrow 2CO \quad - 160$ MJ/kmol

$C + H_2O \rightarrow CO + H_2 - 118$ MJ/kmol

② Absorption of sensible heat of product char

③ Heat transfer to the furnace wall (the contribution of this factor to decreased gas temperature is large in such a small test system as used for the present study.) Gas temperature in the reductor can be lowered by increasing coal2 feed.

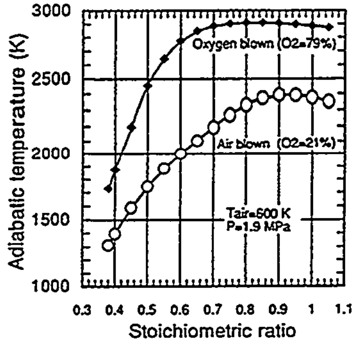

Fig.13 Theoretical daiabatic flame temp. Vs. Air/fuel stoichiometric ratio λ_1

Fig. 13 shows adiabatic flame temperature in coal gasification, as obtained by chemical equilibrium calculation. Adiabatic flame temperature depends largely on stoichiometric air ratio. Theoretically, it drops by 260 K with a stoichiometric air ratio decrease from 0.6 to 0.5. The corresponding temperature drop measured in our test was about 100 K. Drop in measured temperature was smaller, presumably due to thermal absorption by various parts of the combustor, as well as to slow progress in endothermic reaction for gasification. In the test, CO_2 concentration did not drop below 8% even when total coal feed was increased to 350 kg/h, although the equilibrium concentration of CO_2 at a stoichiometric air ratio of 0.5 and a gas temperature of 1,273 K was not above 5%. CO_2 corresponding to this difference in concentration remained in the product gas, producing sensible heat. This is why the T_2 temperature drop was smaller.

Progress in coal gasification reaction is hampered when coal2 feed ratio is increased. In two stage coal supply, therefore, coal2 feed ratio should be decreased as low as possible, so far as stable slag discharge is secured. The reaction in the CPC furnace is more active, presumably because it takes place under high-temperature conditions suitable for fusion of slag, and because the residence time of pulverized coal increases due to swirling flow.

EFFECT OF TWO STAGE COAL SUPPLY

Stable Molten Slag Discharge

In a gasifier with a single stage coal supply, gas temperature drops as stoichiometric air ratio is decreased. When gas temperature drops to a certain level, stable slag discharge is hampered, as is obvious from the theoretical adiabatic flame temperature calculation shown in **Fig. 13**. If coal is supplied at two stages, it is possible to increase the stoichiometric air ratio λ_1 [Equation (2)] of the CPC furnace (to which coal1 is supplied) while maintaining the total stoichiometric ratio λ_t at a certain level. Accordingly, it is possible to operate the system with low total stoichiometric air ratio without sacrificing the melting performance.

This also applies to the oxygen (or oxygen-enriched air)-blown gasification process. However, with the same stoichiometric air ratio, oxygen-blown process results in higher gas and slag temperatures than the air-blown process, due to lower N_2 concentration. Therefore, stable slag discharge is attained more easily with the oxygen-blown process. In addition, since the product gas from the oxygen-blown process has sufficiently high heating value, it is not necessary to operate the system with very low stoichiometric ratio in order to secure a certain heating value. In this sense, two stage coal supply is more effective in the air-blown gasification process than in the oxygen-blown process.

Prevention of Slagging

With single stage coal supply, molten slag from the CPC and fly molten slag in the product gas are cooled gradually after leaving the CPC. When the molten slag temperature drops to a point (between ash fluid point and ash softening point), slag does not provide fluidity but deposits and grows (so-called slagging), possibly hampering system operation. One possible means of preventing slagging is, as described in Section 2, to install a slag-free duct at a point where the product gas flows vertically upward, so that slag deposit can be removed periodically. Another way is to rapidly lower the product gas temperature. Rapid mixing of low temperature gas, and two stage coal supply, as employed in the present study, are also effective means .

A realistic means of preventing slagging problem is to combine various measures selected from the following.

① CPC is operated at lowest possible temperature (with low stoichiometric ratio).
② Slag-free duct is installed at a point where product gas flows vertically upward.
③ Wall surface of slag-free duct is designed to be flat, smooth, and low temperature, so as to minimize slag adhesiveness.
④ Coal is supplied to effect gasification endothermic reaction, thereby lowering the product gas temperature.
⑤ Coal is supplied through axial flow so that it does not collide with the duct wall.
⑥ Low temperature gas is mixed in product gas line to lower the product gas temperature.
⑦ Growing slag deposit is mechanically removed.

Measures ① through ⑤ are employed in operating our bench scale test . No slagging problem has so far been observed. For the oxygen-blown process, operation with low stoichiometric ratio is not guaranteed to prevent slagging, and can cause the furnace temperature to rise excessively. For this process, therefore, it is appropriate to use measures ⑥ and ⑦ while supplying coal in a single stage.

PLAN OF 25T/DAY PILOT TEST

To make a longer continuous operation test with a higher gasification pressure, 25 t/d coal pilot scale test is planed. Design pressure of CPC is 2.6 MPa, and normal operating pressure is 2.0 MPa. This system comprises mostly the same facilities as the bench scale test.

The operation starts in October 1996. Not only air-blown PCPC but oxygen-enriched air-blown one can be tested.

Fig.14 shows the structure of the 25 t/d coal capacity PCPC.

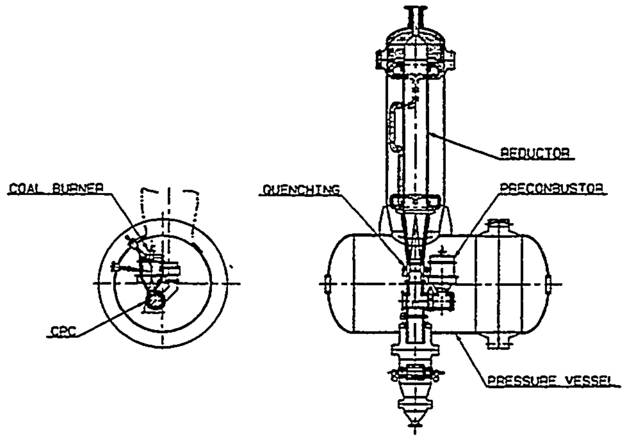

Fig. 14 Structure of 25 t/d PCPC

CONCLUSIONS

Coal gasification characteristics of pressurized coal partial combustor with two-stage coal supply were studied at various coal feed ratios, using the CPC bench scale test facilities of 7 t/d coal capacity, 0.4 MPa gasifier. Our conclusions are as follows.

(1) In the air-blown coal gasifier, as the gasification reaction progresses, adiabatic flame temperature drops, causing loss of slag fluidity. To maintain a temperature that secures stable slag discharge, it is effective to supply coal at two stages: coal1 feed to CPC, and coal2 feed to CPC outlet. The coal2 feed triggers gasification endothermic reaction, quickly decreasing product gas temperature (prevention of slagging).

(2) When coal2 feed ratio is increased with total coal feed fixed, the pressurized CPC

provides the following gasification characteristics:

① Concentration of CO and H_2 (primary combustible components) decreases.
② As a result, heating value of product gas decreases.
③ Hydrocarbons such as methane and ethane increase, possibly producing the problem of increased tar discharge.
④ One-through coal conversion efficiency drops, blocking the coal gasification reaction.

(3) Accordingly, in two stage coal supply, coal2 feed ratio should be decreased, so far as stable slag discharge is secured.

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Pilot Performance of the Integrated Flue Gas Treatment Condensing Heat Exchanger with Orimulsion™ - *R. T. Bailey, K. H. Schulze and J. J. Warchol, Babcock & Wilcox - Research and Development Division; D. W. Johnson, Babcock & Wilcox - Environmental Equipment Division, USA*

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REGENERATION OF SODIUM WASTES FROM FLUE GAS DESULFURIZATION PROCESSES

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The damaging effects that sulfur dioxide emissions are having on the environment are well documented and acknowledged by most scientists, policy makers, and the public at large. These emissions stem primarily from electric utilities and industrial plants with relatively little contribution from the transportation sector. In order to prevent additional environmental damage, clean air legislation has been passed by Congress culminating in the Clean Air Act Amendments of 1990. Ultimately, this legislation will cut sulfur dioxide emissions in half from 1980 levels, and substantially reduce nitrogen oxides emissions from electric utility plants. This legislation should therefore result in a cleaner environment than we have today.

Title IV of the Clean Air Act sets as its primary goal the reduction of annual SO_2 emissions by 10 million tons below 1980 levels[3]. Phase I of the program, which begins this year, will affect 110 utility plants in mostly midwestern and eastern states. Phase II, which begins in the year 2000, tightens the annual emissions allowances on these large, higher emitting plants and also sets controls on smaller units with capacities of greater than 25 megawatts and on all new utility plants. It is estimated that 2,200 plants nationwide will be affected by Phase II. In many cases, Phase I compliance can be met by coupling allowance trading with the blending of low sulfur western coals and regional coals. But there will be other cases in which stack gas cleanup will be the economically most attractive option. Phase II compliance will require virtually all utilities to install some form of flue gas desulfurization equipment. Retrofitting will therefore become of increasing importance as Phase II is implemented.

