GASIFICATION OF A HIGHLY CAKING COAL IN THE GEGAS PRESSURIZED GAS PRODUCER

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INTRODUCTION

The coal gasifier program currently underway at the General Electric Corporate Research and Development Center has as its goal the development of an advanced coal-to-gas conversion system which will permit subsequent removal of pollutants before combustion and permit maximum utilization of the fuel in a combined gas turbine/steam turbine powerplant. The development of such an advanced conversion and clean-up system is required if the advantages of the new generations of ultra-high temperature gas turbines is to have a wide impact in mid- to base-load power generation based on coal(1). In fact, the requirements of the environment have encouraged integration of advanced fuels supply and utilization systems. Emissions levels of SO_x, NO_x, and particulates demand a coal conversion process which imposes an efficiency penalty on the overall cycle which the new utilization systems must counterbalance with efficiency increases. On the other hand, the generally higher temperatures of the advanced utilization systems seem to require an ever cleaner fuel due to accelerated consequences of corrosion, erosion, and deposition on gas path parts, and tendency toward increased NO_x production.

The work presented here describes some of the early commissioning trials of an advanced gas producer called GEGAS-D. This unit accepts a reactant blast of air and steam and produces a fuel gas at 300 psi (20 ata) with a composition of approximately 25% CO, 15% H_2, 3% CH_4, 7% CO_2, 50% N_2 and a heating value of about 160 Btu/sft^3 (1500 kcal/Nm^3). A well-burned ash is the only solid product. The current trials are part of a program which addresses four major needs of the fuels conversion system in an advanced, combined-cycle based on coal:

(1) A broad acceptance of coals with high caking and swelling properties.
(2) The ability to handle crushed, run-of-mine coal with high contents of fines.
(3) Minimum use of process steam.
(4) A reduction in the labor force traditionally required to operate coal gasifiers.

These four major needs are being addressed in distinguishable portions of the GEGAS program.

A deep bed agitator is employed at the top of the GEGAS reactor to allow use of caking coals. The agitator is an advanced version of the type pioneered at ERDA's Morgantown Energy Research Center in the 42" diameter fixed bed which has operated there since 1968(2). As the coal cakes, the agitator slowly breaks the large char masses to permit the gas-solid contact required for gasification.

A coal extruder is being developed to handle coal fines. It accepts underscreenings from the crushed run-of-mine product, blends these fines with byproduct tar, and simultaneously compacts and feeds this blend through a seal directly to reactor pressure without requiring a lockhopper. In addition to avoiding blowover of the fines, this technique also reduces the lockhopper feeding requirement and provides a means by which the byproduct tar can be reinjected (as binder) with the feed and cracked. Work on the extruder has been presented elsewhere[3] and will not be covered here.

The reduction in steam consumption is being enabled by the use of a novel grate system at the bottom of the gasifier. Traditionally, steam consumption has been set empirically by reactor operators who would usually employ a large excess of steam so as to avoid problems with the formation of ash clinkers in the fuel bed[4]. Normally, even locally-formed clinkers would necessitate reactor shutdown. The GEGAS grate permits direct mechanical access to such clinkers while the gasifier is on-line. It then breaks and discharges them from the reactor with the rest of the ash. With this feature, the operator can now reduce the margin of excess steam, and steam utilization in the reactor will be increased from 40% to close to 80%.

Finally, the program aims to reduce manpower requirements for the gas production plant by use of minicomputers for data acquisition, process optimization, and process control. With equipment now available to automatically "poke" the bed (stirrer), remove ash from the bed, and operate lockhopper valves for charging coal and discharging ash, minicomputers can actuate these mechanical functions on a routine basis. However, even beyond this, the computer can evaluate producer performance by comparing it to predicted performance in gasifier mathematical models[5], proceed to optimize the producer efficiency, and alert the operator to any out-of-spec parameters that might arise before they become serious.

Beyond simply controlling the gas producer, the computer system can become a part of the overall combined-cycle master control system. While the gas producer is but a small unit operation in the overall cycle, it must be closely integrated with the rest of the powerplant. Figure 1 shows conceptually how a GEGAS gasifier could be integrated into a General Electric STAG combined cycle. Calls for increased fuel gas will require more air extraction from the gas turbine and more feed water from the steam cycle. Variations in gas load will require direct compensation by the gas clean-up system because of the impracticality of storing large volumes of this rather lean gas. Computed, rather than hard-wired, logic will be employed in the master control system to give the powerplant operator much more flexibility in meeting variable load demands with coal of lowest cost but of varying quality.

This paper will describe work which addresses the needs of this plant to handle caking coal (#1) at low steam consumptions(#3). It follows work done in this laboratory at low pressure in a smaller scale gasifier[6] whose operation provided a design base for the present equipment.

THEORY

The purpose of this paper is primarily to describe the results from the GEGAS producer operating with a highly caking coal at low steam consumption. The assumptions and theory leading to the mechanical equipment design will not be discussed. However, it is desirable to discuss the gasification process itself to describe the various reaction zones in the gasifier.

The fixed bed gas producer operated here has the traditional counterflow configuration in which coal slowly settles toward the grate while it is consumed by the uprushing reactant gases (the steam and air "blast"). Ash is withdrawn from the bottom of the open shaft of fuel by the grate. Coal is charged as required to maintain a present solids level in the reactor, usually between four and twelve feet. The amount of steam admitted with the air is adjusted to such a proportion of the air that maximum bed temperatures lie somewhere below the point where ash fusion begins to give problems. Operator judgement in the latter setting is based on the condition of the ash being discharged from the system, the mechanical "feel" of the bed (from poking or other feedback indication such as grate torque), and local temperature measurements, usually from indicators in the walls of the shaft cavity.

Schematically, the reactor system can be represented as shown in Figure 2. The temperature profile shows a local maximum at the end of the oxidation zone. Then the gas temperature drops through the reduction zone as the sensible heat drives the reactions of

$$C + H_2O \rightarrow CO + H_2$$

and

$$C + CO_2 \rightleftarrows 2CO.$$

The point at which these reactions "freeze" marks the end of the reduction zone. It can be usefully characterized as a point of reaction equilibrium for the purposes of producer modeling even though it is known to be only a pseudo equilibrium[7]. Above that point the gas simply transfers sensible heat to the incoming coal. The effect of this regenerative heat exchange at each end of the gasifier shaft is to force sensible heat toward the center of the reactor shaft where it is utilized to drive the above reduction reactions. Conversely, heat loss from the reactor has a deleterious influence on the efficiency of converting latent heat of the coal to latent heat in the producer gas. An estimate of this effect has been made using a simple, one-dimensional model of the reacting system. The results indicate that about two Btu's of latent heat in the product gas are displaced for every Btu lost from the conversion process[6]. As a result, it is important to minimize thermal losses in the conversion process.

EQUIPMENT

Vessel

The GEGAS-D gasifier vessel itself is a 5' O.D. shell with 1" thick walls of 285 grade C steel. It is shown in Figure 3. Since the vessel is an experimental apparatus, ready access to the vessel interior and mechanicals is provided by several 5' flanges which allow the vessel to be separated easily by means of a hydraulic lifting system and roller scheme. Since the vessel proper is 24' in height with coal lockhoppers above and ash lockhoppers below, a 60' enclosed tower was constructed integrally with the gasifier supports with a floor every 10 feet to provide working access to the vessel. The top hemi-head of the vessel is even with the forty foot level of the tower as shown in Figure 4. (The extruder is not connected to the gasifier in the arrangement shown.)

