U.S. Coal Test Program on BGC-Lurgi Slagging Gasifier

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EPRI PERSPECTIVE

PROJECT DESCRIPTION

The Lurgi moving-bed gasifier, as currently used commercially in South Africa and elsewhere, is a pressurized countercurrent reactor with the coal fed and the ash withdrawn via lock hoppers. Excess steam is injected at the bottom to keep the ash below its fusion temperature so that it operates with dry ash removal. This excess steam reduces the thermal efficiency and produces large volumes of contaminated water, which require treatment. The British Gas Corporation (BGC) is developing a slagging version of the Lurgi gasifier. By operating at higher slagging temperature, only the steam for the gasification reaction is required. The steam consumption is therefore much lower, the overall efficiency is greatly improved, and the waste water treatment is markedly reduced. In addition, the higher slagging temperature increases the reaction rate. These factors result in throughputs three to four times that of the dry ash units.

In moving-bed gasifiers, both dry ash and slagging, a sized coal, typically ¼ inch to 1½ inch, is used as top feed through the lock hopper. The slagging version of the gasifier, however, also provides the opportunity for gasification of by-product tar and fine coal by injection into the bottom slagging zone.

In 1975 EPRI and 14 U.S. oil and gas companies joined with BGC to convert one of their commercial Lurgi dry ash gasifiers to slagging operation at Westfield, Scotland. This initial program, completed in mid-1977, culminated in a 23-day run on Scottish coal and confirmed the throughput (350 tons/day) and efficiency advantages. Economic evaluations conducted for EPRI, using Illinois No. 6 coal and based on extrapolated performance of the Westfield Slagging Gasifier, confirmed the economic competitiveness of this process for gasification-combined-cycle power generation (EPRI Final Reports AF-642 and AP-1725), fuel gas (EPRI Final Report AF-782), and methanol (EPRI Final Report AF-523) production.

The work reported here covers the results of a further test program on caking coals (British and U.S.) using the 6-foot-diameter, 350-tons/day gasifier. EPRI undertook this project (RP1267-1) in 1979 to examine the technical suitability of the technology for power generation applications.

A total of 17 days of testing was accomplished in essentially three runs. In the first run a British coal, Rossington, was used. This coal, with which BGC had considerable prior run experience, has caking qualities and other properties similar to the abundant U.S. Illinois No. 5 and No. 6 coals. Load turndown and turnup were investigated over the range of 30 to 100% of full load. Measurements of operational stability at various load levels and during transients were completed.

The second run utilized Pittsburgh No. 8 coal with a similar series of load-change tests to those conducted earlier with the Rossington coal. In addition, this run included going to and returning from standby conditions.

In the third and final run, also using Pittsburgh No. 8 coal, the following additional parameters were investigated: 20% fines addition to feed coal, tar recycle to bottom slagging zone, limestone flux addition, and extended periods of system control in the gasifier follow mode.

PROJECT OBJECTIVES

The main objective was to test the capability of the BGC-Lurgi Slagging Gasifier to respond to the type of load changes which will be demanded from gasification units integrated with combined-cycle power-generating systems. The four specific project objectives were:

- Measure and analyze steady-state performance of the gasifier on strongly-caking U.S. coal under various process conditions including various throughputs, tar recycle to distributor and to tuyeres, and with fines addition to the feed coal
- Identify any steady-state fluctuations in gas flow or heating value which might adversely impact the gasifier's suitability for use in an integrated power system
- Determine the response of the gasifier to various rates of load change including extremely rapid step changes
- Assess the general state of the technology

PROJECT RESULTS

The high efficiency of the BGC-Lurgi Slagging Gasifier at full throughput and at turndown was documented through good quality heat and material balances on both weakly-caking and strongly-caking coals. The unit performed extremely well over the short term, meeting all dynamic and steady-state stability requirements of combined-cycle power-generating systems. Dramatic bed instabilities and gas-heating value fluctuations did not occur even during rapid (10%/minute) rate changes.

All relevant steady-state fluctuations were quantified and can be accommodated through appropriate design. Simple control in the gasifier follow mode can be used for all transitions even while incorporating lock hopper pressurization with raw gas. The dynamic response was excellent. Fluxing with limestone (more widely available than blast furnace slag), gasifying coal containing 20% fines, and total gasification of net tar were all demonstrated with no short-term difficulties. Some adverse operating conditions were encountered (most of which can be designed out of a commercial plant), and the gasifier proved itself to be very forgiving. Recoveries from upsets were made with minimum deviation from the planned test schedule and, in most cases, with no disruption of gas production.

The following report accurately describes the BGC-Lurgi Slagging Gasifier's excellent short-term performance and operability during the EPRI tests. The performance results have been utilized under a separate contract with General Electric Company (RP914) to confirm the viability of several proposed schemes for control of gasification-combined-cycle plants using the BGC-Lurgi Gasifier. A report on this work will be issued shortly.

Subsequent to this EPRI test program, BGC has conducted tests on additional fines injection through the tuyeres to the bottom-slagging zone.

Not yet demonstrated are the performance on lower rank coals and the long-term mechanical integrity of the gasifier. BGC has announced plans to address this latter issue by conducting a long-term endurance run at Westfield in summer 1981.

Another concern with all gasification technology is that of scale-up to full commercial size. A particular concern in the BGC-Lurgi technology is flow distribution at larger diameters than the 6-foot unit already demonstrated. However, the stability of performance at part-load conditions documented herein for the 6-foot unit gives reason for optimism that the unit should also be operable when scaled up. BGC is proceeding with plans to install a larger 8-foot-diameter commercial-sized unit at Westfield capable of 600 to 800 tons/day throughput for commissioning in early 1983. Together with the demonstrated short-term performance documented in this report, the extended run and scale-up to 8-foot diameter will confirm the position of the BGC-Lurgi Slagging Gasifier as a most promising candidate for power generation applications. The technology development is therefore proceeding on a logical plan, which, within a few years, should result in full commercial availability.

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Section 1

INTRODUCTION AND SUMMARY

The main aims of the EPRI project were to orientate the performance of the Slagging Gasifier, with its already proven record of success on a variety of coals, to the special requirements of the combined cycle operation. The three main special requirements were:

- For the gasifier to perform with stability and reliability at a variety of loadings, perhaps over the range 30 - 110% of standard loading.
- For changes between the above loadings to be made reliably and quickly, without producing gasifier upsets or transient behaviour with variation in make gas CV.*
- For the back end of the plant to produce a constant CV gas at a steady flow, the latter being capable of rapid alteration when desired.

The range of loadings required and the time required to change from one to the other were proposed by British Gas (from data supplied by EPRI) and accepted by EPRI before the project and these are shown in Table 1-1.

Table 1-1

LOADINGS AND LOAD RATE CHANGES REQUIRED FOR SLAGGING GASIFIER IN A COMBINED POWER CYCLE

Loading (SCF/H Oxygen)	% of Standard Load	Time to Change to Given Load From Standard Load
158,080	98.8	6 seconds
147,200	92.0	60 seconds
112,000	70.0	10 seconds
48,000	30.0	1 hr 40 minutes

*Calorific Value

Table 1-1 assumes a standard loading of 160,000 SCF/H oxygen and the load change rates correspond to the combined power cycle functions of frequency regulation, tie-line back up and daily load following.

Before the runs, plant controls were modified to hold product gas flow steady. Using this "Flow Control" system, steam and oxygen are automatically fed to the gasifier as needed to maintain plant pressure. This system is described in Appendix A.

The project was to be carried out on the British Gas standard reference coal, Rossington, a 702 rank coal, and on the EPRI chosen coal, Pittsburgh 8, from the Champion Plant in Pennsylvania. Ancillary programme objectives included running this coal unscreened (with 25% material less than ¼ inch) and tar injection to the tuyeres was planned.

EPRI RUN - 01

Run EPRI - 01 was on Rossington coal and planned to commission new gasifier instrument systems in parallel with the investigations into load change. The run was started on 14 October 1979 at 11:12 hours when steam/oxygen was injected into the gasifier and the gasifier was set up at standard loading of 160,000 SCF/H oxygen, with steam/oxygen 1.30. Successive test periods looked at gasifier performance at 110,000, 80,000 and 50,000 SCF/H oxygen loading, with the loading being returned to standard between each test period. The period of running at 80,000 SCF/H oxygen loading was prolonged in order to obtain data on gasifier performance at different stirrer speeds and different rates of tar injection to the bed top. At 50,000 SCF/H loading the steam/oxygen ratio was high at 1.88 (due to an error in the sets), but this caused no operational problems. All load changes were carried out at or greater than the recommended rates and the gasifier was introduced to the flow control mode with gasifier rates being brought up to 180,000 SCF/H oxygen loading in this mode before the run was terminated at 12:04 on 19 October.

EPRI RUN - 02

No major gasifier changes were made for EPRI - 02, with the run aiming to consolidate the position obtained on Rossington coal with Pittsburgh 8. The run started at 03:03 hours on 7 November and, after a standard start up on Rossington coal, the loading was brought to 80,000 SCF/H oxygen loading and Pittsburgh 8 coal brought to the gasifier. There were some initial problems with underfluxing on limestone, but eventually satisfactory operation was obtained with BFS* fluxing at

*Blast Furnace Slag

130,000 SCF/H oxygen loading. At this juncture, however, premature run termination occurred, after only 36½ hours on line, due to a flange failure at a tuyere. The leak was rectified and all other tuyeres were checked before the gasifier was ready for the continuation of run - 02, redesignated EPRI - 02B. This run started on 21 November at 16:45 with Pittsburgh 8, fluxed with BFS, being introduced to the gasifier at 130,000 SCF/H oxygen loading. The loading was then brought to 160,000 SCF/H oxygen and then down to 110,000 SCF/H oxygen before being returned to 160,000 SCF/H oxygen, some load spikes down to the intermediate loading being carried out. At this stage in the run there was a problem in the quench chamber. At 10:11, 24 November, it was decided to carry out an atmospheric pressure standby to cure the problem. This was done after depressurising the gasifier to standby conditions. The gasifier was boxed up, repressurised, and the run restarted at 01:18 on 25 November, no problems being encountered. Test periods at 160,000, 110,000, 80,000 and 50,000 SCF/H oxygen loading were carried out with extensive running in flow control in order to determine optimum controller settings. The gasifier again demonstrated its ability to change rapidly from one loading to another with no process upsets and no gasifier transients, before the run was terminated at 12:20 on 27 November.

EPRI RUN - 03

EPRI - 03 aimed to amplify the successes of run EPRI - 02B and to carry out the ancillary programme objectives of running on unscreened coal and of tar injection to the tuyeres. Limestone fluxing of Pittsburgh 8 was also to be attempted.

The run was started at 13:03 on 13 December, and the gasifier established on screened Pittsburgh 8 coal, fluxed with BFS, at 130,000 SCF/H oxygen loading. Performance data was obtained at this loading and then tar injection to the Tuyeres was brought on, initially at about 1,000 lbs/hr and then at an estimated 1,500 lbs/hr. Gasifier performance was satisfactory during these tests and following these unscreened Pittsburgh 8 coal was charged to the gasifier. Performance on this fuel differed little to that on the screened fuel, and it was run at loadings of 130,000, 80,000, 50,000 and 160,000 SCF/H oxygen. Most of the run was carried out in the flow control mode and the gasifier was able to change rapidly from one load to another without any process upsets. Limestone fluxing was also successfully introduced for the first time. The run was terminated at 10:43 on 18 December.

The runs are shown in table form in 1-2. The short programme had satisfactorily demonstrated that the Slagging Gasifier is suitable for use in a combined power cycle, some of the important points being:

TABLE 1-2. RUNS CARRIED OUT DURING EPRI CONTRACT, OCTOBER - DECEMBER 1979

Run Number	Total Hours Running Time (Hours)	Approximate Fuel Usage* (Long Tons)			Flux Used	Tar Injected		Range of Loadings '000 SCF/H Oxygen	H_2O/O_2 Used (v/v)	Control		Achievements
		Rossington	Pitts. 8 Screened	Pitts. 8 Unscreened		Top	Bottom			Flow	Pressure	
EPRI–01	121	996	–	–	Limestone	✓		50 –180	1.39 –1.88	✓	✓	Smooth running on Rossington successfully demonstrated with stable operation at a variety of loads. Change from one load to another achieved rapidly with no process upsets and no transients during change. No significant change in make gas characteristics at all loads.
EPRI–02A	36½	202	102	–	Limestone BFS	✓		80 –160	1.10 –1.30	✓	✓	Pittsburgh 8 briefly introduced to the gasifier but detailed testing curtailed by flange failure at a tuyere
EPRI–02B	139½	83	1021	–	Limestone (Rossington only) BFS	✓		50 –160	1.10 –1.30	✓	✓	Good running on Pittsburgh 8 demonstrated Similar performances to Rossington with bed behaviour slightly less steady. Gasifier recovered rapidly from a 13 hour cold standby period with no problems. Good operation in flow control demonstrated over substantial period of run.
EPRI–03	118	70	409	602	Limestone BFS	✓	✓	50 –160	1.10 –1.30	✓	✓	Good running on Pittsburgh 8 coal both screened and unscreened with little difference in performance between the two fuels. Limestone fluxing on Pittsburgh 8 successfully demonstrated. Tar injection to the tuyeres successfully demonstrated with high injection rates. Most of the run operated smoothly in flow control
TOTALS	415 (17.3 Days)	1351	1532	602	Limestone BFS	✓	✓	50 –180	1.10 –1.88	✓	✓	*Excludes 10 12 tons Wondercoal used during start up.

- The gasifer can run steadily, with no signs of degenerative behaviour, at a variety of loads in the range 50 - 180,000 (180,000 demonstrated only on Rossington coal but probably certain on Pittsburgh 8 as well) SCF/H oxygen (30 - 110% of standard).

- Average calorific value of the make gas was nearly constant, whatever the loading; small differences which did arise could be attributed to different concentrations of nitrogen from the slag system at varied loads.

- Changes from one loading to another can be carried out rapidly without upsetting the gasifier operation and without producing any significant transients (see Section 12) in make gas quantity. Indeed, load changes in the 100 - 30 - 100% range were carried out at up to and over ten times the designated programme rates of change on occasion.

- Flow control, with back end of plant held steady, works well on the Slagging Gasifier. A simple closed loop circuit was shown to be satisfactory; as the gas CV is steady and only influenced in a very minor way by any bed upsets which did occur, the above referred to simple closed loop appears to be adequate for gasifier control in the combined power cycle mode.

- As a consequence of flow control, steam/oxygen flows to the gasifier will fluctuate slightly; more so if pressurisation of the coal lock is done by crude gas from downstream.

- Tar can be re-injected to extinction down in the gasifier tuyeres with no detriment to gasifier performance and with only small changes in gasifier characteristics, including a slight drop in make gas CV.

- Fluxing can be carried out with either BFS or limestone.

- Locking can be on nitrogen or on raw gas. Locking has no significant effect upon gasifier performance.

- Stirrer revolution speed and tar injection to the gasifier top contributes to bed behavior in general and there are optimum values for the above for given conditions on a given coal. However, the gasifier exhibits a wide tolerance of the above parameters.

- Pittsburgh 8 coal, as delivered to Westfield, is a suitable coal for Slagging Gasification. The gasifier will tolerate fine material of less than ¼" at levels up to at least 25% in the top coal feed. This level, on the evidence of EPRI - 03, could be significantly higher.

It can therefore be concluded that the runs carried out in the short programme sponsored by EPRI successfully achieved all major objectives. Pittsburgh 8 coal can be regarded as a suitable fuel for the Slagging Gasifier in a combined cycle system. Complete optimisation of operating conditions on this fuel was not possible during the contract period, and this, coupled with gasifier design improvements aimed particularly at Pittsburgh 8 coal, is likely to result in an even better performance on a commercial plant dedicated chiefly to this fuel.

Section 2

STEADY STATE HEAT AND MASS BALANCES

The production of heat and material balances during the EPRI project were subject to errors arising out of the inaccuracies discussed in Appendix E. In the preparation of these balances, some of these potential inaccuracies can be taken into account, and the data can be manipulated in a consistent manner to produce a mass and heat balance which will be close to within 2% for the major elements C, H, and O. It is this philosophy which has been applied to the thirteen heat and material balances produced across the EPRI projects, and whose details can be found in the Appendix.

