Industrial Fuel Gas Demonstration Plant Program

Prepared For
The Department of Energy
Under Contract ET-77-C-01-2582

MEMPHIS LIGHT, GAS AND WATER DIVISION
P.O. BOX 430, MEMPHIS, TENNESSEE 38145

In Association with
FOSTER WHEELER ENERGY CORPORATION
INSTITUTE OF GAS TECHNOLOGY
DELTA REFINING COMPANY

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1078-625

Industrial Fuel Gas Demonstration Plant Program

MONTHLY AND QUARTERLY PROGRESS REPORT

MASTER

FOR THE PERIOD
1 APRIL 1978 - 30 JUNE 1978

DISCLAIMER

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MEMPHIS INDUSTRIAL FUEL GAS

DEMONSTRATION PLANT PROJECT

QUARTERLY TECHNICAL PROGRESS REPORT
(Deliverable #12)

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MEMPHIS LIGHT, GAS & WATER DIVISION
P.O. Box 430
MEMPHIS, TENNESSEE 38101

PREPARED FOR THE
U. S. DEPARTMENT OF ENERGY--
ASSISTANT SECRETARY OF FOSSIL ENERGY

UNDER CONTRACT DE-AC02-77ET13406
(FORMERLY CONTRACT ET-77-C-01-2582)

MLGW/DOE INDUSTRIAL FUEL GAS
DEMONSTRATION PLANT PROGRAM

FOSTER WHEELER

MTSR-8
July 1978

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**MLGW/DOE INDUSTRIAL FUEL GAS
DEMONSTRATION PLANT PROGRAM**

Report No. MTSR-8 Technical Status Report
Period of April 1, 1978 through June 30, 1978

F FOSTER WHEELER

Contract ET-77-C-01-2582

1.0 Contract Objective

The objective of the Industrial Fuel Gas Demonstration Plant Program is to demonstrate the feasibility of converting coal to fuel gas and utilizing this gas in a commercial application. Specific objectives of this program are to conduct process analysis, design, construction, testing, operation, and evaluation of a plant based on the U-Gas process for converting coal to fuel gas.

2.0 Contract Tasks

2.1 Task I - Design and Evaluation of Commercial Plant

During the reporting period, the design of the Commercial Plant was completed and cost estimate prepared. As reported in earlier reports, most work remaining on Task I relates to the following:

1. Demonstration Plant recommendations, including a configuration study, at DOE's request, to evaluate alternatives for the Demo Plant configuration to achieve the required reliability with a coal based solution. The specific alternatives being considered are:
 - a. A base U-Gas Plant designed to produce 50 MMM Btu/day 365 days per year (the "High Reliability" case).
 - b. A base U-Gas Plant plus product storage.
 - c. A base U-Gas Plant, plus a small methanation facility to upgrade a small portion of the product to obtain "credits."
2. Other studies in support of the Demonstration Plant recommendation these studies include:
 - a. Economic study of U-Gas versus coal fired boilers for internal steam generation.
 - b. An assessment of the cost/benefit of designing the Demo Plant for an operating pressure higher than the current data allows.
 - c. Evaluate the costs of providing a "zero discharge" wastewater treatment system.

MLGW/DOE INDUSTRIAL FUEL GAS
DEMONSTRATION PLANT PROGRAM

FOSTER WHEELER

Contract ET-77-C-01-2582

- d. Conduct preliminary investigation of different methanation processes and the impact on the process if such a configuration is selected.

3. Risk Analysis and Economics

2.2 Task II - Demonstration Plant Process Design

The Process Design of the Demonstration Plant has not been able to proceed as planned during the reporting period because of a lack of a basic Plant Configuration, which is being studied as part of the Demo Plant recommendations.

All work on this task has ceased.

NOTE: At the Monthly Review Meeting held on June 23, 1978, DOE presented MLGW with a STOP WORK order, effective at that time, for the following tasks:

1. Task II, III, VI, VII, VIII, X
2. Task IV except soil activities needed for the environmental analysis.

2.3 Task III - Demonstration Plant Process and Mechanical Design Package

This task was scheduled to begin July 1, 1978, but is currently affected by the STOP WORK order.

2.4 Task IV - Site Evaluation and Selection

MLGW formally selected the Allen Site on May 2, 1978. The site evaluation report was being revised with the latest information, to be reissued in July. However, all work on this report has ceased.

Negotiations have begun with the TVA, who holds an easement for ash disposal on the site, for the release from that easement. This work has also ceased. Survey and soil work are proceeding. Quotations from prospective bidders will be solicited during July.

2.5 Task V - Demonstration Plant Environmental Analysis

The status of the environmental effort is as follows:

Plant Systems Description

FWEC provided EIA with Task I plant design inputs and a schedule for providing input to the ER. EIA is reviewing this schedule in light of the current project slowdown.

A review of the plant design information supplied to EIA by FWEC is also in progress. These materials include:
a) Volumes I and II of "The Basis of Estimate for Commercial Plant Operation," b) a list of estimated catalyst and chemical requirements, and c) new drafts of the simplified plant block-flow diagram, Section 2.0, "Nominal Plant Material and Energy Balance" and the Table of Contents for the Task I Report.

Water Quality and Hydrology

The data from the first season of water quality work has been evaluated. Work will proceed on the surface water portions of the hydrology task but the groundwater portions is deferred until a meeting with EIA and ORNL could be arranged.

Air Quality and Meteorology

EMP&E, the local subcontractor to EIA, voluntarily terminated its involvement in the program as of June 15 and EIA took over the operation of the field program. EIA hired a local technician. Operation of the air van has continued while long-term equipment is manufactured, shipped and tested.

Plant engineering and emission inputs (from FWEC), Allen Station emission data (from MLGW) and climatological information have been gathered and are being evaluated to identify the location of the instrument site for long-term monitoring.

Plume rise calculations for the fuel gas plant stacks, were completed and the identification of a location for the long-term monitoring site is proceeding on schedule.

Terrestrial Ecology

A work plan has been established with Memphis State University.

Aquatic Ecology

Ichthyoplankton samples were collected for the last times during the weeks of June 5 and June 12 in the vicinity of the Allen site. This concluded the ichthyoplankton studies. Arrangements for conducting the summer field programs were pursued with representatives of MSU.

Socioeconomics

Activities were limited to reviewing scope with the Shelby County Planning Commission.

Noise

RAMCON, a subcontractor to EIA, began researching the existing noise regulations and examining the site vicinity for noise sensitive locations. A listing of the applicable noise regulations was received.

Environmental Report Production

A review is underway to assess the impact of the project slowdown on the schedule for preparation of the ER. EIA will be identifying critical input items.

2.6 Task VI - Materials, Agreements and Licenses

Work continued in May on the marketing plan for the disposition of products from the plant. All work on this task has ceased.

2.7 Task VII - Planning for Final Design Construction and Operation Planning

This task was scheduled to begin August 1, 1978.

2.8 Task VIII - Economic Assessment

This task was scheduled to begin February 1, 1979.

2.9 Task IX - Technical Support

FWEC is preparing a design basis for the gasification section of the Demonstration Plant, in order to determine where significant data gaps exist, and to suggest appropriate Pilot Plant work to obtain the required data.

Further discussion of Task IX is found in the Appendix.

2.10 Task X - Long Lead Time Items

Work on this activity has stopped.

2.11 Task XI - Program Management

Management of the program has been transferred to the DOE Chicago Operations office as of June 19, 1978.

3.0 Project Status

3.1 Schedule

The Phase I schedule has been affected by the STOP WORK order. Task I work is being completed. The final report is anticipated to be issued to DOE in early to mid-August.

Task V is on schedule, but will be affected as time progresses, as the STOP WORK order continues.

3.2 Cost Status

The Cost Status remains the same as reported on the last Monthly Report and at the May 25, 1978 Monthly Briefing.

3.3 Technical Status

A complete review of the Pilot Plant design and equipment was conducted during April and May. The modifications scheduled for that period were completed. A complete review of this work was conducted on May 5, 1978, with DOE participation.

As of the end of May, it was felt that Pilot Plant confirmation of Design data will not be possible to allow a significant portion of total engineering effort to proceed and that a further evaluation should be conducted in mid-July.

3.4 Contractual Status

No significant contractual problems exist.

3.5 Management Status

A meeting at DOE was held in early May to help resolve the problems associated with DOE's decision to not allow Monsanto to enter into a secrecy agreement with MLGW for their monitoring work. This meeting resulted in DOE's consent for MRC to execute an agreement. A draft was prepared, reviewed by FWEC, and resubmitted to MRC on May 24, 1978. As of the end of June, agreement appears likely.

DOE's stop work order (See 2.2) was received on June 23, 1978.

4.0 Technical Approach or Work Plan Changes

No such changes were undertaken during the period except for the stopping of certain work.

5.0 Program Manager's Assessment

An evaluation of the Pilot Plant is scheduled for July 18, 1978 to review recent activity and to determine further action. Both DOE and the industrial team will attend.

U.S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION
CONTRACT MANAGEMENT SUMMARY REPORT

1. Contract Identification Industrial Fuel Gas Demonstration Plant		2. Reporting Period 6/1/78 through 6/30/78	3. Contract Number W-77-C-01-2582
4. Contractor (name and address) Memphis Light Gas and Water Division P. O. Box 430 Memphis, Tennessee 38101		5. Contract Start Date September 19, 1977*	
		6. Contract Completion Date May 19, 1979	

7. Months	O	N	D	J	F	M	A	M	J	J	A	S	8. FY 1978
-----------	---	---	---	---	---	---	---	---	---	---	---	---	------------

9. Cost Status															
a. Dollars (Thousands)	7,000												e. Actual Costs Prior FYs		
	6,000												2		
	5,000												f. Planned Costs Prior FYs		
	4,000												2		
	3,000												g. Total Estimated Accrued Costs for Contract		
	2,000												7,118		
	1,000												h. Total Contract Value		
Accrued Costs		b. Planned	7	54	112	362	421	476	511	560	535	350	363	373	6,460
		c. Actual	7	44	79	209	290	452	371	792	503				Phase I
		d. Variance		(10)	(33)	(153)	(131)	(24)	(140)	232	(32)				

10. Manpower Status															
a. Manhours (Hundreds)	1,200												e. Actual Manpower Prior FYs		
	1,000												2		
	800												f. Planned Manpower Prior FYs		
	600												2		
	400												g. Total Estimated Manpower for Contract		
	200												2201		
Manpower		b. Planned	3	18	37	108	121	125	142	119	116	114	114	115	h. Total Contract Manpower (hours)
		c. Actual	3	21	25	68	86	135	123	151	131				202,900
		d. Variance		3	(12)	(40)	(35)	10	(19)	2	(15)				

11. Major Milestone Status		
a.	TASK I	JULY 15, 1978
b.	TASK II	AUG. 25, 1978
c.	TASK III	MID MAY, 1979
d.	TASK IV	AUG. 25, 1978
e.	TASK V	MAR. 2, 1979
f.	TASK VI	MID MAY, 1979
g.	TASK VII	MAY 4, 1979
h.	TASK VIII	MAR. 9, 1979
i.	TASK IX	MAR. 9, 1979
j.	TASK X	END DEC. 1978
k.	TASK XI	END MAY 1979

12. Remarks	
*Subcontractor start dates October 24, 1977 and November 11, 1977.	

13. Signature of Contractor's Project Manager and Date <i>R. W. Gray / RJS 7-17-78</i>	14. Signature of Government Technical Representative and Date
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1. Contract Identification Industrial Low Btu Fuel Gas Demonstration Plant Program						2. Reporting Period 6/1/78 through 6/30/78						3. Contract Number ET-77-C-01-2582																	
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7. Identification Number		8. Reporting Category (e.g., contract line item or work breakdown structure element)		9. Fiscal Years and Months												10. Percent Complete													
				F. Y. 1978												F. Y. 1979													
Task				O	N	D	J	F	M	A	M	J	J	A	S	O	N	D	J	F	M	A	M	J	J	A			
I	Comm Plt Des and Eval			██████████												REPORT ▽												93%	
I-1	Process Eng Des			██████████																								96%	
I-2	Risk Analysis			██████████																								50%	
I-3	Proj Eng Des			██████████																								84%	
I-4	Cost Est			██████████																								95%	
I-5	Econ/Env. Impact			██████████																								20%	
I-6	Demo Plt Proc Recomm.			██████████												L ▽												20%	
II	Demo Plt Proc Des			██████████												██████████													
II-1	Basis of Design			██████████												A B C D E F G H I J K L M N O P Q R S T U V W X Y Z AA AB AC AD AE AF AG AH AI AJ AK AL AM AN AO AP AQ AR AS AT AU AV AW AX AY AZ BA BB BC BD BE BF BG BH BI BJ BK BL BM BN BO BP BQ BR BS BT BU BV BW BX BY BZ CA CB CC CD CE CF CG CH CI CJ CK CL CM CN CO CP CQ CR CS CT CU CV CW CX CY CZ DA DB DC DD DE DF DG DH DI DJ DK DL DM DN DO DP DQ DR DS DT DU DV DW DX DY DZ EA EB EC ED EE EF EG EH EI EJ EK EL EM EN EO EP EQ ER ES ET EU EV EW EX EY EZ FA FB FC FD FE FF FG FH FI FJ FK FL FM FN FO FP FQ FR FS FT FU FV FW FX FY FZ GA GB GC GD GE GF GG GH GI GJ GK GL GM GN GO GP GQ GR GS GT GU GV GW GX GY GZ HA HB HC HD HE HF HG HH HI HJ HK HL HM HN HO HP HQ HR HS HT HU HV HW HX HY HZ IA IB IC ID IE IF IG IH II IJ IK IL IM IN IO IP IQ IR IS IT IU IV IW IX IY IZ JA JB JC JD JE JF JG JH JI JJ JK JL JM JN JO JP JQ JR JS JT JU JV JW JX JY JZ KA KB KC KD KE KF KG KH KI KJ KK KL KM KN KO KP KQ KR KS KT KU KV KW KX KY KZ LA LB LC LD LE LF LG LH LI LJ LK LM LN LO LP LQ LR LS LT LU LV LW LX LY LZ MA MB MC MD ME MF MG MH MI MJ MK ML MN MO MP MQ MR MS MT MU MV MW MX MY MZ NA NB NC ND NE NF NG NH NI NJ NK NL NM NO NP NQ NR NS NT NU NV NW NX NY NZ OA OB OC OD OE OF OG OH OI OJ OK OL OM ON OO OP OQ OR OS OT OU OV OW OX OY OZ PA PB PC PD PE PF PG PH PI PJ PK PL PM PN PO PP PQ PR PS PT PU PV PW PX PY PZ QA QB QC QD QE QF QG QH QI QJ QK QL QM QN QO QP QQ QR QS QT QU QV QW QX QY QZ RA RB RC RD RE RF RG RH RI RJ RK RL RM RN RO RP RQ RR RS RT RU RV RW RX RY RZ SA SB SC SD SE SF SG SH SI SJ SK SL SM SN SO SP SQ SR SS ST SU SV SW SX SY SZ TA TB TC TD TE TF TG TH TI TJ TK TL TM TN TO TP TQ TR TS TT TU TV TW TX TY TZ UA UB UC UD UE UF UG UH UI UJ UK UL UM UN UO UP UQ UR US UT UU UV UW UX UY UZ VA VB VC VD VE VF VG VH VI VJ VK VL VM VN VO VP VQ VR VS VT VU VV VW VX VY VZ WA WB WC WD WE WF WG WH WI WJ WK WL WM WN WO WP WQ WR WS WT WU WV WW WX WY WZ XA XB XC XD XE XF XG XH XI XJ XK XL XM XN XO XP XQ XR XS XT XU XV XW XX XY XZ YA YB YC YD YE YF YG YH YI YJ YK YL YM YN YO YP YQ YR YS YT YU YV YW YX YZ ZA ZB ZC ZD ZE ZF ZG ZH ZI ZJ ZK ZL ZM ZN ZO ZP ZQ ZR ZS ZT ZU ZV ZW ZX ZY ZZ													
11. Remarks A Stop Work Order was received 6/23/78 for Tasks II, III, Part of IV, VI, VII, VIII and X.																													
12. Signature of Contractor's Project Manager and Date RW Gray / R 7-17-78														13. Signature of Government Technical Representative and Date															

U. S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION
MILESTONE PLAN AND MANAGEMENT REPORT

1. Contract Identification Industrial Low Btu Fuel Gas Demonstration Plant Program		2. Reporting Period 6/1/78 through 6/30/78		3. Contract Number ET-77-C-01-2582																						
4. Contractor (name, address) Memphis Light, Gas and Water Division P.O. Box 430 Memphis, Tennessee 38101		5. Contract Start Date September 19, 1977		6. Contract Completion Date May 19, 1979																						
7. Identification Number	8. Reporting Category (e.g., contract line item or work breakdown structure element)	9. Fiscal Years and Months												10. Percent Complete												
		F. Y. 1978												F. Y. 1979												
Task		O	N	D	J	F	M	A	M	J	J	A	S	O	N	D	J	F	M	A	M	J	J	A		
III	Demo Plt Process and Mech Design Pkg																									
III-1	Prelim Eng and Layouts																									
III-2	Plant Specifications																									
III-3	Cost Est																									
III-4	Proc. Eng Rpt																									
III-5	Failure Analysis																									
III-6	Procurement Req																									
IV	Site Eval/Selection																									
IV-1	Site Data																									
IV-2	Site Agreement																									
11. Remarks																										
12. Signature of Contractor's Project Manager and Date <i>R W Gray / RJS</i> 7-17-78													13. Signature of Government Technical Representative and Date													

U. S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION
MILESTONE PLAN AND MANAGEMENT REPORTFORM APPROVED
OMB NO. 38R-
Page 3 of 6

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		F. Y. 1978										F. Y. 1979														
Task		O	N	D	J	F	M	A	M	J	J	A	S	O	N	D	J	F	M	A	M	J	J	A		
V	Env. Analysis																									
V-1	Basis for Env. Analysis																									
V-2	Proj. Desc. for Env. Anal.																									
V-3	Demo Plt Environment																									
V-4	Proj. Opn.																									
V-5	Mitigation Measures																									
V-6	Adverse Env. Effects																									
V-7	Long Term Aspects																									
V-8	Irreversible Commitments																									
V-9	Alternatives																									
V-10	Env. Report																									
11. Remarks																										
12. Signature of Contractor's Project Manager and Date R W Gray / RJS 7-17-78													13. Signature of Government Technical Representative and Date													

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U. S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION
MILESTONE PLAN AND MANAGEMENT REPORT

