COMBINED ULTRAFINE GRINDING AND DRYING OF A LOW RANK COAL

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ABSTRACT

An experimental evaluation of combined ultrafine grinding and drying of a Martin Lake, Texas lignite has been made. Analysis of the energy requirements for the process, determination of various properties of the product, and evalaution of ash isolation are presented in this paper.

A sample of a Martin Lake, Texas lignite was obtained and a series of tests were conducted in a fluid-energy mill at the development facility of the Micro-Energy Division of Ergon, Inc. in Vicksburg, MS. The grinding fluids used were air at 116°F and steam at 225, 310, 350, 400 and 488°F as measured in the mill. The products of these tests were analyzed for volatile matter, ash, total moisture, equilibrium moisture, heating value, density distribution, aerodynamic particle size classification, angle of repose, porosity, density, and particle size distribution. The mill was instrumented and measurements were made to determine the energy requirements for the process. Ultimate analyses and ash mineral content analyses were also made.

In general, the fluid-energy mill was used successfully in simultaneous ultrafine grinding and drying of the lignite. Particle size reduction to less than 10 microns on a population basis was achieved. The equilibrium moisture of the samples decreased with increasing grinding fluid temperature. Density distribution studies showed that a significant fraction of the ground particles were ash rich. This ash rich fraction was about two percent of the mass of the sample and contained approximately fifty percent of the ash in the sample. The energy required for the process increased with increasing temperature.

INTRODUCTION

Research on ultrafine coal has shown that when ultrafine grind coal is burned, much less slagging occurs than with conventional boiler grind pulverized coal (1-5)[1]. The small ash particles in the ultrafine grind coal tend to follow the flow streams around the heat transfer surfaces rather than impacting the tubes and causing slagging. This phenomena also reduces the erosion potential of the ash particles. These ultrafine particles then pass through the plant and are caught in a baghouse. The benefit of such a fuel is that a solid fuel, either as a dry powder or in a slurry form, can be burned in a boiler designed for oil or gas without major boiler modifications. The tube spacing in oil and gas boilers is closer than that in coal-fired boilers. These ultrafine grind coal and lignite fuels, which are potentially less expensive than oil and gas, allow for a relatively inexpensive retrofit for using coal in existing power plants and boilers.

In Australia, tests were conducted with the use of ultrafine grind lignite in a diesel engine (6). The grinding energy required for ultrafine grinding was established at approximately 500 kWh/ton for median particle sizes of about 7 microns.

The authors of this paper have conducted research on the ultrafine grinding of Mississippi lignite in a fluid-energy mill (7-9). These results have shown that lignite, even with its high moisture content, can be ground to a product with a mean diameter, based on population, of 2 to 5 microns and a mean diameter, based on volume, of 7 to 25 microns. In-the-mill drying tests were also conducted, with the results showing that some permanent drying of the lignite can be accomplished.

It has been shown (10, 11) that if lignite is heated to a temperature of about 310°C for 15 minutes, the inherent moisture reabsorbed will be only about 10 percent as compared with the 35 percent for the original sample. Also, only a small additional drying effect results at a temperature greater than 310°C. Simultaneous ultrafine grinding and permanent drying with steam at moderately high (>240°C) temperature has also been studied (8). The level of inherent moisture was reduced from 40.9 percent to 14.1 percent.

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[1]Numbers in parentheses refer to references.

Keller and Simmons (12) have reported that for coals with good washability (ash distribution with specific gravity) characteristics, good cleaning of -325 mesh coal in a true heavy liquid medium of Freon-113 can be accomplished in a two inch hydrocyclone with an 85 psi pressure drop. Washability studies of gulf coast lignites indicate that some gulf region lignites do have reasonably good washability characteristics while some exhibit almost no ash content variation with specific gravity (13). Paulson and Baria (14) have done extensive work on ion-exchange processes which result in a reduction in the sodium content of the ash. Sodium is apparently the predominant ash component contributing to tube fouling problems in boilers.

In the work reported here, a Martin Lake, Texas lignite was ground in a fluid energy mill to an ultrafine size. Grinding with high temperature steam accomplished a simultaneous drying while grinding. Density distribution tests on the products showed that a significant portion of the ash is contained in particles which comprise ten percent of the total mass. The discussion of the tests and results are presented in the following sections.

EXPERIMENTAL PROCEDURES

The ultrafine grinding of the samples of the lignite was performed in the development facility of Ergon-Micro Energy which is located at Vicksburg, Mississippi. In general, the facility consists of a fluid-energy mill with a design capacity of 2,000 lbm/hr of solid material. The grinding medium may be either compressed air or steam. The air supply is capable of furnishing up to approximately 4,000 lbm/hr of air at approximately 100 psig. The steam supply is capable of furnishing up to approximately 4,000 lbm/hr of superheated steam at 100 psig and 750 degrees Fahrenheit. The pulverization unit is based on the multiple jet cylindrical body design concept present in the Taylor patent (15).

In order to evaluate the energy consumption requirements for the ultrafine grinding with simultaneous drying of low rank coals, it was necessary to instrument the facility to obtain energy and mass balances on the fluid and the solids streams. The instrumentation on the fluid stream consisted of pressure and temperature measurements upstream of the orifice in the supply line, pressure drop across the orifice, pressure and temperature at the nozzle ring of the pulverizer, pressure and temperature in the pulverizer, and pressure and temperature on the exhaust stream. The instrumentation required for the solids stream included a mass flow rate based on a weight belt device, the temperature of the feed, and the temperature of the product. Commercial instrumentation was available on the system and this was augmented by pressure transducers and thermocouples which were compatible with a computer based data acquisition system which allowed experimental data to be recorded throughout the experimental period.

The energy balance required that the heat losses from the system be determined. This was accomplished by running only the fluid stream through the system at various temperatures and calculating the appropriate heat loss from an energy balance on the system. Then a heat loss correlation was developed which was used in the data reduction program to ascertain the process energy requirements.

The speed of the belt feeder to the pulverizer was used as a feed rate transducer and was calibrated prior to the conduct of the tests. The feed passed through a gate and was of constant cross section on the feeder belt for the speed ranges used in this test. Temperature measurements of the feed material were also taken during the tests.

Prior to the data runs all of the transducers were calibrated. The pressure transducers were calibrated against standard devices over their appropriate ranges. The thermocouples were calibrated against known reference temperatures.

Prior to each test the system was preheated by allowing fluid only to flow through the mill. Then the low rank coal feed rate was adjusted to the desired value. Monitoring of the various temperatures and pressures allowed determination that a steady state condition had been reached in the system, after which data were recorded for use in the energy consumption determinations as well as for use in describing the conditions of operation for the process.

The data acquisition system used was based on an Apple III microcomputer with a Micro-Mac data acquisition module which allowed 12 channels of input data. A schematic diagram of the pilot plant and the location of the various instruments is shown in Figure 1.

The runs were continued for approximately 30 minutes after steady state had been reached. The conditions and runs which were conducted during the experiment are presented in Table 1.
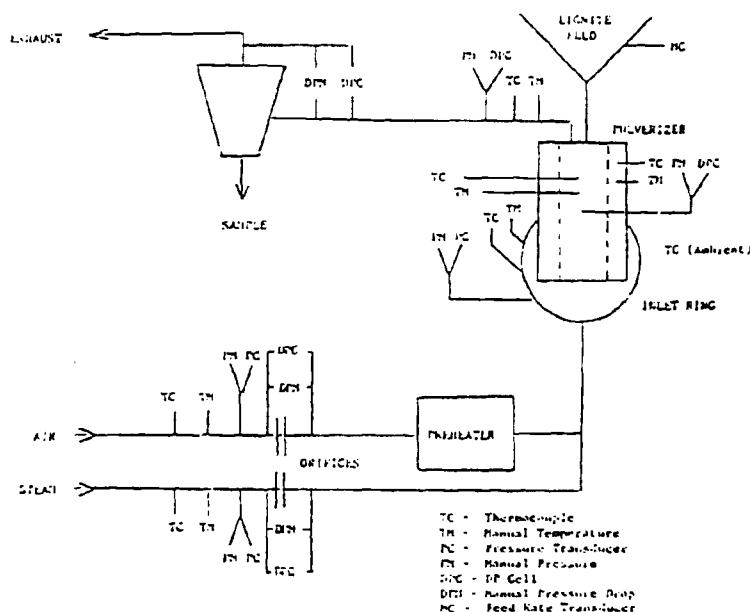
Figure 1.   Schematic Diagram of the Instrumentation on the Fluid Energy Mill

TABLE 1.   Conditions of Operation for the Experimental Studies

Medium	Temperature	Number of Runs
Air	116	1
Steam	225	1
Steam	310	3
Steam	350	2
Steam	400	3
Steam	486	1

The temperature in the list above represents the mixture temperature of the fluid and low rank coal in the pulverizer vessel.

Feedstock samples and product samples of the micropulverized lignite were taken during each run. These were collected and sealed in air tight cans.

ANALYSIS PROCEDURES

The samples from these tests on the Martin Lake, Texas lignite were analyzed to determine the effects of grinding and drying on the products. The tests conducted were volatile matter, ash, total moisture, equilibrium moisture, heating value, and density. Porosity tests, particle size distribution, ultimate analyses, and electron microscope photographs were also conducted by the University of North Dakota Energy Research Center (UNDERC), Ergon, Inc., Commercial Testing and Engineering Co., and Mississippi State University respectively. Density distribution, aerodynamic particle size classification, and angle of repose tests were also conducted. The volatile matter, ash, total moisture, and heating value tests were conducted according to ASTM procedures D-3175, D-3174, D-3173, and D-2015 respectively.

The equilibrium moisture value for a coal usually represents the in-place coal seam moisture value. This moisture value is obtainable in the laboratory by placing a coal sample in a 97% relative humidity environment for a specified period of time. The laboratory results agree well with the actual bed moisture for higher rank coals; however, the equilibrium moisture results produced in the laboratory for the lignite investigated are low compared to the actual bed moisture. This problem is common with lower grade coals according to Luppens (16). Even though the equilibrium moisture prediction of the actual bed moisture is low for this lignite, it is a reproducible property which is used as a basis of comparison.

Two techniques were used to determine the equilibrium moisture content of the coal. Both of these techniques are given in ASTM standard D-1412. The first technique consisted of mixing an approximate 1/3 solid-to-liquid ratio coal-water slurry and agitating it for thirty minutes. The slurry was then placed in a water bath maintained at 30°C for three hours. The excess water was then filtered from the mixture and the solids were placed in a humidity chamber according to ASTM standard D-1412. The second technique consisted of taking samples directly from the airtight containers and then placing them into the humidity chamber according to the alternate method given in ASTM standard D-1412. This technique was used so that the equilibrium moisture value could be approached from above and below thereby increasing the certainty in a final equilibrium moisture value.

Density tests were performed by measuring the liquid displacement of a weighed lignite sample. The density measured by this method represents the mass of the material per unit volume of ethanol displaced when the material is submerged in ethanol. This is an apparent density of the individual particles because of the probability of gas pockets in the highly porous material. Lignite samples were completely dried in an oven and weighed before placement into a flask of known volume. Ethanol was then dispensed into the flask from a burrette until a known calibrated volume was obtained and all of the sample was submerged. Ethanol was used because the micropulverized lignite floated in water. The density measurements reported herein are based on six samples at a given run condition. This density measurement is defined as an apparent density uncorrected for mineral matter.

Porosity tests were conducted at the UND Energy Research Center using a heats of immersion determining calorimeter. The wetting liquid used was methanol. When an outgassed sample is immersed in a liquid which wets its surface, a rise in temperature of the system results. The increase is proportional to the product of the total surface area wetted and the heat released on wetting a unit area of the absorbent (17, 18).

The particle size distribution (based on population and volume) was determined at Ergon, Inc. with a Coulter Counter model Ta II particle analyzer. In addition to the particle size tests, mineral ash analyses and ultimate analyses were conducted by the Commercial Testing and Engineering Co. Electron microscope photographs were taken of samples at each test condition. These samples were mixtures of each of the runs at a given condition. Photographs were taken as magnifications of 1,000 and 10,000.

Samples of each of the runs were used to determine the density distribution for the product of the ultrafine grinding process. Because the particles are small, various surface forces become significant in comparison with the gravity or body force on the particle. Consequently, a centrifuge was used to amplify the bouyant and acceleration body forces in order to facilitate separation according to specific gravity ranges. The sample was placed in a commercially prepared organic liquid of appropriate specific gravity in a centrifuge bottle. Starting with the lowest specific gravity fluid desired, the sink and float fractions were recovered. Then the sink fraction was successively separated in progressively higher specific gravity fluids into further sink and float fractions. The fluids used in this test had specific gravities of 1.2, 1.3, 1.4, and 1.6. Thus, there were five density ranges for the particles to report to, viz. 1.2 float, 1.3 float, 1.4 float, 1.6 float, and 1.6 sink. The resultant samples were then analyzed for particle size distribution and for ash content.

A particle size separation was also conducted by an air classification to yield an aerodynamic particle size distribution. A simple laboratory scale apparatus was fabricated to allow vertical airflow at different rates to carry over particles which were sufficiently small to be carried out of a six inch cylindrical section. The larger particles would settle back through a conical section into the sample chamber. Three different size ranges were obtained for further analysis in this apparatus. Ash analyses were run on these particles to ascertain if there was an ash rich fraction in one of the size ranges.

Also measured in the analysis of the products was the angle of repose. The angle of repose is defined as the angle of the surface of a pile of material which will result when the material is piled and allowed to stand in its natural state. Two techniques were used to establish this angle. The first was to gently add material to the center of a pile and allow the material to accumulate at a surface angle which was then measured. The second method was to establish a pile of material and then remove some material from the toe or bottom edges of the pile until the sloping surface began to move as a whole. Also, the tests were repeated after allowing the material to be exposed to the atmosphere for 30 days.

PRODUCT ANALYSIS RESULTS OF COMBINED MICROPULVERIZATION AND DRYING TESTS

The equilibrium moisture, proximate and ultimate analyses, and heating value of the feed coal are given in Table 2. The results of the analyses of the products of the steam and air pulverization tests are shown in Figures 2-9. The feedstock analysis results are also shown on each figure. The mean values and the data scatter bands are given on each

TABLE 2. Martin Lake Texas Lignite Feedstock Analyses

Equilibrium Moisture (%)   25.24

Proximate Analysis (%)	As Received	Dry
Moisture	30.20	--
Ash	10.29	14.74
Volatile Matter	30.25	43.36
Fixed Carbon	29.25	41.90
Total	100.00	100.00

Ultimate Analysis (%)		
Moisture	30.20	--
Carbon	44.51	63.73
Hydrogen	3.11	4.45
Nitrogen	0.72	1.03
Chlorine	0.03	0.04
Sulfur	1.03	1.47
Ash	10.29	14.74
Oxygen	10.11	14.49
Total	100.00	100.00
Gross Calorific Value (BTU/LB)	7279	10429

figure. These bands represent ± 2 standard deviations of the data. In addition, electron microscope photographs of the ground products are shown in Figure 10.

The scatter in the data around the mean value as shown by the uncertainty bands is partly the result of precision error in making the specific analyses. However, the major reason for the scatter is the variation in the products from different replications of the test conditions. The 310°F and 400°F steam ground conditions were replicated three times and the 350°F steam ground condition was replicated twice. In addition, ash variation in the feedstock had a significant effect on the feedstock analyses and on the analyses of the conditions that were replicated.

The as-received moisture of the feedstock and the test products is shown in Figure 2. These samples were stored in air-tight containers during the test, and moisture analyses were run on them within a week after the completion of the test. These values should be indicative of the moisture that the products would have immediately after being ground and prior to significant exposure to air.

The equilibrium moisture versus steam pulverizing temperature is plotted in Figure 3. The initial moisture of the feedstock is about 25% and the minimum equilibrium moisture value of 17% occurs at a steam temperature of 488°F. Air pulverization at 116°F yields a final equilibrium moisture value of 20%. These equilibrium moisture values are higher than those obtained at UNDERC, but the drying temperatures were higher for those tests (19, 20). A hot-water drying process was used in some of the UNDERC tests.

The dry basis ash versus pulverizing temperature for both steam and air grinding mediums is plotted in Figure 4. The variation in the ash with pulverization temperature appears to be only the result of variation in the feedstock.

The dry basis volatile matter versus temperature is shown in Figure 5 for the steam and air grinding mediums. There is very little change in the values for either pulverization process.

The dry basis higher heating value versus temperature for both steam and air pulverization is plotted in Figure 6. The mean heating value remains essentially constant with temperature with small variations resulting from feedstock variations.

The porosity versus pulverization temperature is plotted in Figure 7. The feedstock porosity value is 109.5 square meters per gram (m^2/g). The maximum porosity value of 279.2 m^2/g occurs at a steam temperature of 488°F. The porosity of the products increases with drying temperature. This may be due to the removal of tars and waxes from the pores during the drying and grinding process (19).

The apparent particle density versus temperature is shown in Figure 8. It should be noted that there is an increase in the density value for all grinding temperatures compared to the feedstock value. The product density is between 1.44 and 1.50 g/cc.
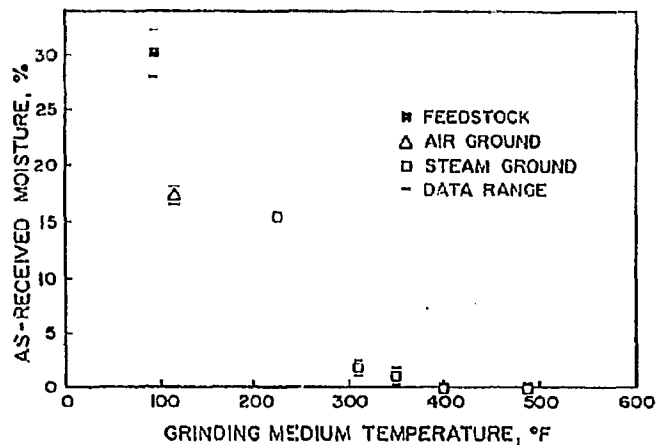
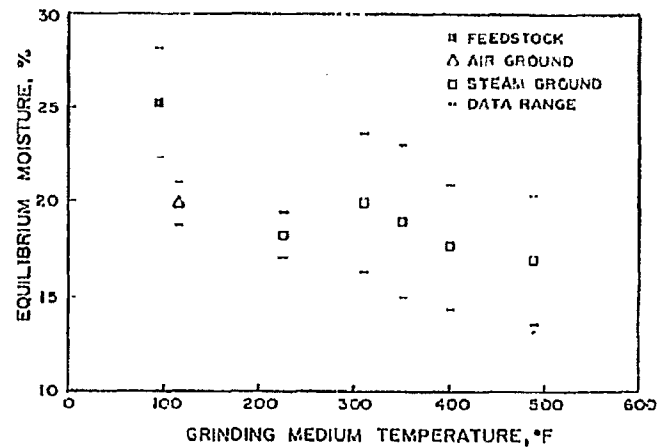Figure 2. Moisture as a Function of in the Mill Treatment Temperature

Figure 3. Equilibrium Moisture as a Function of in the Mill Treatment Temperature
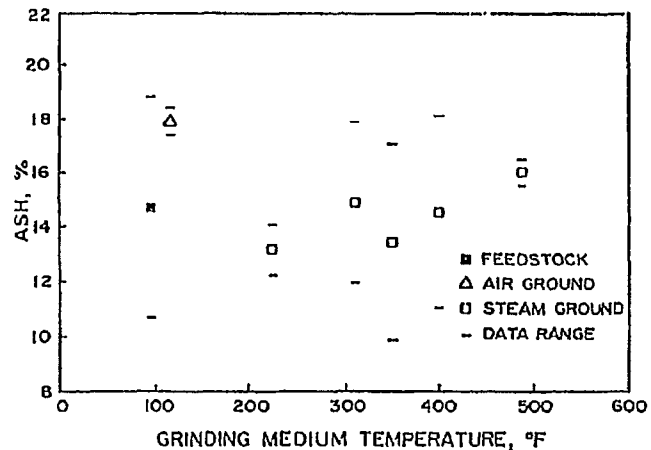

Figure 4. Ash Content as a Function of in the Mill Treatment Temperature
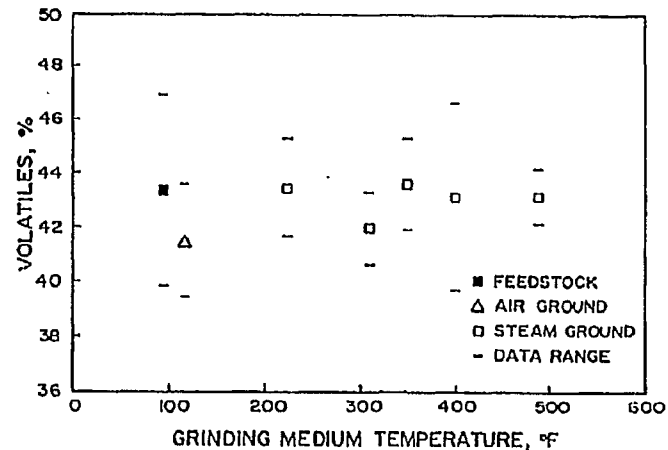
Figure 5. Volatile Content as a Function of in the Mill Treatment Temperature

Figure 6. Higher Heating Value as a Function of in the Mill Treatment Temperature

Figure 7. Porosity of the Sample as a Function of the in the Mill Treatment Temperature
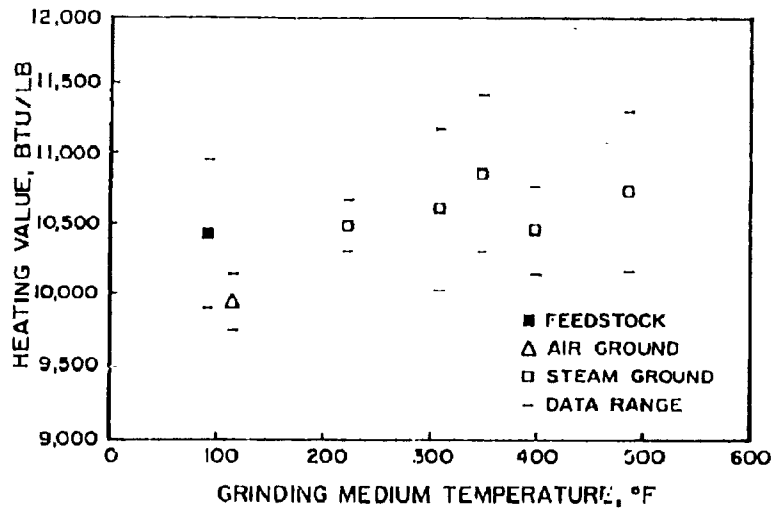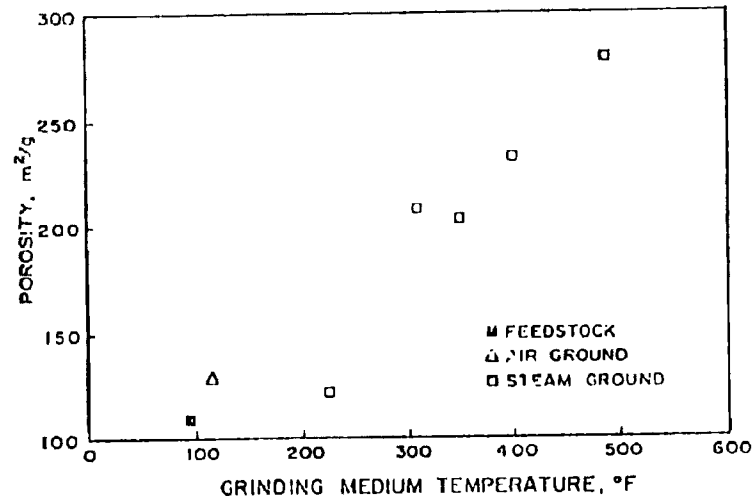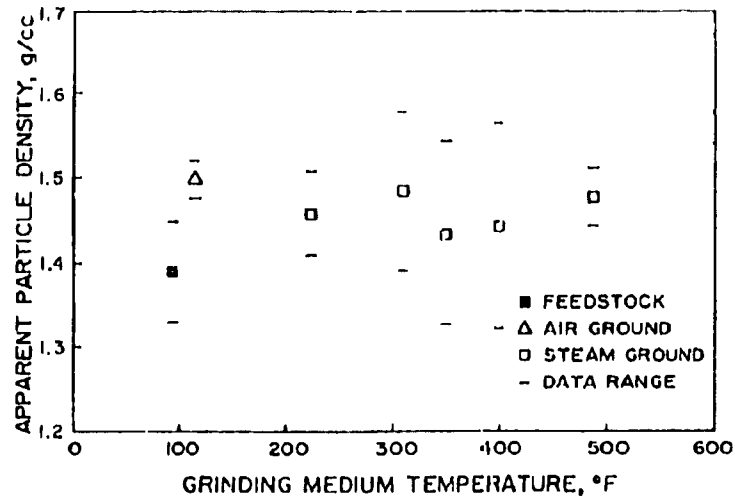

Figure 8. Apparent Particle Density as a Function of the in the Mill Treatment Temperature
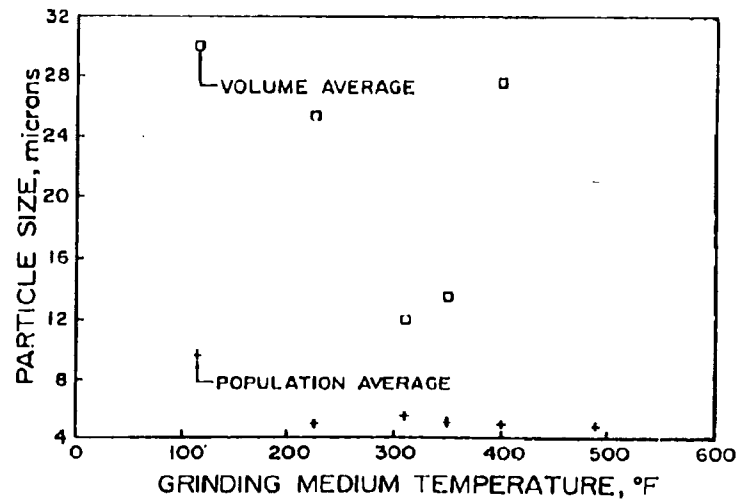

Figure 9. Median Particle Sizes as a Function of the in the Mill Treatment Temperature

The particle size distributions based on volume and population versus temperature are shown in Figure 9. The volume based particle size means are all above 20 μm except for the 310°F and 350°F steam ground products. The population based particle size means are less than 10 μm for all cases.

The ultimate analysis results for the 310°F, 400°F, and 488°F steam ground materials are given in Table 3. As expected there is little difference among the three run conditions. There is also little difference among these values and the feedstock ultimate analysis given in Table 2.