The results of these mass and heat balances are summarised in Table 2-1. The results during the EPRI contract serve to illustrate the consistency of the Slagging Gasifier performance data from coal to coal (particularly true for bituminous coals) and the relative insensitivity of the process to changes in process parameters. The steam usage of the process between 1.0 and 1.30 steam/total oxygen input ratio is shown in Figure 2-1, where it can be seen that, within experimental error, consistent steam usage is obtained across the three runs, despite variations in coal type, load, fluxing and tar injection to the tuyeres. The oxygen consumption is plotted against gasifier loading in Figure 2-2, and this shows the only discernable trend, that of increased specific oxygen consumption towards lower loadings. This effect is likely to be due to relatively greater heat losses expressed as a percentage of total gasifier energy output.

Changes in process condition did produce minor effects upon gasifier performance data, although these effects were very small. Thus, tar injection to the tuyeres resulted in a slightly higher gasifier outlet temperature, a lower methane concentration in the make gas and a slightly higher oxygen consumption and lower gas HHV. Coal with more fines in perhaps showed a slight increase in oxygen consumption.

Little effect was seen of different coal types, different fluxing agents and tar injection to the gasifier top.

TABLE 2-1. SUMMARY OF GASIFIER PERFORMANCE DATA

Run No.	EPRI-01 MB1	EPRI 01 MB3	EPRI 01 MB4	EPRI 01 MB5	EPRI 02 MB1	EPRI 02 MB2	EPRI 02 MB3	EPRI 02 MB4	EPRI 03 MB1	EPRI-03 MB2	EPRI 03 MB3	EPRI 03 MB4	EPRI 03 MB5
Coal	Rossington	Rossington	Rossington	Rossington	Pitts. 8	Pitts. 8	Pitts. 8	Pitts. 8	Pitts. 8	Pitts. 8	Pitts. 8	*Pitts. 8	*Pitts. 8
Flux	Limestone	Limestone	Limestone	Limestone	BFS	BFS	BFS	BFS	BFS	BFS	BFS	BFS	Limestone
H_2O/O_2	1.23	1.15	1.15	1.58	1.17	1.13	1.14	1.00	1.15	1.13	1.14	1.11	1.70
O_2 10³ SCF/H (Inquired)	170.3	89.7	88.7	49.5	161.9	113.4	80.1	50.9	131.3	13'.9	131.5	132.6	179.8
O_2 Consumption SCF/Therm	58.15	60.70	61.85	62.25	57.43	59.88	61.14	64.33	57.33	59.40	59.00	60.94	58.67
DAF bed Loading lbs/hr/ft²	853	445	440	273	816	558	384	250	681	606	592	636	678
Steam Consumption lb/therm	3.29	3.30	3.38	4.87	3.19	3.21	3.32	3.05	3.1J	3.20	3.19	3.22	3.35
Efficiency †	87.07	90.96	87.52	89.34	85.09	84.42	85.25	86.53	80.91	83.70	83.28	80.25	87.65
Tar (% DAF Coal) D = Distributor T = Tuyeres	3.6 D	9.1 D	6.3 D	8.5 D	3.8 D	4.2 D	4.6 D	6.9 D	0	4.7 T	5.4 T	4.5 D	5.2D

* Unscreened coal (over 20% < ¼")

†(Total crude gas therms out)/(Coal thrown in)

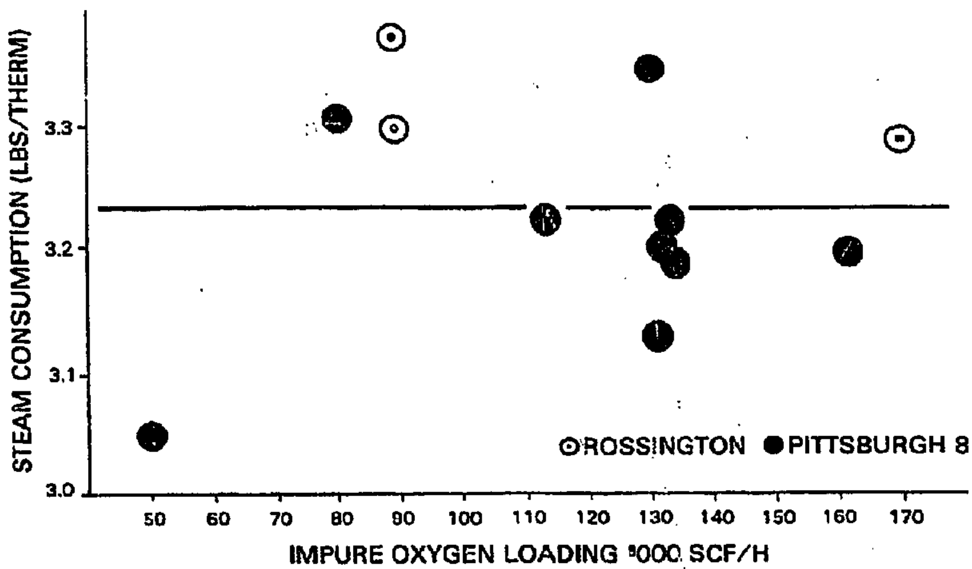

Figure 2-1. Steam Consumption as a Function of Load

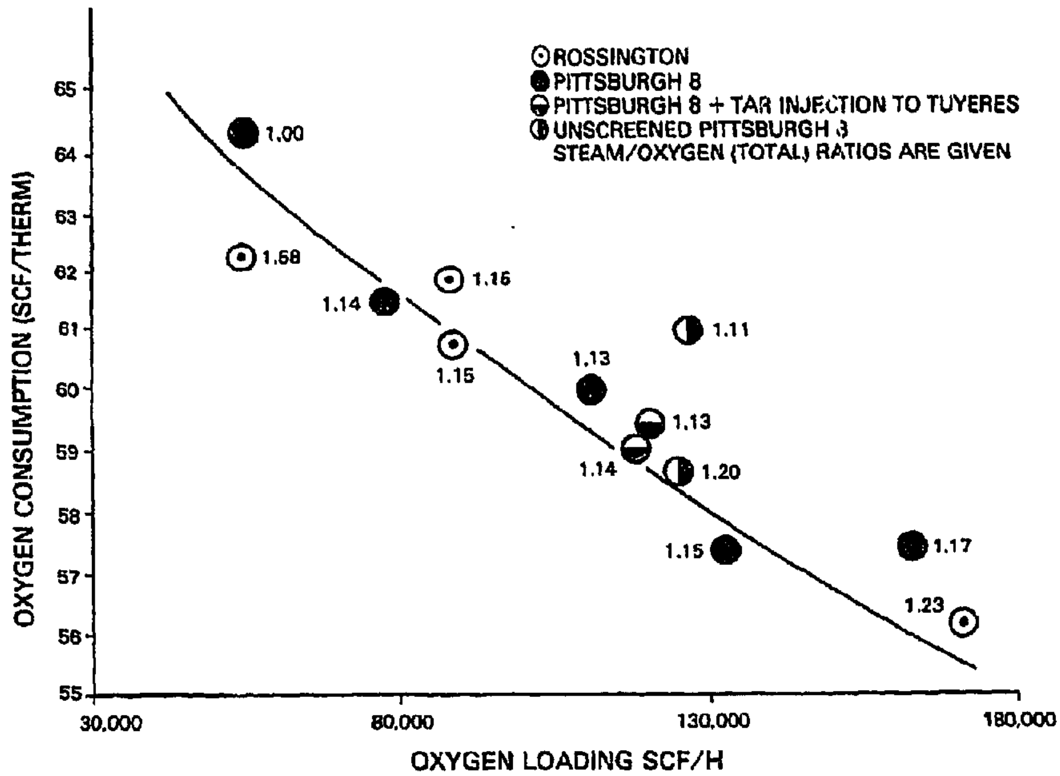

Figure 2-2. Gasifier Loading V. Oxygen Consumption

The consistent performance of the gasifier across a wide variety of conditions can be expressed in another way by examining the gas analyses across the various mass balance periods, as Table 2-2 illustrates. Gas LHV was very steady across a range of conditions, showing a reduction towards low loadings because of extra contribution from slag system Nitrogen as Figure 2-3 shows.

Fluxing agent choice will have a marginal effect on gasifier performance. If we compare the last two mass balances on EPRI - 03, 0.81% of the total output gasifier heat went out in the slag with limestone fluxing, whereas the BFS case was higher at 1.44%. Thus, limestone fluxing will result in a slightly better efficiency (reflected mainly in oxygen consumption). The oxygen consumptions on limestone (58.62 SCF/therm) and BFS (60.94 SCF/therm) bear this out, although in view of the inaccuracies in the latter figures the trend may be fortuitous.

There are minor differences in gas analysis when comparing Pittsburgh 8 and Rossington coal. In general, for the former coal, CO_2 levels in the raw gas tend to be higher by about 1%, whereas CO figures are correspondingly lower. On a N_2-free basis, the methane concentration in the gas goes up with reducing load on Pittsburgh 8 wherease no effect is noticeable for Rossington, as Figure 2-3 shows.

In general, therefore, analysis of the heat and mass balance and general performance data shows the general constancy of performance of the Slagging Gasifier over a range of conditions, confirming observations made over a significant number of earlier runs.

TABLE 2-2. GAS ANALYSES AT VARIOUS CONDITIONS

Run, MB No. and coal	EPRI 01 MB1 Rossington		EPRI 01 MB3 Rossington		EPRI 01 MB4 Rossington.		EPRI 01 MB5 Rossington		EPRI 02 MB1 Pitts. 8		EPRI 02 MB2 Pitts 8	
H_2O/O_2	1.73		1.15		1.15		1.58		1.17		1.14	
Loading '000 SCF/H O_2	170		89		89		50		167		113	
Flux	Limestone		Limestone		Limestone		Limestone		BFS		BFS	
CH_4	6.77	6.95	6.42	6.72	6.34	6.62	6.75	6.89	7.07	7.38	6.92	7.32
CO	58.97	60.53	58.25	60.86	59.05	61.68	50.07	53.57	56.47	58.91	56.27	59.44
H_2	28.46	29.21	27.19	28.41	26.86	28.08	29.34	31.39	28.02	29.25	27.42	28.99
CO_2	1.87	1.92	2.61	2.73	2.34	2.44	6.58	7.04	2.96	3.09	3.52	3.72
N_2 + Ar	2.58	--	4.79	-	4.27		6.53		4.22	-	5.42	-
C_2H_4	0.17	0.17	0.16	0.17	0.15	0.16	0.11	0.12	0.08	0.08	0.11	0.12
C_2H_6	0.36	0.37	0.33	0.34	0.32	0.33	0.33	0.35	0.34	0.35	0.44	0.47
H_2S	0.50	0.51	0.49	0.51	0.47	0.49	0.47	0.50	0.60	0.63	0.62	0.06
NH_3	0.34	0.35	0.19	0.20	0.14	0.15	0.29	0.31	0.26	0.27	0.26	0.27
COS	0.03	0.03	0.07	0.07	0.07	0.07	0.05	0.05	0.03	0.03	0.03	0.03
LHV	332		324		324		301		325		320	

Run, MB No. and coal	EPRI 02 MB3 Pitts 8		EPRI 02 MB4 Pitts 8		EPRI 03 MB1 Pitts 8		EPRI 03 MB2 Pitts. 8		EPRI-03 MB3 Pitts. 8		EPRI-03 MB4 Pitts. 8		EPRI-03 MB5 Pitts. 8	
H_2O/O_2	1.14		1.00		1.15		1.13		1.14		1.11		1.20	
Loading '000 SCF/H O_2	80		51		131		132		132		133		130	
Flux	BFS		BFS		BFS		BFS		BFS		BFS		Limestone	
CH_4	7.31	7.74	7.34	7.87	6.42	6.70	5.82	6.07	5.64	5.89	6.16	6.39	6.02	6.32
CO	54.70	57.93	53.61	58.20	55.83	58.77	55.64	58.01	56.58	58.03	57.89	60.03	56.50	59.24
H_2	27.23	28.85	26.14	28.38	28.43	29.67	29.37	30.62	29.43	30.73	27.40	28.41	27.34	28.51
CO_2	3.68	3.90	3.80	4.16	3.31	3.45	3.53	3.68	3.52	3.68	3.47	3.60	4.03	4.23
N_2 + Ar	5.58		7.88	-	4.19		4.09	-	4.23	-	3.56	--	4.79	-
C_2H_4	0.11	0.12	0.05	0.05	0.18	0.19	0.17	0.18	0.16	0.17	0.16	0.17	0.14	0.15
C_2H_6	0.49	0.52	0.31	0.34	0.70	0.73	0.57	0.59	0.63	0.66	0.51	0.53	0.50	0.53
H_2S	0.59	0.62	0.53	0.58	0.62	0.65	0.54	0.56	0.51	0.53	0.56	0.58	0.49	0.51
NH_3	0.27	0.29	0.27	0.29	0.30	0.31	0.25	0.26	0.26	0.27	0.27	0.28	0.38	0.40
COS	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03
LHV	332		312		325		320		319		324		317	

Notes: (1) Tar recycle was operating on all balances except during run 3, when there was no recycle during balance 1, and injection at the tuyeres during balances 2 and 3

(2) Lower Heating Value. This value excludes H_2S in the product gas, and is calculated for saturated gas at 30" Hg and 60°F.

Section 3

PROCESS DYNAMICS AT STEADY STATE

Gasifier behavior at steady state under a variety of operating conditions is summarized in Table 3-1. General gasifier behaviour centres around the fixed bed of coal, which in the Westfield Slagging Gasifier is approximately six feet in diameter. The "fixed" or more correctly, countercurrent, bed works best when smooth fuel flow down the shaft to the reaction zone is coupled with good well distributed gas flow up the bed, allowing intimate contact with the descending fuel and proper promotion of various physical and chemical reactions of the coal which must occur before the carbon rich char is delivered to the reaction zone.

At the top of the bed, there must be a system which enables coal to be laid down smoothly and continuously at the bed top, keeping this bed top level "fixed"; and a suitable escapement cross-sectional area must be provided for the ascending gases before they leave the reactor via one or more offtakes. At the bottom of the reactor there must be a system for injecting gasification medium which must allow the ash to separate as molten slag and be continuously removed from the reactor.

Imperfect bed behavior was observed during the EPRI contract and can be divided into four categories:

1. Poor coal feed to the gasifier top, resulting in a low fuel bed level.
2. Irregular flow of solid fuel down the gasifier shaft.
3. Abnormally high torque on the stirrer/distribution.
4. Bed channelling.

Throughout the EPRI project, the only example of poor coal feed to the bed top was during EPRI - 01 on Rossington coal. It was marked by a continuous rise in offtake temperature with concentrations of CO_2, CH_4, C_2H_4 starting to fall.