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7. Identification Number Task	8. Reporting Category (e.g., contract line item or work breakdown structure element)	9. Fiscal Years and Months F. Y. 1978											F. Y. 1979												10. Percent Complete
		O	N	D	J	F	M	A	M	J	J	A	S	O	N	D	J	F	M	A	M	J	J	A	
VII	Const/OPN Planning																								
VII-1	Conf. Mgt Plan																								
VII-2	Staffing Plan																								
VII-3	Safety																								
VII-4	Procurement Plan																								
VII-5	Sched/Controls																								
VII-6	Labor Surveys																								
VII-7	Summary Plans																								
VIII	Economic Assessment																								
VIII-1	Commercial PIT																								
VIII-2	Demo PIT																								

11. Remarks	
12. Signature of Contractor's Project Manager and Date RW Gray / R. 7-17-78	
13. Signature of Government Technical Representative and Date	

Page _____ of _____

[illegible]

COST MANAGEMENT REPORT

1. Contract Identification Industrial Low BTU Fuel Gas Demonstration Plant Program					2. Reporting Period <u>6/1/78</u> through <u>6/30/78</u>			3. Contract Number EF-77-C-01-2502					
4. Contractor Name and Address Memphis Light, Gas and Water Division P.O. Box 430 Memphis, Tennessee 38101					5. Cost Plan Date December 20, 1977			6. Contract Start Date September 19, 1977					
								7. Contract Completion Date May 19, 1979					
8. Government Funding \$745,000		9. Contractor Funding 0		10. Number of Invoices Billed 5		11. Frequency Monthly		12. Number of Invoices Paid 2		13. Total Invoice Amounts Billed \$87,609.24		14. Total Payment Received \$19,059.48*	
15. Identification Number	16. Reporting Category (e.g., contract line item or work breakdown structure element)	17. Costs Accrued				18. Projected Accrued Costs			19. Total Contract Value	20. Variance			
		During Reporting Period		Cumulative to Date		a. Subsequent Reporting Period	b. Balance of Fiscal Year	c. Total Contract					
Task		a. Actual	b. Planned	c. Actual	d. Planned								
I	Comm. Pkt Des. and Eval	7	4	7	24	5	0	15	28	(13)			
II	Demo Pkt Process Des	0	4	1	20	0	9	16	24	(8)			
III	Mech. Design Pkg	0	8	1	8	0	20	63	96	(33)			
IV	Site Eval and Selection	0	8	25	80	4	43	108	98	10			
V	Env. Analysis	2	2	4	7	2	4	17	16	1			
VI	Materials and Licenses	0	4	0	14	0	5	36	58	(22)			
VII	Const and OPN Planning						0	9	13	(4)			
VIII	Economic Assessment												
IX	Tech. Support												
X	Long Lead Items	0	1	0	1	0	5	23	25	(2)			
XI	Project Mgt	21	27	133	148	16	65	458	387	71			
21. Total		30	58	171	302	27	151	745	745	0			
22. Remarks *MIGW through 6/30/78									23. Dollars Expressed In: Thousands				
24. Signature of Contractor's Project Manager and Date R W Gran 1/8/ 7-17-68				25. Signature of Contractor's Authorized Financial Representative and Date				26. Signature of Government Technical Representative and Date					

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COST MANAGEMENT REPORT

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U.S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION
COST MANAGEMENT REPORTFORM APPROVED
OMB NO. 38R.

Page _____ of _____

1. Contract Identification Industrial Low BTU Fuel Gas Demonstration Plant Program					2. Reporting Period <u>6/1/78</u> through <u>6/30/78</u>		3. Contract Number EF-77-C-01-2582			
4. Contractor Name and Address Memphis Light, Gas and Water Division P.O. Box 430 Memphis, Tennessee 38101					5. Cost Plan Date December 20, 1977		6. Contract Start Date October 24, 1977			
							7. Contract Completion Date May 19, 1979			
8. Government Funding \$2,825,000		9. Contractor Funding		10. Number of Invoices Billed 10	11. Frequency Monthly	12. Number of Invoices Paid 6	13. Total Invoice Amounts Billed \$1,136,280.41		14. Total Payment Received \$416,258.85	
15. Identification Number	16. Reporting Category (e.g., contract line item or work breakdown structure element)	17. Costs Accrued				18. Projected Accrued Costs			19. Total Contract Value	20. Variance
		During Reporting Period		Cumulative to Date		a. Subsequent Reporting Period	b. Balance of Fiscal Year	c. Total Contract		
Task		a. Actual	b. Planned	c. Actual	d. Planned					
I	Comm. Plt Des. and Eval	84	14	530	349	52	0	525	349	176
II	Demo Plt Process Des	9	57	24	199	0	10	241	241	
III	Mech. Design Pkg	0	57	0	81	0	10	1,282	1,282	
IV	Site Eval and Selection	2	6	7	43	0	16	51	51	
V	Env. Analysis	3	1	16	32	3	6	32	32	
VI	Materials and Licenses	0	2	1	10	0	4	27	27	
VII	Const and OPN Planning					0	6	82	82	
VIII	Economic Assessment					0	6	28	28	
IX	Tech. Support	13	6	91	17	5	10	141	141	100
X	Long Lead Items	0	25	0	65	0	19	182	182	
XI	Project Mgt	41	32	287	250	28	59	510	510	
21. Total		152	200	956	1,046	88	146	3,101	2,825	276
22. Remarks FWEC Invoice figures include EIA charges.									23. Dollars Expressed In: Thousands	
24. Signature of Contractor's Project Manager and Date R W Gray / Rpl. 7-17-78				25. Signature of Contractor's Authorized Financial Representative and Date			26. Signature of Government Technical Representative and Date			

U.S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION
COST MANAGEMENT REPORT

1. Contract Identification Industrial Low BTU Fuel Gas Demonstration Plant Program					2. Reporting Period 6/1/78 through 6/30/78		3. Contract Number EF-77-C-01-2582		
4. Contractor (name and address) Memphis Light, Gas and Water Division P.O. Box 430 Memphis, Tennessee 38101					5. Cost Plan Date December 20, 1977		6. Contract Start Date January 31, 1978		
							7. Contract Completion Date May 19, 1979		
8. Government Funding		9. Contractor Funding		10. Number of Invoices Billed 3		11. Frequency Monthly		12. Number of Invoices Paid 1	
								13. Total Invoice Amounts Billed \$179,765.08	
								14. Total Payment Received \$179,765.08	
15. Identification Number	16. Reporting Category (e.g., contract line item or work breakdown structure element)	17. Costs Accrued		18. Projected Accrued Costs		19. Total Contract Value	20. Variance		
		During Reporting Period		Cumulative to Date					
Task		a. Actual	b. Planned	c. Actual	d. Planned	e. Subsequent Reporting Period	f. Balance of Fiscal Year	g. Total Contract	
I	Comm. Plt Des. and Eval								
II	Demo Plt Process Des								
III	Mech. Design Pkg								
IV	Site Eval and Selection								
V	Env. Analysis	109	24	216	48	75	150	750	350
VI	Materials and Licenses								
VII	Const and OPN Planning								
VIII	Economic Assessment								
IX	Tech. Support								
X	Long Lead Items								
XI	Project Mgt								
21. Total		109	24	216	48	75	150	750	350
22. Remarks		*The variance shown is an estimate only. A more detailed estimated variance will be determined in June after finalization of work scope based on new laws and regulations. After determining the final work scope, EIA/FWEC will detail the changes and costs in a request for contract modification.							23. Dollars Expressed In: Thousands
24. Signature of Contractor's Project Manager and Date R W Gray 1 Rgt. 7-77-78				25. Signature of Contractor's Authorized Financial Representative and Date				26. Signature of Government Technical Representative and Date	

MANPOWER MANAGEMENT REPORT

1. Contract Identification Industrial Low BTU Fuel Gas Demonstration Plant				2. Reporting Period 6/1/78 through 6/30/78				3. Contract Number EF-77-C-01-2582						
4. Contractor (name and address) Memphis Light, Gas and Water Division P.O. Box 430 Memphis, Tennessee 38101				5. Manpower Plan Date December 20, 1977				6. Contract Start Date September 19, 1977						
								7. Contract Completion Date May 19, 1979						
8.		9.		10.		11. Monthly		12.		13.		14.		
15. Identification Number		16. Reporting Category (e.g., contract line item or work breakdown structure element)		17. Manpower Expended				18. Projected Manpower Expenditures			19. Total Contract Manpower		20. Variance	
Task				During Reporting Period		Cumulative to Date		a. Subsequent Reporting Period	b. Balance of Fiscal Year	c. Total Contract				
				a. Actual	b. Planned	c. Actual	d. Planned							
I		Comm. Plt Des. and Eval		37	8	196	150	31	2	211	153	58		
II		Demo Plt Process Des		8	27	114	96	0	97	126	130	(14)		
III		Mech. Design Pkg		0	25	1	33	0	36	531	545	(14)		
IV		Site Eval and Selection		0	4	10	35	1	13	38	44	2		
V		Env. Analysis		30	7	61	33	21	42	207	110	97		
VI		Materials and Licenses		1	6	3	25	0	7	40	64	(24)		
VII		Const and OPN Planning		0	0	1	0	0	7	49	46	3		
VIII		Economic Assessment							8	21	21			
IX		Tech. Support		19	36	256	254	22	20	326	310	16		
X		Long Lead Items		0	8	0	22	0	7	76	78	(2)		
XI		Project Mgt		36	29	201	211	28	85	572	528	44		
21. Total				131	150	743	859	103	324	2,205	2,029	176		
22. Remarks Program total for Contractor and Subcontractor												23. Manpower Expressed In: 100 Man-hours		
24. Signature of Contractor's Project Manager and Date R W Gray/RLH 7-17-78							25. Signature of Government Technical Representative and Date							

Page ____ of ____

1. Contract Identification Industrial Low BTU Fuel Gas Demonstration Plant						2. Reporting Period 6/1/78 through 6/30/78			3. Contract Number EF-77-C-01-2582						
4. Contractor (name and address) Memphis Light, Gas and Water Division P.O. Box 430 Memphis, Tennessee 38101						5. Manpower Plan Date December 20, 1977			6. Contract Start Date September 19, 1977						
									7. Contract Completion Date May 19, 1979						
8.		9.		10.		11. Monthly		12.		13.					
15. Identification Number		16. Reporting Category (e.g., contract line item or work breakdown structure element)		17. Manpower Expended				18. Projected Manpower Expenditures			19. Total Contract Manpower		20. Variance		
Task				During Reporting Period		Cumulative to Date		a. Subsequent Reporting Period		b. Balance of Fiscal Year		c. Total Contract			
				a. Actual b. Planned		c. Actual d. Planned									
I Comm. Plt Des. and Eval				4 3		4 16		5 0		11		19		(8)	
II Demo Plt Process Des				0 3		1 13		0 8		12		16		(4)	
III Mech. Design Pkg				0 6		1 6		0 16		-52		66		(14)	
IV Site Eval and Selection				0 2		8 22		1 7		30		28		2	
V Env. Analysis				2 1		4 5		1 1		10		7		3	
VI Materials and Licenses				0 3		0 11		0 4		26		42		(16)	
VII Const and OPN Planning								0 0		8		9		(1)	
VIII Economic Assessment															
IX Tech. Support															
X Long Lead Items								4 4		17		19		(2)	
XI Project Mgt				10 10		75 78		10 40		242		202		40	
21. Total				16 28		93 151		17 80		408		408		0	
22. Remarks MLGW										23. Manpower Expressed In: 100 Man-hours					
24. Signature of Contractor's Project Manager and Date R W Gray / RGH 7-17-78						25. Signature of Government Technical Representative and Date									

MANPOWER MANAGEMENT REPORT

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MANPOWER MANAGEMENT REPORT

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MANPOWER MANAGEMENT REPORT

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INDUSTRIAL FUEL GAS

DEMONSTRATION PLANT

Page 1 of 2

INTERMEDIATE EVENT LOG

TASK	SUBTASK	SUBTASK DESCRIPTION	INTERMEDIATE EVENTS	SCHEDULED RELEASE	ACTUAL RELEASE
I	I-1	Process Eng. Design	A) Process Units	4/15/78	4/15/78
			B) Offsites	4/28/78	4/28/78
	I-3	Project Eng. Design	C) Wastewater Treatment	4/15/78	4/15/78
			D) Plot Plans	4/28/78	4/28/78
			E) Buildings	4/28/78	4/28/78
			F) Coal Ash Handling	4/15/78	4/15/78
			G) Dock Facilities	4/28/78	4/28/78
			H) Risk Analysis	5/26/78	6/30/78
	I-2	Risk Analysis	J) Cost Estimates	6/02/78	6/30/78
	I-4	Cost Estimates	K) Econ/Env. Impact	6/15/78	
	I-5	Econ/Env. Impact	L) Demo Plant Recomm.	6/30/78	
II	I-6	Demo Plant Recomm.	A) Basis of Design	4/28/78	4/28/78
	II-1	Basis of Design	B) Onsites	6/30/78	
			C) Offsites	8/15/78	
	II-2	Process Eng. Design	D) Wastewater Treatment	6/23/78	
			E) Utilities	6/30/78	
			F) Flare	6/09/78	
			G) Plant and Instr. Air	6/30/78	
			H) Glycol Cooling	6/30/78	
			J) Tankage	6/23/78	
			K) Interconnecting Ppg.	6/02/78	
			L) Firewater	6/23/78	
	II-3	Overall Process Descrip.	Overall Process Descrip.	8/26/78	
III	III-1	Prel. Eng. and Layouts		2/23/79	
	III-2	Plant Specifications		4/13/79	
	III-3	Cost Estimating	To be developed	4/13/79	
	III-4	Process Eng. Report		5/04/79	
	III-5	Failure Analysis		2/23/79	
	III-6	Procurement Req.		2/23/79	
IX	IX-1	Design Data	A) -Commercial Plant H and M Balance	12/30/77	12/30/77
			B) Confirm H and M Balance	2/24/78	2/24/78
			C) Confirm Gasifier Design	3/24/78	3/24/78
			D) Confirm H and M Balance Gasifier Product	6/09/78	
			E) Final Gasifier Design	9/22/78	
			A) Atmospheric Tests	3/24/78	3/24/78
	IX-2, 3, 4	Pilot Plant	B) Atmospheric Operating Data from Pressure Modifications	5/19/78	
			C) Operating Data for Adjust to Gasifier Design	7/21/78	
			To be Used in Conjunction with Milestone Plan and Management Report.

INDUSTRIAL FUEL GAS

DEMONSTRATION PLANT

Page 2 of 2

INTERMEDIATE EVENT LOG

TASK	SUBTASK	SUBTASK DESCRIPTION	INTERMEDIATE EVENTS	SCHEDULED RELEASE	ACTUAL RELEASE
	IX-5	Mechanical Configuration	A) Define Scope B) Detailed Equip Sizing D) Equipment Rec. E) Site Eval. and Selection F) Construction (Start) G) Construction (Compl) H) Venturi Arrangement J) Feed Loc. and Dust Recycle	2/17/78 4/07/78 7/07/78 3/17/78 4/07/78 7/28/78 9/08/78 11/17/78	2/17/78 4/07/78 4/07/78 3/17/78 4/07/78
	IX-6	Mathematical Model	K) Turn Down A) Final Cases and Reports B) Modify Exist Prog. for Atmospheric Pressure C) Final Cases for Adjust to Gasifier Design D) Write Mathematical Equations E) Run Program F) Final Cases	2/02/79 3/24/78 3/24/78 7/21/78 9/08/78 12/8/78 2/16/79	3/24/78 3/24/78
	IX-7	Combustion and Sales	A) Customer Combustion Equip. Identif. B) Confirm Customer Combustion Equip. C) Detail Base Case Alternate D) Est. Ceiling Price Based on Commercial Plant E) Prepare Sales Offer F) Secure Customer Letter of Intent G) Final Project Decision	4/14/78 1/19/79 3/24/78 7/21/78 7/21/78 12/22/78 3/2/79	4/14/78

APPENDIX TO JUNE MONTHLY REPORT PILOT PLANT TEST PROGRAM

For the Period April Through May 1978

INDUSTRIAL LOW-Btu FUEL GAS DEMONSTRATION-PLANT PROGRAM

**Submitted by
MEMPHIS LIGHT, GAS AND WATER DIVISION
P.O. Box 430
Memphis, Tennessee 38101**

**Prepared by
Institute of Gas Technology
IIT Center, 3424 S. State Street
Chicago, Illinois 60616**

**Prepared for the
UNITED STATES DEPARTMENT OF ENERGY**

Under Contract No. ET-77-C-01-2582

PILOT PLANT TEST PROGRAM

Summary

During April the pilot plant was shut down for modifications and repairs. The major modification was the reduction of the gasifier diameter from 4 to 3 feet up to a height of 6 feet from the bottom gas distributor grid. Some of the changes required for operating the pilot plant at a higher pressure were also completed. In addition, a detailed engineering and system analysis of the entire pilot plant facility was conducted by the industrial team members. Changes for more reliable and safer operation that were recommended as a result of this analysis were also implemented during this period. Details of the activities during April were presented in a separate report – AAG Pilot Plant – Design Analysis, Risk Analysis, and Configuration Control, May 1978.

Two tests (Runs 123 and 124) were conducted during May and the early part of June. In Run 123, Western Kentucky coal was fed for 38 hours during 70 hours of operation. Gasification of the coal was conducted with an air-steam mixture at a temperature of 1840°F and a superficial velocity of 4.3 ft/s. During the test, the bed ash content was maintained at about a 40% ash level by using the alternate bed discharge line. An instrumentation malfunction misrepresented the bed height as 4 feet instead of 8 feet. This error was caused by the erosion of an internally extended pressure tap. The internal cyclone dipleg was completely immersed in this 8-foot deep bed, thus imparting excessive cyclical stress, and causing the dipleg to break loose and fall into the bed. The test run was terminated. Physical examination revealed excessive erosion-corrosion of the internal cyclone outlet tube, several thermowells, and the extended pressure tap. The corroded thermowells were replaced and coated with an abrasion-resistant alumina. The dipleg (which was deformed) of the internal cyclone and the outlet tube were replaced. The diameter of the latter was decreased from 9 to 8 inches and was refractory-lined to resist erosion. To provide a smooth surface for the flow of bed solids along the gas distributor grid, the hex-head grid nozzles were replaced by 1/2-in. round bar stock having holes flush with the grid surface, directing the fluidizing gas vertically into the reactor.