The vessel shell and mechanicals, weighing on the order of 55 tons, are supported from a point just below the start of the top hemi head to minimize the effect of thermal expansion and provide easy disassembly. The externals such as lockhoppers are bolted directly to the gasifier but are supplementally supported with coil springs.

The shell of the vessel is protected from the high temperatures of the gasification process by two layers of low iron, castable refractory with properties as shown in Table I. Major penetrations are flange-cooled.

TABLE I

Vessel Refractories Properties

	Hard-Cast Inner Layer	Insulating Outer Layer
Thickness (in)	3 3/4	8 3/4
Density (lbm/ft^3)	165	80
Conductivity ($Btu-in/hr-ft^2-°F$) @1500°F	10.4	1.8
Cold Crushing Strength (psi)	5000	250

Mechanicals

To provide a continuous coal feed to the gasifier, an auger operating at pressure receives 9 cubic foot batches of coal from the feed lockhopper and meters it horizontally into the vessel. In addition to providing continuous rather than batch vessel feed, this arrangement reduces lockhopper elevation with the side feed configuration, and forces the coal into the gasifier under conditions of tar condensation and deposit accumulation.

Conditioning of the bed from the combustion zone upward is accomplished with a stirrer mechanism that enters the gasifier through the center of the top hemi-head. Three horizontal agitator paddles are spaced over the central drive shaft and are series water cooled. The stirrer can precess from the very top of the gasifier where it sweeps deposits from the dome and offtake areas down to the vicinity of the grate to break coke formations.

The grate is equipped with a similar agitator arm which discharges ash in the lower position and can be raised and rotated to break weakly fused clinker formations. Discharged ash falls into a pit from which it is pushed into the chute leading to the ash lockhopper. An ash quench system is provided to cool the ash as needed. This system also serves to recover some of the sensible heat of the ash as steam and reduces dust on discharge.

Gas Handling

Reactant air at up to 500 psi is indirectly heated to 400°F mixed with steam from a 450 psi electric boiler, and ducted under the grate. Pairs of orifices are provided in the steam and air lines to cover the flow ranges desired.

Inside the gasifier the blast of steam and air passes up through the center and over the edges of the grate. The gas passes upward through the various zones in the settling bed of coal and exits through a fixed diameter flow nozzle in the gasifier off-take flange. From this point, the hot but low pressure gas passes through a dust cyclone and upward through mufflers to the flare on the tower roof. All of the gas produced in these runs was flared hot.

Instrumentation

In order to monitor the operating temperatures of the gasifier, thermocouples were placed at one foot increments along two sides of the gasifier with the tips mounted flush with the inside surface of the castable refractory. Thermocouples are also provided to measure gas temperatures throughout the reactant and product gas systems.

Thermocouples were spot welded to the vessel shell in various locations to monitor metal temperatures. An automatic circuit tests these temperatures against an operator-set maximum and provides an audible alarm if over-temperature occurs.

Bed level is monitored by a plumb-bob type system. Resonant-type level indicators have been employed in the lockhoppers. Coal is weighed by a strain gauge transducer immediately prior to each lockhopper filling cycle. Various pressures on the gasifier lockhopper and gas handling systems are measured by individual millivolt output Bourdon tube transducers so that panel meters and the computer can have access to the signals. The quality of the gas produced is measured with a computer-monitored gas chromatograph which analyzes a cooled sample every seven minutes. Volume percentages of H_2, CO, CO_2, CH_4, O_2 and N_2 are reported. The differential pressure across the orifices are measured with strain gauge transducers.

The control panel for the gasifier laid in a semi-graphic manner, has proven very satisfactory in operation. It is laid out in sections so that an operator or operators can sit in front of the panel and operate and/or take data in a very straightforward manner. There is a materials handling section, air-steam flow section, lockhopper sequence section, and a gasifier mechanical section. Control can also be switched to local panels by key-operated switches on the master control panel.

The computer based data acquisition system has proven to be very productive. During the past several runs, the computer has recorded all data inputs on magnetic tape or magnetic disc and presented the operator with a visual display of these inputs converted to appropriate engineering units every 10 seconds. This display feature has been extremely helpful in data taking and gasifier operation. In addition to gas flows and gas analysis, it records operation of gasifier mechanicals including the charged weights of coal.

OPERATION PROCEDURE

For these runs, the gasifier was started on anthracite coal to minimize the tar build up on cold surfaces. In preparation for operation, the gasifier was loaded with 1 1/2 feet of ash, then about 1 1/2 feet of wood scraps and 5 feet of anthracite. Before ignition, the bed was warmed for about an hour with 400° air. Air flow was then cut back and a simple hydrogen-fueled ignition torch inserted through a removable 1" diameter plug located at the base of the wood layer. The wood layer provided for adequate flame spread before the coal was ignited.

After about 30 minutes of torch application, the torch was withdrawn and the air flow increased to a high bank condition with no steam flow. When the lower wall temperatures approached the maximum allowable values, steam was introduced along with the air. Once the exit gas temperature reached 700°F (about four to six hours) the coal feed was switched from anthracite to bituminous.

During typical operation the combustion zone wall temperatures were maintained at maximum values using the steam admission as a control. The steam ratio was not adjusted for every variation in the maximum thermocouple reading. Only obvious temperature trends either upwards or downwards would encourage steam flow adjustment. Operator judgement is an important factor in steam control.

Ash discharge to the pit was intermittent during these runs. Periodically the grate was slowly raised and rotated to be sure that the walls were cleared of any adhering slag. The typical ash output was 1" - 3" diameter chunks of clinkered material.

The stirrer was rotated continuously to agitate the region about two feet below the top of the coal bed. Periodically it was precessed down to an elevation 4 feet above the grate pan and then raised back to its original position.

Using the bed level detector and gas offtake temperature as guides, the coal bed level was maintained constant. The coal feed auger was allowed to run continuously in order to even out the batches it received from the lockhopper.

RESULTS

The GEGAS-D gas producer has been operated successfully to date on Pittsburgh #8 coal at pressures of 200 and 300 psig in run durations of 40 and 10 hours respectively. About 100 hours of operation at 200 psi or above were accumulated by November, in total.

From the first 40 hour run, an intermediate 15-hour period was taken for the mass and heat balances shown in Tables II and III, respectively, Gas composition variations over the period are shown in Figure 5. Gas compositions from the GEGAS tests were obtained by carefully integrating gas analyses over the fifteen-hour run period. Mass balance calculations were made by forcing a nitrogen balance to obtain exit gas flow (to both the exit orifice and lockhoppers) before the carbon balance was computed. Then the oxygen balance was forced to obtain the exit water flow. The good balances on carbon and hydrogen served as a check that major masses were accounted for and resulted in an overall closure within 1% in spite of a poor ash balance which had little impact because of the low mass involved within the fifteen hour period selected.

The heat balance was based on the mass flows in Table II and was closed to within 5 1/2%. All computed losses from the shell to the air (using temperatures from the shell temperature alarm system) and measured losses to cooling agents were included. Shell temperatures of 270°F were measured at the oxidation zone in surprising agreement with 280°F calculated from the properties in Table I.

Heat and mass balances for the shorter 300 psig run were not obtained, but Figure 6 shows the resulting pressure and higher heating value variations. The run was shut down normally after 9 1/2 hours. An average steam/air ratio of 0.2 lbm/lbm was employed during this run and a well-clinkered ash was discharged.