On successful addition of fresh coal, the sudden inrush of fuel to the bed top caused the offtake temperature to fall, and as the offtake temperature approached a minimum value, the concentrations of CO_2, CH_4, C_2H_4 all started to increase again,

TABLE 3-1. BED STABILITY INDICATIONS

Run and (Test Period)	Coal	O_2, 100 SCF/H	H_2O/O_2 (v/v)	Flux	Control Mode (F or P)	Stirrer HP/A %	Fan and % pump stroke	Mean Offtake (°F)		Length of Test Period (Hrs.)	Offtake Excursions per Hour	Calc LHV Btu-SCF	SD LHV	SD on O_2 load SCF/H ÷ 1000
2(A)	P8	110	1.2	BFS	P	85.75	D60,50	933		11.60	1.5	354.7	4.9	5.8
2(B)	P8	110	1.2	BFS	F	75	D50	927		6.52	0.15	334.7		5.7
2(C)	P8	160	1.2	BFS	F	Variable	D50	925		10.30	3.0	332.2	4.6	9.6
2(D)	R	130	1.1,1.3	L	P	90	D60	873		1.55	0.65	334.4	4.8	7.8
2(E)	P8	130	1.3	BFS	P,F	90	D60	965		3.13	1.9	328.6	4.7	4.6
2(F)	P8	80	1.3	BFS	F	56	D50	912		15.82	0.76	316.4		4.9
2(G)	P8	160	1.3	BFS	F	56	D50	914		2.88	7.1	337.7	4.5	6.1
2(H)	P8	50	1.3	BFS	F	37.5	D40,50	884		14.83	0.40	316.2		12.7
2(I)	P8	160	1.3	BFS	F	105	50,0	923		3.95	0.76	330.7		29.7
2(J)	P8	110	1.3	BFS	F	75	0	888		5.50	0.55	331.2	4.7	8.6
2(K)	P8	110	1.3	BFS	F	75	100,0	931		4.50	0.89	325.7		7.2
3(A)	R	160	1.3	L	P	70	0	945		3.40	2.1	336.8	3.7	1.1
3(B)	P8	130	1.3	BFS	F	87.5	0	910		14.13	1.9	320.4	2.3	2.9
3(C)	P8	130	1.3,1.2	BFS	F	87.5	165	966		10.00	2.7	328.8		3.2
3(D)	P8	130	1.2	BFS	F	87.5	195	970		10.00	2.8	325.1		2.7
3(E)	P8*	130	1.2	BFS	F	105	D60	943		10.00	1.9	323.4		13.6
3(F)	P8*	80	1.2	BFS	F O	72.5	D50	891		3.60	0.67	329.7	3.4	
3(G)	P8*	80	1.2	BFS	F O	72.5	D80	915		2.33	0.86	327.8	3.8	7.0
3(H)	P8*	50	1.3	BFS	F O	55	D80	888		2.97	—	318.0	5.7	9.1
3(I)	P8*	50	1.1	BFS	F O	56	D80	942		2.65	0.38	318.2	6.5	14.0
3(K)	P8*	130	1.2	L	F O	105	D60	967		3.05	0.66	331.4	4.7	4.7
3(L)	P8*	130	1.2,1.3	L	F O	125	D60	970		[illegible].30	1.0	334.3	3.5	5.7
3(M)	P8*	130	1.3	L	F O	87.5	D60	944		3.00	1.0	331.9	4.1	4.8
3(N)	P8*	160	1.3	L	F O	87.5	D60	949		3.00	0.67	332.7	3.8	4.7
3(O)	P8*	160	1.3	L	F O	105	D60	1010		3.60	4.3	331.2	3.8	4.4
3(P)	P8*	50	1.3	L	F O	55	D60,0	837		2.98	0	324.0	4.4	4.8
3(Q)	P8*	160	1.3	L	F O	125	0	1007		1.25	1.25	331.3	4.8	7.4

P8 - Pittsburgh 8 H - Rossington * - Unscreened BFS - Blast Furnace Slag L - Limestone P - Pressure F - Flow

O - Distributor T - Tuyeres 6 - Coal locking on new gas.

all tending to overshoot their original value eventually reaching equilibrium again.

This particular incident appeared to be caused by fuel hanging up above the top cone of the coal lock which needed to be released before conditions returned to normal.

The absence of fresh coal feed led to increasing offtake temperatures. Methane, ethane and ethylene are formed by hydrogenative pyrolysis of the volatile matter in the coal in a zone somewhere at the top/middle of the bed and some of the CO_2 will be formed from pyrolysis in this zone and it is clear that the low bed level interfered with this zone, decreasing the amount of pyrolysis products. On restoration of feed the zone was restored and momentarily enlarged, resulting in a greater than normal amount of product.

This phenomenon highlights the importance of good coal feed to the gasifier to maintain the "fixed" bed top level. Poor coal feed can result through mechanical failure, for instance, of the bottom cone shroud, but in the above example the problem appeared to have been caused by wet dusty coal hanging up at the coal lock bottom.

Unfortunately, in the Westfield Gasifier used for these tests, the distributor volume bei , that of the original Lurgi Gasifier now utilised as a Slagging Gasifier in the Westfield Gasifier, is small in comparison with throughput, and cannot adequately act as a buffer against coal hang up in the feed system. This problem is corrected in commercial design.

The irregular flow of solid fuel down the shaft, 2 above, is by far the commonest manifestation of imperfect bed behaviour, and may always be incipiently present, to be only noticed in the more severe instances.

The irregular solid fuel flow phenomenon occurs with greater frequency at high loads and very low steam/oxygen ratios. It is promoted by tar injection to the gasifier top and by low stirrer speeds and by tar injection to the tuyeres.

Optimisations of gasifier conditions to reduce the phenomenon needs to take several variables into account including gasifier design. However, it should be stressed that, during the EPRI contract, whilst little effort was made towards complete optimisation, no serious bed upsets were generated from this phenomenon and it appears unlikely that those that did occur would have interfered at all with a gasifier-combined cycle.

A third type of bed phenomenon can be recognised which appears to be mainly restricted to highly swelling and caking coals of high fines content, and this manifested itself on EPRI - 03. The incidents are characterised by an increasing torque on the stirrer/distributor system. Providing that the system can be restored back to normal operation quickly, the gasification processes carry on smoothly with no marked upsets and interference from the above phenomenon, and this is generally the case.

The third phenomenon, described above, if it persists, may interfere with gasifier operation but can be overcome with careful attention to gasifier design and operating techniques.

The fixed bed generally behaved very well throughout the EPRI project, although all three categories of behaviour described above were present. Given proper attention to design, there appears little likelihood of poor feed through the coal lock and stirrer system happening on the type of Pittsburgh 8 coal used here, and poor bed behaviour for this reason is only likely to occur if the coal itself is of poor quality, perhaps high in fines and very wet. Unwashed coal could present problems if the unwashed element comprises a sticky clay. Irregular flow phenomena, as described above, are likely to be always present to some extent on all coals under most conditions, but on Pittsburgh 8 are unlikely to interfere with normal process operation. Given further optimisation, and attention to design, it is likely that the occurrance of this behavior can be further minimised.

Bed channelling (4), is probably a phenomenon which is always incipiently present. This phenomenon was not markedly present during the EPRI contract and never caused any bed upsets. It is probably characterised by an upward spike in offtake temperature profile which then returns back to normal. In bed channelling, the closing of a channel will simply restore the temperature back to normal and coal flow is not interfered with. Very occasionally the CO_2 at the offtakes may increase slightly and the bed DP's* fall, perhaps indicating a large channel in the center of the bed.

For a given steam/oxygen ratio and loading on a fixed coal type with a standard amount of flux, gas analysis, and hence gas CV+ is relatively very stable for the Slagging Gasifier. Because of the nature of the fixed bed there are small fluctuations in concentrations of all gaseous components which occur even when the gasifier

*Differential Pressure
+Calorific Value - net or gross heating value

is running steadily; the widest observed variation occurs in the minor components CH_4, C_2H_4, and CO_2. The majority of methane from the Slagging Gasifier is thought to be generated from the destructive pyrolysis of the volatile matter in the coal, in the presence of large partial pressures of hydrogen, in the middle/top of the fuel bed. Lesser amounts of methane will be formed further down from direct char hydrogenation in the reaction zone at the bottom of the gasifier. Ethane and ethylene will also arise from similar, concurrent reactions in the middle of the bed; their relative concentrations, in the presence of large amounts of hydrogen, correspond to equilibrium at 800 - 900°C under Slagging Gasifier conditions, which may be indicative of the temperatures prevailing in the above mentioned pyrolysis zone.

Serious interference with the pyrolysis zone can occur if the bed level becomes too low, when methane, ethane and ethylene concentrations dipped sharply as the bed level dropped. Some perturbation of this zone can occur during an irregular fuel flow, although the effect is less marked.

During incidents when the stirrer torque rises sharply there is also evidence for a small rise in methane concentration.

The majority of the carbon dioxide is likely to be generated in the combustion zone in a series of reactions, thus:

$$C + O_2 = CO_2 \qquad (1)$$

$$CO_2 + C = 2CO \qquad (2)$$

$$CO + H_2O = CO_2 + H_2 \qquad (3)$$

Although significant amounts can be formed from pyrolysis reactions involving the oxygen in the coal higher up the bed, a slight fall in carbon dioxide concentration can occur when the bed level is low. Poor gas/solid loading in the middle of the reactor, which is thought to occur during coal flow irregularities, can, however, account for relatively high concentrations of CO_2 formed from reaction (1) escaping the raceway, perhaps by channelling up the walls. High torque incidents appear to have little effect upon the carbon dioxide concentration. Variation of the calorific value is unlikely to be a problem in the operation of a combined cycle system using the Slagging Gasifier unless there is some mechanical breakdown. Examination of the methane chart during the "starved fuel" incident shows that the methane percentage in the make gas was reduced to below half its normal level, which would have caused a reduction in the flare gas CV of approximately 10%. This type of incident is more likely to be caused by a mechanical problem and is not a normal feature of the gasifier's operation.

The coal flow irregularity phenomenon can also have a slight effect upon the control systems. The gas production rate was held constant throughout one such incident and it can be concluded from the above that the effect of the incident is to slightly reduce both the CV of the gas and its production rate. It appears that during the incident the process is less efficient, this loss in efficiency manifesting itself in more than normal sensible heat losses from the bed top, due perhaps to less devolatilisation/pyrolysis/hydrogenolysis reaction occuring in the bed.

In the flow control mode, for this fairly severe incident, the steam/oxygen flow rates increased by about 6%. This perturbation on the gasifier may indeed help to lessen the effect of such incidents on the bed, with the higher flows opening up new channels in the bed.

When operating the gasifier in flow control the steam/oxygen flows will fluctuate slightly although controller settings can be selected to minimise this fluctuation, perhaps to about $\pm$ 3% under the best conditions. However, when locking on raw gas the problem is more severe, with an increase in flow of up to 22,500 SCF/H steam/oxygen blast over a five minute period (four times an hour on standard load). If this system is practised, the oxygen and steam systems will need to have sufficient flexibility to provide this increase, which represents about 14% of standard load and 56% of quarter load. In view of this it would perhaps appear desirable to lock on nitrogen. The swing can be reduced significantly by a much slower pressurisation of the coal lock, which would certainly be done at lower loadings.

Increase in the steam flow to the gasifier will not cause operational problems as considerably higher steam/oxygen ratios than standard were used, with no difficulties arising, during EPRI - 01. It did cause an increase in CO_2 level in the make gas and had a small effect upon gas CV. Increasing the oxygen throughput with regard to steam was practised in EPRI - 02 and 03 when steam/oxygen ratios of 1.1 and 1.2 were used for periods. No process problems were encountered; thus the Slagging Gasifier is not unduly sensitive to steam/oxygen ratios, and precise control of this parameter, although desirable, is not essential for continuous smooth operation of the gasifier.

Section 4

DEMONSTRATED RAPID LOAD CHANGES

The load changes made during the EPRI contract are summarised graphically in Figure 4-1. The notational theoretical responses required are taken from D.N. Ewart et al, 1978 American Power Conference. Note that the Westfield responses are generally faster.

Details of the load changes are given in the individual run reports. Discussion of the dynamics accompanying some of these load changes is presented in the next section and in the run reports.

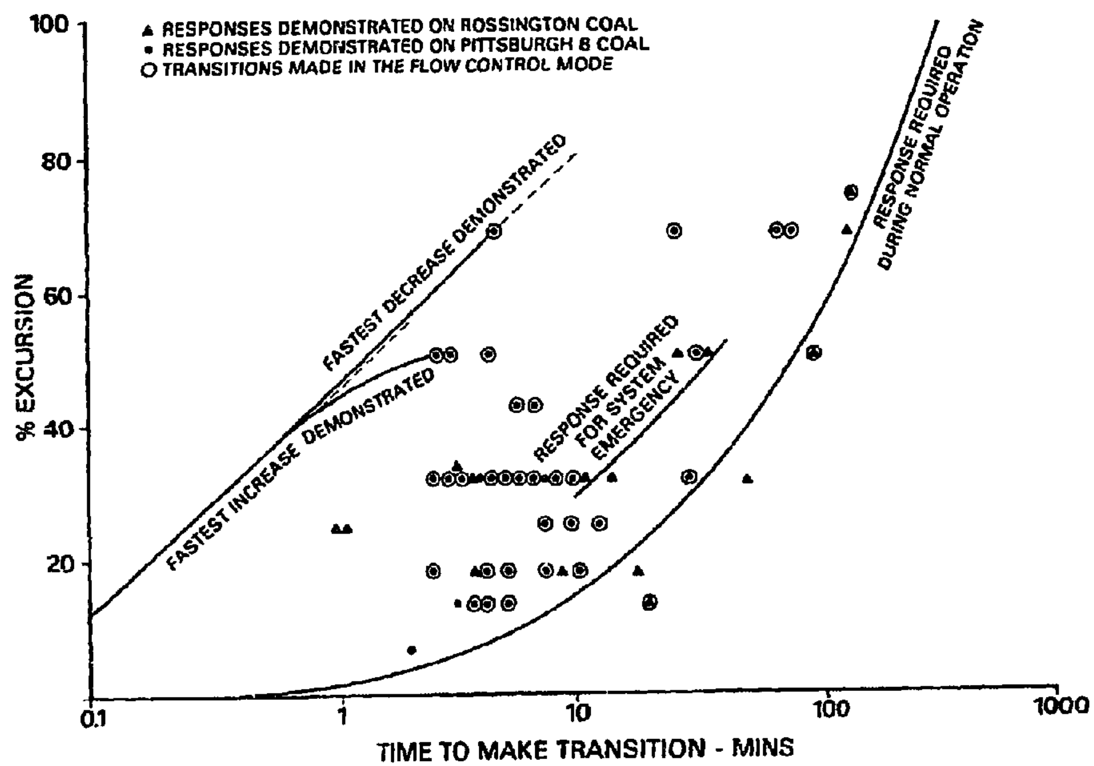

Figure 4-1. Summary of Load Changes

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Section 5

PROCESS DYNAMICS ACCOMPANYING LOAD CHANGES

GENERAL BEHAVIOUR DURING TRANSIENTS

During the EPRI project the gasifier showed that it was capable of making rapid changes in load without interfering with gasifier stability; these rapid changes exceeding the original specifications, before the contract, by EPRI. It thus appears that the Slagging Gasifier can respond reliably to load changes at a rate which will more than satisfy the worst case of a single gasifier associated with a single gas turbine.

There is virtually no change in the gas produced due to load variations except for an increase in nitrogen content as the load decreases because this particular gasifier used nitrogen in the slag removal system. Slight variations in downstream pressure and CO_2 and H_2S concentrations will be easily dealt with by acid gas removal systems such as Rectisol and Selexol.

The effects of rapid flow variations on the gasifier will be described in the following section and in the individual run reports. A flow spike or flow reduction has no significant effects, but there is a noticeable effect upon bed pressure drop when the gas production rate is increased rapidly. They remain at about double their normal values for more than 15 minutes. There is no evidence that these rapid load changes promote bed disturbances.

The flare gas CV was nearly constant through all the changes imposed upon the gasifier. The greatest variation observed was a 5% increase for approximately 20 minutes after a very rapid load reduction; this was caused by an increase in methane concentration, and a slower change would probably reduce this small increase.

In conclusion, the effect of rapid change of throughput on the major performance variables during a change are virtually indistinguishable from the normal running situation except for the gasifier pressure which shows a typical slight increase or decrease at the change. The differential pressure levels through the bed do show a significant increase when the gas make is increased rapidly. They then take at least 15 minutes to reduce to their normal values at the new throughput. Bed DP's, although of value in understanding the behaviour of the gasifier, do not affect the gasifier's output in any way and, therefore, are not regarded as major performance variables. The surge in bed DP's is probably caused by the narrow gas flow channels of a particular size and shape established in the bed at the low rate which are difficult to break down, but as this restricting bed reacts new upper layers will be laid down so as to allow the high gas flow rate and, therefore, the pressure drop slowly reduces to normal.

A microprocessor would appear to be unnecessary to accommodate the required rates of change; but if immediate, virtually instant changes are required, then it could be tuned to give an exact response for a particular coal and plant.

THE RESPONSE TO PULSE CHANGES

At the conception of the EPRI project, the Sponsors were concerned that the dynamic response functions of the gasifier and its systems be measured. The preferred method was one of pulse testing which, in the simplest case, would consist of making a sharp downward spike in the gasification medium, ideally lasting one to two minutes; this spike would then produce a similar reduction spike in overall gas flow as measured at the back end of the plant just before the final flare control valve. The transfer function for the gasifier and its systems could then be derived by using a computing method developed by EPRI.

The above ideal approach depended upon the gasifier working in pressure control at the back end and giving a steady flow output under constant conditions. However, the gasifier and its control systems do not produce this ideal behaviour and in pressure control the flow profile at the back end is irregular and corrupts the expected downward flow spike. In fact, after initial commissioning, the run programme concentrated on working in the back end flow control mode, which meant that such a spike could not be done. In this mode a spike in flow at the back end of the plant, produced by moving the flow set point, has the effect of producing a pressure rise at the gasifier, which is countered by fall in gasification medium input to the gasifier.

However, during normal running, a concentration spike in nitrogen in flare gas can be introduced into the base of the gasifier, and the effects of this spike were investigated in detail during EPRI - 03, and the results have been analysed.