Run 124 is by far the most successful test run: 305 hours of operation were logged. Western Kentucky No. 9 run-of-mine coal was fed for 168 of these hours. Coke ash agglomerates were produced at 1950°F using a 4 ft/s superficial velocity. Coal feeding was conducted at an operating temperature of 1875°-1900°F while gasifying with

oxygen-enriched steam-air at a superficial velocity of 4 ft/s. During this test steady periods of coal operation were produced at operating temperatures of 1820°F. Later, operations were conducted in the steam-oxygen mode at an average temperature of 1810°-1815°F with coal feed at a superficial velocity of 4 ft/s. During operations with coal and steam-oxygen, the product gas heating value varied between 250 and 270 Btu/SCF. In both operating modes the bed ash content was maintained at 40%-50% by means of the bed withdrawal system. The same procedure was used to operate at 50%-60% ash content in the latter part of the steam-oxygen operating phase. When the coal silo ran empty and sufficient data had been obtained, a voluntary shutdown was initiated, and the test run was terminated. Upon reactor examination about 100 pounds of deposit was found partially covering the grid and the venturi throat. Sintered fines flakes were found lodged among the gas straightening vanes located below the neck of the venturi, so we decided to remove these vanes prior to Run 125.

PILOT PLANT OPERATIONS

Run 123

The test plan for Run 123 consisted of the following:

1. Gasify 1/4 x 0 in. Western Kentucky No. 9 run-of-mine coal.
2. Maintain a fluidized-bed temperature of 1850°F and a superficial velocity of 4 ft/s with steam-air operation.
3. Determine the effect of an expanded reactor freeboard on the fines carryover rate in absence of the internal cyclone.
4. Recirculate second-stage cyclone fines directly into the bed.
5. Achieve coal ash agglomerate formation at 1850°-1875°F using an oxygen-enriched steam-air gasifying medium. Upon attaining sufficient data, change the gasification operation to the steam-oxygen mode.

Actual operation of Run 123 consisted of the following:

1. The test run encompassed 70 operating hours with 38 hours of coal feed.
2. Air-steam gasification was conducted at an operating temperature of 1840°F and a superficial velocity of 4.28 ft/s.
3. While gasifying coal, a 40% bed ash level was maintained by using the alternate bed discharge line.
4. Because the dipleg was immersed in the bed, the internal cyclone fell onto the grid of the reactor. The presence of the cyclone caused defluidization of the bed, and the run was terminated.

Run 123 Operation

Run 123 was initiated on Monday, May 8, at approximately 1900 hours. A fluidized-bed temperature of 1860°F was achieved at 1600 hours on May 9. At this time operations were being conducted with coke as the feed and air-steam as the gasifying medium. The superficial velocity was controlled at 4 ft/s and the bed temperature at 1850°-1860°F by adjustment of the process gas (air-steam) rates and ratios. After 6 hours at this temperature with coke feed the bed ash content had exceeded 25%, and the formation of ash agglomerates was not apparent. Operations were switched to coal feed at midnight on May 9. The bed ash level was allowed to reach 40%, and the bed withdrawal system was then activated.

On May 10, shortly after noon, there was a process upset, and a loss of feed occurred. The unit was brought down in temperature and idled while the source of the

feed interruption was located. Nitrogen purge gas was leaking from a valve located on an alternate discharge line attached to the lockhopper feeding the reactor screw feeder. This leak caused reactor back-gassing up through the L-valve and stopped the coal feed. A blind flange was installed on this valve, and the back-gassing ceased. Inadvertently, coal feed was restored when the bed temperature was still around 1000°F: When this was discovered, coke feed was restored. There was some evidence of possible defluidization, although the chart traces did return to fairly normal settings within an hour. However, there was a significant increase in the fluidized-bed density reading, which, after a check of the transmitter impulse lines, persisted. Upon restoration of coke feed some oversize lumps discharged from the venturi. At 1700 hours, the Joy process air compressor again developed water leakage into the oil sump, and the auxiliary rental unit was started. Because of an excessive amount of fines being carried out of the reactor, the internal cyclone was activated at 2100 hours. Shortly thereafter, coal feed was restored because all recordings and discharge streams appeared normal. The use of the bed withdrawal line was continued to limit the bed ash content to 40%-50%.

Operations continued in this manner overnight from May 10 to 11 with an average of 200 lb/hr of fines being removed from the system. The feed rate was about 1038 lb/hr. At about 1040 hours on May 11, all pressure-drop readings for the internal cyclone fell to zero. Immediately, a drop in bed level and a simultaneous large increase in fines carryover to the external cyclones developed. An increase in feed rate and air flow was made and partially compensated for the drop in temperature to 1800°F.

During the balance of the test run, which was terminated at 2200 hours on May 11, a low (6 lb/hr) venturi discharge rate persisted. (It had been 20-25 lb/hr before 1200 hours on May 11.) The low discharge rate from the venturi gave evidence that something was blocking the venturi. At 2130 hours an examination of the bed view port revealed a darkened stagnant zone, and the unit was shut down.

When the reactor was opened following Run 123, significant damage to some high-alloy metal components was apparent. Bed thermowells No. 5, 6, and 8 were heavily corroded as was the element itself. The bed density low-pressure tap, which had an extension down the reactor wall, was corroded-eroded with a hole located approximately 15 inches above the bottom of the tap.

The internal cyclone had fallen into the reactor bed. The cyclone dipleg had collapsed, and the reverse seal pot had become corroded. The cyclone had pulled loose by disengaging its outlet tube from the upper cyclone shell sheet. A small amount of sinter accompanied by about 100 pounds of clinker was found on the grid. This

accumulation was the result of the reverse seal pot blocking the venturi and part of the grid. A deposit of relatively high carbon content was found surrounding the new feed nozzle. The third-stage cyclone, which had developed a small leak during the test, was found to be severely worn at the bottom of the cone where it joined with its dipleg.

The following repairs were implemented to correct those problems:

- Thermowells and thermocouples No. 5, 6, and 8 were replaced and were plasma coated with aluminum oxide to increase their erosion-corrosion resistance.
- To reinforce the internal cyclone, four special gussets were installed to connect the new outlet pipe, the eccentric support cone, and the top sheet of the cyclone.
- The dipleg of the internal cyclone was replaced.
- The 9-in. diameter outlet of the internal cyclone was eroded on the outside. This erosion was similar, but not as severe, as that suffered by the previous cyclone with an 11-in. outlet. This outlet pipe was replaced by an 8-in. diameter pipe and was refractory lined on the outside.
- The cone of the third-stage cyclone was replaced because it was severely worn from 1-2 feet up from the bottom. In addition, to minimize wear through improved solids disengagement, the barrel portion of the cyclone was lengthened by 1 foot.

The following work, some of it modification completion work, was accomplished before starting Run 124:

- The 3-in. pipe nitrogen supply line connection to the steam-iron facility was completed. A larger capacity will now be available for rapid cooling of the bed.
- The hex-head grid nozzles that had been relatively standard in this unit were replaced by 1/2-in. round bar stock with holes drilled in such a manner as to direct the fluidizing gas vertically into the reactor. This nozzle change is expected to provide the smooth grid surface needed for the flow of solids so as to eliminate sintering on the grid.
- Because of the corrosion-erosion of the bed density low-pressure tap, the unit had been operated with an 8-foot instead of a 4-foot deep bed after the feed upset. At this depth the dipleg of the internal cyclone was immersed. Because of excessive vibration and low mechanical strength (at 1800°F) the cyclone suffered fatigue failure. Although previously in use for some time, this tube extension will be cut off and another tube installed (through a lower nozzle) to provide an equivalent density reading.

Table 1 shows a summary of the operating conditions for a 10-hour period during Run 123 prior to the dislodgment of the internal cyclone. Table 2 shows test data for the entire duration of Run 123. The solids analysis shown with ash composition only was conducted at the pilot plant during operation of the test run as a control method. All other listings on solids compositions were obtained from the IGT State Street analytical laboratory.

Table 1. ASH AGGLOMERATING GASIFIER OPERATING CONDITIONS
SUMMARY FOR RUN 123
(5/10/78-5/11/78; 2400-1000 Hours)

Run No. <u>123</u> Date <u>5/10-5/11/78</u> Time Period <u>2400 to 1000</u>		
Venturi Diameter <u>3</u> inches Feed Material <u>W. Ky. No. 9 Coal</u>		
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>		
1	Average Fluidized Bed Temperature, °F	1838
2	Average Reactor Pressure, psig	8.0
3	Venturi Pressure Drop, psi	0.77
4	Superficial Velocity, ft/s	4.28
5	Freeboard Velocity, ft/s	--
6	Venturi Velocity, ft/s	77.91
7	Grid Gas, mol/hr	66.84
8	Grid Gas Composition: %N ₂	63.17
9	%O ₂	16.79
10	%Steam	20.04
11	Venturi Gas, mol/hr	27.72
12	Venturi Gas Composition: %N ₂	64.68
13	%O ₂	17.21
14	%Steam	18.11
15	Fluidized Bed Density, lb/ft ³	31.0
16	Fluidized Bed Height, ft	7.01
17	Feed Composition, <u>1037.5</u> lb/hr, %C/%Ash	68.80/19.16
18	Average Fluidized Bed Composition, <u>30.46</u> lb/hr, %C/%Ash	60.70/36.05
19	Bed Discharge Composition, <u>198.56</u> lb/hr, %C/%Ash	51.50/44.47
20	Venturi Discharge Composition, <u>33.78</u> lb/hr, %C/%Ash	29.25/65.09
21	2nd Stage Fines Composition, <u>150.7</u> lb/hr, %C/%Ash	74.25/21.77
22	3rd Stage Fines Composition, <u>26.3</u> lb/hr, %C/%Ash	66.15/30.46
23	Carbon/Oxygen (Weight Basis)	1.37
24	Carbon/Steam (Weight Basis)	2.12

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.2	10.6	18.9	20.3	18.1	13.9	10.5	4.3	3.2							
Fluidized Bed	0	2.9	18.4	25.7	19.9	16.1	13.9	2.8	0.3							
Discharged Bed	0	0.8	7.4	17.2	35.5	19.3	17.2	2.5	0.1							
Ash	0.4	21.3	32.8	18.9	11.1	7.8	--	--	7.7							
2nd Stage Fines					0.5	1.5	6.7			8.4	3.7	12.3	37.6	19.0	9.5	0.8
3rd Stage Fines																

Table 2, Part 1. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 123

DATE 5/9-5/10/78 RUN NUMBER 123

TIME	1200	1400	1600	1800	2000	2200	2400	5/10 0200	0400
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	618	588	576	664	675	673	582	699	750
TE 9206.2 GAS TEMP ENTERING GRID, °F	534	520	341	304	482	438	426	402	331
TE 9206.3 BED TEMP, °F	1372	1765	1862	1860	1855	1851	1840	1805	1825
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1140	1541	1732	1727	1735	1750	1728	1687	1698
FR 903 STEAM FLOW TO GRID, mol/hr	13.45	14.07	14.65	14.48	25.86	28.22	21.10	13.52	13.56
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 AIR FLOW TO GRID, mol/hr	39.61	39.81	54.20	47.42	43.98	44.38	46.79	55.01	48.59
FR 906 AIR FLOW TO VENTURI, mol/hr	29.16	28.70	25.66	22.55	24.99	25.04	25.68	28.28	27.05
FR 909 SWEEP GAS (N ₂), mol/hr	2.71	4.41	4.46	4.29	4.17	3.31	3.23	3.95	3.74
FR 938 OXYGEN TO AIR HEATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	2.58	3.16	3.53	3.43	3.41	2.96	4.37	4.52	4.29
FR 937 ATOMIZING N ₂ , mol/hr	2.53	2.53	2.46	2.46	2.46	2.40	2.40	2.40	2.40
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	0	2.29	36.58	38.86	40.01	41.15	37.72	49.15	44.58
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	4.91	4.97	5.97	5.97	7.46	7.46	7.46
FR 995 AIR TO AIR HEATER, mol/hr	73.86	74.18	87.77	77.53	75.96	75.21	77.20	88.87	81.46
FR 9304 STEAM TO VENTURI, mol/hr	0	0	0	0	0	0	0	2.46	2.41
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	1.05	2.65	2.75	4.80	4.30	7.30	6.25	3.75	4.00
NITROGEN PURGE TO REACTOR, mol/hr	10.67	11.02	9.73	10.19	10.11	10.11	10.11	10.30	10.22
BED DENSITY, lb/ft ³	35.3	32.5	28.2	28.2	29.6	31.0	31.0	26.8	29.6
BED HEIGHT, ft	3.1	3.2	5.11	5.11	4.67	4.27	4.55	5.16	4.38
REACTOR GAS COMPOSITION (dry), %									
N ₂									
CO									
CO ₂									
H ₂									
CH ₄									
H ₂ S									
COKE OR COAL FEED RATE, (wet) (avg), lb/hr	COKE→ 555	555	555	555	555	555	COAL→ 1025	1025	1025
C-7 BED DISCHARGE RATE, lb/hr	0	0	0	151.9	100.9	63.4	63.4	0	0
C-4 ASH DISCHARGE RATE, lb/hr	27.7	8.3	11.0	10.0	12.3	10.0	8.0	9.7	18.7
2nd STAGE FINES REMOVAL RATE, lb/hr	73.4	73.4	73.4	73.4	73.4	73.4	417	417	417
3rd STAGE FINES REMOVAL RATE, lb/hr	60.5	60.5	60.5	60.5	60.5	29.2	29.2	29.2	29.2
FEED COMPOSITION, %: C					84.30				
ASH					12.48				
REACTOR BED COMPOSITION, %: C					73.20				
ASH		14.0	17.0	27.0	25.56	22.0	28.0	30.0	47.0
DISCHARGED BED COMPOSITION, %: C					71.40				
ASH					22.24				
ASH COMPOSITION, %: C					54.30				
ASH				62.0	44.60				
2nd STAGE FINES COMPOSITION, %: C					83.90				
ASH			14.0		14.32				
3rd STAGE FINES COMPOSITION, %: C					79.70				
ASH					18.34				

Table 2, Part 2. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 123

DATE 5/10/78 RUN NUMBER 123

TIME	0600	0800	1000	1200	1400	1600	1800	2000	2200
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	679	641	553	667	463	650	625	710	615
TE 9206.2 GAS TEMP ENTERING GRID, °F	488	306	475	256	473	441	445	436	434
TE 9206.3 BED TEMP, °F	1789	1836	1657	1846	1810	1839	1854	1842	1833
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1658	1705	1639	1695	1578	1682	1711	1722	1735
FR 903 STEAM FLOW TO GRID, mol/hr	13.72	13.92	13.70	13.62	19.30	41.17	39.96	43.53	16.19
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 AIR FLOW TO GRID, mol/hr	48.87	52.60	53.74	56.65	61.65	83.49	56.28	53.14	69.84
FR 906 AIR FLOW TO VENTURI, mol/hr	26.57	26.73	27.31	26.49	26.50	24.43	24.73	24.29	22.50
FR 909 SWEEP GAS (N ₂), mol/hr	2.58	2.60	3.37	3.06	4.23	3.15	3.78	3.12	3.00
FR 938 OXYGEN TO AIR HEATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	4.03	3.90	4.90	4.15	5.13	1.99	4.73	3.98	3.78
FR 937 ATOMIZING N ₂ , mol/hr	2.40	2.40	2.40	2.40	2.21	2.21	2.40	2.40	2.40
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	43.43	41.15	48.01	38.86	45.72	68.58	68.58	64.01	45.72
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	7.46	7.46	6.96	7.46	7.46	7.46	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	81.80	83.62	90.57	90.57	106.64	102.63	98.17	98.23	95.91
FR 9304 STEAM TO VENTURI, mol/hr	3.95	3.92	3.91	3.86	4.34	7.82	10.96	10.85	7.93
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	9.60	2.65	2.10	2.10	2.20	2.10	2.10	5.15	5.25
NITROGEN PURGE TO REACTOR, mol/hr	10.10	9.92	10.30	10.30	10.30	10.30	10.30	10.30	10.30
BED DENSITY, lb/ft ³	33.9	33.9	31.0	31.0	31.0	31.0	--	31.0	31.0
BED HEIGHT, ft	3.23	3.23	3.48	3.71	2.78	3.25	--	4.45	6.03
REACTOR GAS COMPOSITION (dry), %									
N ₂				61.20				73.10	
CO				12.70				4.90	
CO ₂				8.70				13.00	
H ₂				15.00				7.30	
CH ₄				1.70				0.24	
H ₂ S				0.00				0.00	
COKE OR COAL FEED RATE, (wet) (avg), lb/hr	COAL+ 1025	1025	1025	1025	COKE+ 1200	1200	1200	1200	COAL+ 1037.5
C-7 BED DISCHARGE RATE, lb/hr	49.7	143.0	147.7	0	85.2	56.8	0	0	240.7
C-4 ASH DISCHARGE RATE, lb/hr	25.3	21.3	13.7	19.7	90.3	21.3	25.0	27.0	27.7
2nd STAGE FINES REMOVAL RATE, lb/hr	417	417	417	417	597.2	597.2	597.2	597.2	150.7
3rd STAGE FINES REMOVAL RATE, lb/hr	29.2	29.2	29.2	29.2	5.7	5.7	5.7	5.7	5.7
FEED COMPOSITION, %: C									
ASH									
REACTOR BED COMPOSITION, %: C									
ASH	46.0	58.0	61.0	48.0	54.0	43.0	--	32.0	37.0
DISCHARGED BED COMPOSITION, %: C									
ASH		46.0							
ASH COMPOSITION, %: C									
ASH			83.0		89.0				
2nd STAGE FINES COMPOSITION, %: C									
ASH								20.0	
3rd STAGE FINES COMPOSITION, %: C									
ASH									