FLUE GAS DESULFURIZATION WITH SODIUM SORBENTS

Sodium based sorbents for Flue Gas Desulfurization are orders of magnitude more reactive than are the calcium based sorbents, and furthermore, the sodium sorbents react with SO_2 to near completion; whereas, reaction rates for calcium sorbents diminish with

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[1] Professor of Chemical Engineering.

[2] Graduate Student.

[3] anonymous, "Acid Rain Program Overview", U.S. Environmental Protection Agency Rept., EPA430/F-92/019, Dec., 1992.

conversion due to product layer diffusion resistances[4]. If advantage is taken of the higher reactivity, lower capital costs may be realized, and the complete utilization of sorbent helps to off-set the higher cost of the sodium sorbent.

Dry Sorbent Injection (or Dry Sodium Injection - DSI), is a simple process whereby pulverized nahcolite (sodium bicarbonate) is injected dry into the flue gas duct of an electric utility where the reaction:

$$2NaHCO_3 + SO_2 + \frac{1}{2} O_2 \rightarrow Na_2SO_4 + H_2O + 2CO_2$$

takes place. The product sodium sulfate is collected along with fly ash in the baghouse or electrostatic precipitator. The simplicity of the process makes it an ideal candidate for retrofitting. DSI has been the subject of much study over the last twenty years, and several large scale tests have been successfully completed. A number of commercial installations in both the power and manufacturing industries have evolved from these tests. The flagship installation is Wisconsin Electric's Port Washington facility where DSI has been operating on Unit #1 for over a year, and Unit #4 came on stream this past summer. Public Service Company of Colorado in cooperation with the U.S. Department of Energy and the Electric Power Research Institute is currently evaluating integrated dry NO_x/SO_2 emissions control systems[5]. An interesting feature of this study is that both calcium and sodium sorbents are under evaluation on a common ground.

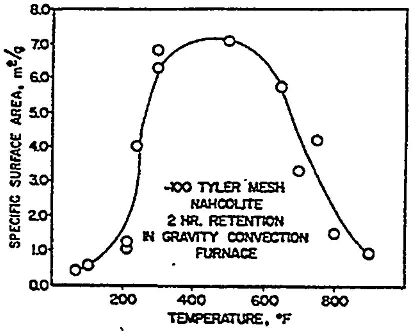

Figure 1. Effect of activation temperature on nahcolite surface area[6].

As indicated above, the preferred sorbent for DSI is sodium

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[5] Hunt, T., Smith, R., Muzio, L., Jones, D., Mali, E. and J. Stallings, "Current Progress with the Integrated Dry NO_x/SO_2 Emissions Control System", 3rd Clean Coal Conf., Chicago, IL, Sept. 6-8, 1994.

bicarbonate. However, trona or sodium sesquicarbonate - $Na_2CO_3 \cdot NaHCO_3 \cdot H_2O$, is sometimes an attractive alternative due to its lower cost. The explanation for sodium bicarbonate's higher reactivity lies in the manner in which it decomposes to produce a high surface area sodium carbonate. This is illustrated Figure 1 taken from the thesis by Stern[6]. Generally it has been observed that sodium bicarbonate or nahcolite decomposes rapidly when heated to temperatures above about 100°C (212°F), and that sintering occurs at temperatures much above 300°C (571°F). Provided that sintering has not occurred, the particle size is unchanged, and a pore structure develops due to the decrease in molar volume of Na_2CO_3 (20.93 cm^3/mol Na) compared to $NaHCO_3$ (38.92 cm^3/mol Na). It is this ability to develop internal porosity and surface area that is responsible for the higher reactivity of the bicarbonate in DSI applications.

Another flue gas desulfurization process which may advantageously employ sodium sorbents is spray drying. Spray drying of the sorbent often provides an attractive alternative to conventional wet scrubbing. In this process a fine mist composed of an aqueous slurry or solution of the reagent is sprayed into the flue gas duct. The water evaporates and the dry product is collected by the baghouse or electrostatic precipitator. Since the product is a solid, this too is a "dry" process, not to be confused with DSI. By comparison to wet scrubbers, spray driers are less capital intensive, and maintenance costs tend to be lower. Many installations employing this technology are in operation world wide. The usual reagent is hydrated or "slaked" lime, but finely ground limestone is also effective[7]. In the latter case, humidity control is important, and the system must operate at 60% relative humidity or higher. This can lead to operational problems in the fabric filter[5].

Spray drier scrubbing of SO_2 with trona solution is a very effective means of flue gas desulfurization[8]. High SO_2 removals and sorbent utilization efficiencies are observed, and there is no need for humidification. In contrast to the DSI results, sodium carbonate appears to be at least as effective as sodium bicarbonate. At low temperatures only the carbonate reacts; whereas, the reaction of bicarbonate requires that temperatures be above the decomposition temperature.

Sodium sorbents can also be employed in conventional wet scrubbing processes. In western Wyoming a somewhat unique situation exists due to the availability of inexpensive waste streams from the processing of trona. The Jim Bridger and Naughton

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[6] Stern, F. R., "Bench-Scale Study of Sulfur and Nitrogen Oxides Adsorption by Nahcolite and Trona", M.S. Thesis, Univ. of North Dakota, Dec., 1978.

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plants operated by Pacific Power both employ wet scrubbing with sodium sorbents.

Despite the successful demonstrations of sodium-based FGD technology and the recent commercializations, there are disadvantages of these processes that have worked to slow development. The first and perhaps most serious problem is sorbent cost. Estimated prices of several SO_2 sorbents are listed in Table 1. For a once through process, soda ash cost is two to three times more expensive than the cost of calcium based sorbents. The cost of SO_2 removal using sodium bicarbonate seems prohibitive. However, natural reserves of nahcolite, the mineral form of sodium bicarbonate, are available, and recently a solution mine was opened near Rifle, CO, by NaTec Resources, Inc. This may bring the cost down from the figure in the table, but there is still a considerable cost advantage ascribed to calcium based sorbents. Recognizing that most of the natural deposits of sodium compounds are located in remote regions of the country, an EPRI study concludes that transporting dry sodium compounds to coal-fired power plants also constitutes a major expense[9].

Table 1. Bulk Prices Estimates for SO_2 Sorbents

	\$/ton	\$/ton SO_2
Limestone, $CaCO_3$	25	39
Quicklime, CaO	40	35
Hydrated Lime, $Ca(OH)_2$	53	61
Soda Ash, Na_2CO_3	70	116
Sodium Bicarbonate, $NaHCO_3$	250	656

A second major problem associated with sodium based sorbent systems is the waste disposal problem. Sodium sulfate is water soluble, and some might argue that an air pollution problem has been transformed into a water pollution problem. Indeed, a study by the Electric Power Research Institute explored a number of waste disposal alternatives and concluded that "most conventional waste management practices will need to be altered before dry sodium wastes can be adequately managed"[10]. They recommend reuse options for sodium injection wastes such as recovery of sodium compound. Once separated from fly ash, sodium sulfate has a market value in several industries such as the kraft paper industry, the detergent

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[9] Bronzini, M. S., Middendorf, D. P. and J. L. Abbott, "Transportation Cost Estimates for Sodium Compounds", Elec. Power Res. Inst. Rept., CS-4764, Sept., 1986.

[10] Dawson, G. W., Eklund, A. G., Delleney, R. D., Achord, R. D. and J. B. Owens, "Laboratory Characterization of Advanced SO_2 Control By-Products: Dry Sodium and Calcium In-Duct Injection Wastes", Elec. Power Res. Inst. Rept., EPRI GS-6622, Dec., 1989.

industry and the glass industry. The existence of vast geological deposits of water soluble materials such as sodium chloride, trona, nahcolite and sodium sulfate testifies that conditions do exist in nature to enable safe storage for indefinite periods of time. However, the practical disposal of sodium wastes will no doubt tend to be site specific.

If an inexpensive means could be found to regenerate the FGD sorbent (Na_2CO_3 or $NaHCO_3$) from the waste product (Na_2SO_4), and at the same time produce sulfur in a saleable or readily disposable form, the problems stated above would be essentially solved. Elemental sulfur would be the most desirable sulfurous product. Regeneration facilities might be built in locations that could service several utility plants. Soluble sodium salts would be recovered on the utility plant site, and the crude waste product would be transported to the regeneration facility. Sodium sorbent would be recovered and shipped back to the utility. Elemental sulfur would be sold on the market. For those manufacturing industries that utilize sodium based sorbents for SO_2 removal, it might be possible to integrate the regeneration step into the particular manufacturing process. Petroleum refineries, for example, will have available a number of reducing gas streams, and sulfur recovery facilities will already be on site.