TABLE 3. Ultimate Analysis for Ultrafine Grind Products - Martin Lake, Texas Lignite

	310°F S.G.*		400°F S.G.		488°F S.G.	
	As-Received**	Dry	As-Received	Dry	As-Received	Dry
Moisture	4.45		1.43		1.11	
Carbon	59.55	62.32	61.19	62.08	62.91	63.61
Hydrogen	3.83	4.01	4.16	4.22	3.97	4.01
Nitrogen	0.99	1.04	1.12	1.14	1.09	1.10
Chlorine	0.01	0.01	0.01	0.01	0.08	0.08
Sulfur	1.23	1.29	1.23	1.25	1.17	1.18
Ash	15.14	15.84	15.42	15.64	15.23	15.40
Oxygen	14.80	15.49	15.44	15.66	14.44	14.62
Total%	100.00	100.00	100.00	100.00	100.00	100.00

* S.G. = Steam Ground
** After storage and handling.

The mineral analyses of ash results are given in Table 4. As with the ultimate analyses there is little change in ash constituents with pulverization condition.

TABLE 4. Mineral Analysis for Ash for Ultrafine Grind Test Samples - Martin Lake, Texas Lignite

Constituent (%)	Feedstock	310°F S.G.*	400°F S.G.	488°F S.G.
Al_2O_3	14.98	15.40	15.09	14.87
Fe_2O	8.15	9.05	8.84	7.60
TiO_2	1.03	1.07	1.04	0.99
MgO	3.61	3.41	3.30	3.78
K_2O	0.90	0.91	0.88	0.91
Na_2O	0.70	0.55	0.55	0.60
SO_3	14.80	9.55	12.05	12.28
P_2O_5	0.16	0.13	0.15	0.12
SrO	0.35	0.33	0.33	0.35
BaO	0.24	0.24	0.23	0.20
Mn_3O_4	0.16	0.14	0.13	0.17
Undetermined	1.14	2.88	2.01	2.41
Total	100.00	100.00	100.00	100.00
Base/Acid Ratio	0.50	0.49	0.51	0.48

* S.G. = Steam Ground

The electron microscope photographs shown in Figure 10 are at a nominal magnification of 10,000. This magnification was determined to show the most descriptive surface characteristics of the lignite particles. The steam and air pulverized materials show the general trend of some surface smoothing from the condition of the feedstock material.

Figure 10.   Electron Microscope Photographs of Various Samples
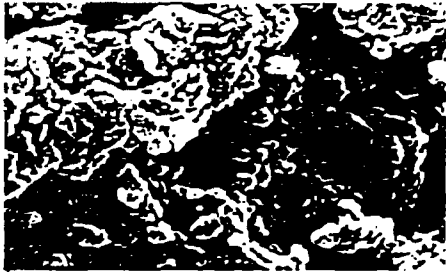
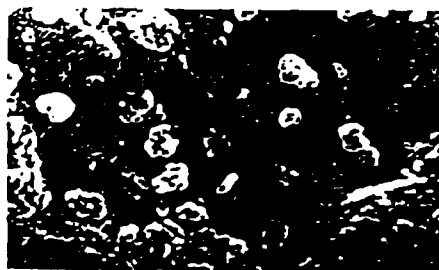Feedstock

116°F Air Ground

225°F Steam Ground

310°F Steam Ground

350°F Steam Ground

400°F Steam Ground

488°F Steam Ground

The density distribution of the Martin Lake, Texas lignite was determined for the various grinding conditions studied. Because there was very little material below the 1.3 specific gravity, the 1.2 specific gravity fluid was not used. The mass in the various specific gravity ranges was relatively consistent for the different grinding mediums and temperatures.

Analyses for ash fraction were performed on the density fractions with the significant result that the heavy fraction (>1.6 specific gravity) contained over 65 percent ash. Considering that this density fraction comprises approximately 9 percent of the total mass, then there is approximately 6 percent of the total product material as ash in separate particles in the >1.6 specific gravity range. This represents a little less than half of the total ash in the material.

A bar chart of the various specific gravity fractions is presented as Figure 11. Also, a bar chart of the ash content of the various temperature treatments is presented as Figure 12. It is very convincing that the heavy fraction (>1.6 specific gravity) contains much more ash than does the lighter fractions.
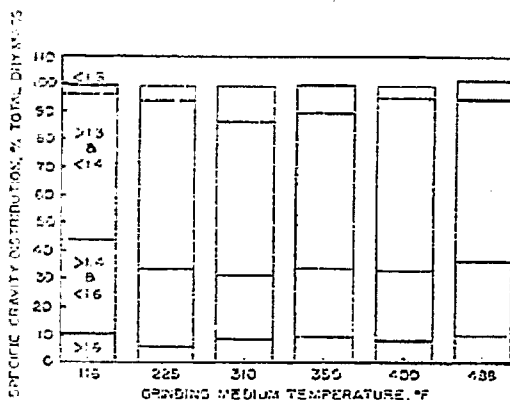

Figure 11. Specific Gravity Distribution of
Samples for Various in the Mill
Treatment Temperatures
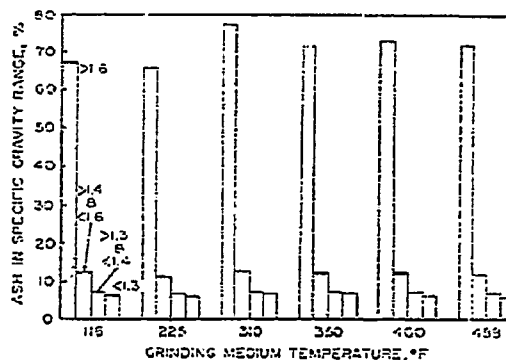
Figure 12. Ash Content of Samples for
Various in the Mill Treatment
Temperatures

The particle size distributions for the ash rich fractions are presented as Table 5. A comparison of the particle sizes for these materials with those of the total product samples (Figure 9) shows that these heavy particles are smaller than the general size of the particles in the aggregate. This suggests that the ash particles would be sufficiently small to minimize the slagging effect on boiler tubes.

TABLE 5. Particle Sizes for the Ash Rich Fractions (>1.6
Specific Gravity) of the Density Distribution Tests.

Test	Particle Size (Microns)			
	Volume Average	Standard Deviation	Population Average	Standard Deviation
116°F A.G.*	15.30	2.95	4.85	1.37
225°F S.G.**	16.87	2.57	5.60	1.39
310°F S.G.	8.91	1.90	5.24	1.45
350°F S.G.	10.18	2.13	5.20	1.46
400°F S.G.	14.03	2.25	5.29	1.36
488°F S.G.	11.60	2.24	5.16	1.45

*   A.G. = Air Ground
** G.S. = Steam Ground

Aerodynamic classification tests were conducted to determine if an ash rich particle size fraction could be obtained. The products of the grinding/drying tests were separated into three particle size fractions (fine, medium and course). Very distinct particle size splits were obtained, but there was no significant ash variation among the three fractions.

The results of the angle of repose tests are presented in Table 6. 5 th as-received material and material that was exposed to an ambient environment for three weeks were investigated.

TABLE 6. Angle of Repose Results

The angle of repose is defined as the angle of the pile with the horizontal as measured from the bottom of the pile. Four replicate tests were made for each condition and averages are reported.

| Temperature | 116°F | 225°F | 310°F | 350°F | 400°F | 488°F |
Grinding Fluid	Air	Steam	Steam	Steam	Steam	Steam
SAMPLES AS RECEIVED FROM TESTS						
Moisture (%)	17.5	15.5	2.0	1.2	0.1	0.0
Piled Slope	41.2	40.0	32.5	36.2	36.3	37.5
Fallen Slope	62.5	63.7	58.2	63.7	61.2	58.7
SAMPLES EXPOSED TO AIR FOR THREE WEEKS						
Moisture (%)	9.4	8.9	4.2	3.7	4.2	3.0
Piled Slope	29.5	28.7	35.5	35.0	36.2	37.0
Fallen Slope	57.7	53.2	56.7	54.0	57.7	53.5

As may be observed from the moisture data, the wetter, air-ground materials dried out after being exposed to air and the drier, steam-ground products gained some moisture. The loss of moisture during air exposure for the low-temperature ground products caused the angle of repose to decrease. Also, some degradation of the surface in the air exposed samples could contribute to the change in the angle of repose.

The increase in the angle of repose from the piled tests to the fallen slope tests probably relates to the method of piling the material on the cone. The fallen slope values are more indicative of the related surface and body forces on the material.

ENERGY REQUIREMENTS FOR ULTRAFINE GRINDING AND DRYING OF LIGNITE IN A FLUID ENERGY MILL

During the grinding/drying tests on the Martin Lake, Texas lignite, data were collected in order to determine the energy consumption requirements for the ultrafine grinding/drying. As discussed above, the instrumentation system consisted of thermocouples, total pressure transducers, and a lignite feed rate transducer connected to a microcomputer through an analog-to-digital interface board. Data were recorded on a disk every two minutes during each test run.

The flow rate of the air or steam grinding fluid, $\dot{m}_F$, was determined from an orifice plate pressure drop. The orifice constant was determined from standard curves for concentric orifices in pipes (21). The density for the air runs was calculated from the ideal gas relationship using the measured temperature and pressure at the orifice plate. The density for the steam runs was calculated from a correlation for superheated steam specific volume given by Keenan and Keyes (22) which is also a function of temperature and pressure.

The lignite feed rate was determined from the calibration curve for the lignite mass flow rate versus feed belt speed. The feed passed through a gate to maintain it at a constant cross section on the feed belt.

Samples of the feedstock and products for each run were analyzed for moisture in order to determine the amount of water driven out of the lignite. The flow rates of the product lignite (solid + final water in lignite) and the moisture removed from the lignite were determined by the following relationships:

$$\dot{m}_{LP} = \dot{m}_{LF} \left[\frac{1 - n_f}{1 - n_p} \right] \tag{1}$$

and

$$\dot{m}_{WV} = \dot{m}_{LF} \left[\frac{n_f - n_p}{1 - n_p} \right] \tag{2}$$

where

$\dot{m}_{LF}$ = Lignite feed flow rate
$\dot{m}_{LP}$ = Lignite product flow rate
$\dot{m}_{WV}$ = Flow rate of water vapor driven out of the lignite
η_f = Moisture fraction of feed
η_p = Moisture fraction of product.

An energy balance on the fluid-energy mill is illustrated in Figure 13. The energy lost by the grinding fluid minus the heat loss from the system is equal to the energy gained by the lignite and the water vapor driven out of the lignite:

$$\dot{m}_F(h_{F_1} - h_{F_2}) - Q_{HL} = \dot{m}_{LP}(h_{L_2} - h_{L_1}) + \dot{m}_{WV}(h_{WV_2} - h_{WV_1}) \qquad (3)$$

$\dot{m}_F h_{F2} + \dot{m}_{LP} h_{L2} + \dot{m}_{WV} h_{WV2}$
(Product)

$\dot{m}_{LP} h_{L1}$
(Lignite)

$\dot{m}_{WV} h_{WV1}$

Q_{HL}

$\dot{m}_F h_{F1}$
(Steam or Air)

$\dot{m}_F$ = Grinding fluid flow rate

$\dot{m}_{LP}$ = Lignite product flow rate

$\dot{m}_{WV}$ = Flow rate of water vapor driven out of the lignite

h_F = Grinding fluid enthalpy

h_L = Lignite product enthalpy

h_{WV} = Enthalpy of water driven out of the lignite
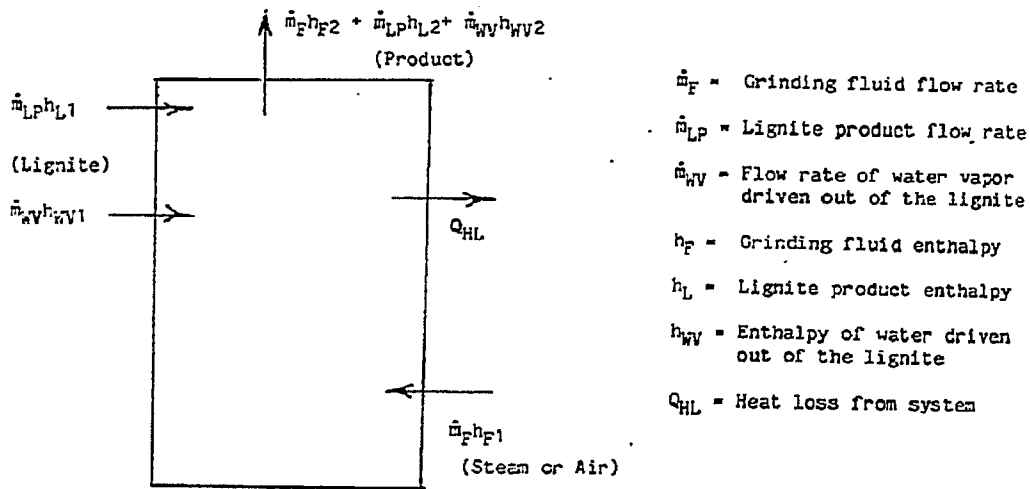
Q_{HL} = Heat loss from system

Figure 13. Energy Balance of Fluid-Energy Mill

The enthalpy differences were determined from appropriate tables and thermodynamic property correlations (22, 23).

During the calibration phase of the test program, a heat loss correlation was developed which was a function of the difference between the mixture temperature in the pulverizer and ambient temperature. Data for this correlation were obtained by running only steam through the system at various temperatures and calculating the steam enthalpy change.

With the above expressions it was possible to determine the energy required to dry and grind the lignite in the fluid-energy mill. It should be noted that there is no way to separate the grinding energy from the drying energy since the grinding process creates heat which goes into heating the lignite and the water vapor.

The left hand side of equation 3 is the power required to dry and grind the lignite as determined from measurements on the grinding fluid. This quantity served as the primary calculation of the power requirements. The nominal conditions for the test runs are given in Table 7.

TABLE 7. Nominal Conditions for Martin Lake Texas Lignite Tests

Grinding Fluid	Mill Temperature (°F)	Inlet Fluid Temperature (°F)	Outlet Fluid Temperature (°F)	Fluid Flow Rate (lbm/hr)	Feedstock Flow Rate (lbm/hr)
Air	116	320	116	1600	1600
Steam	225	689	225	1525	1675
Steam	310	688	302	1925	740
Steam	350	700	329	2100	738
Steam	400	703	350	2275	737
Steam	488	730	438	*	737

* Flow orifice pressure drop transducer was overranged. Flow rate was greater than 2400 lbm/hr.

The grinding/drying energy value for each run is given in Table 8. These results are expressed in kWh/Ton and are the ratio of the power requirements divided by the lignite feed flow rate. Also given in Table 8 are the energy-per-ton values determined from the lignite feed and product moistures and temperatures (the right hand side of equation 3). These second calculations served as a balance check on the grinding/drying energy measurements.

At the higher temperatures the comparison is good between the energy per ton calculated from the steam and from the lignite data. At the 225°F steam ground condition, the energy determined from the lignite is somewhat higher. As seen in Table 7, the exit temperature in this case was 225°F and the exit pressure was close to atmospheric. Since this condition is near saturation, there was probably liquid water on some of the product when it left the mill. The model used to calculate the change in enthalpy of the water removed from the lignite assumed that a superheated steam existed at the exit. Therefore, for the case close to saturation, the model would overpredict the enthalpy increase. For the 116°F air ground condition, the exit was below saturation and no calculations could be made from the lignite data.

The grinding/drying energy versus the medium temperature is presented in Figure 14. The energy requirements are approximately linear with respect to grinding temperature over the range tested.
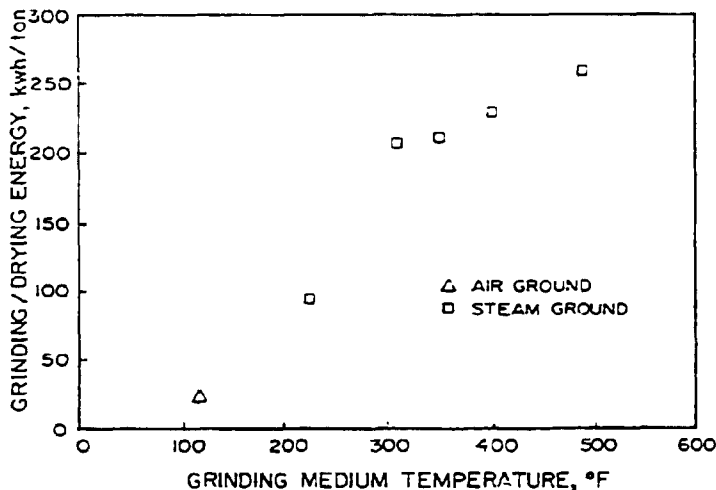

Figure 14. Process Energy Required form Simultaneous Grinding/Drying as a Function of the in the Mill Treatment Temperature

It should be stressed that these energy data only account for the process inside the fluid-energy mill. They do not include the compressor/boiler efficiency, the transmission losses, and the other inefficiencies that are part of the entire grinding circuit.

The exit temperatures given in Table 7 show that at the higher temperature grinding conditions, there is still significant energy in the exit steam. For an economical process, this hot exhaust would have to be used to preheat the boiler water in order to reduce the overall system losses.

TABLE 8. Energy Requirements to Grind and Dry Martin Lake, Texas Lignite in a Fluid-Energy Mill.

Test Condition	Energy/Ton Lignite (Calculated from Fluid) kWh/Ton	Energy/Ton Lignite (Calculated from Lignite) kWh/Ton
116°F A.G.*	24	--
225°F S.G.**	95	129
310°F S.G.		
Run 1	215	219
Run 2	212	216
Run 3	198	217
350°F S.G.		
Run 1	196	228
Run 2	227	226
400°F S.G.		
Run 1	***	238
Run 2	232	237
Run 3	226	237
488°F S.G.	***	259

*   Air Ground
** Steam Ground
*** Fluid flow orifice pressure drop transducer was overranged.

CONCLUSIONS

Combined ultrafine grinding and drying of a Martin Lake, Texas lignite yields a product with a permanent reduction in moisture and a mean particle size (based on population) of less than ten microns in diameter. Both of these results are essential and significant in the search for a retrofit fuel for gas and oil boilers.

Property measurements of the ultrafinely ground products show that the ash and volatile contents as well as the heating value of the lignite are not affected by the grinding/drying process up to temperatures of 500°F. The equilibrium moisture of the product decreases with steam and air grinding over the value for the feedstock. The density of the products is greater than that of the feedstock and the porosity of the product increases with grinding medium temperature. Ultimate analysis of the products and mineral analysis of the ash show little change with grinding temperature.

The product was separated according to particle density and the >1.6 specific gravity particles contained approximately 70 percent ash or approximately 50% of the ash content of the feedstock. Aerodynamic classification was not successful in obtaining an ash rich fraction.

The energy required to grind and dry the lignite was determined from temperature, pressure, and flow rate measurements made during the tests. The energy usage ranged from 24 kWh/Ton of lignite for air grinding to 260 kWh/Ton for 488°F steam grinding.

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GASIFICATION ISLAND SYSTEMS

by

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ABSTRACT

The technical feasibility of using integrated gasification combined-cycle
(IGCC) technology for the production of coal-based electricity is currently
being demonstrated at commercial scale (100 MWe) in the Cool Water Coal Gasi-
fication Program. This successful project is demonstrating that IGCC systems
can meet stringent environmental emission standards and operate with high
availability. The Department of Energy's Morgantown Energy Technology Center
(METC) has explored the performance and economic incentives for advanced gasi-
fication/gas turbine power production systems which build on the technology
base generated by first-generation IGCC systems such as the Cool Water project.
The term "gasification island" is used to describe an electric plant using this
advanced gasification/gas turbine power production technology. The gasifica-
tion island itself encompasses all of the processing steps required to convert
coal into a clean fuel gas. In this paper, the term "gasification island" is
analogous to the term "power island" used to describe the power generation
equipment in conventional coal-fired steam plants.

Gasification island systems use coal gasifiers together with hot gas cleanup
processes, either internal or external to the gasifier, to produce a clean,
low-Btu fuel gas. The solids handling, gasification, and gas cleanup processes
are factory-fabricated modules designed to match the requirements of specific
gas turbine systems. By using air-blown gasification systems (which consume
less auxiliary power than oxygen-blown systems), hot gas cleanup, innovative
tail gas treatment processing schemes, and high-efficiency gas turbine cycles,
the thermal efficiency of gasification island systems is projected to be higher
than first-generation IGCC systems. Because of standardized designs and simple
process configurations, the engineering and equipment costs for gasification
island systems are also projected to be significantly lower than first-
generation systems. This is particularly true for new capacity additions or
life-extension modifications in the 100 MWe size range. This paper describes
the results of the gasification island concept systems studies for low-rank
coal and reviews the systems implications of the Department of Energy-sponsored
research programs being conducted in support of the gasification island
concept.

INTRODUCTION

While currently there is excess electric power generation capacity in many parts of the country, the effective reserve margin is declining. A load growth projection from Edison Electric Institute (1) assumes a conservative 2.2 percent annual growth rate in peak load and shows that, on a national basis, shortfalls in generating capacity are possible beginning in 1992. Projections by other groups show similar trends. To avoid major shortfalls, significant amounts of new generating capacity will need to come on line during the mid to late 1990's. Utilities are beginning to assess the available technological options for providing this new capacity. A changing financial and regulatory environment may lead utilities to select technologies for the 1990's which are very different from those selected during the 1970's, the era of the last major surge in plant construction. In the 1990's, the most attractive candidates for new capacity additions may be those technologies which are available in cost-competitive small module sizes, are fuel flexible, have low capital cost and short construction time, are highly efficient, and are capable of meeting stringent environmental regulations.

Several power generation technologies are available for utilities considering building new plants during the 1990's, including first-generation IGCC, a technology which offers several significant improvements over pulverized coal plants with flue gas desulfurization (PC/FGD). As is currently being demonstrated by the Cool Water Gasification Project, it is possible to meet very stringent environmental emission standards using this technology. Because IGCC is based on gas turbine systems, this technology lends itself to the sequential introduction of power generating components -- first, natural gas-fired gas turbines, followed by a steam turbine bottoming cycle, and finally, by a coal gasification system. Heat rates for mature versions of first-generation IGCC systems using gas turbines with 2,200°F firing temperatures are in the 8,500 to 9,200 Btu/kWhr range (37 to 40 percent efficiency). The total capital requirement for mature systems is projected to be similar to PC/FGD plants; the cost of electricity (COE) is projected to be approximately 10 percent lower. This paper describes a gasification island system, a system with the potential for further reductions in the COE, particularly for small-size plants.

GASIFICATION ISLAND SYSTEM DESCRIPTION

The term "gasification island" describes a simplified approach for converting coal into a clean fuel gas for subsequent combustion by a gas turbine system. This approach uses air-blown, pressurized gasifiers and hot gas cleanup processes to produce a fuel gas containing very low levels of particulate and sulfur contaminants. The specific processes utilized in gasification island flowsheets are selected from a system point of view. The "ideal" flowsheet would incorporate a gasifier whose offgas temperature would exactly match the design requirements of downstream cleanup processes. Likewise, pressure requirements would also be matched. Both fluidized-bed and fixed-bed gasifiers are candidate fuel producers for gasification island concept systems. The temperature of the offgas from these gasifiers is generally low enough to minimize or eliminate the need for gas cooling prior to treatment in hot gas cleanup processes. For both options, the gasification and cleanup system is sized to match the requirements of specific U.S. manufactured gas turbine systems in a nominal 50 to 100 MWe size range.

Figure 1 compares the projected performance of four first-generation IGCC systems (320 to 390 MWe size range) with three gasification island systems (110 to 224 MWe size range). (The basis for this figure is discussed later in this paper.) The simplified process configurations used in the gasification island concept result in systems with efficiencies which tend to be higher, and total capital requirements which are 30 to 40 percent lower, than first-generation systems. These trends result in a COE for gasification island systems which is lower than that of other coal-based technologies, including first-generation IGCC systems. The gasification island COE is also competitive with natural gas-fired gas turbine systems, when all systems operate at the same capacity factor.

While the target performance estimates for gasification island systems are extremely promising, the eventual commercialization of this concept is contingent on the successful resolution of several technical issues:

- Compatibility of the fuel gas with the requirement of the gas turbine. Fuel gas contaminants could lead to gas turbine corrosion, erosion, or deposition. Maintaining stable combustion with low-Btu gas is another concern.

- Ability to meet or exceed existing environmental regulations. The ability of the hot gas cleanup process or integrated gas turbine components (e.g., staged combustors) to meet environmental regulations for sulfur dioxide (SO_2), nitrogen oxide (NO_x), and particulate emissions needs to be demonstrated.

- System integration. The individual system components (i.e., fuel production, fuel cleanup, contaminant disposal, and gas turbine) need to be integrated into an overall system which is reliable and exhibits good load-following characteristics.

The Department of Energy is systematically addressing these issues in research programs being conducted both under external contracts and at METC. Results from several of these programs are summarized in the following section.

HOT GAS DESULFURIZATION

Processes to effectively remove sulfur contaminants from the fuel gas at high-temperature, high-pressure conditions are a keystone of gasification island systems. METC has conducted research on hot gas desulfurization processes for a number of years. Iron oxide processes were pioneered at METC during the late 1970's. More recently, METC research has centered on zinc ferrite, a metal oxide sorbent which absorbs the reduced sulfur compounds in the fuel gas. Zinc ferrite, an iron and zinc oxide compound with the chemical formula $ZnFe_2O_4$, has the following advantages:

- High sulfur removal efficiencies. In addition to hydrogen sulfide (H_2S), zinc ferrite removes sulfur compounds such as carbonyl sulfide (COS) and carbon disulfide (CS_2) from the fuel gas to levels generally less than 10 parts per million (2).

- Regenerability. Zinc ferrite regenerates easily using a mixture of air and process steam and shows little performance degradation over multiple cycles. The offgas from the regeneration, consisting of SO_2, steam, and nitrogen is further treated in a sulfur recovery or disposal process.

- High sulfur-bearing capacity. The theoretical loading of sulfur for complete utilization of the sorbent is 35 percent of the initial sorbent weight. In laboratory tests, sulfur loadings as high as 25 percent have been achieved.

- Relatively low sorbent cost. Sorbent costs for large-scale systems are projected to be $2 to $4 per pound.

The optimum (in terms of sulfur capture efficiency and sorbent integrity) operating temperature range for the absorption step of the zinc ferrite process is 1,000° to 1,200°F. The sorbent catalyzes the exothermic water gas shift reaction which results in an approximately 100°F temperature increase in the fuel gas. Thus, the fuel gas entering the zinc ferrite unit needs to be in the 900° to 1,100°F temperature range. This temperature range has several advantages:

- It is consistent with the equipment constraints imposed by downstream equipment, i.e., the fuel gas control valve on the gas turbine.

- Alkali species in the fuel gas are in the condensed phase over this temperature range, enabling the use of a particulate removal process for the simultaneous removal of alkali and particulates.