At the conclusion of the 200 and 300 psig runs, the gasifier was disassembled and the inner refractory inspected. The lower part of the shaft was coated with slagged ash to a depth allowed by the paddle diameters on the coal and grate agitators. There were many signs that the ash was fractured by the agitators but no evidence that the refractory was failing due to internal fractures, i.e., ash was breaking from ash rather than refractory breaking from refractory when agitator forces were applied.

DISCUSSION

The initial test results presented in this paper show that the GEGAS-D gasifier can operate successfully with this highly caking coal at low steam consumptions. The carryover of solids to the cyclone was only about 1% of the coal input even though crushed, run-of-mine coal was employed with agitation.

Continuous agitation with the upper stirrer covering the two-feet vertical region below the top of the coal bed with only periodic deep-bed penetrations was adequate for this difficult fuel. At the agitation rates employed during these two runs, the frequency of deep-bed penetration was satisfactory when so timed that one and one-half to two feet of axial coal movement was permitted between penetrations. Penetration by the (lower) ash zone agitator was required about every three-to-six inches of ash movement.

The resulting gas analyses compare favorably with results obtained in the experiments of Lewis, Liberatore, and McGee[2]. Table IV compares the results. The air/coal ratio is 20% less in the present experiments, with only 8% of this reduction attributable to the amount of fixed carbon in each fuel and about 2% for difference in steam usage leaving a 10% real improvement in air utilization. This improvement, and the overall increase in cold gas efficiency, is attributable largely to the difference in coal bed containment in the two systems. The present gasifier was well insulated and had measured and computed losses of 3.7% of total input enthalpy. The heat losses estimated by Lewis, et. al. for their tests was 10-13%.

The ability of the stirred, fixed bed gasifier to produce a continuous flow of a rich gas in a reliable fashion with a highly caking coal at pressure was affirmed in these preliminary GEGAS-D trials. Moreover, the operating characteristics of the gasifier seem well-suited for combined-cycle powerplant applications. Instantaneous demands for gas can be met initially by only increasing flows of air and steam to the reactor. The accompanying necessary increase in coal flow can be deferred for several minutes because of the large carbon inventory in the system. The ability to abort operation, bank for long periods of time, and even bottle up hot for hours without reignition on restart has been achieved without either bed "dumping" phenomena or specialized solids handling procedures. This is very attractive for load-following operations in power plants. The fact that this high coal-to-gas conversion is obtained in a single process step is viewed as an additional advantage.

CONCLUSIONS

It is concluded from these preliminary trials that crushed, run-of-mine, highly caking Pittsburgh #8 coal can be gasified successfully in a fixed-bed gasifier with periodic deep-bed agitation. This verifies and extends results obtained on the Morgantown Energy Research Center gasifier with continuous deep-bed agitation. Results from both experiments confirm that this difficult coal can be gasified with a low steam consumption with steam-to-air mass ratios near or less than 0.2 in the reactor blast at pressures between 6 and 20 atmospheres. Moreover, the present experiment has given a preliminary indication that refractory-lined shafts are compatible with low steam/air, stirred-bed operations with this difficult coal. The use of insulated, rather than water-cooled shafts and periodic vs. continuous bed agitation seems to be worth from 5 to 10 points in cold gas conversion efficiency in 3 to 3 1/2 foot diameter pilot units.

Further work will attempt to quantify the ability to respond quickly to load demands and to measure the effects of demand on gas quality and process efficiency.

TABLE II **MASS BALANCE SUMMARY**

Conditions: S/A = 0.17 lbm/lbm
Pressure = 225 psig
Continuous Period = 15 hours

FLOWS:
(lbm/hr)

	Coal	Air	Steam	Ash Quench	TOT. In	Gas	Water	Dust	Residue	Tars &Oils (3)	TOT. Out	O/I
C	890.				890.	788.	-----	13.8	49.	54.	905.	1.02
H	65.7		66.	0.6	132.	95.1	30.3	----	2.8	3.	131.2	.99
O	97.5	800.	528.	4.4	1430.	1165.	241.(2)	----	23.8	0.5	1430.	1.00
N(1)	17.8	2670.			2690.	(2690.)	----	----	----	----	2690.	1.00
S	50.7				50.7	(50.7)	----	----	----	----	(50.7)	1.00
Ash	198.0				198.	------	----	2.2	114.	----	116.	0.59
TOT.	1320.	3470.	594.	5.0	5385.	4790.	271.	16.0	190.	57.5	5320.	0.99

(1) Basis for mass balance

(2) Forced finally to yield water content of gas

(3) Assumed as 4% of coal (per Ref. 8) at $CH_{0.8}O_{0.1}$

AVERAGE GAS COMPOSITION:

	Fixed Gas (Vol.%)	Raw Gas (Vol.%)
CO	23.8	22.1
H_2	17.0	15.8
CH_4	3.2	3.0
CO_2	6.7	6.2
N_2	49.2	45.7
O_2	0.1	0.1
H_2O	----	7.2
	100.0	100.0

Table III - Heat Balance Summary

	Flow (lbm/hr)	Latent: Specific (Btu/lbm)	Latent: Total (Btu/hr)	Sensible: T (°F)	Sensible: Total (Btu/hr)	Total (Btu/hr)	% of Input
INPUT							
Coal	1320	12,303	$16.24x10^6$	70	-------	$16.24x10^6$	94.5
Air	3470		---------	350	$0.235x10^6$	$0.235x10^6$	1.4
Steam	594		---------	350	$0.709x10^6$	$0.709x10^6$	4.1
Ash Quench	5.0		---------	70	-------	0	
TOTAL IN	5390					$17.2x10^6$	100
OUTPUT							
CO	1300.	4,347	$5.651x10^6$	1020	Fixed Gases		
H_2	66.9	61,100	$4.087x10^6$		=4790x0.26		
CH_4	99.4	23,879	$2.374x10^6$		x(1020-32)		
CO_2	57.6	------	-----		$=1.246x10^6$		
N_2	2690.	------	-----				
O_2	3.2	------	-----				
NH_3*	(2.2)	Neglect					
H_2S*	(51.)	3,983	$0.203x10^6$				
Tar*	(58.)	16,500	$0.957x10^6$				
H_2O	271.	------			$0.320x10^6$		
*Assumed							
Gas	5120		$13.27x10^6$		$1.566x10^6$	$14.84x10^6$	86.4
Residue C	49	13,500	$0.661x10^6$	500°			
Residue Other	141	------	-----	"	$0.031x10^6$		
Residue	190		$0.661x10^6$		$0.031x10^6$	$.69x10^6$	4.0
CW (65° Inlet)							
Stirrer				100	$0.365x10^6$	$.365x10^6$	
Grate				80		$.184x10^6$	3.1
Shell Loss						$.112x10^6$	0.6
TOTAL OUT						$16.2x10^6$	94.3

Table IV. Comparison with Results of Lewis et al.