SOLVAY (AMMONIA-SODA) PROCESS FOR Na_2CO_3 RECOVERY

For many years, the Solvay, or ammonia-soda process was the means by which soda ash was manufactured in the United States and abroad. A process based on commercially proven Solvay technology can be easily devised for regenerating sodium waste from FGD. In the early 1980s laboratory work on such a process was performed by Versar, Inc., under contract by the U.S. Dept. of Energy/Grand Forks Technology Center[11]. This was followed by operation of a 10 lb/hr pilot plant at the Grand Forks Energy Technology Center[12]. Briefly, the sorbent regeneration process may be divided into three parts: 1. leaching of the spent sorbent, 2. conversion of the leachate to a NaCl brine and 3. production of $NaHCO_3$ from this brine via commercially proven Solvay process technology. The first part consists of aqueous leaching of spent sorbent material, and separation of the leachate from the residual insoluble solids which are essentially coal-derived flyash materials. The leachate is a concentrated solution of Na_2SO_4 and Na_2CO_3 with other minor components. In the second part of the process, a recycled liquor from the end of the process consisting mainly of dissolved $CaCl_2$ and unconverted sodium constituents is added to the leachate. A

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[11] Kapsalopoulou, A. J., Sargent, D. H. and E. F. Rissman, "Regeneration of FGD Dry-Sorbent Materials. Phase I - Final Report", U. S. Dept. of Energy rept. DOE/FC/10179-2, May, 1982.

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waste sludge of $CaCO_3$ is formed, and the NaCl brine is isolated by filtration. This brine is analogous to the feed in a conventional Solvay process. In the third part of the process, the NaCl brine is first ammoniated. Next, carbon dioxide is added. The presence of ammonia facilitates the absorption of CO_2 by converting it to dissolved NH_4HCO_3. As more CO_2 is added, crystals of sodium bicarbonate precipitate from solution because $NaHCO_3$ is the least soluble of the sodium salts in the system. The product $NaHCO_3$ is filtered and dried (regenerated FGD sorbent). Ammonia is recovered from the filtrate for recycle by addition of lime and steam stripping. The product liquor is a solution of calcium chloride along with some sodium chloride which did not precipitate as $NaHCO_3$. This stream is recycled back to the second part of the process.

The recovery of $NaHCO_3$ by this process will be expensive. A preliminary cost estimate (in 1982 dollars) for the recovery process alone set revenue requirements at $658 per ton of SO_2 removed[12]. This might be compared to the cost of sulfur allowances under Phase I of the Clean Air Act which are expected to trade in the range of $235-$295 per ton of SO_2 removed[13]. The prospects thus seems remote that sodium recovery based on Solvay technology will become economically attractive.

DIRECT REDUCTION PROCESS

Direct reduction is potentially an attractive means of converting sodium sulfate to sodium carbonate and hydrogen sulfide:

$$Na_2SO_4 + 3H_2 + CO \rightarrow Na_2CO_3 + 2H_2O + H_2S$$

The hydrogen sulfide can then be converted to elemental sulfur as discussed later. In order to insure good selectivity to sodium carbonate, conditions must be found to eliminate the competing reaction to form sodium sulfide:

$$Na_2SO_4 + 4H_2 \rightarrow Na_2S + 4H_2O$$

As a matter of fact, the chemicals recovery furnace of a kraft paper mill is designed to produce sodium sulfide by the second reaction. Sodium carbonate is also produced, but this is only because the furnace operates with excess sodium over that required to react with the sulfur. A discussion of the kraft recovery furnace will be deferred until later.

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[13] anonymous, "Creative Clean Air Act Compliance Strategies: Combining Dry Sorbent Injection, Coal Blending and Allowance Trading", unpublished study by NaTec Resources, Inc., 1177 West Loop South, P.O. Box 56571, Houston, TX 77256.

Thermodynamics of Direct Reduction Processes:

A thermodynamics analysis of the direct reduction of sodium sulfate was presented some years ago by Bauer and Dorland[14]. Their subject was the kraft recovery furnace, and consequently certain constraints were imposed on the system which are not applicable to the present system. Still, some of the conclusions from this study are of interest. For example, it was shown that Na_2CO_3 is prominent at all temperatures investigated from 800 K to 1600 K. Also, it was determined that NaOH and Na_2O cannot exist at any region of the phase diagram at temperatures above 1000 K. Sodium sulfate, Na_2SO_4, is the principal sulfur compound in an oxidizing atmosphere, and the amounts of elemental sulfur as S_2 are negligible over the whole range of conditions investigated.

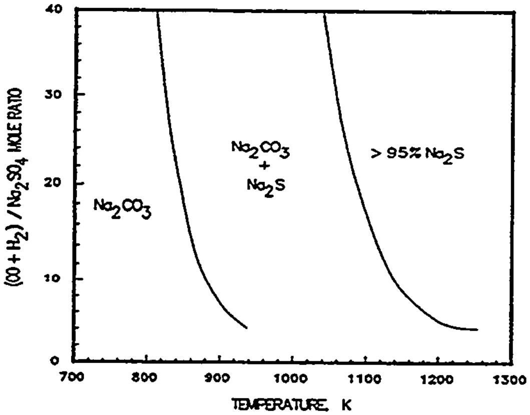

Figure 2. Sodium sulfate reduction products distribution as a function of temperature. H_2/CO mole ratio = 1. P = 1 atm.

We have performed our own thermodynamics calculations aimed principally at identifying a region where the formation of sodium carbonate is favored over sulfide formation. The following components were considered in the analysis: Na_2SO_4, Na_2CO_3, Na_2S, H_2, H_2O, CO, CO_2, H_2S, S_2, S_6, S_8, O_2, SO_2 and SO_3. With a total of 14 components composed of 5 elements (Na, S, O, C, and H), the analysis requires that a total of 9 independent stoichiometric reactions be written. The complex system of equations that results was solved by the Reactors-in-Series method[15] using a commercial

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computer package, THERMOPAK™[16].

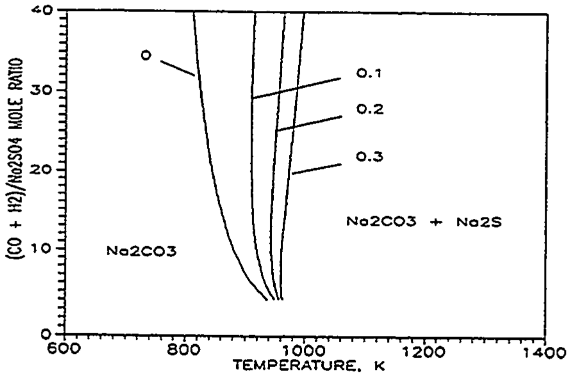

Figure 3. Effect of H_2O in feed. Parameter = moles H_2O/mole (CO + H_2). P = 1 atm.

The results of these calculations are plotted in Figures 2 to 4. Referring first to Figure 2, where the moles of reducing gas in the feed are plotted vs. temperature, a region has been found where only Na_2CO_3 is present. No sulfide can exist in the region identified as Na_2CO_3, i.e. below about 825 to 900 K. Some Na_2CO_3 is present at higher temperatures and vanishingly small amounts are present even in the region identified as Na_2S. The curve which establishes this region corresponds to points at which 95% of the sodium is present as Na_2S. Between the two curves, both Na_2CO_3 and Na_2S are present in substantial amounts.

The presence of water in the feed has the effect of expanding the Na_2CO_3 zone to higher temperatures, Figure 3. The addition of carbon dioxide to the feed has a similar effect, Figure 4. Therefore it is not necessary to dry the sulfate leachate, and on the contrary, the presence of water is beneficial. It is not shown in these figures, but the addition of water and/or carbon dioxide to the feed also shifts the Na_2S zone to the right. The feed H_2/CO ratio has a relatively minor, but still significant, effect on these curves.

The thermodynamics analysis has shown that no elemental sulfur is formed under the reducing conditions required for the conversion of Na_2SO_4 to Na_2CO_3. If oxygen is added to the system, all that in excess of the amount required to oxidize H_2 and CO reacts to convert the sulfide back to the sulfate. Thus, for:

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[16] available ChemE Computations, P. O. Box 4056, Laramie, WY 82071.

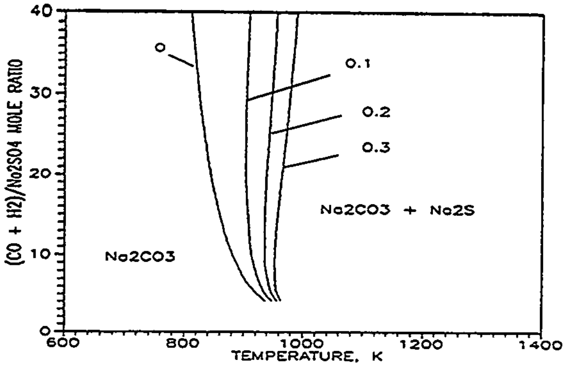

Figure 4. Effect of CO_2 in feed. Parameter = moles CO_2/mole ($CO + H_2$). P = 1 atm.

$$\Delta = P_{O2} - \frac{1}{2}P_{H2} - \frac{1}{2}P_{CO}$$

greater than zero, all the sulfur recombines with sodium to form the sulfate[14]. However, if the reduction can be carried out at sufficiently low temperatures as identified in Figure 2-4, gaseous H_2S is readily separated from the solid products where it can be processed separately.