- For a number of gasifier/coal type combinations, this temperature range minimizes the amount of heat exchange equipment required to cool the fuel gas between the gasifier exit and the zinc ferrite process inlet. Compared to entrained-flow gasifiers, the offgas from fluidized-bed gasifiers (1,600° to 1,850°F) requires a modest amount of temperature reduction either in radiant/convective heat exchange or a direct quench, to reach the zinc ferrite operating temperature.

    For bituminous coal, the offgas temperature (1,000°F) from a fixed-bed gasifier is ideally matched to the requirements of the zinc ferrite process. In this temperature range, the tar and oils pass through the zinc ferrite unit in the vapor state and are essentially incinerated along with the fuel gas in the gas turbine combustor. Most data from fixed-bed gasifiers operating on low-rank coal indicate that the offgas temperature is lower than the requirements of the zinc ferrite unit. There are several possible options for obtaining a higher temperature fuel gas. The gas could be heated by introducing a small amount of air into the fuel gas in a precombustor. Alternatively, either the gasifier operation or hardware could be modified with the aim of producing a higher temperature fuel gas. As a specific example of the latter, General Electric (GE) Company discussed an air-blown, slagging composite gasifier in a recent report (3). This gasifier contains two stages: a high-temperature, entrained-flow first stage and a slagging fixed-bed (supported on a static grate) second stage. The composite gasifier's projected offgas temperature is significantly higher than that from fixed-bed gasifiers using more conventional designs.

At METC, zinc ferrite has been tested in a pressurized bench-scale unit operating on slipstreams from both fluidized-bed and fixed-bed gasifiers. Laboratory-scale tests in a 2-inch reactor have also been conducted. Test results showed sulfur removal efficiencies in excess of 99.9 percent (2) for the gas-phase sulfur in the fuel gas stream. Sorbent regenerability in the METC tests was demonstrated over a limited number of absorption and regeneration cycles, with no significant change in the sorbent's sulfur removal effectiveness following regeneration.

A number of technical issues concerning zinc ferrite are currently being studied either under Department of Energy sponsored contracts or at METC. The following issues significantly affect the performance and economics projected for commercial versions of this process.

Long-term sorbent durability. Amax Extractive R&D, Inc. is studying long-term sorbent durability, and the trade-offs between zinc ferrite sorbent porosity and crush strength. They tested a total of 13 different zinc ferrite sorbent formulations for up to 92 absorption/regeneration cycles (2). The most promising Amax sorbent, tested for 43 absorption/regeneration cycles (over 600 total hours), demonstrated twice the sulfur-bearing capacity and physical crush strength as the original METC sorbent formulation supplied by United Catalyst. Further improvements in sorbent formulation are anticipated.

Reactor configuration for large-scale system. Bench-scale zinc ferrite tests at METC were conducted in a 6-inch diameter packed-bed reactor containing 3/16-inch diameter extruded zinc ferrite pellets. The selection of the packed-bed reactor configuration was primarily dictated by experimental expediency. If this configuration were selected for a commercial system, it would require two or more packed-bed reactors installed in parallel, and operated with alternating absorption and regeneration cycles (Figure 2). Other reactor designs are possible. For example, a moving-bed reactor configuration is an option, but this may require further improvement in the sorbent crush strength. A moving-bed zinc ferrite reactor could conceivably also function as a particulate removal device, i.e., a type of granular-bed filter. The chemical reactions occurring during regeneration are highly exothermic; therefore, a method to control the reactor temperature during regeneration needs to be incorporated into any reactor design. In the METC tests, the temperature was controlled by mixing a diluent (steam) with the regeneration air, but other approaches are possible.

Ammonia cracking. Ammonia (NH_3) in the fuel gas stream is a NO_x precursor and, therefore, a species of environmental interest. In preliminary tests at METC, zinc ferrite sorbents doped with small amounts of other metal oxides (copper and nickel) have reduced the NH_3 concentration in a simulated fuel gas by as much as 50 percent. The ammonia is cracked to elemental nitrogen and other products.

Scale-up. Scale-up issues are being addressed in tests using a large-scale (4.5-ft diameter by 16-ft height), packed-bed zinc ferrite reactor. This unit is currently being demonstrated by KRW Energy Systems, Inc. (KRW), at their Waltz Mill, Pennsylvania, site in full-stream tests on a fluidized-bed gasifier process demonstration unit (PDU).

COMBUSTION CHARACTERISTICS OF HOT, LOW-BTU FUEL GAS

Turbine compatibility issues were addressed in a recent experimental program (4) conducted at the GE Corporate Research and Development Center to study the integration of a fixed-bed gasifier, a hot cyclone for particulate removal, and a gas turbine simulator. In the tests, a nominal 1 ton per hour, pressurized fixed-bed gasifier, was operated on Illinois No. 6 coal. Particulates were removed from the hot (1,000°F), low-Btu fuel gas exiting the gasifier was in a single cyclone stage. The fuel gas was then ducted in heat-traced lines to a gas turbine simulator where it was burned in a commercial-scale gas turbine combustor. The resultant products of combustion (POC) then passed through a first-stage air-foil cascade. A total of 280 hours of tests were conducted over a matrix of test conditions: gas turbine firing temperatures of 1,800° and 2,100°F, and cascade metal temperatures of 600° and 1,300°F. These metal temperatures are consistent, respectively, with water and air cooling of the airfoil cascade.

Several very positive results emerged from the GE test (Figure 3).

- The turbine airfoils had no measurable deposits.

- Cyclone cleanup coupled with carbon burnout in the combustor was capable of meeting both gas turbine equipment specifications and environmental regulations (New Source Performance Standards or NSPS for steam coal) for particulates.

- Alkali levels in the POC were extremely low, indicating that alkali induced corrosion should not be a problem in the turbine. Sodium and potassium levels were 20-40 parts per billion -- near the specification for residual fuels. Additionally, results from related research suggest that the alkali in coal behaves differently during the combustion process than the alkali in residual fuels, with the coal alkali exhibiting a lower corrosion potential.

- Combustor operation was stable. Stable operation was a concern because, in hot gas cleanup systems, there are fewer vessels to dampen gasifier transients than in cold cleanup configurations.

- NO_x emissions from the unoptimized combustor were 0.7 lb/MMBtu, 10-20 percent higher than the NSPS for NO_x. This value suggests that staged combustion or other combustor modifications, or the use of an ammonia cracking additive with the zinc ferrite process, may be an adequate NO_x control strategy.

The conclusion from these tests is that the ash and alkali level in fuel gas produced by a fixed-bed gasifier operating on eastern coal is compatible with existing gas turbines, following cleaning of the fuel gas in cyclones.

GASIFIER IN SITU DESULFURIZATION USING CALCIUM-BASED SORBENTS

Over the years, a number of research groups investigated the possibility of producing a low-sulfur fuel gas by adding limestone or dolomite to a coal gasifier. Sulfur species, released during the gasification process, react with the sorbent material and are removed from the gasifier with the coal ash. Chemical

equilibrium calculations indicate that sulfur removal efficiencies as high as 50 to 95 percent of the sulfur in the feed coal are possible in the temperature and pressure ranges encountered in fluidized-bed and fixed-bed gasifiers.

In 1985, KRW conducted a series of tests which showed that in situ desulfurization is a very viable concept for fluidized-bed gasifiers. During the tests, the KRW PDU gasifier was operated in an air-blown mode. Particulates were removed from the fuel gas using two stages of recycle cyclones and a sintered metal filter which operated at 1,200°F. The tests evaluated the effect on gasifier operation of a number of process variables, including sorbent type, particle size, and feed location. The most promising results from these tests were obtained when larger bed inventories were maintained in the gasifier which provided greater fines residence times and resulted in a lower elutriation rate of fines (5). Results given below are representative of those obtained during the tests (Figure 4).

- A 91 percent sulfur capture was achieved when operating with Wyoming subbituminous coal and dolomite at a calcium- (dolomitic calcium only) to-sulfur (Ca/S) molar ratio of 2.0 (6). Sulfur capture in excess of 90 percent was achieved when operating with high-sulfur eastern coal and limestone at Ca/S molar ratios of 1.5 to 2 (7).

- The heating value of the fuel gas was high -- 150 Btu/scf (5). Typical values from previous air-blown tests at lower bed levels and without in situ desulfurization were 90 Btu/scf. The reason for this increase in heating value has yet to be determined. Possible reasons are calcium-induced catalysis of the gasification reactions or more efficient gasification of recycled fines in the deep bed.

- NH_3 levels in the fuel gas were very low, a result partially associated with air-blown operation. Assuming all the NH_3 is converted to NO_x, the resulting NO_x emissions would be 0.2 lb/MMBtu, one-third the NSPS for NO_x (5). Recent test results indicate that emissions could be even lower.

- The vapor-phase alkali concentration measured downstream of the sintered metal filter was less than 20 ppb (5). The particulate loading at the same location was essentially zero. This observation implies that at 1,200°F, the alkali and particulates can be simultaneously removed from the system.

The relative inertness of the gasifier ash is a technical uncertainty of the in situ desulfurization concept. Prior to disposal, the calcium sulfide in the ash must be oxidized, either within the gasifier or in an external process, to calcium sulfate ($CaSO_4$) or other environmentally acceptable, nonleachable form. This issue is being explored in tests at the KRW PDU and in planned tests with the METC fixed-bed gasifier.

SO$_2$ RECYCLE TO GASIFIER

The method selected to treat the SO_2-containing offgas stream produced during the regeneration of the zinc ferrite sorbent has a significant impact on the economics of gasification island concept systems. Potentially, the zinc ferrite process could be used in tandem with in situ desulfurization to provide a low capital cost system capable of meeting very stringent sulfur emission

standards. In this concept, bulk sulfur removal would occur in the gasifier. The zinc ferrite unit serves as a sulfur polishing step, and removes essentially all of the remaining gas-phase sulfur in the fuel gas. The steam and SO_2-containing offgas from the zinc ferrite process is recycled to the gasifier. The sorbent added to the gasifier captures the sulfur; the sulfur exits the gasifier with the gasifier ash and spent sorbent mixture. This concept tends to improve the system efficiency because of the effective steam integration. The SO_2 recycle concept will be evaluated in upcoming test runs of the KRW PDU and METC gasifiers.

ADVANCED GAS TURBINE SYSTEMS

Gas turbines, an integral part of gasification island concept systems, have great potential for performance growth. There is a direct relationship between turbine firing temperature and efficiency. The firing temperature of state-of-the-art stationary gas turbines is in the 2,000° to 2,300°F range. The U.S., however, has an active research program to develop high-temperature turbines for military applications. In the long term, successful development of high-temperature ceramic blades and/or advanced turbine blade cooling techniques will allow firing temperatures to be 3,000°F or higher. Fossil energy technologies using stationary gas turbine systems, particularly those using aircraft derivative machines, will ultimately share in the gas turbine technology improvements resulting from defense-related research.

In addition to the potential for performance growth, gas turbine systems have a number of other inherent advantages; namely, these systems are pre-engineered, factory fabricated, reliable, and modular. The performance and economic projections for gasification island concept systems given later in this paper assume that these inherent advantages of gas turbine systems will also apply to the front portion of the system -- the gasification island.

Steam-injected gas turbine (STIG) systems which inject superheated steam into the turbine combustor and between some of the downstream stages have several features which make them an attractive option for gasification island concept systems. Steam injection is not a new concept; it has been used for NO_x control and for power augmentation. The present application, however, uses more massive quantities of steam. The steam, which is produced in a heat recovery steam generator (HRSG), increases the amount of power produced by the turbine by increasing mass flow through the expansion stages. Steam injection also results in efficiency improvements because of effective recovery of heat from the turbine exhaust gas. Gas turbines which operate at high-pressure ratios, i.e., aircraft derivative turbines, show greater efficiency gains with steam injection than industrial gas turbines which operate at lower pressure ratios.

Adding an intercooler between the low- and high-pressure compressors results in further efficiency improvements and higher power outputs for a given size core assembly. A 6,560 Btu/kWhr heat rate (52 percent efficiency), lower heating value basis, is projected for a natural gas-fired intercooled steam-injected gas turbine (ISTIG) cycle (3).

A STIG system is a near-term option requiring minimal modification of existing machine designs -- the concept is currently being tested in a cogeneration system fired with natural gas. The ISTIG is a longer term option. It is

estimated that the ISTIG would require a 4- to 5-year and $100 million engineering development program for commercialization (3).

GASIFICATION ISLAND PROCESS FLOWSHEETS

Two gasification island systems, both operating with low-rank coal, were evaluated. Both of the cases studied used air-blown, pressurized, fluidized-bed gasifiers to produce fuel gas for a gas turbine-based power generation system. Data from a recent KRW PDU test were used as a basis to develop the gasification island portion of the system. In this PDU test, Wyoming subbituminous coal was used as the coal feedstock. Dolomite was added to the gasifiers to effect high in situ sulfur capture. This subbituminous coal contained 2.0 percent sulfur and 15.1 percent moisture (as-received basis). The partially desulfurized fuel gas exited the gasifier at 1,820°F (5).

In the cases evaluated in this paper, Wyoming subbituminous coal was dried to 10 percent moisture prior to feeding to fluidized-bed gasifiers, a step thought to be required to ensure reliable operation of the pressurized feed system. Clean fuel gas, extracted from the fuel gas stream to the gas turbine, was used for coal drying. Bulk sulfur capture occurred in the gasifiers through the use of limestone as the in situ sulfur capture medium. [Limestone rather than dolomite was used because limestone is typically lower cost (at a given Ca/S molar ratio) and KRW tests showed that the sulfur capture efficiency of limestone is similar to dolomite.] It was assumed that 90 percent of the sulfur in the coal would be retained by the ash and by sorbent in the gasifiers (limestone Ca/S molar ratio of 2.0) and that 0.25 pounds of steam per pound of dry, ash-free coal were required for the operation of the gasifiers. The air blast for the gasifiers was extracted from the gas turbine compressor discharge, cooled in an intercooler, and then further pressurized in a motor-driven booster compressor. The heat and material balances around the gasifiers were calculated using a computer model which used a minimization of Gibbs Free Energy algorithm to calculate gas composition. A total carbon loss from the gasifiers (in the gasifier ash and tertiary filter fines) of 2.5 percent was assumed, based on data from Reference 7.

On exiting the gasifiers, the fuel gas was treated in two stages of recycle cyclones and then cooled to 1,100°F using a direct water quench. (The quench has a lower capital cost than radiant and convective heat exchangers.) The fuel gas then passed sequentially through a tertiary particulate removal device, in this case a ceramic cross-flow filter, and a zinc ferrite desulfurization process. It was assumed that the fuel gas leaving the zinc ferrite unit was sufficiently particulate free to meet both gas turbine manufacturer's specifications and NSPS regulations for particulates. Because zinc ferrite also serves as a catalyst for the exothermic water gas shift reaction, it was assumed that equilibrium with respect to this reaction was achieved in the reactor with a resultant increase in the fuel gas temperature.

The zinc ferrite system was regenerated using a mixture of air and process steam. The hot, SO_2-containing offgas from the regeneration step was cooled by heat exchange with the incoming air/steam mixture (thereby heating this mixture to 900°F) and then recycled to the gasifiers. It was assumed that solid waste mixture of spent sorbent and coal ash produced in the gasifiers could successfully be oxidized to a nonleachable, easily disposable material in the base or oxidizing region of the gasifiers.

Gas turbine performance was based on ISO conditions (59°F, sea level pressure, 60 percent relative humidity). Hot path metal parts in the expansion turbine were cooled using air extraction bleed streams from the compressor.

Case 1: Fluidized-bed gasifier ISTIG gas turbine system. Case 1 (Figure 5) used one ISTIG operating at a pressure ratio of 35 and a firing temperature of 2,400°F for power production. Steam produced in a HRSG was injected into the gas turbine combustor at a rate of 15 percent of the total mass flow of air to the compressor. It was assumed that the expander portion of the gas turbine could accommodate the additional volumetric flow resulting from the steam injection and the use of low-Btu gas as a fuel. The gas turbine compressor was intercooled, and the heat from this intercooler was used to preheat the boiler feed water. To accommodate the high-pressure ratio gas turbine, the gasifiers were operated at 600 psig. The net output for this system was 128 MWe; the heat rate was 8,240 Btu/kWhr (41.4 percent coal pile to busbar efficiency, including coal drying).

Case 2: Fluidized-bed gasifier/combined cycle. The gasification island in Case 2 (Figure 6) is similar to Case 1 except that the gasifiers operated at a pressure of 450 psia, and the ISTIG turbine was replaced by a combustion turbine combined-cycle (CC) system. The gas turbine, operating at a pressure ratio of 12 and a firing temperature of 2,300°F, produced 169 MWe. A condensing steam turbine bottoming cycle, with turbine throttle conditions of 1,800 psia and 1,000°F, and a condenser pressure of 2-inch HgA, produced an additional 88 MWe. Steam for the gasifier and the zinc ferrite regeneration was extracted at appropriate locations from the steam turbine. The combined-cycle system produced 224 MWe net at a heat rate of 8,790 Btu/kWhr (38.8 percent efficiency).

INVESTMENT COST

The total capital requirement (TCR) for gasification island Case 1 and 2 was estimated. These results are shown in Table 1. For comparison, Table 1 also shows the TCR for a fixed-bed gasification island case from a recent GE study (3). This GE study used Illinois No. 6 coal, an ISTIG, and included a sulfuric acid plant for disposal of the SO_2-containing offgas stream for the zinc ferrite unit. All costs in Table 1 are reported in 1985 dollars and were updated using the Chemical Engineering Plant Cost Index.

The TCR for the two KRW gasification island systems shown in Table 1 are considered to be cost targets. The plant section capital costs on this table were estimated using costs reported in the literature and appropriate scaling relationships. In the METC study, it was assumed that the gasification island technology was mature and that equipment vendors marketed the plant as a standardized design. Thus, process contingencies were assumed to be zero and only minimal engineering costs were factored into the estimates.

The solids preparation plant section, including facilities for handling both coal and limestone, was costed based on material flow. A 0.7 scaling factor was applied. Costs for this plant section were derived from those reported by Fluor (8). Costs for the cyclones and zinc ferrite unit were derived from Reference 3. This reference included the cost of a sulfuric acid plant with the zinc ferrite plant section cost; the cost of the zinc ferrite unit was

TABLE 1. Comparison of Coal Gasification Power Systems

	Gasification Island			First-Generation IGCC			
	Fluid-Bed/ ISTIG	Fluid-Bed/ CC	Fixed-Bed/ ISTIG	Texaco CC	BGC/Lurgi CC	KRW CC	KRW CC
COAL TYPE	WY Sub	WY Sub	IL No. 6	IL No. 6	IL No. 6	IL No. 6	TX Lig
Performance							
Thermal Efficiency, %	41.4	38.8	42.1	37.9	39.4	37.7	34.0
Capacity, MW	128	224	110	390	340	365	320
Capital, $/kW							
Plant Section							
Solids Handling	75	70	40	65	60	50	70
Gasification	225	210	95	245	110	310	375
Fuel Gas Physical Cleanup	20	20	10	--	--	--	--
Fuel Gas Chemical Cleanup	60	55	165[1]	100	50	70	50
Power Cycle	250	300	280	415	275	370	385
Oxygen Plant or Compressor	10	10	10	180	90	150	170
Balance of Plant	160	165	155	195	350	185	195
Process Contingency	--	--	40	40	55	70	80
Project Contingency	120	125	120	180	150	175	200
Start-Up and Working Capital	40	40	55	105	110	105	100
AFDC	45	45	20	100	70	100	110
Total Capital Requirement	1,005	1,040	990	1,625	1,320	1,585	1,735
COE, Mills/kWhr							
Capital	33	34	33	53	43	52	57
Fuel	13[4]	13[4]	16[2]	18[2]	17[2]	18[2]	13[3]
O&M	7	7	7	11	15	11	13
Total COE	53	54	56	82	75	81	83

[1] Includes H_2SO_4 plant for SO_2 recovery.
[2] Coal cost $1.60/MMBtu; 0.7 percent escalation.
[3] Coal cost $1.05/MMBtu; 1.0 percent escalation.
[4] Coal cost $1.20/MMBtu; 1.0 percent escalation.

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35 percent of the total reported cost (11). The costs for a tertiary particulate removal device, a ceramic cross-flow filter, were based on a Gilbert/Commonwealth study (12).

The fluidized-bed gasifier plant section was costed using information reported in Reference 7. The gasifiers were sized on the basis of the moisture and ash-free coal feed, with cost increased by 25 percent to account for the use of an air-blown system. Based on Reference 8, the gasifier plant section cost was reduced by $55/kW to reflect the replacement of radiant/convective heat exchangers with a quench. The power generation block consisted of the gas turbines, HRSG, steam turbine for the combined-cycle case, generator, and associated equipment. The costs for the combined-cycle and ISTIG systems were developed from Reference 3.

The balance of plant included the following components: site work, foundations, control and administrative buildings, fire protection, fencing, compressed air systems, distillate fuel storage and handling systems, auxiliary steam, site wastewater disposal and drainage system, raw water, water treatment system, and circulating water system. The balance of plant was estimated to be 25 percent of the total of the direct equipment, material, and labor cost of the other plant sections. Factors of 15, 5, and 5.5 percent of the total direct plant investment were used to estimate the project contingency, startup and working capital, and allowance for funds during construction (AFDC) categories.

In order to compare gasification island systems with first-generation IGCC systems using cold gas cleanup, Table 1 also contains information for four first-generation IGCC systems. The TCR for these four IGCC systems was based, respectively, on data from Electric Power Research Institute (EPRI) conceptual designs for Texaco, BGC/Lurgi, and KRW IGCC systems (8,9,7) operating on bituminous coal and a KRW IGCC system operating on lignite (7). The plant sizes given in the EPRI designs were downscaled to plants containing what was believed to be two gasification/gas turbine trains, i.e., plant sizes of 390, 340, 365, and 320 MWe for the Texaco, BGC/Lurgi, and two KRW systems, respectively. Factors for this downscaling were abstracted from an EPRI study (10) showing the effect of plant size on the economics of coal gasification combined-cycle plants. For the Texaco and KRW designs, the engineering costs were prorated among the individual plant sections; in the BGC/Lurgi design, engineering cost appears in the balance of plant category. The TCR of the first-generation IGCC systems is approximately 50 percent higher than the TCR of the gasification island systems.

COST OF ELECTRICITY

The COE for the three gasification island cases and the four first-generation IGCC systems shown on Table 1 was determined using the methodology outlined in EPRI's Technical Assessment Guide (13). In the METC study, all COE values are reported in tenth year, current dollars, levelized. The economic assumptions used in this study are given in Table 2. Table 1 shows that the COE of the gasification island systems is approximately 30 percent lower than the COE first-generation in IGCC systems.

TABLE 2. Economic Assumption for Gasification
Island Study

General Inflation Rate, %/yr	4
Discount Rate, %/yr	11.5
O&M Escalation Rate, %/yr	0.7
Book Life, Years	30
Tax Life, Years	15
Capacity Factor, %	65
Federal and State Income Tax Rate, %/yr	38
Property Taxes and Insurance, %/yr	2
Investment Tax Credit, %	0
Levelized Capital Charge Factor, %/yr	18.7
Coal Cost, $/MMBtu; Escalation Rate, %/yr	

Illinois No. 6	$1.60; 0.7
Texas Lignite	$1.05; 1.0
Wyoming Subbituminous	$1.20; 1.0

The gasification island cases operating on subbituminous coal were also compared
to PC/FGD plants using the same coal as a feedstock. Based on Reference 13,
the heat rate of a 330 MWe size PC/FGD plant was estimated to be 10,070 Btu/kWhr
(33.9 percent efficiency). The TCR was $1,600/kW; the levelized operating and
maintenance (O&M) cost was 10 mills/kWhr. With these assumptions, the resultant
COE from a PC/FGD plant operating on subbituminous coal is 78 mills/kWhr --
approximately 45 percent higher than the gasification island cases.

CONCLUSION

A screening study has shown that small size (nominally 100 to 200 MWe) gasi-
fication island concept systems operating on low-rank coal have the potential
for a significantly lower COE than first-generation IGCC systems or PC/FGD
plants. For the gasification island system, simple process configurations and
standardized designs tailored to specific gas turbine systems lead to low
capital cost. Thermal efficiency is high, particularly for the steam-injected
cycles. Very positive results are emerging from experimental programs support-
ing gasification island, indicating the technical viability of the concept.
Using high-temperature cleanup processes, the concept has the potential for
superior environmental performance. Gasification island provides utilities
with a new technological path for economically attractive power generation
systems in small capacity size increments.

U.S. industry recognizes the advantages of the gasification island concept as
evidenced by the fact that three of the proposals submitted to the Department
of Energy (DOE) under the Clean Coal Technology Program were for power genera-
tion systems utilizing this approach. This program, with $400 million in
Federal funds, requires a minimum of 50 percent cost sharing by the private
sector firms receiving awards. DOE selected two proposals for final negotia-
tions: the "Integrated Gasification-Steam Injected Gas Turbine" proposal
submitted by General Electric and the "Appalachian Project" submitted by the
M. W. Kellogg Company.