Table 2, Part 3. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 123

DATE 5/10-5/11/78 RUN NUMBER 123

TIME	2400	5/11 0200	0400	0800	1000	1200	1400	1600	1800	2000
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	750	783	521	572	519	421	434	340	383	376
TE 9206.2 GAS TEMP ENTERING GRID, °F	439	439	399	429	464	496	498	295	344	442
TE 9206.3 BED TEMP, °F	1850	1831	1826	1841	1841	1807	1802	1785	1807	1805
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1744	1730	1730	1743	1745	1726	1706	1695	1713	1724
FR 903 STEAM FLOW TO GRID, mol/hr	12.87	11.59	10.94	11.72	11.09	11.01	11.52	11.55	11.44	11.59
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0	0
FR 905 AIR FLOW TO GRID, mol/hr	55.00	55.37	54.36	56.62	54.67	55.35	57.35	51.55	75.37	79.86
FR 906 AIR FLOW TO VENTURI, mol/hr	22.76	22.91	25.56	22.98	22.97	22.81	22.76	24.57	26.71	24.71
FR 909 SWEEP GAS (N ₂), mol/hr	2.99	2.06	1.15	2.04	2.52	2.48	2.77	2.77	2.23	2.27
FR 938 OXYGEN TO AIR HEATER, mol/hr	0	0	0	0	0	0	0	0	0	0
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.95	4.08	4.01	4.08	4.05	3.55	3.63	3.66	3.51	3.56
FR 937 ATOMIZING N ₂ , mol/hr	2.40	2.40	2.40	2.40	2.34	2.40	2.40	2.34	2.40	2.34
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	46.86	45.72	46.86	38.86	35.43	52.58	48.01	49.15	54.86	65.15
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	4.97	4.97	4.97	2.49	2.49	0.99	0.99	4.97	0	0
FR 995 AIR TO AIR HEATER, mol/hr	94.65	95.65	96.16	95.75	95.49	94.58	94.98	91.44	104.05	115.29
FR 9304 STEAM TO VENTURI, mol/hr	4.37	5.62	3.90	3.73	3.74	3.13	4.04	0	0	0
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	4.00	1.05	1.05	2.10	0	4.30	3.75	4.35	3.15	3.15
NITROGEN PURGE TO REACTOR, mol/hr	10.92	11.02	11.16	11.09	11.06	11.06	10.83	9.77	9.65	9.65
BED DENSITY, lb/ft ³	31.0	31.0	31.0	31.0	31.0	31.0	31.0	31.0	31.0	31.0
BED HEIGHT, ft	6.03	7.05	7.01	7.52	7.42	6.12	6.03	5.57	5.57	5.57
REACTOR GAS COMPOSITION (dry), %										
N ₂	69.30					62.70		61.70		62.70
CO	11.20					10.10		8.40		9.00
CO ₂	5.80					8.40		9.10		9.50
H ₂	12.00					15.00		13.60		14.20
CH ₄	0.62					2.60		3.00		3.30
H ₂ S	0.00					0.00		0.10		0.00
COAL FEED RATE, (wet) (avg), lb/hr	1037.5	1037.5	1037.5	1037.5	1037.5	1680	1680	1680	1680	1680
C-7 BED DISCHARGE RATE, lb/hr	49.8	134.2	298.7	151.0	359.1	91.3	141.6	234.5	189.5	624.8
C-4 ASH DISCHARGE RATE, lb/hr	28.0	33.0	64.3	23.3	20.3	12.0	6.7	6.7	7.0	4.0
2nd STAGE FINES REMOVAL RATE, lb/hr	150.7	150.7	150.7	150.7	150.7	512.0	512.0	512.0	512.0	512.0
3rd STAGE FINES REMOVAL RATE, lb/hr	5.7	5.7	5.7	5.7	108.7	108.7	147.1	147.1	147.1	147.1
FEED COMPOSITION, %: C			68.80			61.40		63.30		
ASH			19.16			22.32		20.38		
REACTOR BED COMPOSITION, %: C	66.90		54.50			48.00		53.80		
ASH	31.13	36.0	40.96	44.0	46.0	47.06	52.0	41.67	47.0	44.0
DISCHARGED BED COMPOSITION, %: C	53.90		49.10					47.00		
ASH	42.49		46.44					49.17		
ASH COMPOSITION, %: C	32.10		26.40			11.40		15.90		
ASH	62.49		67.68	79.0		79.92		75.76		
2nd STAGE FINES COMPOSITION, %: C	74.80		73.70			56.30		66.40		
ASH	21.50		22.04			40.07		28.42		
3rd STAGE FINES COMPOSITION, %: C	66.50		65.80			58.40		66.80		
ASH	31.08		29.83	33.0		35.83		27.56		

Run 124

The test plan for Run 124 consisted of the following:

1. Gasify Western Kentucky run-of-mine coal.
2. Using oxygen-enriched steam-air, maintain bed temperatures up to 1900°F with a 4-foot bed height and a 4 ft/s superficial velocity to produce ash agglomerates from coke for at least 12 continuous hours. If satisfactory ash agglomerates are not formed at 1900°F, lower the bed ash to 20% by discharging bed material and raise the temperature up to 1950°F in two 25°F increments to obtain steady and sufficient ash agglomerate discharge.
3. With the coal feed, use oxygen-enriched air-steam to maintain a temperature of 1875°F with a 4-foot bed height and a 4 ft/s superficial velocity.
4. Maintain up to 50% bed ash content by withdrawal of bed material, and achieve an overall ash balance around the reactor system for 12 hours.
5. Allow the bed ash content to rise to 60%, and determine the extent of ash agglomerate production. In any event, maintain an ash balance for 24 hours.
6. Reinject second- and third-stage fines into the bed.
7. Use the steam-oxygen mode of gasification to produce ash agglomerates at a 4 ft/s superficial velocity and a 4-foot bed height.

Actual operation of Run 124 consisted of the following:

1. The test run encompassed 305 operating hours with 168 hours of coal feed. (This is the longest operation with Western Kentucky coal to date.)
2. At about 1950°F and a 4 ft/s superficial velocity coke ash agglomerates were produced and continued to discharge for about an 8-10 hour period.
3. Operations were then conducted with coal feed at 1875°-1900°F using steam-air plus oxygen process gas at a 1 ft/s superficial velocity. Several steady periods of coal operation were achieved at a temperature of 1820°F.
4. The steam-oxygen mode of coal gasification also produced several steady-state periods at temperatures of 1810°-1815°F and a 4 ft/s superficial velocity. Operations were maintained at 40%-50% bed ash initially and 50%-60% bed ash content in the later part of the run by alternate withdrawal of bed material to achieve an overall ash balance. The product gas heating value varied between 250 and 270 Btu/SCF.
5. A voluntary shutdown was initiated after 10 days of continuous operation.

Run 124 Operation

Run 124 was initiated on Tuesday, May 23, at 2330 hours. After reactor preheating was completed, coke was fed and a 3-4 foot deep fluidized bed accumulated. The start-up burner was then ignited at 0600 hours on May 28. The bed temperature rose

slowly, reaching only 800°F in 14 hours. At 2330 hours a hot spot was observed on the reactor wall above the start-up burner nozzle. It was decided to turn off the start-up burner, drop the bed, open the reactor, and check the burner. Upon removing the burner tube 4 hours later, we saw that the flame retention nozzle along with the reactor nozzle enclosing it had been partially destroyed. The burner flame had penetrated back through the light refractory and overheated the reactor wall. The destroyed light weight (LW-20) refractory has been replaced with Kaowool, and the two nozzles were replaced on May 29. This equipment was damaged because of insufficient cooling nitrogen flow in the annular space between the two nozzles. To prevent a recurrence, a solenoid valve was used to assure nitrogen flow at any time the burner was operational. In addition, a thermocouple was used to monitor this annular temperature so that flow can be adjusted as needed.

During this aborted start-up of Run 124, a very high grid pressure drop (10-12 psi) was observed. This pressure drop was significantly greater than that determined from standard calculations by correlation with previous bolt-type nozzles. A sample of this new grid nozzle (Figure 1) installed for this run was tested in the laboratory and produced a pressure drop close to the predicted value (1-2 psi) at the operating flow conditions. Therefore, it was decided to ream out all the grid nozzles to eliminate any blockage created when they were welded to the grid.

Run 124 was reinitiated on Saturday, May 27, 1978. The unit was brought to an operating temperature of 1872°F around 1800 hours on May 28. The operation continued up to 0600 hours on Monday, May 29, without any interruption. Early that morning, a failure in drive mechanism in the reactor screw feeder occurred. The diagnosis was that either replacement parts or machining of the new parts was necessary to complete the repairs. No vendors or machine shops were available on Monday, May 29, because of the holiday; consequently, the reactor was brought down to below char ignition temperature (1000°F) and idled.

Tuesday, May 30, was spent on locating parts, getting dimensions, and machining of a replacement part. The unit was repaired on Wednesday, May 31. It was restarted at approximately 1600 hours on May 31 and brought to a bed temperature of 1875°F at 2400 hours with air-steam operation. On Thursday, June 1, oxygen was added, and the unit temperature was raised to 1925°F around 0400 hours. The bed ash content was near 50%. At this point, the bed withdrawal system was activated to reduce the bed ash content to 30%. At 1800 hours on June 1 the bed temperature was raised to 1950°F using oxygen. Shortly after 2200 hours, feed problems developed because of low plant air supply

Figure 1. DETAILS OF GRID NOZZLE

pressure. The feed was restored in approximately 3 hours. Significantly, ash agglomerates were produced from coke while operating at 1950°-1975°F.

When the capability to produce agglomerates from coke was reestablished and feed capability restored, operations were switched to coal feed at about 0230 hours on Friday, June 2. The test procedure for the first phase of coal feed was now being implemented. Operations were conducted at approximately 1875°F at a 4 ft/s superficial velocity and continuous bed withdrawal to sustain a 40%-50% bed ash content. Ash agglomerates first observed during coke feed at 2100 hours on June 1 continued to discharge during coal feed through 1100 hours on June 2. The feed rate was about 1000 lb/hr. These operating conditions were to be maintained until all ash discharged through the venturi or the need for alternate bed discharge was confirmed to achieve an ash balance.

The bed temperature was raised to about 1900°F on June 3 using oxygen-enriched steam-air. The bed ash content was maintained between 40% and 50% using the alternate discharge line, and the temperature was maintained for about 12 hours. Fines removal rate was about 160 lb/hr, and feed rate was 1275 lb/hr (0200-1600 hours on June 3). At this stage some sinter formation was apparent because of the presence of sinter-like flakes discharging from the venturi. On June 4, operations continued at somewhat lower temperatures of 1870°-1880°F (0200-1800 hours). The feed rate was reduced to 900 lb/hr, and the fines rate came down to 93 lb/hr. At 2000 hours on June 4 venturi discharge ceased. An attempt was made to unplug the venturi, but this was not successful.

From 2000 hours on June 4 to 2000 hours on June 6 the bed temperature in the reactor was lowered to 1825°-1850°F. By alternate bed discharge, the bed ash content was maintained at 40%-50%, and the feed rate was approximately 800 lb/hr with the fines rate at 90 lb/hr.

Early on June 7, operations with coal were switched to steam-oxygen, and the test run was continued. Even though the venturi was blocked and ash agglomerates were not being produced, the test was providing very useful data on direct feeding of coal, gas yields and composition, trace impurities in the product gas, and waste water composition. The bed ash content was maintained at 40%-50% by alternate bed withdrawal, and the bed temperature was maintained at about 1820°F. The change in gasification mode produced an increase in feed rate to 1020 lb/hr, and operations continued up to 1200 hours. Shortly after 1200 hours, a feed interruption occurred because the lockhopper valves malfunctioned. The problem was corrected; coke feed was initiated, and the

temperature was raised to 1827°F at about 1600 hours on June 7. Shortly after 1800 hours, coal feed replaced coke feed.

Gasification operations continued on June 8 at a temperature of approximately 1810°F with coal feed. The bed ash content was allowed to increase to 55%-60%: The objective of increasing the ash level was to operate the unit at higher carbon conversions in the steam-oxygen mode. The feed rate was 1036 lb/hr, and the fines rate was approximately 124 lb/hr.

Shortly after 2200 hours on June 8 the internal cyclone was deactivated by turning off the aeration gas to the seal pot. The deactivation of the cyclone was evident by a decrease in the pressure drop across the cyclone accompanied by an increase in the fines carried over to the second-stage cyclone from 118 lb/hr to about 309 lb/hr. This operating objective was to ascertain the effectiveness of this internal cyclone. After 0200 hours on June 9 the internal cyclone was reactivated, and the recycle of external fines collected in the second-stage cyclone was started. The recycle of second-stage cyclone fines worked satisfactorily after some adjustment in the reverse seal gas rate.

At about 1200 hours on June 9 the coal supply was exhausted, and an orderly shutdown was initiated to terminate the test run at 1900 hours on June 9.

Examination of the reactor upon cooling revealed that some sintering had occurred, and, in some areas, clinkers had formed. Approximately 100 pounds of sintered clinker were removed from the venturi and grid areas. Complete examination of the internal cyclone showed no evidence of erosion-corrosion on any portions of the outer or inner shell, but showed some accumulation of ash particulate on the cyclone refractory lining itself, mostly on the scroll inlet and on a small area of the outlet tube. These materials were readily removed by light scraping.

The internal cyclone reverse seal pot contained only dust and a few of the scraped-down ash particles mentioned above. There was no damage to the grid start-up burner, thermowells, pressure taps, or feed nozzles.

The oxygen injection pipe, which terminates above the venturi, was severely eroded on its bottom side directly above the center of the venturi. Accordingly, the only work to be done inside the reactor was that of relocating this pipe, such that it projects through the 45° conical grid directly up a distance of 4 feet normal to the grid surface and terminates at that point. An examination below the venturi neck did reveal that some sintered fines discharging during the test run through the venturi had apparently accumulated among the gas straightening vanes located below the neck of the venturi. It was decided to remove these vanes prior to Run 125.

The sintered material amounted to only about 100 pounds and had permitted operation without any hindrance; this indicated that the deposit was not growing continuously during the test but was formed early in the test when operating at a relatively high temperature with coal. After lowering the temperature, the deposit growth was checked, and we were able to continue the test for several more days.

During Run 124, two recorded single malfunctions in the scale cycling system had occurred. Some leakage at an automatic block valve flange had been detected and was rectified by improving the gasketing at that point. Upon recommendation from IGT's operations people, indicating lights are being furnished on the control panel illustrating the valve positions in this scale system so that the cause of any malfunction resulting in the loss of feed to the reactor can be more rapidly diagnosed.

The various phases and modes of operation for Run 124 are shown in Figure 2. Table 3 contains test data for the entire duration of Run 124. Detailed material and energy balances for two different modes of gasification (steam-air and steam-oxygen) for Run 124 are presented in Tables 4 and 5 and Figures 3 and 4.

A reexamination of data from Run 122 has indicated another steady-state period. The heat and material balance for this period is shown in Figure 5, and the relevant operating data are shown in Table 6.

Petrographic Studies

Samples of C_4 discharge from Run 122, obtained during periods when the Kentucky coal was fed, have been examined petrographically. The particles have dispersed ferrous sulfide and a characteristic crystal pattern in the iron-rich low-melting zones; in both characteristics they are very different from agglomerates obtained from coke. Some other structural features of the particles differ from those of particles obtained from the Illinois coal (as well as those from coke). About three-quarters of the particles have an iron-rich, low-melting coating on the outside of an iron-poor aluminosilicate-rich interior (Figure 6). The uncoated particles and the interior regions of coated particles are similar: some are uniformly light in color, others are uniformly light in color except for dark streaks or cracks that suggest the boundaries of agglomerated subparticles, and still others are uniformly dark or mottled (Figure 7). Most of these particles or regions appear to be well sintered or even partially melted. A few particles of char and particles consisting only of iron sulfide and/or oxide are also present. The same kinds of particles are also present in the side discharge (C-7) samples.

Figure 2. TIME SUMMARY OF RUN 124

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Table 3, Part 1. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE	5/28-5/29/78	RUN NUMBER	124						
TIME	1400	1600	1800	2000	2200	2400	5/29 0200	0400	0600
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	639	590	494	516	615	619	596	617	627
TE 9206.2 GAS TEMP ENTERING GRID, °F	474	528	472	445	459	437	567	474	415
TE 9206.3 BED TEMP, °F	1590	1818	1886	1891	1867	1885	1890	1873	1897
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1326	1662	1772	1794	1783	1791	1774	1775	1790
FR 903 Steam FLOW TO GRID, mol/hr	9.01	10.89	19.44	19.52	19.54	20.67	11.30	13.64	11.30
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 Proc. Gas FLOW TO GRID, mol/hr	21.31	48.78	42.07	45.67	44.67	49.40	54.65	57.24	61.18
FR 906 Proc. Gas FLOW TO VENTURI, mol/hr	25.70	25.69	26.06	26.22	22.86	23.49	23.66	22.88	22.80
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	0	0	4.27	0.86	0.85	0	0	0	0
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.06	3.05	2.95	2.97	2.97	2.81	2.82	2.94	2.93
FR 937 ATOMIZING N ₂ , mol/hr	3.30	3.04	2.87	2.89	2.61	2.62	2.92	2.75	2.88
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	11.43	52.00	48.01	45.72	45.72	54.86	22.86	46.86	41.15
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	53.92	82.43	68.88	77.10	74.92	75.23	84.26	84.35	91.52
FR 9304 TO VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	0	2.91	3.74	3.85	3.74	4.26	4.16	5.20	5.20
NITROGEN PURGE TO REACTOR, mol/hr	7.72	8.86	8.88	8.88	8.88	8.89	8.79	8.65	8.65
BED DENSITY, lb/ft ³	45.19	42.53	39.88	42.53	42.53	45.19	46.52	42.53	42.53
BED HEIGHT, ft	2.8	3.1	3.97	4.20	3.99	3.95	3.84	3.72	4.06
REACTOR GAS COMPOSITION (dry), %									
N ₂				60.40		66.10		70.90	
CO				9.70		8.40		10.90	
CO ₂				16.10		14.20		13.90	
H ₂				11.20		10.60		3.60	
CH ₄				1.10		0.11		0.10	
H ₂ S				0.00		0.00		0.00	
Coke FEED RATE, (wet) (avg), lb/hr	300	300	675	675	675	675	637.5	637.5	637.5
C-7 BED DISCHARGE RATE, lb/hr	0	0	0	0	105.7	62.0	62.0	65.0	65.0
C-4 ASH DISCHARGE RATE, lb/hr	3.0	2.7	2.3	2.3	5.0	4.0	5.3	8.0	11.0
2nd STAGE FINES REMOVAL RATE, lb/hr	43.8	43.8	106.5	106.5	106.5	106.5	110.3	110.3	110.3
3rd STAGE FINES REMOVAL RATE, lb/hr	11.8	11.8	15.3	15.3	3.3	3.3	3.3	3.3	3.3
FEED COMPOSITION, %: C									
ASH									
REACTOR BED COMPOSITION, %: C									
ASH		18.0	17.5	25.0	34.0	28.5	32.5	40.0	24.0
DISCHARGED BED COMPOSITION, %: C									
ASH									
ASH COMPOSITION, %: C									
ASH									
2nd STAGE FINES COMPOSITION, %: C									
ASH					22.0				
3rd STAGE FINES COMPOSITION, %: C									
ASH									