		Reference (2) (MORGAS)		These Results (GEGAS)	
Run		1	2	200 psig	300 psig
Coal Analysis					
Ultimate %	C	76.9		67.42	
	H	5.5		4.98	
	O	6.2		7.39	
	N	1.3		1.35	
	S	2.4		3.82	
	Ash	7.7		15.02	
Prox. %	H_2O	1.3		2.53	
	VM	36.8		32.75	
	FC	54.2		49.70	
Ash I.D. Temp.(°F)		(Reducing)		2010	
FSI		8		7	
HHV (Btu/lbm)		13850		12303	
Steam/Air (lbm/lbm)		0.15	0.16	0.17	
Pressure (psig)		80	80	225	300
Air/Coal (lbm/lbm)		3.32	3.31	2.63	↑
Gas (dry, cold %):					
	CO	20.0	20.5	23.8	Averages Not Made
	H_2	15.5	15.6	17.0	
	CH_4	2.8	2.4	3.2	
	CO_2	7.2	8.7	6.7	
	N_2	54.5	53.2	49.2	↓
HHV Btu/sft^3		142	145	160	∿165 (cf. Fig.6)
Efficiency:					
$\frac{\text{Cold Gas HHV}}{\text{Coal HHV}}$ (%)		64	68	74	---

ACKNOWLEDGMENT

The authors would like to acknowledge the efforts of the key engineers who assisted in the operation of the gasifier for the trials reported here. R.A. Shisler and R.M. Pfeiffer provided data acquisition system set-up and programming. J.K. Floess set up, calibrated and supervised operation of the gas analysis system. R.A. Engler designed and supervised installation of the gasifier controls. Additional personnel assisting in operations were A.H. Furman, L.P. Inzinna, R.L. Richard and M.S. Samuels. The authors would also like to acknowledge the Electric Power Research Institute whose complementary program to develop the coal extruder contributed some of the equipment utilized in these trials.

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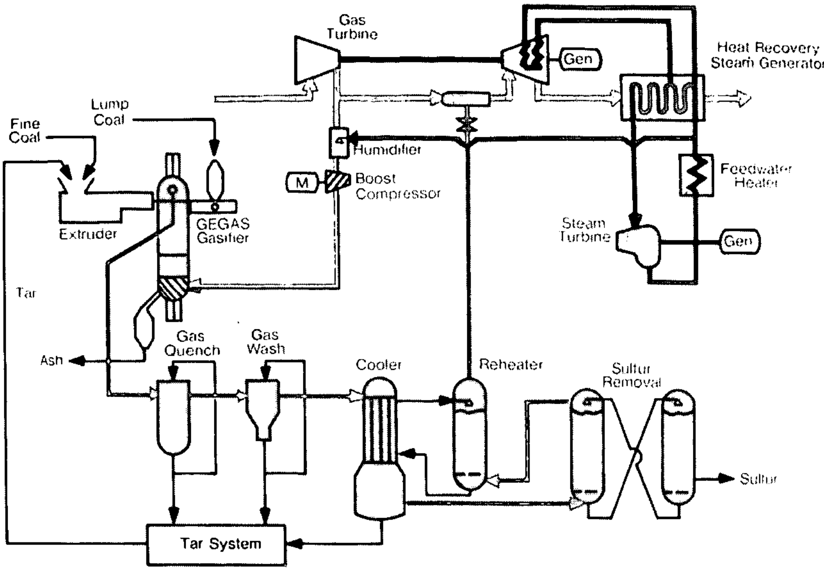

Figure 1. Integration of the GEGAS Gasifier Into An Advanced Combined Cycle.

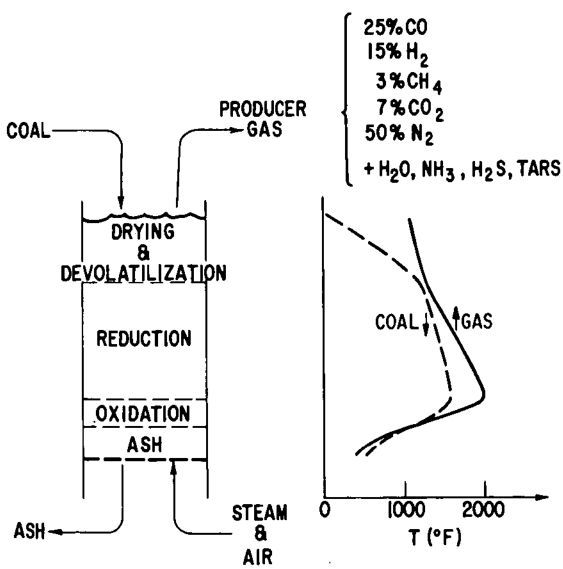

Figure 2. Reaction Zones in a Fixed Bed Producer.

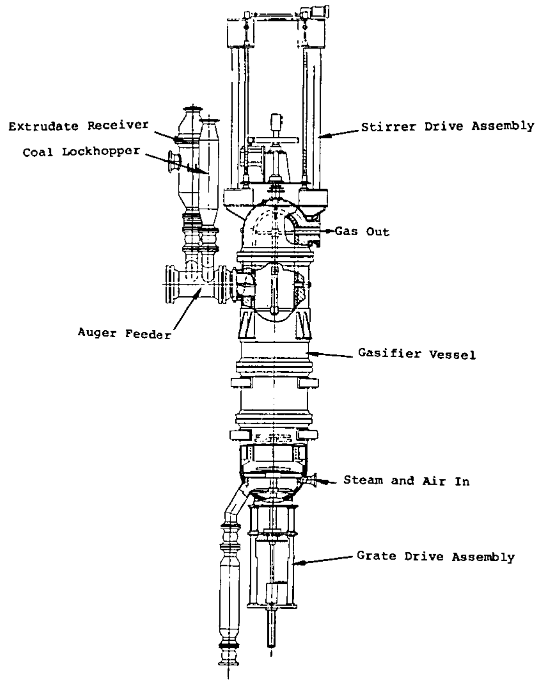

Figure 3. GEGAS-D Coal Gasifier Vessel.

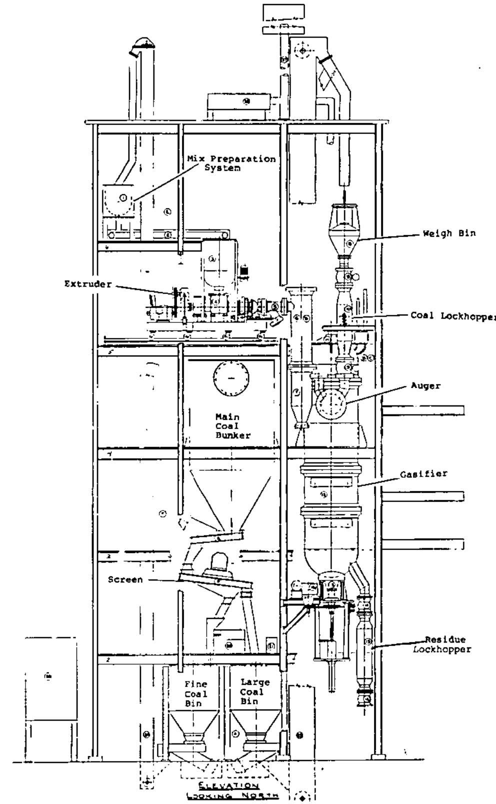

Figure 4. General Electric Gasification Facility.