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Gasification Island Systems

First Generation IGCC Systems
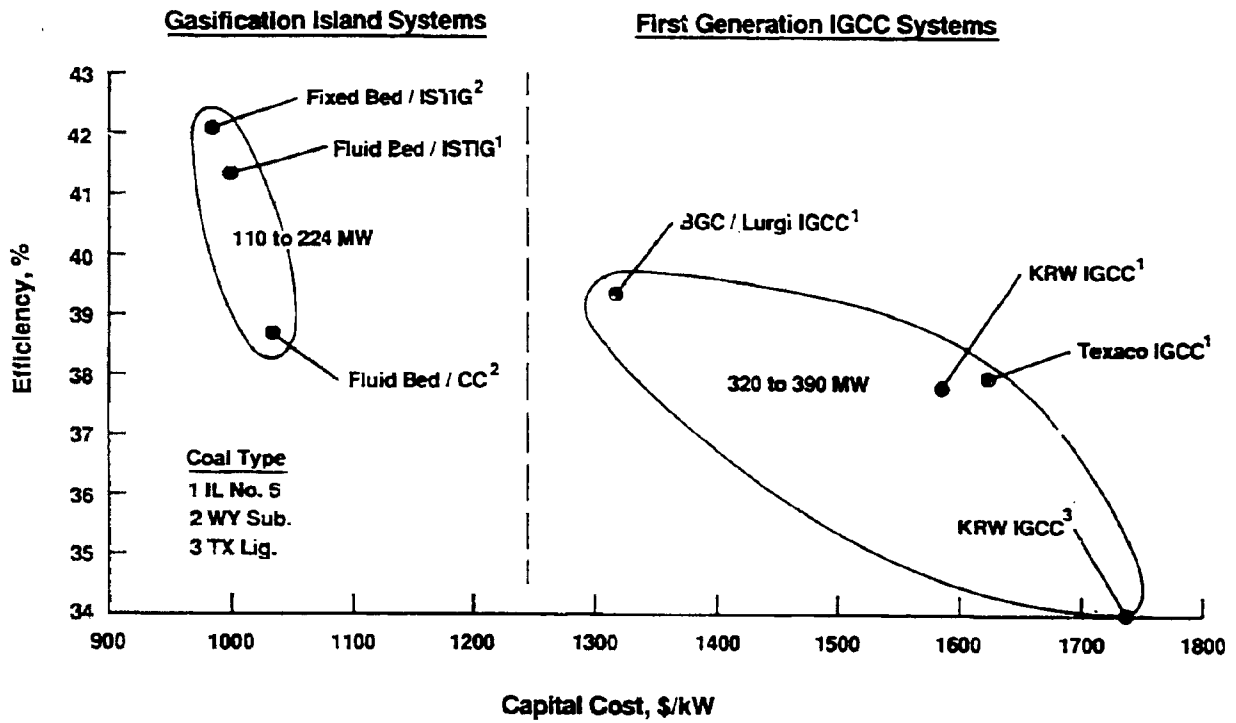
FIGURE 1. Comparison of Gasification Island and Conventional IGCC Systems
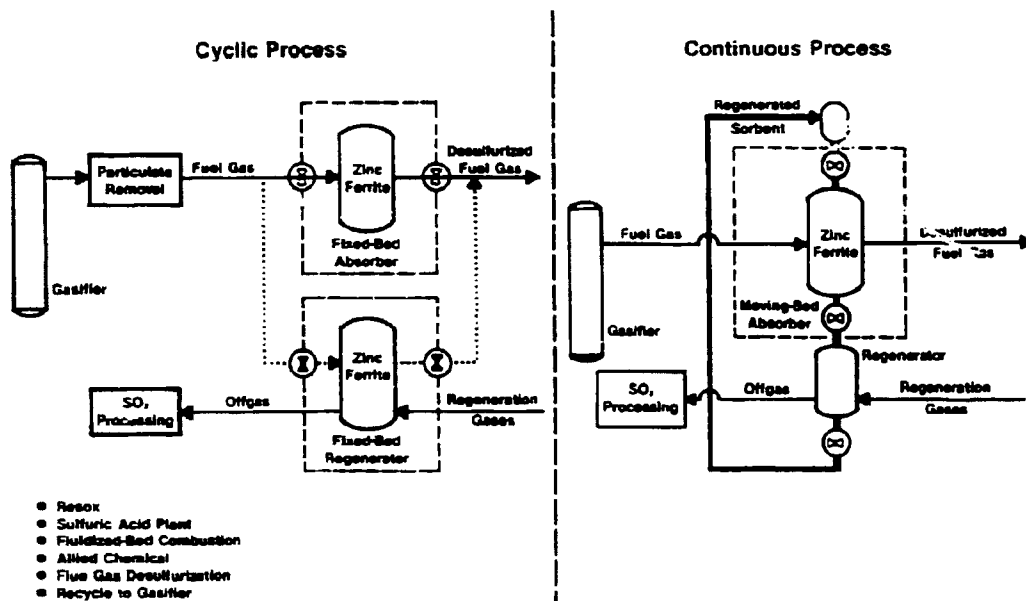
FIGURE 2. Zinc Ferrite Process Configuration Options

Conclusion:
Ash and alkali levels appear compatible with current gas turbines
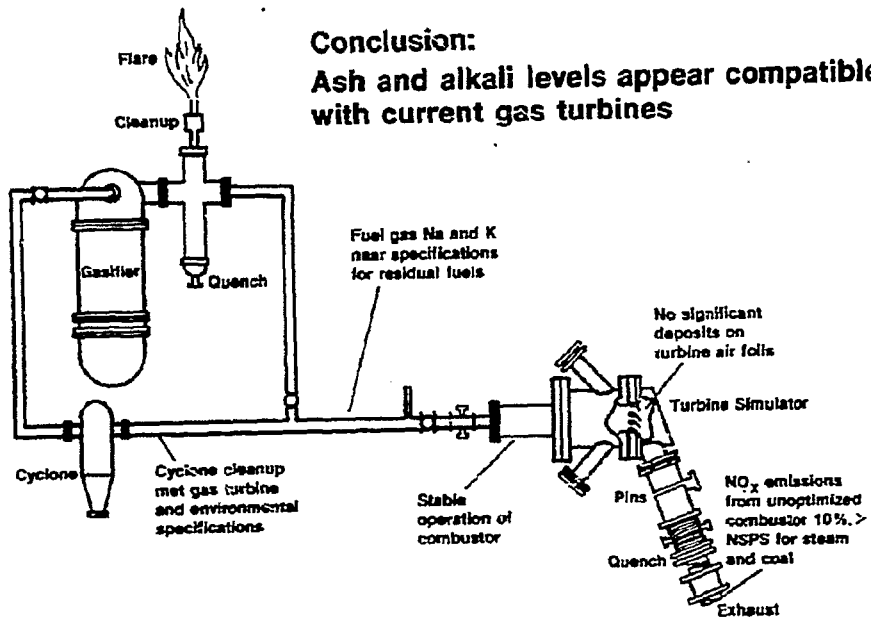
FIGURE 3. Results from GE Components Test
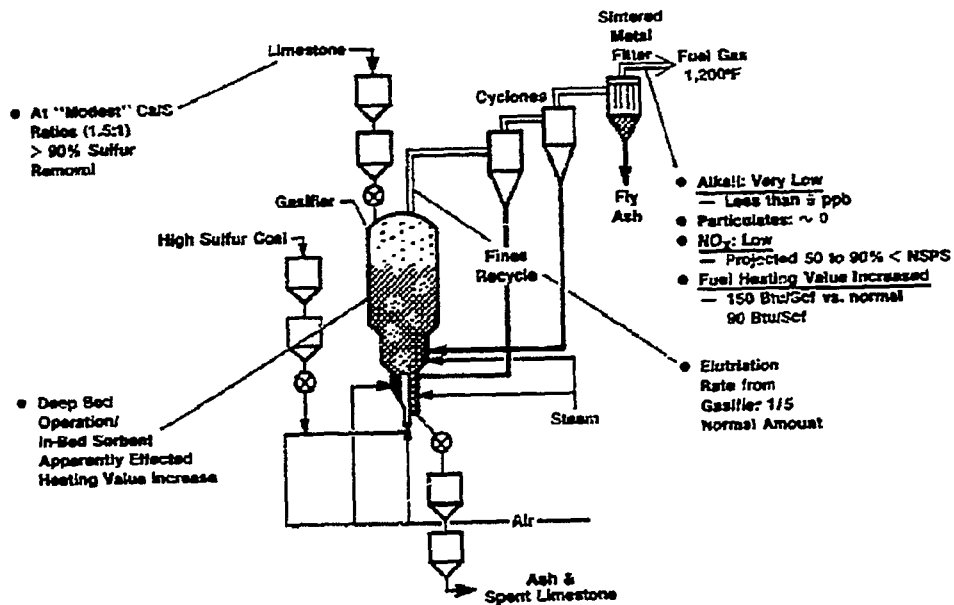
FIGURE 4. Operation of and Results from KRW In-Bed Desulfurization Tests
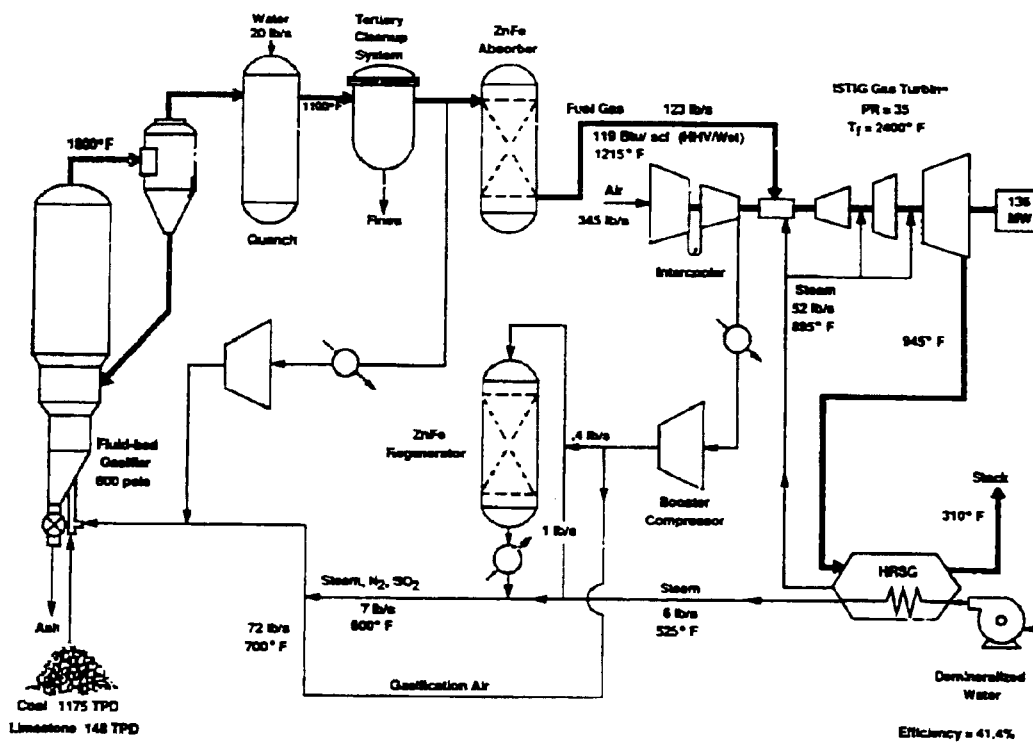
FIGURE 5. Case 1: Fluid-Bed Gasifier/ISTIG -- 224 MW
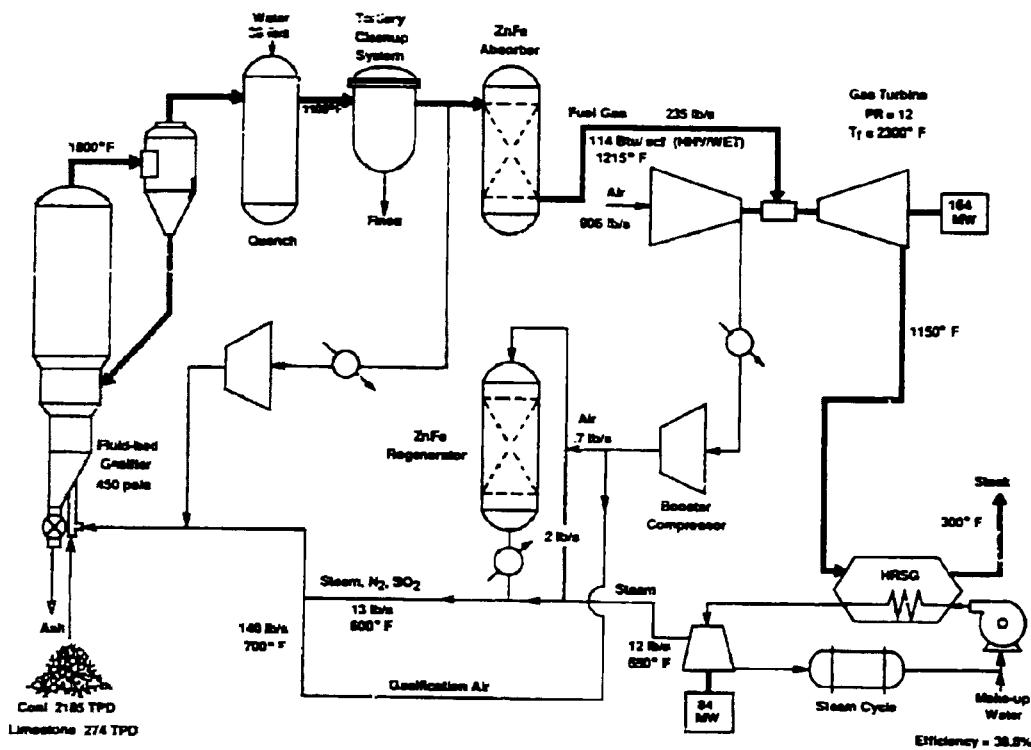
FIGURE 6. Case 2: Fluid-Bed Gasifier/Combined Cycle -- 128 MW

1 B1-17

CONVERSION OF LIGNITE TO LIQUID FUELS,

USING FISCHER-TROPSCH SYNTHESIS

Paper to be presented by

Tino van der Zeyde, CEng; FIChemE

at the

Fourteenth Biennial Lignite Symposium

to be held at the

Dallas Sheraton Hotel, Dallas Texas

on May 19, 1987

(This paper forms part of the

Commercial Utilisation Stream of the Symposium)

SYNOPSIS

An approach is outlined to the application of the SYNTHOL SYNTHESIS PROCESS
as owned and operated by SASOL to using LIGNITE rather than Sasol's young
BITUMINOUS COAL as the raw material. Factors are presented which play a
part in shaping an overall plant complex to this purpose.

Of the many gasification systems available, also for gasifying lignite,
three were chosen to illustrate important different parameters of these,
i.e. the LURGI, HIGH-TEMPERATURE WINKLER and the SHELL GASIFICATION PROCESS.
The effect of temperature, pressure, formation or otherwise of by-products,
nature of the feed and other factors are discussed on raw gas composition,
and on the downstream steps of modifying the gas composition by reforming
or shift-conversion. Purification steps in order to arrive at the specified
Synthol Syngas.

The importance of good energy managment is explained and stressed.

After syngas manufacture follows the synthesis plant, its operation, which
reactions take place and thoughts on why; the nature of the catalyst,
observations on selectivity and reactivity, including loss of activity.
Some notes are given on selectivity and product distribution.

The magnitude of the various factors are illustrated by developing an
example, whereby approx. 14 million metric tons of lignite as received
(33% moisture - 42% Carbon) are converted to 2 million metric tons of
liquid fuels, i.e. gasoline, diesel and LPG. Such a capacity would compare
with a 50,000 bbl/day crude oil refinery, but would cost some $ 50,000 per
daily barrel. a number far in excess of that of a conventional refinery.
The Sasol operation is financially viable; this is explained. Also why not,
at this stage, in the USA. But the time will probably come - it is a good
idea to be prepared for what the future holds!

INTRODUCTION

Synthetic liquid fuels and chemicals have been produced from coal success-
fully at the SASOL PLANTS in South Africa since 1955. Currently some 5 mill-
ion tons of motor fuels are thus manufactured annually. This corresponds
with the capacity of a 120.000 bbl/day refinery. The investment, if incurred
to-day, would amount to about $ 50.000 per daily barrel, a number far in
excess of that for a regular crude oil refinery.

Nevertheless, the South African operation has proved financially viable;
shares are available to the public (the first issue was twelve times over-
subscribed) and this stock continues to show a good record. There are a
few reasons why, in the case of SASOL, this is so. Unlimited quantities of
low-grade coal are available within conveyor-belt distance of the plants.
By levying low excise duties on the products,the government created a favor-
able climate for this operation. And its main markets are reasonably close;
the oil companies take care of the other parts of the country, using import-
ed crude.

Including a talk on synthetic fuels at the Lignite Symposium would suggest
that a comparable situation could exist here in the USA. At to-day's crude
oil prices, this is far from true. But in ten or twenty years' time, crude
oil prices will be very much higher and its availability greatly reduced.
Preparing for the future is the name of the game!

Apart from SASOL, the major oil companies have set the example. SHELL owns
large coal-mining operations. This company developed excellent systems for
converting coal and natural gas to syngas and liquid fuels. MOBIL has one
plant already operating in New-Zealand which converts, according to its
own technology, natural gas to methanol and methanol to gasoline. A more
recent Mobil development allows the conversion of the methanol to a good
quality diesel as well, using the "MOGD-Process". The EXXON-DONOR-System
is another example of a promising development in this field.

OBJECTIVES

An example has been chosen whereby 2 million metric tons of liquid fuels, i.e. gasoline, diesel and LPG, are produced from Texas Lignite annually. The weight ratio between gasoline and diesel is set at 2-to-1.

The author was asked to apply Sasol technology in this talk. But Sasol Coal is not the same as Texas Lignite. According to Sasol-Technology, the solid feed is converted to raw gas which is modified and purified to give the "Synthol"-type of SYNGAS. Producing syngas takes anything between 67 and 75% of the total investment. No wonder therefore that the author has to pay attention to the production of this syngas, even though talks on gasification are offered at the symposium. "Synthol" is the trade-name of the synthesis process operated and owned by SASOL.

The main objectives of this short study are therefore to convert sufficient Lignite of analysis and characteristics as shown in TABLE 1 to Syngas as shown in TABLE 2, in order to arrive at the final product distribution as shown in TABLE 3. (Tables 1 and 2 are on page 4; Table 3 is on page 5.)

Additional objectives, which are also complied with by SASOL:-

> No liquid effluent whatsoever leaves the plant
> Not more than 5% of the sulfur in the lignite
> may leave the plant in gaseous effluent

PRODUCTION OF SYNGAS

The process steps required for the conversion of the selected type of lignite to synthol syngas are primarily gasification, modification and purification. It depends on which type of gasifier is chosen, what the further process steps will look like. There are about fifteen different gasification processes; about half of these are operated commercially; the others are in various stages of development, from pilot plant operation to operating demonstration plants.

TABLE 1 - ANALYSIS OF TEXAS LIGNITE USED FOR THE EXAMPLE

Proximate analysis (%wt)		As received (%wt)
Ash	9.0	9.0
Moisture.	33.0	33.0
Volatile matter	31.0	–
Fixed carbon	27.0	–
	100.0	
Ultimate analysis (maf)		
C	72.6	42.1
H	6.0	3.5
O	19.2	11.1
N	1.2	0.7
S	1.0	0.6
	100.0	100.0

(With kind permission of SHELL INT.)

TABLE 2 - SYNTHESIS GAS COMPOSITION FOR THE SYNTHOL PROCESS

H_2 .... 61.0%(vol) H_2/CO ratio: 2.0

CO .... 30.5 "

CO_2 .... 0.9 "

CH_4 .... 0.1 "

H_2O .... 3.1 "

N_2 .... 4.4 "

100.0 "

N.B. This composition deviates from the one used at Sasol where Lurgi gasification is applied. More methane and CO_2 would be present than when e.g. Shell gasification, followed by shift-conversion and scrubbing is done

TABLE 3 - DISTRIBUTION OF PRODUCTS LEAVING THE REACTOR SECTION

	(1000 metric tons per annum)			
	C	H	O	total
saleable prodducts:				
LPG $(C_{3+} + C_{4+})$	90.0	19.3	–	109.3
gasoline $(C_{5-12} + 0.25 \times CPP^* + 0.70 \times C_{2.5}H_5O_{1.25})$	1043.4	180.3	34.7	1258.4
diesel $(C_{14}H_{28.5} + 0.75 \times CPP(ave.))$	541.4	86.2	–	627.6
wax $(C_{25}H_{51})$ (ave.)	89.0	15.3	–	104.3
alcohols, acetone, MEK(excL.ethanol)	2.8	0.5	1.9	5.1
sub-total (saleable products)	1766.6	301.6	36.6	2104.7
non-saleable products:				
fuel gas:(for internal use)				
unconverted CO	450.6	–	600.7	1051.3
unconverted hydrogen	–	75.1	–	75.1
$C_1 + C_{2-}^{**} + C_{2+}$	416.4	14.4	–	430.8
water vapour	–	16.0	128.0	144.0
sub-total (fuel gas, excL.N$_2$)	867.0	105.5	728.7	1701.2
water (to Cooling W. after treating)	–	130.6	1044.7	1175.3
aliphatic acids (biolog. destroyed)	60.8	10.1	72.9	143.8
sub-total (non-saleable products)	927.8	246.2	1846.3	3020.3
grand total	2694.4	547.8	1882.9	5125.0

The above data is derived from Dr M.E. Dry's article ([1])

* CPP = cat poly product; ** : Ethylene is recovered if it pays to do so. It was omitted from the example.

The following claims, generally valid, are found in gasification marketing literature:

optimum utilisation of the carbonaceous feed

dependability of operation

not polluting the environment

cost-effectiveness

For some types of gasifiers the following additional claims are valid::

        the process accepts any type of carbonaceous feed
        the process operates in the desired pressure range

Gasifiers marketed have good track records. The prospective lignite conv-
ertor analyses his own situation, thereby considering the mass and energy
picture, not of a specific gasifier in isolation, but of the whole arrange-
ment, all the way from lignite, thru raw gas, syngas and synthesis with
product work-up and corresponding utilities, waste water treatment etc. in
order to arrive at his final combination of choices which will give him the
optimum operation as a whole.

This is a short treatise on plant complexes in which the final products are
obtained by using the Synthol type of synthesis; it is not a detailed study
of the different gasifiers. Three types of gasifiers have been chosen for
incorporation into such a plant complex. These types vary considerably and
the differences are instructive. The choice is not meant to express pref-
erence for one type over another. These three types are:-
                Lurgi Pressure Gasification
                High-Temperature Winkler - Uhde
                Shell Coal Gasifcation Process

The main features of these three types of gasifiers have been tabulated in
TABLE 4 on page 7. Some notes follow on the main features listed in the
table.
The most significant difference is that of the temperature. At high temps.
 relatively more oxygen and less or no steam is used. The result is a H_2/CO-
ratio of e.g. 0.5, whereas our syngas should have 2.0. This can easily be
corrected downstream by applying shift conversion. Part of the CO reacts
with steam (either originally present in the lignite, or freshly fed steam)
to give Hydrogen and CO_2. Each reacting mole of CO and/or H_2 is obtained
by consuming one C-atom from the lignite. So - it doen't matter whether
the required number of moles of hydrogen were there already in the raw gas
or - whether they formed later during the shift conversion.

TABLE 4 – MAIN FEATURES OF THREE SELECTED TYPES OF GASIFIER

feature	LURGI	H.T.-WINKLER	SHELL
type of bed	fixed	fluid	entrained
suitable for lignite	can be made suitable	yes	yes
type of feed	pieces (6-50mm) (¼ – 2in.) if fines pres. outlet to be found	pulverised suff. dry for pneum.conv.	two options: dry slurry < 100 micron
operating pressure	17 – 34 bar	10 bar*	20 – 40 bar
operating temperature	± 500°C(900°F)	1100°C(2000°F)	1200 – 1400°C 2200 – 2500°F
solid ash or slag	solid	solid	gran.slag
gas composition (%vol)(these are merely examples; compositions can be varied)			
H_2	35.0	much between	32.4
CO	21.0	LURGI	61.8
CO_2	27.7	and	4.6
CH_4	12.0	3.6	0.1
H_2S – COS (made equal)	0.3	SHELL	C.3
N_2 + Argon	4.0		0.9

In a LURGI G. the exact H_2/CO-ratio required can be produced. Since the LURGI Gasifier (L) produces methane, which the high-temperature G.'s do not produce, it is considered advantageous to "reform" the methane, formed in gasification with further methane formed in the synthesis reactors; by reacting methane (and ethane or even ethene) with oxygen and steam, about half these light hydrocarbons are converted to CO and H_2; the other half ends-up as CO_2 and its carbon is lost to the process. ·Therefore, it may be concluded that , if L.is used, a reforming plant forms part of the gas train and no shift-converting plant is required. If U. or S. is used, there is insufficient methane to justify reforming, but - a shift conversion plant is needed. And the steam to go with it.

The next significant difference is that L. produces tar and ammonia liquor; the others don't. Tar and ammonia liquor are more a liability than an asset. Tar is distilled; the creosote can be hydrogenated and become part of the diesel production. If the gasoline/diesel-ratio has to be below 2.0, this could be a factor. The ammonia liquor contains phenols, which are removed by countercurrent extraction in a Lurgi Phenosolvan Plant to 300 ppm in H_2O. The ammonia is recovered by steam-stripping and converted to sulfate. With the other two gasifiers there is no such by-product problem. And a problem it is, considering considerable investment and operating costs involved. That would seem to indicate that there is no point in selecting Lurgi. But that would not be true either. It should be remembered that Sasol has used Lurgi G.'s ever since the beginning. The Sasol people are well aware of the other options. Sasol coal is brittle and up to 40% coal fines may arise during handling. Therefore, Sasol could easily have Lurgi gasifiers in parallel with (an)other type(s) and thus use up the fines in the other gasifiers. By blending the gas and chosing volumes from Lurgi and other streams such that the ratio would come out right, would e.g. obviate shift convertors. The author could harldy imagine that Sasol wouldn't have thought of that; there must be other good reasons to stick to the Lurgi type of gasifier.

As regards the formation of CO_2 the three types of G.'s are not significantly different. Each mole of CO_2 is lost to the process, at least nearly so; the synthol catalyst does convert CO_2 as well as CO if a small percentage of it is present in the feed.

Undesirable by-products are formed in all gasification processes. At the higher temperature there tends to be more HCN and NH_3. In all cases there are significant amounts of H_2S and, to a lesser extent, COS. Selective scrubbing allows for the removal of CO_2, H_2S and COS to a greater or lesser extent. But for Synthol Synthesis, the lower the sulfur content in the feed, the better. Until recently 100 ppb (0.1 ppm) was acceptable; a recent development in Holland boasts 20 ppb; and synthesis operators would love that!

Nitrogen should also be considered in this comparison. If HTW.or S.G.'s
are used, it was said already, there would be no need for reforming; i.e.
for recycling synthesis tailgas. The tailgas could be added to the fuel gas
system. There might even be sufficient tailgas to keep the fuelgas system
going, considering that aproximately 20%(vol) leaves the reactor section
after removing condensible products. And if there is no tailgas recycle,
there is no fear of the nitrogen concentration building up in the gas system.
Then the nitrogen need not be.as pure as it would otherwise have to be. And
this would save power. A look at TABLE 3 makes it clear that masses of O_2
are produced and the nitrogen volumes are four times as large! What to do
with all that nitrogen. There are some rather minor essential in-plant uses.
Reduced catalyst is conveyed and blanketed with nitrogen; most fractions in
the work-up section are subjected to some form of hydro-treating; to make
plant safe for welding, etc. etc. But a major use Sasol makes of it is the
production of nitrogenous fertilisers, such as ammonium nitrate with or
without calcium.

In order to obtain the very low sulfur limit, mentioned earlier, there is
the Lurgi Rectisol Process which worked wonderfully well at Sasol ever
since the button was pressed the very first time. And lots of Rectisol
Plants have been built since all over the world. But - always a but -
there has been some trouble with the process in North Dakota. In the Rect-
isol, the whole of the gas stream is cooled to minus 60°C (76°F below) and
washed with methanol. This methanol removes sulfur almost completely but,
obviously, it must be stripped of the dissolved sulfur compounds, i.e. re-
generated. The problem in ND was in the methanol failing to regenerate, un-
til a solution (of which the author is not aware) was found.

The Stretford Process is also well-known for sulfur removal from gas streams.
And there are various others, depending how clean the gas has to be.

