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**HIGH TEMPERATURE STAGED FLUIDIZED BED
COMBUSTION (HITS) BENCH SCALE EXPERIMENTAL
TEST PROGRAM CONDUCTED DURING 1980**

Final Report

By

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April 1981

Work Performed Under Contract No. AC21-79ET15167

**Aerojet Energy Conversion Company
Sacramento, California**

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FINAL REPORT

DESCRIPTION OF

HIGH TEMPERATURE STAGED FLUIDIZED BED COMBUSTION (HITS)
BENCH SCALE EXPERIMENTAL TEST PROGRAM CONDUCTED DURING 1980

**Contract Nos. DE-AC-01-79-ET15167 and
DE-AC-21-79-ET15167**

April 1981

Prepared For:

**Morgantown Energy Technology Center
U.S. Department of Energy**

By:

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**Aerojet Energy Conversion Company
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HITS Final Report - 1980

FORWARD

This report describes a program undertaken at Aerojet Energy Conversion Company (AECC) under Contract No. DE-AC-21-79-ET15167 with the Department of Energy (DOE), the objectives of which were to demonstrate the feasibility of a High Temperature Staged Fluidized Bed Combustion System (HITS) through design, fabrication and operation of a small-scale test system.

DOE Program Management and Technical Direction was by Arthur W. Hall of the Combustion Projects Branch of Project Management Division of the Morgantown Energy Research Center. John W. Byam and John S. Wilson are the respective managers of these organizations and Ralph Schafer is Deputy Manager of the Project Management Division.

Aerojet Energy Conversion Company's Program Management was by Mark Rudnicki and Richard Colman. Donald Jassowski of the Advanced Development Department headed by Ronald Newton was Project Engineer. The process design and data analysis were done by Roger Anderson, and testing was conducted at the Energy Engineering Laboratories under the direction of Thomas Harrington. Charles Gracey was responsible for test facility activation and test facility operation was the responsibility of Dr. Eugene Vander Wall and James Juchau.

This document comprises the final report under the contract.

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ABSTRACT

An experimental program was conducted to evaluate the process feasibility of the first stage of the HITS two-stage coal combustion system. Tests were run in a small (12-inch ID) fluidized bed facility at the Energy Engineering Laboratory, Aerojet Energy Conversion Company, Sacramento, California.

The first stage reactor was run with low (0.70%) and high (4.06%) sulfur coals with ash fusion temperatures of 2450° and 2220°F, respectively. Limestone was used to scavenge the sulfur. The produced low-Btu gas was burned in a combustor. Bed temperature and inlet gas percent oxygen were varied in the course of testing. Key results are summarized as follows:

- o The process was stable and readily controllable. It generated a free-flowing char product using coals with low (2220°F) and high (2450°F) ash fusion temperatures at bed temperatures of at least 1700° and 1800°F, respectively.
- o The gaseous product was found to have a total heating value of about 120 Btu/SCF at 1350°F. The practicality of cleaning the hot product gas and delivering it to the combustor was demonstrated.
- o Sulfur capture efficiencies above 80% were demonstrated for both low and high sulfur coals with a calcium/sulfur mole ratio of approximately two.
- o Gasification rates of about 5,000 SCF/ft²·hr were obtained. Coal input rates ranging from 40 to 135 lbm/hr were tested, as required to maintain the desired bed temperatures.
- o The gaseous product yielded combustion temperatures in excess of 3000°F when burned with preheated (900°F) air.

The above test results support the promise of the HITS system to provide a practical means of converting high sulfur coal to a clean gas for industrial applications. Sulfur capture, gas heating value, and gas production rate are all in the range required for an effective system. Planning is underway for additional testing of the system in the 12-inch fluid bed facility, including demonstration of the second stage char burnup reactor.

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1.0 INTRODUCTION AND SUMMARY

This report describes a fifteen-month program which was conducted to verify the feasibility of and obtain design data for a fluid bed coal combustion process. The High-Temperature Staged Combustion (HITS) Bench-Scale Test Program, being conducted under contract DE-AC-21-79-ET15167, is a study which has:

- o Demonstrated the feasibility of the HITS process through design, fabrication and operation of the bench-scale test system.
- o Yielded data relevant to char reaction rates, chemical equilibria, sulfur capture efficiency, and process gas combustion characteristics
- o Provided the foundation for an engineering design data base for subsequent scale-up to larger systems.

Empirical study of critical process parameters was required to verify process feasibility and to provide the basis for later efforts to scale-up system operation.

The HITS test system has served as a laboratory apparatus for conducting the bench-scale test program. The planned test program has been previously described in High Temperature Staged Fluidized-Bed Combustion (HITS) Bench-Scale Test Plan, Report AECC-2324-TR-01, dated 15 February 1980. This final report, submitted in partial satisfaction of contract requirements, includes: (1) description of program activities leading up to bench scale testing, (2) the test program, (3) data analysis procedures, (4) test results and conclusions, and (5) recommendations for further work towards commercial development of the HITS system.

1.1 The HITS Process

HITS is a two-stage process for environmentally sound combustion of coal. A high temperature, low-Btu gas is produced in the first stage fluid bed reactor in the presence of limestone for sulfur absorption. The unreacted char and spent bed material are then burned in a second-stage fluid bed combustor.

A concise process block diagram is shown in Figure 1. Several innovative aspects of the HITS process include:

- o Simplicity of design enhancing commercialization potential.

Figure 1. HITS Process Block Diagram

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1.1 (continued)

- o High-temperature (>3000°F) product gas combustion, extending the range of potential industrial applications.
- o Ability to utilize high-sulfur, low-quality coal in an environmentally acceptable manner.
- o Process stages are decoupled, allowing for changes in product gas char/ash production levels without impacting system stability.

Process characteristics are summarized in Table L

1.2 The HITS Bench-Scale Test System Design, Set Up and Activation

The HITS bench-scale test system is a laboratory prototype of the first stage reactor and product gas burner. With some modification, the same system can also be used for testing of the second stage combustor. The first-stage reactor has an internal diameter of 12-5/8 inches and operates at atmospheric pressure.

This one-ton per day bench-scale plant has been designed, fabricated and set up at AECC's Energy Engineering Laboratory at Nimbus, California. This system permits evaluation of critical aspects of the first-stage process on a reasonable scale prior to construction of a Process Development Unit (PDU). Figure 2 is a map of Aerojet's Nimbus, California facility, showing the location of areas utilized by the HITS program. The 12 sq. mile plant is located 15 miles east of Sacramento and is served by major rail transportation. Figure 3 is a plot plan of AECC's Energy Engineering Laboratories, the site of the HITS bench-scale test system. The laboratory has extensive engineering test and support equipment capabilities. Figure 4 is a plan view of the HITS facility, showing major system components. An isometric drawing of the system, showing the arrangement of the equipment, is given in Figure 5. A photograph of the test site after equipment installation is shown in Figure 6.

1.3 Testing

The objectives of the HITS bench-scale test program were to verify the first stage process design assumptions with regard to:

- o Char reaction rates
- o Chemical equilibria

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TABLE I

HITS Process Characteristics

First Stage

Nature of fluidization	Bubbling bed
Nature of reaction	Partial air oxidation/gasification
Fluidizing gas	Air
Air/fuel stoichiometry	Approx. 34%
Reaction temperature	Approx. 1600°F
Carbon conversion	Variable (nominally 65%)
Char residence time	Approx. 100 min
Ash and char removal	Primarily underflow
In-bed sulfur capture	Yes, with limestone or dolomite
Off-gas cleanup	As required

Second Stage

Type of reactor and reaction	Single-stage char and CaS burnup cell with in-bed water cooling for steam and/or power generation
Operating temperature	1600°F
Method of temperature control	In-bed boiler/steam generator system
Limestone/dolomite addition	No
Steam/power generation	Yes
Excess air	Low (10-30%)

Product Gas Burner

Air supply	Recuperatively preheated air
Nominal flame temperature	To 3400°F

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Figure 2. Aerojet Energy Conversion Company Sites Utilized by HITS Program

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Figure 3 AECC Energy Engineering Laboratories - HITS Bench-Scale Test System Site

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Figure 4 HITS Bench-Scale Test System - Equipment Layout

Figure 5 HITS Bench-Scale Test System - Isometric Layout

Figure 6 Photograph of HITS Test System

1.3 (continued)

- o Sulfur capture efficiency
- o Product gas flame characteristics

To verify these design assumptions, data were obtained during reaction of coal in a subscale fluidized bed reactor and subsequent combustion of the product gas in a burner. The bench-scale system processes up to one and a half ton per day of coal in a 12-inch diameter first stage reactor. This system is designed to operate over a range of test conditions so that the influence of process variables can be determined.

A total of fifteen operating conditions were tested in four groups of tests. Not including starting time and tests with coke, the system was operated with coal for a total of 185 hours. The longest test duration was 66 hours. The test program is summarized in Table 2. Test variables were:

- o Coal - two types: low sulfur (0.7%) high volatile A bituminous (UTAH H1B #1) and high sulfur (4.06%) high volatile A bituminous (Kentucky Elcome seam)
- o Bed temperature - 1500 to 1800°F
- o Oxidizing species - normal, depleted and enriched oxygen

1.4 Results and Conclusions

System performance closely followed design and pretest predictions in most areas. Sulfur capture, gas heating value and char reaction rates as required by the full scale process were obtained. Optimum operating temperature for the coals tested is between 1700 and 1800°F, rather than 1650 as originally planned. Extent of approach to equilibrium in the gaseous products was lower than theoretical, but in line with results with similar systems. Reaction of over 70% of the coal carbon was achieved. The char product formed was a free flowing, unconsolidated material which can be readily transported to the second stage reactor.

The reactor can be operated at bed temperatures of 1800°F for the high ash softening temperature coal and to 1700°F for the low ash softening temperature coal. Above these temperatures consolidation and clumping of the char can occur with eventual defluidization of the bed.

TABLE 2

Summary of HITS Bench Scale Testing

Test Group	Date 1980	Purpose	Coal	Test Variables Bed Temp. °F	Oxygen Conc, wt%	Number of Test Plan Points Run	Time on Coal for This Group, Hours	Cumulative Time on Coal, Hours
1	15-17 Aug.	System Checkout	Low Sulfur	1550 1650 1750	22	3	45	45
2	22-25 Aug.	Run Low Sulfur Tests	Low Sulfur	1500 1600 1700 1800	19, 22, 28	9	66	111
3	17-19 Sept.	Additional Low S. Tests	Low Sulfur	1600 1700 1800	19	3	49	160
4	1-2 Oct.	Run High Sulfur Tests	High Sulfur	1600 1700 1800	22	3	25	185

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1.4 (continued)

One unexpected result was noted. Relatively small reductions in oxygen concentration (decreasing input oxygen concentration from 22 wt% to 19 wt%) caused large reductions in sulfur capture efficiency with the low sulfur coal (from above 80% efficiency to below 40%).

The process was easily controlled. Bed temperatures could be set very accurately and maintained by either manual or automatic control of solids feed rate. Process performance can be controlled by changing bed temperature to vary the extent of char reaction and therefore the fraction of the coal energy which appears in the product gas from the first stage. This fraction approaches 0.7 as bed temperature is raised to 1800°F.

1.5 Recommendations for Further Work

Additional first stage tests should be run with a wider range of coals and limestone to better define the limits of the process. Maximum allowable bed temperatures should be more closely defined for each coal. Tests should be run with coal introduction at mid bed since there is potential that this will both permit an increase in average bed temperature and will provide higher gas heating values.

Testing should be conducted with larger, single-screened material to match what will be used in commercial practice. Some system sizing tests should be run with varying bed height and superficial velocity.

In addition to the above first stage tests, similar tests should be conducted with the unit modified to burn char to provide data on second stage operation.

2.0 TEST PLANNING, SYSTEM DESIGN, SETUP AND ACTIVATION

This section describes the program activity which preceded testing. It included planning of the overall program, development of a detailed test plan and detailed system specification. From these, detailed system and component designs were prepared. Following procurement and fabrication, the equipment was installed, checked out and calibrated, bringing the test system to the point where checkout tests could be run with coal.

2.1 Test Plan

A detailed test plan, Ref. 2, was prepared early in the program to define testing, and therefore system requirements. This plan defined the Test Matrix, Figure 7, where the nominal test variables or requirements are shown versus the test data objectives for each corresponding test. Note that the tests were divided into three test series, designated Test Series A, B and C. Each of the test series is discussed in detail in Sections 2.1.1, 2.1.2 and 2.1.3 respectively.

2.1.1 Test Series A

This test series was designed to establish gasification rates, complete process mass balances, sulfur capture efficiencies and the flame characteristics of the gas burner for a range of process operating conditions or variables. The variables planned for study in this series were of three types: (1) gasifier feed variables, (2) gasifier operating conditions, and (3) gas burner operating conditions. The gasifier feed variables included coal type and oxygen concentration in the reactant gas. The primary gasifier operating variable to be studied was reaction (bed) temperature, and the primary gas burner operating variable was air/fuel ratio. This entire test series was to be comprised of eight tests where were, in turn, divided into three parts as shown in Figure 7. The test series would, in fact, provide a total of twenty-four sets of data. The test variables and corresponding tests directed toward the evaluation of each variable are discussed in detail in the following sections.

NOMINAL TEST VARIABLES												TEST PURPOSE								
TEST			TYPE OF COAL	GASIFIER FEED		GASIFIER OPERATING CONDITIONS			BURNER CONDITIONS											
SERIES	NO.	ORDER		Ca/S MOLE RATIO	INLET GAS, % VOL O ₂	BED TEMP., °F	FEED LOCATION	SUPERFICIAL VEL. FT/SEC	BED DEPTH, FT	AIR INLET TEMP., °F	AIR/FUEL RATIO									
A	1a	4	LOW-S	2	21	1450	LOW	6	5	1100	VARIABLE	DETERMINE EFFECT OF BED TEMPERATURE AND INLET O ₂ CONCENTRATION ON SYSTEM OPERATION WITH LOW-SULFUR COAL								
	1b	9			30	↓														
	1c	12			15	↓														
	2a	2			21	↓														
	2b	7			30	↓														
	2c	11			15	↓														
	3a	1			21	↓														
	3b	6			30	↓														
	3c	5			15	↓														
	4a	3			21	↓														
	4b	8	30	↓																
	4c	10	15	↓																
	5a	16	HI-S	2	21	1450	LOW	6	5	1100	VARIABLE		DETERMINE EFFECT OF BED TEMPERATURE AND INLET O ₂ CONCENTRATION ON SYSTEM OPERATION WITH HIGH-SULFUR COAL							
	5b	21			30	↓														
	5c	24			15	↓														
	6a	14			21	↓														
	6b	19			30	↓														
	6c	23			15	↓														
	7a	13			21	↓														
	7b	18			30	↓														
	7c	17			15	↓														
	8a	15			21	↓														
	8b	20			30	↓														
	8c	22			15	↓														
B	9	27			LOW-S	2						21		TBD*	HIGH	6	5	1100	VARIABLE	EFFECT OF HIGH FEED INLET LOCATION FOR LOW AND HIGH SULFUR COALS
	10	28			LOW-S							↓		↓						
	11	25	HI-S	↓	↓															
	12	26	HI-S	↓	↓															
C	13	29	LOW-S	1.5 3.0 1.5 3.0	21	TBD*	TBD*	6	5	1100	TBD*	EFFECT OF Ca/S RATIO FOR LOW AND HIGH SULFUR COALS								
	14	30	LOW-S		↓	↓														
	15	31	HI-S		↓	↓														
	16	32	HI-S		↓	↓														

*TO BE DEFINED BASED ON THE RESULTS OF TESTS IN PRECEDING TEST SERIES; BEST VALUE WILL BE USED

TEST DATA OBJECTIVES
FOR ALL TESTS MEASURE:

- GASIFICATION RATES
- GAS/SOLID PRODUCT CHARACTERISTICS
- SULFUR CAPTURE EFFICIENCY
- PRODUCT GAS FLAME CHARACTERISTICS

Figure 7. Test Matrix

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2.1.1.1 Gasifier Feed Variables

Type of Coal

Two types of coal were to be evaluated in this test series: (1) a low sulfur, nonagglomerating, high volatile bituminous coal in Test 1 through 4, and (2) a low-quality high-sulfur, high volatile bituminous coal in Tests 5 through 8. These coals would have nominal particle sizes in the range of 16-35 US-Std mesh (0.5-1.19 mm). The coals and the basis for their selection are discussed in further detail in Section 2.2.2.1.

Reactant Gas

Reactant gases for gasifying the coals in this series were to be evaluated at three different oxygen concentrations: (1) normal air (approx. 21% vol. O₂) in test 1a, 2a,8a, (2) oxygen-enriched air (approx. 30% vol. O₂) in test 1b, 2b,8b, and (3) nitrogen-enriched air (approx. 15% vol. O₂) in tests 1c, 2c,8c. These variations were intended primarily to establish the effect of reactant gas concentration on gasification rates, i.e., establish apparent reaction order.

Other Feed Conditions

A single limestone was to be employed in all of these tests. The limestone would have a nominal particle size in the range of 16-35 mesh (0.5 to 0.19 mm) and would be utilized in proportion to the coals such that Ca/S mole ratio is nominally 2:1. The limestone selection is discussed in further detail in Section 2.2.2.2.

Coal and limestone feed rates would be controlled automatically in each test to maintain both the desired Ca/S mole ratio and the selected gasifier bed temperatures. Reactant gas feed rate would be fixed in each test to maintain the desired superficial velocity in the gasifier (nominally 6 ft/sec based on bed discharge rate).

2.1.1.2 Gasifier Operating Conditions

The effects of gasifier bed temperature on gasification rates, material balances, sulfur capture, and burner flame characteristics would be evaluated for each coal at four temperatures (nominally 1450, 1550, 1650 and 1750°F). The low bed temperature was to be utilized in tests 1a, b and c, and in 5a, b and c. The next higher bed temperature (approx. 1550°F) would be established in tests 2a, b and c, and in 6a, b and c.

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2.1.1.2 (continued)

Bed temperatures of approx. 1650°F would be utilized in tests 3a, b and c, and in 7a, b and c, and of approx. 1750°F, in tests 4a, b and c, and 8a, b and c. Bed temperatures would be maintained automatically by a control system which will modulate coal and limestone feed rates in parallel fashion in response to a sensed deviation between the actual bed temperature and a selected set-point.

Other Operating Conditions

Coal and limestone feed would be introduced directly into the bed of the gasifier by pneumatic transport. For all Series A tests, the feed inlet would be deep (low) in the bed, i.e., approximately 6 inches above the discharge plane of the reactant gas distributor. This location provides for near maximum interaction between initial coal devolatilization products, the char-ash in the bed, and the other gasification products. A later test series was planned for a high feed inlet location.

The superficial velocity in the bed increases from the bottom to the top as the reactant gas is converted to producer gas. For this reason, the bed superficial velocity at the top of the bed is most critical in terms of bed carryover. All tests were anticipated to be operated at a single value of superficial velocity (nominally 6 ft/sec) at the top of the bed. This velocity would be controlled by adjustment of the reactant gas feed rate. The velocity was selected based on previously conducted tradeoff analyses for a large-scale gasifier at AECC and upon experience in atmospheric fluidized bed combustors and gasifiers.

Bed depth in the gasifier was planned to be maintained in all tests at a nominal value of 5 ft (as fluidized). This value was selected based on anticipated gasification rates and tradeoff analyses conducted at AECC for a large-scale gasifier and consultants recommendations (Ref. 1). In view of the modest diameter of the test reactor (1 ft), some slugging was anticipated at a bed depth of 5 ft.

2.1.1.3 Burner Operating Conditions

Air/Fuel Ratio

The effect of the burner air-to-fuel ratio on the burner flame characteristics was planned to be determined in each test in Test Series A.

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2.1.1.3 (continued)

The range of air-to-fuel ratios to be evaluated was such as to define the maximum value for each of the measured flame characteristics, i.e., flame temperature, apparent emissivity and total radiant heat flux.

Other Operating Conditions

The air employed in the gas burner was to be preheated to 1100°F in each case. This value represents a temperature that can reasonably be anticipated in an actual application such as a cement kiln and is near the limits of readily available test facility air preheaters. If flame temperatures in the gas burner should exceed 3400°F (the upper design limit of the burner), a lower air preheat temperature would be selected to avoid burner damage.

2.1.2 Test Series B

Test Series B was planned to consist of four tests specifically designed to establish the effect of feeding coal and limestone into the gasifier bed at a location high in the bed. The effects on gasification rates, process mass balances, sulfur capture, and burner flame characteristics were to be established by direct comparisons of results from these tests with those obtained in directly-comparable Test Series A tests with the lower feed inlet location. The points of particular concern were how strongly feed input location affects: (1) product gas composition, especially in terms of gaseous hydrocarbons and tars, and (2) sulfur capture efficiency. Feed introduction deep in the bed should tend toward the desirable reduction of complex devolatilization products (e.g., tars) and sulfur-bearing components in the producer gas by virtue of relatively longer gas residence times. On the other hand, feed introduction high in the bed should tend toward increased gas production rates and/or gases of higher heating value since devolatilization would occur in a relatively oxygen-poor region and with relatively less oxygen consumption.

The test conditions planned for these four tests (Tests 9-12) are shown in the Test Matrix given in Figure 7. Two tests at different gasifier bed temperatures were to be conducted with each of the two coals, i.e., the low-sulfur and high-sulfur coals. The feed inlet location for these four tests was planned to be approximately one foot below the mean surface of the fluidized bed. The specific temperatures were to be defined based on the results of tests conducted in Test Series A. The basis for making the temperature selection would be primarily a trade-off between high gasification rates and high sulfur capture efficiencies. It was anticipated that the two selected temperatures would be near 1600°F.

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2.1.2 (continued)

These tests would involve only the use of air as the gasifier reactant gas. All other operating conditions would be the same as for those in Test Series A, as discussed in the preceding section.

2.1.3 Test Series C

Four tests were planned for this test series. They were specifically designed to establish the effect of the Ca/S mole ratio in the coal/limestone feed to the gasifier, especially on the sulfur capture efficiency, but also on gasification rates, process mass balances, and burner flame characteristics. The effects of the Ca/S mole ratio were to be defined by direct comparisons of performance data from these tests with that obtained in the directly comparable tests in Test Series A or B.

The test conditions for these four tests (Tests 13-16) are shown in the Test Matrix. Two tests would be conducted with each of the two coals (low-sulfur coal and high sulfur coal) at two different Ca/S mole ratios. The nominal Ca/S mole ratios to be evaluated in these tests are 1.5:1 and 3:1. Air would be used in each of these tests as the reactant gas. Each test would also be conducted at a selected gasifier bed temperature, solids feed location, and gas burner air/fuel ratio. These selected conditions would be based on analyses of data from tests in Series A and B. The other operating conditions (superficial velocity, bed depth and burner air preheat temperature) would be held the same as in the tests in Series A and B.

2.2 Bench-Scale Test System

The bench-scale HITS test system was developed to provide critical process data under approximately full-scale operating conditions. The design requirements of the bench-scale test system are presented in the following sections, while the testing to be accomplished in this system is summarized in Section 2.1 and documented in detail in Reference 2. The process design has been chosen to eliminate complex handling of hot solids and gaseous streams with complete decoupling of major system components. This serves to eliminate the possibility of system instabilities caused by process interactions. In addition, the added complexity and expense of steam, oxygen or flue gas injection is avoided. Finally, the process does not require any complex scheme for regeneration of the sulfur sorbent.

2.2.1 Design Tradeoffs/Requirements

2.2.1.1 Primary Reactor

Optimization of design and operating conditions for the bench-scale primary reactor involves a number of variable factors, including test scale, temperature, pressure, air/coal ratio, limestone/coal ratio, bed particle size, superficial velocity, and bed depth. Since many of these factors are related, consideration of tradeoffs is required. Primary reactor design criteria are listed in Table 3.

The primary system requirement involves production of a fuel gas which can be used in high temperature industrial processes requiring flame temperatures above 3000°F. The system must do this efficiently and inexpensively while meeting environmental emission requirements. To be commercially practical, the process must function with a range of commonly available coals.

Operating pressure was one of the simplest parameters to establish for this process since it does not have a profound influence on the process chemistry. The producer gas from the reactor need be only slightly above atmospheric pressure. Thus, average bed operating pressure was established at approximately 2.5 psig. A higher pressure could reduce reactor diameter, but only with a significant increase in fluidization air compression cost and reactor design complications including pressurized solids feed and bed removal subsystems and producer gas-throttling or expansion equipment.