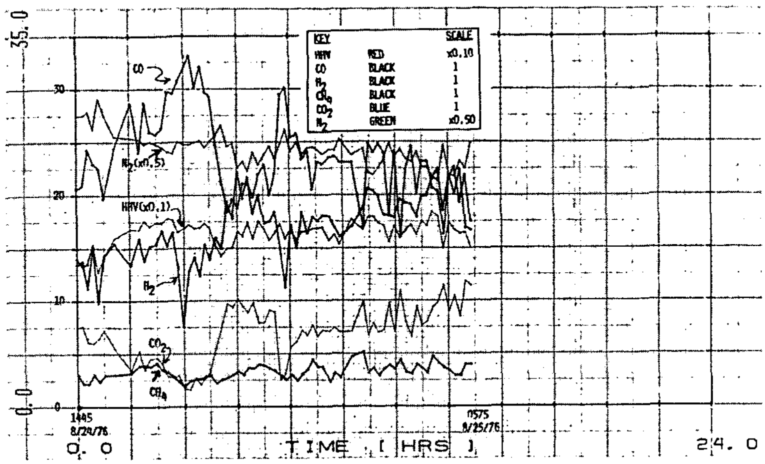

Figure 5. Gas Compositions (Vol. %, Dry) From Run at 225 psig.

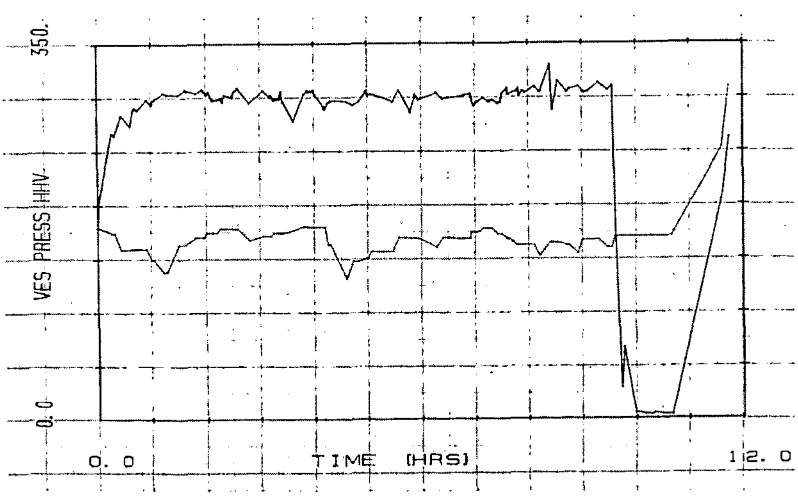

Figure 6. Pressure (psig) and Gas Heating Value (Btu/sft^3) From Run at 300 psig.

MASS AND HEAT BALANCE FOR COAL GASIFICATION BY ATOMICS INTERNATIONAL'S MOLTEN SALT GASIFICATION PROCESS. Charles A. Trilling, Atomics International Division, Rockwell International Corporation, 8900 De Soto Avenue, Canoga Park, CA 91304.

Rockwell International's Atomics International Division is presently developing molten salt processes for the gasification of coal. In these processes the coal is partially oxidized and completely gasified by reaction with air or with oxygen and steam in a bed of molten sodium carbonate. The gasification takes place at temperatures of 1700 to 1800°F and pressures of 10 to 30 atmospheres. The sulfur and ash of the coal are retained in the melt. A nonpolluting low- or medium-Btu gas is produced which can be used as fuel gas for electric utility or industrial applications or as a synthesis gas for the production of pipeline quality gas, methanol or liquid hydrocarbons. A sidestream of melt is withdrawn from the gasifier and processed in an aqueous regeneration system for removal of ash, recovery of elemental sulfur, and return of the regenerated sodium carbonate to the gasifier.

This report describes the mass and heat balance around the molten salt gasifier and the composition of the fuel gas produced as a function of air-to-coal or oxygen- and steam-to-coal feed ratios and system heat losses. Calculated values are compared with the experimental data obtained in laboratory and small scale pilot plant tests.

Operation of the Westinghouse Fluidized Bed Devolatilizer with a Variety of Coal Feedstocks

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Under Energy Research & Development Administration sponsorship, Westinghouse is conducting a diverse program to develop a low Btu coal gasification, combined cycle electrical power generating process. The total program includes work in gas turbine combustor development, studies of turbine tolerance to erosive and corrosive fuel, gas cleaning and coal gasification process development. As part of the gasification work, Westinghouse is operating a process development unit (PDU) at Waltz Mill, Pennsylvania. It is this aspect of the program that will be discussed.

PROCESS DESCRIPTION

Before we discuss specific test results, a brief introduction to the process is in order. Essentially, the Westinghouse Fluidized Bed Process consists of two reactors (Figure 1). Coal is fed by pneumatic transport from lock hoppers to the devolatilizer-desulfurizer reactor where it is fluidized by hot reducing gases produced in the gasifier-agglomerator reactor. The coal and hot gas are transported at relatively high velocity upward in a draft tube along the reactor centerline. Devolatilized coal or char product is also entrained in the upward flow of solids and gases in the draft tube. This dilution - on the order of 30 to 1 - of fresh coal with char in the entrained bed of the draft tube prevents the fresh coal from sticking together or caking as it is heated through its plastic stage. When the coal leaves the draft tube, it enters a second portion of the fluidized bed where devolatilization is completed and where desulfurization takes place. The latter is achieved by absorption of the hydrogen sulfide with dolomite which is also circulating in the fluidized bed with the char product.

Char from the devolatilizer is continuously drawn from the bed and fed to the gasifier-agglomerator reactor. A portion of the char is combusted with air in the combustor zone at the bottom of the reactor. This zone is operated at a temperature of about 1950°F at which ash particles stick together or agglomerate and become defluidized. Ash is continuously removed from the bottom of the reactor after being cooled with steam. This steam is used to gasify the remainder of the char and to moderate the combustor temperature. The heat produced in the combustor is carried to the gasification zone by circulating solids and fluidizing gases composed essentially of CO, CO_2, H_2, H_2O and N_2. Eventually this gas exits the gasifier and enters the devolatilizer where it provides a heating and fluidizing medium for the bed.

The hot product gases leave the devolatilizer at about 1600°F and 225 psig and go through various stages of cleaning for particulate removal prior to being combusted with air in a gas turbine - steam turbine combined cycle generating plant. Nominally, a 50 T/H coal gasification plant produces sufficient gas for 130 MW of electricity plus the compressed air and steam used in the process. The low Btu gases have a heating value of about 120 Btu/scf.

PDU RESULTS

In August 1976, the initial series of tests of the devolatilizer reactor were completed on the PDU scale of nominally 15 T/D. These tests were conducted with a variety of feedstock materials and conditions and culminated with the "feasibility demonstration" of the system with two highly caking Eastern bituminous coals. These coals were processed for over 200 hours in the devolatilizer without pretreatment. This was a major accomplishment in coal gasification development because the use of costly and inefficient pretreating operations (usually by surface oxidation) to decake Eastern coals was not necessary.

The devolatilizer test program was comprised of three types of test: plant start-up/shakedown, system sensitivity and feasibility demonstration runs. Essentially the work began with non-caking coal feedstocks, progressed to mildly caking bituminous coal and concluded with highly caking Pittsburgh and Upper Freeport seam coals. This test sequence is summarized in Table I. Coal properties are shown in Table II. Typical char product properties are given in Table III.

The principal product of the devolatilizer reactor is de-caked coal or char. To understand and predict the dynamics of the integrated gasification plant, the operating characteristics of the devolatilization process must be considered. Because the Westinghouse gasifier is a fluidized bed reactor, the effect of devolatilization on the char production rates and on the fluid dynamic properties of the char particles are critical. These properties include char particle size distribution, the fraction of coal feed that becomes char product, and the split of that product between drawoff from the bed and overhead product taken from the gas stream in the particulate removal cyclone.