It should be mentioned that there are shift conversion catalysts which can
operate well when the sulfur is still in the gas. But it is most important

that the dust (flyash) leaving the gasifier with the gas is virtually com-
pletely removed. And in the case of high-temperature gasifiers, the passa-
ge where the gas leaves must be cooled in order to prevent drops of liquid
slag reaching the inside of the Waste-Heat-Boilers downstream of the G.'s.

That brings us to "Energy Management" of such systems. They stand or fall by
judicial use and transportation of heat and compression energy. It starts
with the G.-walls; depending on the operating temperature, medium or high
pressure steam is generated by heat extraction from the G.walls and from
the W.H.B.'s where the raw gas loses most of its sensible heat. Gas-to-gas
heat exchangers are obviously avoided; otherwise anything goes! Heat ex-
change between hot gas and BFW under pressure is one favorite way of coll-
ecting heat from the gas stream.

A prospective lignite conversion operator has a complex problem in sorting
out to what extent it pays to collect heat in a decentralised fashion or
rather by centralising and sending enthalpy-rich BFW to a central boiler-
cum-power station. Whether to run steam turbines rather than explosion-
proof electric motors, e.g. for the oxygen or the recycle compressors in
the synthesis plant. There is no universally applicable solution for this.

If Lurgi G.'s are chosen the volume of liquid effluent to be treated is much
larger than in the other two cases, there being no gas liquor with the latt-
er. The gas liquor (in the case of Lurgi), from which most of the phenols
and ammonia have been removed and the water of reaction from the synthesis
plant, which contains the aliphatic acids, formed by the synthesis reaction,
are biologically oxidised in an activiated sludge plant in the case of the
Sasol Secunda Plant. It would be possible to apply aneaerobic conversion
and produce methane rather than CO_2. But even if methane is produced, about
40% of the carbon converted by this anaerobic action ends-up as CO_2 as well.

It is not possible to treat each and every aspect and plant section of such
a plant. It is now time to deal with the Synthol Synthesis Process.

SYNTHOL SYNTHESIS

The working principles of the Synthol Reactor are shown in FIGURE 1. A flow of catalyst of thousands of tons per hr, at reaction temperature falls into the entering gasstream, which consists of approx. one third of fresh feed and two thirds of recylcle gas. The reaction starts immediately. When the first HE is reached, a large proportion of the evolved heat of reaction is transferred to a flow of oil which conveys the heat to a MPSteam Boiler (±12 bar)

FIGURE 1 - SYNTHOL REACTOR [1]
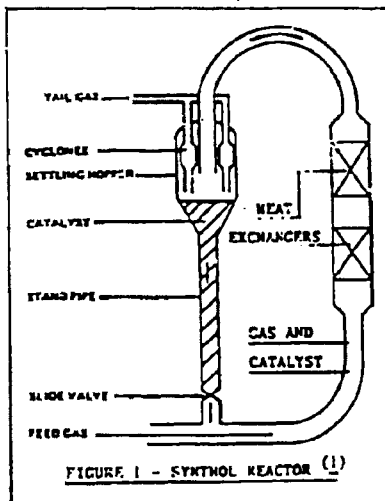
The reaction continues and more heat is extracted in the 2nd HE. Partial pressure then become low and the reaction stops. The reactor operates virtually under isothermic conditions, a very narrow temp.band being in evidence. This is an important factor in maintaining the desired selectivity. The product gasstream enters the cat settling hopper; gas and cat are separated virtually completely, by passing thru a series cyclones; a small portion of cat fines gets thru. The cat settles in the hopper cone and works its way down the standpipe. It then enters the gasstream again thru a controlled slide valve. The cat stream should be seen as a "heat flywheel". The system is a remarkable case of balancing gas and solids' mass flows, quantities of sensible and reaction heat, velocities and concentrations.

Downstream of the cat settling hopper, cat fines and a heavy oil fraction are removed in a scrubber tower. After further HE's and Coolers, an oily and an aqueous stream are collected. The oily stream contains the gasoline fraction and part of the diesel. The aqueous stream contains the oxygenates in solution, alcohols, aldehydes, ketones and aliphatic acids. Most of these products are recovered and refined; acet aldehyde is treated and, with the etanol, becomes part of the gasoline product. Cat poly product can be run up to 75% diesel by recycling. The C3+4(sats) are marketed as LPG. The oily fractions are subjected to some further hydrotreating and become internationally accepted grades of gas and diesel. Alcohols are separated and marketed; most of the ethanol is added to gasoline. The acids are destroyed.

TABLE 5 - SYNTHOL REACTIONS

$$(2n+1)H_2 + nCO \longrightarrow C_nH_{2n+2} \text{ (aliphatics)} + n\ H_2O$$

$$2n\ H_2 + nCO \longrightarrow C_nH_{2n} \left\{ \begin{array}{l} \text{olefins +} \\ \text{naphthenes} \end{array} \right\} + n\ H_2O$$

$$2n\ H_2 + nCO \longrightarrow C_nH_{2n+1}OH \text{ (alcohols)} + (n-1)H_2O$$

$$(2n-1)H_2 + nCO \longrightarrow C_nH_{2n}O \left\{ \begin{array}{l} \text{aldehydes} \\ \text{ketones} \end{array} \right\} + (n-1)H_2O$$

$$(2n-2)H_2 + nCO \longrightarrow C_nH_{2n}O_2 \left\{ \begin{array}{l} \text{aliphatic} \\ \text{acids} \end{array} \right\} + (n-2)H_2O$$

$$(2n-3)H_2 + nCO \longrightarrow C_nH_{2n-6} \left\{ \begin{array}{l} \text{aromatic} \\ \text{hydrocar's} \end{array} \right\} + n\ H_2O$$

These reactions are influenced by the shift equilibrium reaction:-

$$CO_2 + H_2 \rightleftharpoons CO + H_2O$$

The series of reactions which take place simultaneously in a synthol Reactor are shown in TABLE 5. The Synthol reduced iron catalyst is also a good shift cat; the oxygen in the CO, inasfar as this not end-up in oxygenates, becomes part of the water of reaction. The synthol cat has a very favorable selectivity, as well as activity. By keeping conversion per pass low, temps. can be controlled very rigorously. When trying to answer the question: "Why do these reactions take place", the author quotes Prof.Ponec of Leiden Un.:"The driving force in the dissociation of CO is the formation of Metal-C and Metal-O compounds. If these are not sufficiently strong,(Pd, Pt, Ir) then the diss. is very sluggish. Where these are strong,the dissociation is rapid - but - the succeeding hydrogenation of C- and O-, which amounts to regeneration of the cat surface, is too slow.(III-V categories of metals)

Some metals are conductive to CO dissociation and yet, are able to retain part of the metallic surface; those are the best F-T-synthesis metals: Fe, Co, Ni and Ru." This was the dissociation of the CO part. Then follows the formation of CH_2-radicals. These radicals then polimerise to form longer or shorter chains. Hence there is a "chain-formation-termination-force. The Schulz-Flory equation,generally used in predicting the chain lengths of polymers, applies to this chain formation by and large. The distribution pattern concerns the relative lengths of chains formed, but also to what extent unsaturated and saturated hydrocarbons are formed; furthermore, to

12
11
10
9
8
7
6
5
4
3
2
1

%C-atoms

EXAMPLE

C_{10} appears in product-mix:
   2.7% as unsat.(C_{10-})
   1.3% as satur.(C_{10+})

(as % C-atoms of C converted; the
the conditions (325°C) which
apply in Dr Dry's article ([1]))

unsaturates

saturates

C_5 C_{10} C_{15} C_{20} C_{25} C_{40}
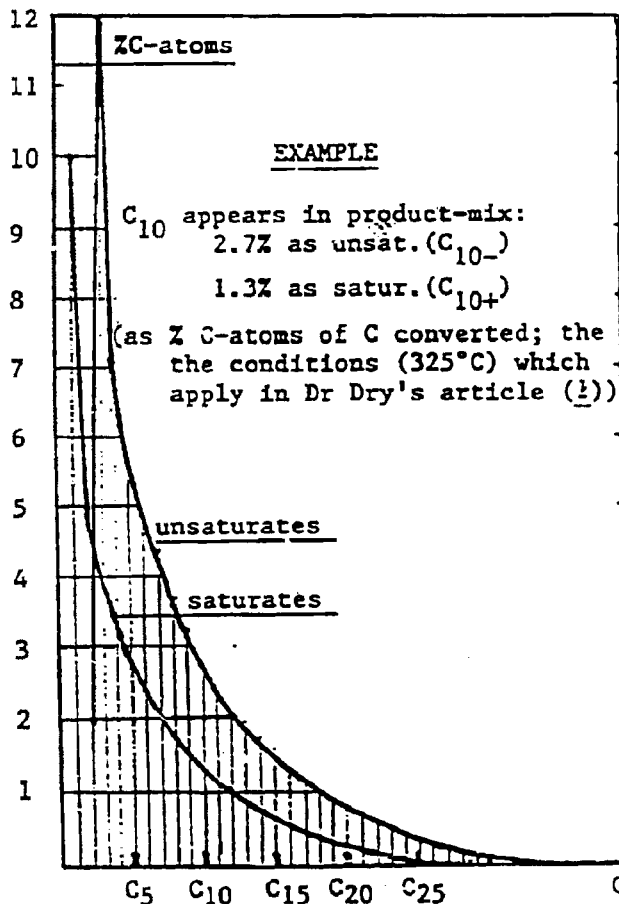
FIG. 2. - SYNTHOL PRODUCT
DISTRIBUTION

what measure are the chains straight
or branched? It can be said conf-
idently that the Synthol product
distribution is favorable compared
with several other catalyst systems.
FIGURE 2 shows the typical distrib-
ution curves for unsats and sats at
reactor temp. of 325°C(617°F). Note
that much more unsats are made than
sats; that C_3H_6 is the most prolific
product and, unfortunately, CH_4 is
of the sats! Work on improving sel-
ectivities continues; at Wits Un
a Co/MnO-cat made 4% CH_4 only with-
out detriment to remaining pattern.
(Asst.Prof.Hutchings) A reduction
of selectivity towards acids would
also be very valueable. (now 2.7%
of C-atoms converted)

Finally, the matter of maintaining catalyst activity. Activity declines with
time. As from early 1986, Sasol had a break-thru on this score which caused
plant capacity increase of 16% without having to add any plant. Pure cat-
science! Causes for activity decline are deposition of carbon and high-mol-
ecular substances on the cat surface - and - the presence of sulfur! Each
ppb less prolongs the cat life noticeably.

In conclusion on Synthol, it is a tricky process, needs close watching by
good operators but performs wonderfully well; so well in fact, that inter-
ested parties tend to chose Synthol in preference to other systems with
higher thermal efficiencies, e.g. 36% vs 33% on the overall complex.

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                          "The Shell Coal Gasification Porcess" (SCGP),
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                          Kiangsu, China - November 1986

CONVERSION OF LIGNITE TO LIQUID FUELS

USING FISCHER-TROPSCH SYNTHESIS

ERRATA

Page 1 : Center - "manag<u>e</u>ment"

Page 5 : Left top - "prod<u>u</u>cts"

Page 7 - HTW Center Column -
"Repeat asterisk after 10 bar
on next line up , followed by
"25 in future"

Pages 11 - Line 8 - "pressure<u>s</u> become"

Page 12 - Third line below Table -
"this <u>does</u> not"

My apologies for any inconvenience

<u>Tino van der Zeyde - Author of paper</u>

THE DOW SYNGAS PROJECT — DESCRIPTION AND STATUS REPORT

R. H. FISACKERLY

D. G. SUNDSTROM

THE DOW CHEMICAL COMPANY

INTRODUCTION

In the early 1970's, as a result of increasing energy costs and supply
uncertainties, Dow recognized the need to establish alternative sources
of energy to natural gas. In accomplishing this objective, Dow sought
ways to take advantage of the advancements in combustion turbine
technology and the United States' most abundant natural resource, coal.
The coal gasification research and development effort that followed
included construction of a 36 TPD Pilot Plant in 1979 and a 1600 TPD
Proto Plant in 1982. The unique Dow Coal Gasification Process
represents the results of Dow's research and development efforts. The
2400 TPD Dow Syngas Project is the first commercial scale demonstration
of that technology.

Project Description

The Dow Syngas Project is owned by Louisiana Gasification Technology,
Inc. (LGTI), a wholly owned subsidiary of The Dow Chemical Company
(Dow), the Project Sponsor. The Project is located near Plaquemine,
Louisiana, in Iberville and West Baton Rouge Parishes, and within the
existing Dow Louisiana Division. (Figure 1) The Project utilizes Dow-
developed coal gasification technology to convert Western sub-bituminous
coal, or other suitable coal, including lignite, into medium Btu
synthetic gas. The synthetic gas is purchased by Dow and is utilized in
the Dow Louisiana Division as fuel in combined cycle gas turbines co-
generating electricity and steam.

DOW SYNGAS PROJECT

FACILITIES AT THE
DOW LOUISIANA DIVISION SITE
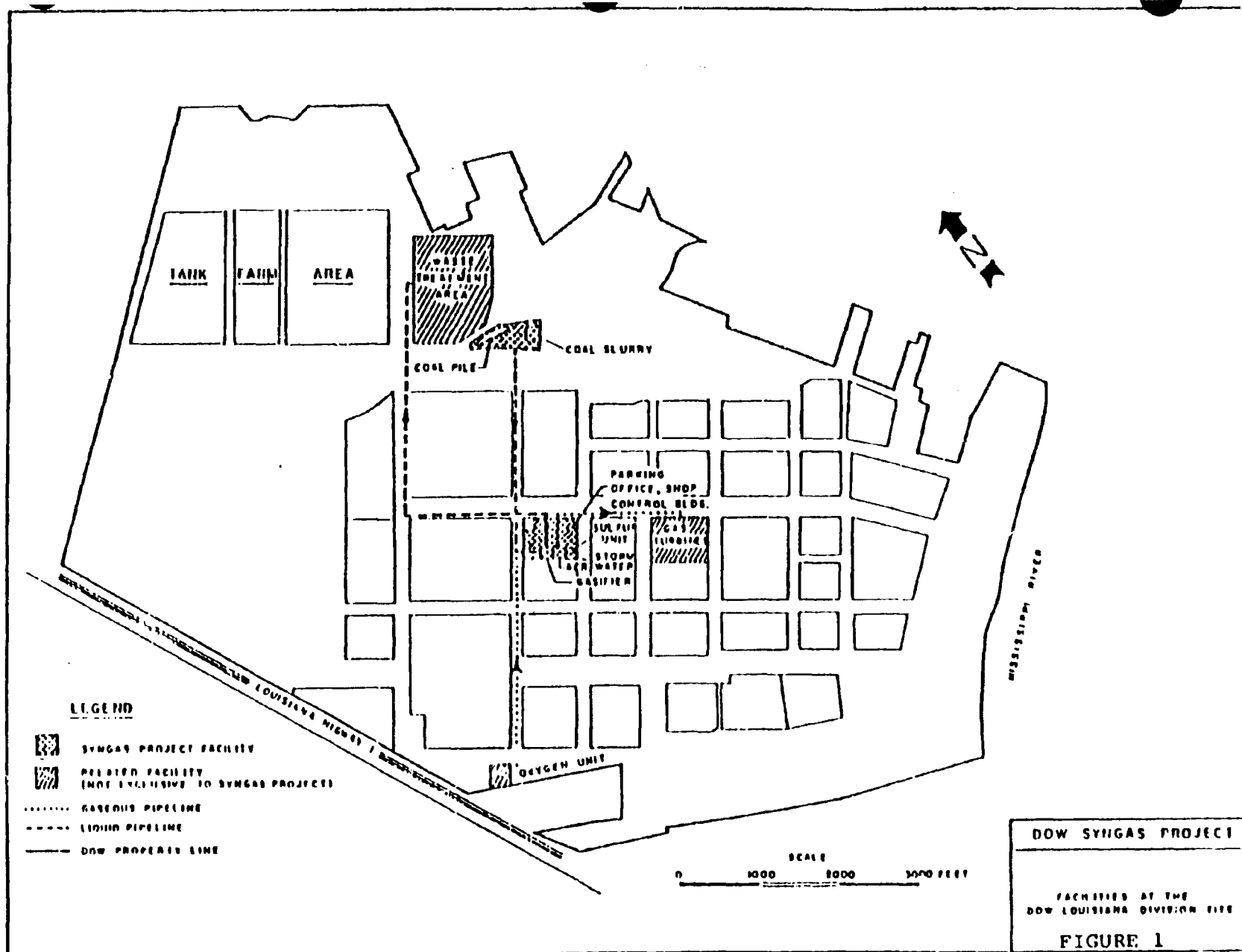
FIGURE 1

Project Description cont'd.

At full capacity, the Project operates at a sub-bituminous coal feed
rate of about 2,200 tons per day which will produce 30 billion Btu per
day of synthetic gas (See Figure 2). Additionally, at full capacity the
Project produces 2467 tons per day of steam. It is anticipated that,
within four years of start-up, the Project will be operating at full
capacity 85% of the year. The synthetic gas output is equivalent to
5170 barrels of oil per day. Sulfur is sold in the local market and slag
is sold in the local market or used on site for structural fill. The
feed for the Project is sub-bituminous coal from the Powder River Basin,
Wyoming, and the Project will continue to use Western sub-bituminous
coal while it is the economic feed of choice. The sub-bituminous coal
is transported from the mine to the Project by unit trains consisting of
from 70 to 110 rail cars per unit. The Project is designed to utilize
other coal, including Gulf Coast lignite, as an alternative feed.
Should supply or economic conditions dictate, the Project maintains the
capability and flexibility to switch to lignite for its feedstock. This
feedstock flexibility enhances the potential for replication of the Dow
Coal Gasification Process. Typical Material Balances for Western Sub-
bituminous Coal (Figure 3) and Gulf Coast Lignite (Figure 3A) are
included for reference.

Coal Grinding and Slurry Preparation

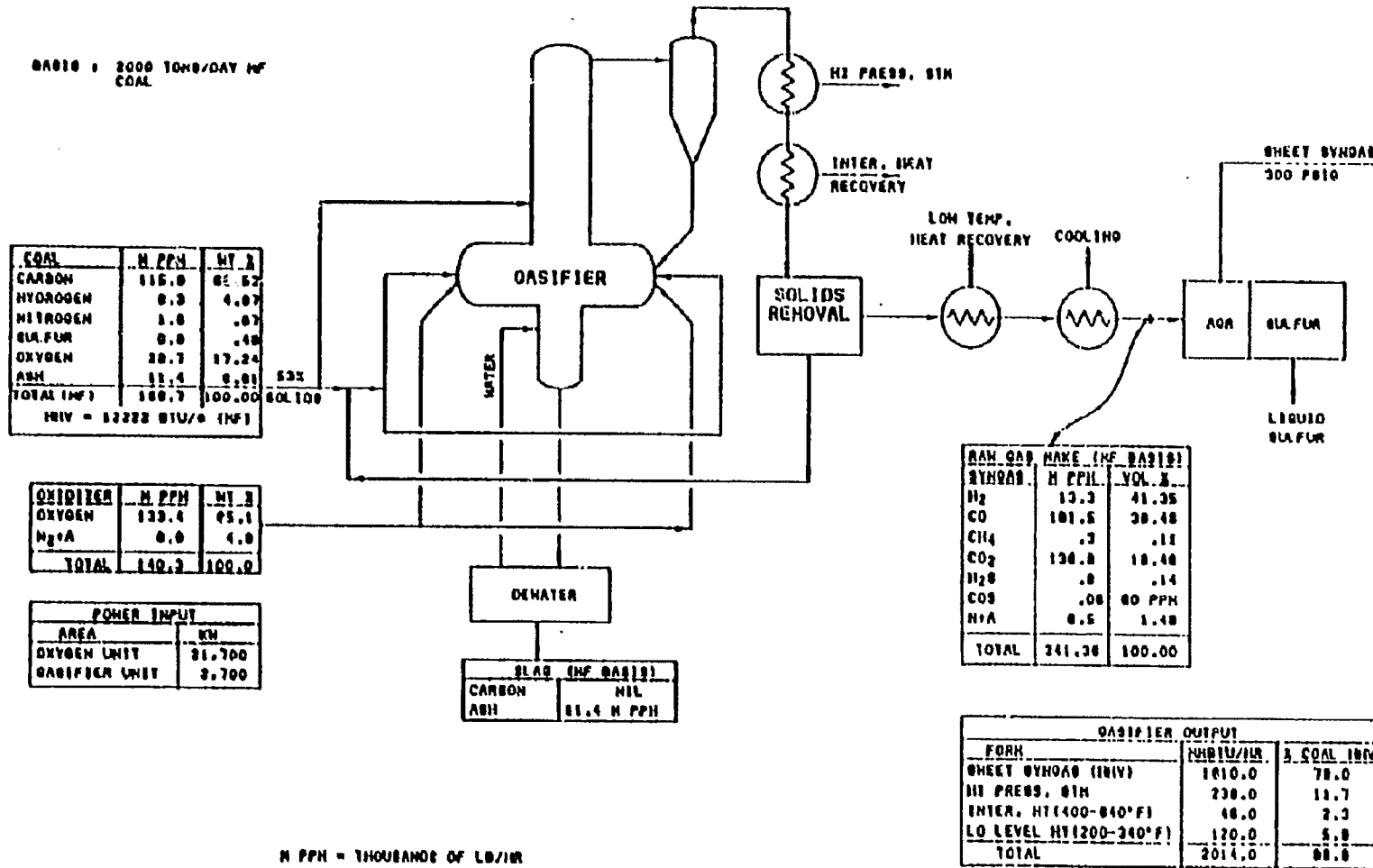The sub-bituminous coal (or lignite) is received in railroad hopper
cars, unloaded and then ground and slurried with water recycled from the
synthetic gas cleanup process. The slurry is then transferred to a
slurry storage tank to provide hold-up for transfer to the gasification
area.
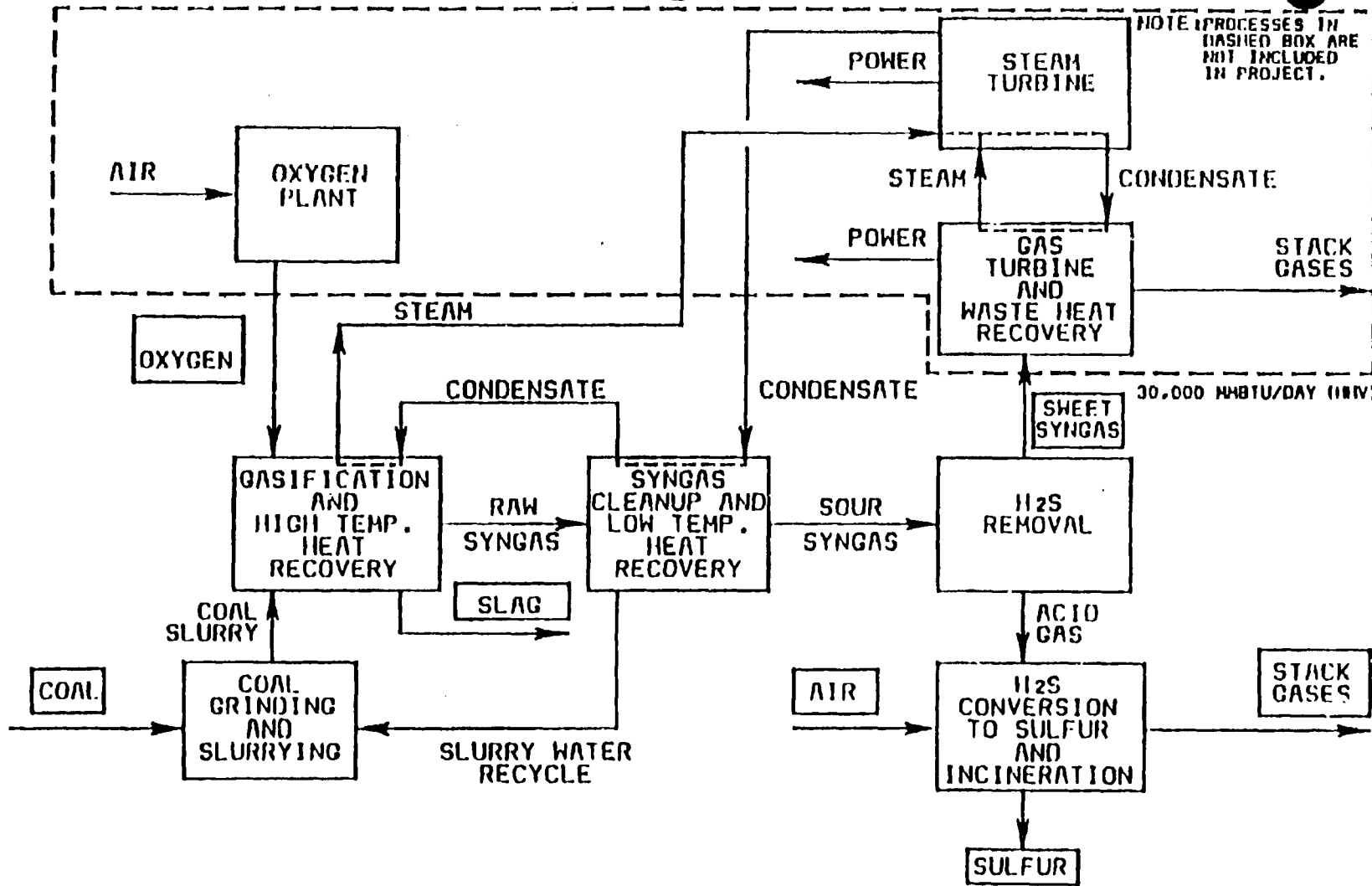
DOW SYNGAS PROJECT BLOCK DIAGRAM
FIGURE 2

FIGURE 3

DOW COAL GASIFICATION PROCESS TYPICAL MATERIAL BALANCE
WESTERN SUBBITUMINOUS COAL

BASIS : 2000 TONS/DAY MF
COAL

COAL	M PPH	WT X
CARBON	101.5	61.01
HYDROGEN	7.5	4.48
NITROGEN	1.8	1.07
SULFUR	1.8	1.13
OXYGEN	26.4	15.81
ASH	27.8	16.58
TOTAL (MF)	166.7	100.00

HHV = 10465 BTU/LB (MF)

OXIDIZER	M PPH	WT X
OXYGEN	136.0	95.1
H2O+A	7.0	4.8
TOTAL	142.8	100.0

POWER INPUT	
AREA	KW
OXYGEN UNIT	22.100
GASIFIER UNIT	3.000

SLAG (MF BASIS)	
CARBON	NIL
ASH	27.8 M PPH

GASIFIER

WATER

SOL
SOLIDS
REMOVAL

DEWATER

HI PRESS. STM

INTER. HEAT
RECOVERY

LOW TEMP.
HEAT RECOVERY

COOLING

SWEET SYNGAS
300 PSIG

AOR   SULFUR

LIQUID
SULFUR

RAW GAS MAKE (MF BASIS)		
SYNGAS	M PPH	VOL X
H_2	11.1	38.82
CO	134.7	33.74
CH_4	.1	.04
CO_2	158.5	25.28
H_2O	1.8	.204
COS	.1	.018
N+A	8.8	1.82
TOTAL	315.2	100.0