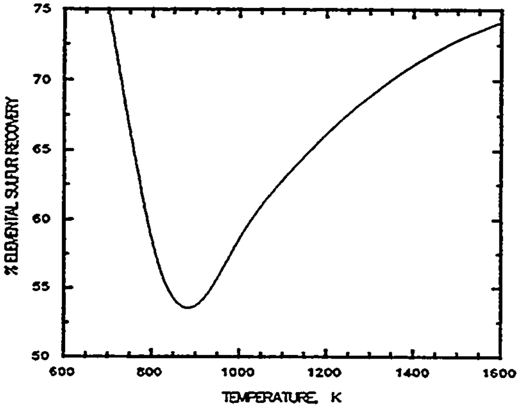

Figure 5. Percent elemental sulfur recovery as a function of temperature. Stoichiometric feed ratio. P = 1 atm.

A thermodynamics calculation on the system H_2S, O_2, H_2, H_2O, S_2, S_6, S_8, SO_2 and SO_3 is shown in Figure 5. Here the percent sulfur recovered as elemental sulfur is plotted as a function of temperature. The feed is assumed to be in stoichiometric proportion according to the reaction:

$$2H_2S + O_2 \rightarrow S_2 + 2H_2O$$

The minimum in this curve is a consequent of the fact that S_6 and S_8 are prominent at low temperatures; whereas, S_2 is the only form of elemental sulfur present at high temperatures. Sulfur recovery by the Claus and related processes is widely practiced in the chemical industry.

Experimental Studies:

Favorable thermodynamics is a necessary, but not sufficient requirement for reaction to take place. Thus, while our thermodynamics analysis has identified a region where sodium sulfate may be converted to sodium carbonate, there is no guarantee that the reaction will take place, or if it does that the reaction rates will be sufficient for a viable process.

A search of the literature has identified only one study devoted to the reduction of Na_2SO_4 to Na_2CO_3 and H_2S. This was a Russian study conducted in the mid-1950s. In these experiments, a reducing gas was passed through a fixed bed of Na_2SO_4. The reducing gas was generated by several different procedures. In the first paper, CO_2 was passed through a bed of hot charcoal and the CO produced was bubbled through water[17]. In the second series of experiments, air was fed to the hot charcoal instead of CO_2[18]. In both cases the product from the gas generator contained little hydrogen. In the third paper, water and air were fed to the gas generator producing a product gas containing H_2 and CO in roughly equal proportions[19]. And finally, in a fourth paper the process was studied on a larger scale in a moving bed reactor measuring 0.286 m in diameter and 1.8 meters in height[20].

Due to analytical limitations, the experimentation was rather crude. No material balances were reported, and seems impossible to obtain a quantitative measure of reaction rates. Still, useful information is contained in these reports. The solid phase products present in significant quantities were: Na_2CO_3, Na_2S, Na_2SO_3 and unreacted Na_2SO_4. As expected from the thermodynamics discussion, no NaOH was reported in the products, but reference was made to its possible presence as an intermediate. The only

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[17] Gorbanev, A. I., Zhur. Priklad. Khim. 27, 804(1954).

[18] Gorbanev, A. I., ibid 27, 921(1954).

[19] Gorvanev, A. I., ibid 27, 1033(1954).

[20] Gorbanev, A. I., ibid 8, 1270(1958).

sulfurous product reported other than Na_2S and Na_2SO_3 was hydrogen sulfide.

The presence of H_2 in the feed gas lowered the reaction temperature and extended the range over which the reaction occurred from about 590°C (863 K, 1094°F) to 680°C (953 K, 1256°F). This corresponds approximately to the phase boundary for the pure Na_2CO_3 region on Figures 2-4. At equal concentrations of CO and H_2 the briquets did not melt or stick together, but when the H_2 content considerably exceeded that of CO, surface melting became a problem. This was attributed to the production of excessive Na_2S[21]. At the optimum conditions of roughly 630°C (903 K, 1166°F), $P_{H2O} = 0.35$ atm, 16 to 19% CO, the same H_2, and 10 to 13% CO_2, 97.2% reduction of the Na_2SO_4 was obtained with a Na_2CO_3 yield of 92.7%. When the H_2 content was raised above 20%, the surface began to melt.

Other noteworthy observations were that Fe_2O_3 in small amounts (~0.2%) served as an effective catalyst and lowered the reaction temperature. At temperatures of the order of 590°C to 600°C (1094°F to 1112°F) carbon formation was sometimes observed. This could be due to the reverse Boudouard reaction. Finally, gas utilization early in the experiment was normally high and tapered off with time suggesting that mass transfer resistances might be coming into play. Cylindrical briquets 15 mm in diameter were employed in the large scale tests.

Kraft Recovery Furnace Operation:

A process for regeneration of sodium waste products has been in operation for many years in the paper industry. In the chemicals recovery furnace of the kraft (sulfate) paper process, concentrated black liquor from the digesters is burned to recover Na_2CO_3 and Na_2S[22]. Black liquor contains a wide variety of organic and inorganic constituents including sodium salts of the organic acids formed from carbohydrates and lignin during the alkaline cook, sodium salts (soaps) of the resin and fatty acids from the wood, sodium carbonate, sodium sulfate, sodium thiosulfate and other residual alkali components. More than enough sodium is present to tie up all the sulfur as Na_2S or Na_2SO_4. This material along with makeup sodium, usually as Na_2SO_4, is burned in the recovery furnace where the inorganic sodium salts, principally Na_2CO_3, Na_2S and some unreacted Na_2SO_4, form a "smelt" which flows out of the furnace for recausticizing with slaked lime.

Conditions in the recovery furnace are variable and complex. Concentrated black liquor is injected into the furnace through spray guns just above the hearth, and the droplets as they descend

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[21] The melting point of Na_2S is 950°C, well above reaction temperatures. However, Na_2S will form solution - melts with other components of the mixture, most notably Na_2CO_3, at lower temperatures.

[22] Casey, James P., ed., *Pulp and Paper Chemistry and Chemical Technology*, 3rd ed., vol. I, Wiley-Interscience, New York, 1980.

are dried, pyrolyzed, and gasified with H_2O, CO_2 and substoichiometric oxygen. The char product falls to a bed in the lower part of the furnace. Temperatures in the bed are generally well in excess of 815°C (1500°F), where conversion of the black liquor compounds to Na_2CO_3 and NaS_2 begins to take place. Actual bed temperatures are difficult to measure or predict, but in one experimental simulation of the char burning region, temperatures of the order of 1180°C (2155°F) were observed[23]. Air introduction to the char bed is carefully controlled to maintain a reducing atmosphere thereby keeping the smelt in the reduced state. Thus conditions in the lower region of the furnace are thermodynamically favorable for Na_2S formation, Figures 2-4. The smelt, melting point below 760°C (1400°F), flows out at the bottom of the furnace through water cooled smelt spouts.

Additional air is injected into the upper section of the furnace to assist in the combustion of gases and organic volatiles to SO_2, CO_2 and H_2O. The SO_2 reacts with Na_2CO_3 "fume" and oxygen to form Na_2SO_4. This material is then captured by electrostatic precipitators and recycled back to the reduction zone. Recapture of sulfur by this mechanism is important for emissions control. Too much fume is detrimental to performance, however, since it also serves to foul heat transfer tubes. The source of fume appears to be elemental sodium which forms in the reducing zone and then reacts with CO_2 and O_2 in the oxidizing zone to produce Na_2CO_3[24].

CONCLUSIONS

The direct reduction of Na_2SO_4 to Na_2CO_3 is not only thermodynamically feasible, but the reactions are known to take place at reasonable rates along with several competing reactions. In the only commercial process for recovering sodium wastes, no effort is made to maximize Na_2CO_3 production since the "byproduct" Na_2S is needed in the pulping operation. New processing conditions and configurations must be found which maximize carbonate production. The Russian study suggests that such conditions exist. Hydrogen sulfide is readily converted to elemental sulfur by the Claus process.

More research is needed to characterize the low temperature region where H_2S and Na_2CO_3 are the principal products. In this region no information is available on the reaction kinetics, or for that matter on the principal reaction paths. A research program designed to obtain information on reaction rates would assist in the selection of process options to be investigated in future bench scale or pilot plant studies.

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A SIMULTANEOUS SO_X AND NO_X REMOVAL SYSTEM FOR BURNING ALL GRADES OF COAL

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Burning coal for cooking, heating, industrial uses and for electricity generation is nothing new and has been around for centuries. With the recent advent of global environmental awareness the method of burning and cleaning up the gases from boilers has come under scrutiny by regulatory agencies and even the general public.

Throughout history basically premium coals were used for combustion, but as regulations got more strict, the word premium has constantly changed. There was a time when soot was the most common complaint about coal combustion, but as devices such as cyclones, electrostatic precipitators and baghouses appeared, that problem was solved. However, the unseen elements in gases were being studied and identified and regulations began to appear for SO_2, NO_X and other air toxins.