The selection of nominal operating temperature primarily involves a tradeoff between the higher temperatures, desired to obtain improved char gasification rates, and the somewhat lower temperatures, desired to achieve maximum sulfur scavenging. The selected nominal temperature is 1600°F, but was subject to variations from approximately 1500 to 1800°F during testing.

The desired bed operating temperature essentially defined the required air/coal feed ratio since only sufficient air was desired to balance the exothermic heat of combustion with the endothermic heat of gasification and reactor heat losses. For the nominal bed operating temperature of 1600°F, this amounts to approximately 2.5 lb air/lb coal (neglecting reactor heat losses which must be minimized).

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TABLE 3

PRIMARY REACTOR DESIGN CRITERIA

	Nominal	Range
Operating pressure	2.5 psig	—
Nominal bed temperature	1600°F	1450-1750°F
Limestone/coal ratio		
— Low sulfur coal	.050 lb/lb	.033-.066 lb/lb
— High sulfur coal	.35 lb/lb	.23-.47 lb/lb
Bed depth	5 ft	—
Superficial velocity	6 ft/sec	—
Mean bed particle size	.0315 in	—

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2.2.1.1 (continued)

The limestone/coal ratio depends primarily on the sulfur content of coal and secondarily on the degree of sulfur scavenging desired. As a nominal baseline design value, a limestone/coal ratio corresponding to a Ca/S molar ratio of 2 was selected. For a low-sulfur coal containing 0.7% wt sulfur, the corresponding limestone/coal ratio is about 0.044 lb/lb, and for a high-sulfur coal containing 4% wt sulfur, the limestone/coal ratio is about 0.25 lb/lb.

Bed particle size, superficial velocity, and bed depth are very strongly related and require careful consideration of: (1) minimum fluidization, terminal settling, and approximate optimum superficial velocities versus particle diameter, (2) combustion/-gasification reaction kinetics, and (3) the combined effects of these factors on reactor size (bed diameter, bed depth, and freeboard height) and cost. Based on a detailed analysis of a full-scale system conducted prior to the current program, the approximate optimum superficial velocity at the upper bed surface was defined to be 6 ft/sec, the bed depth to be 5 ft, and the mean bed particle diameter to be 0.0315 inches (800 microns). That primary reactor design tradeoff analysis is presented in Appendix B. Because the basic approach involves testing under full-scale conditions, the values for superficial velocity, bed depth and particle size given above are also appropriate for the design of the bench-scale test system.

The physical size of the reactor and scale of testing are practically synonymous. The most appropriate size or scale factor for bench-scale work is largely established by a judgment taking into account both economic and technical factors. For this program, the most suitable reactor size was judged to be one having a nominal inside diameter of 12 inches. That size corresponds to a nominal coal feed rate of about 1 ton/day, which is approximately 1/275th of full-scale.

The reactor must be capable of accepting feed injection at several different bed levels and of discharging char/ash from different bed levels. It was required to operate with minimal heat losses and to have extensive instrumentation ports.

2.2.1.2 Auxiliary Components

The major auxiliary components required to support the primary fluidized-bed reactor include: (1) limestone and coal feed systems, (2) gas supply system, (3) off-gas ducting and cleanup system, (4) char/ash removal system, (5) off-gas combustor, and (6) process instruments and controls.

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2.2.1.2 (continued)

The limestone and coal feed systems are required to continuously supply the primary reactor at controlled rates. Materials are continuously metered from separate, pressure-isolated feed hoppers by variable-speed screw conveyors. A pneumatic line feeds the primary reactor. The system includes weighing equipment and remote speed controls for the metering conveyors.

The gas supply is required to provide fluidization gases (air, oxygen-enriched air, and nitrogen-enriched air) to the primary reactor at controlled rates and concentrations. It also must provide air for firing the primary reactor preheater, gases for pneumatically transporting the coal and limestone to the reactor, and air for burning the reactor off-gases. Air is supplied from a 100 CFM compressor, nitrogen from a 6000 gal. vaporizer and oxygen from 100-liter dewars. The gas supply system includes appropriate gas metering equipment.

The off-gas ducting and cleanup system transports the reactor off-gas to the off-gas combustor with minimal heat losses. A centrifugal separator, with a recycle loop for conveying fines back into the reactor, removes the larger particulates. Off-gases are sampled before and after the cleanup step. The system also includes a gas measuring orifice.

The char/ash system removes hot char/ash from the primary fluidized-bed reactor at a controlled rate, cools the material, and transfers it to a closed receiver. This system includes water-cooled ducting, a water-cooled variable-speed screw conveyor, and an isolatable closed receiver arrangement.

The off-gas combustor system burns the reactor off-gases with preheated air for minimal heat losses. The design provides means for determining the resulting flame temperatures and emissivities at controlled air-to-fuel ratios. It includes a burner and combustion chamber, an air preheater, flame measurement instruments, and an exhaust stack.

The instrumentation and control system is required to provide the necessary basic process operating parameters and controls for evaluating the HITS process and operating the overall system semi-automatically in a fail-safe manner.

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2.2.2 Process Design Data

Based on the design requirements and tradeoffs appropriate to the bench-scale test system as described in Section 2.2.1, a basic process flow diagram has been developed and is presented in Figure 8. This diagram defines all the primary process streams, along with the nominal anticipated operating conditions and flowrates (mass balances). It has been used as a basis for equipment sizing and developing the detailed designs of all of the major process equipment. The various process streams identified in the process flow diagram are discussed in further detail in the following subsections.

2.2.2.1 Coal Feeds

Two types of coal were evaluated in this program: (1) a baseline low-sulfur, non-agglomerating, high-volatile bituminous coal and (2) a high-sulfur, high-volatile high swelling index bituminous coal. These coals were ground and screened to nominal particle sizes of 15-35 mesh (0.5-1.19 mm). Their characteristics and the basis for their selection are discussed below.

System activation and shakedown, along with system baseline operational characteristics, were determined with the first coal. This is a high-quality, Utah high-volatile 'A' bituminous coal. Its low-sulfur content and low-agglomerating characteristics were chosen to minimize operational problems. An analysis of this coal is given in Table 4.

The second coal tested in the bench-scale unit to demonstrate system performance over a range of coal types was an Eastern Kentucky coal. This coal had a high sulfur content and a greater tendency to agglomerate, with a relative low ash softening temperature and high free swelling index. The HITS process is intended to provide trouble-free operation with this type of coal, with environmentally acceptable emissions levels. An analysis of this coal is given in Table 5.

	A	B	C	D	E	F	G	H	I	J	K	L	M	N
TEMPERATURE, °F	77	77	77	77	386	1400	1300	3200	77	1100	77	200	77	77
PRESSURE, PSIA	19.7	19.7	19.7	19.0	19.0	15.4	15.1	14.8	15.2	15.1	100	17.7	19.7	19.7
FLOW RATE:														
GASES:														
SCFM (@70°F)	—	—	44.57	3.86	2.45	77.89	76.06	141.90	78.89	78.89	0.66	—	5.93	14.86
ACFM	—	—	33.58	3.03	3.03	260.96	245.92	973.58	77.31	226.15	0.098	—	4.48	11.23
LB/HR	—	—	200.26	17.35	9.85	297.66	290.67	645.17	345.50	354.50	2.86	—	29.44	64.58
SOLIDS:														
LB/HR	85.68	24.85	—	—	17.45	18.33	0.88	0.88	—	—	—	39.45	—	—

Figure 8 HITS Bench-Scale Process Flow Diagram

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TABLE 4

Laboratory Analysis of Low Sulfur Coal

Utah HIB #1 (High Volatile A, Bituminous)

<u>Proximate Analysis</u>			<u>Ultimate Analysis</u>		
	<u>As Rec'd.</u>	<u>Dry Basis</u>		<u>As Rec'd.</u>	<u>Dry Basis</u>
% Moisture	2.79	—	% Moisture	2.79	—
% Ash	11.30	11.62	% Carbon	68.73	70.70
% Volatile	41.56	42.75	% Hydrogen	5.29	5.44
% Fixed Carbon	<u>44.35</u>	<u>45.63</u>	% Nitrogen	1.04	1.07
	100.00	100.00	% Chlorine	0.03	0.03
Btu/lb.	12440	12800	% Sulfur	0.70	0.72
% Sulfur	0.70	0.72	% Ash	11.30	11.62
			% Oxygen (diff)	<u>10.12</u>	<u>10.42</u>
				100.00	100.00

Free Swelling Index (ASTM D-720) = 1.5

Ash softening temperature = 2450°F

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TABLE 5

Laboratory Analysis of High-Sulfur, Moderately Agglomerating Coal
Used in Test Program

Eastern Kentucky, Elcome Seam (High Volatile A, Bituminous)

<u>Proximate Analysis</u>			<u>Ultimate Analysis</u>		
	<u>As Rec'd.</u>	<u>Dry Basis</u>		<u>As Rec'd.</u>	<u>Dry Basis</u>
% Moisture	1.48	—	% Moisture	1.48	—
% Ash	14.25	14.46	% Carbon	69.10	70.14
% Volatile	34.57	35.08	% Hydrogen	4.87	4.95
% Fixed Carbon	<u>49.70</u>	<u>50.46</u>	% Nitrogen	1.49	1.51
	100.00	100.00	% Sulfur	4.06	4.12
Btu/lb.	12,573	12,761	% Ash	14.25	14.46
% Sulfur	4.06	4.12	% Oxygen (diff.)	<u>4.75</u>	<u>4.82</u>
Mineral matter free Btu/lb.		15,290		100.00	100.00

Free Swelling Index (ASTM D-720) = 6.0

Fusion Temperature of Ash (°F)	<u>Reducing</u>
Initial Deformation	2220
Softening	2280
Hemispherical	2440
Fluid	2500

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2.2.2.2 Limestone Feed

A single limestone, Colorado-Lien "high calcium granules" was employed in all the bench-scale tests conducted. It had a nominal particle size range of 16-35 mesh (0.5-1.19 mm). The analysis of the limestone is given in Table 6.

2.2.2.3 Fluidization Gas

The primary reactor fluidization gas, process stream 'C', is normally air, used in a quantity such that the superficial velocity at the top of the fluidized-bed reactor is 6 ft/sec. That condition is the design basis for the bench-scale process shown in Figure 8. In certain tests, the air was to be enriched with oxygen, process stream 'M', to produce a fluidizing gas containing approximately 30% vol. O₂. In other tests, the air was to be enriched with nitrogen, process stream 'N', to produce a fluidizing gas containing approximately 15% vol. O₂. The nominal quantities of oxygen and nitrogen required for these purposes are shown in Figure 8.

2.2.2.4 Producer Gas

The producer gas, process stream 'F', generated in the fluidized-bed reactor, was assumed as a design basis to be an equilibrium composition of the elements C, H, O and N at 1600°F and to be free of sulfur-containing gases. It was also assumed to carry a nominal particulate burden of 27.5 grains per standard cubic foot. In actual practice, the producer gas was anticipated to be appreciably richer in CO₂, H₂O, hydrocarbons, and sulfur compounds, and leaner in CO and H₂. The assumed particulate loading was believed to be a reasonable worstcase value. The design basis producer gas composition is summarized in Table 7. The composition of the particulates was assumed to be the same as that of the char/ash.

2.2.2.5 Char/Ash

The char/ash, process stream 'L' and the solid particulate fraction of process streams 'E', 'F', and 'G' were assumed as a design basis to be composed of a mixture of char, coal ash, and spent excess calcined limestone. The particular composition expected for the mixture is summarized in Table 8.

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TABLE 6

Analysis of Limestone Used in HITS Bench Scale Testing

Identity: Colorado-Lien "High Calcium Granules"

A. Laboratory Analysis

<u>Component</u>	<u>As-Received, % wt</u>
CaCO ₃	94.60
MgCO ₃	0.80
CaSO ₄	0.30
Other (Mineral Matter)	4.00
Moisture	<u>0.30</u>
	100.00

B. Composition Used for Data Reduction

<u>Component</u>	<u>As-Received, % wt</u>
CaO	53.00 *
MgO	0.38 *
CO ₂	42.01
Sulfur	0.07
Mineral Matter	4.30 **
Moisture	0.30

*Equivalent to 38.26 % wt Ca (includes both CaO & MgO)

**CaSO₄ + Other

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TABLE 7

Product Gas Composition
Design-Basis: Theoretical Equilibrium

<u>Species</u>	<u>% Volume</u>	
	<u>Overall Basis</u>	<u>Moisture-Free Basis</u>
N ₂	51.55	53.38
CO	25.18	26.07
H ₂	14.79	15.31
CO ₂	5.06	5.24
H ₂ O	<u>3.42</u>	<u>—</u>
	100.00	100.00
Mean Mol. Wt.	24.64	24.87
Particulate Loading = 27.5 gr/SCF		

TABLE 8

Composition of Char/Ash - Design-Basis

<u>Species</u>	<u>% wt</u>	<u>mole %</u>
Carbon, C	29.34	69.76
CaS	22.21	8.79
CaO	17.26	8.79
SiO ₂	16.14	7.67
Al ₂ O ₃	8.44	2.36
FeO	<u>6.61</u>	<u>2.63</u>
	100.00	100.00

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2.2.2.6 Fines Recycle

As a design basis, the fines recycle stream, 'E', was a mixture of producer gas, char/ash particulates, and nitrogen (process stream 'K'). The nitrogen served as the motive gas for eduction of the mixture of producer gas and the cyclone fines back to the fluidized-bed reactor. For design purposes, 95% wt. of the particulates in the producer gas, stream 'F', were assumed to be captured by the cyclone and returned to the reactor via the fines recycle stream.

2.2.2.7 Cleaned Producer Gas

The producer gas, after being cleaned of most of the particulates by the cyclone, process stream 'G', is fed directly to the gas burner. For design purposes, this stream was assumed to have the same composition as the raw producer gas, but a particulate loading of only 5% of that of the uncleaned gas.

2.2.2.8 Burner Combustion Products

The combustion products of the gas burner, process stream 'H', were assumed for baseline design purposes to be in chemical equilibrium and to result from firing with a 10% excess of preheated air (stream 'J').

2.2.3 Component Design and Specification

Design information for major HITS subsystems is contained in the following sections. Major components are summarized in Table 10, which also identifies the equipment with its symbol on the Piping and Instrumentation Diagram (P&ID), Figure 9. Major subsystems include:

- o Fluid bed reactor
- o Solids feeders
- o Char removal system
- o Off-gas piping
- o Product gas combustor

Particle and gas analysis systems are described here; combustor temperature and heat flux instrumentation are described in Appendix A. Finally, the interrelation of these systems through the process controls and instrumentation is described by reference to the P&ID.

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TABLE 9

Major Equipment List
Bench-Scale First-Stage Test Program

<u>Item</u>	<u>P&ID Identification</u>
A. Fluid Bed Reactor	
1. Windbox/Preheat Burner	B-1
2. Tuyere Plate	
3. Bed Section	R-1
4. Disengaging Section	
5. Feed, Removal and Instrumentation Ports	
B. Solids Feed System	
1. Coal Refill Hopper	H-1
2. Coal Feed Hopper	H-3
3. Limestone Refill Hopper	H-2
4. Limestone Feed Hopper	H-4
5. Coal Screw Feeder	G-1
6. Limestone Screw Feeder	G-2
7. Eductor and Plumbing	G-3
C. Char Removal System	
1. Char Screw Feeder	G-4
2. Char Removal Piping	M-4
3. Char Storage Hopper	H-6
D. Off-Gas Piping and Fines Recovery	
1. Insulated Pipe Sections	M-1, M-2
2. Cyclone	F-1
3. Fines Injector	
4. Fines Hopper	H-5
E. Product Gas Combustor	
1. Fuel Gas Flow Orifice	FE-602
2. Burner Assembly	B-2
3. Combustor Assembly	B-4
4. Pilot Igniter	B-3
5. Exhaust Gas Stack	S-1
F. Gas and Particulate Analysis System	
1. Gas Sample Probe	
2. Particle Collection System	
3. Gas Analysis System	

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Table 9 (cont.)

<u>Item</u>	<u>P&ID Identification</u>
G. Support Structure	
H. Controls & Instrumentation	
1. Gage Panel	
2. Control Console	
3. Data Logger	
4. Temperature Recorder	
5. Optical Pyrometer	
6. Calorimeter	
L. Auxiliary Equipment/Systems	
1. Fluidizing Air Compressor	C-1
2. Combustor Air Blower	C-2
3. Combustor Air Heater	E-1
4. Propane Supply	
5. Nitrogen Supply	
6. Oxygen Supply

Figure 9 HITS Bench - Scale Piping and Instrumentation Diagram

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2.2.3.1 Fluidized-Bed Reactor

Based on process requirements, a set of design criteria was established for the fluidized-bed reactor early in the program. These criteria are presented in Table 10. Reactor design procedures and tradeoffs, detailed in Appendix B, resulted in a system which consists of four refractory-lined assemblies: (1) a propane-fired preheater, (2) a flanged fluidized-bed section, (3) a flanged freeboard section, and (4) a flanged top closure. The detail designs of these assemblies are discussed in the following subsections and are shown in Figures 10 and 11.

Reactor Preheater Assembly

The preheater assembly is a flanged unit which mates with the fluidized-bed section of the reactor. It consists of a horizontally-fired burner section and a vertically intersecting windbox section enclosed by a carbon steel shell. Both parts are internally insulated with two layers of castable refractories. The entire assembly is mounted on castors to permit movement of the unit when it is disconnected from burner supply lines and at the preheater/bed flange joint.

The preheater assembly is designed to operate at pressures from 0-10 psig. It includes a propane-fired burner, capable of firing at pressures up to 10 psig and having a rated capacity of approximately 250,000 Btu/hr. The burner includes an igniter, flame safety supervision, valves, manual controls, and appropriate pressure reducers.

The assembly also includes: (1) a flame view port with a quartz window, (2) a carbon steel coupling for burner/fluidization air, (3) two horizontal instrumentation ports approximately 6 inches below the preheater/bed flange joint, (4) a port to accommodate the burner, its supplies and accessory components, and (5) jacking pads for raising the preheater.

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TABLE 10

Reactor Design Criteria

Size:

Inside reactor diameter, in.	12-5/8
Bed height, ft.	5
Freeboard height, ft.	6

Operating Conditions:

	<u>Nominal</u>	<u>Range</u>
Inlet (windbox) pressure, psia	20	14.7-25
Exit (discharge) pressure, psia	15	10-20
Bed temperature, °F	1600	1000-1800
Fluidization gas inlet temp, °F		
During startup	1000	800-1200
During steady state	Ambient	50-100
Fluidization gas(air) flowrate, lb/hr	200	140-280
Feed rates, lb/hr		
Primary solid (16x35 mesh coal)	85	25-250
Secondary solid (16x35 mesh limestone)	10	3-30
Pneumatic transport air	18	10-30
Solids (char/ash) discharge rate, lb/hr	30	10-100

Other Design Goals:

Startup period	≤ 8 hr
Steady state heat losses	$(T_{bed} - T_{exit}) \leq 200^{\circ}F$
Materials of construction	Compat. with reducing, oxidizing & sulfur-containing atmospheres.
Distributor	Non-sifting type
Bed drain arrangement	To minimize bed residual

Interfaces:

Solid feed nozzles (3)	1" flanged pipe
Solid discharge nozzles (3)	3" flanged pipe
Instrumentation ports (10)	1" pipe couplings
Cyclone fines recycle port (2)	1" flanged pipe coupling
Fluidization air inlet nozzle (1)	2" flanged pipe
Hot gas discharge nozzle (1)	4" flanged pipe
Spare reactor head nozzles (2)	2" flanged pipe

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Figure 10 HITS Reactor Assembly Drawing

Figure 11 Reactor Assembly

2.2.3.1 (continued)

Fluidized-Bed Section

The fluidized-bed section is a flanged unit which joins the preheater assembly and the freeboard section. It consists of a 1/4-inch thick carbon steel shell approximately 7'-9" inches long by 36" O.D., which is lined with two layers of insulating castable refractories to an inside diameter of 12 5/8". The refractory linings in the assembly are retained by anchors welded to the shell. This section includes a tuyere plate as shown in Figure 12. The plate is retained between the lower bed section and preheater flanges. This section also includes a structurally-reinforced section and attachment pads for supporting the entire fluidized-bed reactor system.

This assembly is designed to operate at pressures from -5 to +10 psig, fluidization gas inlet temperatures up to 1200°F, and bed temperatures up to 1800°F. The stainless steel tuyere plate is of the tapered-top-hat type design; it is essentially non-sifting. It is designed to have a nominal pressure drop of 12" W.C. at a nominal airflow rate of 35 ACFM, and to be capable of withstanding a pressure drop of up to 5 psi at 1200°F. The pitch on the tuyere risers is approximately 2-1/2". The tuyere plate is retained at, and affixed to, the lower flange joint by easily removed bolts capable of supporting the plate when the preheater/bed flange joint is disconnected.

The assembly includes seven 1-inch instrumentation ports, three 1-inch feed ports, and two 1-inch cyclone underflow ports; all ports include stainless steel pipe extensions through the shell and outer layer of refractory. It also includes three 3-inch flanged stainless steel discharge nozzles with internal extensions through the shell and outer layer of refractory, oriented 45° upward from the shell. The orientations and elevations of the nozzles are as indicated in the table on Figure 10. All nozzles and instrumentation ports extend outward from the shell a sufficient distance to protrude through one inch of external insulation and to permit ready accessibility for installing mating connectors.

Freeboard Section

The freeboard section is a flanged unit which fits above the fluidized-bed section. It consists of a 1/4-inch thick carbon steel shell approximately 6' long by 36" O.D., which is lined with two layers of castable refractories to an inside diameter of 18 5/8".

Figure 12 Tuyere Assembly

2.2.3.1 (continued)

This assembly is designed to operate at pressures from -5 to +5 psig and temperatures up to 1800°F. It includes five 1-inch instrumentation ports (1" pipe couplings with internal stainless steel extensions through the shell and outer layer of refractory, oriented in the horizontal plane). The elevations and radial orientations of these ports are as indicated in the table on Figure 10. The ports extend outward from the shell a sufficient distance to protrude through one inch of external insulation and to permit ready accessibility for installing mating connectors. The refractory linings in the assembly are retained by anchors welded to the shell.

Reactor Top Closure

The top closure of the fluidized-bed reactor is a flat, refractory-lined, flanged assembly penetrated by three vertical flanged discharge nozzles. This assembly is designed to operate at pressures from -5 to +5 psig, and temperatures up to 1800°F. The assembly includes one 4-inch and two 2-inch flanged stainless steel nozzles. The 4-inch nozzle is designed to operate hot, but to minimize heat loss to the shell and thermal stresses at the nozzle/shell joint. The two 2-inch nozzles are supplied with blinds. The 4-inch nozzle is placed off-center. The closure is equipped with lifting lugs designed to permit hoisting of the top closure/freeboard section as a unit. The refractory linings in the assembly are retained by anchors welded to the shell.

2.2.3.2 Feed System

The solids system for the HITS bench-scale unit consists of three main components: (1) the volumetric solids feeders, (2) the active hopper weighing system, and (3) the pneumatic transport to the reactor. It is designed to deliver a steady quantity of measured coal and limestone to the fluid bed reactor. The overall system is shown schematically in Figure 13. Figure 14 is a photograph of the feeders. Table 11 summarizes operating parameters for the solids feed systems.

Each of the volumetric feeders in the system is attached to a series of two hoppers. Each volumetric feeder consists of a 2-inch diameter screw feeder, driven by a SCR unit attached to an active feed hopper. The SCR control unit varies feeder speeds relative to the input control signal. This controller maintains proper bed temperature by varying coal and limestone feed rates.

Figure 13 Solids Feed System Schematic

Figure 14 Solids Feed System

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TABLE 11

Operating Parameters for Solids Feed System

Coal

1. Minimum flowrate, ft ³ /hr.	0.4
2. Maximum flowrate, ft ³ /hr.	4.0
3. Nominal flowrate, ft ³ /hr.	1.8
4. Approximate density, lb/ft ³ .	50
5. Active hopper capacity, ft ³ .	6
6. Refill hopper capacity ft ³ .	4
7. Particle size, mesh	16 x 35

Limestone

1. Minimum flowrate, ft ³ /hr.	0.06
2. Maximum flowrate, ft ³ /hr.	0.35
3. Nominal flowrate, ft ³ /hr.	0.18
4. Approximate density, lb/ft ³ .	85
5. Active hopper capacity, ft ³ .	2
6. Refill hopper capacity, ft ³ .	1-1/2
7. Particle size, mesh	16 x 35

Pneumatic Transport

1. Air/nitrogen available for transport, lbm/hr. (ambient temp.)	25
2. Nominal line size, in. L.D.	1/2

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2.2.3.2 (continued)

The active feed hoppers are sized to yield approximately 3 hours of operation without refill. Each active feed hopper is designed to operate at 10 psig. The normal maximum operating pressure was about 3 psig. During hopper refill operations, the active hopper is isolated from the feeder and then vented to the stack to depressurize. A small reservoir above the screw feeder allows continuous feeding during refill (approximately 3 minutes). Once vented, the active hopper is filled from the refill hopper. The active hopper is then pressurized to operating pressure, and flow is resumed to the feeder. Each active hopper has a low-bin indicator warning which alerts the operator that the hopper is about 80% depleted.