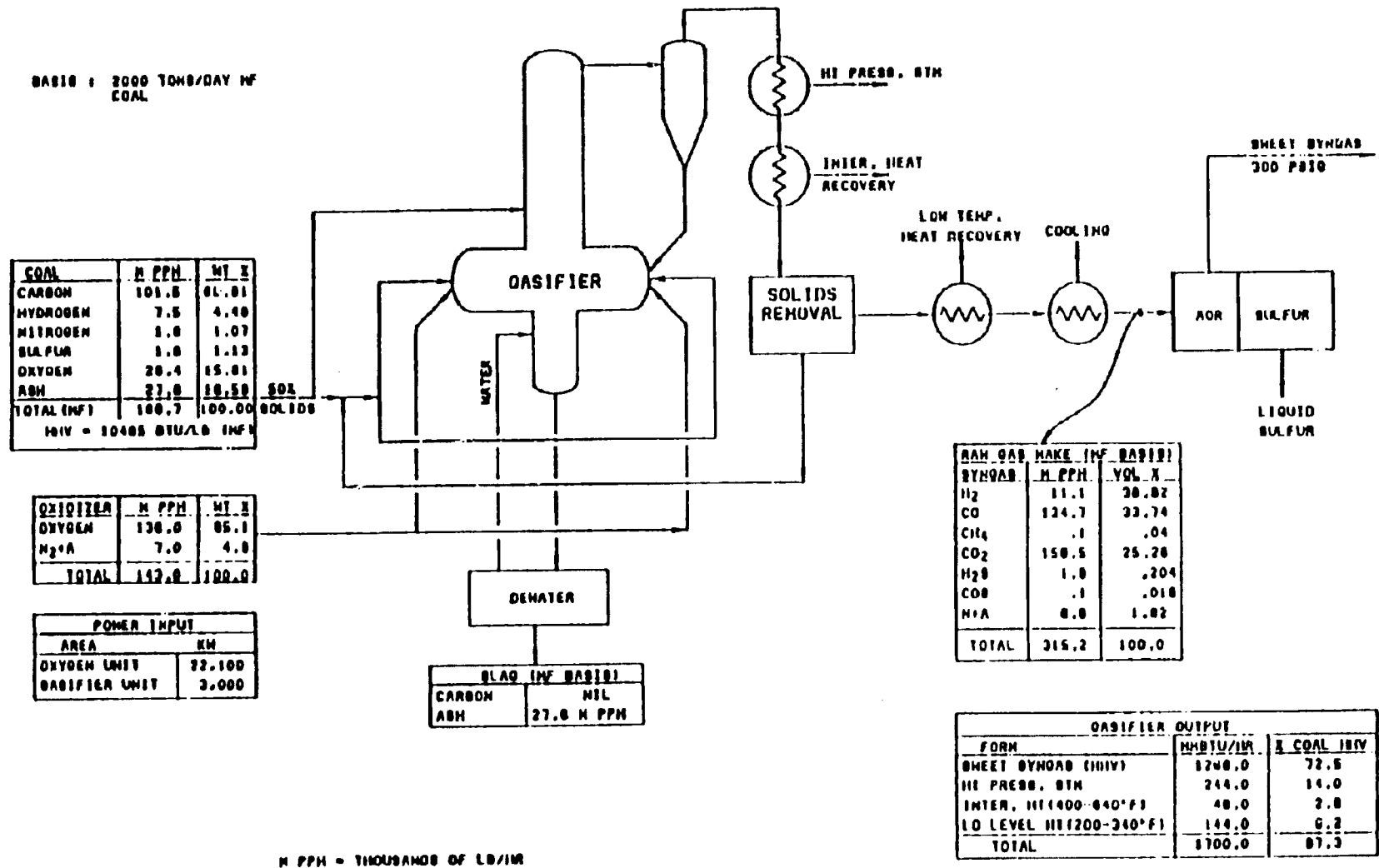| GASIFIER OUTPUT		
FORM	MMBTU/HR	X COAL HHV
SWEET SYNGAS (HHV)	1248.0	72.5
HI PRESS. STM	244.0	14.0
INTER. HT (400-640°F)	48.0	2.8
LO LEVEL HT (200-340°F)	144.0	6.2
TOTAL	1700.0	87.3

M PPH = THOUSANDS OF LB/HR

FIGURE 3A

DOW COAL GASIFICATION PROCESS TYPICAL MATERIAL BALANCE
LIGNITE

9-86

Gasification and High Temperature Heat Recovery

The ground coal slurry is pumped to a slurry feed tank in the gasification area. Provisions are made for back flushing lines and recirculating coal slurry to prevent plugging. The slurry is pumped to he slurry preheaters, where it is heated to within 50 to 100 degrees Fahrenheit of the boiling point of the slurry at the reactor pressure. Positive displacement pumps capable of handling liquid-solid suspensions at high pressure are used to control the slurry feed rate to the preheaters. After the preheating, the slurry is fed to the reactor where it is mixed with oxygen in the burner nozzles. The feed rate of oxygen is carefully controlled to maintain the reactor temperatures in a specific range depending on the properties of the coal. Under these conditions, the coal is almost totally gasified by partial combustion to produce synthetic gas consisting principally of hydrogen, carbon monoxide, carbon dioxide, and water. The sulfur in the sub-bituminous coal (or lignite) is converted almost totally to hydrogen sulfide with small amounts of carbonyl sulfide. The ash is fused in the flame, and the molten ash is drained from the bottom of the gasifier into a water quench. The slag is withdrawn continuously as a slurry through grinders and a pressure letdown system. The hot synthetic gas is cooled in an integral heat recovery system to about 1800 degrees Fahrenheit. This 2nd stage is unique to the Dow Coal Gasification Process. The hot gases leaving the 1st stage are cooled by additional slurry being introduced into the 2nd stage. The benefit being that more of the energy available in the coal is converted to chemical energy to be released in the gas turbine rather than recovered by radiant boilers from the hot gas. The raw gas is then passed through a cyclone separator to effect a separation of the entrained particles. Dow has included in the Project a spare gasifier, cyclone and slag crusher in order to help insure the projected 85% availability.

Gasification and High Temperature Heat Recovery cont'd.

A high temperature heat recovery train consisting of steam boilers and a steam superheater generates steam for use in the Dow steam system. The synthetic gas is cooled to within 50 degrees Fahrenheit of the condensation temperature to prevent condensation in the boiler.

Synthetic Gas Cleanup and Low Temperature Heat Recovery

After the synthetic gas is cooled to near its condensation point, the synthetic gas is fed to a wet particulate scrubber. The wet scrubber is operated at the boiling point of the recirculating water, and the dilute slurry produced is concentrated and blended with the reactor feed stream. Water collected from other parts of the process is used in the wet scrubber.

The scrubbed synthetic gas is then cooled through a series of heat exchangers to about 120 degrees Fahrenheit prior to H2S removal. Water condensed from the synthetic gas as it is cooled is recycled to the process and slurry unit after removal of NH3, H2S and other soluble gases.

H2S Removal

H2S is removed from the synthetic gas in the GAS/SPEC ST-1 process licensed from Dow. The sweetened synthetic gas is suitable for fuel for the gas turbine power generation system. The separated acid gas, consisting of H2S, CO2 and water, is suitable for feed to the sulfur recovery unit.

Sulfur Recovery and Incineration

Sulfur is recovered from the acid gas produced in the H2S removal
section. The Selectox Process, licensed from Union Oil Company of
California through the Ralph M. Parsons Company, is being used. The
acid gas is preheated, mixed with a controlled flow of air and fed to
the catalytic reactor, which partially oxidizes the H2S to sulfur and
water. The effluent gas from the reactor is cooled to condense the
sulfur. The Tail-gas is fed to an incinerator which burns the remaining
H2S to SO2 and vents it to the atmosphere.

Small amounts of acid gas collected from the various points in the
process, where dissolved gases are flashed from recirculated water, are
also disposed of through the incinerator. The sulfur removal and
recovery process meet existing environmental regulations.

Combined Cycle Plant

The sweetened syngas is pipelined approximately 2000 feet to Dow's
existing gas turbines which have been modified to accept the medium Btu
synthetic gas. Two Westinghouse 501D5 gas turbines have been modified
to accept 100% Syngas. This provides flexibility of operations for the
Power Plant and there is no sacrifice in heat rate when running on 100%
natural gas. These gas turbines have a capacity of 110 MW each.

Oxygen Plant

Oxygen is being supplied to the Project by Air Products Corporation. The oxygen plant is co-located with Air Products' existing hydrogen plant, adjacent to the Dow Louisiana Division. The Project built certain minimum facilities needed to connect the co-located oxygen plant. At capacity, the Project receives about 1500 tons per day of oxygen.

Project Plan

A history of the implementation of the project is shown on Figure 4. At the date of this paper's submittal the project is in the process of refractory curing in preparation for coal feed the first week of April, 1987.

Commercial Application

The Dow Syngas Project is the largest Gasification Combined Cycle power plant in operation. This plant will demonstrate the reliability and energy efficiency of the two stage Dow Coal Gasification process. Capital, operating and maintenance costs are being confirmed and the result is a system that reflects Dow's owner/operator perspective. Capital Costs have already been confirmed and are reflected in the following projection of a generic base load power plant constructed on the Gulf Coast. (Figure 5).

	1983	1984	1985	1986	1987	88	89	90	91	
	1 2 3 4	1 2 3 4	1 2 3 4	1 2 3 4	1 2 3 4					

DEVELOPMENT
PILOT PLANT — OPER
PROTO PLANT AIR — CONST / OPER
PROTO PLANT O₂ — CONST / OPER
PROCESS ENG.

GASIFICATION — DESIGN; INQUIRY PROCUREMENT; CONST — (DESIGN, PROCUREMENT, CONSTRUCTION)

FEED PREP. — DESIGN; PROCUREMENT; CONST

OSBL SERVICES — DESIGN; PROCUREMENT; CONST

TURBINE CONVERSION — DESIGN; PROCUREMENT; CONST

OXYGEN PLANT — PRELIMINARY DESIGN; DETAIL DESIGN; PROCUREMENT; CONST & START-UP

PLANT START-UP — TEST; TRAIN

FIGURE 4

FIGURE 5

1392 MW IGCC PLANT UTILIZING THE DOW COAL GASIFICATION PROCESS
TYPICAL LIGNITE
CAPITAL COST

Capacity

GT 6 @ 145 MW	870 MW
ST 2 @ 361.5 MW	722 MW
Total Generation	1592 MW
Gasification Use	- 23 MW
Oxygen Use	-162 MW
Auxiliary Use	- 15 MW
Net Power Generated	1392 MW

Capital	MM 86$	86$/KW
Gasification Units	341	245
Oxygen Units	147	106
Combined Cycle Units	482	346
Site Facilities	140	101
TOTAL	1110	798

Net System Heat Rate (HHV):   9418 BTU/KWH

Basis:

1. 1986$

2. No interest during construction.

3. Contingency included is approximately 10% of the Total Capital.

4. Engineering, procurement and contract administration are included at 9.3% of total direct cost.

5. "Owner's Costs" are not included as they can be variable.

6. Cost estimate is the type that Dow would do for presentation to its Board for capital authorization.

7. P&ID's, Instrument Lists and Equipment Designs were taken from the LGTI project where recent actual purchases were made.

MONITORING PROGRAMS FOR UNIT OPERATIONAL PROBLEMS

DUE TO

HIGHLY FLUCTUATING LIGNITE QUALITY

By Dan Andrew, Eric Darmstaedter and John Vidovich

GUS, Inc.

Dallas, Texas

ABSTRACT

Mine mouth power generating stations burning Texas lignite can experience operational problems due to periods of widely fluctuating lignite quality. Intermittent problems can occur in the areas of dust generation, lignite handling system pluggage, slagging, and flyash collection.

Lignite quality changes affecting operations in these areas can be grouped into primary and secondary categories. Field quality monitoring equipment is available that can measure these quality changes, usually in less than one hour. Monitoring equipment results can then be correlated with existing operations information. General lignite quality range limits can be identified which predict the onset of operational problems.

With the additional capabilities of programmable logic controllers and PC data acquisition systems, diagnostic and long term monitoring programs can be run effectively.

INTRODUCTION

This paper examines the problem of fuel quality fluctuation at lignite mine-mouth power generation sites from an operational viewpoint. The purpose of this paper is to review various field monitoring equipment, procedures, and programs that can provide operations personnel with early warning signals concerning problem lignite. With this knowledge, modified operational procedures can be instituted to control impending problems before they occur.

Information is based on research, field trials, and existing monitoring programs at several Texas lignite fired mine mouth power sites from 1983 to 1987. Mining areas are typically 4-12 miles from plant sites. Dedicated rail or haul road systems connect each plant with its respective mining areas. Between 15,000 and 40,000 tons per day are mined, delivered and burned at each site.

Although these generating units were originally designed with acceptable margins for lignite quality variations, 10% - 15% of unit operating time is far enough outside design margins to result in fuel cycle problems. Fluctuations in primary qualities of moisture, ash, sulfur, silica, calcium and particle size are common. These variations can be due to natural seam formations, mining delivery and storage procedures, and/or weather patterns. Combinations of these changing parameters result in wide variations in secondary qualities such as bulk shear strength, dustiness, ash softening temperature, and flyash resistivity. Resulting problems occur in the areas of dust generation, lignite system pluggage, furnace slagging and/or flyash collection.

Monitoring procedures have been set up at some locations to help predict these fuel quality related problems. Some procedures simply involve scheduled analysis of a combination of available operating parameters. Other procedures involve field analysis equipment capable of results readout of primary and secondary qualities in one hour or less.

Moisture analyzers, x-ray diffraction analyzers, bulk coal shear testers, real time dust monitors, and resistivity probes are a few examples of the equipment discussed in this paper.

Standard operations data and signal outputs from these monitoring tools can be tied together in various combinations through use of programmable logic controllers and PC data acquisition systems. The correct mix can result in a useful diagnostic program and/or a full scale continuous monitoring program. If data is also logged in a data base format, relationships between monitored parameters and operational problems can be continuously refined and improved upon. Eventually, the information loop can be closed by directly controlling problem solving equipment.

LIGNITE QUALITY FLUCTUATIONS

- **Overview**

    Operations personnel are concerned with lignite quality fluctuation at the generating plant. Three major factors influence these fluctuations at the mine mouth sites described in this paper.

    - Variations in lignite deposit quality
    - Mining, delivery and storage procedures
    - Weather patterns

    When variations in all three areas are combined, extensive fluctuation in as-received lignite quality occurs.

    The lignite quality parameters of interest in this paper are divided into the primary and secondary qualities shown below.

    Primary — Silica, Calcium, Sulfur, Ash, Moisture, Particle size

    Secondary — Dustiness, bulk shear strength, ash fusion and flow temperatures, ash resistivity

    Certain combinations of primary quality variations can result in multiplicative fluctuations in secondary qualities. This paper addresses the periods when secondary qualities have varied enough to result in operational problems.

- **Variations in Lignite Deposit Quality**

    Lignite deposit quality can fluctuate widely from seam to seam, pit to pit, and within a pit. The pit core analysis in Figure 1 shows primary qualities of sulfur and ash (dry basis) varying by factors of 2.2 and 3.2 respectively. Pit variations in higher quality deposits are somewhat less.

PIT CORE   FIGURE 1

INTERVAL FROM - TO	FT	ASH	SULFUR	BTU
63.0 - 63.5	0.5	70.94	0.71	2574
63.5 - 66.0	2.5	36.52	4.17	8160
66.0 - 66.5	0.5	76.79	1.53	1505
(3.6 ft parting but only top and bottom 0.5 analyzed)				
69.1 - 69.6	0.5	78.59	1.88	1504
69.6 - 74.0	4.4	16.45	2.34	10736
74.0 - 74.9	0.9	66.34	0.63	2872
74.9 - 78.6	3.7	26.70	1.48	9027
78.6 - 79.4	0.8	46.18	1.07	6207
79.4 - 84.0	4.6	16.25	1.31	10659
84.0 - 84.5	0.5	72.34	0.32	2320
(5.0 ft parting cored 0.5 ft top and 0.5 ft bottom)				
88.5 - 89.0	0.5	78.29	0.54	1484
89.0 - 90.0	1.0	20.79	1.99	10145

The thin partings shown may be blackjack, clays, sandstone or other contaminants, depending on lignite deposit quality. These partings are usually high in ash, primarily containing aluminum and silica. Based on parting thickness, this material may be mined as part of the lignite seam or removed as it is encountered. The additional variable of partings being present or absent increases fluctuations in primary lignite qualities as lignite is transported out of each pit.

Mining, Delivery, and Storage

Typically, each plant site is fed simultaneously from several mining areas. Each mining area may include several pits. Lignite from different pits can be combined differently as it is transported by haulers, rail cars and/or conveyors to the plant site. Planned blending operations and delivery schedules can reduce the number of pit combinations possible. However, transport equipment outages and other unexpected mining problems minimize the chances of continuous known pit blends reaching the plant site.

Once on site, lignite can normally be directed to four general areas; the unit, short term outside storage, short term storage silos, or dead storage. Dead storage construction procedures add another source of lignite quality fluctuation. Dead storage piles are typically built by compacting lignite in layers and wetting down each layer as it is built. This helps structurally stabilize the pile for traffic and reduces spontaneous combustion problems. It also substantially changes the primary qualities of moisture and particle size.

As dead storage is reclaimed and delivered to the units alternately with short term storage or run of mine lignite, secondary qualities of dustiness and bulk shear strength may fluctuate widely.

Weather Patterns

Texas lignite mine mouth sites are subjected to sudden variations in weather. The extremes of above 100°F temperatures, summer direct sunlight and twenty four hour rainstorms in excess of three inches cause the majority of problems.

Lignite in hot dry air can spall to fine particle size as it loses moisture. A lignite chunk can easily change from 35% moisture to 25% moisture in less than a week of hot dry sunny weather.

Rainstorms affect lignite quality variation in two ways. A twenty four hour, three inch rainstorm can add 4% to 6% surface moisture to the top layer of outside storage piles.

If rainstorms are severe, mining operations can be inter-
rupted due to temporarily flooded pits or impassable ramps.
This situation forces changes in the source of lignite
delivered to the unit, contributing further to lignite
quality fluctuation.

Lignite Quality Fluctuation at the Plant Site

Figures 2 and 3 represent lignite quality variation measured
after secondary crushers. A wide range of lignite quality
variation during two different time periods is shown in
Figure 2. A significant shift in the ranges occurred over a
thirty year period due to the opening of additional mining
areas.

PLANT SITE LIGNITE QUALITY VARIATION
SITE 1

FIGURE 2

PRIMARY LIGNITE QUALITY	1955 RANGE	1981-1983 RANGE
S (%)	1.6 - 2.0	1.7 - 2.5
Ash (%)	14.5 - 18.1	10.8 - 23.4
SiO_2 (%)	8.4 - 28.9	56.6 - 65.2
Al_2O_3 (%)	9.3 - 25.1	10.3 - 21.8
Fe_2O_3 (%)	9.0 - 13.9	2.9 - 8.7
CaO (%)	13.8 - 30.3	7.5 - 16.2
MgO (%)	3.0 - 12.6	1.9 - 3.0
SO_3 (%)	17.9 - 24.1	2.4 - 3.0

Primary lignite quality variations at site 2 are shown for
two separate years during a ten year period in Figure 3. The
two ranges are similar, however they cannot be directly
compared since the 1982 data shows only variations to two
standard deviations for moisture and ash.

PLANT SITE LIGNITE QUALITY VARIATION
SITE 2   FIGURE 3

	TOTAL RANGE (1972)	RANGE TO TWO STD. DEVIATIONS (1982)
Moisture (%)	22.9 - 38.1	34.1 - 34.7
Ash (%)	4.3 - 13.2	8.8 - 14.7
Sulfur (%)	0.5 - 3.6	0.44- 3.5*

* Total Range

The effect of the three major influencing factors discussed
in the first section can also be seen in day to day lignite
quality fluctuation. Primary qualities of moisture, ash and
sulfur (dry basis) are graphed for a typical ten day period
at site 2 in Figure 4. Data is from daily samples routinely
collected from tripper belts.

SITE 2
MONTHLY REPORT FOR UNIT #1
JULY

FIGURE 4
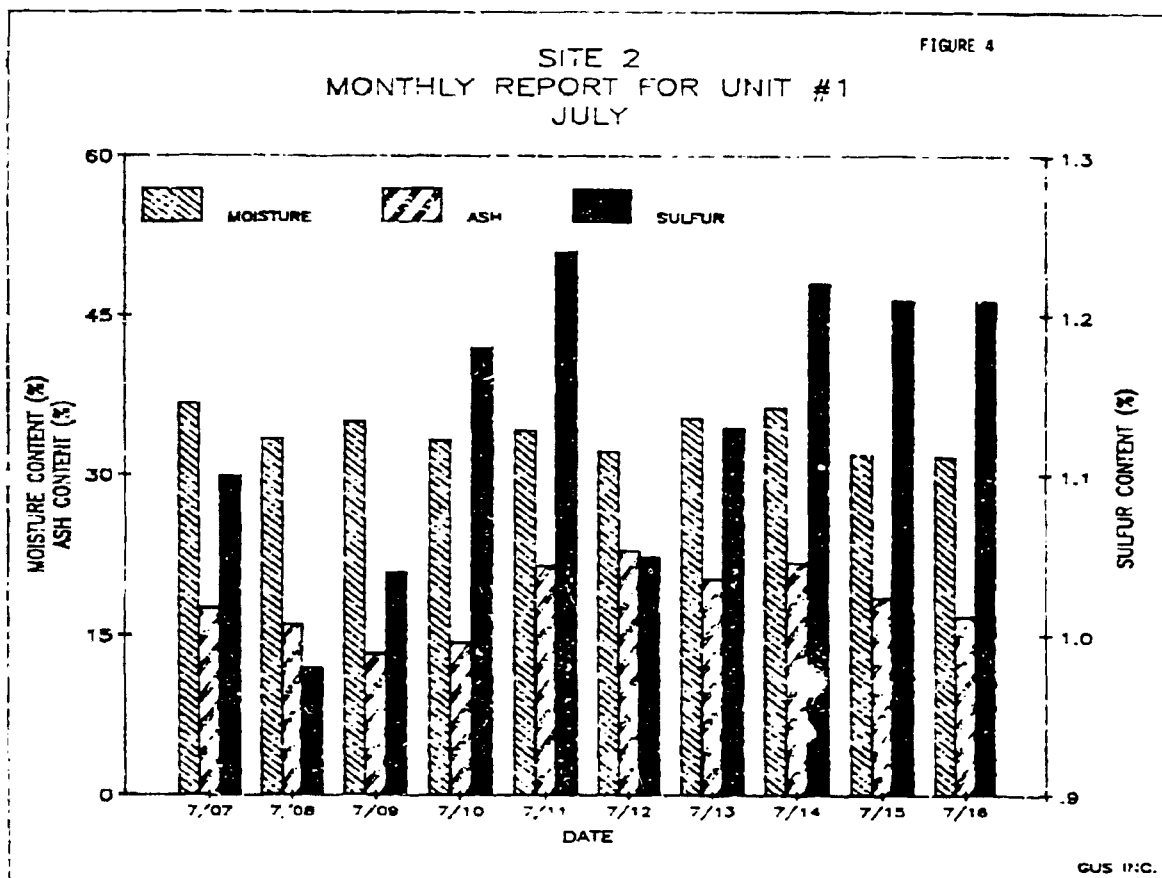
Lignite quality fluctuations that cause operational problems are usually the result of two or more primary quality change's occurring in the wrong direction simultaneously. The difference in moisture between July 14th and July 16th (36.4-31.8) together with the change in ash content (22.0-16.8) can result in an extensive shift in bulk shear strength.

The difference in sulfur content between July 12th and July 16th (1.05-1.21) together with the change in ash content (23.0-16.8) can cause significant changes in ash resistivity. Combinations of primary lignite qualities such as these result in secondary quality fluctuations that are of concern to operations personnel.

LIGNITE MONITORING PROCEDURES

Overview

From an operational viewpoint it is beneficial to know when changes in secondary lignite quality parameters are occurring. This can be accomplished by monitoring related primary lignite qualities or directly measuring secondary qualities. General relationships between primary and secondary qualities are shown in Figure 5, along with field monitoring equipment used to measure both. This section describes each monitoring instrument. Examples are given to show how equipment can be used to diagnose and predict operational problems due to fluctuating lignite quality.

MONITORING LIGNITE QUALITY
FOR OPERATIONS   FIGURE 5

LIGNITE		OPERATIONS	
PRIMARY QUALITY	SECONDARY QUALITY	MONITORING EQUIPMENT	OPERATIONAL SIGNALS
Ash, Silica, Moisture	Bulk Shear Strength	A, B, C	A, B, D
Moisture, Particle Size	Dustiness	C, D	A, D, C
Ash, Silica, Calcium, Sulfur	Ash softening temperature or ash flow point temperature	A, E	E, F
Ash, Sulfur	Resistivity	A, F	E, F, G

MONITORING EQUIPMENT

A. X-ray fluorescence
B. Rotational sheartester
C. Microwave moisture analyzer
D. Real Time Aerosol monitor
E. Optical pyrometer
F. Resistivity probe

OPERATIONAL INFORMATION

A. Storage pile source
B. Coal on/off belts
C. Surge bin levels
D. Chute buildups
E. Excess O_2, MW, TPH, Burnrate
F. Exit gas temperature, ESP resistance factors, SO_2 inlet ppm
G. Opacity

X-Ray Fluorescence Lignite Analyzer

The x-ray fluorescence analyzer is used to monitor primary lignite qualities of ash, sulfur, silica and calcium. As a portable unit, it is capable of operating on battery or AC power and is well suited for field use. After establishing a model by initial calibration with known samples, the analyzer provides accurate results within five minutes of sample preparation.

The unit operates by exciting the elements within the lignite sample using a radioactive source. Emissions from the lignite are counted and results are displayed on an LCD.

The analyzer has proven to be valuable in determining lignite properties because of its short turnaround time. As shown in Figure 6, ash and silica values can be plotted with lignite moisture levels and bulk shear strengths (FFV) as often as once every sixty minutes. This information is used to predict when buildups and/or pluggage will occur in the lignite handling system.