Unfortunately mother nature did not give us only one grade of coal which is perfect, so mankind had to look for ways to improve the burning and cleanup so that the earths most economical and abundant fuel supply could be used with efficiency and cleanliness. Table 1 shows the typical parameters of various fuels. Concentrating only on the coal portion of the solid fuels, it can be seen that the heating value, sulfur content and ash contents vary tremendously for the three mentioned types of coal.

Table 2 shows coal production in various countries. Many countries use coal predominantly for power generation, but as can be seen the coal is not always cleaned before it is burned. Additionally, the cleaning has focused on reducing ash-forming minerals and up until recently, the removal of sulfur has not been a prime aim of the coal cleaning process.

Table 3 shows the properties of selected international coals, and as can be seen the composition varies a great deal from region to region. Some low sulfur coals have properties that are different from those of high sulfur coals which may make low sulfur coals unsuitable for use in plants designed for high sulfur coal or vice-versa.

Since sulfur in coal my be present in either an inorganic form (as pyrite) or organic form, coal cleaning techniques are limited since at the present time they only remove inorganic sulfur. This leaves the organic sulfur in the coal which often times is such a percentage that it is still a high sulfur coal. Many countries have imposed regulations on the importation of types of coal to be used in their countries, and in most cases, the sulfur is limited to less than 1%. Additionally, as time goes on it is felt that the availability of exportable coals will decrease as the demand for low sulfur coal increases. This would then create a price difference between what would be called premium coal vs. a non-premium coal. If this is the case many countries will export the cleaner burning coals, and use the residual coals for their own power generation. The necessity of obtaining hard currency will drive the markets for these countries. It is conceivable that the need to burn different grades of coal in many areas will persist, including industrialized countries.

Consequently, the necessity to remove unwanted elements from the flue gases will be important for all types of coal. Up until the present time air pollution control equipment operated at relatively low sulfur content levels. There is very little experience in operating processes for high SO_2 content. However, electron-beam process pilot plant tests indicate a linear relationship between NO_x removal efficiency and increased SO_2 concentrations.

The two basic systems, currently in use for SO_2 and NO_x removal are shown in Figure 1. The wet flue gas desulfurization unit (FGD) and selective catalytic reduction (SCR) have been used extensively, but basically for premium grade fuels with low SO_2 and NO_x concentrations in the flue gas. Work is being done in many areas to perfect simultaneous removal systems which operate with a minimum of environmental impact. The electron-beam process, which has been under development for two decades, is one such process that holds promise for the future since it meets many of the requirements that will be necessary for a system of the future.

A basic requirement will be to simultaneously remove SO_2 and NO_x in a dry manner where no residual waste and disposal problems develop. Additionally, it will be necessary to have a by-product produced that will not require any additional waste disposal. The electron beam process fits these requirements very well, as demonstrated by the pilot-plant testing that has been completed in the last several years. Pilot plant tests have been completed in the United States, Japan and Poland. In the United States a pilot plant test was successfully completed in 1988 by a consortium consisting of Ebara Corportation, Indianapolis Power and Light, C. F. Industries, AMAX Coal Corporation and other partners, in a cost-sharing agreement with the U.S. Department of Energy. In Japan, Ebara Corporation, in partnership with Chubu Electric Company and JAERI successfully completed a pilot plant test at the Chubu Electric Utility in Nagoya, Japan. In Poland, other pilot-plant testing work is ongoing at the Institute of Nuclear Chemistry and Technology in Warsaw, Poland. As a result, it is anticipated that the commercial use will soon appear, since it does meet the present and future requirements for an environmentally compatible system.

The electron beam process is a dry-scrubbing process, shown in Figure 2, which simultaneously removes sulfur dioxide (SO_2) and nitrous oxides (NO_x) from combustion flue gases. The irradiation of the flue gas produces active radicals which react with the SO_2 and NO_x to form their respective acids. In the presence of ammonia (NH_3), these acids are converted to ammonium sulfate $(NH_4)_2SO_4$ and ammonium sulfate nitrate $(NH_4)_2SO_4$-$2NH_4NO_3$). Other organic compounds such as VOC's can also be treated using the same principles.

When the electron beam process is used to clean the flue gas from an electric utility boiler, as depicted in Figure 3, the flue gas is first cleaned of flyash by a particle collector. The gas then passes through an evaporative spray cooler where the gas temperature is lowered, as the humidity is increased. The gas then passes to a process vessel where it is irradiated by a beam of high-energy electrons, in the presence of a near-stoichiometric amount of ammonia which was injected upstream of the process vessel. SO_2 and NO_x are oxidized to form H_2SO_4 and HNO_3, respectively. These acids subsequently react with the added ammonia to form ammonium sulfate and ammonium sulfate nitrate. These salts are recovered as a dry powder using a conventional particle collector. The collected powder is potentially salable as an agricultural fertilizer.

It is important that systems utilizing a singular technology be developed to meet the future requirements for the simultaneous removal of both SO_2 and NO_x from both low and high sulfur fuels. The electron beam process fits very well into this category since it utilizes the same basic technology to simultaneously remove both pollutants. The process is a low temperature, dry scrubbing process which does not create any waste water or secondary pollutant problems. Since the utilization and disposal of by-products is becoming extremely important, the electron beam process becomes valuable, both environmentally and economically, since the by-product that is obtained from the process can be used as an agricultural fertilizer.

As far as removals are concerned, the most stringent regulatory requirements would probably require removals of 95% of SO_2 and 80-85% for NO_x removal. The electron beam process economically achieves these values since the process tests have determined that the highest energy requirement is for the NO_x removal. Several new techniques were developed at the Chubu Electric Co. plant, including zone irradiation, and the orientation of the accelerators, have been developed to lower the energy requirements for the NO_x removal.

From Figure 4, it can be seen that the SO_2 removal efficiency is less than 10 kilograys for removals over 95%. This is much less than early tests which indicated that at high temperatures approximately 4 megarads (40 kilograys) were required for similar SO_2 removal.

As far as the NO_x removal is concerned, it can be seen in Figure 5 that at the same dose of less than 10 kilograys, over 85% of the NO_x removal can be achieved. The low dose levels are easily translated into power-generation operating costs which would make the process operating costs equivalent to 2% or less of the total power generated, depending on the inlet SO_2 and NO_x concentrations. For example, if it were installed on a 100 megawatt power station, the total requirements for the simultaneous removal process would be in the range of 2 megawatts.

NO_x removal is becoming very critical in many areas; therefore, it is anticipated that reductions in the allowable limits will be instituted for NO_x control in the future. The electron-beam process will respond to these regulatory changes very well because it has been proven that the system operates very well, even with low NO_x levels in the gas. Shown in Figure 6, recent pilot-plant tests indicate that NO_x removal efficiency is a function of NO_x concentration in the inlet gas; therefore low NO_x concentrations mean high removal efficiencies at low electron-beam doses.

The recent test facility, at Chubu Electric Company, had both a bag filter and a dry ESP as collectors. From the previous figures you can see that the reductions of SO_2 and NO_x would meet any regulatory requirements. The particulate emissions from the system are also within the regulations as shown in Figure 7. In order to minimize pressure drop and operating energy, it was demonstrated that the dry ESP outlet conditions met all of the requirements. This would eliminate the necessity of a bag filter in the system.

Figure 8 is a comparison and evaluation of the $DeSO_x$, $DeNO_x$ and Electron-beam processes, illustrating the operating characteristics, the by-products, the operatability of the system and technical evaluations of each of the three systems. From evaluations, it can be seen that the construction costs are less for an E-beam system due to its simplicity. Additionally, the operating costs are also comparable to conventional systems. It is felt that conventional systems have basically reached their improvement level in potential cost reductions. The E-beam system is now entering the stage where reductions, for installation costs and operating costs, can be made based on new accelerator developments and other system components. Therefore, it is foreseen that costs will be further reduced in years to come.

The following shows the major advantages of the E-beam process, which are:

1) Efficient simultaneous removal of SO_x and NO_x.
2) Easy to operate with an easy system start-up and shut-down.
3) Dry process without waste water treatment.
4) Produces a valuable fertilizer by-product.
5) Lower capital investment and operational expenses are predicted and will be further reduced in the future.

A photograph of the by-product collected from the E-beam process is shown in Figure 9. The by-product powder is collected in a dry ESP and then is granulated for use as a fertilizer. The composition of the by-product, from recent testing, is shown in Figure 10. This composition is comparable to a commercial grade of fertilizer. Figure 11 shows some comparisons of commercial fertilizers versus the by-product, in recent plant growing tests.

Using accelerators for cleaning up flue gases is a reliable and simple method. Electron beam accelerators have progressed in reliability and efficiency throughout the years and many are currently used in other radiation processing applications, such as sterilization, wire and cable manufacturing, food irradiation, cross linking and other applications. Accelerators are safe for operating personnel and the environment. With the growing interest in environmental preservation and remediation, the electron beam process for the treatment of combustion flue gases will continue to grow. The systems are easy to install and use.