Table 3, Part 2. ASH AGGLOMERATING GASIFIER- DATA SUMMARY
FOR RUN 124

DATE 5/31-6/1/78 RUN NUMBER 124

TIME	5/31 2200	2400	6/1 0200	0400	0600	0800	1000	1200	1400
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	606	573	549	689	657	913	711	742	513
TE 9206.2 GAS TEMP ENTERING GRID, °F	509	511	513	511	466	455	295	320	260
TE 9206.3 BED TEMP, °F	1826	1888	1893	1923	1928	1899	1863	1869	1886
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1663	1760	1765	1786	1782	1782	1769	1775	1792
FR 903 Stream FLOW TO GRID, mol/hr	12.21	12.21	13.04	13.03	13.42	15.81	16.21	14.49	14.15
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 Proc. FLOW TO GRID, mol/hr	49.73	51.37	52.29	52.58	52.20	52.32	52.57	50.36	54.01
FR 906 Proc. FLOW TO VENTURI, mol/hr	25.08	24.32	24.23	24.36	24.72	25.48	19.56	17.62	17.71
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	0	0	3.27	2.61	2.60	0	0	0	0
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.08	3.15	3.27	3.28	3.24	3.14	3.21	3.11	3.17
FR 937 ATOMIZING N ₂ , mol/hr	2.66	2.45	2.51	2.41	2.46	1.83	1.90	1.82	1.84
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	37.72	42.29	43.43	45.72	41.15	38.86	51.44	40.01	44.58
FR 9330 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	84.60	81.55	78.48	78.93	78.68	98.92	75.54	71.12	77.96
FR 9304 Steam TO VENTURI, mol/hr	0	0	0	0	0	4.48	10.26	9.40	9.17
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	4.89	4.26	4.06	4.16	5.62	5.10	5.41	4.58	4.58
NITROGEN PURGE TO REACTOR, mol/hr	8.75	8.75	8.27	8.66	8.66	8.77	8.83	8.46	8.46
BED DENSITY, lb/ft ³	41.21	39.88	37.22	37.22	37.22	45.19	46.52	46.52	43.86
BED HEIGHT, ft	3.84	3.97	3.87	3.71	4.02	3.92	4.95	3.71	3.41
REACTOR GAS COMPOSITION (dry), %									
N ₂				59.30	59.30	69.00		67.40	
CO				19.20	19.80	9.20		6.40	
CO ₂				9.20	8.90	12.60		13.10	
H ₂				11.60	11.30	8.40		9.00	
CH ₄				0.09	0.09	0.13		0.13	
H ₂ S				0.00	0.00	0.00		0.00	
Coke FEED RATE, (wet) (avg), lb/hr	450	450	725	725	725	725	725	725	725
C-7 BED DISCHARGE RATE, lb/hr	0	0	0	0	0	127.1	127.1	174.1	174.1
C-4 ASH DISCHARGE RATE, lb/hr	11.0	4.0	4.3	7.3	10.0	23.3	17.0	37.0	27.0
2nd STAGE FINES REMOVAL RATE, lb/hr	82.7	82.7	116	116	116	116	116	116	116
3rd STAGE FINES REMOVAL RATE, lb/hr	5.8	5.8	5.8	5.8	5.8	5.8	10.4	10.4	10.4
FEED COMPOSITION, %:									
ASH									
REACTOR BED COMPOSITION, %:									
ASH	22.0	17.0	20.0	27.0	47.0	40.6	39.5	36.5	32.5
DISCHARGED BED COMPOSITION, %:									
ASH									
ASH COMPOSITION, %:									
ASH									
2nd STAGE FINES COMPOSITION, %:									
ASH									
3rd STAGE FINES COMPOSITION, %:									
ASH									

Table 3, Part 3. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE 6/1=6/2/78 RUN NUMBER 124

TIME	1600	1800	2000	2130	2400	6/2 0400	0600	0800	1000
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	510	459	610	710	605	424	408	426	371
TE 9206.2 GAS TEMP ENTERING GRID, °F	310	321	334	415	442	431	406	404	409
TE 9206.3 BED TEMP, °F	1888	1888	1954	1940	1753	1881	1882	1898	1869
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1786	1793	1809		1701	1792	1797	1759	1782
FR 903 Steam FLOW TO GRID, mol/hr	13.38	13.31	14.88	15.50	24.41	13.95	13.24	13.73	13.95
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 Proc. Gas FLOW TO GRID, mol/hr	52.56	53.36	47.64	50.63	48.16	56.67	56.74	57.70	56.99
FR 906 Proc. Gas FLOW TO VENTURI, mol/hr	18.81	19.09	19.84	20.29	19.37	19.87	19.90	19.95	19.41
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	0	0	6.11	10.63	0	4.64	4.66	2.22	0
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.0	2.86	2.95	3.07	3.07	3.19	3.20	3.20	3.19
FR 937 ATOMIZING N ₂ , mol/hr	1.88	1.72	1.72	1.56	1.58	1.81	1.84	1.75	1.85
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	43.43	44.58	42.29	49.15	38.86	51.44	53.72	50.29	43.43
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	77.21	76.63	64.27	65.01	72.0	72.53	72.65	73.45	76.56
FR 9304 Steam TO VENTURI, mol/hr	8.70	8.97	4.38	5.50	7.93	7.93	7.93	7.94	7.93
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	4.47	4.37	4.47	4.37	6.55	6.03	6.34	6.76	6.03
NITROGEN PURGE TO REACTOR, mol/hr	8.46	8.46	8.46	8.81	8.81	8.57	8.05	7.92	7.86
BED DENSITY, lb/ft ³	45.19	42.53	38.55	38.55	42.53	37.22	37.22	38.55	39.88
BED HEIGHT, ft	3.06	3.66	3.66	4.18	3.25	4.87	5.18	4.48	3.90
REACTOR GAS COMPOSITION (dry), %									
N ₂	69.90		51.50		74.30	51.60		54.90	
CO	8.10		22.10		3.90	16.30		16.50	
CO ₂	12.90		9.60		13.90	9.90		9.10	
H ₂	8.20		16.10		7.00	20.90		18.30	
CH ₄	0.11		0.12		0.24	0.76		0.56	
H ₂ S	0.00		0.00		0.00	0.00		0.20	
Coke or Coal FEED RATE, (wet) (avg), lb/hr	Coke 725	725	725	900	900	Coal 1012.5	1012.5	1012.5	1012.5
C-7 BED DISCHARGE RATE, lb/hr	174.1	186.5	79.60	79.60	79.60	94.4	43.9	43.9	178.3
C-4 ASH DISCHARGE RATE, lb/hr	13.0	16.0	7.7	34.3	41.3	18.7	14.7	16.7	9.5
2nd STAGE FINES REMOVAL RATE, lb/hr	116	116	116	137.9	137.9	139.3	139.3	139.3	139.3
3rd STAGE FINES REMOVAL RATE, lb/hr	10.4	10.4	10.4	10.4	10.4	9.0	9.0	9.0	9.0
FEED COMPOSITION, %:									
C									
ASH									
REACTOR BED COMPOSITION, %:									
C									
ASH	32.5	29.0	34.5	51.0	44.0	30.0	35.0	53.0	56.0
DISCHARGED BED COMPOSITION, %:									
C									
ASH									
ASH COMPOSITION, %:									
C									
ASH							86.0		
2nd STAGE FINES COMPOSITION, %:									
C									
ASH									
3rd STAGE FINES COMPOSITION, %:									
C									
ASH									

Table 3, Part 4. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

Date 6/2-6/3/78 RUN NUMBER 124

TIME	1200	1400	1600	1800	2000	2200	2400	6/3 0200	0400
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	382	362	343	317	317	324	275	288	293
TE 9206.2 GAS TEMP ENTERING GRID, °F	408	428	399	389	383	360	347	376	377
TE 9206.3 BED TEMP, °F	1873	1876	1874	1862	1843	1828	1829	1870	1887
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1779	1784	1784	1774	1762	1739	1749	1782	1809
FR 903 Stream FLOW TO GRID, mol/hr	13.95	13.96	15.51	15.51	19.33	19.49	19.38	13.96	14.34
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 Proc. Gas FLOW TO GRID, mol/hr	56.35	55.00	55.13	55.66	53.41	52.09	54.10	56.94	60.10
FR 906 Proc. Gas FLOW TO VENTURI, mol/hr	19.37	19.29	20.09	20.65	18.06	19.12	19.03	18.74	19.14
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	5.74	5.26	5.44	4.78	5.22	6.07	5.65	8.76	8.82
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.81	3.72	3.62	3.89	4.04	3.97	3.88	3.78	3.60
FR 937 ATOMIZING N ₂ , mol/hr	1.78	1.96	1.80	1.80	1.87	1.90	1.85	1.72	1.84
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	51.44	51.44	51.44	48.01	51.44	49.15	52.58	57.15	60.58
FR 9330 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	72.64	71.93	71.63	73.7	68.39	67.13	69.36	73.62	75.60
FR 9304 Steam TO VENTURI, mol/hr	7.93	8.50	8.01	6.94	2.36	4.45	6.91	6.59	6.71
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	6.97	6.34	3.12	2.60	3.64	3.54	4.68	4.16	4.26
NITROGEN PURGE TO REACTOR, mol/hr	10.15	10.15	10.30	10.30	10.30	10.52	11.38	11.19	11.14
BED DENSITY, lb/ft ³	38.55	33.23	37.22	33.23	31.90	34.56	31.90	30.57	27.91
BED HEIGHT, ft	3.88	4.68	4.18	3.55	3.79	3.17	3.97	4.80	4.64
REACTOR GAS COMPOSITION (dry), %									
N ₂	53.00		54.90				57.60		50.70
CO	13.00		17.00				14.60		18.50
CO ₂	7.70		8.10				8.90		7.00
H ₂	23.20		18.90				14.60		21.10
CH ₄	2.10		0.66				1.60		0.93
H ₂ S	0.40		0.00				0.00		0.00
Coal FEED RATE, (wet) (avg), lb/hr	1012.5	1012.5	1012.5	1012.5	1012.5	1012.5	1012.5	1275	1275
C-7 BED DISCHARGE RATE, lb/hr	244.7	121.8	179.6	250.4	313.0	40.6	357.4	251.2	289.0
C-4 ASH DISCHARGE RATE, lb/hr	10.0	2.7	2.0	2.0	2.3	4.3	3.3	2.0	2.3
2nd STAGE FINES REMOVAL RATE, lb/hr	139.3	139.3	139.3	139.3	139.3	139.3	139.3	147.5	147.5
3rd STAGE FINES REMOVAL RATE, lb/hr	9.0	9.0	9.0	9.0	9.0	9.0	9.0	11.2	11.2
FEED COMPOSITION, %:									
C									66.80
ASH									16.37
REACTOR BED COMPOSITION, %:									
C									41.80
ASH	62.5	50.0	50.0	65.0	50.0	59.0	50.0	43.0	52.17
DISCHARGED BED COMPOSITION, %:									
C									
ASH									
ASH COMPOSITION, %:									
C									25.70
ASH									66.45
2nd STAGE FINES COMPOSITION, %:									
C									72.10
ASH			20.5		23.0				23.62
3rd STAGE FINES COMPOSITION, %:									
C									65.00
ASH									29.43

Table 3, Part 5. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE 6/3/78 RUN NUMBER 124

TIME	0600	0800	1000	1200	1400	1600	1800	2000	2200
TE 9206.1									
GAS TEMP ENTERING VENTURI, °F	272	279	347	345	347	355	370	316	357
TE 9206.2									
GAS TEMP ENTERING GRID, °F	151	361	390	425	398	434	408	387	351
TE 9206.3									
BED TEMP, °F	1896	1893	1901	1898	1886	1900	1848	1834	1846
TE 9206.16									
GAS TEMP ENTERING QUENCHER, °F	1813	1812	1817	1802	1794	1801	1758	1748	1750
FR 903									
Steam FLOW TO GRID, mol/hr	14.26	14.32	15.26	14.33	14.33	14.33	14.26	15.28	13.92
FR 904									
FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905									
Proc. Gas FLOW TO GRID, mol/hr	59.90	59.55	60.86	59.95	60.27	59.54	53.44	51.72	46.51
FR 906									
Proc. Gas FLOW TO VENTURI, mol/hr	19.07	18.96	18.66	18.38	18.35	18.26	17.35	18.90	17.49
FR 909									
SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938									
OXYGEN TO AIR HEATER, mol/hr	8.81	8.62	8.64	8.61	8.62	8.52	8.15	3.45	3.90
FR 9303									
AUX. N ₂ TO REACTOR, mol/hr	3.80	3.88	3.95	3.88	3.86	3.87	4.11	4.28	4.20
FR 937									
ATOMIZING N ₂ , mol/hr	1.84	1.84	1.94	1.92	2.01	1.94	1.96	2.07	2.06
FR 918 (1st stage quencher)									
RATE OF QUENCH WATER, mol/hr	56.01	53.72	54.86	51.44	59.44	59.44	51.44	44.58	42.29
FR 9350 (2nd stage quencher)									
RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995									
AIR TO AIR HEATER, mol/hr	72.27	73.24	73.34	73.45	73.24	72.73	71.71	66.43	58.32
FR 9304									
Steam TO VENTURI, mol/hr	6.61	6.86	7.01	6.86	6.23	6.67	6.71	6.73	6.27
FR 9361									
N ₂ TO BED DISCHARGE, mol/hr	4.37	3.95	2.70	1.56	3.33	3.12	3.12	3.12	3.74
NITROGEN PURGE TO REACTOR, mol/hr	11.0	11.02	11.14	11.10	11.14	11.09	11.09	11.01	11.01
BED DENSITY, lb/ft ³	27.91	27.91	26.58	25.25	25.25	26.58	25.25	25.25	25.25
BED HEIGHT, ft	5.05	4.23	4.55	4.56	4.56	4.44	4.90	5.02	5.02
REACTOR GAS COMPOSITION (dry), %									
N ₂				47.50		50.90		53.80	
CO				16.30		16.70		15.70	
CO ₂				10.30		7.50		8.10	
H ₂				22.30		22.60		19.00	
CH ₄				2.20		1.70		0.90	
H ₂ S				0.50		0.00		0.10	
Coal FEED RATE, (wet) (avg), lb/hr	1275	1275	1275	1275	1275	1275	1075	1075	1075
C-7 BED DISCHARGE RATE, lb/hr	128.6	335.6	328.0	479.6	266.7	396.6	401.0	440.0	225.5
C-4 ASH DISCHARGE RATE, lb/hr	2.0	1.3	2.3	3.0	2.7	3.3	3.7	4.0	3.7
2nd STAGE FINES REMOVAL RATE, lb/hr	147.5	147.5	147.5	147.5	147.5	147.5	127.2	127.2	127.2
3rd STAGE FINES REMOVAL RATE, lb/hr	11.2	11.2	11.2	11.2	11.2	11.2	1.1	1.1	1.1
FEED COMPOSITION, %: C						66.40			
ASH						16.43			
REACTOR BED COMPOSITION, %: C						48.3			
ASH	50.0	52.0	55.0	52.0	41.5	47.25	50.0	36.5	45.5
DISCHARGED BED COMPOSITION, %: C						47.10			
ASH						47.56			
ASH COMPOSITION, %: C						11.90			
ASH						78.66			
2nd STAGE FINES COMPOSITION, %: C						74.66			
ASH						21.16			
3rd STAGE FINES COMPOSITION, %: C						62.70			
ASH						31.78			

Table 3, Part 6. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE 6/3-6/4/78 RUN NUMBER 124

TIME	2400	6/4 0200	0400	0600	0800	1000	1200	1400	1600
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	350	359	351	350	350	362	372	395	378
TE 9206.2 GAS TEMP ENTERING GRID, °F	345	338	339	340	372	407	411	403	393
TE 9206.3 BED TEMP, °F	1844	1862	1860	1865	1884	1880	1865	1878	1882
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1748	1749	1755	1760	1770	1767	1757	1767	1773
FR 903 Steam FLOW TO GRID, mol/hr	13.92	13.94	13.94	13.89	13.95	13.95	14.09	14.17	14.09
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 Proc. Gas FLOW TO GRID, mol/hr	46.40	49.55	48.16	47.06	47.60	44.34	43.80	43.38	42.31
FR 906 Proc. Gas FLOW TO VENTURI, mol/hr	17.58	17.96	17.87	17.84	17.62	16.15	15.79	15.04	15.15
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	3.93	4.41	5.62	5.50	3.77	4.19	4.69	4.64	4.64
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	4.56	3.90	3.91	3.91	3.92	3.86	3.96	3.86	3.07
FR 937 ATOMIZING N ₂ , mol/hr	2.08	2.01	2.03	2.02	2.02	2.08	2.09	2.08	2.08
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	48.01	43.43	44.58	44.58	37.72	36.58	34.29	36.58	41.15
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	58.41	62.39	60.55	60.80	63.24	57.59	56.35	59.07	55.97
FR 9304 Steam TO VENTURI, mol/hr	6.55	5.84	6.48	6.50	6.67	6.71	5.61	6.70	7.19
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	3.12	3.33	3.33	3.33	2.29	2.50	0	3.12	2.70
NITROGEN PURGE TO REACTOR, mol/hr	11.05	11.37	11.32	11.14	10.95	11.14	11.14	10.92	10.92
BED DENSITY, lb/ft ³	22.60	21.27	21.27	21.27	19.94	22.60	19.94	21.27	22.60
BED HEIGHT, ft	5.10	6.10	6.29	6.23	5.05	4.46	4.33	4.20	4.01
REACTOR GAS COMPOSITION (dry), %									
N ₂			63.90		53.90		55.30		55.40
CO			14.40		16.80		15.20		16.80
CO ₂			5.70		9.00		7.50		7.60
H ₂			14.70		17.90		19.80		18.90
CH ₄			0.79		1.20		1.30		0.58
H ₂ S			0.10		0.30		0.00		0.30
Coal FEED RATE, (wet) (avg), lb/hr	1075	900	900	900	900	900	900	900	900
C-7 BED DISCHARGE RATE, lb/hr	215.1	147.5	207.1	295.5	262.5	176.0	275.7	86.5	161.0
C-4 ASH DISCHARGE RATE, lb/hr	3.7	2.0	3.7	1.7	2.3	3.0	2.3	3.7	4.7
2nd STAGE FINES REMOVAL RATE, lb/hr	127.2	82.5	82.5	82.5	82.5	82.5	82.5	82.5	82.5
3rd STAGE FINES REMOVAL RATE, lb/hr	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4
FEED COMPOSITION, %:									
ASH									
REACTOR BED COMPOSITION, %:									
ASH	43.0	43.0	41.0	45.0	40.0	55.5	47.0	47.0	54.5
DISCHARGED BED COMPOSITION, %:									
ASH									
ASH COMPOSITION, %:									
ASH									
2nd STAGE FINES COMPOSITION, %:									
ASH									26.0
3rd STAGE FINES COMPOSITION, %:									
ASH									

Table 3, Part 7. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE 6/4-6/5/78 RUN NUMBER 124