To reduce the time to refill the active hopper, a refill hopper is provided directly over each active hopper. This hopper holds approximately two thirds of the volume of the connected active hopper to prevent overfilling. A six mesh per inch screen is also provided on the top of each refill hopper to keep oversize material out of the feed system.

Mechanical scales provide an accurate measure of the weight of the coal and limestone feed hoppers to within $\pm 1/2$ lb. The feeders and active hoppers of both the coal and limestone are each supported by their own weigh platform. Plumbing and electrical connections to the feeders and hoppers are by stiff flexure-type couplings, perpendicular to the movement of the weigh platforms. In this way, the overall weighing system can be calibrated to give accurate weight changes. A 1000 lb. capacity dial scale is used to determine coal flowrates, and a 1000 lb. capacity single beam balance is used to measure the limestone flowrates.

The feed streams from the coal and limestone feeders are mixed and conveyed by air at the solids pickup. This mixture of solids and gas is conveyed to the reactor through a flexible, 1/2 inch I.D. nylon-reinforced tube. This tube is connected to a 1/2 inch stainless steel tube at the reactor; the latter penetrates through the refractory wall into the bed. Two feed points are located 6 and 54 inches above the tuyere discharge; the lower point was used in all testing.

Air and nitrogen are both available to pneumatically transport material to the reactor. A flow controller assures a constant flow of transport gases. The solids pickup can also be bypassed to flow only gas through the feed circuit to the reactor when required.

2.2.3.3 Product Gas Combustor

The product gas combustor has two major functions. First, it operates as a flare to burn the gasifier output to reduce discharge of hazardous gases to the atmosphere. Second, it serves as a high temperature burner to permit evaluation of the combustion characteristics of the product gas.

The combustor is designed to provide a stable, high-temperature flame from the product gas. To this end, the combustor features both swirl and bluff body stabilization, along with intense mixing of the fuel and air streams to maximize reaction rates. This design results in a stable, high-intensity, high-temperature flame. The resultant flame shape is a Type II flame, as described by the International Flame Research Foundation (IFRF) classification system (Ref. 3). This flame type has been documented to achieve high flame temperatures and complete burnout of the fuel in a short length (Ref. 4). Stabilization by swirl and bluff bodies is a well-known technique, documented in the literature (Ref. 5) and in industrial experience (Ref. 6). The product gas combustor is shown schematically in Figure 15. Figure 16 is a photograph of the combustor assembly installed in the test facility. Table 12 summarizes product gas combustor design characteristics.

The product gas is injected through an annular injector at low momentum and zero swirl. This configuration, based on International Flame Research Foundation studies (Ref. 7), provides stability and minimum pressure drop across the injector. The combustion air is introduced through a co-axial injector with high velocity and high swirl. The combined flows of gas and combustion air have a swirl number of approximately 0.5, as calculated by the Mathur and Maccullum method for annular swirlers (Ref. 7). This is a swirl number recommended by the IFRF for stable, high-intensity gas flames.

The combustion chamber itself is a well-insulated volume, sized for a reactant residence time of 20 milliseconds, sufficient for complete burnout of the gases (Ref. 8). The chamber insulation is of laminated construction. Provisions have been made in the chamber design to allow access for flame temperature and heat transfer measurements. From the combustion chamber, the burned gases are ducted through a stack and discharged to atmosphere.

Figure 15 HITS Product Gas Combustor

Figure 16 Combustor

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TABLE 12

Product Gas Combustor Design Characteristics

Maximum flame temperature	3400°F
Nominal heat release	600,000 Btu/hour
Fuel heating value	130 Btu/ft ³
Fuel temperature	1300-1600°F
Fuel flowrate	291 lb/hr
Air temperature	1100°F
Air flowrate	355 lb/hr
Igniter	Cowan Model 78-500
Igniter fuel	propane
Igniter heat release	40-60,000 Btu/hr
Igniter safety	conductivity probe
Instrumentation	8 ea 1" dia. ports
Exhaust	18" dia. x 34 ft stack

2.2.3.3 (continued)

As part of the product combustor design analysis, adiabatic equilibrium flame temperatures were calculated, using Aerojet's TRANS-72 computer program. The equilibrium product gas composition was used in this calculation, along with the estimated product gas temperature at the burner of 1300°F and design combustion air temperature of 1100°F. Excess air levels were varied at this condition. As shown in Figure 17, a theoretical flame temperature of 3690°F could be achieved with a stoichiometric mixture. The calculated values are in agreement with figures published by the U.S. Bureau of Mines (Ref. 9). Since flame temperatures with low excess air and 1100°F combustion air preheat could exceed the melting temperature of available castable refractories, the flame temperature was limited by lowering air preheat to about 900°F.

The exhaust gas from the product gas combustor is directed into a 34 ft. high stack with an induced draft of fresh air for cooling and dilution. A section of the stack is refractory-lined for thermal protection. A portion of the blower air bypasses the combustor and is used to induce additional cooling air around the burner exhaust. Cleanout capability is provided in the event excess fines pass the cyclone and deposit in this portion of the system. The stack is shown schematically in Figure 18; Figure 19 is a photograph of the stack installed in the facility.

THEORETICAL
EQUILIBRIUM
FLAME TEMPERATURE,
F.

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Figure 17 | Burner Design Operating Conditions

Figure 18 HITS Exhaust Stack Schematic

Figure 19 Exhaust Stack

2.2.3.4 Char/Ash Removal System

Design criteria for the char/ash removal system are presented in Table 13.

TABLE 13

Char/Ash Removal System Design Criteria

Size:

Char/Ash inlet line	3" pipe
Transfer screw conveyor (water jacketed)	4" barrel
Char/Ash receivers	4 ft ³

Operating Conditions:

	<u>Nominal</u>	<u>Desired Range</u>
System pressure, psia	19	14-24
Inlet temperature, °F	1600	To 1800
Char/Ash flowrate, lb/hr	40	15-125 (manually variable)
Char/Ash discharge temp. °F	200	Ambient-400

Other Design Criteria:

Isolation	Provide isolation between conveyor and char/ash receiver
Receiver change-out	System to be capable of online change-out of char/ash receivers
Char/Ash cooling	System must provide sufficient char/ash cooling to safely permit short-term exposure of the char/ash to ambient air during the receiver change-out and capping operation.
Materials of construction	Compat. with reducing, oxidizing, or sulfur-bearing atmospheres and solids

Interfaces:

With reactor solid discharge nozzle	3" flanged pipe
With char/ash receiver	30-gal drum closure

2.2.3.4 (continued)

The char/ash removal system consists of: (1) a manifold/inlet piping assembly which connects the discharge ports of the bed section of the fluidized-bed reactor to the inlet of the char/ash conveyor, (2) a water-cooled, inclined screw conveyor for the char/ash, and (3) a discharge duct, which connects the discharge port of the conveyor to a closed char/ash receiver. This system is shown schematically in Figure 20; Figure 21 is a photograph of the char/ash removal system. The detail design of these assemblies are discussed in the following subsections.

Manifold/Inlet Piping Assembly

This assembly is composed of three 3-inch 316 stainless steel pipe flanges. These parts include: (1) two flanged spools, each approximately 2-1/2 ft long, and (2) one flanged 45° cross with two blinds and an extended vertical leg, approximately 1-1/2 ft long, with an integral water cooling jacket. The 45° flanged cross mates with one of the three 3-inch flanged discharge ports of the bed section of the fluidized-bed reactor (normally the lowest discharge) and with the 3-inch flanged inlet to the water-cooled conveyor. One or both of the uncooled spools fit between the 45° cross and the water-cooled conveyor when the 45° cross is attached to one of the higher fluidized-bed reactor discharge ports. This assembly is designed to withstand pressures from -5 to 10 psig and inlet temperatures up to 1600°F.

Water-Cooled Screw Conveyor

This conveyor is designed to withstand pressures from -5 to +10 psig and temperatures consistent with the discharge temperature of the manifold/inlet piping assembly defined above. It has a nominal char/ash capacity of 40 lb/hr, but is manually variable over the range of 15-125 lb/hr. The conveyor has a nominal 4-inch diameter and is approximately nine (9) feet in length. The water-cooled jacket (nominally 6-inch in diameter) is rated at 100 psia and has inlet and outlet ports consisting of 1-inch pipe couplings. The conveyor inlet and outlet are 3-inch flanged joints angled to the conveyor body to provide vertical downward inflow and outflow of the char/ash. The conveyor body is inclined (nominally 15°) to accommodate the placement of a standard 30-gal steel drum beneath the conveyor discharge. The conveyor includes an electric drive and speed reducer operating on 110-115V, 60 Hz, single phase current. The drive assembly is a manually controlled variable-speed unit.

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Figure 21 Char Removal System

2.2.3.4 (continued)

Conveyor bearings are of the heavy-duty type and provided with dust seals. Pressure-sealing packing glands are externally adjustable and readily replaceable. Bearings and seals are placed outboard of the conveyor inlet and discharge, with sufficient distance to eliminate adverse heating effects. The drive assembly is placed at the upper (discharge) end of the conveyor.

Conveyor Discharge Duct

The discharge duct is a short vertical section (approximately two feet long) of 3-inch carbon steel. It is flanged at the upper end and contains a 3-inch full-port screwed valve near its midpoint. The duct is terminated at its lower end with a quick-disconnect coupling. The lower half of the coupling is attached to a 30-gal drum lid. The upper portion of the duct contains a 1-inch pipe coupling port, inclined 45° downward, for instrumentation. An appropriate bleed port and bleed valve is installed between the valve and the quick-disconnect coupling.

2.2.3.5 Off-Gas System

Based on process requirements, a set of design criteria was established for the off-gas system. Those requirements basically defined a set of cyclone design criteria, since the cyclone is the primary component in that system. Cyclone design criteria are presented in Table 14.

The off-gas system consists of: (1) a duct from the fluidized-bed reactor system to the cyclone, (2) the cyclone, (3) a duct from the cyclone gas discharge to the interface with a gas burner, (4) a fines recycle subsystem, including an eductor and a subsystem for removing and receiving fines from the fines recycle loop, and (5) appropriate insulation materials for the above. This system is shown in Figure 23 without insulation; Figure 24 is a photograph of the cyclone.

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TABLE 14

Cyclone Design Criteria

Size:

Capacity, ACFM	260
Barrel diameter, in.	6

Operating Conditions:

	<u>Nominal</u>	<u>Desired Range</u>
Inlet temperature, °F	1400	800-1600
Inlet pressure, psia	15	10-20
Gas flowrate, ACFM	260	185-370
Pressure drop, in. W.C.	6	3-12
Gas density, lb/ft ³	0.019	0.011-0.037
Gas viscosity, lb/ft-sec	2.7×10^{-5}	$2 \times 10^{-5} - 3 \times 10^{-5}$
Particulate loading, grains/ACF	8	1-20
Particle size distribution	See Figure 22	

Other Design Criteria:

Cyclone efficiency	>95%
Fines handling	Provide a means of pneumatically recycling the fines to the reactor with N ₂ and/or batchwise removing them via an isolatable catch pot (1 cu. ft)
Materials of construction	Compatible with reducing, oxidizing or sulfur-containing atmospheres
Heat losses	Design to minimize heat losses and heat-up time

Interfaces:

Cyclone inlet with reactor gas dischg.	4" flanged pipe
Cyclone gas outlet with burner inlet	3" flanged pipe
Cyclone fines outlet	Compatible with the pneumatic fines recycle system and the fines catch pot

Figure 22 Projected Particle Size Distribution

Figure 23 Off-Gas System Piping

Figure 24 Photograph of Cyclone

2.2.3.5 (continued)

Reactor/Cyclone Duct

The duct connecting the reactor top closure to the cyclone inlet is designed to operate at pressures from -5 to +5 psig and temperatures up to 1600°F. It includes a flanged 4-inch 316 stainless steel cross with two blinds and a straight, flanged 4-inch 316 stainless steel duct approximately 4-1/2 feet long. The flanged cross mates with the 4-inch flanged nozzle on the reactor top-closure of the fluidized bed reactor and with the flanged duct. The straight duct includes a 3-inch flanged sampling port nozzle with blind and a 1/2-inch pipe and coupling type instrumentation port. These ports are diametrically opposite from each other and oriented perpendicular to the duct axis. They are placed downstream of the forward flange by ≥ 10 duct diameters and upstream of the aft flange by ≥ 3 duct diameters. These ports are made sufficiently long to penetrate the external duct insulation.

Cyclone/Burner Duct

The duct connecting the gas discharge of the cyclone with the burner gas inlet is designed to operate at pressures from -5 to +5 psig and temperatures up to 1600°F. It includes a flanged 316 stainless steel cross with two blinds, a straight, flanged 316 stainless steel duct approximately 3 feet long, a flanged 316 stainless steel tee, an expansion joint, a gas metering section, and another expansion joint. The flanged cross mates with the flange on the cyclone gas discharge and with the straight duct. The straight duct includes a flanged sampling port nozzle with blind and a 1/2-inch pipe-nipple and coupling instrumentation port. These ports are diametrically opposite from each other and oriented perpendicular to the duct axis. They are located approximately ≥ 10 duct diameters downstream of the forward flange and approximately 3 duct diameters upstream of the flanged tee.

Cyclone Fines Recycle Subsystem

This assembly is designed to withstand pressures from -5 to +5 psig and temperatures up to 1600°F. It includes a 2-inch 316 stainless steel duct, approximately 12-feet long, containing a block valve, a nitrogen-driven fines eductor, a nominal 1/2-inch L.D. 316 stainless pneumatic transfer line, approximately 4 feet long with an appropriate shutoff valve, and a closed fines receiver. The 2-inch flanged duct branches just upstream of a normally closed block valve, where it mates with the suction side of the eductor.

2.2.3.5 (continued)

The eductor is designed to educt approximately 3 ACFM of recycle gas (and fines) at approx. 0 psig and discharge it at bed injection point pressure, using gaseous nitrogen as the motive gas. The gas discharge side of the eductor mates with the nominal 1/2-inch L.D. pneumatic transport line. The nominal 1/2-inch L.D. pneumatic transport line mates with the eductor and at the 1-inch cyclone underflow port in the fluidized-bed section of the reactor. It includes an extension (nominal 1/2-inch L.D.) which passes through and seals at the 1-inch flange, extending to the inside refractory surface of the fluidized-bed reactor. It includes a full-port shutoff valve close-coupled to the eductor gas discharge.

When the block valve in the 2-inch cyclone fines discharge duct is opened and the motive gas inlet to the eductor is valved-off, the fines drop vertically through the duct where they are caught either in a closed fines receiver (approx. 1 ft³ in capacity) or in a 30 gallon drum. The receivers are attached to the fines discharge duct via a quick-disconnect coupling. A bleed port and bleed valve are installed between the valve and the quick-disconnect coupling.

Off-Gas System Insulation

The entire off-gas system, except for the fines subsystem below the recycle eductor, is insulated with 4-inches of fibrous insulation having a thermal conductivity of 0.5 Btu/hr.ft².°F/in. The insulation is enclosed with a weatherproof covering.

2.2.3.6 Process Equipment and Controls

Control of the process is straightforward. Gas flowrates are maintained constant at predetermined values by conventional process controllers. Reactor temperature is maintained automatically by monitoring the bed temperature and increasing the solids feed rate to decrease bed temperature. Combustor air preheat temperature is maintained automatically. Other process variables are preset at constant values.

The interrelation of the process equipment and controls is shown in the P&ID, Figure 9. The majority of system controls and display instruments are installed in the control room. A schematic process layout with system alarm points and the pressure purge controls is mounted on a control panel. A control console contains electrical and pneumatic controls, data logger and other visual readout. Figure 25 is a photograph of these units with the data logger removed.

Figure 25. Photograph of HITS Control Room

2.2.3.7 Support Structure

A support structure was designed and built to support the major equipment while providing ready access for operation and maintenance. Figures 5 and 6 show the major features of this portion of the facility. The structure consists of a two-deck steel I-beam tower, 30 feet high, with appropriate access stairs and guard rails. It is designed to withstand the following live loads:

A.	Stairs and platforms	75 PSF
B.	Concentrated load	1000 lbs
C.	Wind	13 PSF basic + 2 PSF for shape factor + 50% for gusts
D.	Seismic	Zone 3

The structure has adequate work space around the equipment and storage area for a 24-hour supply of coal and limestone. Since, for safety considerations, the structure is not enclosed, controls and instrumentation are housed in an adjacent 12 x 24 building. This building also provides office space for the test crew and serves as a data reduction and analysis center.

2.2.3.8 Gas and Particulate Sampling and Analysis Equipment

The acquisition of meaningful design data on the HITS process from bench-scale testing is strongly dependent upon both the techniques and equipment employed in obtaining gas and particulate analyses. The location and design of sampling ports has been carefully planned, along with the selection and use of appropriate test equipment.

Sample Ports and Retractable Sample Probes

Sample ports have been located in the straight, constant-cross-section duct connecting the reactor gas discharge to the cyclone inlet, and in the duct connecting the cyclone gas discharge to the burner inlet. These ports are located ≥ 10 diameters downstream and ≥ 3 diameters upstream of a change in duct geometry or direction. The sample ports consist of 3-inch pipe tees with blind flanges. Retractable sample probes pass through the blind flanges via sealing tubing fittings. An instrumentation port is located in the duct diametrically opposite the sample port. That port is utilized for determining the temperature and/or pressure at the sampling point.

2.2.3.8 (continued)

Sampling Equipment

Particulate moisture content and samples are taken from the gas ducts via the sample port and probe, using a Sierra Instruments Constant Flow Air Sampler (Series 110, Model 113C). Particulate sampling is accomplished by adjusting the flowrate of the air sampler so that isokinetic conditions exist at the sample probe inlet. When analyzing only for moisture content, isokinetic conditions are not required.

Samples of noncondensable gases are similarly taken via the sample ports and probes described above, but the flows are normally controlled by the sample control system. This is a part of an existing online Emission Measurement Console. In most tests, grab samples of noncondensable gases for off-line analyses were withdrawn via the sample probe and pumped into large Tedlar sample bags, as shown in Figure 26. The sample was filtered and condensibles removed before measurement.

Analysis Equipment

The equipment used for the analysis of particulate loading and gas composition is described below.

Alundum Thimble Filter

A Research Appliance Co. Alundum Thimble Filter (Cat. #996142) was used for determining particulate loading. It was operated with the sampling ports, probes and a constant flow sampler under isokinetic conditions.

Emission Measurement Console

This console is an assembly of specific gas analyzers, along with auxiliary components for sample control, conditioning, and for calibrating the individual analyzers. The control and conditioning components include filters, a refrigeration bath (approx. 0°C), vacuum pumps, and the miscellaneous equipment necessary to deliver the sample gas to each analyzer at specified conditions. Figure 27 is a photograph of the system. The specific gas analyzers in the unit are as follows:

Figure 26. Gas Analysis Sample Being Taken During Test

Figure 27 Continuous Gas Analysis System

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2.2.3.8 (continued)

<u>Analyzer</u>	<u>Description</u>
CO	Beckman Infrared Analyzer, Models 315A and 315B
CO ₂	Beckman Infrared Analyzer, Model 315A
O ₂	Beckman Oxygen Analyzer, Model 741
NO-NO _x	Thermo Electron Corp. Chemiluminescent NO-NO _x Analyzer, Model 10A with NO _x Converter
SO ₂ -H ₂ S	Thermo Electron Corp. Pulsed Fluorescent SO ₂ Analyzer, Series 40 with Model 340 H ₂ S Converter
Hydrocarbons	Scott Heated Total Hydrocarbon Analyzer, Model 215

Gas Chromatograph

To complement the online Emission Measurement Console described above, the gas composition of bag samples was also determined by using a gas chromatograph. The instrument used is a Hewlett-Packard 5830 Gas Chromatograph with an 18850A G.C. Terminal. This unit provided quantitative measurements of CO, H₂, O₂ and CH₄ as well as N₂.

3.0 BENCH SCALE TESTING

Described in this section are all testing activities conducted during the 1980 bench scale program. These activities included:

- o Checkout and calibration of subsystems
- o Refractory cure
- o All-up system checkout
- o HITS bench scale system performance tests

3.1 Calibration of Subsystems

Components and subsystems were prepared and operated to verify satisfactory performance. Where required (flow orifices and solids feeder scales), the units were calibrated. Details of some calibration procedures are given in Appendix C-2.

Results of the orifice calibrations are shown in Figures 28 through 31. Feeder scale calibrations are shown in Figures 32 and 33.

3.2 System Checkout

Detailed test procedures were prepared for checkout and operation of the system. These procedures are given in Appendix C-1 and C-3. They cover initial checkout and startup of the system as well as normal operation. Procedures to correct abnormal conditions and emergency shutdown are also included.

As part of system checkout, the reactor, combustor and exhaust stack refractories were run through a two-week cure cycle. During this period the system was heated with the windbox propane burner. System temperatures were gradually increased to above 1800°F in the windbox. Appendix C-4 gives the details of this cure procedure.

The final stage of system checkout was operation of the complete system with coal. Table 2 describes these tests. The system was operated according to the procedure outlined in Appendix C-3.C. The reactor was initially charged with coke and heated to its ignition temperature with the propane burner. This point was reached in about twelve hours. Temperature then rose smoothly and approached operating temperature in about two hours. During this period char and limestone input was initiated to obtain control on bed temperature, while char removal was started to control bed height.

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Figure 28 Fluidization Gas Orifice Calibration

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Figure 29. Transport Air Orifice Calibration

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Figure 30 Combustor Air Orifice Calibration

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$$Q \text{ (SCFM AT 70°F)} = -0.576 + 6015 (F_1) + 13491 (F_1)^2$$

RMS DEVIATION 2.70 SCFM OR 1.05%

Figure 31. Producer Gas Flow Nozzle Calibration

Figure 32. Coal Feeder Weighing System Calibration

Figure 33 Limestone Feeder Weighing System Calibration

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3.2 (continued)

Overall process data were logged about once an hour; critical operating temperatures were recorded every ten minutes on the data logger. Solids input weights were recorded two to three times per hour and char and fines weights were taken generally every hour.

After reaching a bed temperature of about 1800°F on coke, low sulfur coal feed was started and the bed temperature was reduced to about 1660°F. After two and a half hours at this condition, the bed temperature was reduced to about 1560°F and held for an hour. Bed temperature was then increased to 1760°F and held for five hours.

At this point the objectives of the checkout series had been met so the system was shut down and secured. The results and conclusions obtained from this and subsequent system operations are described in Section 5.0.

3.3 System Performance Tests

The balance of the testing was conducted in three test groups, as shown in Table 2. Two of these test groups were with low sulfur coal; the third group was run with high sulfur coal. The conduct of these tests is described below. The results of these tests and conclusions derived from them are discussed in Section 5.0

3.3.1 Low Sulfur Performance Tests

After minor modifications and maintenance to the system, which included completion of the off-gas line insulation system, the system was started up on 21 Aug. 1980. The system was operated at four bed temperatures, approx. 1500, 1600, 1700 and 1800 with air. During this testing the exhaust stack was partially blocked for several hours. Overheating of the combustor at three of the instrumentation ports began at this time and slowly increased throughout the run.

To determine the effects of reduced oxidizing species concentration on char gasification rate, nitrogen was added to the fluidizing air to reduce the oxygen concentration to about 19 wt percent. The four nominal bed temperatures were repeated under these conditions.

Next, the oxygen concentration was increased to about 28 weight percent to determine the effect on char gasification rate. The first point, run at a bed temperature of 1780°F, resulted in excessive combustor temperatures.

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3.3.1 (continued)

Since this aggravated the overheating problem noted earlier, the system was returned to operation with air after one hour at 28% O₂.

The final data points in this test series were to be at low and high Ca/S ratio. In running these tests at 1700°F it was noted that the bed temperature distribution was changing. This change, which started with the enriched oxygen test condition, resulted in the decision to terminate the test run since the bed was apparently defluidizing.