During this run, a sharp rectangular pulse in the nitrogen supply to the gasifier purge/control systems was produced over 20 seconds. Since nitrogen is inert this pulse can be used to identify details of the flow pattern through the gasifier and its systems, as continuous analysis for nitrogen is done just before the flare valve, using a mass spectrometer coupled to a small side stream purification system.

Typical output peaks taken during EPRI - 03 are shown in Figure 5-1. These are at 160,000 SCF/H oxygen loading on Pittsburgh 8 coal and the input pulse is followed by an N_2 peak measured by the mass spectrometer. Examination of this output trace suggests that in many cases a second minor peak occurs, sometimes with a sufficient delay to overlap the next pulse. This ghost peak may be caused by interaction of the two cooling streams.

Figure 5-2 shows the nitrogen output concentration from four such pulses for a period of running at 160,000 SCF/H oxygen loading. The figure also shows an "average curve" which is representative of all the curves and is arranged to start and finish at the same level, ignoring any secondary peak.

The transportation lag in the system can be calculated from the known volumes and conditions of the cooling streams. These have been calculated as follows:

Reduced volume of:

(a) (½ gasifier + No. 1 cooling train) = 20557 SCF

(b) (½ gasifier + No. 2 cooling train) = 26178 SCF

Table 5-1 shows the calculated times to traverse the cooling train using the above data and compares it with the actual times measured from the charts at three different loadings.

Table 5-1

TIME TAKEN TO TRAVERSE COOLING STREAMS

Oxygen Loading (SCF/H)	Time to Traverse (Train-(Calctd) secs (a)	(b)	Time to peak N_2 Variation (Secs)	Time to first N_2 Variation (Secs)
50,000	557	710	330 - 360	270 - 300
130,000	214	272	170 - 180	130 - 140
160,000	174	220	192	152

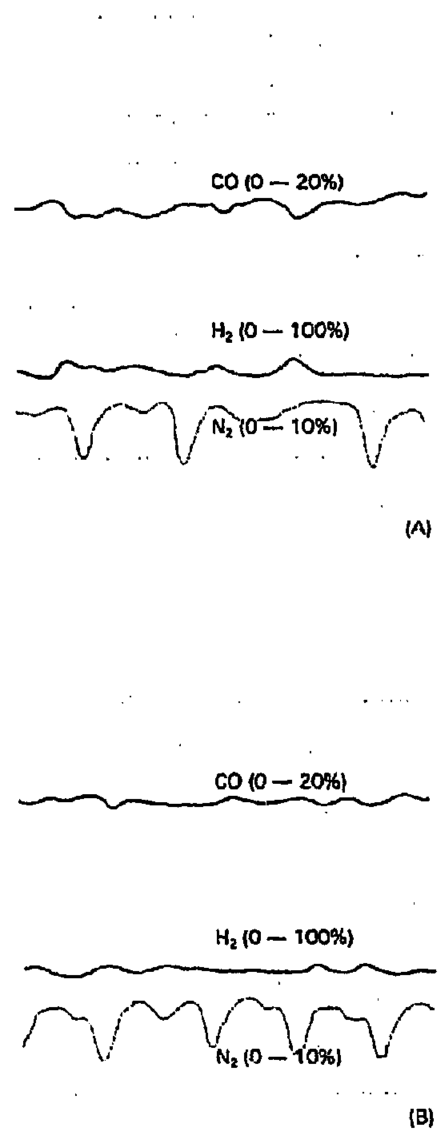

Figure 5-1. N_2 Concentration Following Pulses
Chart Speed: 2 Min./CM

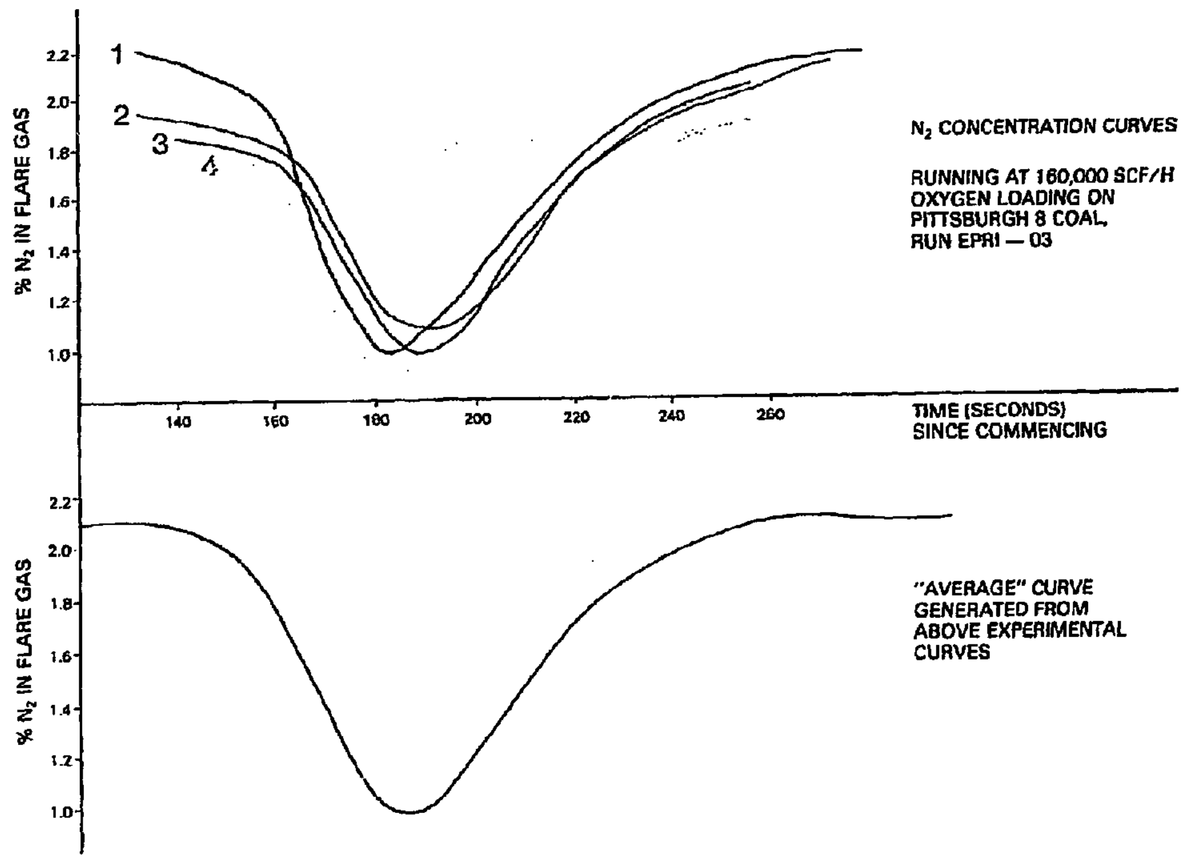

Figure 5-2. Nitrogen Profiles in Flare Gas, EPRI - 03

Table 5-1 suggests that the vessel volumes have been greatly overestimated by ignoring dead space and assuming plug flow, as the agreement between observed and calculated results is not good.

Frequency response analysis was carried out from the curve in Figure 5-2 by using a rectangular pulse for input of duration 20 seconds. The height of the input pulse should be adjusted so as to give conservation of nitrogen, i.e., so that the net area under each pulse is the same. This has not been done since this only assures that when input and output nitrogen are measured in the same units the overall gain of the system is 1. The transportation lag of 130 seconds in the system has been ignored so as to give an instantaneous output. The Bode diagram in Figure 5-3 was obtained using the programme supplied by EPRI.

An approximate transfer function has been obtained by a general least square analysis of the two Bodo plots, which yields

$$e^{-14.4s}/(14s + 1)^3$$

The above function is also plotted in Figure 5-3. With the additional 130 second lag the function becomes

$$e^{-144.4s}/(14s + 1)^3$$

Since the high frequency data is doubtful this transfer function should be treated with caution. The precise utility of the function is not apparent since it might be expected that the cooling train would be represented as a series of first order lags and a transportation delay. If further details of this function are needed it would be useful to obtain information to define the Bode diagrams at higher frequencies by using a short input pulse and, were it possible, at a larger variation in N_2 flow, so as to retain accuracy in the mass spectrometry measurements.

Attention was also given to the behaviour of three gasifier variables across a load spike shown. The gasifier pressure is a function of both steam/oxygen inlet rate and the down stream valve position. In the experiment a proportional/integral controller is used to eliminate one dependency by linking inlet flow to existing gasifier pressure. The constants of the controller are known.

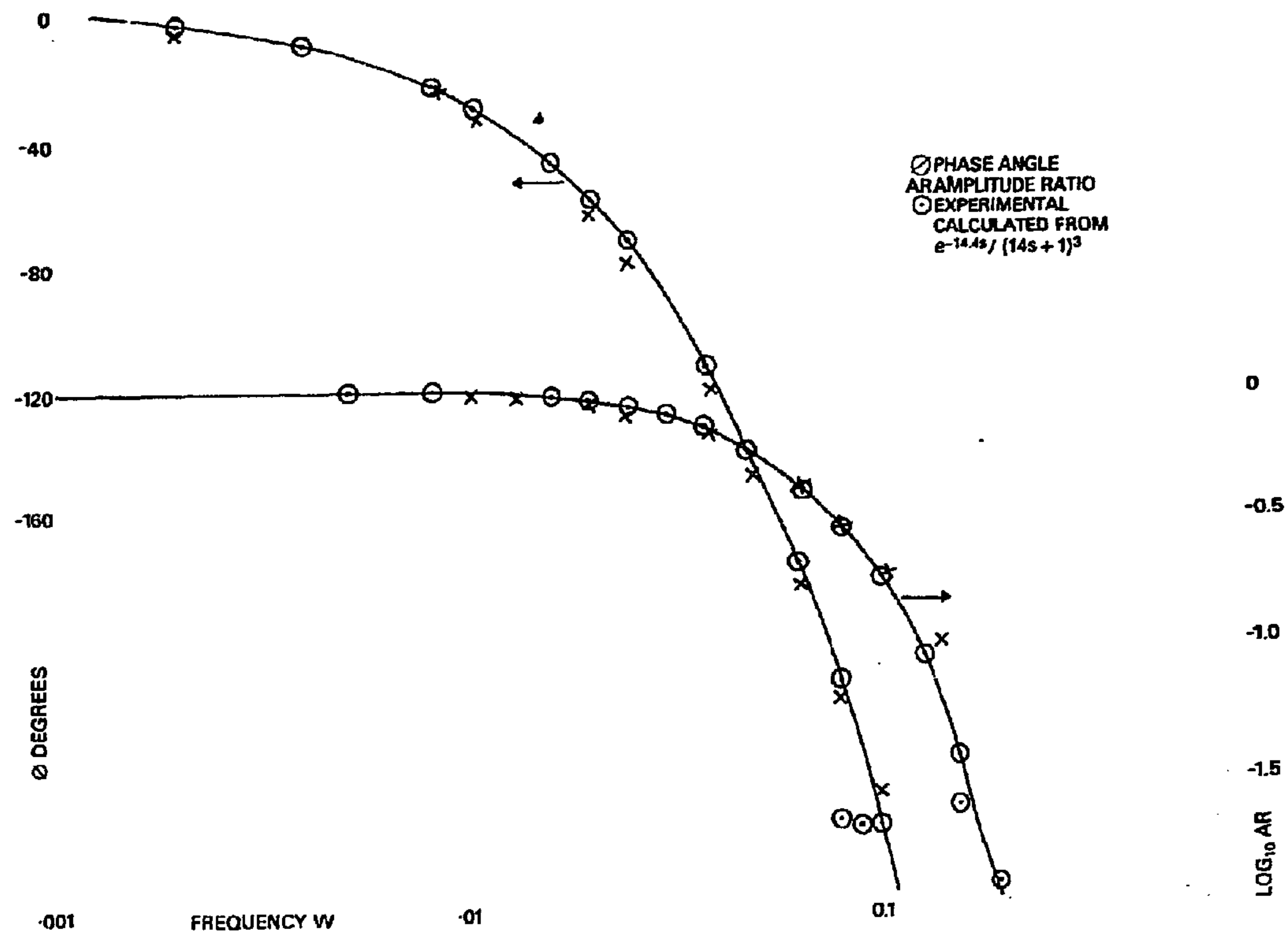

Figure 5-3. Bode Plot for Output Curve of N_2 Concentration in Previous Figure

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For pulse testing it would be desirable that the gasifier is not disturbed in some extra manner beyond the input pulse. It is obvious, in the example given, that coal locking has produced significant changes in gasifier pressure and gasification medium flow and therefore it is difficult to relate the oxygen flow changes directly to the output flow level.

Smoothed data from Figure 5-4 are plotted on a common time axis in Figure 5-5, allowing for differences in zero time in the chart recorders. This suggests that the pressure change caused by altering the outlet flow takes about 12 seconds to travel up the cooling train and shows that coal locking has a major influence on inlet flow rate. There may be a possibility that the PI controller gain is slightly too high, causing unwanted flow fluctuations, but this would require much more investigation and more details of the gasifier system pressure history.

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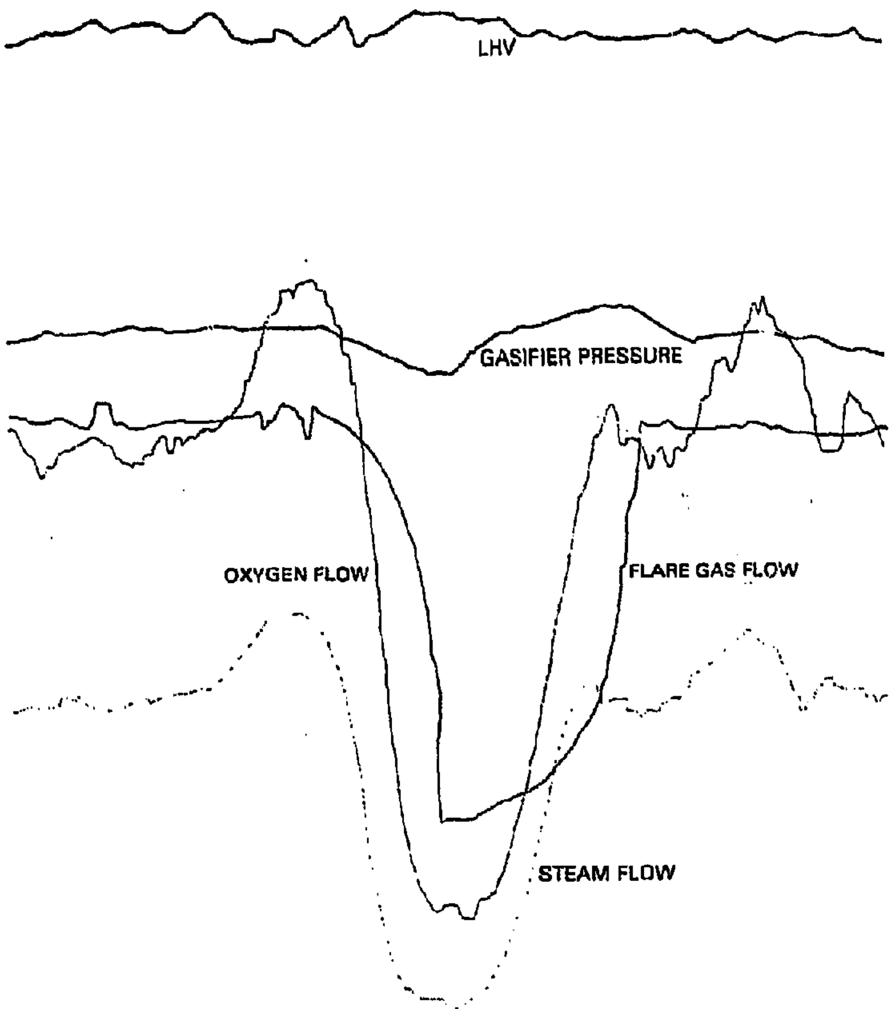

Figure 5-4. Gasifier Parameters Across a Spike in Load Nominally from 160,000 - 110 - 160,000 SCF/H Oxygen