Tests have been conducted in various locations for water purification, VOC removal, low NO_x concentrations from tunnel off-gasses, and other potential applications for remediation and chemical processing. Tests have been conducted on heavy metal reductions in liquids and gases, which again would play a beneficial role in cleaning up combustion flue gases. Published data, summarized in the NIST Radiation Kinetics Data Base and from studies made by High Voltage Environmental Applications Inc. (Miami, FL), detail various reactions which provides the basis for understanding the likely interactions of high energy electrons and heavy metals in aqueous solutions, and would be similar to heavy metals in gases. The radiation chemistry of the heavy metals is that of oxidation/reduction. The key to removing metals of interest is to manipulate the radiation chemistry to form insoluble precipitates. For example, Mercury Hg(I) and Hg(II) is somewhat unique in that Hg(II) salts are reasonably soluble whereas the Hg(I) are insoluble. Therefore, the processes that will lead to the removal of Hg are reducing processes. i.e. involving the e-aq or $H^{\cdot}$. A way to facilitate this would be to add an $OH^{\cdot}$ radical. The combination of the electrons, humidity and low temperature in the flue gas could affect heavy meatal reductions.

It is also conceivable that in the future it will be possible to treat coal slurries to remove some of the impurities from the fuel before it is even burned. Removal of these impurities from coal, in such a fashion, would make a very environmentally compatible combustion system, when combined with a post-combustion, electron-beam, flue-gas-treatment process.

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TABLE 1

TYPICAL PARAMETERS OF VARIOUS FUELS[a]

Type of Fuel	Heating Value		Sulfur	Ash
	kcal	BTU	% (by weight)	% (by weight)
Solid Fuels				
Bituminous Coal	7,200/kg	13,000/lb	0.6-5.4	4-20
Anthracite Coal	6,810/kg	12,300/lb	0.5-1.0	7.0-16.0
Lignite (@ 35% moisture)	3,990/kg	7,200/lb	0.7	6.2
Wood (@ 40% moisture)	2,880/kg	5,200/lb	N	1-3
Bagasse (@ 50% moisture)	2,220/kg	4,000/lb	N	1-2
Bark (@ 50% moisture)	2,492/kg	4,500/lb	N	1-3[b]
Coke, Byproduct	7,380/kg	13,300/lb	0.5-1.0	0.5-5.0
Liquid Fuels				
Residual Oil	$9.98 \times 10^6/m^3$	150,000/gal	0.5-4.0	0.05-0.1
Distillate Oil	$9.30 \times 10^6/m^3$	140,000/gal	0.2-1.0	N
Diesel	$9.12 \times 10^6/m^3$	137,000/gal	0.4	N
Gasoline	$8.62 \times 10^6/m^3$	130,000/gal	0.03-0.04	N
Kerosene	$8.32 \times 10^6/m^3$	135,000/gal	0.02-0.05	N
Liquid Petroleum Gas	$6.25 \times 10^6/m^3$	94,000/gal	N	N
Gaseous Fuels				
Natural Gas	$9{,}341/nm^3$	1,050/SCF	N	N
Coke Oven Gas	$5{,}249/nm^3$	590/SCF	0.5-2.0	N
Blast Furnace Gas	$890/nm^3$	100/SCF	N	N

[a] N = negligible.
[b] Ash content may be considerably higher when sand, dirt, etc. are present.

TABLE 2

COUNTRIES (WORLD ENERGY CONFERENCE, 1989)

Country	Total Coal Production Mt	Bituminous, Subbituminous, & Lignite Mt		Approx.% of Bituminous Cleaned
		Mt	Mt	
China	1018	985	33	20
USA	833	581	252	55
USSR	760	550	210	60
Germany	487	77	410	95
Poland	266	193	73	40
Australia	224	179	45	75
South Africa	214	214	0	60
India	191	180	11	20
Czechoslovakia	126	26	100	nd
UK	100	100	0	75
Canada	61	33	28	95+
Other countries	485	164	320	nd
Total	4765	3282	1482	nd

These figures, although published in 1989, refer to the year 1987. Those for Germany are the sum of what was then East and West Germany.

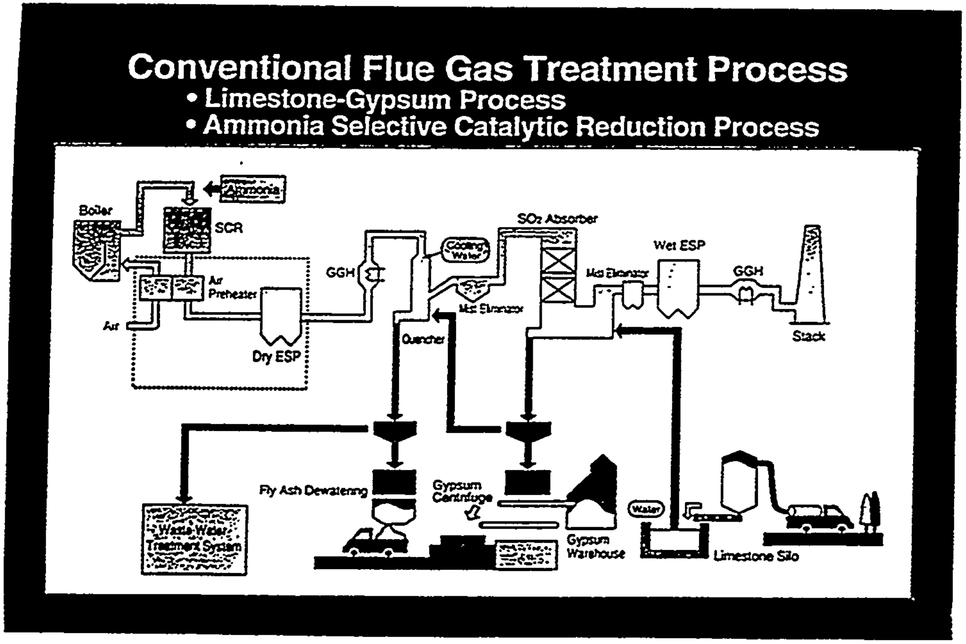
TABLE 3

PROPERTIES OF SELECTED INTERNATIONAL COALS

Source	Australia	China	France	S.Africa	Indonesia	Korea	Spain
Ultimate							
Carbon	56.60	62.67	74.60	69.70	56.53	68.46	37.02
Hydrogen	3.50	3.86	4.86	4.50	4.13	0.90	2.75
Nitrogen	1.22	0.83	1.39	1.60	0.88	0.20	0.88
Sulfur	0.35	0.46	0.79	0.70	0.21	2.09	7.46
Ash	24.00	4.71	8.13	10.10	1.77	23.48	38.65
Oxygen	7.43	10.34	9.42	9.10	12.58	4.38	11.39
Proximate							
Moisture	6.90	17.13	0.80	4.30	23.90	0.50	1.80
Volatile matter, dry	24.80	30.92	36.11	35.30	45.57	7.46	45.27
Fixed carbon, dry	44.30	47.24	54.96	50.30	28.76	68.56	14.24
Ash, dry	24.00	4.71	8.13	10.10	1.77	23.48	38.69
Higher heating value, Btu/lb	9,660	10,740	13,144	12,170	9,840	9,443	6,098

From Steam, 40th Edition, Published by Babcock & Willcox

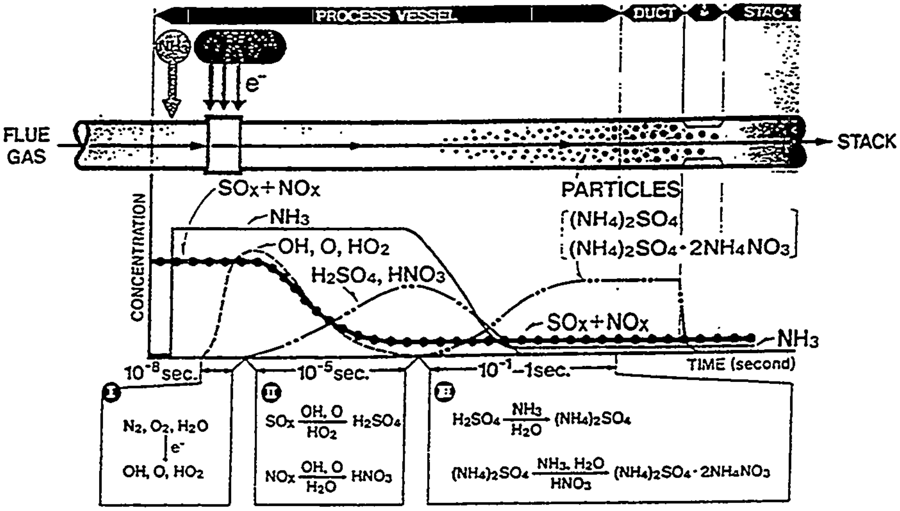
FIGURE 1



Conventional Flue Gas Treatment Process

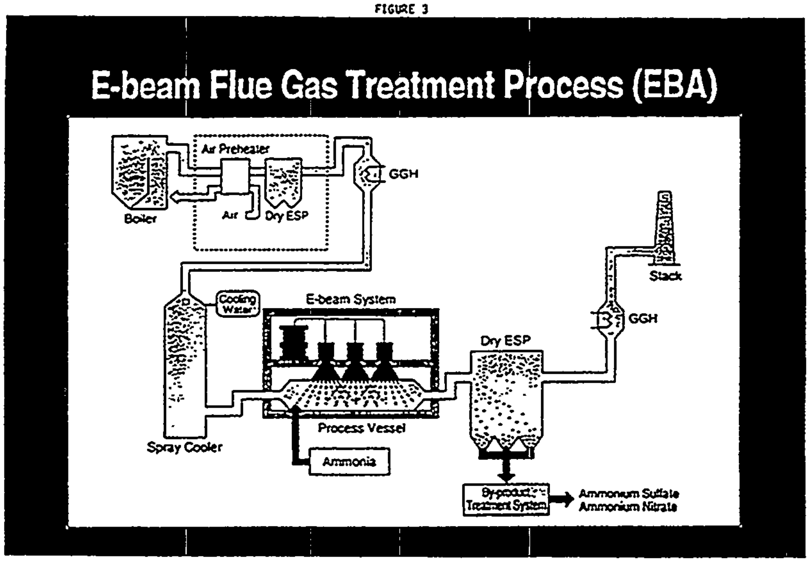
- Limestone-Gypsum Process
- Ammonia Selective Catalytic Reduction Process