TIME	1800	2000	2200	2400	6/5 0200	0400	0600	0800	1000
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	411	364	271	340	345	318	311	322	309
TE 9206.2 GAS TEMP ENTERING GRID, °F	400	417	446	405	365	381	372	335	346
TE 9206.3 BED TEMP, °F	1883	1773	1819	1848	1854	1844	1833	1844	1839
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1782	1692	1710	1744	1756	1752	1738	1739	1739
FR 903 Steam FLOW TO GRID, mol/hr	17.62	16.36	16.36	16.36	18.71	19.76	19.37	14.72	14.72
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 Proc. Gas FLOW TO GRID, mol/hr	42.42	42.45	43.78	44.23	47.94	49.17	52.11	46.24	47.03
FR 906 Proc. Gas FLOW TO VENTURI, mol/hr	18.72	19.79	10.59	16.51	16.93	17.40	17.55	16.48	16.36
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	5.87	0	4.18	8.56	7.99	5.41	4.50	4.45	4.46
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.91	3.50	3.65	4.03	3.71	3.89	3.69	3.84	3.79
FR 937 ATOMIZING N ₂ , mol/hr	2.08	2.06	2.07	2.08	2.07	2.07	2.07	2.08	2.06
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	43.43	38.86	38.86	43.43	45.72	45.72	45.72	41.15	40.01
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 993 AIR TO AIR HEATER, mol/hr	57.43	66.37	53.46	57.76	59.57	64.30	64.30	61.18	59.72
FR 9304 Steam TO VENTURI, mol/hr	2.81	9.20	15.96	8.89	9.95	10.30	9.92	8.35	8.35
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	2.50	2.29	2.50	4.68	5.30	5.41	5.30	7.28	6.66
NITROGEN PURGE TO REACTOR, mol/hr	10.92	11.07	11.07	10.93	10.95	10.84	10.94	10.98	11.00
BED DENSITY, lb/ft ³	22.60	26.58	23.93	23.93	23.93	23.93	23.93	23.93	23.93
BED HEIGHT, ft	3.57	4.22	4.45	4.69	4.81	4.81	4.81	5.17	5.05
REACTOR GAS COMPOSITION (dry), %									
N ₂		66.40		55.80		58.90		60.10	
CO		10.40		14.40		14.60		14.50	
CO ₂		7.90		9.30		8.40		7.60	
H ₂		12.30		16.90		16.90		16.00	
CH ₄		1.50		0.52		0.64		0.72	
H ₂ S		0.00		0.00		0.00		0.00	
Coal FEED RATE, (wet) (avg), lb/hr	900	825	825	825	825	825	825	825	825
C-7 BED DISCHARGE RATE, lb/hr	278.1	247.1	235.0	134.8	224.0	101.0	134.3	134.1	195.6
C-4 ASH DISCHARGE RATE, lb/hr	4.0	0	1	1	5	0	0	0	0
2nd STAGE FINES REMOVAL RATE, lb/hr	82.5	85.4	85.4	85.4	85.4	85.4	85.4	85.4	85.4
3rd STAGE FINES REMOVAL RATE, lb/hr	6.1	6.1	6.1	6.1	6.1	6.1	6.1	6.1	6.1
FEED COMPOSITION, %: C									
ASH									
REACTOR BED COMPOSITION, %: C									
ASH	62.5	40.5	40.0	34.0	41.0	51.0	48.0	50.0	54.0
DISCHARGED BED COMPOSITION, %: C									
ASH									
ASH COMPOSITION, %: C									
ASH									
2nd STAGE FINES COMPOSITION, %: C									
ASH									
3rd STAGE FINES COMPOSITION, %: C									
ASH									

Table 3, Part 8. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE 6/5-6/6/78 RUN NUMBER 124

TIME	1200	1400	1600	1800	2000	2200	2400	6/6 0200	0400
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	323	325	302	283	259	252	250	250	250
TE 9206.2 GAS TEMP ENTERING GRID, °F	355	365	378	385	374	372	376	373	374
TE 9206.3 BED TEMP, °F	1842	1826	1833	1832	1821	1834	1830	1830	1832
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1736	1727	1733	1726	1726	1710	1706	1707	1709
FR 903 STEAM FLOW TO GRID, mol/hr	14.88	14.72	16.28	17.06	16.27	16.28	16.27	16.66	14.68
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr	0	0	0	0	0	0	0	0	0
FR 905 PROCESS GAS FLOW TO GRID, mol/hr	45.73	45.45	50.70	49.85	50.31	49.30	50.11	50.22	48.68
FR 906 PROCESS GAS FLOW TO VENTURI, mol/hr	16.30	16.20	16.35	11.43	11.98	12.00	11.85	11.91	11.85
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	4.30	4.33	3.89	3.86	3.57	4.15	4.23	4.15	4.28
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.90	3.89	3.90	3.86	3.71	3.86	3.85	3.86	3.06
FR 937 ATOMIZING N ₂ , mol/hr	2.08	2.08	2.08	2.06	1.86	1.99	2.08	2.01	2.00
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	43.43	43.43	43.43	44.58	53.72	44.58	44.58	46.86	42.29
FR 9330 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	60.06	59.71	64.45	64.45	61.52	59.64	59.81	59.47	59.47
FR 9304 STEAM TO VENTURI, mol/hr	8.69	9.28	9.44	14.46	13.87	14.77	14.36	14.36	14.36
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	6.24	4.99	4.37	3.95	3.74	3.12	2.91	2.91	2.91
NITROGEN PURGE TO REACTOR, mol/hr	11.41	11.57	11.07	11.07	11.07	11.07	10.99	10.94	11.01
BED DENSITY, lb/ft ³	23.93	23.93	22.60	23.93	23.93	22.60	23.93	22.60	22.60
BED HEIGHT, ft	5.05	4.69	3.95	4.21	4.33	4.59	4.57	4.08	4.33
REACTOR GAS COMPOSITION (dry), %									
N ₂	60.90		65.00		57.40		60.80		60.70
CO	13.20		11.80		13.30		12.90		14.10
CO ₂	8.00		7.00		8.60		8.50		8.40
H ₂	16.20		14.80		15.80		16.40		15.60
CH ₄	0.86		0.80		0.68		0.79		0.67
H ₂ S	0.30		0.20		0.10		0.20		0.10
COAL FEED RATE, (wet) (avg), lb/hr	825	825	825	785	785	785	785	785	785
C-7 BED DISCHARGE RATE, lb/hr	1024.1	321.7	312.8	240.8	203.0	156.3	173.2	273.7	251.7
C-4 ASH DISCHARGE RATE, lb/hr	0	1	0	0.3	0	0.7	0.3	1.5	0
2nd STAGE FINES REMOVAL RATE, lb/hr	85.4	85.4	18.60	18.60	18.60	77.6	77.6	77.6	77.6
3rd STAGE FINES REMOVAL RATE, lb/hr	6.1	6.1	11.7	11.7	11.7	11.7	11.7	11.7	11.7
FEED COMPOSITION, %:									
C									
ASH									
REACTOR BED COMPOSITION, %:									
C									
ASH	50.0	44.0	35.0	32.0	38.0	36.5	36.0	44.0	47.0
DISCHARGED BED COMPOSITION, %:									
C									
ASH									
ASH COMPOSITION, %:									
C									
ASH									
2nd STAGE FINES COMPOSITION, %:									
C									
ASH									
3rd STAGE FINES COMPOSITION, %:									
C									
ASH									

Table 3, Part 9. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE 6/6/78		RUN NUMBER 124								
TIME		0600	0800	1000	1200	1400	1600	1800	2000	2200
TE 9206.1 GAS TEMP ENTERING VENTURI, °F		231	254	267	278	286	288	293	291	278
TE 9206.2 GAS TEMP ENTERING GRID, °F		353	357	374	424	444	408	379	383	369
TE 9206.3 BED TEMP, °F		1833	1837	1835	1834	1842	1833	1826	1838	1817
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F		1708	1707	1694	1701	1702	1698	1692	1705	1692
FR 903 STEAM FLOW TO GRID, mol/hr		14.72	16.66	16.34	17.14	18.30	17.15	16.33	16.33	16.33
FR 904 FLOW TO GRID/HEATER/VENTURI, mol/hr		0	0	0	0	0	0	0	0	0
FR 905 PROCESS GAS FLOW TO GRID, mol/hr		49.91	50.57	48.62	48.99	48.97	50.28	50.43	50.83	50.69
FR 906 PROCESS GAS FLOW TO VENTURI, mol/hr		11.80	12.04	11.93	12.02	11.75	11.98	12.81	12.83	12.79
FR 909 SWEEP GAS (N ₂), mol/hr		0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr		4.30	4.18	2.95	2.78	2.83	2.83	2.76	3.40	3.08
FR 9303 AUX. N ₂ TO REACTOR, mol/hr		3.84	3.83	3.66	3.56	3.78	3.79	3.79	3.84	3.77
FR 937 ATOMIZING N ₂ , mol/hr		2.08	2.08	2.07	2.07	2.06	1.91	2.05	2.08	1.97
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr		41.15	44.58	42.29	43.43	43.43	41.15	44.58	44.58	41.15
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr		0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr		59.47	59.81	63.02	61.59	60.08	62.47	62.68	61.12	60.69
FR 9304 STEAM TO VENTURI, mol/hr		14.05	13.80	13.69	14.56	12.21	11.77	11.22	11.60	11.50
FR 9361 N ₂ TO BED DISCHARGE, mol/hr		3.12	3.12	1.35	1.56	6.24	4.99	6.45	3.95	3.95
NITROGEN PURGE TO REACTOR, mol/hr		10.99	11.09	11.09	11.09	11.09	11.09	11.09	11.09	11.09
BED DENSITY, lb/ft ³		22.60	21.27	22.60	21.27	23.93	23.93	22.60	23.93	22.60
BED HEIGHT, ft		4.46	4.74	3.82	4.06	3.85	3.97	4.33	3.97	4.14
REACTOR GAS COMPOSITION (dry), %										
N ₂						60.10	62.40		63.30	
CO						12.80	13.00		14.20	
CO ₂						8.50	8.30		8.40	
H ₂						17.20	15.00		12.90	
CH ₄						0.83	0.67		0.71	
H ₂ S						0.08	0.10		0.00	
COAL FEED RATE, (wet) (avg), lb/hr		785	785	785	785	785	785	785	785	785
C-7 BED DISCHARGE RATE, lb/hr		210.0	267.0	288.0	276.5	162.7	236.7	193.0	232.0	354.3
C-4 ASH DISCHARGE RATE, lb/hr		0	0	1.0	1.3	0	0	0.7	0	0
2nd STAGE FINES REMOVAL RATE, lb/hr		77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6
3rd STAGE FINES REMOVAL RATE, lb/hr		11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7
FEED COMPOSITION, %: C										
ASH										
REACTOR BED COMPOSITION, %: C										
ASH		45.0	38.0	36.0	39.0	47.0	41.0	43.5	39.0	53.5
DISCHARGED BED COMPOSITION, %: C										
ASH										
ASH COMPOSITION, %: C										
ASH										
2nd STAGE FINES COMPOSITION, %: C										
ASH							26.0			
3rd STAGE FINES COMPOSITION, %: C										
ASH							30.0			

Table 3, Part 10. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE 6/6-6/7/78 RUN NUMBER 124

TIME	2400	6/7 0200	0400	0600	0800	1000	1200	1600	1800
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	284	301	295	290	290	294	300	346	347
TE 9206.2 GAS TEMP ENTERING GRID, °F	357	561	549	543	550	545	576	465	418
TE 9206.3 BED TEMP, °F	1829	1824	1806	1816	1804	1835	1816	1827	1842
TE 9206.1b GAS TEMP ENTERING QUENCHER, °F	1689	1689	1679	1656	1667	1676	1671	1617	1675
FR 903 STEAM FLOW TO GRID, mol/hr	16.32	14.66	14.66	14.72	13.88	13.85	13.88	14.79	18.31
FR 904 STEAM FLOW TO HEATER, mol/hr	0	44.47	44.47	45.76	45.56	44.47	44.44	0	0
FR 905 PROCESS GAS FLOW TO GRID, mol/hr	49.96	61.21	48.34	49.47	50.83	45.05	48.90	48.37	47.20
FR 906 PROCESS GAS FLOW TO VENTURI, mol/hr	12.89	13.50	13.67	13.68	13.68	16.31	13.29	18.38	19.16
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	6.08	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	3.44	19.29	20.13	19.73	19.33	19.89	20.75	3.52	0
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.84	3.89	3.90	3.84	3.86	3.94	3.98	3.86	3.79
FR 937 ATOMIZING N ₂ , mol/hr	2.00	1.94	1.97	2.01	2.02	1.97	2.00	2.00	2.01
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	51.44	46.86	49.15	52.58	45.72	43.43	45.72	36.58	37.72
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	61.61	0	0	0	0	0	0	75.59	76.05
FR 9304 STEAM TO VENTURI, mol/hr	11.28	11.52	11.85	11.34	11.36	8.99	11.22	11.70	9.70
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	4.37	3.95	4.06	4.37	3.74	4.26	3.33	3.85	5.30
NITROGEN PURGE TO REACTOR, mol/hr	11.09	10.94	11.09	10.96	10.92	10.92	10.90	10.76	10.76
BED DENSITY, lb/ft ³	22.60	22.60	22.60	19.94	19.94	19.94	21.27	33.23	35.89
BED HEIGHT, ft	4.46	4.46	4.59	4.40	4.62	4.91	4.20	3.81	4.01
REACTOR GAS COMPOSITION (dry), %									
N ₂	30.50		63.80			31.90	27.50		
CO	18.90		12.80			19.20	17.80		
CO ₂	15.30		7.10			15.50	13.80		
H ₂	33.50		15.10			31.40	37.80		
CH ₄	1.70		0.74			1.40	2.30		
H ₂ S	0.00		0.00			0.40	0.60		
FEED RATE, (wet) (avg), lb/hr	COAL+ 785	1020	1020	1020	1020	1020	1020	COKE+ 750	750
C-7 BED DISCHARGE RATE, lb/hr	245.6	142.8	333.5	302.0	338.0	253.0	169.0	250.0	336.7
C-4 ASH DISCHARGE RATE, lb/hr	1.7	2.0	1.0	1.7	1.0	2.3	0	2.3	1.0
2nd STAGE FINES REMOVAL RATE, lb/hr	77.6	87.1	87.1	87.1	87.1	87.1	87.1	83.1	83.1
3rd STAGE FINES REMOVAL RATE, lb/hr	6.1	6.1	6.1	6.1	2.7	2.7	2.7	2.7	2.7
FEED COMPOSITION, %:									
ASH									
REACTOR BED COMPOSITION, %:									
ASH	41.0	55.0	63.0	56.0	38.0	41.0	43.0	42.0	37.0
DISCHARGED BED COMPOSITION, %:									
ASH									
ASH COMPOSITION, %:									
ASH									
2nd STAGE FINES COMPOSITION, %:									
ASH			28.0				29.0		
3rd STAGE FINES COMPOSITION, %:									
ASH			32.0				35.0		

Table 3, Part 11. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

* DATE 6/7-6/8/78 RUN NUMBER 124

TIME	2000	2200	2400	6/8 0200	0400	0600	0800	1000	1200
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	336	313	300	286	296	286	293	286	298
TE 9206.2 GAS TEMP ENTERING GRID, °F	304	303	498	487	495	504	507	481	480
TE 9206.3 BED TEMP, °F	1844	1802	1797	1824	1809	1810	1816	1825	1826
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1683	1661	1653	1669	1660	1658	1660	1669	1669
FR 903 STEAM FLOW TO GRID, mol/hr	18.21	17.16	16.89	14.59	14.62	14.96	14.96	14.55	14.97
FR 904 STEAM FLOW TO HEATER, mol/hr	0	57.47	57.68	45.58	48.92	48.89	45.58	45.50	46.73
FR 905 PROCESS GAS FLOW TO GRID, mol/hr	48.43	52.79	57.71	46.84	48.30	48.26	47.57	47.29	46.35
FR 906 PROCESS GAS FLOW TO VENTURI, mol/hr	19.22	18.37	19.01	16.96	21.23	17.05	16.52	17.21	17.21
FR 909 SWEEP GAS (N ₂), mol/hr	0	0	0	0	0	0	0	0	0
FR 938 OXYGEN TO AIR HEATER, mol/hr	5.37	19.88	20.61	20.80	19.72	19.39	21.24	21.60	21.05
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.80	3.82	3.82	3.77	3.81	3.82	3.80	3.83	3.84
FR 937 ATOMIZING N ₂ , mol/hr	1.90	1.94	1.93	1.79	1.94	1.93	1.93	1.93	1.96
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	45.72	53.72	59.44	49.15	52.58	52.58	64.01	66.29	51.44
FR 9330 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr	77.20	0	0	0	0	0	0	0	0
FR 9304 STEAM TO VENTURI, mol/hr	9.70	10.80	10.80	11.20	10.80	10.79	10.80	10.78	10.80
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	3.43	5.82	6.55	6.03	5.72	5.82	5.62	5.51	4.78
NITROGEN PURGE TO REACTOR, mol/hr	10.76	10.72	10.94	11.06	11.21	11.03	10.97	10.96	10.96
BED DENSITY, lb/ft ³	34.56	31.90	31.90	29.24	29.24	29.24	27.91	26.58	26.58
BED HEIGHT, ft	4.17	4.51	4.24	4.82	4.73	4.82	5.16	5.63	5.20
REACTOR GAS COMPOSITION (dry), %									
N ₂	63.30		52.40				28.65		
CO	10.80		15.30				18.45		
CO ₂	9.30		17.50				15.80		
H ₂	14.70		31.80				34.85		
CH ₄	1.20		2.40				1.75		
H ₂ S	0.10		0.40				0.30		
COAL FEED RATE, (wet) (avg), lb/hr	937.5	937.5	937.5	1036	1036	1036	1036	1036	1036
C-7 BED DISCHARGE RATE, lb/hr	291.7	206.5	142.6	123.0	159.9	197.0	168.5	168.5	162.0
C-4 ASH DISCHARGE RATE, lb/hr	0.3	0	0	2.0	0	0	1.7	0.7	0.7
2nd STAGE FINES REMOVAL RATE, lb/hr	127.3	127.3	127.3	118.3	118.3	118.3	118.3	118.3	118.3
3rd STAGE FINES REMOVAL RATE, lb/hr	2.7	5.5	5.5	5.5	5.5	0	0	5.0	5.0
FEED COMPOSITION, %:									
C					63.70				68.90
ASH					19.49				14.39
REACTOR BED COMPOSITION, %:									
C					48.30				40.00
ASH	39.0	39.5	--	56.0	48.87	48.0	54.5	42.5	45.89
DISCHARGED BED COMPOSITION, %:									
C									30.00
ASH									65.59
ASH COMPOSITION, %:									
C									
ASH									
2nd STAGE FINES COMPOSITION, %:									
C					65.60				66.90
ASH					30.82				29.80
3rd STAGE FINES COMPOSITION, %:									
C					61.40				61.90
ASH					34.22				34.38