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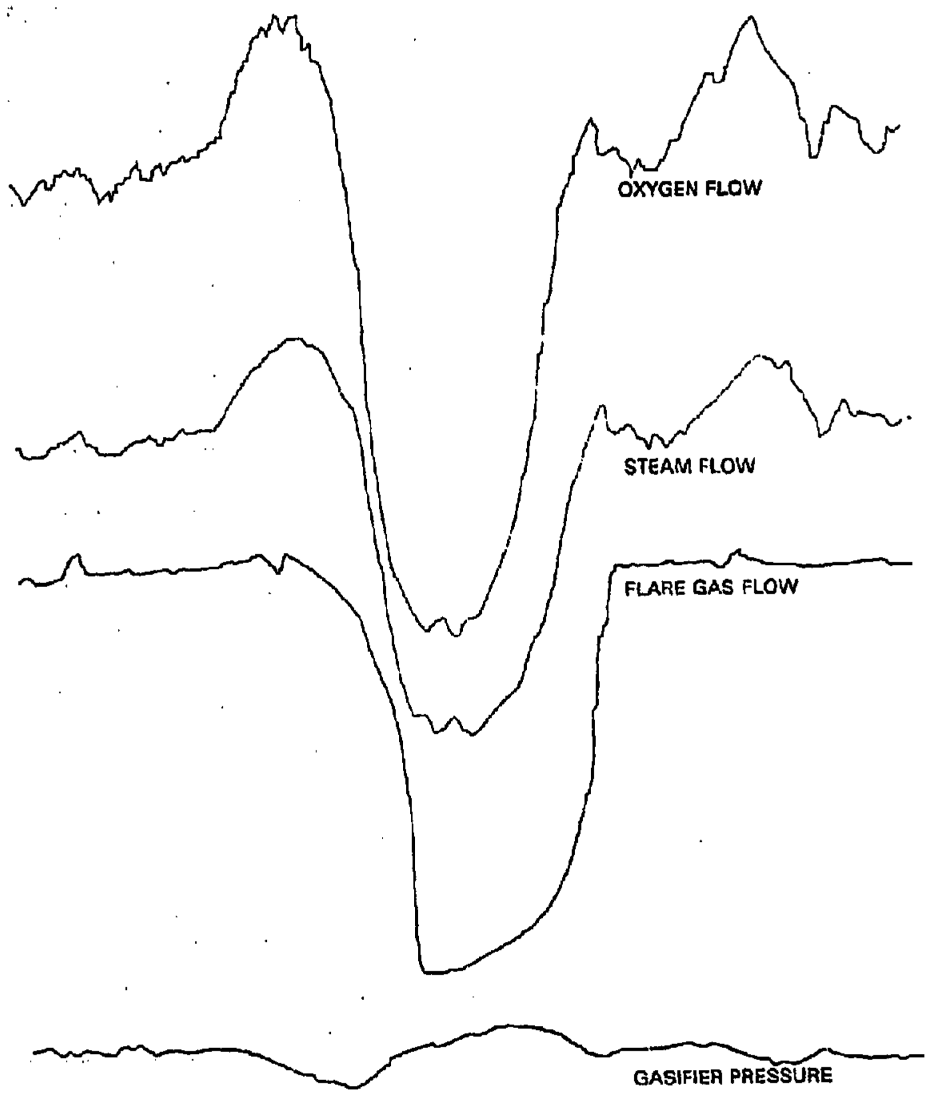

Figure 5-5. Tracing of Chart Data for 10:15 December 79
Vertical Scales Adjusted for Clarity
Time Scale Adjusted for Chart Zero

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Section 6

MECHANICAL PERFORMANCE

Sustained gasifier operation, necessary in commercial situations for periods of the order of nine months, relies heavily on the integrity of the various components of the Slagging Gasifier reactor. Of particular interest and relevance are:

- Coal feed and related systems which include systems for injecting tar to the gasifier top.
- The gasifier hearth.
- Slag tap systems.
- The quench chamber and slag tapping and removal systems.
- Gasifier standby procedures which may be necessary for on line remedies to faults in the above systems.

A more detailed discussion of the top of the bed systems and coal feed systems is found in the Lurgi report (Section 7). The importance of keeping a fixed bed top has been stressed earlier in this report, and this means that coal charging and feeding systems, as well as being of the right design, must be very reliable; and Westfield experience with a few minor exceptions, has proved this to be the case. It should be remembered that the feed system is designed to handle lump coal, so that problems can be anticipated in trying to feed fine coal, particularly if it is contaminated with dirt and dust.

The stirrer system gave a good performance through most of the project, being able to produce a char suitable for gasification from a highly caking and swelling coal. With a high fines content in Pittsburgh 8 coal there was a tendency for a high torque on the stirrer/distributor system to occur, especially at high speed.

Table 6-1 summarizes the properties of the slag which passed through the hearth in the four EPRI runs. For all the runs the slag analysis was fairly consistent and the slag temperatures will have been roughly constant as most of the running was done at about 1.30 steam/oxygen ratio.

TABLE 6-1. HEARTH DATA FROM EPRI PROJECT

Run No.	EPRI-01	EPRI-02A	EPRI-02B	EPRI-03
Coal/ Flux	Rossington Limestone	Pittsburgh 8 BFS/Limestone	Pittsburgh 8 BFS/Limestone	Pittsburgh 8 BFS/Limestone
Total Run Time (Hrs.)	120	$36\frac{1}{2}$	140	110
Slag Analysis (%)				
Al_2O_3	23.8	17.25	16.86	17.55
SiO_2	34.36	38.61	40.26	39.44
Fe_2O_3	7.66	4.92	7.57	5.39
CaO	25.64	28.84	25.47	26.8
MgO	1.94	7.52	6.88	6.25
Na_2O	2.02	0.67	0.44	0.40
K_2O	1.38	1.21	1.13	1.18
Silica Ratio	48.2	47.8	50.19	50.27
Base/Acid	0.66	0.77	0.72	0.70

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The slag tap system at the hearth bottom worked well during the EPRI project. The system had run before the project and added a further 415 hours to their total running time. The slag tap wɾ.s visually examined between runs and appeared not to be wearing or being damaged with time.

For EPRI - 01, on Rossington coal, fluxing levels and slag tapping experience had been gained over previous runs and the performance here can be regarded as being close to optimum. For the given hearth geometry, turndown to 30% of standard load was achieved with good operation of the system.

During EPRI - 01 an error in fluxing led to the addition rate being 5 - 10% below optimum but this had little effect.

During EPRI - 02A, while on Rossington, fluxing was inadequate because of a blockage in the fluxing feed to the coal lock, but this was quickly recognised and easily remedied. Fluxing on Pittsburgh 8 was initially carried out on limestone, and poor slag tapping was obtained due to underfluxing before good conditions were obtained on BFS flux. For run - 02B, fluxing on Pittsburgh 8 was established immediately on BFS and good slag tapping was obtained. This run demonstrated the capability of the hearth and slag tapping systems to make a rapid and complete recovery following a hot standby.

Slag tapping was good throughout EPRI - 03 and changing from screened to unscreened coal did not effect the tapping quality nor did changing fluxes.

The slag tapping systems worked well during the EPRI project and were shown to be capable of dealing with all turn up and turn down conditions used.

Although fluxing on Pittsburgh 8 was not properly optimised during the project, it is apparent that significant amount of limestone flux are required to operate the slagging gasifier on this fuel, with flux/ash ratios likely to be of the order of 1.0/1.0. (in the two runs on Pittsburgh no attempt was made to optimise fluxing which was set at 1.2 to give a good performance. An optimum ratio less than the above 1.2 can therefore be possible).

Tuyere systems generally gave satisfactory performance during the EPRI project.

During EPRI - 03, tar injection through tuyeres was attempted. There was a 24 hour period of tar injection to the tuyeres.

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Slag tapping and slag removal systems worked well during the EPRI contract and these systems are ready for commercial duty.

The problem with the quench chamber during EPRI - 02B was mainly due to control systems not being set up properly. Further improvement can be obtained by due attention to design.

The incident referred to above illustrates the value of a gasifier that can readily be put on standby to effect necessary repairs, either in the gasifier or its ancilliary systems; and then be expected to return on line quickly and reliably. The Slagging Gasifier is an example of such a gasifier, as it has demonstrated often in the past. Providing that the gasifier is running properly, with good bed and hearth conditions, it appears that return from a standby situation is certain, provided the proper simple routines are followed. The standby can either be hot, at pressure, or cold at atmospheric. In the latter case there appears to be no limit to the length of standby period which can be reliably obtained.

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Section 7

PERFORMANCE OF LURGI PROPRIETARY EQUIPMENT
(By Lurgi Kohle und Mineralöltechnik GmbH)

This report is produced with exclusive reference to the general performance of Lurgi proprietary equipment. Special emphasis is given to the influence of the performance of the stirrer/distributor system on the general gasifier behaviour. Depending on the type of coal, an optimum stirrer speed has been identified to give the best gasifier performance for steady state operations.

The stirrer/coal distributor system performed satisfactorily at all load changes.

Tar recirculation on the distributor has, in this particular equipment, a noticeable impact on gasifier performance.

There is no conclusive evidence of the influence of an increased fraction of fines in the feed coal.

Although the gasifier performed well under all operating conditions, several bed upsets occurred in regular intervals. These bed upsets, however, had no lasting effect on the gasifier performance.

OBJECTIVES

The objectives of the EPRI-trials on the slagging BG/Lurgi gasifier in Westfield include the test of the performance of the available stirrer/distributor equipment with respect to gasification of two different coals. Due to the availability of experience in fixed bed gasification, Lurgi was invited to provide some input in outlining the run schedules and in covering and evaluating the trials as far as the Lurgi proprietary equipment is concerned.

Because of the countercurrent operation, fixed bed gasification is inherently more efficient than fluidized bed or entrained bed gasification. Considering the large volume of coal normally present in the reactor, there is reason to assume that fixed bed gasification is operationally safer than other modes of gasification.

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The EPRI-trials consisted of three individual runs. This report covers only those particular aspects of the e.. .riments which are related to the Lurgi proprietary equipment.

The main points considered are:

- Optimizing stirrer speed for different loads of two different coals.
- Evaluation of tar injection to the coal distributor.
- Influence of high fines content in regard to the upper part of the gasifier.
- Evaluation of bed upsets.

Some convenient performance criteria for good fixed bed gasification are:

- Low offtake temperatures of the raw gas.
- Low pressure loss of the fixed bed.
- High calorific value of the raw gas.
- Low standard deviations of all dependent variables.

GENERAL ARRANGEMENT OF COAL DISTRIBUTOR, STIRRER AND TAR RECIRCULATION SYSTEM

Coal is fed from bunkers via a manually operated lock hopper, into a coal distributor. Flux is fed simultaneously with the coal into the lock hopper via a vibrator feeder. The coal distributor is hydraulically driven and feeds the coal in layers onto the coal bed. Tar is fed into the top part of the coal distributor to wet the coal in the distributor. The arms of the stirrer break up any caked lumps which may form during devolatization in the upper part of the reactor.

A particular coal distributor and stirrer system is usually designed for a set of particular coal properties and for particular operation conditions of the reactor. Although the available system in the Slagging Gasifier is certainly not suitable for all coals, serious operational problems did not appear.

ANALYSIS OF EXPERIMENTAL RESULTS

Stirrer speed optimization for Rossington and Pittsburgh 8 coal

Figure 7-1 shows the dependence of offtake temperatures on the stirrer speed (Run No. 1). Under certain conditions the performance of the gasifier improves with greater RPH (Revolutions Per Hour).

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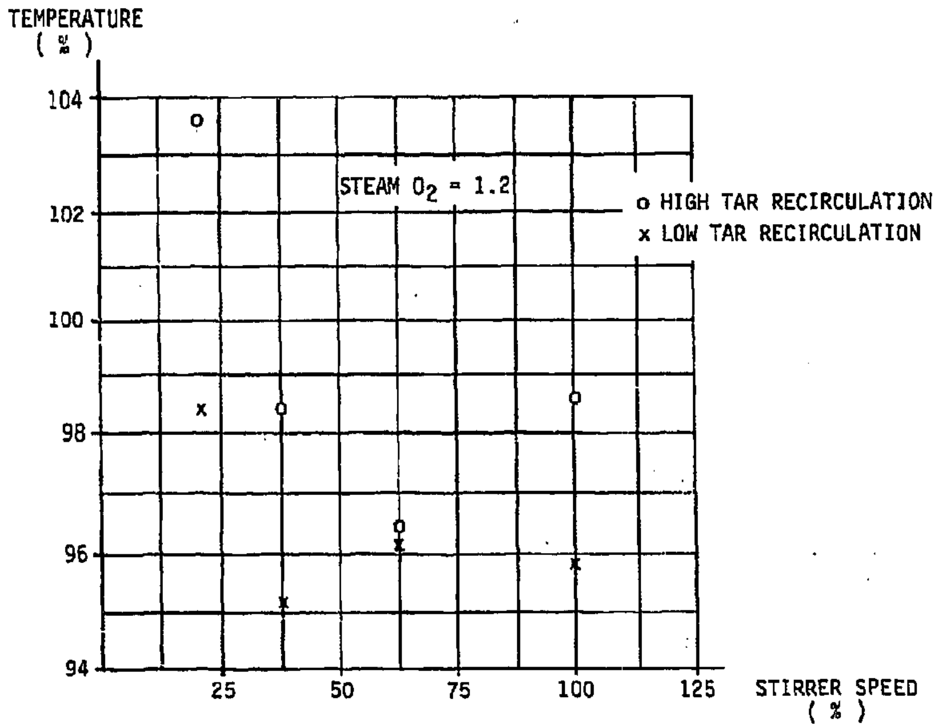

Figure 7-1. Mean Offtake Temperature at Different Stirrer Speeds and Tar Recirculation Rates

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This effect is even more perceptible when evaluating run No. 2. Table 7-1a presents a quantitative evaluation of three stirrer speeds using offtake temperatures, bed Δp and calculated lower calorific value. Table 7-1a shows a rating of nine hourly mean values in which the most favorable value (lowest offtake temperatures and bed Δp's, highest calorific value) is given the number 1, the least favorable, the number 9. The elements of the matrix show the matching stirrer speeds. The table shows consistently the lowest value to have the worst rating and the highest value the best followed closely by the intermediate RPH. In the same manner, the standard deviations are evaluated (Table 7-1b). For the sake of clarity, Table 7-1b shows only the two lowest and two highest ratings of the nine periods investigated.

Considering the fact that the intermediate value is close to the highest value in Tables 7-1a and 7-1b, this intermediate value is assumed as a base for the relation between load and stirrer speed (Figure 7-2). Comparing Rossington and Pittsburgh 8 coal, a distinctive influence of the type of coal is evident.

Stirrer speed settings at increased fines content of feed coal

Based on previous experience of the influence of fines in "dry ash gassifier," the stirrer speed has been increased proportionally with the fraction of fines in run No. 3. However, the fines content is considered to be very moderate compared to similar gasification tests with dry ash gasifier, where the fines content reached 30% < 1/8".

Stirrer tests with high fines content indicated that increasing the stirrer speed above the previously identified value may improve the bed behaviour. However, the fines content fluctuated considerably during the second half of run No. 3, hence a conclusive assessment was not possible

Table 7-1a

EVALUATION OF STIRRER SPEED TESTS

Variable/Rating	1	2	3	4	5	6	7	8	9*
lower bed p	112	112	87	87	112	87	62	62	62
upper bed p	112	87	112	87	87	112	62	62	62
T_3	112	87	112	112	87	87	62	62	62
T_4	112	87	112	87	112	62	62	62	62
calorific value	112	112	112	87	87	62	87	62	62

Matrix elements: stirrer speed (in %)

*Ratings of 9 mean values of bed pressure losses, offtake temperatures and heating values.