FIGURE 2



MODEL DIAGRAM OF REACTION MECHANISM

FIGURE 3



E-beam Flue Gas Treatment Process (EBA)

FIGURE 4

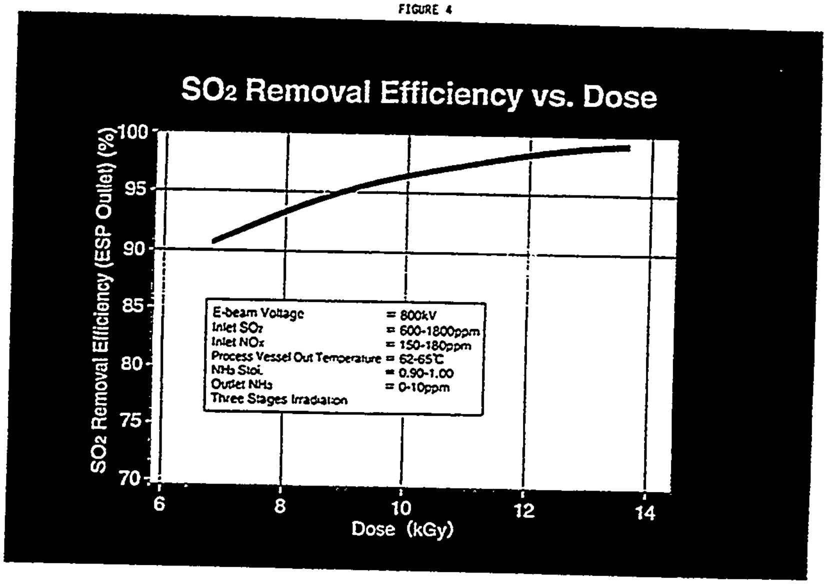


FIGURE 5

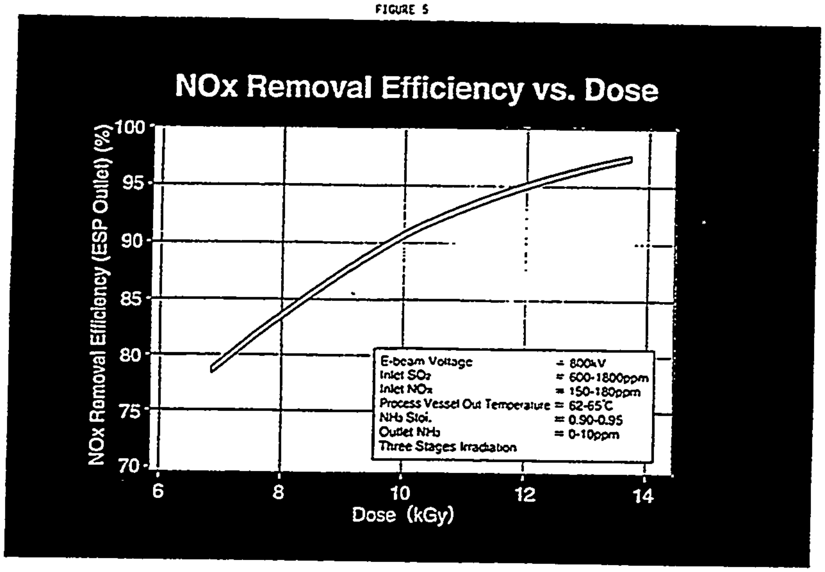


FIGURE 6

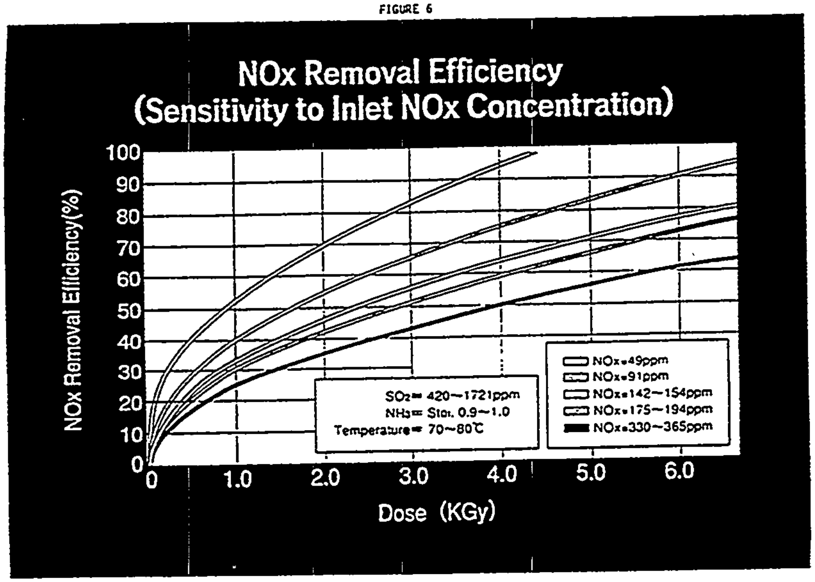
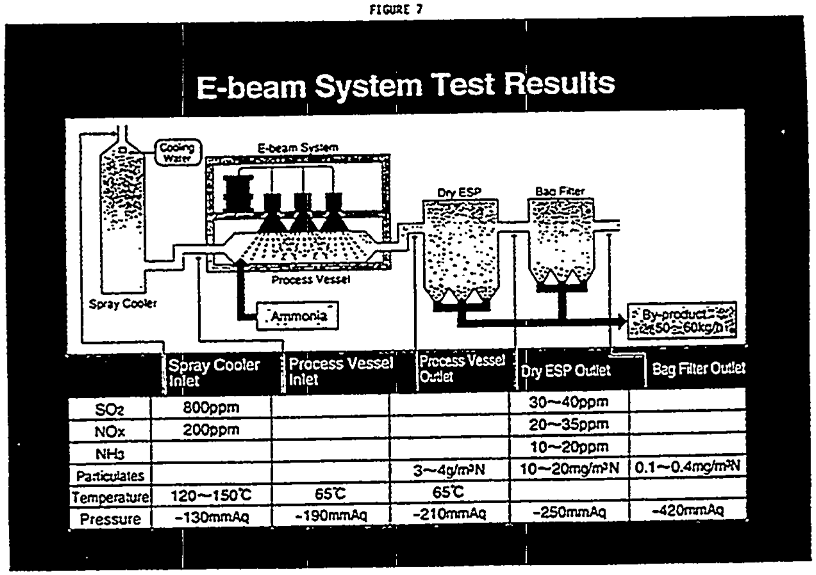
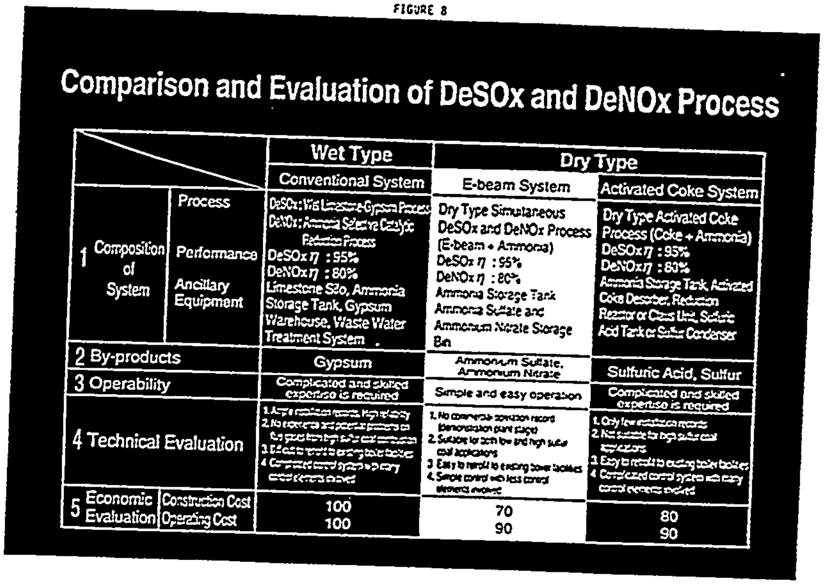