Table 3, Part 12. ASH AGGLOMERATING GASIFIER - DATA SUMMARY
FOR RUN 124

DATE	6/8-6/9/78	RUN NUMBER	124									
TIME				1400	1600	1800	2000	2200	2400	6/9 0200	0400	0600
TE 9206.1 GAS TEMP ENTERING VENTURI, °F				308	319	316	318	313	318	317	367	325
TE 9206.2 GAS TEMP ENTERING GRID, °F				489	494	490	500	496	504	495	472	488
TE 9206.3 BED TEMP, °F				1822	1804	1805	1800	1821	1817	1814	1798	1813
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F				1661	1644	1630	1640	1625	1607	1610	1604	1616
FR 903 STEAM FLOW TO GRID, mol/hr				14.93	13.42	13.58	13.49	13.43	13.46	13.43	13.43	13.82
FR 904 FLOW TO HEATER, mol/hr				45.50	45.56	45.58	46.67	44.47	45.58	45.56	46.70	46.70
FR 905 PROCESS FLOW TO GRID, mol/hr				47.00	46.90	46.17	46.26	45.48	44.12	44.81	38.69	46.69
FR 906 PROCESS FLOW TO VENTURI, mol/hr				17.02	17.51	17.24	17.43	16.98	16.53	16.34	23.36	16.99
FR 909 SWEEP GAS (N ₂), mol/hr				0	0	0	0	0	0	8.66	6.26	5.58
FR 938 OXYGEN TO AIR HEATER, mol/hr				21.02	19.97	20.78	19.03	20.08	18.84	19.06	19.12	21.57
FR 9303 AUX. N ₂ TO REACTOR, mol/hr				3.96	3.97	4.02	3.99	3.98	3.90	3.63	3.92	3.94
FR 937 ATOMIZING N ₂ , mol/hr				2.00	1.98	1.98	1.99	2.00	1.96	1.90	2.04	1.96
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr				51.44	52.58	51.44	51.44	46.86	57.15	45.72	51.44	49.15
FR 9350 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr				0	0	0	0	0	0	0	0	0
FR 995 AIR TO AIR HEATER, mol/hr				0	0	0	0	0	0	0	0	0
FR 9304 STEAM TO VENTURI, mol/hr				10.43	10.31	10.31	10.45	8.78	8.72	9.19	9.62	9.41
FR 9361 N ₂ TO BED DISCHARGE, mol/hr				5.20	3.74	3.85	3.95	4.37	3.95	3.74	4.47	4.78
NITROGEN PURGE TO REACTOR, mol/hr				11.03	11.03	10.28	10.28	10.28	10.27	10.69	10.25	10.70
BED DENSITY, lb/ft ³				29.24	31.90	30.57	31.90	25.25	23.93	26.58	29.24	25.25
BED HEIGHT, ft				4.92	4.69	4.52	4.29	5.02	3.85	4.55	3.64	4.22
REACTOR GAS COMPOSITION (dry), %												
N ₂					25.60		28.20		27.10		29.60	
CO					16.30		16.50		16.10		16.20	
CO ₂					17.80		17.40		18.80		18.40	
H ₂					35.50		34.90		34.00		33.20	
CH ₄					2.20		2.10		2.80		1.80	
H ₂ S					0.60		0.60		0.80		0.60	
COAL FEED RATE, (wet) (avg), lb/hr				1036	1036	1036	1036	1036	1036	956.3	956.3	956.3
C-7 BED DISCHARGE RATE, lb/hr				245.0	265.3	298.7	249.7	249.3	79.5	87.6	190.6	208.5
C-4 ASH DISCHARGE RATE, lb/hr				3.0	3.3	2.3	0	3.0	0	1.0	36.0	0
2nd STAGE FINES REMOVAL RATE, lb/hr				118.3	118.3	118.3	118.3	118.3	309.3	309.3	24.8	24.8
3rd STAGE FINES REMOVAL RATE, lb/hr				5.0	5.0	5.0	5.0	5.0	5.0	56.7	78.9	78.9
FEED COMPOSITION, %: C							65.60					
ASH							18.32					
REACTOR BED COMPOSITION, %: C							49.26					
ASH				56.0	66.0	61.0	47.42	58.0	60.5	61.0	66.0	64.0
DISCHARGED BED COMPOSITION, %: C												
ASH												
ASH COMPOSITION, %: C												
ASH												
2nd STAGE FINES COMPOSITION, %: C							67.80					
ASH							28.54		31.0			
3rd STAGE FINES COMPOSITION, %: C							61.50					
ASH							35.02					

Table 3, Part 13. ASH AGGLOMERATING GASIFIER — DATA SUMMARY
FOR RUN 124

DATE 6/9/78 RUN NUMBER 124

TIME	0800	1000	1400	1600	1800				
TE 9206.1 GAS TEMP ENTERING VENTURI, °F	332	336	408	158	216				
TE 9206.2 GAS TEMP ENTERING GRID, °F	512	534	347	187	234				
TE 9206.3 BED TEMP, °F	1810	1782	1849	1868	1610				
TE 9206.16 GAS TEMP ENTERING QUENCHER, °F	1610	1594	1599	1607	1504				
FR 903 STEAM FLOW TO GRID, mol/hr	13.85	13.85	15.88	24.67	16.42				
FR 904 STEAM FLOW TO HEATER, mol/hr	44.47	44.47	0	0	74.80				
FR 905 PROCESS FLOW TO GRID, mol/hr	45.90	45.05	55.29	44.37	50.52				
FR 906 PROCESS FLOW TO VENTURI, mol/hr	16.77	16.31	23.36	24.01	17.37				
FR 909 SWEEP GAS (N ₂), mol/hr	6.02	6.08	6.50	6.30	5.06				
FR 938 OXYGEN TO AIR HEATER, mol/hr	20.33	19.89	0	5.23	20.52				
FR 9303 AUX. N ₂ TO REACTOR, mol/hr	3.91	3.94	3.91	3.84	3.84				
FR 937 ATOMIZING N ₂ , mol/hr	1.96	1.97	2.08	1.93	1.87				
FR 918 (1st stage quencher) RATE OF QUENCH WATER, mol/hr	48.01	43.43	34.29	46.86	69.72				
FR 9330 (2nd stage quencher) RATE OF QUENCH WATER, mol/hr	0	0	0	0	22.39				
FR 995 AIR TO AIR HEATER, mol/hr	0	0	89.11	90.89	0				
FR 9304 STEAM TO VENTURI, mol/hr	8.99	8.99	7.73	10.51	9.21				
FR 9361 N ₂ TO BED DISCHARGE, mol/hr	3.74	0.83	2.70	5.51	8.01				
NITROGEN PURGE TO REACTOR, mol/hr	10.95	10.97	10.53	10.39	10.45				
BED DENSITY, lb/ft ³	23.93	22.60	35.89	39.88	39.88				
BED HEIGHT, ft	4.57	5.10	4.33	3.97	4.51				
REACTOR GAS COMPOSITION (dry), %				74.60					
N ₂				5.30					
CO				12.70					
CO ₂				6.60					
H ₂				0.12					
CH ₄				0.10					
H ₂ S									
COAL OR COKE FEED RATE, (wet) (avg), lb/hr	COAL- 956.3	956.3	COKE+ 750	750	750				
C-7 BED DISCHARGE RATE, lb/hr	247.0	215.0	260.0	194.6	65.1				
C-4 ASH DISCHARGE RATE, lb/hr	3.0	1.7	0	0	0				
2nd STAGE FINES REMOVAL RATE, lb/hr	24.8	24.8	0	0	0				
3rd STAGE FINES REMOVAL RATE, lb/hr	78.9	78.9	102.6	102.6	102.6				
FEED COMPOSITION, %: C									
ASH									
REACTOR BED COMPOSITION, %: C									
ASH	65.5	46.5	39.0	43.0					
DISCHARGED BED COMPOSITION, %: C									
ASH									
ASH COMPOSITION, %: C									
ASH									
2nd STAGE FINES COMPOSITION, %: C									
ASH									
3rd STAGE FINES COMPOSITION, %: C									
ASH									

Table 4. ASH AGGLOMERATING GASIFIER OPERATING CONDITIONS
SUMMARY FOR RUN 124
(6/3/78; 0200-1600 Hours)

Run No. <u>124</u> Date <u>6/3/78</u> Time Period <u>0200 to 1600</u>		
Venturi Diameter <u>3</u> inches Feed Material <u>W. Ky. No. 9 Coal</u>		
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>		
1	Average Fluidized Bed Temperature, °F	1891
2	Average Reactor Pressure, psig	9.2
3	Venturi Pressure Drop, psi	0.56
4	Superficial Velocity, ft/s	4.88
5	Freeboard Velocity, ft/s	6.61
6	Venturi Velocity, ft/s	48.82
7	Grid Gas, mol/hr	74.0
8	Grid Gas Composition: %N ₂	56.92
9	%O ₂	23.64
10	%Steam	19.44
11	Venturi Gas, mol/hr	25.4
12	Venturi Gas Composition: %N ₂	52.03
13	%O ₂	21.62
14	%Steam	26.35
15	Fluidized Bed Density, lb/ft ³	27.25
16	Fluidized Bed Height, ft	4.60
17	Feed Composition, <u>1275.0</u> lb/hr, %C/%Ash	66.60/16.40
18	Average Fluidized Bed Composition, <u>-16.8</u> lb/hr, %C/%Ash	45.05/49.71
19	Bed Discharge Composition, <u>309.4</u> lb/hr, %C/%Ash	47.10/47.56
20	Venturi Discharge Composition, <u>2.4</u> lb/hr, %C/%Ash	18.80/72.56
21	2nd Stage Fines Composition, <u>147.5</u> lb/hr, %C/%Ash	73.35/22.39
22	3rd Stage Fines Composition, <u>11.2</u> lb/hr, %C/%Ash	63.85/30.61
23	Carbon/Oxygen (Weight Basis)	1.12
24	Carbon/Steam (Weight Basis)	2.18

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	ASH	CARBON	NITROGEN	OXYGEN	HYDROGEN	Btu/hr
Feed W. Ky #9 Coal	203.66	827.07	16.39	105.84	61.80	14,902,821
To Venturi	0	0	369.88	282.72	13.38	168,409
To Grid	0	0	1179.92	790.24	28.78	428,209
To Reactor	0	0	107.24	0	0	0
Purges	0	0	406.84	0	0	0
TOTAL INPUT, lb/hr	203.66	827.07	2080.27	1178.80	103.96	15,499,439
Wall Losses	0	0	0	0	0	250,000
Venturi Discharge	1.74	0.45	0	0	0	11,196
Bed Discharge	147.15	145.73	1.49	0	0.87	2,400,230
2nd Stage Fines	33.03	108.19	0.07	0	0.62	1,710,932
3rd Stage Fines	3.43	7.15	0.05	0	0.05	116,060
Bed Inventory Change	-8.34	-7.56	-0.04	0	-0.04	-134,285
Reactor Gas	0	552.36	2077.88	1086.56	102.44	11,462,708
TOTAL OUTPUT, lb/hr	177.01	806.32	2080.25	1086.56	103.94	15,816,841
NET (OUTPUT-INPUT)	-26.65	-20.75	0	-92.24	0	317,402
% Balance, Output/Input	86.91	97.49	100	92.18	100	102.0

Figure 3. MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 124
(6/3/78; 0200-1600 Hours)

Table 5. ASH AGGLOMERATING GASIFIER OPERATING CONDITIONS
 SUMMARY FOR RUN 124
 (6/8/78; 0200-2200 Hours)

Run No. <u>124</u> Date <u>6/8/78</u> Time Period <u>0200</u> to <u>2200</u>		
Venturi Diameter <u>3</u> inches Feed Material <u>W. Ky No. 9 Coal</u>		
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>		
1	Average Fluidized Bed Temperature, °F	1815
2	Average Reactor Pressure, psig	7.66
3	Venturi Pressure Drop, psi	0.35
4	Superficial Velocity, ft/s	5.28
5	Freeboard Velocity, ft/s	5.97
6	Venturi Velocity, ft/s	56.93
7	Grid Gas, mol/hr	61.27
8	Grid Gas Composition: %N ₂	0
9	%O ₂	23.47
10	%Steam	76.53
11	Venturi Gas, mol/hr	27.99
12	Venturi Gas Composition: %N ₂	0
13	%O ₂	19.15
14	%Steam	80.85
15	Fluidized Bed Density, lb/ft ³	28.88
16	Fluidized Bed Height, ft	4.89
17	Feed Composition, <u>1036.0</u> lb/hr, %C/%Ash	66.07/17.40
18	Average Fluidized Bed Composition, <u>6.61</u> lb/hr, %C/%Ash	45.83/50.73
19	Bed Discharge Composition, <u>207.9</u> lb/hr, %C/%Ash	30.00/65.59
20	Venturi Discharge Composition, <u>1.52</u> lb/hr, %C/%Ash	0/100
21	2nd Stage Fines Composition, <u>118.3</u> lb/hr, %C/%Ash	66.77/29.72
22	3rd Stage Fines Composition, <u>5.75</u> lb/hr, %C/%Ash	61.60/34.54
23	Carbon/Oxygen (Weight Basis)	1.05
24	Carbon/Steam (Weight Basis)	0.53

Figure 4. MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 124
(6/8/78; 0200-2200 Hours)

	ASH	CARBON	NITROGEN	OXYGEN	HYDROGEN	Btu/hr
Feed W. Ky #9 Coal	174.86	663.95	12.96	89.16	47.94	11,916,341
To Venturi	0	0	0	533.60	45.26	259,361
To Grid	0	0	0	1223.84	93.78	768,930
To Reactor	0	0	108.92	0	0	0
Purges	0	0	442.12	0	0	0
TOTAL INPUT, lb/hr	174.86	663.95	564.00	1846.6	186.98	12,944,632
Wall Losses	0	0	0	0	0	250,000
Venturi Discharge	1.52	0	0	0	0	978
Bed Discharge	136.36	62.37	0.37	0	0.40	1,063,421
2nd Stage Fines	35.16	78.99	0.49	0.12	0.58	1,247,598
3rd Stage Fines	1.99	3.54	0.02	0.02	0.03	56,327
Bed Inventory Change	-3.35	-3.03	-0.02	0	-0.02	-49,181
Reactor Gas	0	438.72	563.08	1721.28	185.98	10,045,862
TOTAL OUTPUT, lb/hr	171.68	580.59	563.94	1721.42	186.97	12,615,005
NET (OUTPUT-INPUT)	-3.18	-83.36	0	-125.18	0	-329,627
% Balance, Output/Input	98.18	87.44	100	93.20	100	97.45

Table 6. ASH AGGLOMERATING GASIFIER OPERATING CONDITIONS
SUMMARY FOR RUN 122
(3/29/78; 1200-2400 Hours)

Run No. <u>122</u> Date <u>3/29/78</u> Time Period <u>1200 to 2400</u>		
Venturi Diameter <u>3</u> inches Feed Material <u>W. Ky. No. 9 Coal</u>		
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>		
1	Average Fluidized Bed Temperature, °F	1792
2	Average Reactor Pressure, psig	11.58
3	Venturi Pressure Drop, psi	0.57
4	Superficial Velocity, ft/s	3.41
5	Freeboard Velocity, ft/s	4.62
6	Venturi Velocity, ft/s	52.74
7	Grid Gas, mol/hr	134.42
8	Grid Gas Composition: %N ₂	57.51
9	%O ₂	15.28
10	%Steam	27.21
11	Venturi Gas, mol/hr	30.9
12	Venturi Gas Composition: %N ₂	76.49
13	%O ₂	20.33
14	%Steam	3.18
15	Fluidized Bed Density, lb/ft ³	24.83
16	Fluidized Bed Height, ft	3.95
17	Feed Composition, <u>1650.0</u> lb/hr, %C/%Ash	68.50/15.77
18	Average Fluidized Bed Composition, <u> </u> lb/hr, %C/%Ash	47.80/47.84
19	Bed Discharge Composition, <u>277.0</u> lb/hr, %C/%Ash	41.00/54.45
20	Venturi Discharge Composition, <u>23.10</u> lb/hr, %C/%Ash	6.30/85.29
21	2nd Stage Fines Composition, <u>365.9</u> lb/hr, %C/%Ash	70.60/25.21
22	3rd Stage Fines Composition, <u>45.1</u> lb/hr, %C/%Ash	69.80/27.44
23	Carbon/Oxygen (Weight Basis)	1.29
24	Carbon/Steam (Weight Basis)	1.62

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.4	10.7	18.8	19.4	15.9	13.6	9.5	4.8	6.9							
Fluidized Bed	0.1	4.3	18.6	25.5	23.4	21.0	6.2	0.8	0.1							
Discharged Bed	0	2.6	12.8	22.8	24.5	24.3	9.4	2.6	1.0							
Ash	1.3	30.3	45.8	19.0	1.2	0.8	--	--	1.6							
2nd Stage Fines					0.5	6.4	23.9	--	--	13.8	4.2	9.7	26.6	11.1	3.5	0.3
3rd Stage Fines						0.2	1.7	--	--	1.9	0.6	5.7	24.9	20.5	23.5	21.0

Figure 5. MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 122
(3/29/78; 1200-2400 Hours)

	ASH	CARBON	NITROGEN	OXYGEN	HYDROGEN	Btu/hr
Feed	254.22	1104.25	20.80	98.50	74.02	19,528,374
To Venturi	0	0	661.92	216.64	1.96	66,900
To Grid	0	0	2164.40	1242.56	73.16	983,602
To Reactor	0	0	71.12	0	0	0
Purges	0	0	389.48	0	0	0
TOTAL INPUT, lb/hr	254.22	1104.25	3307.72	1557.70	149.14	20,578,876
Wall Losses	0	0	0	0	0	250,000
Venturi Discharge	19.70	1.46	0	0	0	45,879
Bed Discharge	150.85	113.59	0	0	0	1,840,489
2nd Stage Fines	92.24	258.33	0	0	0	4,068,918
3rd Stage Fines	12.38	31.48	0	0	0	490,070
Bed Inventory Change	--	-4.05	0	0	0	-31,033
Reactor Gas	0	550.32	3,307.64	1610.72	149.14	11,648,405
TOTAL OUTPUT, lb/hr	275.17	951.13	3,307.64	1610.72	149.14	18,312,728
NET (OUTPUT-INPUT)	+20.95	-153.12	0	53.02	0	-2,266,148
% Balance, Output/Input	108.24	86.13	100	103.40	100	89

Figure 6. PHOTOMICROGRAPH OF CROSS-SECTION OF A
SHALE PARTICLE WITH AN IRON-RICH COATING
FROM A RUN 122 C-4 DISCHARGE

Very few particles are composed of a continuous iron-rich low-melting phase containing small unmelted particles of ash.

The iron-rich low-melting coating of the particles contains mainly crystals of calcium aluminosilicate, iron silicate, and ferrous sulfide, with lesser amounts of calcium iron silicate and magnetite (Figure 8). These compounds are the same as those found in agglomerates from the Illinois coal (see February work report), although the crystal habit pattern is different in that very few calcium aluminosilicate blades are present. Sintering tests on some of these particles in a 60:40 carbon monoxide/carbon dioxide mixture showed that the coating melts at $1945 \pm 10^{\circ}\text{F}$ and forms a neck between adjacent particles. At 1915°F the particles showed only slight adhesion. Similar tests on agglomerates from coke have shown that the iron-rich phase melts at $1960^{\circ} \pm 20^{\circ}\text{F}$.