After inspection and clean-out of the bed, repair of the product gas combustor and other minor system maintenance, the system was started up for a repeat test series with the low sulfur coal. The intent was to repeat the air-only tests at 1600, 1700 and 1800°F. However, because of a change in the fines eductor, increased eductor nitrogen flow decreased oxygen concentration to about 19.6 wt. percent for most of the tests, essentially a repeat of the earlier low oxygen tests. The tests were conducted in the order 1700, 1600, and 1800°F. After operation for about nine hours at 1800°F, the solids transfer line to the bed was plugged by improper valve sequencing during a coal hopper refill cycle. This caused a transient bed temperature excursion to at least 1880°F. Subsequent to this excursion bed temperature profile changes indicated gradual defluidization of the bed, so testing was terminated to permit cool down, examination and clean-out of the bed.

3.3.2 High Sulfur Performance Tests

A one-week delay was experienced in startup of these tests due to late shipment of the high sulfur coal. This was caused by difficulties at the grinding and classifying facility due to problems with their drying equipment. On 1 Oct. the system was started up for tests with coal containing 4.06% sulfur. Table 15 is a summary of the test log indicating the activity during a typical test cycle. During this test series tests were run at 1700, 1600 and 1800°F bed temperatures. After changing bed temperature from the 1600 to 1800°F conditions, operation appeared normal for about four hours, at which point blockage of the char output was noted. The system was shut down to clear the blockage; during restart a feed line blockage resulted which eventually lead to a bed temperature overshoot of about 1860°F. Since the evidence was that 1800°F was already too high a bed temperature for this coal, and since bed defluidization had been initiated, the testing was terminated.

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TABLE 15

Typical Test Sequence Log
Test Series #4 - High Sulfur Coal

1 Oct, 1980

07:00 Start test preparation
09:50 Purges balanced. Start pilot propane burner for bed heatup
10:51 Start char feed into bed
10:55 Combustor air preheater on to 600°F
11:15 Main propane burner on
12:18 Total of 82 pounds of char delivered to reach 5 ft bed depth
17:00 Data point #108
18:36 Data point #109
18:38 Windbox at 1267°F, bed at 650
20:00 Bed self heating begun (at 775°F); data point #110
20:28 Main burner off
20:30 Pilot off; bed temperature is 1280°F
20:37 Start limestone feed; bed temperature is 1474°F
20:46 Start char removal, bed depth approaching 79"
20:50 Increase char removal rate
20:56 Start char
21:00 Increase char feed rate; data point #111. Bed temperature = 1663°F
21:21 Bed low; stop char removal
22:00 Data point #112. Bed temperature = 1698°F
22:23 Switch to high sulfur (Eastern Kentucky) coal
23:00 Data point #113. Bed temperature = 1698°F
23:22 Start char removal

2 Oct, 1980

00:15 Data point #114. Bed temperature = 1699°F
01:42 Data point #115. Bed temperature = 1701°F
03:45 Data point #116. Bed temperature = 1705°F
05:40 Data point #117. Bed temperature = 1703°F
Change to new operating conditions
08:25 Data point #118. Bed temperature = 1624°F

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TABLE 15 (cont.)

09:09	Data point #119. Bed temperature = 1625°F
10:56	Data point #120. Bed temperature = 1607°F
12:01	Data point #121. Bed temperature = 1604°F
12:55	Data point #122. Bed temperature = 1609°F
14:03	Data point #123. Bed temperature = 1599°F
15:08	Data point #124. Bed temperature = 1604°F
15:20	Change to new operating condition
16:05	Data point #125. Bed temperature = 1800°F
16:28	Adjust limestone flow
17:00	Data point #126. Bed temperature = 1802°F
17:58	Data point #127. Bed temperature = 1806°F
18:08	Noted eductor block valve left closed after last flush cycle Opened and cleared fines system
18:08	Fines recycle operating properly
18:39	Data point #128. Bed temperature = 1804°F
18:45	Sample lines being cleaned; plugged by fines recycle block
19:12	Stop fluidizing air to unblock char discharge
19:46	Blocked transport line during shutdown. Unblocked
20:06	Data point #129. Bed temperature = 1801°F
20:45	Coal feeder overloaded and shutoff. Bed temperature overshoot to above 1860°F. Restarted feeder
21:00	Data point #130. Bed temperature = 1802°F
21:18	Char removal line obstructed
21:57	Data point #131. Bed temperature = 1798°F
22:20	Switch to nitrogen and shutdown system. End of test.

4.0 DATA ANALYSIS PROCEDURES

This section describes the general manner by which the raw experimental data acquired from testing have been reduced to meaningful engineering data and how those data have been translated into useful process design information.

4.1 General Approach

Essentially all of the major process design information derived from the raw experimental data have been obtained from mass balances around the system when operating stably at various given sets of operating conditions. The mass balances have been established from reduced data on the flow rates and compositions of the various input and output streams to and from the process. The flow rates and compositions of the input and output process streams have, in turn, been determined from raw experimental data in the manner described in Sections 4.2 and 4.3. The derivation of process data from mass balance information is discussed in Section 4.4.

4.2 Process Inputs

The process inputs are comprised of separate solids feeds of coal and limestone and an overall gas feed comprised of four streams: (1) fluidization gas (air, N₂-enriched air, or O₂-enriched air), (2) air for pneumatically transporting the solid feeds into the fluidized bed, (3) N₂ for driving a fines-recycle-educator, and (4) N₂ for purging pressure taps entering the fluidized-bed reactor. The transport air comprised about 9% of the combined fluidization and transport gas streams. The N₂ flow to the fines-recycle-educator varied from one test series to the next but generally was 0, approx. 13, and approx. 1% of the total gas input in test series #2, #3, and #4, respectively. The N₂ for purging the pressure taps amounted to about 2% of the total input gas.

4.2.1 Coal and Limestone Feeds

Average feed rates of coal and limestone to the process were determined for the various data periods from changes in the weights of the coal and limestone feed hoppers which were recorded at regular intervals. Checks of the weighing systems showed them to be sufficiently accurate and repeatable to avoid the need for applying corrections or biases to the indicated weights (see Figures 32 and 33).

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4.2.1 (continued)

The compositions of the low- and high-sulfur coals and the limestone were established from chemical analyses performed on each feed by independent analytical laboratories (see Tables 5, 6, and 7). The samples analyzed represented composites prepared by careful sampling of each feed supply. Multiple sulfur and ash analyses on different samples of the low-sulfur coal established that that feed supply was very uniform and well represented by the composite sample.

4.2.2 Gas Feeds

The fluidization gas entering the process was air, air plus nitrogen, or air plus oxygen. The flow rate of the air portion was determined via a flow orifice using measured upstream temperatures and pressures, the orifice pressure drops, and a calibration curve (Figure 28) established by flowing the orifice against a calibrated orifice. The flow rates of the added nitrogen or oxygen were determined using standard commercial rotameters and by applying gas density corrections which were based on measured temperatures and pressures of the metered streams. The composition of the air was assumed to be that of normal atmospheric dry air since its source was uncontaminated ambient air which was filtered, compressed, and dried. The oxygen was assumed to be pure since its source was commercial liquid oxygen. The nitrogen was assumed to also be pure since it was obtained by vaporization of commercial liquid nitrogen.

The flowrate of the air used to pneumatically transport the coal and limestone feeds into the fluidized-bed reactor was determined via a flow orifice using measured upstream temperatures and pressures, the orifice pressure drops, and a calibration curve (Figure 29) established by flowing it against a calibrated orifice. The air composition was the same as for the fluidization air since they shared a common source.

The flow rates of nitrogen used to drive the fines-recycle-educator and to purge the reactor pressure taps were determined using commercial rotameters and by applying gas density corrections which were based on measured temperatures and pressures of the metered streams. The nitrogen was assumed to be pure since it was obtained by vaporization of commercial liquid nitrogen.

4.3 Process Outputs

The process outputs are comprised of two streams of solids and a stream of producer gas. A stream of solid char/ash is removed from the bed region of the fluidized-bed reactor via a water-cooled screw conveyor. A stream of fines is recovered from the underflow of the producer gas cyclone separator which does not get recycled back to the reactor. The stream of producer gas exits the gasification portion of the process from the downstream side of the cyclone separator. This gas stream is ultimately burned with preheated air in the product gas combustor and is exhausted to the atmosphere via a stack but for mass balance purposes the producer gas stream is considered to be the product.

4.3.1 Product Gas

A flow venturi for measuring the flow rate of the product gas was installed in the system downstream of the cyclone and calibrated (see Figure 31). However, in the course of early tests it was found to foul too rapidly to permit its use as a reliable flow metering device. It has thus been necessary to utilize an alternate means for establishing the flow rate of the product gas. That alternate approach is based on a nitrogen balance around the system. The input rate of nitrogen is accurately known based on coal and input gas flow rates and nitrogen contents. Assuming the char/ash and fines contain essentially zero nitrogen, which is a reasonable assumption, all the input nitrogen must appear in the product gas. Since the nitrogen content of the product gas is a measured value, the flow rate of product gas is thus derived by a forced nitrogen balance.

The compositions of the product gases are experimentally determined on samples of the gas stream. The samples are analyzed with online component-specific instruments for CO, CO₂, hydrocarbons, H₂S, O₂, and NO_x and off-line by gas chromatography for H₂, O₂, N₂, CH₄, and CO. The results from both types of analyses are merged to yield best overall dry gas compositions. Particulate concentrations in the dry gas are also experimentally determined on isokinetically sampled portions of the product gas stream. The moisture contents of the product gases were experimentally determined but often led to results which had to be questioned because of hydrogen imbalances around the system. Ultimately it was decided to define the moisture content of the product gas on the basis of a forced hydrogen balance rather than to accept the questionable experimental results. In making hydrogen balances, it was assumed that the char/ash and fines streams contained essentially zero residual hydrogen.

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4.3.2 Char/Ash and Fines Products

The average rates of fines production were directly determined for the various data periods from the weights of fines collected and recorded at regular intervals. The weights of char/ash removed from the reactor were similarly determined by a weighing process which was repeated at regular intervals. However, in the case of the char/ash product, the average production rates could not be defined reliably in a similar direct manner since significant char/ash inventory changes within the fluidized-bed reactor occurred from time to time as the speed of the char/ash discharge conveyor was changed in attempts to maintain a relatively constant bed level. The char/ash production rate was thus defined by a forced overall mass balance around the systems since the input or output of all other process streams were known with good reliability.

The compositions of the char/ash and fines products were determined only in regards to their content of residual mineral matter. This was accomplished by ashing the samples in a muffle furnace at approximately 1650°F (900°C) for about eight hours.

4.4 Derived Process Data

Having established the process input and output flow rates and compositions for the various data periods as described in Sections 4.2 and 4.3 and knowing temperatures and pressures throughout the system, a variety of process data were derived. These derived data are discussed below and are presented in Section 5.0

4.4.1 Gasifier Operating Conditions

Gasifier operating conditions have been characterized in terms of the three parameters: bed temperature, superficial velocity, and bed depth. Bed temperatures have been taken to be the mean value indicated by five thermocouples immersed in the bed at depths from 9 to 50 inches above the tuyere and radially inward approximately 4 inches from the wall. Superficial velocities have been defined based on the product gas flow rate corrected to average bed temperature and freeboard pressure and upon the bed cross-sectional area. Bed depths have generally been difficult to establish with good reliability since they were intended to be defined from a group of bed pressure drop measurements via nitrogen-purged pressure taps in the bed region. The pressure taps have tended to plug and thus often gave erroneous pressure drops. Ultimately, bed depths have been estimated from combined analyses of reactor temperature profiles and various reactor pressure drop measurements.

4.4.2 Coal-Carbon Gasification Rates and Fractions Gasified

Two different coal-carbon gasification rates and fractions gasified have been derived from the data. They differ from one another by being based on either the total coal-carbon or the fixed carbon fraction of the coal. Gasification of fixed carbon has been assumed to occur only to the extent that the total carbon gasified exceeds the volatile carbon content of the input coal feed.

The total coal-carbon gasification rate is thus the rate at which carbon appears in volatile form in the product gas stream as CO , CO_2 , and CH_4 minus that which is attributable to the CO_2 resulting from complete limestone calcination. The fraction of total coal-carbon gasified is the ratio of the total coal-carbon gasification rate to the total coal-carbon input rate.

The fixed coal-carbon gasification rate is the total coal-carbon gasification rate minus the volatile coal-carbon input rate wherein the volatile coal carbon input rate is defined by the difference between the total carbon and fixed carbon fractions in the input coal. The fixed coal-carbon gasification rate can be a negative value which could be interpreted to indicate that the coal devolatilization is incomplete. The fraction of fixed coal carbon gasified is the ratio of the fixed coal-carbon gasification rate to the fixed coal-carbon input rate.

4.4.3 Sulfur Capture Efficiencies

The sulfur capture efficiency is intended to indicate the degree to which the sulfur in the incoming coal is retained in the solid char/ash and fines products rather than being discharged with the product gas. It is expressed as a percentage of sulfur in the coal. Analytically it is defined by the difference between 100 and the percentage of the input sulfur in the coal that is discharged in the product gas.

4.4.4 Product Gas Heating Values

The heating values of the product gas are determined from its compositions and from the known heating values of its combustible components (CO , H_2 , and CH_4). The small H_2S fraction is neglected. The sensible enthalpy of the gas has also been neglected in calculating tabulated heating values but would add about 25-30 Btu/SCF (the SCF being measured at 70°F) to the values. The heating values are gross or upper values and are calculated on both total gas and dry gas bases.

4.4.5 Energy Generation Rates

The rate at which the process generates energy in the form of the combustible product gas is determined from the gas production rate and the gross heating value of the gas (on a total gas basis). The value is normalized by dividing by the reactor cross-section area thus providing a rapid means of ascertaining the approximate size of reactor needed to supply any given heat load. The tabulated energy generation rate, again, does not include the sensible enthalpy of the gas.

4.4.6 Fraction of Coal Energy in the Product Gas

The fraction of the coal energy that is produced in the form of product gas by the process is defined as the ratio of the product of the gas production rate and the gross heating value of the gas to the product of the coal input rate, and the gross heating value of the coal. This fraction does not reflect the sensible enthalpy that the product gas carries.

5.0 RESULTS AND CONCLUSIONS

The test results obtained during the 1980 bench scale test program are presented here and discussed along with conclusions derived from the test results.

5.1 Basic Data Presentation

The basic test data have been reduced to engineering/process terms and tabulated for each of the three performance test series. Limited data are also presented from the first test series, system checkout. The procedures used to obtain these data have been discussed in Section 4.

5.1.1 Checkout Tests

The results of the checkout test series are summarized in Table 16. These results are considered to be only indicative of trends since the major objective of these initial tests was system shakedown, not data acquisition. For example, the system was generally not held on a data point long enough to obtain true equilibrium data.

Most conclusions drawn from these initial tests relate specifically to the functioning of the bench scale equipment itself although by extension they relate to what can be expected for the overall process. Some of the factors evident at this point were:

- o Process temperature is easily controlled. Although operated manually for this test series, no difficulty was encountered in setting or holding bed temperature by controlling coal flow rate (see Figure 34).
- o Char product is free-flowing and easily transportable. Figure 35 shows the char stream during sampling. The char cooler provides product with a temperature close to ambient. The char auger is a good gas seal in normal system operation.
- o The product gas burns stably. Combustion temperatures were not measured during this first test series.
- o Bed depth control from bed pressure measurement is unreliable, making it difficult to hold a constant bed height. Higher than design bed levels can cause overloading of the fines cyclone and passage of bed material through the system.
- o Stack effluent appearance in normal operation was excellent, giving a completely transparent exhaust. Figure 36 shows the system during operation.

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TABLE 16
Results of Test Series #1, System Checkout Tests

	Test Segment Number				
	<u>0</u>	<u>2</u>	<u>1</u>	<u>3</u>	<u>4</u>
Bed Temperature, °F	1800	1558	1660	1761	1760
Superficial Velocity, ft/sec	4.38	5.15	5.30	5.32	4.71
Bed Depth, ft	5.8	8.6	8.4	6.7	5.3
Fuel Type	Coke	Utah Low Sulfur Coal			
Fuel-Feed Rate, lb/hr	105	110.5	92.7	48.6	54.6
Fuel-Carbon Feed Rate, lb/hr	82.8	75.9	63.7	33.4	37.5
Limestone Feed Rate, lb/hr	4.0	9.7	3.2	4.3	4.8
Total Fuel-Carbon Gasification Rate					
Lb Carbon Gasified/hr	—	25.9	27.4	27.9	23.9
Lb Carbon Gasified/hr-lb Bed	—	.147	.162	.247	.310
Fixed Fuel-Carbon Gasification Rate					
Lb Carbon Gasified/hr	20.9	0	4.8	16.0	10.6
Lb Carbon Gasified/hr-lb Bed	.132	0	.028	.142	.137
Fraction of Total Fuel-Carbon Gasified	—	.34	.43	.83	.64
Fraction of Fixed Fuel-Carbon Gasified	.25	0	.12	.74	.44
Product Gas Composition (dry basis)					
CO, % Vol	10.2	10	12	14.95	15.4
CO ₂ , % Vol	11.9	12.5	11.35	8.97	9.0

34. Coal Flowrate Versus Bed Temperature

35. Char Stream Being Sampled

Figure 36 HITS System During Operation

5.1.1 (continued)

- o CO/CO₂ ratios were lower than originally predicted, but in line with recent British work (Ref. 10).

The overall conclusion reached from the 45-hour checkout tests was that the bench scale system and process were functioning very well.

5.1.2 Low Sulfur Coal Tests

The second test series was conducted to obtain performance data for the low sulfur coal. The basic system performance data obtained in these tests are presented in Table 17, which shows process inputs, outputs, gasification rate, sulfur capture, and heat balance data. Product gas composition data for this test series are presented in Table 18.

It should be noted that data points both at and approaching equilibrium are shown in these tables, so care should be taken in using the data directly. Generally, the data plotted in subsequent figures are taken as the average from the near end of each steady test period.

A repeat test group was conducted, Test Series 3, to obtain further data for operation with the low sulfur coal. The performance data from these tests are shown in Table 19 and gas composition data are given in Table 20.

5.1.3 High Sulfur Coal Tests

The fourth and final test series was run to obtain performance data for the high sulfur, low ash melting point coal. Performance data from Series 4 are given in Table 21; gas compositions are listed in Table 22.

5.2 Test Results Relative to Major Program Objectives

The basic data shown above were obtained to investigate major features of the HITS process to permit determination of its feasibility and scale up to larger systems. The major areas of concern were:

- o Sulfur capture efficiency
- o Char reaction rates
- o Approach to chemical equilibria
- o Process gas combustion characteristics

TEST SERIES #2
HITS PERFORMANCE DATA

Data Point	GASIFIER FEED						GASIFIER OPERATING CONDITIONS			GASIFIER PRODUCTS				Coal-Carbon Gasification Rate, lb/hr		Fraction of Coal-Carbon Gasified		Sulfur Capture Eff., %	Energy Gen Rate Btu/hr ft ²	Fraction of Coal Energy in Prod. Gas	Gross Product Gas Heating Vol., Btu/SCF	
	Coal TYPE	lb/hr	Total Input Gas lb/hr	lb/hr	Climestone Co/S Ratio	lb/hr	Bed Temp., °F	Sup. Vel., ft/sec	Bed Depth ft	Producer Gas SCFM	lb/hr	Char/Ash, lb/hr	Fines, lb/hr	Total Carbon	Fixed Carbon	% Total Carbon	% Fixed Carbon				Total Basis	Dry Basis
57	Low-S	81.7	19.3	192.3	5.4	12.0	1498	4.19	7.7	59.47	239.2	35.5	11.3	23.69	4.74	.409	.122	68.0	400,000	.38	-99	-105
59	"	84.8	19.3	199.4	5.3	12.2	1494	4.28	7.7	60.98	244.5	42.3	9.3	22.90	3.30	.383	.082	-	390,000	.36	-94	-100
61	Low-S	60.9	19.0	215.3	7.2	11.9	1594	4.56	7.0	61.78	255.6	24.5	8.0	23.54	9.41	.546	.324	-	350,000	.45	-83	-85
63	"	89.8	19.2	213.9	0	0	1594	4.78	7.7	64.95	260.1	36.3	7.3	23.33	2.50	.367	.059	-	360,000	.31	-81	-90
65	"	77.8	19.3	217.6	4.9	10.4	1591	4.76	7.5	64.66	259.5	40.1	6.2									
66	"	60.4	19.3	218.7	6.4	10.4	1617	4.59	7.0	61.54	250.2	33.3	6.0									
67	Low-S	47.9	19.3	218.0	8.0	10.4	1696	4.61	5.0	59.68	246.0	26.3	4.0									
69	"	57.9	19.0	220.1	6.6	10.4	1696	4.81	5.8	62.12	254.2	15.5	18.7									
73	Low-S	47.0	18.8	221.0	2.5	3.1	1793	5.00	5.3	61.83	254.4	9.8	6.7	25.44	14.54	.764	.649	31.6	370,000	.61	-86	-68
75	"	47.5	18.9	219.4	2.5	3.2	1798	4.96	5.3	61.19	251.4	11.1	7.4	24.78	13.77	.737	.608	9.6	360,000	.59	-86	-68
81	Low-S	134.4	21.3	174.8	3.4	12.3	1497	4.67	8.7	66.44	254.7	46.8	20.0	26.17	-5.00	.275	-.078	75.4	440,000	.26	-97	-118
94	"	127.7	22.6	162.8	3.4	11.9	1491	4.36	10.1	62.29	238.3	48.8	15.3	25.11	-4.51	.278	-.074	91.7	420,000	.26	-98	-121
77	Low-S	120.6	23.9	166.7	1.1	3.7	1789	5.20	8.9	5.14	250.3	31.5	9.2	36.56	8.57	.428	.149	74.0	570,000	.37	-127	-142

Table 17 System Performance Data, Test Series 2

Table 18 Product Gas Composition, Test Series #2

Product Gas Composition	DATA POINT								
	57	59	61	63	73	75	51	54	77
% Vol. (total basis)									
CO	6.96	6.35	8.08	6.67	13.68	13.39	6.58	7.17	11.61
H ₂	(11.01)	(10.92)	(9.88)	(9.14)	(8.59)	(8.59)	(11.26)	(11.16)	(16.11)
CO ₂	11.47	11.11	10.91	9.46	7.18	6.99	11.59	11.57	14.85
H ₂ O	(5.92)	(6.64)	(2.62)	-	(2.29)	(2.26)	(17.80)	(18.53)	(9.99)
H ₂ S	.0736	.1951	-	-	.0870	.1173	.0830	.0285	.0806
CH ₄	4.19	3.92	2.65	3.15	1.53	1.68	4.11	4.07	4.01
NO _x	.0049	.0056	.0041	.0065	.0007	.0009	.0066	.0057	.0051
O ₂	.63	.47	.97	.65	.49	.68	.97	1.02	.95
N ₂	59.73	60.39	64.46	-	66.15	66.28	47.61	46.44	42.40
Particulates, gr/SCF	-	-	-	-	-	-	-	-	-
Gross heating value, Btu/SCF	~99	~94	~83	~81	~86	~86	~97	~98	~127
% Vol. (dry basis)									
CO	7.40	6.80	8.30	7.40	14.00	13.70	8.00	8.80	12.90
H ₂	(11.70)	(11.70)	(10.15)	(10.15)	(8.79)	(8.79)	(13.70)	(13.70)	(17.9)
CO ₂	12.20	11.90	11.20	10.5	7.35	7.15	14.10	14.20	16.50
H ₂ S	.0782	.2090	-	-	.0890	.1200	.1010	.0350	.0895
CH ₄	4.45	4.20	2.72	3.53	1.57	1.72	5.00	5.00	4.45
NO _x	.0052	.0060	.0042	.0072	.0007	.0009	.0080	.0070	.0057
O ₂	.67	.50	1.00	.72	.50	.70	1.18	1.25	1.05
N ₂	63.50	64.68	66.19	~64.6	67.70	67.82	57.92	57.01	47.10
Particulates, gr/SCF	-	-	-	-	-	-	-	-	-
Gross heating value, Btu/SCF	~105	~100	~85	~90	~88	~88	~118	~121	~142

NOTE: Values in parentheses are estimated. H₂ estimates are based on data from Test Series #3. H₂O values estimated based on a hydrogen balance around the system assuming the char/ash and fines are essentially hydrogen-free.