To study these effects, the geometric weight mean of char samples withdrawn from the bed (this does not include char product which goes overhead with the product gas) expressed as a dimensionless ratio, geometric weight mean of char to geometric weight mean of coal, has been explored as a function of the operating parameters involved.

TABLE I

PDU Devolatilizer Test Program

Type of Test	Type of Feedstock	Name of Feedstock	No. of Hrs. Coal Processed
PDU Shakedown	Lignite Derived Char	Husky Char	17
	Sub-Bituminous-C	Sorensen	13
	High Volatile Bituminous	Minnehaha/Indiana #7	30
System Sensitivity	High Volatile Bituminous	Minnehaha/Indiana #7	191
	Medium Volatile Bituminous	Champion/Pittsburgh	30
Feasibility Demonstration	High Volatile Bituminous	Minnehaha/Indiana #7	131
	Low Volatile Bituminous	Renton/Freeport	96
	Medium Volatile Bituminous	Champion/Pittsburgh	91
		TOTAL	599

TABLE II

Coal Raw Materials

Coal Company Mine Seam	Kemmerer Sorensen Adaville	Amax Minnehaha Indiana 7	Consol Montour Pittsburgh	Consol Renton Upper Freeport
Analysis (%)				
Volatiles	36.4	32.1	35.0	35.6
Carbon	41.0	43.3	49.0	53.8
Moisture	19.9	16.2	6.5	1.7
Ash	2.7	8.4	9.5	8.9
Sulfur	0.4	0.5	1.9	1.4
Ash Fusion (Reducing) °F				
I.D.	NA	2170	2270	2510
H=W	NA	2270	2310	2570
H=1/2W	NA	2320	2350	2600
Fluid	2160	2380	2400	2650
Free Swelling Index	0	1-1/2 - 2	7 - 9	8 - 9
Gieseler Plasticity ddm	NA	250	25,000	30,000
Heating Value, Btu/lb, MAF	13,217	14,250	12,570	13,740
Bulk Density, lb/ft^3	45.0	43.8	43.6	44.6

TABLE III

Char Product Properties

Coal Company Mine Seam	Kemmerer Sorensen Adaville	Amax Minnehaha Indiana #7	Consol Montour Pittsburgh	Consol Renton Upper Freeport
Analysis (%)				
Volatiles	6.2	2.7	2.9	2.7
Carbon	83.1	77.1	76.4	78.0
Moisture	1.7	1.0	0.6	1.5
Ash	9.0	19.2	18.2	16.6
Sulfur	0.3	0.2	1.9	1.2
Free Swelling Index	NA	NA	0	NA
Gieseler Plasticity ddm	NA	NA	No Fluidity	NA
Bulk Density, lb/ft^3	14.7	24.2	29.0	22.0

At this time, the data does not allow sophisticated prediction of reactor behavior; however, empirical correlations of results have been made to identify critical parameters. Essentially, relationships were sought between the diameter ratio and the operating parameters presented in Table IV as first order effects for each parameter and for every combination of paired parameters described in Table IV. For example, the first entry in the table (on the coal feed rate row and coal feed rate column) indicates no correlation was found for the available data for the diameter ratio versus the coal feed rate alone. However, proceeding to the next column, the table indicates a fair correlation for the diameter ratio versus coal feed rate when the reactor freeboard velocity is used to parameterize the data. The general criteria used to judge the extent of the correlation was $\pm 10\%$ scatter for a strong correlation, $\pm 15\%$ scatter for a weak correlation, and no correlation for scatter beyond 15%.

As can be seen from Table IV, the gas velocity through the reactor (Figure 2) and the rank of the coal feedstock (Figure 3) correlate the data. In order to get a more complete picture, the data have been correlated in Figure 4 to include all of the pertinent effects. Several observations can be made from this plot. Due to the narrow temperature spread for the reactor gas, the constant freeboard velocities lines drawn through the data are essentially constant gas input rate lines. Thus, proceeding to the right along a freeboard velocity line indicates the effect due to increasing the coal feed rate. The increasing slope of the three lines drawn indicates greater sensitivity to the coal feed rate as the freeboard velocity and/or rank of the coal are increased. Because the reactor freeboard velocity and the coal rank were changed simultaneously, it will be necessary to conduct further tests and analyses to separate the effects of freeboard velocity and coal rank.

Summing the char product stream flow rates (drawoff from the reactor and the char separated from the product gas stream) and plotting the data as in Figure 5, we see that approximately 65% of the coal feed leaves the reactor as char product regardless of the freeboard velocity. However, the split in the two streams is indeed dependent on the freeboard velocity. In Figure 6, it has been shown that increasing the freeboard velocity will cause a relative decrease in the amount of char in the drawoff product stream. To distill these facts, increasing the reactor freeboard velocity appears to strip increased amounts of char from the bed leaving behind a larger mean particle.

With regard to the effect of the coals' caking and swelling properties on the char particle size, the data does not allow any strong conclusions. One would expect the higher free swelling coals to grow more during devolatilization. In addition, it has been proposed that as the coal goes through the sticky phase it is likely to gather a coating of fines on its surface. Looking at photomicrographs of char particle cross sections (see Figure 7), reveals the pore structure, but does not indicate any strong differences between comparable size char particles of different coals. In order to make a rigorous comparison of pore structure, one should look at the char product for identically sized coal. Because of the tenfold size spread in the coal feedstock, we cannot accomplish this from PDU char samples.

TABLE IV

Summary of Bed Material Particle Diameter Correlations

$\frac{D_p \text{ Bed Char}}{D_p \text{ Coal}}$ vs. Coal	$\dot{m}$ Feed Rate	U_{fb} Freeboard Velocity	Coal Rank	Bed Temp.	Char Residence Time	$\dot{m}/\dot{G}$
$\dot{m}$	NF	Fair	Fair	NF	NF	NF
U_{fb}	NF	Strong	NF	NF	NF	NF
Rank	NF	NF	Fair	NF	NF	NF
Temperature	NF	NF	NF	NF	NF	NF
Residence Time	NF	NF	NF	NF	NF	NF
$\dot{m}/\dot{G}$	NF	Fair	Fair	NF	NF	NF

NF - No correlation found for the available data

$\dot{G}$ - Total reactor inlet gas flow rate.

There appears to be a different wall structure on some of the Champion coal char particles (see arrow, Figure 7) which could be a result of the condensing and coking of the tars from this highly fluid coal or an accumulation of fines. At the time of this writing, it has not been determined if this wall structure difference is significant. This phenomena will be investigated further during future tests.

A question germaine to fluidized bed operation and to particle cleaning requirements for the gas is how much attrition or growth of coal and char takes place in the bed. Figures 8 and 9 are plots of overhead product, bed product char and coal versus particle size. The bottom curves combine the two product chars into a "blended" product for comparison with the coal raw material. This presentation illustrates several facts:

1) Both particle growth and reduction take place in the devolatilization process as a result of inter-particle impact devolatilization, gasification, agglomeration and thermal expansion.

2) Both fines and oversize char fractions are produced from the mid-range coal particle sizes (note the bi-modal distribution of blended product).

3) Net production of -200 micron material is on the order of 10 percent of the coal feed (15 percent of char products).

CONCLUSION

The results of this study of char product characteristics along with the other results achieved during the past year of testing with the devolatilizer reactor indicate that the design concept for this portion of the process is feasible. Highly caking coals were processed for over two-hundred hours without pre-treatment utilizing the draft tube and recirculating fluidized bed concept. Char product produced in the process was adequately devolatilized and was in an acceptable size range, for both overhead and bed material fractions, to be used in the gasifier-agglomerator reactor. The attrition growth of particles which occurred was within acceptable limits with respect to overall process dynamics. To some extent, the resulting char particle size distribution depends on freeboard gas velocity, coal feed rate and coal rank.