NOx Removal Efficiency
(Sensitivity to Inlet NOx Concentration)

FIGURE 7


E-beam System Test Results

	Spray Cooler Inlet	Process Vessel Inlet	Process Vessel Outlet	Dry ESP Outlet	Bag Filter Outlet
SO_2	800ppm			30~40ppm	
NOx	200ppm			20~35ppm	
NH_3				10~20ppm	
Particulates			3~4g/m³N	10~20mg/m³N	0.1~0.4mg/m³N
Temperature	120~150℃	65℃	65℃		
Pressure	-130mmAq	-190mmAq	-210mmAq	-250mmAq	-420mmAq

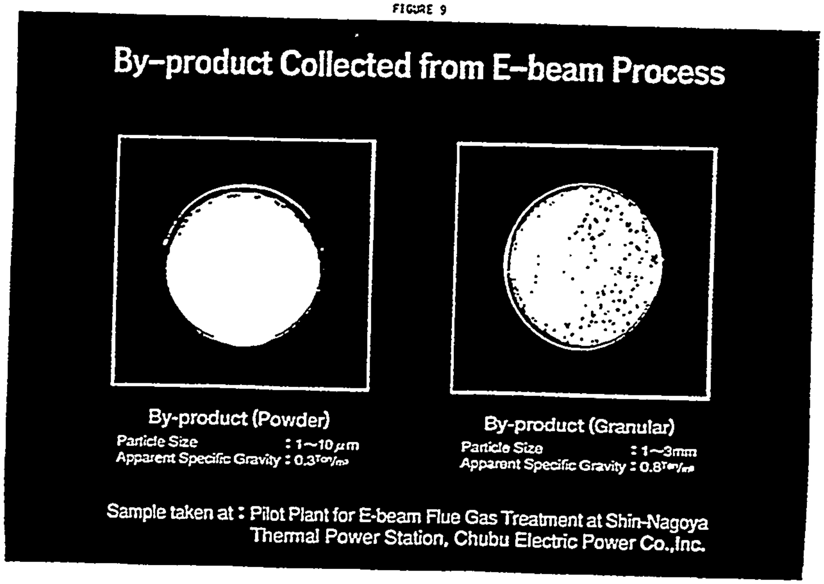
FIGURE 8



Comparison and Evaluation of DeSOx and DeNOx Process

		Wet Type	Dry Type	
		Conventional System	E-beam System	Activated Coke System
1 Composition of System	Process	DeSOx: Wet Limestone-Gypsum Process DeNOx: Ammonia Selective Catalytic Reduction Process	Dry Type Simultaneous DeSOx and DeNOx Process (E-beam + Ammonia)	Dry Type Activated Coke Process (Coke + Ammonia)
	Performance	DeSOx η : 95% DeNOx η : 80%	DeSOx η : 95% DeNOx η : 80%	DeSOx η : 95% DeNOx η : 80%
	Ancillary Equipment	Limestone Silo, Ammonia Storage Tank, Gypsum Warehouse, Waste Water Treatment System	Ammonia Storage Tank Ammonia Sulfate and Ammonium Nitrate Storage Bin	Ammonia Storage Tank, Activated Coke Desorber, Reduction Reactor or Claus Unit, Sulfuric Acid Tank or Sulfur Condenser
2 By-products		Gypsum	Ammonium Sulfate, Ammonium Nitrate	Sulfuric Acid, Sulfur
3 Operability		Complicated and skilled expertise is required	Simple and easy operation	Complicated and skilled expertise is required
4 Technical Evaluation		1. Ample installation records, High reliability 2. No experience and potential problems on the gases from high sulfur coal combustion 3. Difficult to retrofit to existing boiler facilities 4. Complicated control system with many control elements involved	1. No commercial operation record (Demonstration plant stage) 2. Suitable for both low and high sulfur coal applications 3. Easy to retrofit to existing boiler facilities 4. Simple control with less control elements involved	1. Only few installation records 2. Not suitable for high sulfur coal applications 3. Easy to retrofit to existing boiler facilities 4. Complicated control system with many control elements involved
5 Economic Evaluation	Construction Cost	100	70	80
	Operating Cost	100	90	90

FIGURE 9



By-product Collected from E-beam Process

By-product (Powder)

Particle Size : 1~10 μm

Apparent Specific Gravity : $0.3^{\mathrm{Ton}}/_{\mathrm{m}^3}$

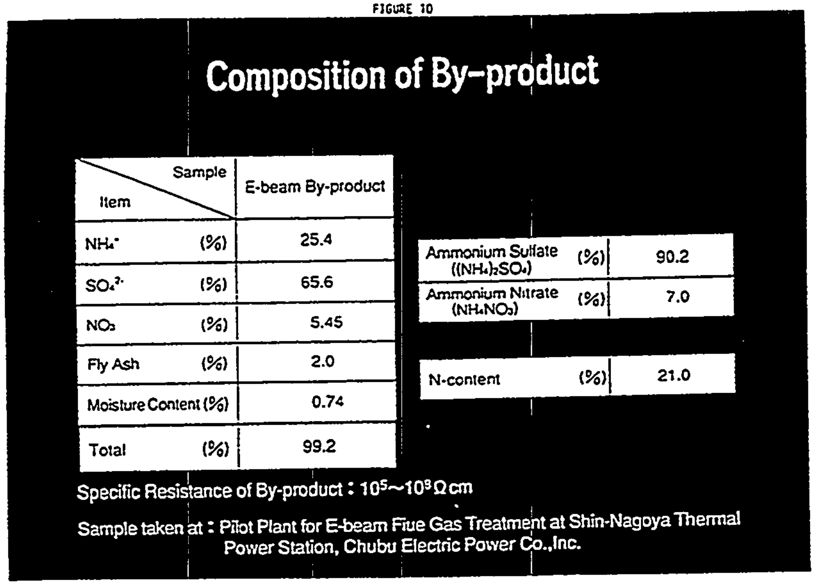
By-product (Granular)

Particle Size : 1~3mm

Apparent Specific Gravity : $0.8^{\mathrm{Ton}}/_{\mathrm{m}^3}$

Sample taken at : Pilot Plant for E-beam Flue Gas Treatment at Shin-Nagoya Thermal Power Station, Chubu Electric Power Co.,Inc.

FIGURE 10



Composition of By-product

Item \ Sample	E-beam By-product
NH_4^+ (%)	25.4
SO_4^{2-} (%)	65.6
NO_3 (%)	5.45
Fly Ash (%)	2.0
Moisture Content (%)	0.74
Total (%)	99.2

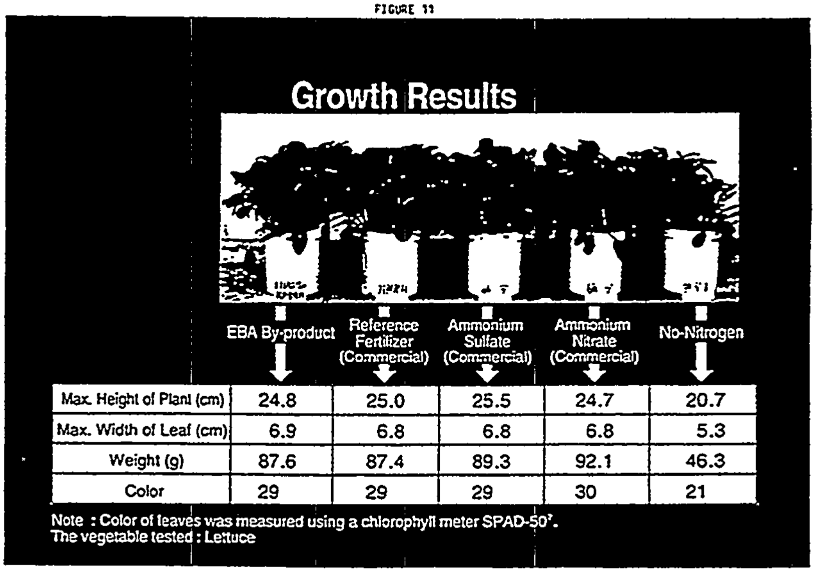
Ammonium Sulfate ($(NH_4)_2SO_4$) (%)	90.2
Ammonium Nitrate (NH_4NO_3) (%)	7.0

N-content (%)	21.0

Specific Resistance of By-product : 10^5~10^9 Ω cm

Sample taken at : Pilot Plant for E-beam Flue Gas Treatment at Shin-Nagoya Thermal Power Station, Chubu Electric Power Co.,Inc.

FIGURE 11



Growth Results

	EBA By-product	Reference Fertilizer (Commercial)	Ammonium Sulfate (Commercial)	Ammonium Nitrate (Commercial)	No-Nitrogen
Max. Height of Plant (cm)	24.8	25.0	25.5	24.7	20.7
Max. Width of Leaf (cm)	6.9	6.8	6.8	6.8	5.3
Weight (g)	87.6	87.4	89.3	92.1	46.3
Color	29	29	29	30	21

Note : Color of leaves was measured using a chlorophyll meter SPAD-50'.
The vegetable tested : Lettuce