TEST SERIES #3 SEPT 17 THROUGH 19, 1980
LOW SULFUR COAL

Data Point	GASIFIER FEED					GASIFIER OPERATING CONDITIONS			GASIFIER PRODUCTS			Total Coal-Carbon Gasification Rate lb/hr	Fixed Coal-Carbon Gasification Rate lb/hr	Fraction of Coal-Carbon Gasified		Sulfur Capture Efficiency	Energy Gen. Rate Btu/hr-ft ²	Fraction of Coal Energy in Prod. Gas	Gross Prod. Gas Heating Value, Btu/SCF			
	Coal Type	lb/hr	Total Input Gas Wet Oz	Limestone Ca/S Ratio	lb/hr	Bed Temp., °F	Sup. Vel. ft/sec	Bed Depth ft	Producer Gas SCFH	Char/Ksh lb/hr	Fines, lb/hr			Of Total %	Of Fixed %				Total Basis	Dry Basis		
104	Low-S	113.1	19.70	207.81	1.99	~6.1	1608	4.99	6.2	66.89	260.28	56.8	9.9	23.29	-2.95	0.2907	-0.0547	69.4	432,000	0.257	93.5	108.2
105	"	115.6	19.45	210.47	1.95	6.12	1595	5.04	6.0	67.59	264.42	57.8	~10	21.48	-5.33	0.2624	-0.0968	72.4	377,000	0.220	80.8	96.4
106	"	115.9	19.79	206.48	2.30	7.23	1598	4.97	6.0	66.82	260.98	58.6	~10	21.98	-6.90	0.2678	-0.0888	71.5	392,000	0.228	85.0	101.1
87	Low-S	47.2	22.52	178.28	4.24	5.43	1628	3.81	4.7	50.58	215.54	9.6	5.6	21.89	10.94	0.6548	0.4066	34.9	242,000	0.345	69.2	73.6
88-89	"	50.3	22.52	178.11	1.05	2.53	1678	4.02	6.0	52.00	218.22	6.6	6.1	23.85	12.19	0.6696	0.5087	38.5	287,000	0.384	79.9	84.4
90	"	75.0	22.51	177.30	1.20	2.44	1714	4.41	6.0	56.17	229.50	18.0	7.2	25.61	8.22	0.4822	0.2300	59.9	341,000	0.306	87.9	97.0
97-92	"	82.9	22.49	176.03	1.08	2.44	1687	4.27	6.3	55.16	219.97	32.5	8.9	23.11	3.88	0.3937	0.0984	69.6	365,000	0.297	95.8	106.8
93	"	68.2	19.55	207.46	1.31	2.42	1697	4.75	6.6	59.59	243.23	34.9	0	21.25	5.43	0.4400	0.1672	61.2	314,000	0.310	76.3	82.7
94	"	67.5	19.62	210.42	1.66	3.05	1684	4.76	6.6	60.23	246.46	34.5	0	21.39	5.73	0.4474	0.1783	59.0	315,000	0.314	75.7	81.8
96	Low-S	~46	19.61	212.63	2.40	~3	1780	4.87	6.9	59.25	242.17	18.2	1.3	23.40	12.81	0.7207	0.5846	40.7	331,000	0.406	81.0	84.3
97	"	38.4	19.63	212.32	2.06	2.17	1785	4.70	6.4	57.15	234.79	15.7	2.4	22.07	13.16	0.8114	0.7195	51.2	312,000	0.548	79.1	80.9
99	"	41.1	19.70	215.07	1.94	2.17	1792	4.80	5.6	58.52	242.27	11.1	5.0	23.00	13.46	0.7900	0.6878	-	312,000	0.512	77.3	79.1
101	"	52.9	20.05	203.85	1.99	2.86	1783	4.76	5.2	57.20	233.71	19.3	6.6	22.61	10.34	0.6034	0.4103	-	332,000	0.424	84.2	87.4
102	"	66.3	20.02	204.60	1.59	2.86	1783	4.92	5.3	59.90	243.45	22.4	7.9	22.94	7.56	0.4885	0.2394	-	329,000	0.334	79.5	85.7
103-104	"	~66.1	19.55	210.27	1.88	2.86	1785	4.92	5.4	59.76	242.46	17.5	9.3	22.87	9.88	0.5756	0.3690	-	341,000	0.410	82.8	86.8

Table 19 System Performance Data,
Test Series 3

Table 20

TEST SERIES #3, 9/17-19/1980 LOW SULFUR COAL
PRODUCT GAS COMPOSITIONS

Product Gas Composition % Vol. (Total Basis)	DATA POINT														
	104	105	106	87	88-89	90	91-92	93	94	96	97	99	101	102	103-104
CO	7.48	6.24	6.45	8.72	10.76	9.41	8.63	7.94	7.92	14.36	14.08	13.30	11.66	10.22	12.15
H ₂	9.59	8.36	8.71	7.45	8.69	9.51	9.86	8.20	8.13	8.57	8.62	8.56	9.44	8.95	9.40
CO ₂	7.78	7.88	8.13	13.34	12.30	12.50	10.36	8.90	8.93	6.36	6.14	7.18	8.09	8.72	7.20
H ₂ O	13.57	16.15	15.76	6.04	5.37	9.40	10.28	7.76	7.46	3.96	2.25	2.26	3.63	7.22	4.60
H ₂ S	.0864	.0788	.0826	.1015	.0994	.0894	.0763	.0742	.0768	.0768	.0547	-	-	-	-
CH ₄	3.99	3.50	3.74	1.82	1.85	2.83	3.77	2.55	2.52	.86	.74	.83	1.76	1.90	1.48
NO _x	.0071	.0071	.0072	.0056	.0031	.0051	.0052	.0041	.0033	.0004	.0002	.0004	.0017	.0022	.0013
O ₂	.33	.29	.30	.42	.57	.45	.54	.65	.88	.11	.15	.20	.29	.19	.25
N ₂	57.17	57.49	56.82	62.09	60.36	55.80	56.48	63.92	64.09	65.71	67.97	67.18	64.92	62.33	64.54
Particulates, gr/SCF	-	-	2.46	-	-	-	-	-	1.01	-	-	-	-	.63	-
Gross heating value, Btu/SCF	93.5	80.8	85.0	69.2	79.9	87.9	95.8	76.3	75.7	81.0	79.1	77.3	84.2	79.5	82.8
% Vol. (Dry Basis)															
CO	8.65	7.44	7.66	9.28	11.37	10.39	9.62	8.61	8.56	14.95	14.40	13.61	12.10	11.02	12.74
H ₂	11.09	9.97	10.34	7.93	9.18	10.50	10.99	8.89	8.78	8.92	8.82	8.76	9.80	9.65	9.85
CO ₂	9.00	9.40	9.65	14.20	13.00	13.80	11.55	9.65	9.65	6.62	6.28	7.35	8.40	9.40	7.55
H ₂ S	.1000	.0940	.0980	.1080	.1050	.0987	.0850	.0805	.0830	.0800	.0560	-	-	-	-
CH ₄	4.62	4.17	4.44	1.94	1.96	3.12	4.20	2.76	2.72	.90	.76	.85	1.83	2.05	1.55
NO _x	.0082	.0085	.0085	.0060	.0033	.0056	.0058	.0045	.0036	.0005	.0002	.0004	.0018	.0024	.0014
O ₂	.38	.35	.36	.45	.60	.50	.60	.70	.95	.11	.155	.20	.30	.20	.26
N ₂	66.15	68.57	67.44	66.09	63.78	61.59	62.95	69.31	69.25	68.42	69.53	68.73	67.37	67.18	67.65
Particulates, gr/SCF	-	-	2.92	-	-	-	-	-	1.10	-	-	-	-	.68	-
Gross heating value, Btu/SCF	108.2	96.4	101.1	73.6	84.4	97.0	106.8	82.7	81.8	84.3	80.9	79.1	87.4	85.7	86.8

Table 21 System Performance Data,
Test Series 4TEST SERIES #4
SYSTEM PERFORMANCE TEST RESULTS - HIGH SULFUR COAL

Data Point	GASIFIER FEED					GASIFIER OPERATING CONDITIONS			GASIFIER PRODUCTS			Total Coal-Carbon Gasification Rate lb/hr	Fixed Coal-Carbon Gasification Rate lb/hr	Fraction of Coal-Carbon Gasified		Sulfur Capture Efficiency	Energy Gen. Rate Btu/hr-ft ²	Fraction of Coal Energy in Prod. Gas	Gross Prod. Gas Heating Value, Btu/SCF	
	Coal Type	lb/hr	Total Input Gas Btu/hr	Limestone lb/hr	Ca/S Ratio	Bed Temp., °F	Sup. Vel., ft/sec	Bed Depth, ft	Producer Gas SCFM	Char/Ash lb/hr	Fines, lb/hr			Of Total C	Of Fixed C				Total Basis	Dry Basis
119	H1-S	103.7	22.4	199	2.08	29.6	1611	4.79	6.4	64.8	264	52.7	16	.374	.130	90.5	408,000	.313	91.2	103
120-121	"	103.4	22.4	199	2.13	30.2	1592	4.69	6.0	64.1	261	52.6	18.5	.366	.118	96.4	404,000	.311	91.3	103
122	"	96.7	22.4	198	2.26	30.0	1597	4.68	6.1	63.7	261	43.1	21	.398	.163	91.7	399,000	.328	90.7	102
124	"	96.7	22.4	198	2.26	30.0	1592	4.66	5.9	63.8	261	37.4	26	.404	.172	92.1	397,000	.327	90.1	101
114	H1-S	51.6	22.4	203	2.92	20.7	1693	4.40	6.3	56.5	239	20.2	16	.658	.524	82.0	300,000	.462	76.9	81.0
115	"	68.3	22.4	203	2.13	20.0	1696	4.64	6.4	59.5	245	29.8	17	.503	.309	81.9	333,000	.388	81.1	88.2
116	"	67.3	22.4	204	2.24	20.7	1698	4.67	6.5	59.8	246	26.9	19	.534	.352	65.5(?)	362,000	.428	87.7	94.0
117	"	67.0	22.4	203	2.25	20.7	1698	4.65	6.5	59.5	243	27.4	20.5	.536	.355	86.8	373,000	.443	90.7	97.1
126-126	H1-S	52.9	22.4	197	0.99	7.2	1785	4.59	5.9	57.8	235	~0	~21.5	.764	.672	91.5	388,000	.583	98.3	102
127	"	54.8	22.4	197	2.10	15.8	1789	4.66	5.9	57.8	237	7.2	23	.713	.601	93.0	386,000	.560	96.7	101
128	"	54.8	22.4	199	2.10	15.8	1788	4.69	5.8	58.3	239	7.8	23	.716	.606	85.7	390,000	.566	96.9	101
129-130	"	66.3	22.4	203	1.67	15.2	1785	4.84	5.4	60.3	245	19.8	19	.596	.438	88.6	401,000	.481	95.5	102
131	"	67.1	22.4	203	1.65	15.2	1780	4.85	5.6	60.7	247	19.7	19	.596	.438	91.3	406,000	.481	96.8	103

Table 22

TEST SERIES #4, 10/1-2/1980
PRODUCT GAS COMPOSITIONS - HIGH SULFUR COAL TESTS

Product Gas Composition	Data Point Number (Grouped By Bed Temperature)												
	Bed Temperature ~1598 F				Bed Temperature ~1696 F				Bed Temperature ~1785 F				
	119	120-121	122	124	114	115	116	117	125-126	127	128	129-130	131
% Vol (Total Basis)													
CO	11.30	10.51	11.93	11.78	13.01	13.59	14.62	15.23	18.40	18.28	18.11	16.69	17.03
H ₂	8.41	8.66	8.10	8.21	6.63	7.17	7.69	7.98	8.38	8.63	8.76	8.83	8.80
CO ₂	10.81	11.15	10.62	11.12	10.02	8.37	8.07	7.52	7.17	7.28	7.33	7.57	7.37
H ₂ O	11.43	11.13	10.74	10.71	5.01	8.05	6.74	6.54	3.69	4.27	3.90	5.58	5.84
H ₂ S	.124	.047	.103	.098	.134	.170	-	.122	.064	.054	.110	.102	.078
CH ₄	2.92	3.10	2.74	2.72	1.50	1.58	1.73	1.76	1.39	1.18	1.22	1.61	1.54
NO _x	.0022	.0027	.0017	.0018	.0008	.0012	.0008	.0008	.0006	.0004	.0005	.0006	.0005
O ₂	.443	.471	.571	.402	.323	.644	.672	.458	.241	.239	.240	.236	.235
N ₂	54.56	54.93	55.11	54.96	63.37	60.48	60.19	60.39	60.68	60.05	60.34	59.38	59.11
Particulates, gr/SCF	-	-	2.96	1.90	-	-	-	-	-	-	-	-	0.63
Gross heating value, Btu/SCF	91.24	91.33	90.70	90.12	76.91	81.07	87.67	90.73	98.27	96.66	96.91	96.51	96.84
% Vol (Dry Basis)													
CO	12.76	11.83	13.37	13.19	13.70	14.72	15.68	16.30	19.10	19.10	18.84	17.68	18.08
H ₂	9.49	9.74	9.16	9.19	6.98	7.80	8.25	8.54	8.70	9.02	9.11	9.35	9.34
CO ₂	12.20	12.55	11.90	12.45	10.55	9.10	8.65	8.05	7.44	7.61	7.63	8.02	7.83
H ₂ S	.140	.053	.115	.110	.141	.185	-	.130	.067	.057	.114	.108	.083
CH ₄	3.30	3.49	3.07	3.05	1.58	1.72	1.86	1.88	1.44	1.23	1.27	1.70	1.64
NO _x	.0025	.0030	.0019	.0020	.0008	.0013	.0009	.0008	.0006	.0004	.0005	.0007	.0006
O ₂	.50	.53	.64	.45	.34	.70	.72	.49	.25	.25	.25	.25	.25
N ₂	61.61	61.80	61.74	61.56	66.56	66.71	65.77	64.61	63.00	62.73	62.79	62.89	62.78
Particulates, gr/SCF	-	-	3.32	2.13	-	-	-	-	-	-	-	-	.673
Gross heating value, Btu/SCF	103.02	102.76	101.61	100.94	80.96	88.16	94.00	97.07	102.04	100.98	100.84	102.21	102.84

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5.2.1 Sulfur Capture Efficiency

Sulfur capture efficiency data are presented in Figure 37, where sulfur capture efficiency for the first stage is defined as $100 - \frac{\text{Sulfur out in product gas}}{\text{total sulfur input}} \times 100$.

This does not distinguish between sulfur still in the char and sulfur captured by the limestone.

Several conclusions can be drawn from these data. First, the process can be operated so that sulfur capture efficiencies well above 80% can be obtained, for both low and high sulfur coals. The capture curve is generally flatter with temperature than had been expected, so that loss in sulfur capture efficiency is not necessarily the parameter which will set the limit on upper process temperature.

One result noted in the data is an apparent influence on capture efficiency of nitrogen concentration in the fluidizing gas. This is not expected, especially the large influence for relatively small changes in nitrogen concentration. It may be an artifact of the measuring technique although it persisted in repeat tests. No firm conclusions can be drawn from the limited number of test points available.

The overall conclusion from the initial bench scale tests with respect to sulfur capture is that satisfactory sulfur capture performance is obtained, with some margin to allow for process variation and use of less than optimum limestone.

5.2.2 Char Reaction Rates

The overall energy production from the first stage system depends on coal devolatilization and char reaction. The fraction of total carbon gasified is shown in Figure 38. The fraction of input coal energy appearing in the product gas is shown in Figure 39, which includes the sensible heat of the product gas stream. Generally, over 70% of the input energy can be available in the process stream at bed temperatures approaching 1800°F. The data indicate that the Eastern Kentucky high-sulfur coal was more reactive at a given temperature than the Utah low-sulfur coal. This is more apparent in Figure 40, which shows a comparison of measured to predicted char fraction reacted. These data were obtained by comparing predicted extent of char reacted with measured amount of char reacted. To do this the expression of Appendix B. IV was solved using the measured gas temperature and char/ash fraction along with reaction constants from Ref. 11. Measured char and fines ash fractions are given in Table 23. As can be seen, the two coals bracketed the prediction. The low sulfur coal had an overall extent of char reaction of about 40% of predicted, while the high sulfur coal was at about 80% of predicted at about 1600°F, reaching over 200% of predicted at about 1800°F.

FIRST STAGE SULFUR CAPTURE EFFICIENCY, PERCENT

Figure 37. Sulfur Capture Efficiency

FRACTION OF TOTAL CARBON GASIFIED

SYMBOL	TEST SERIES	COAL	FLUIDIZING GAS
□	2	LOW S	AIR (22 WT % O ₂)
◻	2	LOW S	AIR + N ₂ (19.1 WT % O ₂)
△	3	LOW S	AIR + N ₂ (19.6 WT % O ₂)
○	4	HIGH S	AIR (22.4 WT % O ₂)

Figure 38. Fraction of Total Carbon Gasified

Figure 39. Fraction of Input Energy
Appearing in Product Gas

Figure 40..Comparison of Measured To
Predicted Extent of Char Reacted

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TABLE 23

Char and Fines Analyses for Residual Mineral Matter

<u>Data Point(s)</u>	<u>Test Series</u>	<u>Nominal Bed Temp., °F</u>	<u>Residual Mineral Matter, % wt</u>	
			<u>Char/Ash</u>	<u>Fines</u>
51	2	1500	48.4	19.8
54,57,59	2	1500	38.3	—
61,63	2	1600	52.5	—
105	3	1600	38.6	39.9
93,94	3	1700	39.9	~25
96-101	3	1800	21.6	23.6
124	4	1600	65.2	41.3
116,117	4	1700	70.8	47.4
129-131	4	1800	72.1	—

5.2.3 Chemical Equilibria

The measured gas compositions are shown in Figure 41, 42 and 43 for three sets of operating conditions: low sulfur coal with both normal air and reduced oxidizing species and high sulfur coal with normal fluidizing air. Generally the ratios of CO to CO₂ and H₂ to H₂O were lower than predicted by equilibrium calculations. Tables 24, 25 and 26 show a comparison of experimental to theoretical results which are also plotted in Figure 44. These results appear to be general for this stage of the combustion process and represent a loss in useful gas production. The excess CO₂ and H₂O production is the result of inadequate reaction due to low rates or poor mixing. As will be shown, the end result required, a product gas with high combustion temperatures, is still achieved.

5.2.4 Process Gas Combustion Characteristics

The process gas combustion characteristics were defined in terms of:

- o Composition
- o Heating value
- o Combustor flame temperature
- o Emissivity
- o Combustor heat flux

5.2.4.1 Gas Composition

Gas composition data were presented in the previous section. Particulate concentrations, which impact the emissivity and heat flux, are shown in Figure 45. It should be noted that these particulates are ahead of the combustor; most of the particulates are either burned in the combustor or removed by centrifugal slagging on the hot combustor walls. Particulate concentration dropped substantially with increased operating temperature. Although no useful particulate measurements were obtained at 1500°F, qualitatively it was apparent there was a significant level of particulates. In addition, some tars and liquid products collected in the gas sample stream at 1500°F. These factors, coupled with the difficulty of limiting bed depth at the high coal flow rates required to maintain low bed temperature, resulted in the decision to limit the lower bed temperature to about 1600°F.

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Figure 41. Product Gas Composition
Test Series #2

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NOTES: 1. LOW SULFUR COAL
2. 19.6 WT % OXYGEN

Figure 42. Product Gas Composition
Test Series #3

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NOTES: 1. HIGH SULFUR COAL
2. 22 WT % OXYGEN

Figure 43 Product Gas Composition
Test Series #4

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TABLE 24

Comparison of Product Gas Composition with Theoretical Equilibrium Composition

Test Series #2, 8/22-25/1980 Low Sulfur Coal

	<u>Data Point</u>								
	<u>57</u>	<u>59</u>	<u>61</u>	<u>63</u>	<u>73</u>	<u>75</u>	<u>51</u>	<u>54</u>	<u>77</u>
Avg. Bed Temp.									
°F	1498	1494	1594	1594	1793	1798	1497	1491	1789
°K	1087.6	1085.4	1140.9	1140.9	1251.5	1254.3	1087.1	1083.7	1249.3
$K_{p3}^{(1)}, (CO)^2/(CO_2)$									
Experimental x $10^2 =$	4.22	3.63	5.99	4.7	26.1	25.7	3.73	4.44	9.08
Calculated ⁽²⁾ =	9.020	8.645	21.56	21.56	103.8	107.6	8.943	8.434	100.8
Expt./Calc. x $10^2 =$	.47	.42	.28	.22	.25	.24	.42	.53	.09

(1) $C(s) + CO_2 = 200$

(2) Based on JANAF Thermochemical Data

Table 25

COMPARISON OF PRODUCT GAS COMPOSITION WITH THEORETICAL EQUILIBRIUM COMPOSITION
TEST SERIES #3, 9/17-19/1980 - LOW SULFUR COAL

Ave Bed Temp	Data Point														
	104	105	106	87	88-89	90	91-92	93	94	96	97	99	101	102	103-104
°F	1606	1595	1598	1628	1678	1714	1687	1697	1684	1780	1785	1792	1783	1783	1785
°K	1147.6	1141.5	1143.2	1159.8	1187.6	1207.6	1192.6	1198.2	1190.9	1244.3	1247.1	1250.9	1245.9	1245.9	1247.1
(1) $Kp_1, (CO)(H_2O)/(CO_2)(H_2)$															
Experimental	1.360	1.530	1.436	0.530	0.541	0.744	0.868	0.844	0.814	1.043	0.599	0.489	0.554	0.945	0.826
Calculated ⁽⁴⁾	1.170	1.149	1.155	1.214	1.314	1.389	1.333	1.354	1.327	1.530	1.541	1.556	1.536	1.536	1.541
Expt/Calc	1.16	1.33	1.24	0.44	0.41	0.54	0.65	0.62	0.61	0.68	0.39	0.31	0.36	0.62	0.54
(2) $Kp_2, (CO)(H_2)/(H_2O)$															
Experimental x100 =	5.286	3.230	3.565	10.76	17.41	9.52	8.28	8.39	8.63	31.08	53.94	50.38	30.32	12.67	24.83
Calculated ⁽⁴⁾	20.43	18.94	19.34	23.73	33.00	41.44	34.96	37.27	34.28	61.75	63.60	66.18	62.80	62.80	63.60
Expt/Calc x100 =	0.26	0.17	0.18	0.45	0.53	0.23	0.24	0.23	0.25	0.50	0.85	0.76	0.48	0.20	0.39
(3) $Kp_3, (CO)^2/CO_2$															
Experimental x100 =	7.192	4.941	4.776	5.700	9.413	7.084	7.189	7.084	7.024	32.42	32.29	24.64	16.81	11.98	20.50
Calculated ⁽⁴⁾	23.92	21.76	22.34	28.80	43.38	57.56	46.60	50.45	45.48	94.46	97.98	102.9	96.46	96.46	97.98
Exp/Calc x100 =	0.30	0.23	0.21	0.20	0.22	0.12	0.15	0.14	0.15	0.34	0.33	0.24	0.17	0.12	0.21

Reactions:

(4) Based on JANNAF Thermochemical Data

Table 26

COMPARISON OF MEASURED COMPOSITION DATA TO THEORETICAL EQUILIBRIUM COMPOSITION
TEST SERIES #4, 10/1-2/1980 - HIGH SULFUR COAL

Ave Bed Temp	119	120-121	122	124	114	115	116	117	125-126	127	128	129-130	131
°F	1611	1592	1597	1592	1693	1696	1698	1698	1785	1789	1788	1785	1780
°K	1150	1140	1143	1140	1196	1198	1199	1199	1247	1249	1249	1247	1244
$Kp_1 = (CO)(H_2O)/(CO_2)(H_2)$													
Experimental	1.423	1.212	1.475	1.383	0.981	1.815	1.587	1.658	1.131	1.242	1.099	1.394	1.532
Calculated	1.180	1.143	1.153	1.143	1.345	1.352	1.356	1.356	1.540	1.549	1.547	1.540	1.530
Expt/Calc	1.21	1.06	1.28	1.21	0.73	1.34	1.17	1.22	0.73	0.80	0.71	0.91	1.00
$Kp_2 = (CO)(H_2)/(H_2O)$													
Experimental X100 =	8.309	8.178	9.090	9.020	17.23	12.06	16.70	18.60	41.76	36.94	40.69	26.41	25.66
Calculated	21.15	18.54	19.20	18.54	36.30	37.01	37.48	37.48	63.60	65.08	64.68	63.60	61.75
Expt/Calc X100 =	0.393	0.441	0.473	0.487	0.475	0.326	0.446	0.496	0.657	0.568	0.629	0.415	0.415
$Kp_3 = (CO)^2/CO_2$													
Experimental X100 =	11.82	9.91	13.41	12.48	16.90	21.89	26.51	30.85	47.22	45.89	44.71	36.80	39.31
Calculated	25.02	21.23	22.18	21.23	48.93	50.12	50.91	50.91	98.16	101.0	100.2	99.16	94.64
Exp/Calc X100 =	0.472	0.467	0.605	0.588	0.345	0.437	0.521	0.606	0.481	0.454	0.446	0.375	0.415

Reactions:

- (1) $CO_2 + H_2 \rightleftharpoons CO + H_2O$
- (2) $C(s) + H_2O \rightleftharpoons CO + H_2$
- (3) $C(s) + CO_2 \rightleftharpoons 2CO$

Figure 44 Ratio Between Experimental and Theoretical Equilibrium Constants

SAMPLE POINT: DOWNSTREAM OF CYCLONE,
UPSTREAM OF COMBUSTOR

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Figure 45. Particulate Concentration
in Product Gas

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5.2.4.2 Gas Heating Value

Heating values of the product gas are derived from the measured gas composition. They are presented in Figure 46 for the product gas at standard conditions and in Figure 47 for the gas including sensible heat at bed temperature. This latter value is useful since this represents the way in which the gas is intended to be used, without cooling. The heating values obtained in this way are in the range of 120 to 130 Btu/SCF for the high sulfur coal and about 115 Btu/SCF for the low sulfur coal.