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FIGURE 1 - CONCEPTUAL LOW BTU COAL GASIFICATION COMBINED CYCLE PROCESS

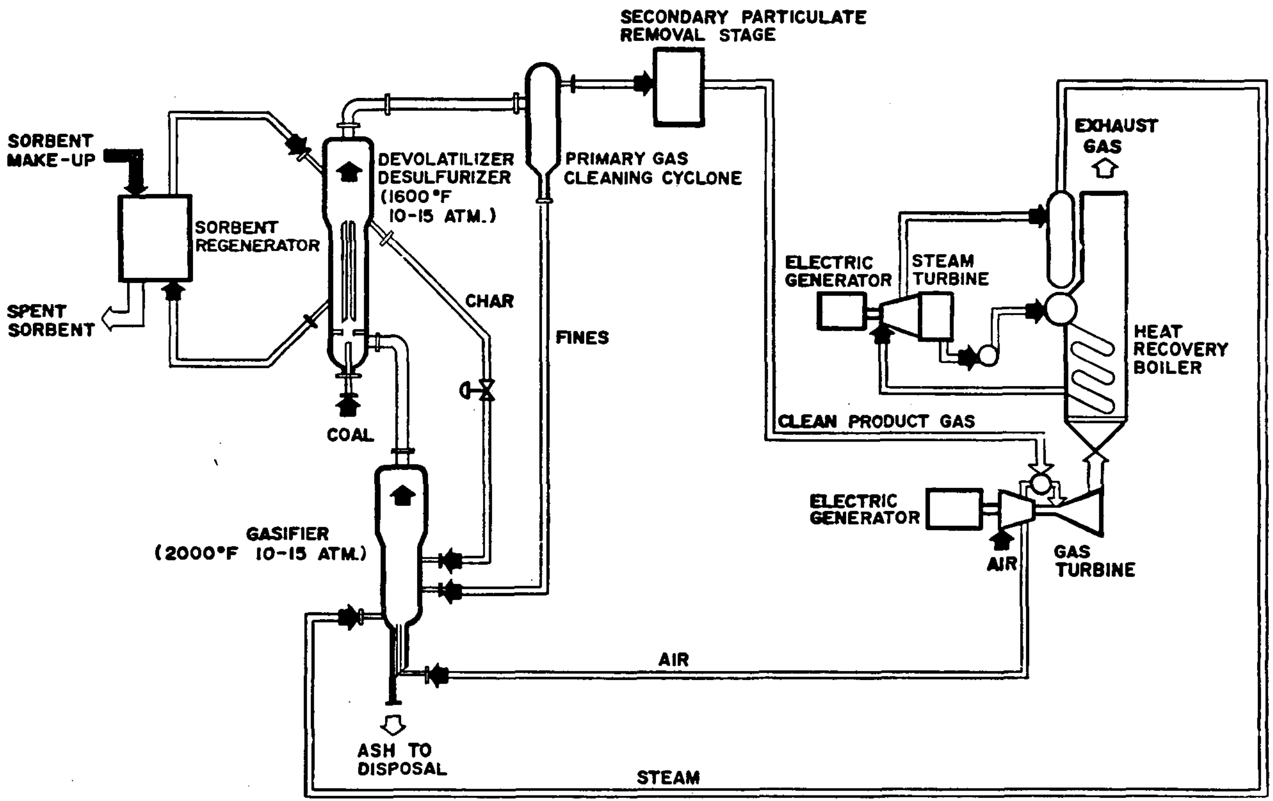


FIGURE 2 - BED PARTICLE SIZE VS REACTOR FREEBOARD VELOCITY

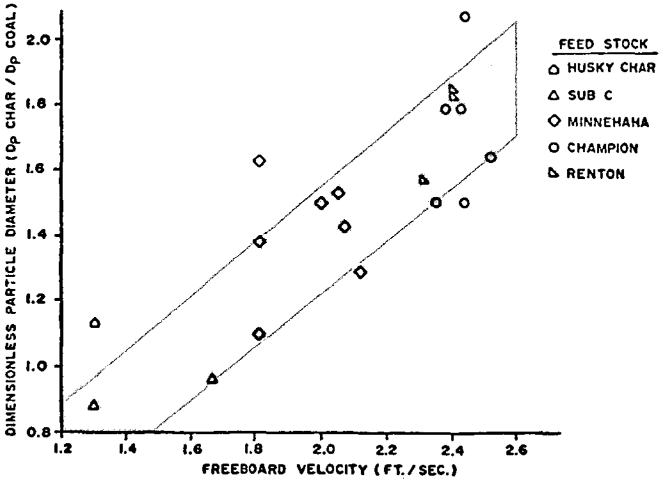


FIGURE 3 - BED PARTICLE SIZE FOR EACH FEEDSTOCK

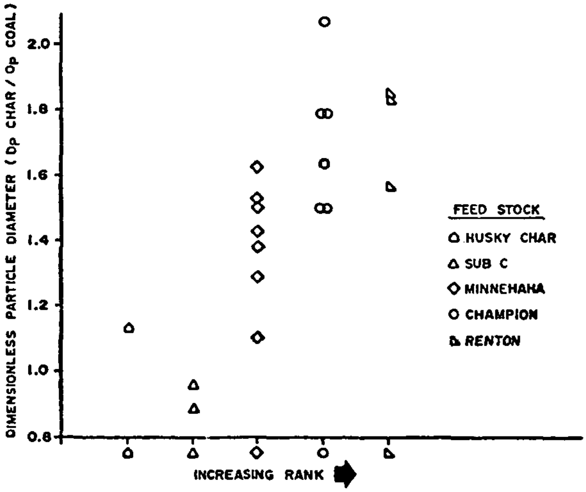


FIGURE 4 - BED PARTICLE SIZE VS LOADING

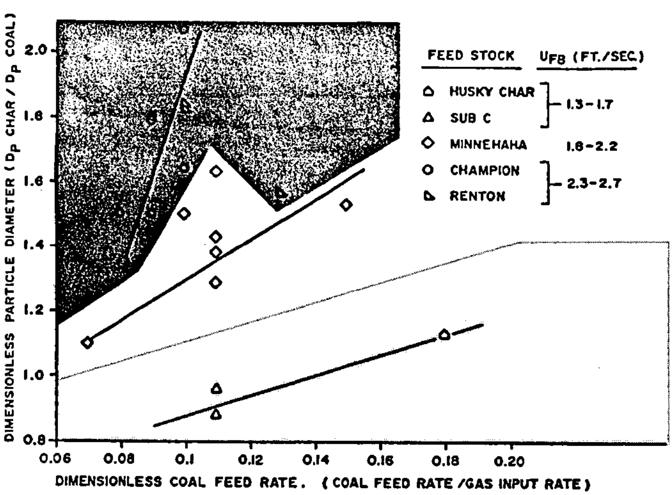


FIGURE 5 - TOTAL CHAR PRODUCTION VS FREEBOARD VELOCITY

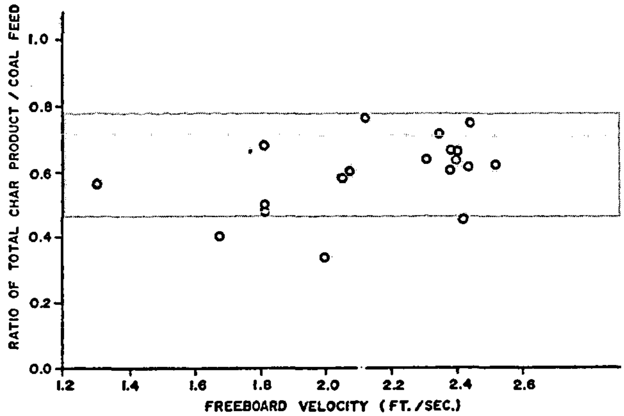
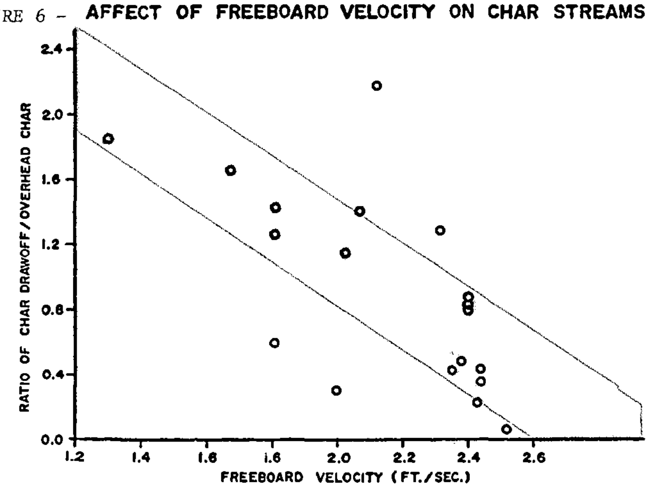