LIGNITE FLOWABILITY EVALUATION   FIGURE 6
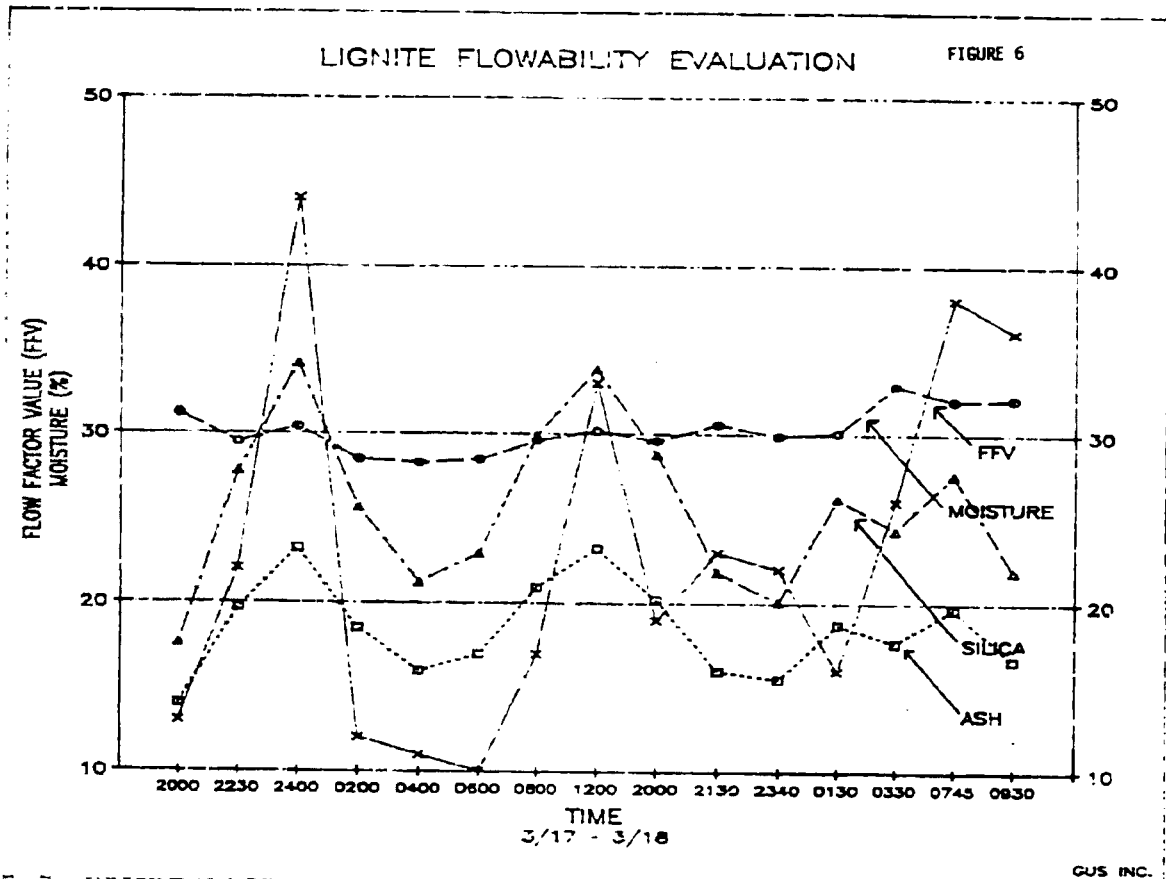
Rotational Bulk Shear Tester

Two major types of bulk shear equipment are in use today –
rotational and linear translational. Due to its ease of
operation and relatively short run time, the rotational shear
tester can be used for field measurement.

The basic function of the rotational sheartester is to set up
an equally distributed shearing force across a horizontal
plane in a lignite sample. This is done while the sample is
placed under varied vertical consolidation loads. As the
sample is twisted (or sheared), the force is transmitted
through the sample via frictional forces to a load cell force
measurement device. Information is then tabulated, processed
(manually or with a computer program) and plotted on a Mohr's
circle analysis graph. Unconfined yield strength and
consolidation pressure are read from the graph. Several
Mohr's circles will define a flow function line. For a given
handling system, a bulk shear strength can be determined and
represented in lbs/ft^2 or an arbitrary relative flow factor
value (FFV).

The entire sampling test time is 30–45 minutes. As shown in
Figure 6, the flow factor value (FFV) is plotted with lignite
ash, moisture and silica values. These primary lignite
qualities can combine to produce significant variations in
the FFV, especially when all qualities fluctuate in the same
direction. The usefulness of this information has been
previously mentioned.

Microwave Lignite Moisture Analyzer

Lignite moisture levels can fluctuate significantly over short periods of time due to influences from all three major contributing factors mentioned in Section 1. Since the minus 8 mesh fraction contributes more to dust and/or bulk shear strength problems, moisture is monitored for this fraction more frequently. The microwave moisture analyzer results are available within twenty minutes of sample preparation.

The analyzer functions by placing a three to four gram, minus 8 mesh lignite sample on a scale within the oven. After the drying cycle is completed, the weight after drying is compared to the initial weight and the water loss is displayed in terms of percent moisture. All weight measurements and calculations are done within the moisture analyzer information processing unit.

Figure 7 shows how moisture fluctuations can be compared to other primary and secondary lignite qualities to monitor material handling problems throughout the handling system. Because of the rapid response time, several lignite sources can be monitored each hour.
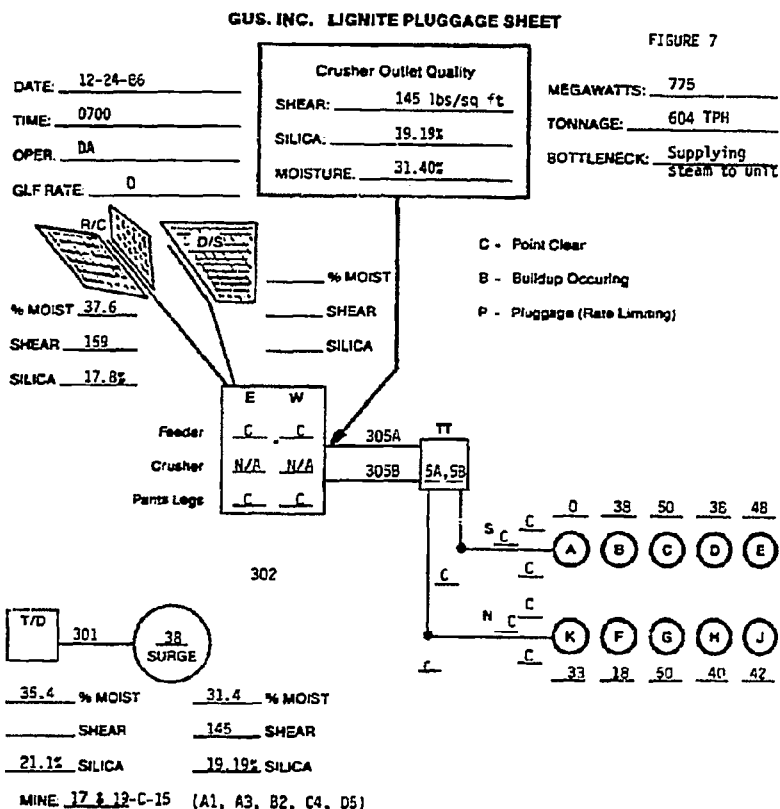
GUS. INC.   LIGNITE PLUGGAGE SHEET

FIGURE 7

DATE: 12-24-86

TIME: 0700

OPER. DA

GLF RATE: 0

Crusher Outlet Quality

SHEAR: 145 lbs/sq ft

SILICA: 19.19%

MOISTURE: 31.40%

MEGAWATTS: 775

TONNAGE: 604 TPH

BOTTLENECK: Supplying steam to unit

R/C

D/S

% MOIST

SHEAR

SILICA

C - Point Clear

B - Buildup Occuring

P - Pluggage (Rate Limiting)

% MOIST 37.6

SHEAR 159

SILICA 17.8%

E   W

Feeder   C   C   305A   TT

Crusher   N/A N/A 305B   5A,5B

Pants Legs C   C

302

S C C   0   38   50   38   48
      C   (A)(B)(C)(D)(E)

C

N C C
      C   (K)(F)(G)(H)(J)
C   33   18   50   40   42

T/D 301   38 SURGE

35.4 % MOIST   31.4 % MOIST

SHEAR   145 SHEAR

21.1% SILICA   19.19% SILICA

MINE: 17 & 13-C-15 (A1, A3, B2, C4, D5)

COMMENTS: Small buildups in chutes of east and west feeder pans of transfer house. Should not cause problems. All points checked at 1930.

Pluggage Detectors

The buildup and pluggage data in Figure 7 was logged manually by operations personnel routinely inspecting the handling system. Automatic pluggage detectors can sense when critical chute buildup levels are reached and send an alarm signal to recording instrumentation. This equipment can reduce the need for extra monitoring manpower during periods of fluctuating lignite bulk shear strength.

Pluggage detectors discussed in this paper function by sending a narrow, low level radiation beam across lignite handling chute work perpendicular to lignite flow. The signal is attenuated by the amount of lignite buildup along the inside of the chute. Critical levels are calibrated and preset.

If equipment of this type is integrated with centralized recording instrumentation, the complete handling system can be monitored for buildup on a real time basis. Manpower time and safety concerns can be reduced.

Real Time Aerosol Monitor

The real time aerosol monitor directly measures ambient particulate levels in terms of mg/m^3. It is more efficient to monitor dust directly than to monitor the primary lignite qualities of moisture and particle size.

The monitor functions by drawing a particulate laden air stream through a pulsed light emitting diode. The scattered radiation measured is proportional to the dust concentration. A signal output is provided for recording purposes.

The aerosol monitor can be used as a portable unit for spot checking areas of concern or as a stationary unit. The aerosol monitor output in Figure 8 is driving a strip chart recorder. Additional signals from feeders, conveyors and a surge silo are being monitored simultaneously according to procedures described in Figure 5. In this example, a real time aerosol monitor is being used to determine which dust problems are caused by lignite quality fluctuation and which are caused by operational changes.

RAM DUST SIGNAL (0-20 mg/m³)

FEEDER 6 STEP SIGNAL
Showing feeder on/off,
tripper positions over
boiler bins

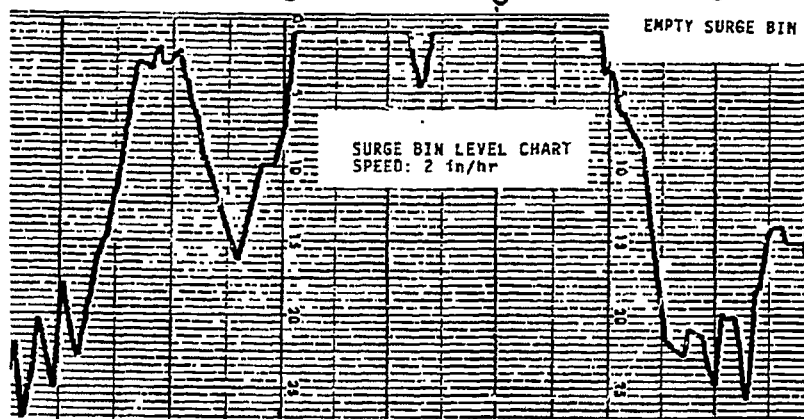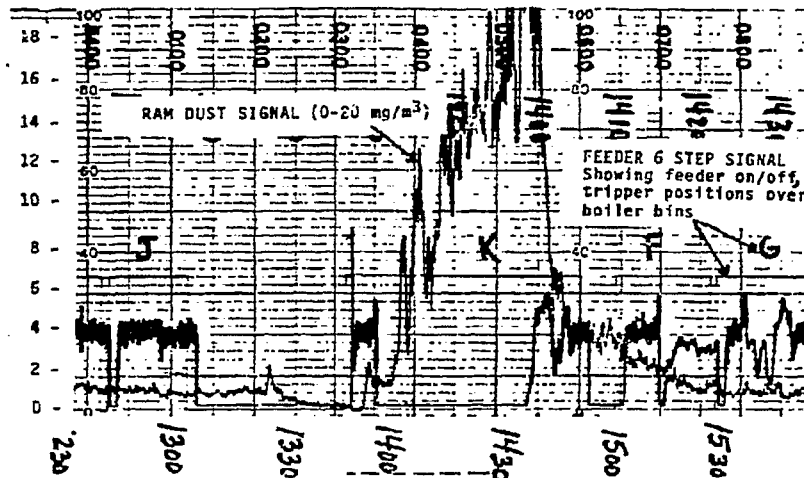EMPTY SURGE BIN

SURGE BIN LEVEL CHART
SPEED: 2 in/hr

CHART 5

DATE: 8/02/85
OPERATING CONDITIONS: Tripper - Reclaiming lignite through an
empty surge bin

PARAMETERS:
GDS-08 @ 0.0025 gpt
Rotoclones on
ROM by train
Using belt 5B
Fuel moisture 32.6%
Handivol Samples N/A

- **High Volume Air Sampler**

This method of measuring ambient lignite dust concentrations
has its main application in outdoor situations or where long
term single sample testing is required.

The method of operation is to draw a know volume of particu-
late laden air through a filter of known weight. At a pre-
determined point, the filter is re-weighed and the particu-
late concentration is calculated.

Several drawbacks to using this measurement device for opera-
tional monitoring include inability to distinguish high and
low dusting conditions during a single sampling run, diffi-
culty in obtaining usable signal outputs and long test turn-
around times. For these reasons, the high volume unit is
seldom used for diagnostic testing of fluctuating lignite
dusting.

Optical Pyrometer

An optical pyrometer is used to measure the temperature of slag deposits inside the furnace. The optical pyrometer is a hand held, battery operated device used to measure slag or furnace wall temperatures in areas not hospitable to other types of measurement. The pyrometer reads the infrared or optical radiation generated by the hot surface and is adjusted to the material via an emissivity value. The emissivity value is a given for the material (slag, tube surface, etc.) being measured in a given temperature range and surface condition.

Pyrometer information, coupled with visual observations of slag quality, can be used to generally estimate secondary lignite qualities of ash softening and ash flow temperatures. However, during combustion, slagging characteristics of lignite are dependent upon operational parameters as well as lignite quality fluctuation.

It is important when monitoring for slagging problems to correctly combine information from x-ray fluorescence, optical pyrometry, slag survey logs, empirical correlations, and unit operations information before reaching predictive conclusions.
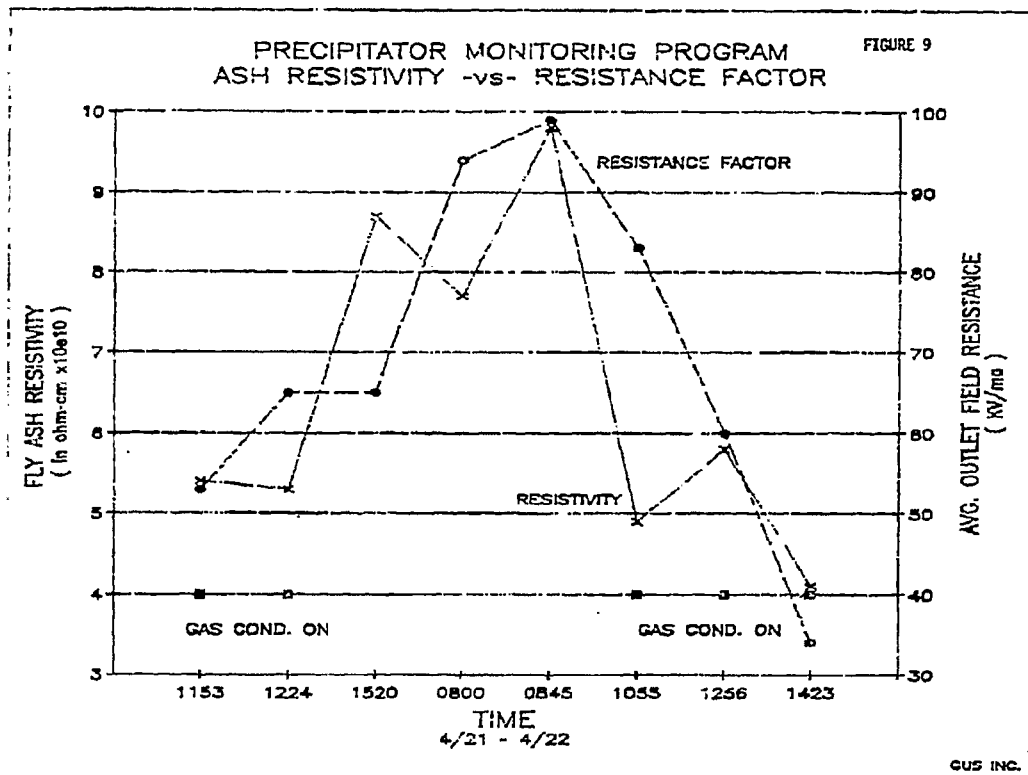
In-situ Resistivity Probe

Flyash resistivity is a secondary lignite quality dependent mostly upon primary qualities of sulfur, silica, calcium, alumina and sodium as well as various unit operating conditions. Flyash resistivity is a key factor in electrostatic precipitator (ESP) performance. The principle tool used to monitor changes in resistivity is the in-situ resistivity probe.

The in-situ resistivity probe operates with the collection section of the device inside the flue gas stream. A flyash sample is collected onto a disc shaped plate electrode through an electrostatic corona discharge emitted continuously from a point source. Flyash particles are charged and deposited in a method similar to that of an ESP.

The voltage versus amperage characteristics of the deposited layer are compared to the voltage versus amperage characteristics without a flyash layer present. The resistivity value is calculated by combining this information with the collected layer thickness.

This test can be extremely valuable in diagnostic programs designed to evaluate ESP performance problems under fluctuating lignite quality conditions. Ash resistivity can fluctuate significantly, as shown in Figure 9. In this example, gas conditioning is being used to artificially change ash resistivity by approximately one half order of magnitude. The ESP resistance factors, a measure of ESP operating performance, generally follow the changes in ash resistivity in this case.

PRECIPITATOR MONITORING PROGRAM — FIGURE 9
ASH RESISTIVITY -vs- RESISTANCE FACTOR

As a measurement tool, the in-situ resistivity probe can supply accurate repeatable and timely results, usually within forty five minutes. Like the optical pyrometer, however, results should be combined with primary lignite qualities and unit operating conditions, especially flue gas temperatures, before predictions on ESP performance are made.

• **ESP Resistance Factors**

ESP resistance factors are calculable values derived by dividing ESP secondary voltage by secondary amperage (kv/ma). This monitoring factor is used for trending and spot checking ESP performance changes due to flyash resistivity and/or collection plate buildups. A general rise in the resistance of all ESP sections usually signals a lignite quality change. A rise in an isolated ESP section usually signals a rapper vibrator malfunction.

Typically, resistance factors are combined with primary lignite quality analysis and in-situ ash resistivity measurement to diagnose and monitor ESP performance. Changes in lignite quality, resistivity and resistance factors usually signal an impending change in ESP performance as measured by opacity (see Figure 10). Resistance factors will trend behind resistivity values due to the time required for ESP ash collection and removal. Opacity problems will normally lag behind changes in resistance factors.

The combination of monitoring resistivity and ESP resistance factors has proven to be an accurate method of tracking and predicting lignite unit ESP performance in most cases.

FIGURE 10

PRECIPITATOR MONITORING PROGRAM
A AND B BOX ESP RESISTANCE FACTORS
DATA SUMMARY - AUGUST 1985
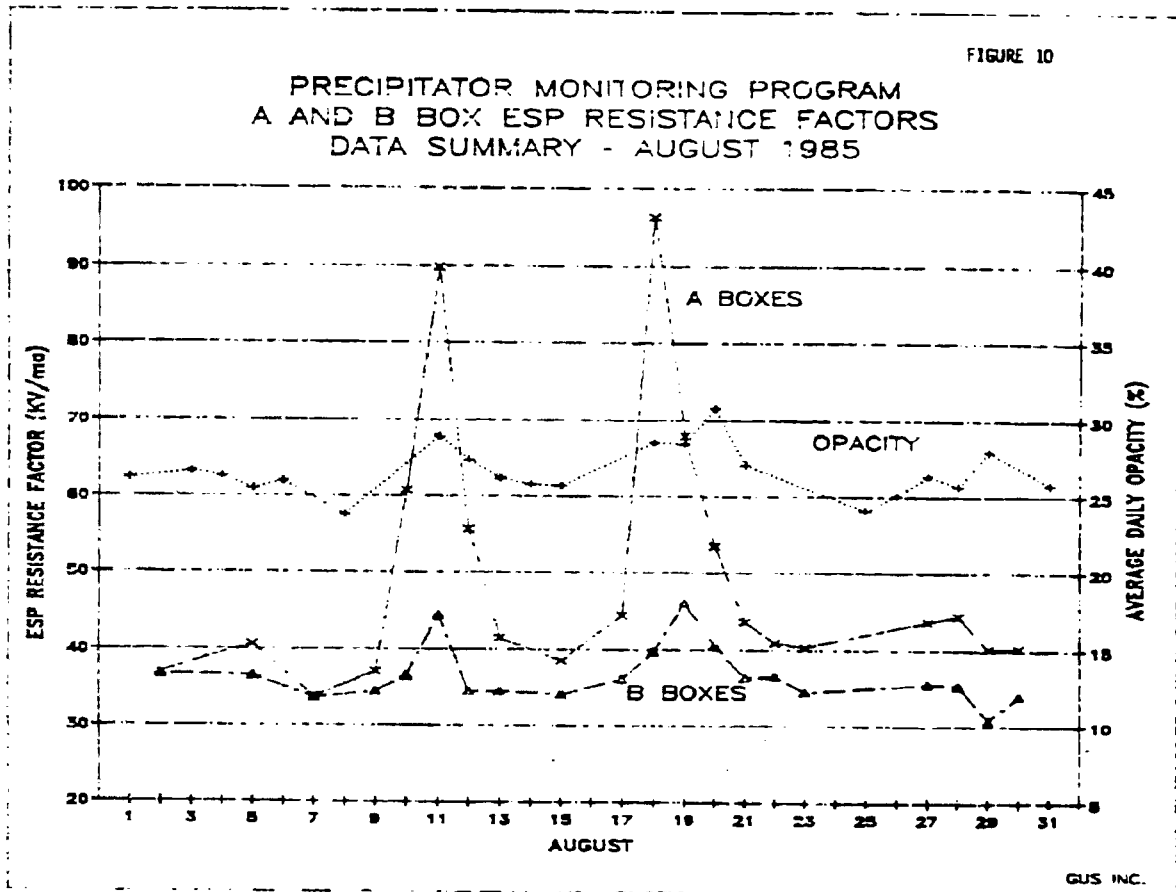
DATA ACQUISITION AND ANALYSIS

* **Overview**

    Historically, lignite plants have used various operational signal outputs to control lignite transfer from the yard to the plant. These signals combined with signal outputs from various monitoring equipment discussed in this paper can provide an engineer with enough information to solve most problems caused by lignite quality fluctuation. However, data acquisition and analysis using strip chart recorders, log sheets and visual observations can be time consuming and cumbersome.

    If equipment and/or available signals are decentralized, control of information becomes difficult and data can be lost or outdated by the time it is analyzed.

    In order to increase information processing efficiency for engineering and provide useful information for operations, additional monitoring tools are recommended. An on-line data acquisition system compatible with most personal computers can monitor and data log all program parameters simultaneously.

A programmable logic controller (PLC) can be used to access and condition all necessary signals as well as provide for flexibility when several different operational procedures must be monitored. The combination of a PLC and data acquisition system with IBM compatible PC interfacing ultimately furnishes the engineer with a powerful monitoring system.

* **Data Acquisition Method Comparison**

A simple example illustrating the comparison between the older monitoring approach and the more efficient method utilizing a PC data acquisition system is shown in Figure 8 and Figure 11 respectively. Both methods document the problem of dust generated by transporting an intermittent supply of lignite through an empty surge bin. The first method (Figure 8) was completed by comparing several strip charts, coordinating several operational signals and manually adjusting and documenting various operating parameters. Three to four days were necessary to complete data acquisition and determine preliminary correlations.

The second method employed a PLC and a PC data acquisition system. The system simultaneously gathers, interprets, displays and logs the same information collected in the first method on a real time basis. Data is automatically stored in a data base format for future correlation analysis. A simple display of several program parameters is shown in Figure 11.

SITE 2
UNIT #3 - CRUSHER OUTLET
DUST GENERATION PROGRAM
MARCH 1987   FIGURE 11

YEAR	MONTH	DATE			
1987	3	20			
			SURGEBIN	BELT	BELT
		DUST	LEVEL	305A	305B
HOUR	MINUTE	(mg/m3)	(tons)	(tph)	(tph)
14	50	2.3	0	0	0
14	51	2.3	0	0	0
14	52	4.8	0	0	0
14	53	4.6	0	10	0
14	54	4.9	13	23	0
14	55	5.3	31	0	0
14	56	6.6	52	7	0
14	57	6.0	73	0	0
14	58	7.0	94	0	0
14	59	6.6	115	2	0
15	0	6.6	136	0	0
15	1	7.5	157	21	0
15	2	7.6	173	0	189
15	3	5.6	184	0	590
15	4	3.8	177	0	884
15	5	3.0	174	0	572
15	6	2.5	183	0	610
15	7	2.3	188	0	614
15	8	1.8	178	0	536
15	9	1.8	172	0	625
15	10	1.6	177	0	883
15	11	1.6	182	0	964
15	12	1.6	184	0	943
15	13	1.4	164	0	1099
15	14	1.5	161	0	1006

Diagnostic efficiency is increased by automatic data analysis and interpretation capabilities.

Data is collected continuously and printed out every minute in this case. Data can be collected, averaged, and/or printed in whatever format the operator chooses. Because information is directly logged into a data base system, immediate trendchecks,correlationsor graphics can be displayed. These features benefit the operator by simplifying data collection, analysis and reporting functions.

SUMMARY

Fluctuations in lignite quality contribute significantly to periodic operational problems. With the use of specialized monitoring equipment, it is possible to maintain a close watch on these primary and secondary lignite qualities. A comprehensive monitoring program can be set up by combining the correct monitoring equipment with existing operational information.

This information can be used by operations personnel to rapidly understand why problems are occurring and make knowledgeable decisions on how to handle these problems. PC data acquisition systems and programmable logic controllers can be used for monitoring, display and data logging. These system enhancements can make overall monitoring programs logistically practical for operations personnel and provide a powerful analysis tool for engineers.

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REPLACEMENT OF TWO MATERIAL CONTROL SYSTEMS WITH
A NEW CENTRALIZED COLOR GRAPHICS SYSTEMS

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ABSTRACT

Utility Fuels, Inc. (UFI) owns and operates coal handling facilities that service Units 5,6,7, and 8 at Houston Lighting & Power's W. A. Parish Electric Generating Station. Nine (9) million tons of coal is transported annually by unit train from mines in Montana and Wyoming and is stacked out over Reclaim Feeders that deliver the coal by conveyor belts to the unit silos. The reclaim system feeding Units 5 and 6 (660 megawatts each) is designated as Phase I while the second system, Phase II, supplies Units 7 and 8 (560 megawatts each).