5.2.4.3 Combustor Flame Temperature

Combustor flame temperature was measured in test series two and three with platinum/platinum-rhodium thermocouples protected with an alumina sheath. In the first series the thermocouple failed after reading an indicated temperature of 2835°F; in the second series a replacement thermocouple failed at about 2975°F.

Figure 48 is a plot of the indicated combustion gas temperature versus time up to the failure point. It should be noted that actual temperatures are higher than the measured temperature because of radiation and stem conduction losses. In both cases the alumina sheath broke at the point at which it penetrated the refractory wall and entered the combustion zone. After the thermocouple failed, the combustor refractory wall temperature or the temperature of an alumina rod projecting slightly from the wall was measured using an L&N optical pyrometer. The highest temperatures measured in this way, 2900°F, are expected to be significantly below the actual combustion gas temperature.

Use of a two-color pyrometer system to directly measure gas temperature and emissivity was not accomplished. It was not available in time for the first tests. It could not be used in test series 3 and 4 because the required optical ports were removed to accomplish repair of the combustor.

It is concluded that combustion temperatures in excess of 3000°F were achieved based on the thermocouple and optical pyrometer data.

5.2.4.4 Flame Emissivity

As discussed above, the two color pyrometer was not used in these tests and therefore no method for measuring emissivity was available. The effective emissivity of the 12" gas column was less than 1.0 since it was relatively easy to see the far wall of the combustor through the gas.

Figure 46. Product Gas Heating Value,
Without Sensible Heat

Figure 47 Product Gas Heating Value,
With Sensible Heat

PLATINUM/PLATINUM - 13% RHODIUM THERMOCOUPLE IN ALUMINA WELL
UNCORRECTED FOR RADIATION AND STEM CONDUCTION LOSSES
TEST GROUP #2

Figure 48. Measured Combustion Gas Temperature

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5.2.4.5 Combustor Heat Flux

Satisfactory heat flux measurements were made in test series 2 using the water-cooled calorimeter described in Appendix A. Figure 49 is a plot of cold wall heat flux data measured with this unit. The values for operation with normal air are above the average of 40,000 to 80,000 which is typical for a combustor of this size. Tests were not run with varying air to fuel ratios and the values used are not necessarily optimum. The combustion temperature is strongly dependent on combustor heat loss and therefore wall temperature. Long thermal time constants at the high temperature made it impractical to determine the effect of A/F ratio.

5.3 Discussion of Results Relative to Process Feasibility

Overall energy production rates per square foot of bed area are above those initially predicted. Figure 50 and 51 show specific energy production excluding and including sensible heat of the gas. Energy production rates of about 0.5×10^6 Btu/Hr-Ft² are achieved. The energy production rates do not appear to be a strong function of bed temperature. The maximum difference in production rate for the two coals tested in this program is less than 15%.

All measurements and significant results of the bench scale test program to date indicate the feasibility of the HITS process. Sulfur capture, gas heating value, and gas production rate are all in the range required for a viable system.

SYMBOL	TEST SERIES	COAL	FLUIDIZING GAS
□	2	LOW S	AIR (22 WT % O ₂)
□	2	LOW S	AIR + N ₂ (19.1 WT % O ₂)
□	2	LOW S	AIR + O ₂ (27.9 WT % O ₂)

Figure 49. Combustor Cold Wall Heat Flux

Figure 50. FUEL Gas Energy Production Rate, Without Sensible Heat

Figure 51. Fuel Gas Energy Production Rate, With Sensible Heat

6.0 RECOMMENDATIONS

Much more useful design data can be obtained with the HITS first stage bench scale system. More tests should be run over a longer period of time, to increase the confidence level of the data, using a wide range of coals and limestones of commercial size. In addition, the system should be operated as a second stage combustor to obtain char combustion data. Some minor changes and improvements should be made to the test system to facilitate further testing.

6.1 Additional First Stage Tests

The test matrix begun in the rests reported here should be completed using commercial-size feed material. Effects of process operating variables such as bed depth, superficial velocity and Ca/S ratio should be determined. In light of the reactivity differences seen between the two coals tested here, it is very important that more coals be evaluated. Limestones with a range of performance qualities should be tested as well. The definite limits found on bed temperature should be more closely defined for each coal.

6.2 Second Stage Testing

The first stage should be modified for operation as a second stage combustor and tested over the same range of conditions as the first stage. In this way a complete set of empirical data will be available for extrapolation to a Product Development Unit.

6.3 Test System Improvements

Some obvious modifications to the system should be made to improve its operation. These include:

- o Modification of the char withdrawal piping to set bed depth by overflow.
- o Improvements to the gas sample system to make it more reliable.
- o Improvements in bed height measurement to permit more accurate determination of bed inventory.
- o Improvement of the fines eductor/reinjection system to make its operation more reliable.
- o Improvement in the combustor gas temperature measurement to permit reliable measurements above 3000°F.

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APPENDIX A

MEASUREMENT OF PRODUCT GAS FLAME CHARACTERISTICS

APPENDIX A

One objective of the test program was to determine the flame characteristics of the product gas. Heat flux from the process and the driving, or flame temperature, are the primary items of interest. The heat flux is determined by the flame temperature, the gas properties, including emissivity, the flame geometry and the flow field. During the bench-scale program, combustion gas temperature, and total heat flux were measured.

Two independent measurements were made to determine the heat flux characteristics of the flame. These were:

- o Combustion gas temperature - high temperature thermocouple
- o Total heat flux - heat flux gage

The approaches chosen, which use relatively simple commercially available equipment, are briefly discussed below.

HIGH-TEMPERATURE THERMOCOUPLE

The straightforward method for flame temperature measurement is use of a thermocouple. Conventional thermocouples are limited to temperatures in the range of 3000° to 3200°F. Platinum alloy thermocouples can be operated reliably for extended periods in an oxidizing environment at temperatures around 3000°F. Operating at above 50% excess air will hold the maximum theoretical flame temperature to safe levels for platinum thermocouples. To protect the platinum couples, they were inserted into permanently installed ceramic thermowells which have safe working temperature limits of 3400°F. In practice, failure of the ceramic thermowells as the temperature approached 3000°F resulted in loss of this measurement. It is believed that thermal stress caused by differential growth of the combustor can relative to the inner refractory wall caused the breakage.

TOTAL HEAT FLUX MEASUREMENT

Total heat flux from the combustion zone was measured with a heat flux gage, a water-cooled heat flux calorimeter manufactured by Hy-Cal, Inc. The unit, shown schematically in its installation in Figure A-1, operates by developing a thermal gradient in a cooled plate exposed to heat transfer. The cooled plate is one leg of a thermocouple and produces a voltage output proportional to heat flux. Water temperature and flowrate must be sufficient to prevent overheating but need not be measured. Table A-1 summarizes characteristics of this sensor.

Figure A-1
HEAT FLUX SENSOR INSTALLATION

TABLE A-1

CHARACTERISTICS OF HEAT FLUX SENSOR

Sensor type	Water-cooled heat flux gage
Mfgr.	Hy-Cal
Model	C-1301-A
Heat flux range available (5 steps)	$0-4 \times 10^5$ to $0-10 \times 10^7$ Btu/Ft ² -Hr
Sensor active diameter	0.2 inches
Output, full scale	10 MV
Cooling water flow required	0.25 Gpm
Surface absorptivity	.89
Spectral range	Flat 0.2 to 10 μ

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The coldwall heat flux measured with this device must be corrected for the difference between the surface temperature and emissivity of the calorimeter and kiln product. This involves making estimates of the separate contributions of convective and radiative heat transfer and correcting them for difference in temperature and emissivity.

Radiative heat transfer alone can also be measured with this device by placing it outside the quartz window and correcting for view factor and quartz absorptivity.

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APPENDIX B

PRIMARY REACTOR DESIGN TRADEOFFS

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The design approach for setting the HITS first-stage reactor design conditions is given below. This approach takes into account the fluid mechanic and chemical reaction constraints of the process. The design procedure which follows is conducted for a unit with a coal combustion capacity of about 2000 tons per week. The bench-scale design conditions are obtained from these results by specifying a bed diameter of one foot while keeping full-scale bed depth, particle size, and superficial velocity.

I. EFFECT OF PARTICLE SIZE ON MINIMUM FLUIDIZATION VELOCITY

The minimum fluidization velocity, U_{mf} , is related to the fluid and solid properties by the following relationship (Reference B-1):

$$\frac{(U_{mf})(\rho_f)(d)}{\mu} = Re_{mf} = 25.7 [(1 + 5.53 \times 10^{-5} Ga)^{1/2} - 1]$$

where

$$Ga = \frac{\rho_f (\rho_s - \rho_f) g d^3}{\mu^2}$$

Ga = Galileo number

d = particle diameter

ρ_f = density of fluid

g = gravitational acceleration

ρ_s = density of solid

μ = viscosity of fluid

For the primary reactor, the temperature is assumed to be 1600°F. The entering fluidizing gas is air, while the exit gas has the following nominal composition (% vol): N_2 - 51.55; CO - 25.18; H_2 - 14.79, CO_2 - 5.06; H_2O - 3.42.

Assuming the fluidized bed to be in the range of 3 to 8 ft deep, the bulk fluidized-bed density to be in the range of 30 to 40 lb/ft³, and the exit gas to leave the bed at about 0.5 psig, the average bed pressure is in the range of 0.8 to 1.6 psig. The average bed pressure is, therefore, assumed to be 1.2 psig and the average gas molecular weight to be ~26.8. The gas density at reactor operating temperature (1600°F) is:

$$\rho_f = (0.07465) \frac{(15.9)(492)}{(14.7)(2060)} = 0.0193 \text{ lb/ft}^3$$

The density of the solid bed material is estimated to contain ~70.7% wt ash (ρ ~2.7 g/cm³) and 29.3% wt char (ρ ~2.0 g/cm³). The true solid density is thus calculated to be:

$$\rho_s = 62.428 / \frac{.707}{2.7} + \frac{.293}{2.0} = 153 \text{ lb/ft}^3$$

B-1. Davidson, J.F. and Harrison, D. Fluidization, p. 35, Academic Press, New York, 1971.

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I, Effect of Particle Size on Minimum Fluidization Velocity (cont.)

The viscosity of air at 1600°F is 3.023×10^{-5} lb (m)/ft sec, while the viscosities of the exit gas components at 1600°F are as follows (lb(m)/ft sec $\times 10^5$): N₂ - 2.912; CO - 3.065; H₂ - 1.479; CO₂ - 2.866; H₂O - 2.777. The viscosity of the exit gas mixture can be estimated according to the following relationship (Reference B-2):

$$\mu_m = [\sum y_i \mu_i (M_i)^{0.5}] / [\sum y_i (M_i)^{0.5}]$$

where

μ_m = mixture viscosity

μ_i = component viscosity

y_i = component mole fraction

M_i = component molecular weight

The exit gas viscosity is therefore estimated to be 2.884×10^{-5} lb(m)/ft sec, and the average viscosity in the bed is considered to be the mean of the air and exit gas values, that is:

$$\begin{aligned}\mu &= (3.023 + 2.884) (10^{-5})/2 \\ &= 2.95 \times 10^{-5} \text{ lb}_m/\text{ft sec}\end{aligned}$$

Assuming bed particle diameters in the range of 100 to 10,000 microns (3.28×10^{-4} to 3.28×10^{-2} ft), minimum fluidization velocities, U_{mf} , are calculated from Equation (1). The results of these calculations are summarized in Table B-I and Figure B-1.

B-2. Reid, R.C. and Sherwood, T.K. The Properties of Gases and Liquids: Their Estimation and Condition, p. 423, McGraw-Hill Co., New York, 1966.

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TABLE B-I

EFFECT OF BED PARTICLE DIAMETER ON MINIMUM
FLUIDIZATION VELOCITY PARTICLE DIAMETER

Microns	ft	Ga	Re _{mf}	U _{mf} ft/sec
100	3.28×10^{-4}	3.85	2.74×10^{-3}	0.0128
300	9.84×10^{-4}	1.04×10^2	7.38×10^{-2}	0.115
1,000	3.28×10^{-3}	3.85×10^3	2.60	1.21
3,000	9.84×10^{-3}	1.04×10^5	41.1	6.38
10,000	3.28×10^{-2}	3.85×10^6	350	16.3

Figure B-1. Primary Reactor Superficial Velocity versus Particle Diameter

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II. EFFECT OF PARTICLE SIZE ON TERMINAL FREE-FALLING VELOCITY

The terminal free-falling particle velocity, U_t , is related to the Galileo number, Ga , by the following expression (Reference B-3):

$$U_t = \frac{(Re_t) (\mu)}{(d) (\rho_f)}$$

where

Ga and Re_t are interrelated as follows

$$Ga = 18 Re_t \quad \text{for } Ga < 3.6$$

$$Ga = 18 Re_t + 2.7 Re_t^{1.687} \quad \text{for } 3.6 < Ga \approx 10^5$$

$$Ga = (Re_t)^{2/3} \quad \text{for } Ga > \approx 10^5$$

Terminal free-falling particle velocities are calculated from the above expressions for a range of particle diameters. The results of these calculations are summarized in Table B-II and Figure B-1.

TABLE B-II

EFFECT OF BED PARTICLE DIAMETER ON TERMINAL
FREE-FALLING VELOCITY PARTICLE DIAMETER

Microns	ft	Ga	Re_t	U_t , ft/sec
100	3.28×10^{-4}	3.85	.2037	0.949
300	9.84×10^{-4}	1.04×10^2	4.134	6.42
1,000	3.28×10^{-3}	3.85×10^3	60.76	28.3
3,000	9.84×10^{-3}	1.04×10^5	5.59×10^2	86.8
10,000	3.28×10^{-2}	3.85×10^6	3.40×10^3	158

B-3. Op. cit., Davidson, J. F. and Harrison, D., p. 51.

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III. EFFECT OF PARTICLE SIZE ON OPTIMUM SUPERFICIAL VELOCITY

There is no single generally accepted relationship between particle size and optimum superficial velocity. It is known, however, that maximum mass and heat transfer coefficients between an object and the fluidized bed go through relatively flat maximums as U goes from U_{mf} to U_t . Empirically, this particular "optimum," U_{opt} , has been related to U_{mf} , U_t , optimum voidage (E_{opt}), and the drag coefficient of an individual particle (C_D) by various expressions which follow (Reference B-4).

$$U_{opt} = U_{mf} (5.0 \pm 0.5) C_D^{3/4} \quad \text{if } C_D < 10 \quad (3)$$

$$= 36 U_{mf} \quad \text{if } C_D > 10 \quad (3a)$$

$$U_{opt} = U_t (0.3 \pm 0.1) C_D^{1/4} \quad \text{if } C_D < 10 \quad (4)$$

$$= 0.6 U_t \quad \text{if } C_D > 10 \quad (4a)$$

$$U_{opt} = U_t E_{opt}^n \quad (5)$$

where

$$C_D = \frac{24}{Re} + \frac{3.6}{Re^{0.313}} \quad \text{for } Re < 2000$$

$$E_{opt} = (0.65 \pm 0.08) C_D^{1/6} \quad \text{if } C_D < 10$$

$$= 0.80 \pm 0.04 \quad \text{if } C_D > 10$$

$$n = f(Ga) \text{ given Ref. B-4, p. 53 (Figure 2.10).}$$

Calculated U_{opt} values based on these alternative correlations (Equations 3-5) for various particle sizes are summarized in Table B-III and Figure B-1.

B-4. Op. cit., Davidson, J.F., and Harrison, D., p. 52, 53, 438, 683.

TABLE B-III

EFFECT OF BED PARTICLE DIAMETER ON "OPTIMUM" SUPERFICIAL VELOCITY

Particle Diameter		Ga	Nominal Re at U_{opt}	Nominal C_D at U_{opt}	Nominal Approx.		U_{opt} (ft/sec) Based on Eqn. ()		
Microns	ft				E_{opt}	n	(3) or (3a)	(4) or (4a)	(5)
100	3.28×10^{-4}	3.85	0.115	216	0.80	4.55	0.46	0.57	$0.34 \pm .08$
300	9.84×10^{-4}	1.04×10^2	2.67	11.6	0.80	4.0	4.1	3.9	2.6 ± 0.5
1,000	3.28×10^{-3}	3.85×10^3	24.4	2.31	0.75	2.9	11 ± 1	10 ± 3	12 ± 4
3,000	9.84×10^{-3}	1.04×10^5	176	0.850	0.63	2.35	28 ± 3	25 ± 8	29 ± 8
10,000	3.28×10^{-2}	3.85×10^6	920	0.451	0.57	2.35	45 ± 4	39 ± 13	42 ± 12

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IV. PRIMARY REACTOR KINETICS AND BED SIZE

It is assumed that the inlet air to the primary reactor becomes heated to reaction temperature (1600°F) and hence reacts essentially stoichiometrically with bed char to form CO₂ within approximately 10% of bed depth. At that point the CO₂ concentration in the gas phase would be approximately 0.21 mole fraction. At the bed exit the concentration of H₂O plus CO₂ would be approximately 0.085 mole fraction, assuming the following equilibrium reaction is achieved:

Thus, the log mean concentration of CO₂ + H₂O in the char gasification portion of the bed would be approximately 0.1334 mole fraction.

It is assumed that the rate equation of Stewart and Diehl (Reference B-5) for the gasification of their char No. 2455 with CO₂ and H₂O is appropriate as a first approximation to the corresponding gasification reaction in the primary reactor. The Stewart and Diehl rate equation is as follows:

$$1 - X = A + (1 - A)e^{-kC^n t}$$

where

$$k = 9.5 \times 10^{-4} e^{+17000/T}$$

X = fraction of char reacted

T = temperature, °K

C = concentration of reacting gas (CO₂ + H₂O),
mole fraction

t = time, min.

k = apparent activity, lb/min C reacted/lb C inventory,
min⁻¹

A = weight fraction ash in unreacted char

n = order of reaction with respect to reacting gas

= 0.58

B-5. Stewart, J.T. and Diehl, E.K. "Fluidized-Bed Gasification - Process and Equipment Development," Proceedings of Third International Conference on Fluidized-Bed Combustion, pp. III-1-1 to III-1-18, EPA-650/2-73-053, Environmental Protection Agency, Washington, D.C., December 1973.

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IV, Primary Reactor Kinetics and Bed Size (cont.)

For the primary reactor, the following values for the variables are established:

$$T = 1144.27^{\circ}\text{K} (1600^{\circ}\text{F})$$

$$k = 0.03354$$

$$A = 0.4682$$

$$x = 0.3374$$

$$C = 0.1334 \text{ (log mean value)}$$

Solving the rate equation for t to define the required char residence time, it is found that $t = 96.5 \text{ min}$.

Since the carbon undergoing gasification must have a residence time of 96.5 min to attain the design extent of reaction and the carbon is to be gasified at a design rate of 5659 lb/hr (this includes the carbon assumed to be oxidized low in the bed), the bed inventory of carbon in the gasifying portion of the bed must be:

$$(5659 \text{ lb/hr}) (96.5 \text{ min}) (1 \text{ hr}/60 \text{ min}) = 9102 \text{ lb of carbon}$$

At steady-state conditions, the bed contains 0.2934 lb C/lb bed; thus, the bed inventory in the gasifying section of the bed is:

$$9102 \text{ lb C} \div 0.2934 \text{ lb C/lb bed} = 31022 \text{ lb bed}$$

Taking the true bed particle density as 153 lb/ft^3 and nominal "optimum" bed voidages (the bed voidage at nominal "optimum" superficial velocity) from Table B-III, the corresponding bulk fluidized-bed densities and required fluidized-bed volumes in the gasifying section of the bed are calculated for range of bed particle sizes. These values are summarized in Table B-IV.

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TABLE B-IV

EFFECT OF BED PARTICLE SIZE ON FLUIDIZED-BED DENSITY
AND REQUIRED BED VOLUME IN GASIFYING SECTION

Particle Diameter		Nominal E_{opt}	Bulk Fluidized Bed Density, lb/ft ³	Required Bed Vol. In Gasifying Sect., ft ³
Microns	ft			
100	3.28×10^{-4}	0.80	31	1010
300	9.84×10^{-4}	0.80	31	1010
1,000	3.28×10^{-3}	0.75	38	810
3,000	9.84×10^{-3}	0.63	57	550
10,000	3.28×10^{-2}	0.57	66	470

The primary reactor design gas flowrate is maximum at the top of the bed (~3219 lb mole/hr) where the pressure is ~0.5 psig. This corresponds to 1300 ft³/sec. Assuming maximum design superficial velocities in the range of 1 to 12 ft/sec, corresponding "optimum" bed particle diameter, required bed cross-sectional areas, and bed depths are determined. These values are summarized in Table B-V.

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TABLE B-V

EFFECT OF DESIGN SUPERFICIAL VELOCITY
ON OPTIMUM PARTICLE SIZE AND BED DIMENSIONS

Design Vel., U, ft/sec	Opt. Particle Dia, Microns	Required Bed Dimensions (Nominal)			
		Cross-Sect. Area, ft ²	Gasification Bed Vol, ft ³	Bed Depth, ft	
				Gasifying Sect.	Total
1	130-200	1300	1010	0.78	0.86
2	190-310	650	1010	1.55	1.73
3	250-420	433	990	2.29	2.54
4	300-540	325	950	2.92	3.25
5	350-670	260	915	3.52	3.91
6	390-800	217	890	4.10	4.56
7	440-950	186	865	4.65	5.17
8	490-1100	162	845	5.22	5.80
10	600-1500	130	795	6.12	6.80
12	700-1900	108	740	6.85	7.61

TABLE B-VI

EFFECT OF DESIGN SUPERFICIAL VELOCITY ON PRIMARY REACTOR
DIMENSIONS AND RELATIVE COST

Design Vel., U, ft/sec	Reactor I.D., ft	Reactor Section Heights, ft				Reactor Surface Area, ft ²	Approx. Relative Reactor Cost ⁽³⁾
		Windbox ⁽¹⁾	Bed	Freeboard ⁽²⁾	Total		
1	40.7	5	0.9	15.5	21.4	4730	1.60
2	28.8	5	1.7	22.5	29.2	3640	1.23
3	23.5	5	2.5	28.1	35.6	3290	1.11
4	20.3	5	3.2	32.8	41.0	3110	1.05
5	18.2	5	3.9	36.9	45.8	3020	1.02
6	16.6	5	4.6	(40.8)	50.4*	2960*	1.00*
7	15.4	5	5.2	(44.2)	54.4*	2920*	0.99*
8	14.4	5	5.8	(47.5)	58.3*	2890*	0.98*
10	12.9	5	6.8	(53.5)	65.3*	2850*	0.96*
12	11.7	5	7.6	(56.3)	68.9*	2700*	0.91*

(1) Assumed to be ~5 ft to permit man-entry for installation of refractory lining, maintenance, and cleanout.

(2) Estimated based on the correlation of Zenz and Weil (Chem. Engr., pp. 170-1, September 17, 1962); values in parenthesis are extrapolations and subject to increased uncertainties.

(3) Assumed to be proportional to reactor surface area.

*These values have increased uncertainty due to the questionable freeboard heights.

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V. REACTOR DIMENSIONS AND RELATIVE COST VERSUS DESIGN VELOCITY

Assuming the primary reactor to be cylindrical in shape with a radially dished top closure and the reactor cost to be approximately proportional to reactor weight (which is, in turn, approximately proportional to reactor surface area), reactor dimensions and relative cost are calculated for a range of design superficial velocities. These values are summarized in Table V-VI

From Table B-VI it can be seen that reactor cost decreases rapidly with increasing design superficial velocity at low velocities but becomes relatively constant at velocities near 5 to 6 ft/sec. Note, however, that bed depth continuously increases with design superficial velocity, and this corresponds to increasing fluidizing air inlet pressures which translates into higher air supply subsystem capital and operating costs. The optimum design superficial velocity is, thus, in the range of about 4 to 8 ft/sec and most probably very near 6 ft/sec.