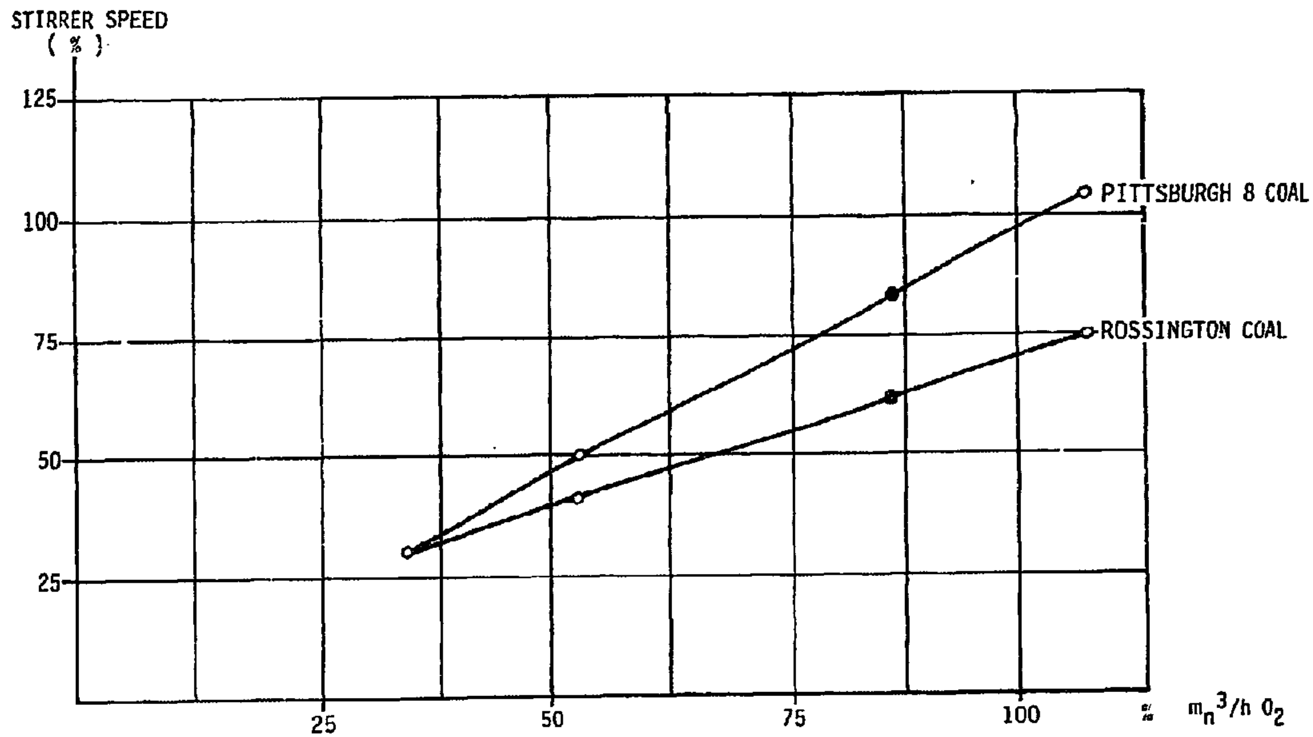

Figure 7-2. Stirrer Speed Vs. Load for Different Coals

Table 7-1b

EVALUATION OF STIRRER SPEED TESTS

Variable/rating	1	2	35	36
hydraulic pressure of stirrer drive	112	87	62	62
lower bed p	112	112	62	62
upper bed p	87	87	62	62
T_3	87	42	62	62
T_4	87	42	62	62
calorific value	42	87	87	87

Matrix elements:
stirrer speed RPH (in %)

EVALUATION OF THE TAR RECIRCULATION

Objectives and analysis of the present system

The objective of the tar recirculation to the top of the reactor is to recycle the tar to extinction in a countercurrent mode with the coal in the moving bed. Furthermore, wetting evenly the coal in the upper part of the reactor decreases the dust content of the offgas.

The tar in the reactor is partially distilled and partially cracked. However, from the experiments conducted within the EPRI-program, no precise information is available concerning the whereabouts of tar fractions.

Impact of tar recirculation on the reactor performance under various operating conditions

Figure 7-1 shows offtake temperatures at two different tar rates; the graph shows consistenly the influence of tar recirculation to the distributor. The performance of the operation with Pittsburgh 8 coal deteriorates with increased tar recirculation.

This is caused by constructional deficiencies and hence uneven wetting of coal. Experimental evidence from other gasifiers has shown that tar recirculation has no adverse operational effects.

The effect of tar on the raw gas analysis is given in Table 7-2. The influence of tar is mainly shown by the increase of standard deviations rather than a change in mean values.

Table 7-2

INFLUENCE OF TAR RECIRCULATION TO THE DISTRIBUTOR ON RAW GAS ANALYSIS

	no tar		max. tar	
	Mean (%)	st. dev.	Mean (%)	st. dev.
N_2	3.36	0.58	3.26	0.56
CH_4	7.13	9.44	6.87	0.89
H_2	27.9	0.62	28.1	3.15
C_2H_6	0.43	0.04	0.41	0.06
C_2H_4	0.14	0.01	0.15	0.02
CO	56.1	1.33	54.8	5.99
CO_2	3.83	0.47	4.16	0.66
H_2S	0.63	0.02	0.59	0.09
COS	0.08	0.002	0.07	0.008

There is no conclusive relation between tar recirculation and dust content of the tar. Test periods were too short compared to the dynamics of settling and sedimentation in the tar separator.

GENERAL REMARKS TO THE OPERATION AT LOAD CHANGES

Load changes, as tested between 100% and 30%, require, in general, adjustments of the stirrer speed according to Figure 7-2.

However, at pulse test (duration up to ten minutes) the stirrer speed can be kept constant without any bed upsets. Ramp changes require approximate proportional changes in stirrer speed. Too high a change leads to instant overfilling of the reactor and hence, to severe upsets (high hydraulic pressure, high offtake temperatures, high bed Δp's).

A typical incidence shows that increasing the stirrer speed too rapidly leads instantaneously to immediate rapid increase in both bed Δp's and hydraulic pressure. The permeability of the bed decreases suddenly and the offtake temperatures rise.

Usually, incidents of this kind can be rectified by lowering the stirrer speed immediately and following gradual increase.

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EVALUATION OF BED UPSETS

The operation of fixed bed gasification is essentially dependent on the dynamics of the coal bed.

Any upset in the smooth downward flow of coal has severe effects on the gasifier performance.

Several times during the experimental runs, bed upsets occurred in irregular intervals which were immediately recognized by strong variations in bed pressure losses and gas temperatures.

These random bed upsets, however, never caused serious operational problems. In most cases, bed upsets rectified themselves without any operational interference from outside.

CONCLUSIONS

It is generally known that the operation of the fixed bed BG-Lurgi Slagging Gasifier is influenced by the performance of stirrer and coal distributor. This phenomenon has been confirmed during these experiments. The stirrer speed has to be adjusted to the particular conditions of this system. Optimizing the stirrer speed leads to significant improvement of the operational behaviour.

The experimental evidence indicates that the operation of the stirrer and coal distributor can be further adjusted for different coal properties.

There is a significant influence of tar recirculation in this particular equipment under any operating conditions. However, tar recirculation never caused any critical operational problems.

There is no conclusive evidence how the increased fines content of the feed coal affects the performance of stirrer and coal distributor.

From the experience with the dry ash gasifier, the stirrer speed is to be increased with increased fraction of fines in the feed coal. However, fines content during the EPRI-trials has been very moderate (approximately 20% < 1/4" vs 20% < 1/8" in "dry ash" gasifier on bituminous coal).

Section 8

PARTICULATES IN RAW GAS

EPRI is concerned about the effect of impurities in the gas from the Slagging Gasifier on the hot components of gas turbines. In particular, two types of attack, namely hot corrosion due to salt layer deposition and erosion due to dust particles, are of interest.

A maximum limit of 0.5 ppm by weight of reactive alkali metals is given as a guideline for fuel impurities. Fuller details of the more important aspects of the above problems are given in Appendix C.

Measurements carried out at Westfield during the EPRI trials allowed the dust concentrations in the gas from the Slagging Gasifier to be calculated for a range of loadings. The dust samples collected were then sent off to the British Gas Corporation London Research Station for further analysis.

DESCRIPTION OF DUST PROBE AND FILTER SYSTEM

The design of a dust probe to sample the flare gas to No. 4 flare isokinetically was based on the British Standard Method for sampling superheated steam from steam generating units (B.S. 3285: 1960).

The location of the dust sampling point in the No. 4 flare line is shown in Figure C-1 in Appendix C. Figure C-2 shows the probe and filter assembly for sampling the gas from the 6" N.B. line, and details about the related pipework and instrumentation are given in Figure C-3 in Appendix C. Steam trace heating of the probe pipework was necessary to prevent blockage of the filter by ammonium carbonats, a problem encountered in preliminary trials prior to EPRI - 01.

The filters were held in the filter holder by a stainless steel mesh disc and a threaded retaining ring. Whatman GF/B glass microfibre filters were used throughout the trials and their technical characteristics are given in Appendix C.

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DESIGN CONSIDERATIONS AND COMMENTS

The design was based on sampling gas produced at an oxygen loading of 160,000 $SCFH^{-1}$ and a steam/oxygen ratio of 1.3, with single flare operation. Orifice plate calculations for the probe were carried out at Westfield on line to the computer at London Research Station, using the orifice plate program developed by British Gas.

Details about the location and diameter of the probe sampling parts calculated as per B.S. 3285: 1960 are given in Appendix C. Eventually, the probe was located in a portion of the flare line of 6" N.B., and the design with four sample parts was adopted.

The probe design, Figure C-2 in Appendix C, had to allow for filters to be changed when the gasifier was on line. Incorporated in the design was a facility for removing the whole probe from the flare line should any blockages occur, allowing it to be cleaned and replaced, again, while on line.

The control system on the probe pipework enabled the flow through the filter to be adjusted so that isokinetic sampling could be achieved for the range of loadings used throughout the EPRI project.

METHOD OF OPERATION OF DUST PROBE

The operating procedures for the dust probe are given in detail in Appendix C. In particular, the valve sequences for the following operations are given:

1. Insertion of probe on line.
2. Removal of probe on line.
3. Changing probe filter on line.

PROBLEMS ENCOUNTERED AND HOW THEY WERE OVERCOME

From preliminary trials prior to the EPRI project, there was a problem with ammonium carbonate crystals blocking the filters. This was overcome to some extent by using steam trace heating on the probe pipework. The effectiveness of this was improved after run EPRI - 01 by lagging the system.

Between test runs, it was noted that liquor had condensed in the flare line at the dust probe sampling point. As there was no drain valve on this section of the No. 4 flare line, the dust probe pipework was purged out for several hours with the flare gas in an attempt to remove the condensed liquor before starting a test.

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Prior to the EPRI trials, no flow controller was used, but the flow was observed to drop off even when reset. This was partly due to ammonium carbonate crystallising out in the pipework. For the EPRI runs, a flow controller and flow control valve were introduced, giving better flow control. After run EPRI - 01, the size of the flow control valve was reduced to give even better control.

In preliminary tests and at the beginning of the EPRI trials, problems were encountered with the filters. On occasion, they either disintegrated or blocked up completely. However, as experience of operating the probe increased, these situations became less frequent. It was discovered that tightening the filter retain ring too much helped cause the filter to disintegrate. It became easier to judge the duration of a test as the EPRI project progressed and so test failure due to blocked filters became less frequent.

RESULTS OF DUST CONCENTRATION MEASUREMENTS

Full details of the dust concentrations measured for the three EPRI test runs are given in Table 8-1. Figures 8-1 and 8-2 show the dust concentrations plotted against oxygen loading, and against gas flow at No. 4 flare for the run EPRI - 03. The data have been corrected for the actual temperatures and pressures being slightly different from the design temperatures and pressures. The criteria for accepting a given result as being isokinetic are given in Appendix C.

Not much useful quantitative information was obtained from EPRI - 01, although valuable experience and expertise at using the probe were gained. In fact, for EPRI - 01, only two isokinetic values were obtained, while for EPRI - 02B, four values were obtained, two of which are suspect because the filter papers were slightly damaged. These suspect results were for the two highest oxygen loadings and may be lower than was actually the case. By far, the most reliable results were from run EPRI - 03, where all the filter papers were intact and all the tests were isokinetic.

Hence, from the results obtained, it is not possible to compare the dust concentrations obtained for the three different coals, namely, Rossington, screened Pittsburgh 8 and as received Pittsburgh 8.

The highest dust concentration measured was for as received Pittsburgh 8 during the test period N of run EPRI - 03. This corresponds to about 3.5×10^{-2} mg. of dust per standard cubic foot of gas, equivalent to 1.3 ppm. w/w.

TABLE 8-1. RESULTS OF DUST IN GAS TEST

Run No.	Dust Sample Number	Oxygen Loading ($SCFH^{-1}$)	St/O_2 Ratio	Corrected Dust Concentration (mg/scf)	Isokinetic Sampling ?	Dust Sample sent for analysis	Coal
EPRI 01	3	80,000	1.3	2.02×10^{-3}	No	Yes	Rossington
	5	80,000	1.3	7.1×10^{-4}	Yes	No	
	7	50,000	1.3	1.06×10^{-3}	Yes	No	
	9	160,000	1.2	4.91×10^{-3}	Yes	Yes	
EPRI 02	10	110,000	1.2	2.60×10^{-3}	Nearly	Yes	Screened Pittsburgh 8
	11	80,000	1.3	3.32×10^{-3}	Yes	Yes	
	12	50,000	1.3	6.93×10^{-3}	Yes	Yes	
EPRI 03	15	80,000	1.2	2.54×10^{-3}	Yes	Yes	As Received Pittsburgh 8
	16	50,000	1.2	9.5×10^{-4}	Nearly	Yes	
	18	130,000	1.2	1.33×10^{-2}	Yes	Yes	
	19	160,000	1.3	3.47×10^{-2}	Yes	Yes	

Dust Sampling Conditions

Pressure 300 psig

Temp. 25°C.