Figure 7. PORTION OF THE 6-12 MESH FRACTION OF A RUN C-4 DISCHARGE
(The particles have been mounted
and sectioned petrographically.)

Figure 8. CRYSTALS IN THE IRON-RICH COATING FOUND ON SOME PARTICLES IN RUN 122 C-4 DISCHARGES

The uniformity within and the diversity among particles of the aluminosilicate- or shale-derived parts of the C_4 discharge particles have been noted. We have considered that each particle might have been derived from a single particle of coal or shale in the coal. We ashed a large (400 gram) sample of the feed coal at a relatively low furnace temperature (1100°F) so that little or no sintering would occur. The sieve analysis of this ash is shown in Table 7 along with results on two bed discharge samples that were ashed at higher temperatures. The particle size distribution of the coal ash, which showed 38% coarser than 40 mesh, indicates that a large part of the particles found in the venturi and side (bed) discharges originates from single particles of coal or shale in the coal. The same conclusion is indicated by the results of an experiment in which a +40 mesh fraction of the coal was subjected to a float-sink separation at a density of 1.6 g/cm^3 . Of the total ash in the original coal, one-third was found in the sink sample.

Table 7. PARTICLE SIZE OF ASH FROM KENTUCKY COAL
AND GASIFIER BED DISCHARGE

	<u>Coal</u>	<u>Bed Discharge</u>	
Time of Sampling	--	1230	1830
Date of Sampling	--	3/30/78	3/30/78
Furnace Temperature, °F	1100	1400	1400
Sieve No., U.S.S., wt %			
3-1/2	3.2	0.0	0.3
6	6.5	5.1	4.0
12	10.7	11.9	13.3
20	9.4	18.0	25.0
40	8.4	28.8	38.8
70	9.6	14.1	10.0
140	12.5	5.4	2.1
270	18.4	4.9	1.5
PAN	21.3	11.7	5.0

We have also ashed some coarse (+6 mesh) particles of the coal spread on a plate so that ash from a single particle can be observed. We found that ash from a single particle is typically uniform in color, but varies greatly from particle to particle. This accounts for the particle-to-particle diversity of the discharge particles.

Elemental composition and fusibility of two samples of the Kentucky coal are shown in Table 8.

Table 8. ELEMENTAL ANALYSIS AND FUSIBILITY OF
ASH OF COAL FROM WESTERN KENTUCKY COAL
(Providence No. 1 Mine)

Elemental Analysis	Sampled at the Mine	IGT Sample
	wt %	
SiO_2	46.9	45.1
Al_2O_3	18.7	17.4
Fe_2O_3	21.0	21.2
TiO_2	0.8	0.8
CaO	6.0	5.4
MgO	1.1	1.1
Na_2O	0.6	0.7
K_2O	2.5	2.4
Ash Fusibility Temperature, °F		
Reducing Atmosphere		
IT	2025	1900
ST	2155	2110
HT	2295	2160
FT	2455	2360
Oxidizing Atmosphere		
IT	--	2160
ST	--	2300
HT	--	2350
FT	--	2520

AAG COLD-MODEL TEST OPERATION

Summary

Seven tests were conducted during April on the half-section, 4-foot diameter ash agglomerator cold-flow model. This work was accompanied by modification of the throat and expander cone of the 3 x 3 in. venturi to determine the effect of obstructions and aeration on the rate of discharge of spheres from the fluidized bed. The test results indicate that these modifications do not significantly affect the discharge rate of spheres from the bed.

Five tests were conducted during May on the half-section 4-1/2 x 2 in. venturi throat. The effect of sphere mass concentration in the bed on the sphere discharge rate was examined. The effect of venturi velocity at a fixed sphere concentration at a superficial velocity of 2.01 ft/s and at two bed heights was determined, and a basis for comparison with future runs was selected (Run 57).

Tests to Determine the Effect of Obstructions and Aeration

Objective

Observations and results of previous test runs on the cold model established that there are various operating parameters influencing the discharge rate of spheres from the bed. Important factors are venturi velocity, average sphere size, particle density of bed material and spheres, bed height, and mass fraction of spheres in the bed. The objective of the tests conducted this month was to determine the influence of some of these variables on venturi discharge rate when the venturi was aerated and when obstructions were implanted in the venturi. Additional effects of parameters related to these two practices were also examined.

Equipment Set-Up

Figure 9 shows the equipment arrangement schematic illustrating the test system. This half-cylindrical 4-foot diameter vessel has a conical grid distributor similar to that used in the ash agglomerating gasifier pilot plant. Figure 10 shows a view of the venturi section used in the test work conducted during April. The obstruction consisted of one extendable 1/4-in. diameter copper tubing. Aeration was produced by nitrogen introduced through a rotameter located at a point just above the venturi throat.

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Figure 9. FOUR-FOOT COLD MODEL OF ASH AGGLOMERATING GASIFIER

Operation and Results

Runs 48 and 49

Runs 48 and 49 were conducted as trials to determine the optimum locations for aeration and the obstruction to generate significant data.

Run 50

This run was conducted using the obstruction location shown in Figure 10 and at varying venturi velocities and weight fractions of spheres in the bed. The results show extremely low discharge at the velocities used.

Run 51

This test was conducted using four different obstructional areas and six different venturi velocities. All other operating parameters were similar to those of Run 50. Figure 11 shows that total discharge through the venturi varied with venturi velocities but the percentage of sphere yield fell sharply below a velocity of 21 ft/s. An obstruction at the venturi has little effect upon solids discharge rate. This figure also shows that net sphere yield is relatively constant at venturi velocities down to 20 ft/s.

Run 52

Run 52 demonstrated the effect of added aeration on venturi discharge rate. Figure 12 shows that discharge rate was inhibited by aeration at lower venturi velocities. As usual, venturi discharge varied inversely with venturi velocity, with the yield of spheres diminishing rapidly at velocities below 21 ft/s.

Runs 53 and 54

These two runs were conducted to determine the effect of superficial velocity on sphere discharge rate at a constant venturi velocity. Two venturi velocities were selected, both near the point at which spheres yield in the discharge decreases. (See Figure 13.) Visually, the improved (greater) discharge rate appeared to be the result of more intensive solids mixing at the higher superficial velocities.

Conclusions

The conclusions of all the tests conducted during the month can be summarized as follows:

- The obstruction of up to 10% of the cross-sectional area of the venturi throat affected the discharge rate by a maximum of approximately 25%, but the effect diminished as venturi velocity was increased and disappeared at venturi velocities above 30 ft/s. (See Figure 14.)

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Figure 10. AAG COLD MODEL: VIEW OF THE
3 x 3 in. VENTURI THROAT

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Figure 11. AAG COLD MODEL: VENTURI VELOCITY VERSUS
 RATE OF TOTAL BED DISCHARGE AND WEIGHT FRACTION
 OF SPHERES IN DISCHARGE (Run 51)

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Figure 12. AAG COLD MODEL: VENTURI VELOCITY VERSUS
 RATE OF SPHERES DISCHARGE (Run 52)

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Figure 13. AAG GOLD MODEL: RATE OF SPHERES DISCHARGE
VERSUS SUPERFICIAL VELOCITY (Runs 53 and 54)

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Figure 14. AAG COLD MODEL: VENTURI VELOCITY VERSUS RATE OF SPHERES DISCHARGE AND WEIGHT FRACTION OF SPHERES IN DISCHARGE (Run 51)

- There appears to be a relationship between venturi velocity and sphere concentration in the discharge and total discharge rate corresponding to other fixed operating parameters. As the venturi velocity is lowered, the discharge rate increases; however, this is accompanied by a lower concentration of spheres in the discharge. Increasing the velocity above the point at which the classification of spheres is virtually complete (approximately 95% spheres in the discharge) has a negligible effect on spheres concentration in the discharge. Lowering the venturi velocity below this "critical" velocity produces a sharp decrease in spheres concentration in the venturi. The effect of venturi velocity on the rate of spheres discharge is, however, linear over the range of venturi velocity. The effect of aeration was slight, and diminished at aeration rates above 13 SCF/hr. In both cases discharge rates were diminished. Therefore, these types of modifications and experiments will not be continued.

Tests on 4-1/2 x 2 in. Venturi

Objective

The main objectives of these test runs were -

- To determine the sphere discharge rate at varying mass fractions of spheres in the bed.
- To determine the sphere discharge rate at varying venturi velocities
- To determine the sphere discharge rate at varying superficial velocities while keeping venturi velocities constant.

Equipment Set-Up and Modification

There was no major change in equipment set-up. The 3 x 3 in. venturi throat was replaced by a half-section 4-1/2 x 2 in. throat. (See Figure 15.) We also replaced the 18 x 3 in. Plexiglas bed discharge collecting tube by a 1-ft³ cylindrical conical steel vessel. This enabled us to operate the unit at much lower venturi velocities at which the discharge rate is higher and large amounts of material are collected in the sampling.

Operation and Results

Run 55

This run demonstrated the effect of three different mass fractions (0.02, 0.07, and 0.09) in the bed on spheres discharge rate. As established earlier, the discharge rate increased linearly with an increased mass fraction of spheres in the bed. (See Figure 16.)

Run 56

This test was conducted to determine the effect of varying venturi velocities on spheres discharge rate in relation to a fixed superficial velocity, bed height, and mass fraction of spheres in the bed. Results indicate that spheres discharge rate varied

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Figure 15. AAG COLD MODEL: VIEW OF THE
4-1/2 x 2 in. VENTURI THROAT

Figure 16. AAG COLD MODEL: WEIGHT FRACTION OF SPHERES IN BED
 VERSUS RATE OF SPHERES DISCHARGE (Run 55)

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inversely with venturi velocity and directly with an increased mass fraction of spheres in the bed. (See Figure 17.)

Run 57

This run was a repetition of Run 56 except that the bed height was increased from 12 to 16 inches. This condition will be the basis of comparison for all our future runs. We also made a film showing the fluidization pattern in the expander at different operating venturi velocities.

Runs 58 and 59

These tests demonstrated the effect of varying superficial velocities on spheres discharge rate. A comparison of the results of this test and those conducted earlier in other venturis will be made and reported in a subsequent report.

Conclusions

- The discharge rate of spheres was found to be linearly proportional to the mass concentration of spheres in the bed up to 9 weight percent mass concentration. The coefficient of proportionality between the discharge rate in lb/min and the concentration in lb spheres/lb total bed was 4.64 for the conditions of Run 55 (venturi velocity of 40.4 ft/s).
- The effects of venturi velocity on the mass concentration of spheres in the discharge and on mass discharge was determined and is shown for Runs 56 and 57 in Figures 17 and 18. The qualitative behavior is as described in the Conclusions section for the 3 x 3 in. venturi. Run 57 will form the basis of cross-venturi comparisons. The regression fit for the discharge rate (lb/min) versus venturi velocity (ft/s) for this run was found to be $\dot{M} = 12.17 - 0.305 U_v$.
- The results of Runs 56 and 57 indicate that bed height does have an effect on the sphere discharge rate. This effect appears very slight; however, a direct determination of the effect of bed height is not possible from these tests because the sphere concentrations in the bed differed by approximately 3%. This will be investigated in more detail at a later time.

Test Plans for June

We plan to repeat the conditions of Run 57 with venturi throat sizes of 3 x 2 in. and 2 x 2 in. and to compare the spheres discharge rate at these conditions at different venturi diameters. We will also install a straight 6-in. flow pipe and determine the effect of such a configuration on the discharge rate.

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Figure 17. AAG COLD MODEL: VENTURI VELOCITY VERSUS RATE OF SPHERES DISCHARGE AND WEIGHT FRACTION OF SPHERES IN DISCHARGE (Run 56)

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Figure 18. AAG COLD MODEL: VENTURI VELOCITY VERSUS
 RATE OF SPHERES DISCHARGE AND WEIGHT FRACTION OF
 SPHERES IN DISCHARGE (Run 57)

BENCH-SCALE ASH AGGLOMERATION TESTS

Introduction

We are initiating a series of experiments to determine the regimes of initiation of ash particle growth and sinter-free operation of Western Kentucky coal in a fluidized-bed environment. The solids particles will be fluidized at high temperature in the absence of oxygen to study the effect of variables under an unchanging environment. The reactor that will be used in these experiments is shown in Figure 19. The reactor is a 2.0-in. Schedule 40 316 SS pipe equipped with pipe caps that have tubing connections to serve as gas inlet and outlet ports and the thermocouple entrance port. A sintered metal porous plate serves as the gas distributor. A bed of about 250-300 grams of material is to be fluidized with N_2 at velocities of 2-5 ft/s. Other details of the experimental set-up are shown in Figure 20. The temperature of the bed will be maintained by heating elements in a furnace that will enclose the reactor. The N_2 will be preheated in a separate furnace and/or in a coil encircling the reactor within the furnace, and the temperature of the N_2 as it enters the reactor will be measured. Pressure taps immersed in the fluidized-bed zone will be used to record a signal representing the density of the fluidized material. Any drastic decrease in this signal will indicate the occurrence of a defluidized state in the reactor, and the conditions pertaining to such a state will be noted. The experiments will be conducted at various superficial velocities and temperatures, with several size ranges of material, and with materials of differing ash contents. The material to be used will be bed material from the AAG pilot plant. A sufficient quantity of a homogenous nature having the required characteristics has been secured beforehand to ensure that all the tests can be conducted with that batch of material. The material removed from the gasifier at the conclusion of the test will be examined for any particle growth or agglomerate or sinter formation.

Summary

During May two gasification tests and a high-temperature fluidization test were conducted in a bench-scale test reactor with char that was obtained from the operation of the ash agglomerating gasifier (AAG Run 122) with coal. The objective of these tests was to determine the agglomeration characteristics of the char at the test reactor operating conditions shown in Table 9. The feasibility of using gasifying environments other than N_2 , i.e., steam and oxygen, was also studied in these initial tests.

Figure 19. BENCH-SCALE TEST REACTOR

Figure 20. DIAGRAM OF THE EXPERIMENTAL SET-UP

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Table 9. OPERATING CONDITIONS FOR AAG-CHAR GASIFICATION TESTS^a

Run No.	AGX-1	AGX-2	AGX-3
Nominal Feed Size, U.S.3.	-3-1/2+325	-3-1/2+325	-3-1/2+325
Reactor Temperature, °F ^b	1710	1812	1830
Reactor Pressure, psig	0-2	0-2	0-2
Weight of Feed, g	269	267.4	268.4
Bed Aspect Ratio, L/D	approx. 3	approx. 3	approx. 3
Fluidizing Gases			
Nitrogen Rate, SCF/hr	20.5	9.5	7.9
Steam, lb/hr	--	0.397	0.401
Oxygen, SCF/hr	--	--	1.9
Superficial Gas Velocity, ft/s ^c	0.91-1.03	0.83-0.94	0.86-0.97
Steady-State Operating Period, min ^d	60	60	7

^aConducted in a 2-inch Schedule 40 316 S.S. reactor fitted with a porous-plate distributor.

^bAverage temperature during the steady-state period as based on the readings of the thermocouple, which is located 1-3/4 inches above the distributor plate.

^cN₂ + Steam + Oxygen (CF/S feed gas at reactor pressure and temperature)/cross-sectional area of reactor.

^dStarted when the reactor attained the operating temperature.

Operation and Results

A typical test was conducted as follows: Approximately 269 grams of the composite char (-3-1/2+325 U.S.S.) was placed in the reactor. Heat-up of the reactor and bed was accomplished by means of a furnace equipped with independently controlled electrical resistance heaters. During the reactor heat-up the char bed was kept fluidized by nitrogen. When the bed reached the operating temperature, the reacting gases (pre-heated steam for Run 1 and oxygen at 70°F for Run 3) were introduced into the bed. Each test except Run 3 was conducted for 1 hour. Because of sintering of the bed material, Run 3 was shut down after 7 minutes. After shutdown, screen analyses of the residue and cyclone fines were conducted for each test.

A summary of the operating results for the three tests is shown in Tables 10 and 11. Screen analyses of the feeds, residues, and the fines collected at the bottom of the cyclone were performed. For AGX Runs 1 and 2, a comparison was based on the weight of +40 mesh material of feed and residue. Run 3 was not included in the analysis because of sintering of the char.

Conclusions

The conclusions from these tests can be summarized as follows:

- There was no char agglomeration at the reactor operating conditions in AGX Runs 1 and 2.
- Char sintering was observed near the gas distributor in Run 3, which was conducted with N₂, steam, and oxygen at 1830°F.

Plan for Future Work

The current experimental equipment is being modified. This will enable operation at temperatures as high as 2100°F. A multi-orifice type distributor has been designed and will be fabricated and installed in the reactor if sintering at the grid remains a problem. The initial phase of the testing will be conducted with N₂ only. Various combinations of temperatures and superficial velocities will be used in the tests. A sample of higher ash material will also be prepared for use in the tests. The effect of using screened fractions of these materials to determine the effect of particle size on agglomeration is also planned. Carbon and ash analyses of all materials and size fractions and fluidization tests are under way; the results will be presented in subsequent reports.

Table 1C. SUMMARY OF OPERATING RESULTS OF AAG-CHAR
GASIFICATION TESTS

Run No.	AGX-1	AGX-2	AGX-3
Nominal Feed Size, U.S.S.	-3-1/2+325	-3-1/2+325	-3-1/2+325
Test Period, min	60	60	7
Weight of Feed, g	269.0	267.4	268.4
Weight of Residual Char, g	224.4	205.8	193.8
Fines Collected, g	28.4	15.5	40.5
Losses, g	16.2	--	
Carbon Conversion, % **	--	50.0	22.7
Condensed Water, g	--	61.0	106.0

* Conducted with N₂, therefore no carbon conversion and no condensate.

** Calculated from the carbon balance.

wt of carbon gasified = wt of carbon in feed - (wt of carbon in residue +
wt of carbon in fines)

$$\% \text{ carbon conversion} = \frac{(\text{wt of carbon gasified})}{(\text{wt of carbon in feed})} \times 100$$

jb/tr

Table 11. CARBON AND ASH CONTENTS OF FEEDS, RESIDUES, AND FINES OF
AAG GASIFICATION TESTS IN THE 2-INCH-DIAMETER REACTOR

Run No.	AGX-1			AGX-2			AGX-3		
	Feed	Residue	Fines	Feed	Residue	Fines	Feed	Residue	Fines
	wt%								
Carbon	--	--	--	34	16	80.7	36	23.5	72.0
Ash**	--	--	--	66	84	19.3	64	76.5	28.0

* This run was conducted with N₂, therefore no samples were given for analysis.

** Samples were analyzed at the pilot plant by the quick ash procedure.