VI. SELECTED PRIMARY REACTOR DESIGN PARAMETERS FOR BASELINE FULL-SCALE UNIT

On the basis of the tradeoff analyses described in the preceding sections, the following design parameters are selected as being the first approximation for an optimum single unit, full-scale reactor.

- ° Superficial velocity (based on exit gas flowrate): ~6 ft/sec.
- ° Mean bed particle diameter: ~800 microns (~003 in.).
- ° Operating temperature: ~1600°F.
- ° Average bed operating pressure: ~1.1 psig.
- ° Inside bed diameter: ~17 ft.
- ° Fluidized bed depth: ~5 ft.
- ° Freeboard height: ~40 ft.
- ° Windbox height: ~5 ft.
- ° Overall reactor dimensions: 18 ft O.D. x 50 ft.

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APPENDIX C

HITS CHECKOUT, CALIBRATION AND OPERATING PROCEDURES

APPENDIX C

HITS CHECKOUT, CALIBRATION & OPERATING PROCEDURES

Included in Appendix C are those procedures which were set up for system checkout, calibration and operation of the HITS system. They are included here for documentation and to make them readily available for future system operation.

- C.1. COMPONENT/SUBSYSTEM CHECKOUT
 - A. General Requirements Summary
 - B. Components/Subsystems Requiring Checkout
 - C. Subsystem Definition
 - D. Checkout Procedure Outline
- C.2. COMPONENT CALIBRATION
 - A. Systems Requiring Calibration
 - B. HITS Gas Flow Orifice Calibration Procedure
- C.3. SYSTEM OPERATING PROCEDURES
 - A. Outline
 - B. HITS System Set Points
 - C. Startup, Shutdown and Emergency Procedures
- C.4. REFRACTORY CURE PROCEDURE
 - A. Reactor
 - B. Combustor and Stack
- C.5. ADDITIONAL INFORMATION SOURCES

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HITS SUBSYSTEM CHECKOUT

C.1.A GENERAL REQUIREMENTS SUMMARY

1. Prior to operating a system, all mechanical, plumbing and electrical components must be checked to verify they have been installed correctly.
2. Any required deviations from system drawings must be marked on the HITS System As-Built Drawings (file copy).
3. All plumbing systems must be blown out, with critical components isolated, to remove dirt and foreign objects.
4. All gas plumbing systems must be leak checked. In particular, lines to all pressure gages and transmitters must be checked.

C.1.B. COMPONENTS/SUBSYSTEMS REQUIRING CHECKOUT

A. MAJOR ITEMS

1. Air Compressor
2. Roots Blower
3. Propane Supply
4. GN_2 Supply
5. GO_2 Supply
6. Cooling Water System
7. Combustion Air Supply System
8. Combustion Air Preheater
9. Fluidization Air Supply System
10. Transport Air Supply System
11. Solids Feed Systems
12. Weigh Systems
13. Instrument Air Supply
14. B-1 Propane Burner
15. Char Removal System
16. Instrument Purge Systems
17. R-1 Reactor
 - a. Windbox
 - b. Tuyere
 - c. Bed
 - d. Disengagement
 - e. Lid
18. Instrumentation - Pressures
19. Instrumentation - Temperature
20. Instrumentation - Other
21. Off Gas Piping
22. Cyclone/Fines Recovery System

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C.1.B. (continued)

- 23. Product Gas Combustor
 - a. Flow Meter
 - b. B-2 Burner
 - c. B-3 Igniter
 - d. Stack
- 24. Gas Measurement Systems
 - a. Particle Collection Systems
 - b. Gas Analysis Systems
- 25. Combustion Measurement Systems
 - a. Optical
 - b. Thermocouple
 - c. Heat Flux
- 26. Control Room
- 27. Pressure Gage Calibration
- 28. Data Reduction/Analysis Procedures
- 29. Carbon Monoxide Monitoring System
- 30. Fire Protection Equipment

C.1.C. SUBSYSTEM DEFINITION

1. AIR COMPRESSOR

1.1 25 HP Compressor Package

- 1.1.1 Motor Switch**
- 1.1.2 Motor**
- 1.1.3 Air Compressor**
- 1.1.4 Aftercooler**
- 1.1.5 Filter**
- 1.1.6 Storage Tank**
- 1.1.7 Pressure Relief**
- 1.1.8 Condensate Dryer**

1.2 Refrigerated Dryer

1.3 Supply Line to Facility & Manual Shutoff

2. ROOTS BLOWER

- Motor Starter**
- Motor**
- Blower**
- Silencers**
- Bypass Plumbing**
- Supply Line to Facility**

3. PROPANE SUPPLY

- 500 Gal. Tank**
- First Stage Regulator (P_{Tank} to 10 psig)**
- Supply Line to Facility & Manual Shutoff**

4. GN₂ SUPPLY

- First Stage Regulation (2400 to approx. 100 psig)**
- Safety Relief**
- Filter**
- Supply Line to Facility & Manual Shutoff**
- Air/GN₂ Control Valves**

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C.1.C. (continued)

5. GO₂ SUPPLY
 - Liquid Dewar
 - Pressure Regulator
 - Pressure Relief
 - Supply Line to Facility with Valves
6. COOLING WATER SYSTEM
 - Plant Supply Line Valves & Meter
 - Storage Tank
 - Pump & Starter
 - Lines to Char Cooler & Return
7. COMBUSTION AIR SUPPLY SYSTEM
 - Flow Section
 - Flow Control Valve
 - Flow Controller/Transmitter
 - Misc. Plumbing
8. COMBUSTION AIR PREHEATER
 - Air Temperature Controller
 - Element Temperature Controller
 - Power Controls
9. FLUIDIZATION AIR SUPPLY SYSTEM
 - Pressure Control
 - Flow Section
 - Flow Control Valve
 - Flow Controller/Transmitter
 - Misc. Plumbing
10. TRANSPORT AIR SUPPLY
 - Flow Section
 - Flow Control Valve
 - Flow Controller/Transmitter
 - Misc. Plumbing

C.1.C. (continued)

11. SOLIDS FEED SYSTEM

Coal

- Refill Hopper
- Run Hopper
- Level Sensor
- N₂ Pressure System
- Vent System
- Coal Feed Auger & Control

Limestone

- Refill Hopper
- Run Hopper
- Level Sensor
- N₂ Pressure System
- Vent System
- Limestone Feed Auger & Control

12. WEIGH SYSTEMS

Mechanical Restraints

- Pressure/Vent Line
- Fill Line Boot
- Output Line Boot
- Drive Motor Power Line

Scale Systems

13. INSTRUMENT AIR SUPPLY

Source: C-1 Compressor

Location: Tied into main air supply at first deck level

Components: Two filters on alternate legs and instrument air regulator

14. B-1 PROPANE BURNER

Consists of:

- Propane Supply
- Air Supply
- Protectifier Controls
- Propane Control System
- Modified Maxon Burner Assembly

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C.1.C. (continued)

15. CHAR REMOVAL SYSTEM

Consists of:

- High, medium and low char removal points on reactor
- Cooled removal cross and interconnecting piping
- Cooled char auger with drive motor and gear box
- Supports (spring hanger and "A"-Frame)
- SCR controller for DC drive motor
- Once-through cooling water system

16. OFF GAS PIPING

Consists of:

- 4" cross elbow with blind-flanged cleanouts
- 4" line with sample port
- Cyclone with 2" down leg and 3" outlet
- 3" cross elbow with blind-flanged cleanouts
- Constant load hanger single point pipe support
- 3" down leg with bellows
- Constant load equalizers around bellows
- Product gas metering nozzle
- 3" line with 3" to 4" adaptor
- 4" bellows with adaptor to burner inlet
- Fines eductor
- 2" fines diversion valve
- 1/2" eductor isolation valve
- Fines collection pot with flexible coupling and quick disconnect
- High temperature pipe and fitting insulation with removeable access ports

17. PRODUCT GAS COMBUSTOR SYSTEM

- B-2 High temperature burner assembly
- B-3 Propane-air pilot igniter
- B-4 Refractory-lined combustor assembly
- 12" Stainless exhaust line

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C.1.C. (continued)

18. EXHAUST STACK

- 18" dia x 34' stack
- Stack eductor
- Stack sample port
- Stack cleanout port

19. COMBUSTOR INSTRUMENTATION

- Flame temperature thermocouple: PT/PT-Rh thermocouple in 3/4" alumina thermal well
- L&N optical pyrometer for hot wall or hot target temperature measurement
- Hy-cal water-cooled calorimeter for combustor heat flux measurement in gas-cooled support/protection tube

20. GAGE/PURGE PANEL

- Pressure gage panel with process layout
- Warning indicators
- Purge/backflush controls

21. CONTROL CONSOLE

- Primary process electrical controls
- Primary process pneumatic controls
- Data logger
- Temperature recorder
- Data display

22. SAMPLE ACQUISITION SYSTEM

- Sample probe
- Line heater
- Gas filter
- Condenser
- Sample pump
- Bag sample port
- Particle collection filter
- Constant volume sampling system
- N₂ purge

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C.1.C. (continued)

23. GAS ANALYSIS SYSTEM

- Flow controls
- Gas cleanup filter
- Gas cooler
- Calibration gas system
- CO analyzer
- CO₂ analyzer
- O₂ analyzer
- SO₂/H₂S analyzer
- NO/NO_x analyzer
- Gas chromatograph system

24. SUPPORT STRUCTURE

- Two-decks 12' x 12'
- 8' x 8' combustor deck
- Access stairs
- Emergency ladder

25. CARBON MONOXIDE MONITORING SYSTEM

- 2 ea. MSA portable CO monitor/alarm units

Usage: General monitoring of potential leak points on system (CO measurement in gas stream is performed by gas analysis system)

Calibration: Standard gas sources are provided with the units for field calibration

Location: HITS control room

26. FIRE PROTECTION EQUIPMENT

- a. Fire hose - located on reel immediately south of facility. To be used for coal or control room structure fires.
- b. CO₂ extinguisher - located in control room. To be used for electrical or small paper fires.
- c. CO₂ extinguisher - located on HITS structure. To be used for electrical or small coal fires.

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C.1.D. HITS COMPONENT/SUBSYSTEM CHECKOUT PROCEDURE OUTLINE

1.0 INITIAL PHASE

- 1.1 Detail written procedures available
- 1.2 Related/dependent components/subsystems checked out
- 1.3 Manpower/supplies/facility available

2.0 ONE-TIME TESTS

- 2.1 Physical/electrical/pneumatic/per design requirements
- 2.2 Installed per Mfgr's directions
- 2.3 Leak/proof/mechanical/electrical tests
- 2.4 Special tests per Mfgr's/other specifications

3.0 OPERATION CHECKOUT TESTS

- 3.1 Startup
- 3.2 Steady state
- 3.3 Calibration
- 3.4 Upset/off design/limits
- 3.5 Shutdown
- 3.6 Secure system

4.0 NORMAL OPERATION PROCEDURE

- 4.1 Startup
- 4.2 Steady state
- 4.3 Modify operating point
- 4.4 Shutdown

5.0 MAINTENANCE

- 5.1 Preoperation
- 5.2 Periodic
- 5.3 Supplies/spares
- 5.4 Trouble shooting

6.0 BACKUP DATA

- 6.1 Manuals
- 6.2 Drawings
- 6.3 Vendor data
- 6.4 Calibration data

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C.2.A. SYSTEMS REQUIRING CALIBRATION

<u>Device</u>	<u>Desired Accuracy</u>
1. All pressure gages used for data	<u>+ 1%</u>
2. Flow orifices used for data	
FO-200 Solids transport air orifice	<u>+ 1%</u>
FO-300 Fluidization air orifice	<u>+ 1%</u>
FO-400 Combustor air orifice	<u>+ 1%</u>
FO-602 Product gas orifice	<u>+ 1%</u>
3. Meters/instruments used for temperature data	<u>+ 1%</u>
4. Coal hopper weigh scale/load platform	<u>+ 1%</u>
5. Limestone hopper weigh scale/load platform	<u>+ 1%</u>
6. Char weigh scale	<u>+ 1%</u>
7. Fines weigh scale	<u>+ 1%</u>
8. Coal auger	<u>+ 10%</u>
9. Limestone auger	<u>+ 10%</u>
10. Char auger	<u>+ 10%</u>
11. Optical pyrometer (L&N)	<u>+ 1%</u>
12. Calorimeter	Use vendor supplied data
13. Flow meters	
FI-300 Nitrogen diluent	<u>+ 2%</u>
FI-302 Oxygen enrichment	<u>+ 2%</u>
FI-606 B-1 Propane	<u>+ 2%</u>
FI-600 System purge	<u>+ 3%</u>

C.2.B. HITS GAS FLOW ORIFICE CALIBRATION PROCEDURE

- 1.0 In-place calibration (end-to-end calibration of orifice, pressure transmitter/controller and gages).**
 - 1.1 Check out gas flow system in which orifice is to be used (see gas flow system checkout procedure).**
 - 1.2 Verify orifice dimensions and proper installation.**
 - 1.3 Provide regulated, high pressure, metered GN₂ source with adequate flow capacity. See Figure C-1 for recommended setup configuration.**
 - 1.4 Determine calibration range required and range of pressure and flow conditions from Table C-1.**
 - 1.5 Determine reference metering nozzle to give sonic flow over calibration range.**
 - 1.6 Setup orifice to be calibrated in actual plumbing to be used, on site. Perform thorough leak check of system and pressure gage check by blanking off system, pressurizing, and monitoring pressure gages.**
 - 1.7 Calibrate over range of conditions and record data. Orifice FO-602 is calibrated over a wider range of back pressure to isolate Reynold's number effects on flow coefficient.**

ORIFICES TO BE CALIBRATED:

FO-200
FO-300
FO-400
FO-602

Figure C-1 Flow Orifice Calibration Setup

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TABLE C-1

HITS Flow Meter Calibration
Set Up Information

Flow Meter	Flow System	Nominal Flow, ACFM	Approx. Equivalent GN ₂ Flow, Lb/Sec	Reference Nozzle	Nominal Pup Ref. Psig
FO-200	Transport Air	3	0.0038	C-2	220
FO-300	Fluidization Air	34	0.043	C-6	135
FO-400	Combustor Air	77	0.096	C-6	325
FO-602	Product Gas	246	0.31	C-5	115
FI-300	Nitrogen Diluent	11	0.014	C-7	205
FI-302	Oxygen Enrichment	4.5	0.0056	C-2	305
FI-606	B-1 Propane	1.8	0.0022	C-2	115

C.3.A. HITS SYSTEM OPERATING PROCEDURE - OUTLINE

- 1.0 PRE-TEST PREPARATION**
 - 1.1 Systems Operational
 - 1.2 Supplies Available
 - 1.3 Personnel Scheduled
 - 1.4 "A" Area Schedule Conflicts Resolved
 - 1.5 Test Conditions Defined
- 2.0 START UP**
 - 2.1 Auxiliaries
 - 2.2 Fluidization Air Startup
 - 2.3 B-1 Burner Startup
 - 2.4 Limestone Feeder Startup
 - 2.5 Char Removal Startup
 - 2.6 Combustor Air Startup
 - 2.7 B-3 Igniter Startup
 - 2.8 Gas and Particle Analysis Systems Ready
- 3.0 SETTING TEST CONDITIONS**
 - 3.1 Ramp Reactor Temperature to 1100°F
 - 3.2 Set Air Flowrate
 - 3.3 Shutdown B-1 Burner
 - 3.4 Coke Feed Startup
 - 3.5 Ramp Reactor Temperature to Test Conditions
 - 3.6 Switch to Coal Feed
 - 3.7 Trim Flowrates and Temperatures
- 4.0 DATA RUNS**
 - 4.1 Verify System is at Equilibrium
 - 4.2 Take Data
 - 4.3 Set New Test Condition
 - 4.4 Repeat 4.1, 4.2 & 4.3 as required

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C.3.A. (continued)

5.0 SYSTEM SHUTDOWN

5.1 Emergency Shutdown

5.1.1 Switch to GN₂

5.1.2 Stop solids feeders

5.1.3 Depending on emergency condition:

Unload bed: Increase char auger rate

Drop bed: Stop GN₂ flowrate

Complete shutdown: Stop combustor air heater, combustor air and B-3 igniter

5.2 Shutdown (for times greater than 1 day)

5.2.1 Stop solids feed

5.2.2 Manually reduce air flow to prevent temperature overshoot

5.2.3 Switch to GN₂

5.2.4 Shutdown B-1 igniter, E-1 air heater and combustor air flow

5.2.5 Increase char auger rate to withdraw bed

5.2.6 When bed material is withdrawn, stop char auger and GN₂

5.2.7 For rapid cooldown, start fluidization air flow

5.3 Shutdown for Short Periods

5.3.1 Stop solids feed

5.3.2 Stop char auger

5.3.3 Manually reduce air flow to prevent temperature overshoot

5.3.4 Switch to GN₂

5.3.5 Shutdown B-1 igniter, E-1 air heater and combustor air flow

5.3.6 Stop GN₂ flow to slump bed

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C.3.B. HITS SYSTEM SET POINTS

1.0 SPECIFIED TEST CONDITIONS

1.1	Bed temperature	_____ °F
1.2	Ca/S ratio	_____
1.3	Coal type	_____
1.4	Fluidization velocity	_____ ft/sec
1.5	Bed height	_____ ft
1.6	Fluidization gas oxygen concentration	_____ %
1.7	Coal flow rate	_____ lb/hr

2.0 DERIVED SYSTEM SET POINTS

2.1	Bed temperature	
	TIC-611/TE-604	_____ °F
2.2	Coal flow rate	
	Coal auger controller nominal setting	_____
2.3	Limestone flow rate	_____ lb/hr
	Limestone auger controller nominal setting	_____
2.4	Fluidization air flow rate	
	$\Delta P = \text{_____ psi @ } P_{up} = 25 \text{ psig/70}^\circ\text{F}$	_____ SCFM
	Transport air flow rate	
	$\Delta P = \text{_____ psi @ } P_{up} = 25 \text{ psig/70}^\circ\text{F}$	_____ SCFM
2.5	Char production rate	_____ lb/hr
	Char auger controller nominal setting	_____
2.6	Combustor air flow rate	
	$\Delta P = \text{_____ psi @ } P_{up} = 4 \text{ psig/70}^\circ\text{F}$	_____ SCFM
2.7	Combustor air temperature	_____ °F

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C.3.C. HITS SYSTEM START UP, SHUTDOWN & EMERGENCY PROCEDURES

A. SYSTEM PREHEAT TO COKE/CHAR IGNITION TEMPERATURE

1. Main Air Supply Turn On

- 1.1 Turn on aftercooler water
- 1.2 Turn on refrigerated dryer
- 1.3 Turn on C-1 air compressor
- 1.4 Open V-102 at air receiver
- 1.5 Verify that V-103 is set to "Air" position
- 1.6 Verify that R-100 is set to 30 psig
- 1.7 Verify that V-207 is closed
- 1.8 Open V-304 & V-306 full
- 1.9 Open V-305 to set air flow
- 1.10 Open V-650, bypass air, until PI-300 starts to fall

NOTE: Air system should remain on at all times during test cycle;
it provides purge of the reactor in the event of a propane valve leak.

2. Propane Supply Turn On

- 2.1 Verify that protectifier control is off and Maxon main propane valve is closed
- 2.2 Open propane valve at tank
- 2.3 Open propane valve at structure
- 2.4 Open 1" propane cock to propane control system
- 2.5 Verify that manual pilot valve is properly set
- 2.6 Close main propane manual valve

3. Burner Ignition

- 3.1 Verify that air is on to burner
- 3.2 Activate data logger to monitor system temperature
- 3.3 Turn power on to protectifier control
- 3.4 Turn protectifier control to "Ign" and release to "On"; timer light will come on and remain on for several minutes for purge
- 3.5 When yellow "Timer" light goes out, turn control switch to "Ign", pilot should light
- 3.6 Start cure cycle with pilot only and full bypass air. Control windbox temperature so that thermocouple inserted into gas stream (TE-602) follows Figure C-2 Temperature-Time Curve

C-20

Figure C-2 Reactor Refractory
Cure Cycle

C.3.C. (continued)

CAUTION: DO NOT EXCEED HEATING RATE SPECIFIED. ERR ON
LOW SIDE OF THE TEMPERATURE CURVE.

- 3.7 If temperature is excessive, decrease pilot propane flow _____
- 3.8 To increase temperature, slowly open main burner manual valve.
Monitor at sight port at rear (north) end of windbox to be sure
burner is not sooting. _____

NOTE: BURNER HAS SHOWN TENDENCY TO SOOT: WHEN MAXIMUM
PROPANE FLOW SHORT OF SOOTING IS REACHED, FURTHER
TEMPERATURE INCREASES MUST BE OBTAINED BY GRADUALLY
CLOSING V-650 TO REDUCE BYPASS AIR FLOW.

4. Emergency Conditions

- 4.1 Excessive gas temperature (above demand curve on Figure C-2)
 - 4.1.1 Decrease propane flow
 - 4.1.2 Increase bypass air flow
- 4.2 Propane leak or other hazardous condition
 - 4.1.1 Shutdown burner
 - 4.1.2 Turn off propane at supply
 - 4.1.3 Continue air flow for purge
- 4.3 Excess metal temperature
 - 4.1.1 If carbon steel temperature will exceed allowable limit when temperature cycle is followed (at present allowable carbon steel maximum temperature is 650°F). Hold at present value and consult project engineer.
- 4.4 Other system failure
 - All likely failure modes can be corrected by turning off the propane burner and/or turning off the propane supply to the burner.

5. General Requirements

- 5.1 Unauthorized personnel must remain off the structure and away from the windbox and burner.

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C.3.C. (continued)

- 5.2 Hot metal parts must be guarded by shielding with perforated metal sheet or other procedure to prevent inadvertent contact while working in the area.

NOTE: NO PERSONNEL ARE AUTHORIZED TO WORK AROUND HOT, UNSHIELDED HARDWARE. PROVIDE APPROPRIATE THERMAL PROTECTION BEFORE WORKING NEAR SYSTEM.

5.3 Off-shift operation

A minimum of two people must be working on HITS during off-shift operation.
A manloading schedule is posted in the control room.

B. AUXILIARY SYSTEMS STARTUP

1. Cooling Water
 - 1.1 Turn on cooling water to char auger
2. Gas Analysis System
 - 2.1 Warm up system electronics.
 - 2.2 Perform system calibration with zero and span gases (see separate procedure Gas Analysis System Operation).
 - 2.3 Perform end-to-end reference gas calibration on both continuous analysis system and gas chromatograph.
3. Particle and Water Analysis Systems
 - 3.1 Activate probe purge and line heater
 - 3.2 Check out and prepare impactor and ceramic filter thimble per Particle Analysis System Operation procedure.
4. Heat Flux Calorimeter
 - 4.1 Turn on calorimeter cooling water.
 - 4.2 Turn on calorimeter purge GN₂.
5. Optical Pyrometer
 - 5.1 Mount pyrometer at viewing port.
6. Instrument Purges
 - 6.1 Turn on main purge flow V-642 and set total purge flow, FI-600.
 - 6.2 Monitor individual purge meters and verify that flow rates are satisfactory.

C.3.C. (continued)

C. COMBUSTOR ACTIVATION

1. Air Supply

- 1.1 Verify V-400 is closed.
- 1.2 Close V-404 and V-405.
- 1.3 Set FIC-400 to design flow.
- 1.4 Open blower bypass.
- 1.5 Turn on blower, C-2.
- 1.6 Close bypass until PI-405 reaches 4.0 psig.
- 1.7 Open V-404 and V-405.

2. Heater

- 2.1 Set JC-401, element limit, to 1500°F.
- 2.2 Set JC-400, gas temperature, to 200°F.
- 2.3 Turn on E-1 heater.
- 2.4 Combustor air temperature TE-402, should rise to 200°F and hold.

3. Pilot Igniter

- 3.1 Verify that air supply is on.
- 3.2 Turn on propane supply to igniter.
- 3.3 Arm igniter control, wait for purge timer to expire.
- 3.4 Ignite pilot igniter.
- 3.5 Verify that igniter flame is satisfactory.

D. SOLIDS FEED SYSTEMS STARTUP

When transferring solids to hoppers wear dust masks to avoid dust inhalation. Minimize dusting and spills. Vacuum up any material spilled during transfer.

1. Fuel Feeder Preparation

- 1.1 Verify that proper feed stock is being used (only coke or char for startup).