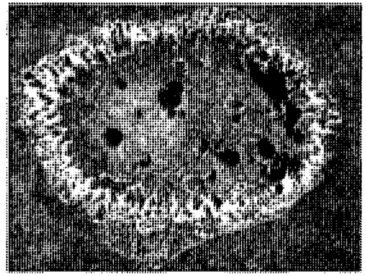


FIGURE 6 - AFFECT OF FREEBOARD VELOCITY ON CHAR STREAMS

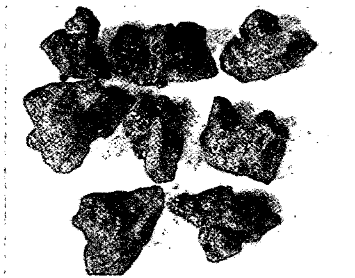


Sub-Bituminous-C 30X



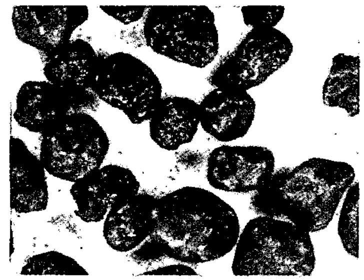
Minnehaha 30X

Champion 30X



Sub-Bituminous-C 10X

Minnehaha 10X



Champion 10X

FIGURE 7 - PHOTOMICROGRAPHS OF TYPICAL CHAR PRODUCTS

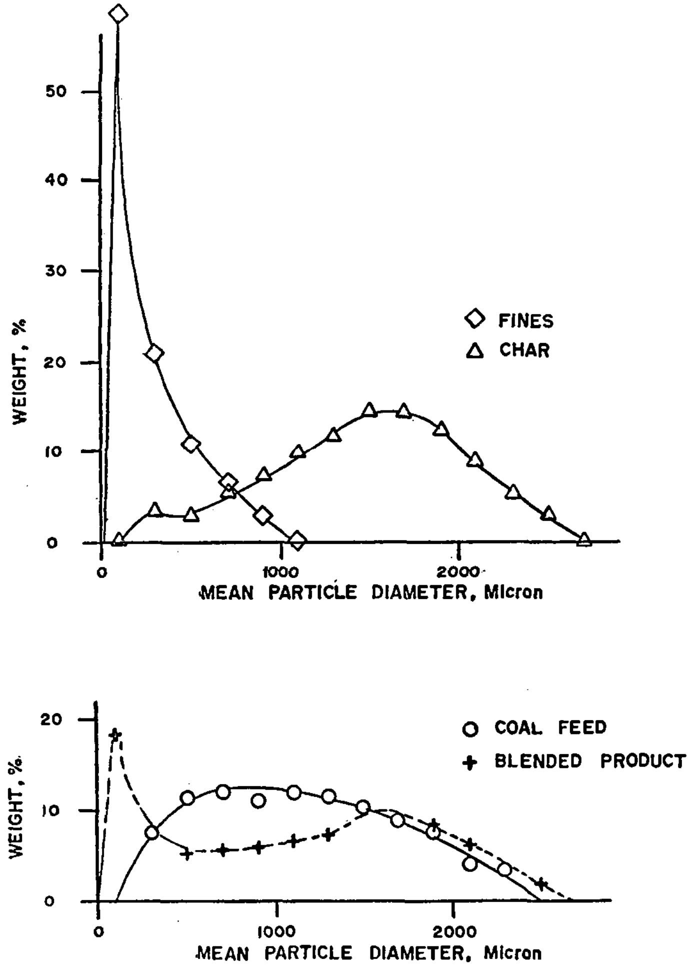


FIGURE 8 - PARTICLE SIZE DISTRIBUTIONS OF CHAR AND FINES PRODUCTS (MINNEHAHA COAL)

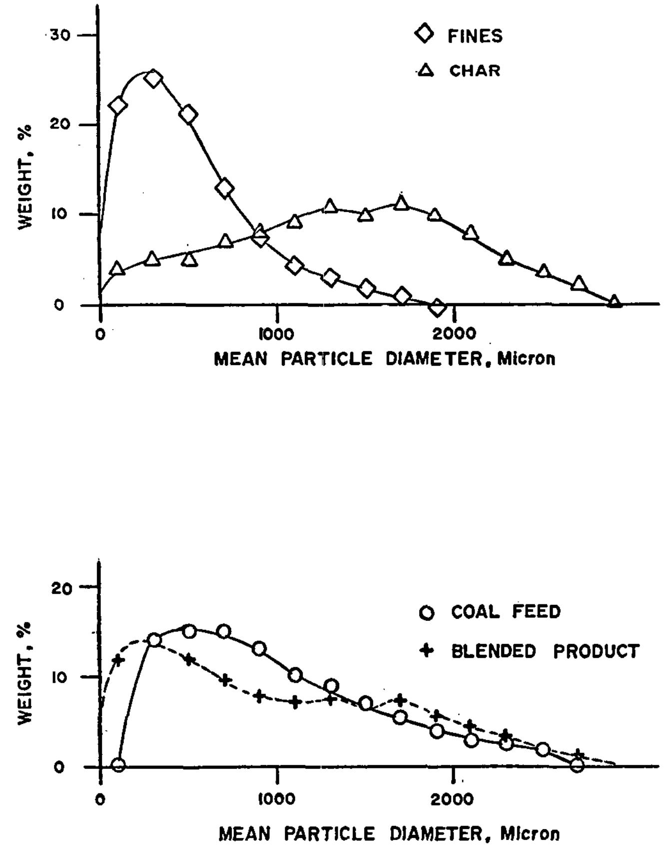


FIGURE 9 - PARTICLE SIZE DISTRIBUTIONS OF CHAR AND FINES PRODUCTS (RENTON COAL)