Conversion Factor

1 mg/scf = 37.9 ppm

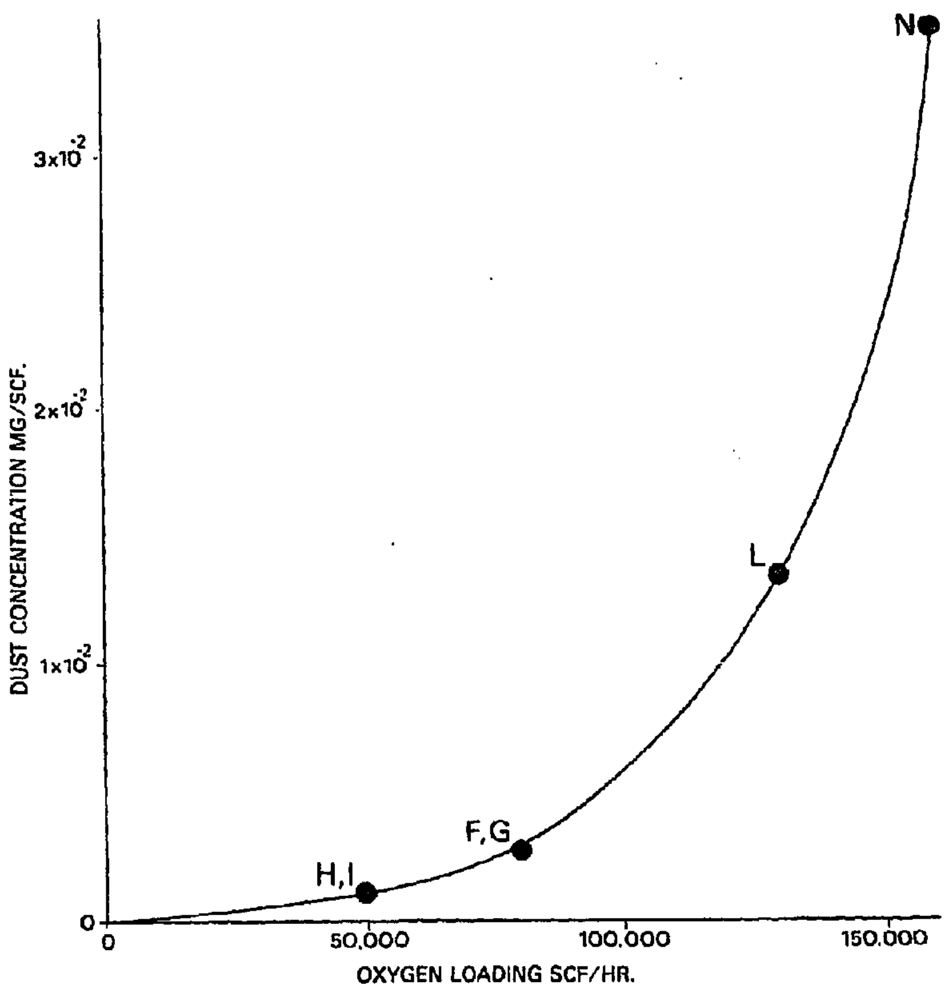

Figure 8-1. Graph of Dust Concentration Against Oxygen Loading
EPRI - 03

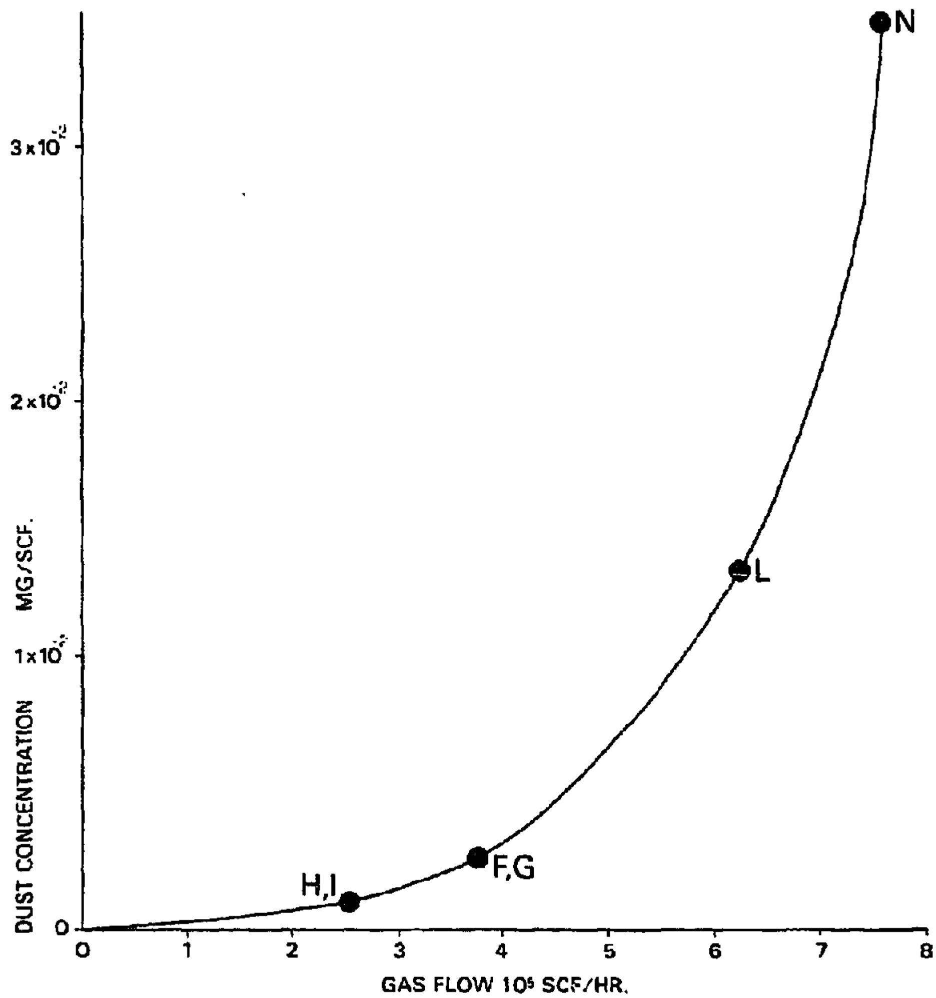

Figure 8-2. Graph of Dust Concentration Against Gas Flow at No. 4 Flare - EPRI - 03

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RESULTS OF DUST ANALYSIS

Dust particle analysis was carried out at the British Gas London Research Station. The particles were observed using a scanning electron microscope, and analysis of individual particles was carried out using an energy dispersive X-ray spectrometer. The material collected on the filter papers was pretreated, before analysis, by solvent washing to remove tar and other organic materials.

Under the contract, two dust samples only were analysed, namely those for samples 3 and 4. The gasifier conditions for sample 3 are given in Table 8-1. Sample 4 was obtained under the same conditions, but in this case, sampling was isokinetic. For sample 4, however, the filter was found to be slightly eroded on removal from the filter holder, making quantitative determination of the dust concentration impossible.

Examples of the dust particles analysed are given in Figures 8-3 to 8-6. Micrographs of a blank filter are given in Figures 8-7 and 8-8. Table 8-2 gives the analyses of the particles numbered in Figures 8-4, 8-6, and 8-8, while Table 8-3 gives a summary of all the particles analysed.

Interpretation of the results should be considered in the light of the information given by EPRI in Appendix C.

None of the particles analysed for samples 3 and 4 had Na as a major component, and only 6% of the particles had K as a major component. This does not take into account the different sizes and weights of the particles. The analyses thus obtained are only qualitative.

An extremely approximate guide to the levels of the various elements is as follows:

major component > 10%
minor component 1 - 10%
trace component > 1%

Care must be taken, however, to avoid reading too much into the above figures.

It therefore seems likely that even for the maximum observed dust loading of about 1.3 ppm, the reactive alkali metal content will be the 0.5 ppm limit specified by EPRI in Appendix C. Only elements with atomic number > 11 were detectable, which

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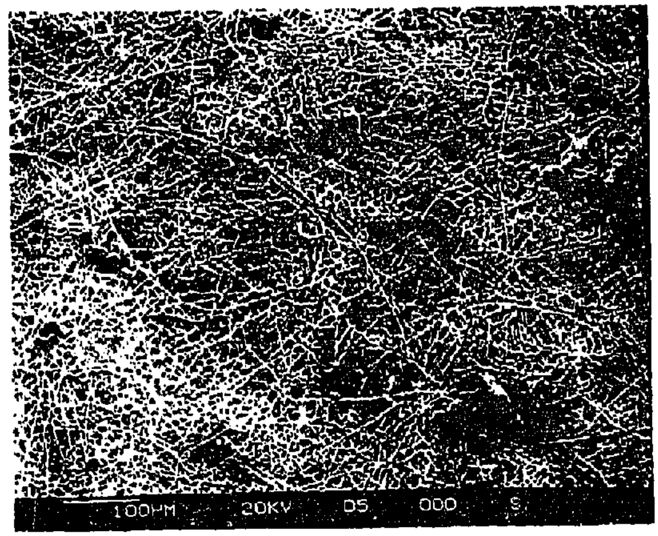

Figure 8-3. Dust Filter Test 3, Magnification x 110*

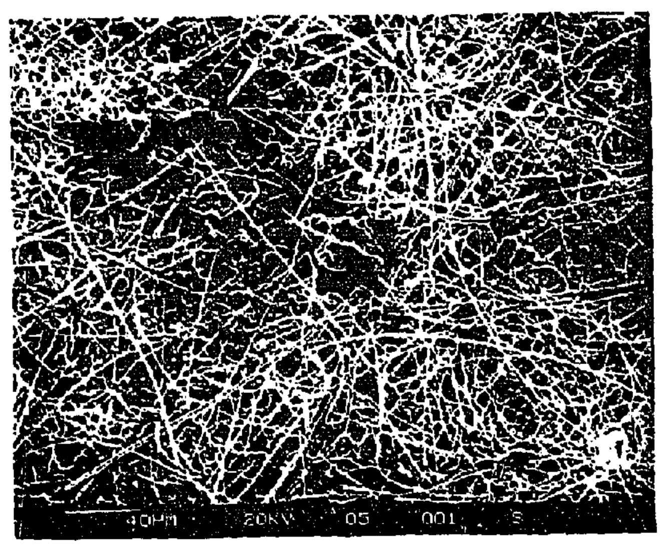

Figure 8-4. Dust Filter Test 3, Magnification x 275*

*Please note that the illustration(s) on this page have been reduced 10% in printing.

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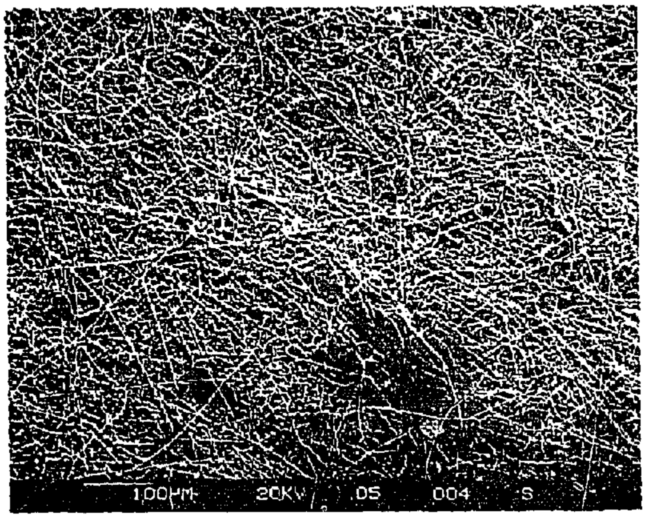

Figure 8-5. Dust Filter Test 4, Magnification x 100*

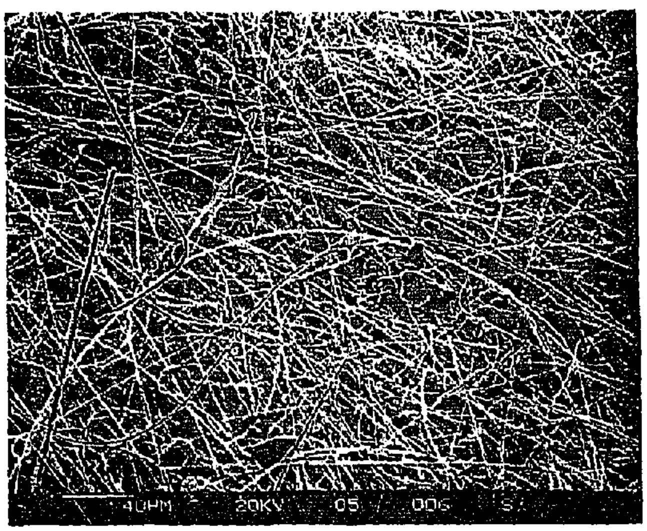

Figure 8-6. Dust Filter Test 4, Magnification x 250*

*Please note that the illustration(s) on this page have been reduced 10% in printing.

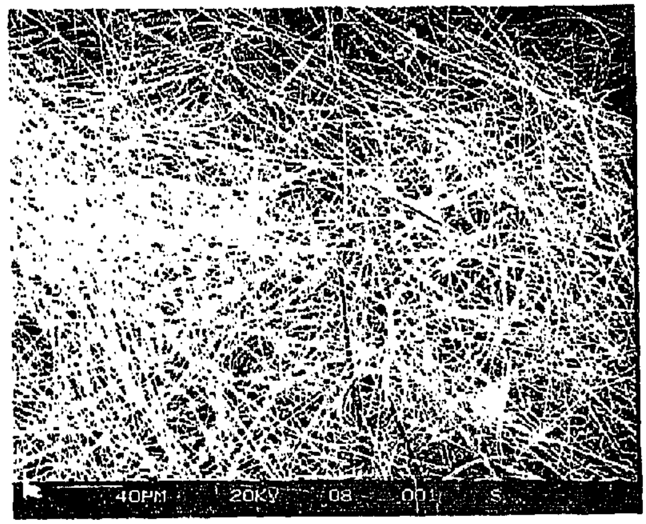

Figure 8-7. Blank Filter, Magnification x 100*

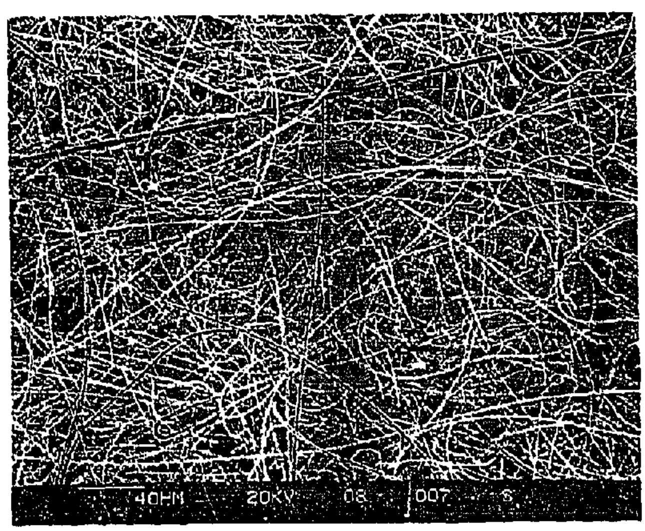

Figure 8-8. Blank Filter, Magnification x 250*

*Please note that the illustration(s) on this page have been reduced 10% in printing.

TABLE 8-2. ANALYSES OF PARTICLES NUMBERED IN MICROGRAPHS

Figure	Test	Particle	Major	Minor	Trace
7.7	3	1	S, Fe	Al, Si	-
		2	S, Fe	Si	Mn
		3	Al, Si, Ca	Mg, S	K
		4	S, Zn	Fe	Si
		5	Si, S, Fe	Na	Al, Ca
		6	S, Fe	Na, Si	Ca
		7	Si, S, Fe	-	-
7.9	4	1	Si, S, Ca	Na, Mg, Al, K	Fe
		2	S, Fe	Si, Cr, Mn	Ni
		3	S	Si	Al, Ca, Fe
		4	S, Fe	Si	Ca, Cr
7.11	Blank	1	Na, Cl	Si	Al, K, Ca
		2	Si. Ca	Na, Mg, K	S, Cl

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TABLE 8-3. SUMMARY OF PARTICLE ANALYSES

	Test 3	Test 4	Blank	units
Number of Particles Analysed	34	24	7	
Proportion with S and Fe major	74	42	0	%
Proportion with Si major	38	83	86	%
Proportion with K major	6	0	0	%
Proportion with K trace	18	25	43	%
Proportion with Na major	0	0	14	%
Proportion with Na trace	21	75	100	%

means it is not possible to comment on the relative levels of akali metal chlorides, hydroxides, oxides and carbonates in the particles.

No vanadium, lead or phosphorus, believed to enhance hot corrosion, were reported to be present.

Complex akali metal sulphates containing Fe, Co and Ni are also thought to contribute to hot corrosion. From the results in Table 8-2, Co and Ni are not important, although Fe was found to be a major component in many of the particles.

PHYSICAL DESCRIPTION OF DUST PARTICLES

Particles were found to range in size from 1 to 50 μm. Attempts to isolate smaller particles were unsuccessful, and because a proportion of the finer particles disappeared into the bulk of the filter, particle size analysis was not possible.

Examples of the individual particles studied are given in Figures 8-3 to 8-6. Small groups of particles were often found, as shown in Figure 8-4 and may have been deposited as part of a tarry droplet.

According to the information in Appendix C, particles below 1 μm in diameter should not cause erosion and in general manufacturers specify no particles > 10μm for a total dust loading of > 35 ppm. The dust particles collected are outside the size restriction, although the total loading should be much less than 35 ppm. Although the hardness of the particles was not measured, problems of erosion may perhaps be encountered.

More quantitative experimental information using better filters is required before the full extent of the erosive nature of dust particles can be satisfactorily assessed.

DISCUSSION AND CONCLUSIONS

The method of sampling gas from No. 4 flare line is worth further consideration, as is the origin of the dust particles collected.

As the gas was required to be sampled just before No. 4 flare, the only reasonable configuration for the probe in this location, bearing in mind the physical constraints at the deaerator roof, was, in fact, with the probe pointing vertically

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upwards from the flare line. It is likely, therefore, that a proportion of the heavier dust particles may not actually have reached the filter.

The filters were found to be discoloured on both sides, suggesting that some material had passed through. It is not, therefore, possible to specify whether the values obtained in this work for the dust concentrations are lower or upper limits.

The filters used were not ideal, but were chosen initially to survive the aggressive environment of the raw Slagging Gasifier gas. Blank filters were observed to contain some particles similar in appearance to those being collected by the test filters. It has been suggested that Nuclepore filters would be more suitable for future experiments.

Considering the limited number of samples, and the uncertainties in the measured dust concentrations, together with the qualitative nature of the analytical results, the overall effect of dust from the Westfield Slagging Gasifier on turbine blades is very much open to debate. It is not possible to draw scientifically valid conclusions, but it would appear that hot corrosion is unlikely to present too much of a problem, particularly since in an industrial plant, there would be at least one more cleaning stage, mainly for sulphur removal, which would reduce the dust concentration even more.

It is likely that most of the dust collected originated from the inside walls of the pipework downstream of the gasifier, as opposed to carry over from the top of the bed of the gasifier. This leaves the interpretation of the result open to speculation without further experiments to determine the origin of the dust. Further analysis did reveal that apprxomately 25% of the solid material was present as hydrocarbon.

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Section 9

LOAD CHANGE TESTS ON OXYGEN PLANTS
(By BOC Limited)

BACKGROUND

British Gas operates two BOC designed and built 100 TPD* Tonnox Internal compression plants supplying oxygen to their gasifiers. The test for EPRI required the gasifiers to be flexed from full load to 65% load and back again in short periods of time, to simulate SNG production for gas turbines generating electricity.