NOTE: COAL IS NOT TO BE INTRODUCED UNTIL SYSTEM HAS REACHED STEADY OPERATING TEMPERATURE IN EXCESS OF 1400°F. IF COAL IS LEFT IN FEED SYSTEM FROM PREVIOUS TEST IT MUST BE REMOVED FROM SYSTEM BEFORE INTRODUCING STARTUP FUEL.

- 1.2 Close valves V-512 and V-500 to isolate H-3 feed hopper pressurization system.

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C.3.C. (continued)

- 1.3 Close hopper outlet valve V-510.
- 1.4 Vent hopper with small vent valve V-509A until PI-501 reads ambient.
- 1.5 Open large vent valve V-509B.
- 1.6 Fill H-1 supply hopper with fuel through bug screen. Record fill on feeder fill log sheet.
- 1.7 Open V-504 and V-503 to transfer coal to H-3 feed hopper.
- 1.8 When coal is transferred, close V-503 and V-504 to isolate fill hopper.
- 1.9 Start up transport air flow by opening V-201, V-204, V-205 and V-206.
- 1.10 Adjust FIC-201 to design flow rate.
- 1.11 Verify that R-500 is set to 90% of PI-204, design feed pressure.
- 1.12 Close V-509A and B.
- 1.13 Open V-512 slowly to pressurize hopper to slightly below feed line pressure, PI-204.
- 1.14 Open V-500 to equalize hopper and line pressure.
- 1.15 Unlock scale WE-500 if locked. Record weight and time on solids feed log sheet.
2. Limestone Feeder Preparation
 - 2.1 Verify that proper feed stock is on hand.
 - 2.2 Close valves V-513 and V-501 to isolate H-4 feed hopper pressurization system.
 - 2.3 Close hopper outlet valve V-511.
 - 2.4 Vent hopper with small vent valve V-508A until PI-502 reads ambient.
 - 2.5 Open large vent valve V-508B.
 - 2.6 Fill H-2 supply hopper with limestone through bug screen. Record fill on feeder fill log sheet.
 - 2.7 Open V-506 and V-505 to transfer coal to H-4 feed hopper.
 - 2.8 When limestone is transferred, close V-505 and V-506 to isolate fill hopper.
 - 2.9 Close V-508A and B.
 - 2.10 Open V-513 slowly to pressurize hopper to slightly below feed line pressure, PI-204.
 - 2.11 Open V-501 to equalize hopper and line pressure.
 - 2.12 Unlock scale WE-501 if locked. Record weight and time on solids feed log sheet.

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C.3.C. (continued)

3. Char Feeder Preparation

- 3.1 Verify that cooling system is operating.
- 3.2 Mount char storage hopper H-6. Identify drum and record tare weight on char log sheet.
- 3.3 Turn on char line purge V-715 and set flow rate FI-702.
- 3.4 Open char transfer valve V-710.
- 3.5 Close vent to stack; V-714.

4. Fines Loop

4.1 Fines Recycle Operation

- 4.1.1 Close fines transfer valve V-703.
- 4.1.2 Open fines injector valve V-705.
- 4.1.3 Open fines GN₂ supply valves V-700.
- 4.1.4 Adjust fines GN₂ flow rate with V-701 and FI-700.

4.2 Fines Collection Operation

- 4.2.1 Install fines collection hopper, H-5.
- 4.2.2 Close V-702, vent to stack.
- 4.2.3 Open V-703, fines transfer valve.
- 4.2.4 Close V-705, fines injection valve.
- 4.2.5 Close V-701, fines GN₂ control valve.

E. SET TEST CONDITIONS

- 1.1 Transfer charge of char to fill bed.
- 1.2 Monitor bed temperature to determine when ignition occurs (approx. 800°F).
- 1.3 When ignition is reached, turn off propane.
- 1.4 As operating temperature is approached, initiate char input from coal feed hopper to control bed temperature.
- 1.5 Start char removal auger to control bed height.
- 1.6 Continue to operate with coke/char input until coldest point in off-gas piping is above 1200°F. At this point load fuel hopper with coal and continue operating with coal.
- 1.7 Trim coal flow rate to hold temperature.
- 1.8 Trim fluidization air flow to design value.
- 1.9 Trim char auger to set design bed depth.

C.3.C. (continued)

- 1.10 Trim limestone feeder to set design calcium/sulfur ratio.
- 1.11 Set controllers to automatic operation.
- 1.12 Operation for twelve (12) hours of steady-state conditions to assure that bed composition has reached equilibrium.
- 1.13 Take steady-state data per Data Acquisition Procedure.
- 1.14 Verify that data are satisfactory.
- 1.15 Go to next test condition.
- 1.16 Remain at steady conditions for four (4) hours after temperatures and flows have stabilized for small changes in test conditions.
- 1.17 Repeat data acquisition procedure and reset test conditions as required.

F. SYSTEM SHUTDOWN AND RECOVERY

1. Emergency Conditions - Automatic

- 1.1 In the event of a sensed out of specification condition, the system will automatically go into a safety shutdown mode. This is initiated by bed over-temperature, loss of fluidization air, or loss of combustor air. Under these conditions the system is switched from air to nitrogen, and the solids feeders are stopped.
- 1.2 Recovery (Hot Bed) - Correct the out of specification condition and bring the system back to operating conditions by:
 - 1.2.1 Dilute the fluidization gas flow with GN₂ by opening V-303 and setting full flow on FI-300.
 - 1.2.2 Switch fluidization gas supply back to air.
 - 1.2.3 Start solids feeders.
 - 1.2.4 Slowly close V-303 to reduce nitrogen dilution.
 - 1.2.5 Bed temperature should rise smoothly to operating conditions.
 - 1.2.6 Trim solids feed, char and air flows, resume testing.
- 1.3 Recovery (bed below 1000°F)
 - 1.3.1 Repeat normal startup procedure with propane heating.

C.3.C. (continued)

2. Emergency Conditions - Manual shutdown to manually respond to emergency conditions.
 - 2.1 Switch V-103 to GN₂.
 - 2.2 Stop solids feeders and char removal drives.
 - 2.3 Correct out of specification condition.
 - 2.4 Restart system as in E 1.2 above.
3. Shutdown and Secure System
 - 3.1 Stop solids feeders.
 - 3.2 Switch to GN₂.
 - 3.3 To shutdown with bed inventory, stop char removal system.
 - 3.4 When char auger has cooled, turn off cooling water.
 - 3.5 Reduce GN₂ flow to about one-half normal flow.
 - 3.6 When bed temperature has dropped to below 500°F, switch back to air flow and continue cooling.
 - 3.7 Turn off fines injection nitrogen.
 - 3.8 Withdraw calorimeter and turn off cooling water and GN₂.
 - 3.9 Turn off manual propane supply valve.

G. CHAR HOPPER OPERATION

NOTE: THE FOLLOWING OPERATION MUST BE ACCOMPLISHED RAPIDLY TO REDUCE THE UPSET IN SYSTEM OPERATION.

Where practical, system upsets such as char drum replacement and feed hopper refill, should be planned to occur during the transient period between steady-state operating levels.

- 1.1 Prepare replacement hopper with quick disconnect and have ready for installation in system.
- 1.2 Stop char auger (SIC-700).
- 1.3 Close V-710, char transfer valve.
- 1.4 Open V-714, char hopper vent to stack.
- 1.5 After venting system disconnect full drum, install empty drum. Seal full drum and remove.
- 1.6 Close vent, V-714.
- 1.7 Open V-710, char transfer valve.
- 1.8 Restart char auger.

C.4.A HITS REACTOR REFRACTORY CURE PROCEDURE

1.0 INTRODUCTION

Satisfactory operation of the refractory in the HITS reactor and in the product gas combustor requires controlled heating to gradually remove moisture and to bring the refractories up to operating temperature for proper cure. Figure C-2 is the temperature-time history recommended for proper cure of the reactor refractory. Long times are required because of the slow diffusion rate of moisture through the 12" thick refractory. The reactor is cured using the B-1 propane burner as a heat source. The product gas combustor refractory will be cured separately using the combustor air preheater.

CAUTION: PERMANENT DAMAGE WILL OCCUR TO THE REFRACTORY LINER IF THE HEATING SCHEDULE IS NOT FOLLOWED. IN PARTICULAR THE 200°F LIMIT MUST NOT BE EXCEEDED UNTIL ALL FREE MOISTURE HAS BEEN REMOVED. HIGHER HEATING RATES OR SHORTER TIMES AT TEMPERATURE ARE DAMAGING.

Times shown in Figure C-2 are minimums judged to be required to reach equilibrium dryness at each temperature level.

2.0 INITIAL SYSTEM CHECKOUT

To conduct the refractory cure procedure requires that the following subsystems be checked out and operational.

2.1 Main Air Supply

This subsystem consists of the air source and flow controls. The air source is the C-1 compressor, its controls, aftercooler, the refrigerated dryer, filters, receiver and condensate trap. The primary flow controls are the R-100 regulator and the flow control station, consisting of flow orifice FO-300, flow control valve V-305, its transmitter PT-304, the indicating flow controller FIC-304, and the instrument air supply.

2.1.1 Air Compressor

This unit is Worthington type - RS 25 HP rotary compressor.

NOTE: BEFORE PROCEEDING WITH COMPRESSOR CHECKOUT, READ
INSTALLATION AND OPERATION INSTRUCTIONS #3408-E3, FEB.
1979, IN HITS EQUIPMENT INSTRUCTIONS.

Completed
By: _____

2.1.1.1 Verify that physical and electrical hookup is proper. _____

.1 Close V-102 to isolate from system. _____

.2 Verify that motor rotation is correct. Jog unit and
determine that fan rotation is per arrow on fan guard
(clockwise when viewed from motor end of unit). _____

.3 Check oil level. Should be above center but not com-
pletely full. This factory fill is good for 500 hrs.
of operation. _____

.4 Turn on refrigerated dryer. _____

.5 Tightly close all three drain valves and cooler vent
valve. _____

CAUTION: NEVER OPEN DRAIN, VENT VALVES OR REMOVE
OIL FILLER PLUG OR OIL FILTER COVER WITH
PRESSURE IN THE SYSTEM.

.6 Turn on water supply to aftercooler. _____

.7 Start unit, pump up to unloaded pressure of 110 psig. _____

.8 Inspect for leaks and repair if necessary. _____

2.1.2 Air Flow Controls

This portion of the system interconnects with other subsystems which should be isolated for these tests. Operation of the flow control system requires that the instrument air supply to PT-305 and FIC-304 and the actuation air supply to V-305 be functioning.

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Completed
By: _____

2.1.2.1 Instrument Air Supply

- .1 Clean out air supply piping with 1" line temporarily removed from V-103 and R-100, open V-102 (& V-105, if installed) and blow out line using appropriate safety precautions. _____
- .2 Connect R-100 plumbing to instrument air supply panel and valves V-304 and V-201. Temporarily disconnect line at instrument air filter F-902 and blow out. Reconnect instrument air filter and downstream plumbing. _____
- .3 Disconnect instrument air supply from instruments, blow out and reconnect. _____

2.1.2.2 Fluidization Air Supply

- .1 Install FO-300. Approximate (+5%) flow accuracy required at this time. Unit can be calibrated against CREESI flow nozzle or measured area and assumed discharge coefficient used for purposes of running burner. Make appropriate air flow vs. P, ΔP , T data available to operator. _____
- .2 Temporarily remove V-305, disconnect air line from burner and blow out. Reconnect to V-305 and burner. _____
- .3 Blow out actuation air supply to V-305 and V-103 and connect. _____

2.1.2.3 Leak Check - All air supply lines and components should now be in place and cleaned out. Pressurize to operating pressure and soap test all joints. Propane supply must be made to B-1 burner supply. _____

- .1 Disconnect line from regulators and blow out with filtered nitrogen. _____
- .2 Cap line, pressurize to operating pressure and soap test all joints. Vent and reconnect to regulators. _____
- .3 Clean out burner gas plumbing, connect air or GN₂ supply at 15 psig and test for leaks up to and thru remote operated valves. _____

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2.1.3 Propane Supply

Propane supply must be made to B-1 burner system.

- .1 Disconnect line from regulators and blow out with filtered nitrogen. _____
- .2 Cap line, pressurize to operating pressure and soap test all joints. Vent and reconnect to regulators. _____
- .3 Clean out burner gas plumbing, connect to air or GN₂ supply at 15 psig and test for leaks up to and thru remote operated valves. _____

2.1.4 B-1 Burner Controls

The B-1 burner controls consist of the Protectofier electronics, the gas plumbing assembly and modified Maxon burner assembly. Read the Protectofier manual and the Maxon manual before proceeding. Verify that all components are in place and properly connected.

2.1.4.1 Valve and Controls Operation. A dry run of the burner system should be made to verify that the pilot and main propane valves igniter and flame sensor circuit function correctly. _____

- .1 With gas cock closed, operate Maxon main propane supply valve 15 times. _____
- .2 Temporarily remove burner/6" pipe assembly from windbox and mount so flame can exit freely and safely. _____
- .3 Set propane pressure to approximately 0.5 in. H₂O. Set air flow rate to approximately 95 SCFM total by energizing V-103 and adjusting FIC-304 controller. _____
- .4 Close 1" needle valve (main propane supply) and throttle 1/2" needle valve. _____

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2.1.4.1 (continued)

Completed
By: _____

- .5 Press "burner start" switch. System should delay for purge and then spark should come on pilot. 1/2" solenoid valve should open providing pilot flame. Adjust 1/2" needle valve to give suitable flame (smallest flame that will be quickly and reliably ignited from the spark igniter). _____
- .6 Main Maxon valve should operate. Open 1" needle valve and adjust to give satisfactory main flame. Propane supply pressure may require adjustment and bypass of air thru V-306 may be required. _____
- .7 Operate burner over range of air and propane flows to determine degree of control available. Adjust propane regulator air bleed to give proper flame over operating range. Develop procedure for starting with low flow and gradually increasing propane flow as will be required for reactor heating. _____
- .8 Check operation of flame detector by closing 1/2" needle valve and extinguishing pilot. Main valve should close. _____
- .9 Attempt restart. System should have timed purge before gas valve will open. _____
- .10 Turn system off. With propane pilot and solenoid valves, recheck for leakage thru valves. No leakage is allowed. If propane leak exists it must be corrected before continuing. _____
- .11 Reinstall burner into windbox. _____

2.1.5 Instrumentation

- .1 Pressure gages and flow meters for air and propane. _____
- .2 Windbox temperature. Must have accurate readings in ambient to 200°F range as well as in 200 to 2000°F range. _____
- .3 Bed temperature. Must be accurate in ambient to 200° range; 200 to 1800°F range will be of interest. _____

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2.1.5 (continued)

Completed
By: _____

- .4 Disengagement section. Must be accurate in ambient to 200°F range and to 1500°F range. _____
- .5 Off gas piping will be blanked off for cure cycle. Hot spot thermocouple must be functioning (Reactor lid, on carbon steel). _____
- .6 Carbon steel hot spot thermocouples. Connect to alarm/shutdown to turn off burner if carbon steel exceeds 650°F. _____

2.1.6 Off Gas Piping and Combustor

- .1 This portion of system blanked off for reactor cure cycle. _____

2.1.7 Special Controls

To permit unattended operation during the steady state portions of the cycle, the air controls must be temporarily modified so that the system can control on an appropriate set temperature. This will be accomplished by controlling burner bypass air from the temperature of interest and increasing bypass air flow as temperature increases. This will reduce burner output and heating rate. In addition, the bed temperature alarm must be wired to shutdown the burner in the event that the desired set temperature is exceeded.

3.0 REFRACTORY CURE

3.1 Set Up

- 3.1.1 Verify that preceding procedures have been carried out satisfactorily. _____
- 3.1.2 Schedule HITS site work and crew to minimize interference and provide 24 hour manning for the 14 day cure cycle. Schedule with balance of Engineering Laboratory operation. _____
- 3.1.3 Remove plugs, caps and lines from ports exiting from reactor; blank line to product gas combustor. On large lines (char removal, spare top gas ports) put blind flange in place on one bolt so that the flow out of the port can be throttled. _____

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3.1.4 Connect fire box to reactor without installing tuyere plate. _____

3.1.5 Install personnel thermal protection around temporary vent flows and around all components which will become hot in operation. _____

3.2- Cure Cycle

Operate burner and bypass air to provide temperature-time cycle as shown in Figure C-2.

CAUTION: AVOID TEMPERATURE OVERSHOTS ABOVE SPECIFIED VALUES ESPECIALLY IN BRINGING EACH REFRACTORY SECTION (WINDBOX, BED DISENGAGING SECTION AND PRODUCT GAS COMBUSTOR) UP TO DRYNESS AT 200°F.

Each section of refractory must be held at 200°F for the specified time and reach equilibrium dryness. If equilibrium is not reached in the time indicated, the holding time period must be extended.

3.3 Post Cure

At the end of the 14 day cure cycle the refractory is ready for use or cool down.

3.3.1 System Operation - Refer to HITS Operating Procedures.

3.3.2 Cool Down.

Extinguish propane burner, turn off propane supply and continue air flow. At present there is no limit on cool down rate.

3.3.3 Restart

There is no restriction on heatup rate after the initial cure unless the unit has been idle for an extended period of time, or water is present in the unit. In this event, it must be taken again through the 200°F cycle before further heatup.

C.4.B HITS COMBUSTOR AND STACK REFRACTORY CURE PROCEDURE

1.0 INTRODUCTION

Satisfactory operation of the refractory in the HITS product gas combustor and stack requires controlled heating to gradually remove moisture and to bring the refractories up to operating temperature for proper cure. Figure C-3 is the temperature-time history recommended for proper cure of the combustor refractory. Moderate times at temperature are required because of the slow diffusion rate of moisture through the thick refractory. The reactor combustor initial cure uses the E-1 electrical heater as a heat source. Intermediate cure (between 1000 and 1600°F) will be accomplished using the B-1 propane burner, as in Procedure -001. Final cure (up to operating temperature) will be done during system tests.

CAUTION: PERMANENT DAMAGE WILL OCCUR TO THE REFRACTORY LINER IF THE HEATING SCHEDULE IS NOT FOLLOWED. IN PARTICULAR THE 200°F LIMIT MUST NOT BE EXCEEDED UNTIL ALL FREE MOISTURE HAS BEEN REMOVED. HIGHER HEATING RATES OR SHORTER TIMES AT TEMPERATURE ARE DAMAGING.

Times shown in Figure C-3 are minimums judged to be required to reach equilibrium dryness at each temperature level. These times are to be verified by monitoring the entrance and exit gas humidity. When exit moisture content reaches inlet moisture content, satisfactory dryness has been achieved for that temperature level.

2.0 INITIAL SYSTEM CHECKOUT

To conduct the refractory cure procedure requires that the following subsystems be checked out and operational.

2.1 Combustor Air Supply

This subsystem consists of the air source and flow controls. The air source is the C-2 roots blower and its controls. The primary flow controls are the flow control station consisting of flow orifice FO-400, flow control valve V-401, its transmitter PT-402, the indicating flow controller FIC-400 and the instrument air supply.

Figure C-3 Combustor Refractory Cure Cycle

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2.1.1 Blower

This unit is a Whispair-Max 3514J Roots Blower. The unit has been used on previous programs. The manual for this unit, Bulletin IRB-131-270, should be read before its operation.

CAUTION: THE BLOWER IS A POSITIVE DISPLACEMENT DEVICE. NOMINAL FLOW CONTROL MUST BE OBTAINED BY BYPASSING AIR FLOW, NOT BY THROTTLING THE OUTLET OR INLET. IN NORMAL OPERATION, THE BYPASS FLOW WILL BE TO THE STACK EDUCTOR. HOWEVER, FOR THE REFRACTORY CURE TESTS, THE BYPASS AT THE COMPRESSOR SHOULD BE USED SO THAT THE REFRACTORY IN THE STACK WILL BE HEATED, DRIED AND CURED.

NOTE: The maximum outlet pressure rise allowed for this unit is 4 psi; maximum temperature rise is 90°F. If the pressure rise is exceeded, temperature rise will exceed 90°F and the blower will overload due to rotor binding caused by thermal expansion. Potential for damage or loss of blower air due to thermal motor shutdown exists because of the way in which flow is controlled to the combustor. If V-401 were to throttle down sufficiently to cause PA-400 to rise above 5-6 psig, rotor binding and blower shutdown will occur. Therefore, for normal operation, the bypass should be opened so that PI-405 reads about 3 psig, to give some control range for V-401. The ideal way to do this is to set PI-405 to just under 4 psig by opening the bypass with V-401 closed. Then opening V-401 will reduce blower outlet pressure and shutting V-401 cannot damage the blower.

Completed
By: _____

2.1.1.1 Blower Lubrication

- .1 Oil level is checked by removing the two oil fill plugs on the side of the unit. With a flexible spout can, fill through both holes until oil runs out each hole. DO NOT CHECK OR FILL WHILE RUNNING. Use a SAE 40 oil below 90°F and SAE 50 above 90°F ambient. Oil is changed every 2000 operating hours.

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2.1.1.1 (continued)	Completed By: _____
.2 Shaft bearings must be lubricated every 700 hours of operation.	_____
2.1.2 Air Flow Controls	
2.1.2.1 Instrument Air Supply The instrument air supply has been checked out as part of Procedure -001.	_____
2.1.2.2 Combustor Air Supply .1 Install FO-400 approximate (<u>±</u> 5%) flow accuracy required at this time. .2 Disconnect, clean out air supply lines and reconnect. .3 Blank off air line at entrance to combustor and leak check.	_____ _____ _____
2.1.2.3 Air Heater .1 Verify proper mechanical and electrical hook up. .2 Set limits on air and element controllers.	_____ _____
2.1.2.4 Combustor .1 Blank off connection to off-gas lines from cyclone. .2 Connect monitoring thermocouples to can hot spot, outline 12" pipe, and install thermocouple in combustor and on refractory in stack. .3 Leave 1" ports on combustor open for bleed of hot gas.	_____ _____ _____
2.1.2.5 Special Controls Temperature alarm TIA-805 should be activated and monitor combustor temperature. It can be wired in series with TA-402, so that heater operation is temporarily stopped while combustor temperature exceeds the desired value.	_____
2.1.2.6 Instrumentation In addition to the instrumentation activated for reactor cure, the thermocouples listed in 2.1.2.4 above should be monitored.	_____

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3.0 REFRACTORY CURE

3.1 Set Up

- 3.1.1 Verify that the preceding procedures have been carried out satisfactorily _____
- 3.1.2 Schedule HITS site work, crew and laboratory for the 3-1/2 day cure cycle. If possible, the combustor cure should start at or before day eight of the reactor cure cycle. _____
- 3.1.3 Install personnel thermal protection around temporary vent flows and around all components which will become hot in operation. _____

3.2 Cure Cycle

- 3.2.1 Turn on roots blower and adjust air flow. _____
- 3.2.2 Set heater element limit to 1200°F. _____
- 3.2.3 Set heater air outlet limit to 90°F above ambient. _____
- 3.2.4 Energize heater. _____
- 3.2.5 Operate heater to give temperature-time cycle as shown in Figure C-3 by slowly increasing JC-400 set point. Verify operation of safety shutdowns by lowering JC-400 set points; heater should turn off and resume heating when temperature falls. Lower JC-401 set point; heater should turn off and remain off until reset is pushed. _____
- 3.2.6 To phase in with reactor cure, when exit temperature from reactor is about 1200°F, temporarily turn off B-1 and fluidization air, and remove blocking plate from off-gas line to combustor. Turn fluidization air and B-1 burner back on, gradually block off top outlet of cross at top of reactor to slowly divert hot gas to combustor over a period of about 1 hour to minimize thermal stresses on piping. Continue cure cycle using hot gas from reactor. _____

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C.5 ADDITIONAL INFORMATION SOURCES ON HITS

1.	HITS Work Plan - Draft Copy #2	18 Dec 1980
2.	HITS Yes-No Grid	31 Jan 1980
3.	HITS Design Review Package	1 Feb 1980
4.	HITS Test Plan	15 Feb 1980
5.	Response to (Environmental) Questions on HITS Coal Gasification Facility, HITS Memo #025, DMJ to PMT	24 Apr 1980
6.	Topical Report: Test System Description	30 Jun 1980
7.	Component and System <u>Drawings</u> (List of Drawings in HITS Control Room Drawing File)	
8.	<u>Equipment Manuals</u> (Vols I, II & III in HITS Control Room)	
9.	<u>HITS Checkout, Calibration and Operating Procedures</u> , (Vol I, in HITS Control Room)	
10.	HITS <u>Component List</u> (HITS Control Room)	
11.	HITS Program Task <u>Schedule</u>	19 Jun 1980