A Mimic Coal Board was located in each of HL&P's Phase I and Phase II control centers for monitoring and controlling the Phase I and Phase II Reclaim Systems. In late 1985, UFI decided to replace the Phase I system and to consolidate the Phase I and II Reclaim Control Systems in a new building.

The project consisted of replacing the control system on Phase I with Modicon Programmable Logic Controllers (PLC) and Remote Input/Output Modules similar to that existing in Phase II. Phase I and Phase II Mimic Control Boards were replaced with a computerized CRT/Color Graphics Operators Console.

In December, 1986, Phase I and II were operational and training was completed. Operations was able to move out of HL&P's two control rooms into a centralized, state-of-the-art Operations Control Center.

The early planning, designing, and implementation of this system, and its description, is the subject of this paper. The equipment was installed and tested without disruption of UFI's ability to deliver coal. From the time a Project Team was established in January, 1986, to the turning over to Operations in January, 1987, UFI never once failed to meet its coal delivery requirements to HL&P.
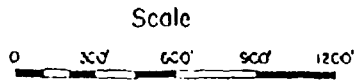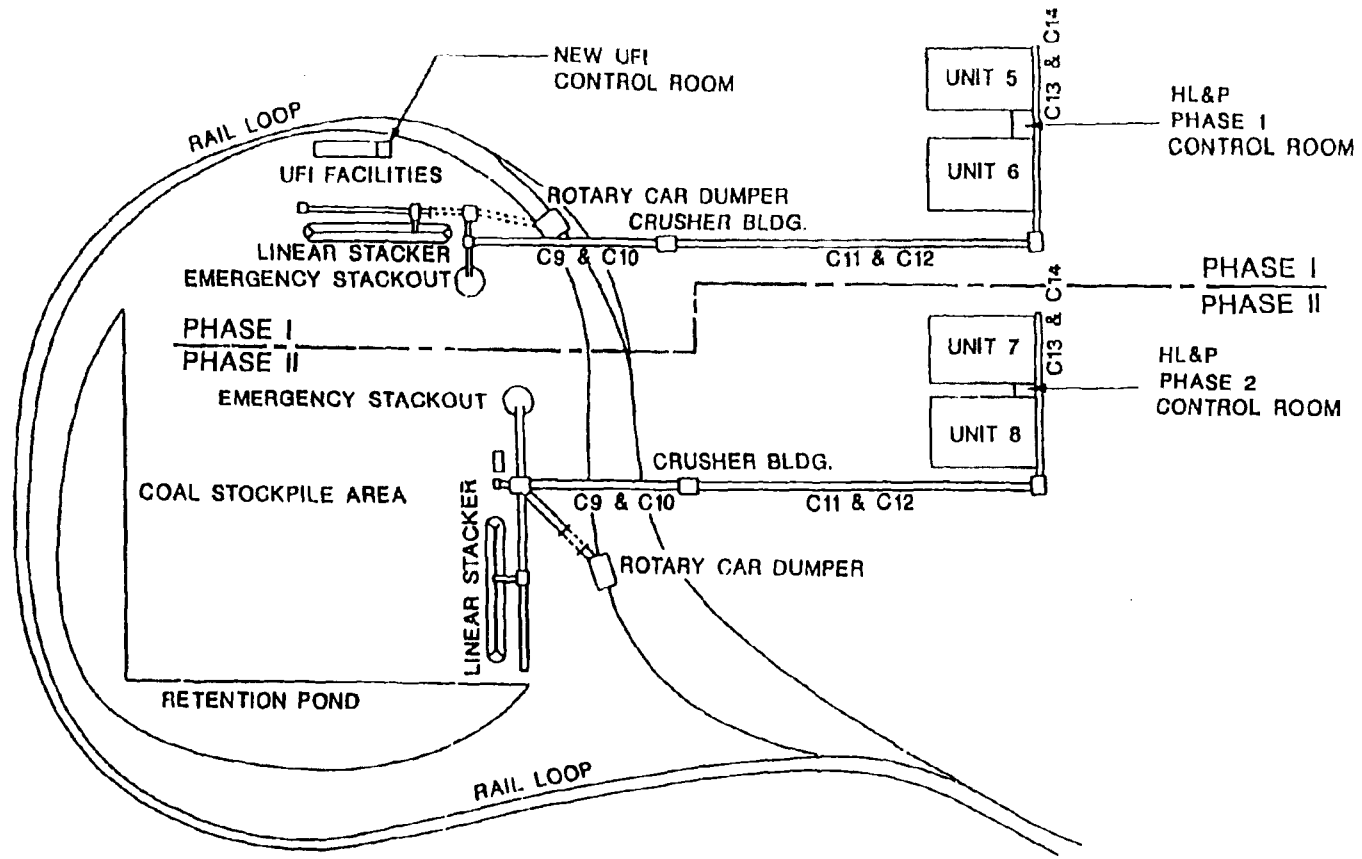
INTRODUCTION

Utility Fuels, Inc. (UFI) owns and operates the coal handling facilities that provide the fuel to the boilers of Houston Lighting and Power Company (HL&P) at their W. A. Parish Electric Generating Station. The Parish plant is located approximately thirty miles southwest of downtown Houston in Fort Bend County, Texas. The initial four units of the plant are gas fired with a second set of four being fired by coal. The Parish Units 5 and 6 are capable of generating 660 megawatts gross each and Units 7 and 8 have 500 megawatts gross capability. The fuel for these Units is obtained from Wyoming and Montana. It is loaded into unit trains composed of rail cars owned by UFI and brought to the Parish facilities where they are unloaded by UFI. The coal is stockpiled until it is required, and then it is reclaimed from the stockpile and transported by conveyors to the boiler silos. (See Figure 1.)

The facilities operated by UFI include a system dedicated to Units 5 and 6 commonly referred to as Phase I and a system dedicated to Units 7 and 8 referred to as Phase II. Each system consists of an unloading/stockpiling component and a reclaim component that conveys the coal to the boiler silos. The control system for the unloading/storing subsystem is separate from the reclaim portion on both phases. Coal is purchased by UFI from both Montana and Wyoming. To optimize the quality of coal burned, a blending capability at the reclaim is necessary. Coal is segregated in the storage areas and both reclaim systems are capable of blending products conveyed to the boiler to meet the requirements given to UFI by HL&P. The total annual coal consumption for the four units is approximately nine million tons.

The Phase I coal handling system was placed into service in 1978 and Phase II was placed into service in 1982. Actual operations of the coal handling facility was performed by an independent contractor until March, 1984. At that time, UFI, an operating company, took over the daily operation of the facility. When this occurred, several evaluations of both equipment and operating methods were undertaken. One area that was identified as being an opportunity for improved productivity and reliability was the Reclaim Control Systems.

In January, 1986, UFI began a project to upgrade the Phase I and Phase II Reclaim Control Systems. The planning, designing, and commissioning of this project is the subject of this paper.

RAIL LOOP

NEW UFI
CONTROL ROOM

UFI FACILITIES

ROTARY CAR DUMPER
CRUSHER BLDG.

LINEAR STACKER
EMERGENCY STACKOUT

C9 & C10

C11 & C12

UNIT 5

UNIT 6

C13 & C14

HL&P
PHASE 1
CONTROL ROOM

PHASE I
—————
PHASE II

PHASE I
—————
PHASE II

EMERGENCY STACKOUT

COAL STOCKPILE AREA

LINEAR STACKER

C9 & C10

CRUSHER BLDG.

C11 & C12

ROTARY CAR DUMPER

C13 & C14

UNIT 7

UNIT 8

C13 & C14

HL&P
PHASE 2
CONTROL ROOM

RETENTION POND

RAIL LOOP

Scale

0   300'   600'   900'   1200'

SKD   3-5-81

W. A. PARISH

COAL HANDLING FACILITIES

FIGURE 1

CS-003

0

EQUIPMENT CONFIGURATION

The Phase I Reclaim Control System consisted of relay logic centralized in three electric equipment buildings. Remote control and monitoring relay contacts were wired to terminal strips located in a single equipment cabinet in each electric building. Also contained in this equipment cabinet was the Forney Remote Telemetry System with input/output boards wired to these terminal strips.

Each of the three Forney Remotes telemetered data over shielded cable to the Forney Master Controller in HL&P's Phase I Computer Equipment Room located just below the Control Room. The Forney Master, which consisted of a NOVA computer and Forney equipment, was hardwired by multiple cables to the Reclaim Control Board upstairs.

The Reclaim Control Board was a fourteen-foot-wide graphic representation of the Phase I Reclaim System. Control pushbuttons, lamp indicators, and an alarm annunciator panel were wired to the Forney System.

The UFI operator was stationed at this board with Gaitronics (a paging/phone system) and radio communication contact to the field personnel. HL&P's control equipment and personnel occupied the bulk of the control room area.

The Phase II Reclaim Control System functions operationally in a similar manner as Phase I. The major difference being that Phase II utilizes a Gould-Modicon Programmable Control Logic (PLC) instead of relay logic.

Three Remote Input/Output (I/O) units are located in three electric equipment buildings. The I/O modules interface directly to the field devices and starters. This data is transmitted via a redundant coax cable to redundant PLC Mainframes. A manual switch places one PLC and one coaxial data highway on active status. The remote I/O data is processed in the active PLC program to perform the permissive and sequential operation.

Adjacent to the dual PLC's in HL&P's Phase II Electronics Equipment Room are additional Remote I/O Units that serve as the interface between the active PLC and the Control Board located upstairs. The layout of the graphic Control Board is very similar in design and function as described for Phase I.

In addition to the two Modicon 584L PLC Mainframes for the Phase II Reclaim System, UFI owns two other 584L PLC's. One was set up as a training unit and the other serves as a site spare.

PROJECT DESCRIPTION

The Phase I control system was becoming increasingly difficult to maintain due to its age and lack of spare parts.

The location of the two control centers in HL&P's main controls rooms, together with the various subsystems being located in HL&P's Electric Rooms created a host of inefficiencies in both operation and maintenance. Centralizing both Control Systems eliminated the need for four operators and placed the heart of Operations in a location accessible by Management, Maintenance, and Operations.

The project was defined to upgrade the Phase I system with Modicon PLC's and I/O's similar to Phase II, and to centralize the control centers in a new UFI Control Room. The Control Board would be replaced by a color-graphic CRT console. A Project Team was established in January, 1986, with a goal of being operational from the new control room by December, 1986.

Major tasks to be accomplished included:

1. Install a new coaxial data highway for Phase I.

2. Extend the existing data highway of Phase II to the new control room.

3. Construct a new control room.

4. Install new remote I/O for Phase I.

5. Install single PLC's for both Phases, to be followed later with backup Unit installation.

6. Install a centralized CRT/Color Graphics Operators Console.

7. Train operators and electricians on the new system.

IMPLEMENTATION

UFI issued contracts for the Control Room Building, Coaxial Highway, and the Control System Equipment and System Design.

It was decided to extend the UFI Administration Building to include the Operators Control Room, the PLC Equipment Room, and the Foreman's Office. Underground conduits and stubups were laid out prior to the slab being poured for the future installation of the control equipment.

The new coaxial data highway for Phase I was installed and tested for the future tie-in of the Modicon equipment. Seventy-five-ohm terminators were installed where Remote I/O Units would be connected. A Frequency Generator and Test Set were connected on the PLC Mainframe end. Simulated signals were placed on the cable and db loss readings were obtained. All losses were measured and found to be within design parameters.

The two spare UFI PLC's were shipped to TXE-Texamation, the Systems Contractor, for configuration and programming into the new Control System. New Remote I/O's were purchased for Phase I as well as for each of HL&P's Control Rooms for the purpose of monitoring silo levels and status.

A computerized CRT/Color Graphics System, called OMNI-MAP, was designed and furnished by Texamation for the man-machine interface between the operator and the PLC System.

The system design was to take the existing Phase II PLC program and change the I/O address from the old Control Board to the OMNI computer. The field addresses and program logic were to remain unchanged.

The Phase I PLC program was to be copied from the Phase II program and then modified. Although several changes had to be made to accommodate unique differences in each Phase, two results were accomplished by using this copy approach. First, a tried and proven program would be used as a basis for change rather than starting from scratch. Second, and perhaps more beneficial for the future, PLC base logic will ultimately replace field relay logic; that is, permissive and shutdown devices would be paralleled between PLC logic and relay logic. A trip device would drop the relays circuit as well as the PLC. Subsequently, relays can be removed with little or no change to the PLC program.

Unique graphic displays and procedures for operator interaction were specified to the Contractor by UFI. A daily working relationship between UFI and Texamation was maintained throughout the design and implementation. An acceptance test procedure was generated and approved.

The Factory Acceptance Test was performed in October, 1986, and the system shipped to W. A. Parish.

The central hardware was installed and powered up as soon as it was in place. The Phase I Remote Units were set in place, powered, and connected to the data highway.

All Phase I remote unit interface wiring was installed in parallel to the old system without any I/O modules plugged into their housing. One by one, Input Modules were plugged in to provide a current data base to the PLC and the console computer. Dynamic values were checked point by point.

When the coal was not being delivered to the Phase I silos, the outputs from the old control system were disabled, and the Modicon Output Modules were inserted and activated. As time permitted, individual control outputs were initiated from the new console. When it was time to start up the old system, Modicon Output Modules were removed and the old outputs activated.

The control swapover continued until all devices were checked out thoroughly on the new system.
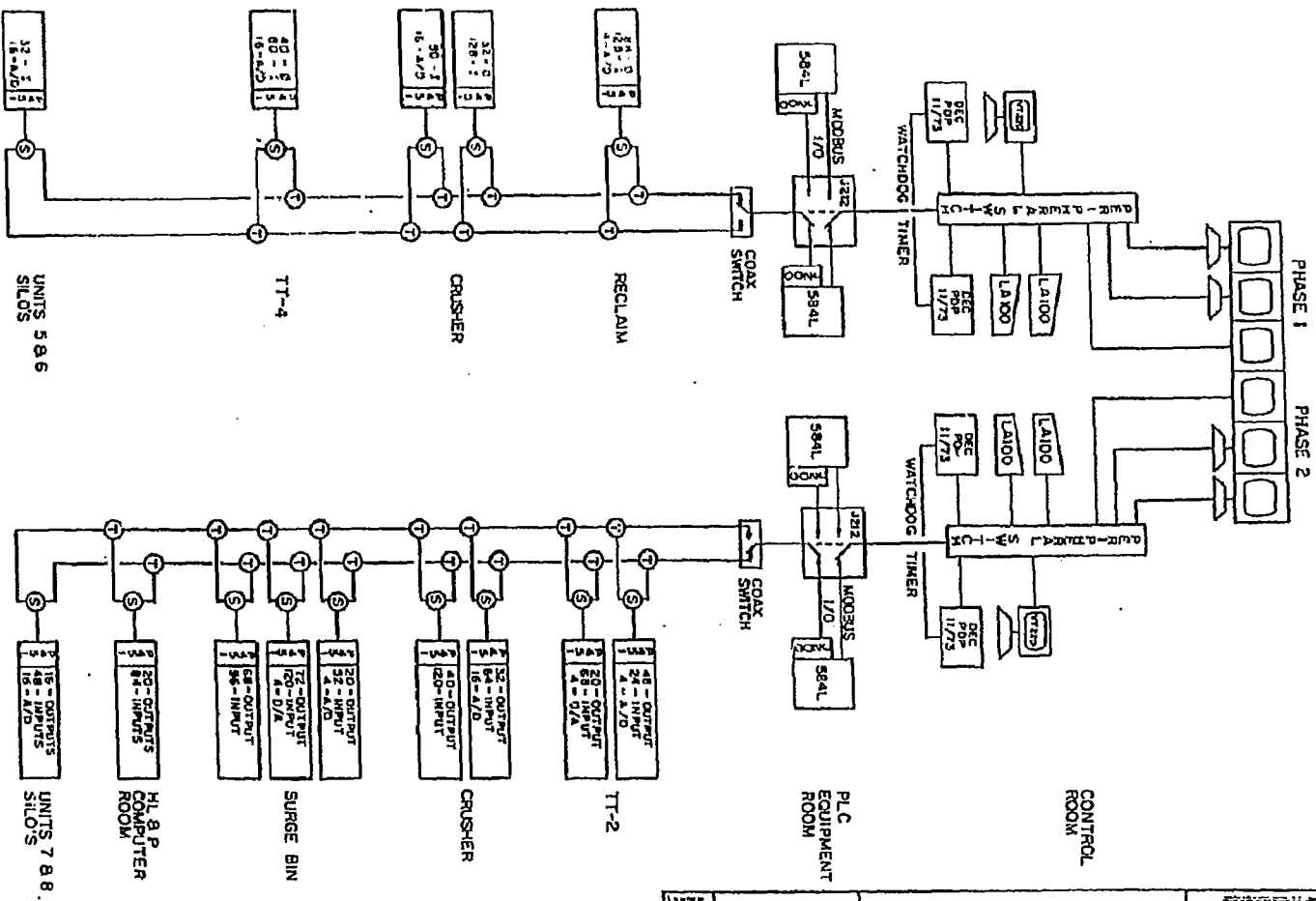
The system control remained on the old system while Operator Training was conducted. Each of the four shift crews received two days of training on the Phase I "monitoring- only" console.

Following the training, the complete swapover of Phase I was performed, and Phase I began operations full time from the new control center.

Phase II could not be paralleled, as was Phase I, because the Remote I/O's could only be addressed from either the old or new PLC Mainframe location. When coal was not being delivered to the Phase II silos, the old PLC was removed from the data highway and replaced with a 75-ohm terminator. The 75-ohm terminator on the new end of the data highway was removed and replaced by the newly configured PLC. Where pushbuttons and lights on the Coal Board monitored and controlled the reclaim system via the old PLC configuration, the OMNI-MAP CRT's and keyboards replaced those functions. Each action had to be checked on the new system. As soon as a coal delivery was required, the reversal of PLC's and terminators was performed to restore the old system. Following several days of this swapping back and forth, the Phase II system became operational from the new control room.
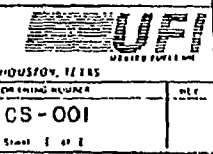
After fine tuning and enhancing the system while on-line, the old PLC's were removed from the old system, installed as backups to Phase I and Phase II, and the new configuration program tapes were loaded.

The Reclaim Control System is shown in the attached UFI Drawing No. CS-001.

PHASE 1

PHASE 2

584L
COAX

MODBUS
I/O

J2R2

COAX
584L

WATCHDOG TIMER

DEC
PDP
11/73

LA36

DEC
PDP
11/73

LA30

LA100

UNITS 5 & 6
SILO'S

TT-4

CRUSHER

RECLAIM

COAX
SWITCH

584L
COAX

J2R2

MODBUS
I/O

COAX
584L

WATCHDOG TIMER

DEC
PDP
11/73

LA30

LA100

DEC
PDP
11/73

LA36

UNITS 7 & 8
SILO'S

HL & P
COMPUTER
ROOM

SURGE BIN

CRUSHER

TT-2

COAX
SWITCH

PLC
EQUIPMENT
ROOM

CONTROL
ROOM

RECLAIM
CONTROL SYSTEM

W.A. PARISH

HOUSTON, TEXAS

DRAWING NUMBER

CS-001

NONE

SYSTEM DESCRIPTION

The Operators Console consists of a center section and 45° wings on each end. Phase I is contained in the left section and Phase II is in the right section. (See Figure 2)

Each Phase contains three Color CRT Monitors. The two on the 45° Section are identical. Each one is a touchscreen with an associated keyboard. The center section contains a fixed display that depicts the silo status.

Each Phase has two printers located in the Operators Control Room. One printer is for all alarm and control action logging. The other printer is for demand logging.

Located in the PLC Equipment room is an Engineering CRT/Keyboard unit that is automatically connected to the off-line, or backup console computer.

Each Phase consists of two DEC-PDP1173 computers. Each computer is connected to a watchdog peripheral switch so as to switch the active computer to the Operators Console and the PLC Mainframe.
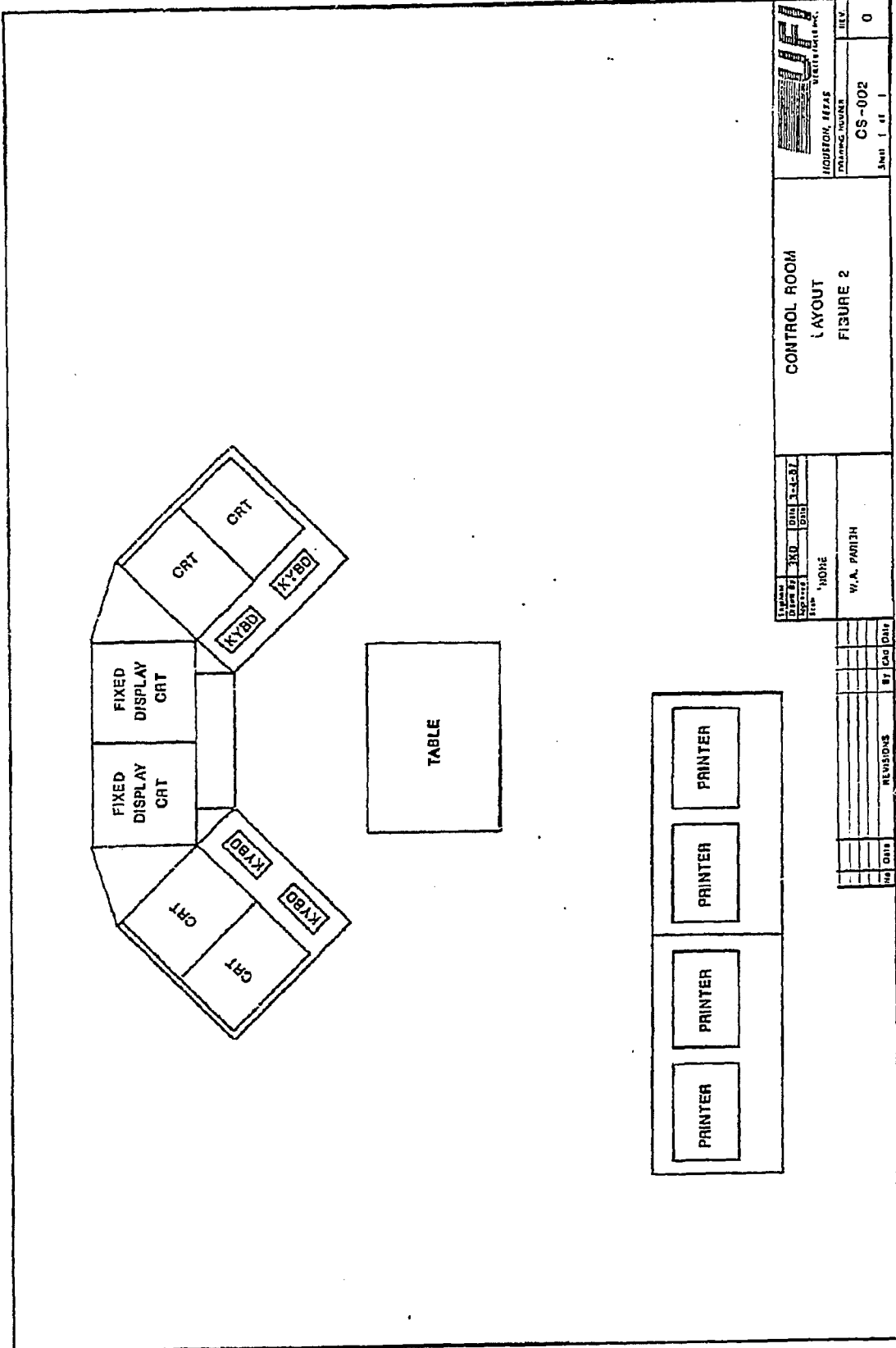
The Operator has access to several different displays on the two CRT's on the 45° section. System displays are arranged in a hierarchy order whereby the overall system summary is at the top and individual subsystems can be selected for display from the Summary. Individual block status displays can be selected from the subsystem.

Dynamic color coding is used to show the status of each device. Red indicates a device is running, green is for off, magenta depicts that the device is in a non-ready or lock-out condition, and other colors for static background.

The two top lines of every display show the two most current alarms or control action. Multiple alarm displays contain all the alarm conditions. When an alarm occurs, an audible alarm sounds and the description is logged on the alarm summary page, the CRT's top line, the printer and on the hard disk. Each alarm, return-to-normal, or operator action is listed by Time/Date, Description, and the Shift Operator's name.

Every analog value can be displayed on Trend Displays that graphically plot historical values. This "historical window" can be adjusted from a minimum of four hours to a maximum of 999 hours.

Other displays show the Fire System Status, Individual Motor Permissive Status, HELP Displays, and Operator Scratch Pads for writing messages to the next Shift Operator.

CRT

CRT

CRT

KYBD

KYBD

FIXED DISPLAY CRT

FIXED DISPLAY CRT

KYBD

KYBD

CRT

CRT

TABLE

PRINTER

PRINTER

PRINTER

PRINTER

HOUSTON, TEXAS

DRAWING NUMBER

CS-002

REV.

0

Sheet 1 of 1

CONTROL ROOM
LAYOUT
FIGURE 2

Drawn By: TKD   Date 3-4-81
Appr'd: Date
Item: 'NONE

W.A. PARISH

REVISIONS

No. Date | By | Chk | Date

2 B2-11

System control can be accomplished in either MANUAL or AUTOMATIC modes. In the MANUAL mode, individual devices are controlled by touch selection and execution. If a device is ready to start, i.e., not locked out and conveyors downstream are on, the device name is displayed in red. If the device is not ready to operate, it will be displayed in white.

In the AUTOMATIC mode, the Operator must designate which conveyor system, odd or even, is to be run, which tripper will be used to fill which Units silos, and whether the Reclaimed Coal will come from the Active or Emergency Pile. If the Flop Gates are in the flow alignment selected, the AUTO MODE will indicate a READY. The Operator then executes a single AUTO START command. Conveyors and the associated crusher are automatically started in the proper sequence. The Operator starts the required feeders. However, the system remembers the initially selected feeders, and when purging of the belts is required, feeders will shut down and restart automatically. Purging comes about when the system detects that the next silo is not to be filled, or when the filling of all silos is complete. When the last silo is filled to its PURGE LEVEL, a sequential purging shutdown commences. When the Tripper has been returned to its PARK position, the automatic sequence is turned off until the next Operator-initiated START occurs. /

SUMMARY

With this computer base CRT Graphics System, UFI will be able to add, delete, and modify system enhancement much easier and less costly than modifying a fixed Mimic Control Board. Record keeping and retrieval will be greatly simplified.

With all the system redundancy and the same PLC I/O equipment in both Phases, UFI's corrective and preventive maintenance should be greatly improved in addition to the savings in labor by manning only one Control Room.

The UFI W. A. Parish Reclaim Control System is a unique, modern, state-of-the-art system of which Management, Engineering, and Operations is very proud.