To demonstrate the ability of typical oxygen plants to cope with the flexible oxygen demands of a gasifer it was necessary to carry out load change tests on one of the Westfield Tonnox plants. The objective of the tests was to identify the rate at which the oxygen plant could be turned up and down, such that the power savings associated with reducing output could be realised.

TEST PROCEDURE

The test was carried out on No. 1 plant.

The plant was running in a turndown condition on 19 December, 1979, making both gaseous and liquid oxygen. Power readings and oxygen make were recorded at this setting.

On 20 December, 1979 the plant was increased to a maximum gaseous oxygen setting. It was then run for 2½ hours at this setting and power and oxygen make recorded.

The plant was then reduced to a turndown gas plus liquid setting.

The oxygen and nitrogen purities were measured by portable analyser since the plant continuous analysers were not operable. Oxygen flow was taken from the plant flow recorder. Power consumption was taken from the KWhr meters. Cooling water power is not included in the L.T. measurement since C.W. is common site supply.

*TPD = long tons (1016.1 kg) per day

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RESULTS

The readings taken whilst increasing and reducing plant output are shown on Figures 9-1 and 9-2 respectively.

- Turndown setting 19 & 20 December, 1979

  --Liquid oxygen make = 5.3 TPD = 5,850 sft^3/hr*

  --Gaseous oxygen make = 66.5 TPD = 73,400 sft^3/hr
  Chart Reading = 5.5

  --Total oxygen make = 79,250 sft^3/hr purity 96.7%

  --Total power = 2,094.8 kW

  --Specific power of oxygen = 26.4 kW/1,000 sft^3/hr

  --Specific power of gaseous oxygen = 25.4 kW/1,000 sft^3/hr (allowing 40 kW/1,000 sft^3/hr for LO)

- Increasing plant output:

  --The gaseous oxygen flowrate was increased from a chart reading of 5.5 to 8.2 in 20 minutes. This represents an increase in oxygen flow from 73,400 sft^3/hr to 109,430 sft^3/hr; i.e., 67% to 100%. The oxygen purity increased over the next 20 minutes and the gaseous oxygen flow was correspondingly increased to a further 3%.

- Maximum gaseous oxygen 20 December, 1979

  --Liquid oxygen make = 0

  --Gaseous oxygen make = 102.2 TPD
  = 112,770 sft^3/hr Purity 96.5%

  --Total power = 2,378.8 kW

  --Specific power of gaseous oxygen = 21.1 kW/1,000 sft^3/hr @ 450 psig

- Reducing plant output

  --The gaseous oxygen flow rate was reduced from a chart reading of 8.55 to 5.5 in 22 minutes. This represents a decrease in oxygen flow from 114,100 sft^3/hr, to 73,400 sft^3/hr; i.e., 104% to 6%.

- The oxygen purity as measured by the portable analyser read low before and during the test such that the rate of oxygen take off was reduced to recover this. The indicated purity partially recovered.

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*sft^3 = cubic feet at: Pressure - 30 inches Hg
Temperature - 60°F

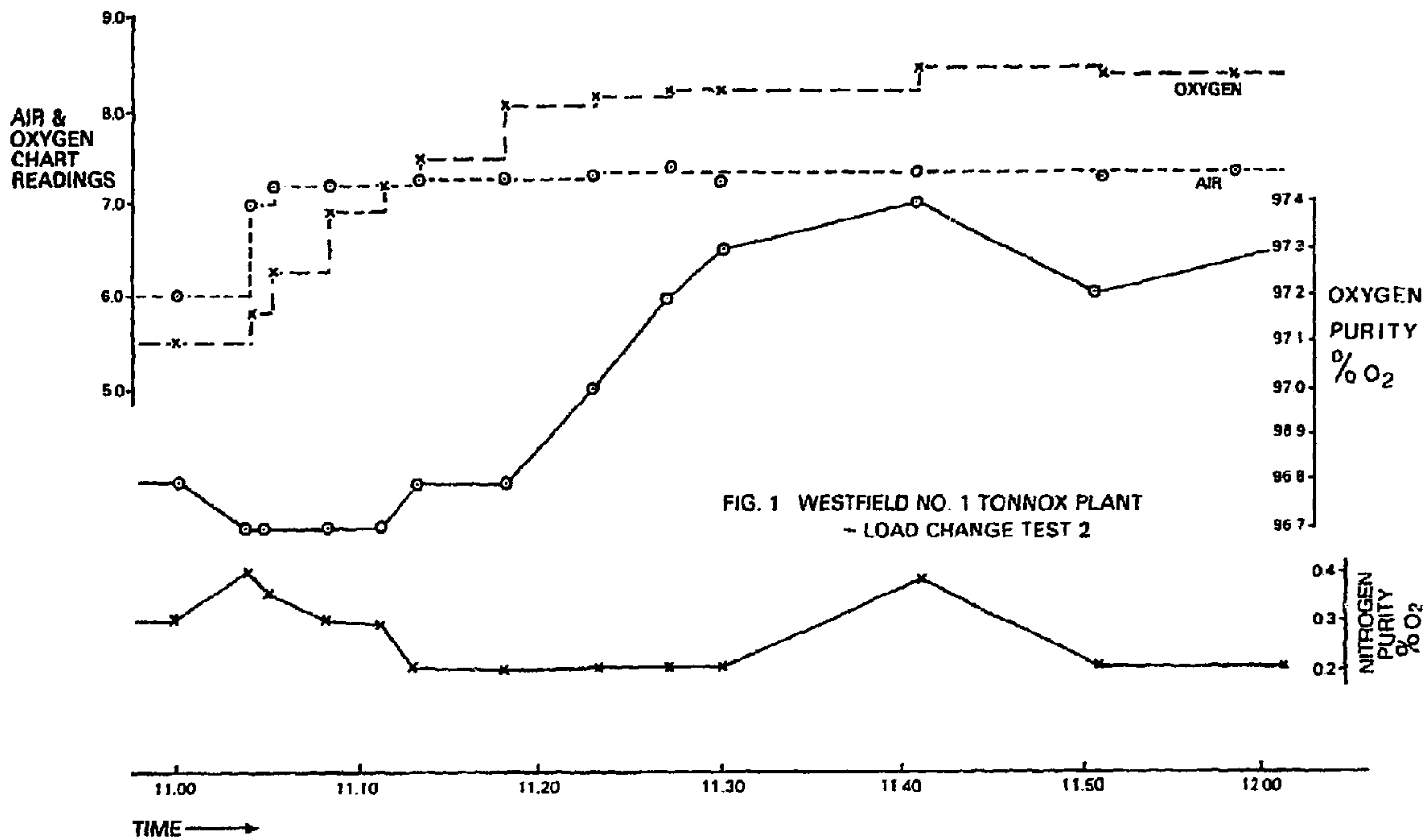

Figure 9-1. Westfield No. 1 Tonnox Plant-Load Change Test 2

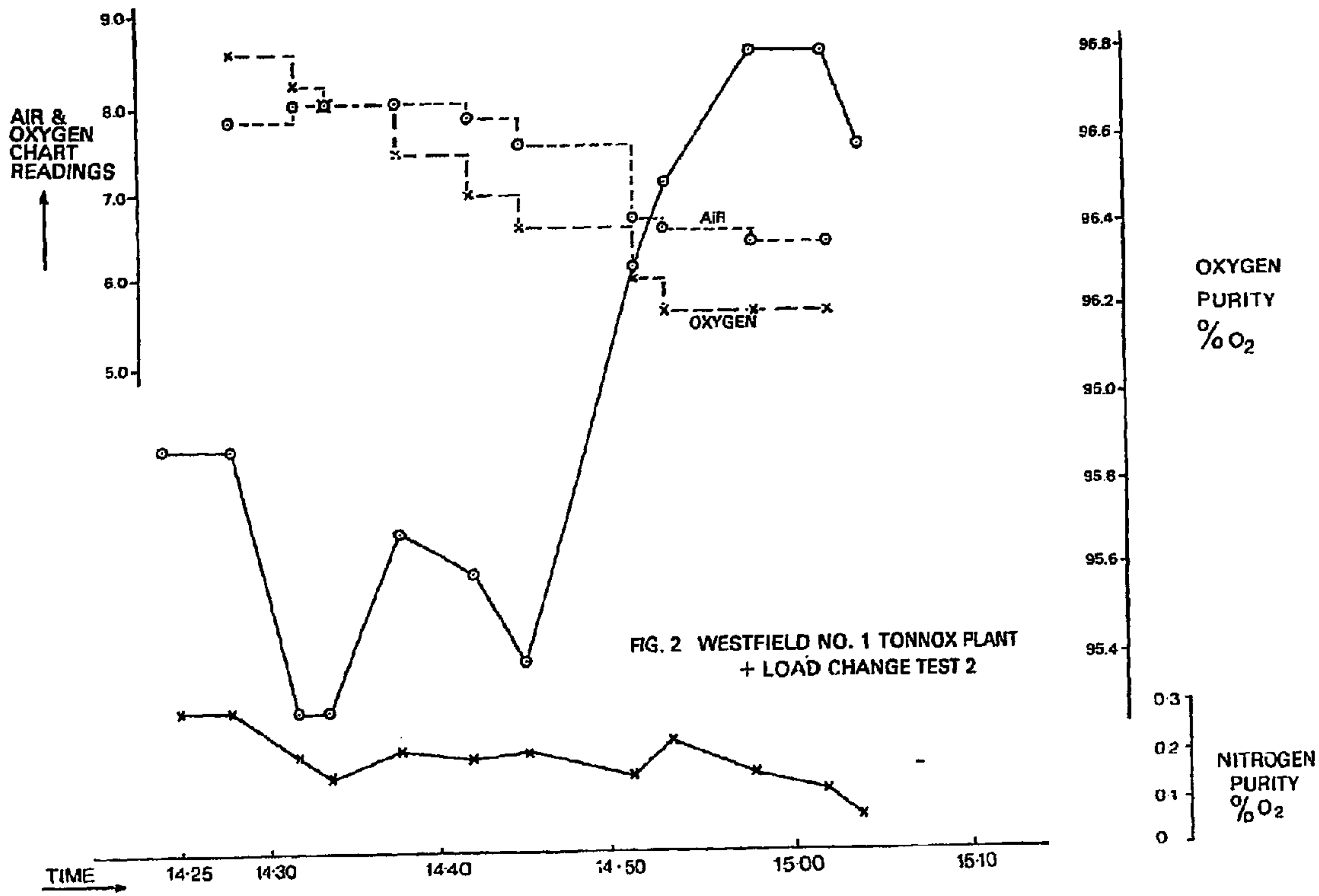

Figure 9-2. Westfield No. 1 Tonnox Plant-Load Change Test 2

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COMMENTS

- No difficulty was experienced in retaining purity or increasing the plant output in 10 minutes. Indeed the increased oxygen purity implies not enough gaseous oxygen was withdrawn. Therefore, with gain in operating experience it is possible that 20 minutes could be improved upon, when increasing plant output.

- The rapid turndown of a plant is usually more difficult than the reverse, therefore, it is unlikely that a turndown time of better than 20 minutes is achievable on this plant.

- The limits of turndown and maximum gas as measured during these short tests cannot be considered to be the absolute limits of this plant's operating range. Certainly there is scope for optimisation which could not be attempted within the timescale of the tests.

DISCUSSION

- No problem was experienced turning the plant up or down in 20 minutes and still retaining the design purities on both oxygen and waste nitrogen. It is probable the turn up rate could be improved upon with practice. The quality of products imposes restraints on the flexing time of plants. If nitrogen purity is not critical, and oxygen purities down to 90% are acceptable then it is likely that similar load changes could be carried out in less than 20 minutes on this particular plant.

- The Westfield Tonnox plants are oversized for their present oxygen demand. A plant specifically designed for the oxygen demand would have a more suitable range of operation.

- The Tonnox plants use a now outdated high pressure Heylandt liquefaction cycle which incurs higher power consumptions than would be acceptable on modern plants. Modern low pressure gas plants deliver gaseous oxygen at a few psig with external oxygen compression.

- Tonnox plants are similar to modern low pressure air separation plants insofar as the distillation columns are concerned. Both use the classic double column system.

- The turn up/down rates of these tests were carried out totally manually on a plant with practically no automation. Automatic computer controlled plants could undoubtedly operate these load changes far more efficiently and effictively than any manual operation.

SUMMARY

- The objective of the tests was to identify the rate at which typical oxygen producing plants could be turned up and down to simulate a variable gasifier demand.

- The tests proved that a 100 TPD Tonnox plant can flex between 65 and 100% of gaseous oxygen make in 20 minutes.

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- Nc difficulties in maintaining design purities were experienced.
- There is considerable scope for optimisation of the plant which was not possible with the limited time and data available. A wider operating range and quicker load changes might be possible with operational experience.
- Automatic computer controlled plants could undoubtedly operate these load changes more efficiently and effectively than manual operation.

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Section 10

REPORT ON TEST RUN EPRI - 01

SUMMARY

Run - 01 was the first of the three run programmes of the EPRI contract and aimed to demonstrate the ability of the gasifier, operating on the standard reference coal, Rossington,* to run at various loadings, and to change to these various loadings quickly and reliably. The dynamic behavior of the gasifier during these changes was to be investigated as was the ability of the gasifier to be controlled from the back end of the plant.

Slagging Gasification on Rossington coal started at 11:12 on 14 October, and with good running obtained under standard conditions.+ Some brief assessments of the effect of stirrer speed on performance were carried out before a mass balance performance test period was initiated. The gasifier was then turned down over ten minutes to 100,000 SCF/H oxygen loading with no problems and held at this loading for four hours, and then brought back up again at 13:17 on 15 October, 1979. Over the next six hours the gasifier was then spiked down to 110,000 SCF/H oxygen and back again on two separate occasions, the total duration of each spike being ten minutes and three minutes. These were carried out without problems and at 22:07 gasifier loading was brought down to 80,000 SCF/H oxygen over half an hour.

There was a prolonged period of running of over 40 hours at this loading. During this time several performance tests were carried out on the gasifier. These included three planned mass balance periods, the successful commissioning of flow control on the gasifier, and a series of short tests with different rates of tar injection and different stirrer speeds. All these tests were successfully carried out and at 19:33 on 17 October the rates were brought back slowly to standard conditions in flow control. After a period of steady running at 160,000 SCF/H oxygen

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*The standard reference coal, Rossington, is a 702 rank coal from a long life single seam colliery in Yorkshire and is thought to be typical of coals available for gasification in Britain in the 21st century.

+Standard Westfield Slagging Gasifier conditions are defined as 160,000 SCF/H (approximately 96% pure) oxygen loading to the tuyeres H_2O/O_2 = 1.30(v/v), and pressure psig.

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loading rate reduction again commenced, with the gasifier being steadied out at 50,000 SCF/H oxygen loading on 18 October. Running at 50,000 SCF/H oxygen loading was planned to be carried out at 1.30 steam/oxygen ratio, but due to a calibration error this was, in fact, 1.88. This high value did not affect smooth operation at the low loading.

The run was concluded with a period in flow control, with rates being brought up to 180,000 SCF/H oxygen before pressure control was restored and a controlled shut-down was carried out.

The run is summarised diagramatically in Figure 10-1. Table 10-1 describes the load changes made during the run.

EQUIPMENT, INSTRUMENTS AND CONTROLS

There were no major gasifier changes for EPRI - 01. However, numerous changes were made on the control and instrumentation side, some of which had started to be commissioned during previous runs. Some of these were:

- Installation of a new flare tip at number 4 flare to provide for quieter running on single flare, full load, operation.
- Replacement of the 4" flare control valve by a 6" valve on number 4 stream to deal with potential higher flows up the flare.
- A new control system which enabled the gasifier to work in back end flow control and front end pressure control as well as the normal mode of flow control on the steam/oxygen and pressure control at the flare.
- Continuous gas analysis at the flare by mass spectrometer for the elements CO, H_2, CH_4, CO_2, N_2, C_2H_4, C_2H_6, COS, H_2S.
- Continuous gas analysis at the upper dimester (either 3 or 4) using discrete individual analysers to monitor CO, CO_2, H_2, CH_4, H_2S and total hydrocarbons.
- Continuous measurement of gas CV and Wobbe number at the upper demister.
- Metering of tar injection to gasifier.
- Installation of data logging system.

Instrumentation apart, systems were operated as per previous runs and it was planned to use the system developed to facilitate running at low loads.

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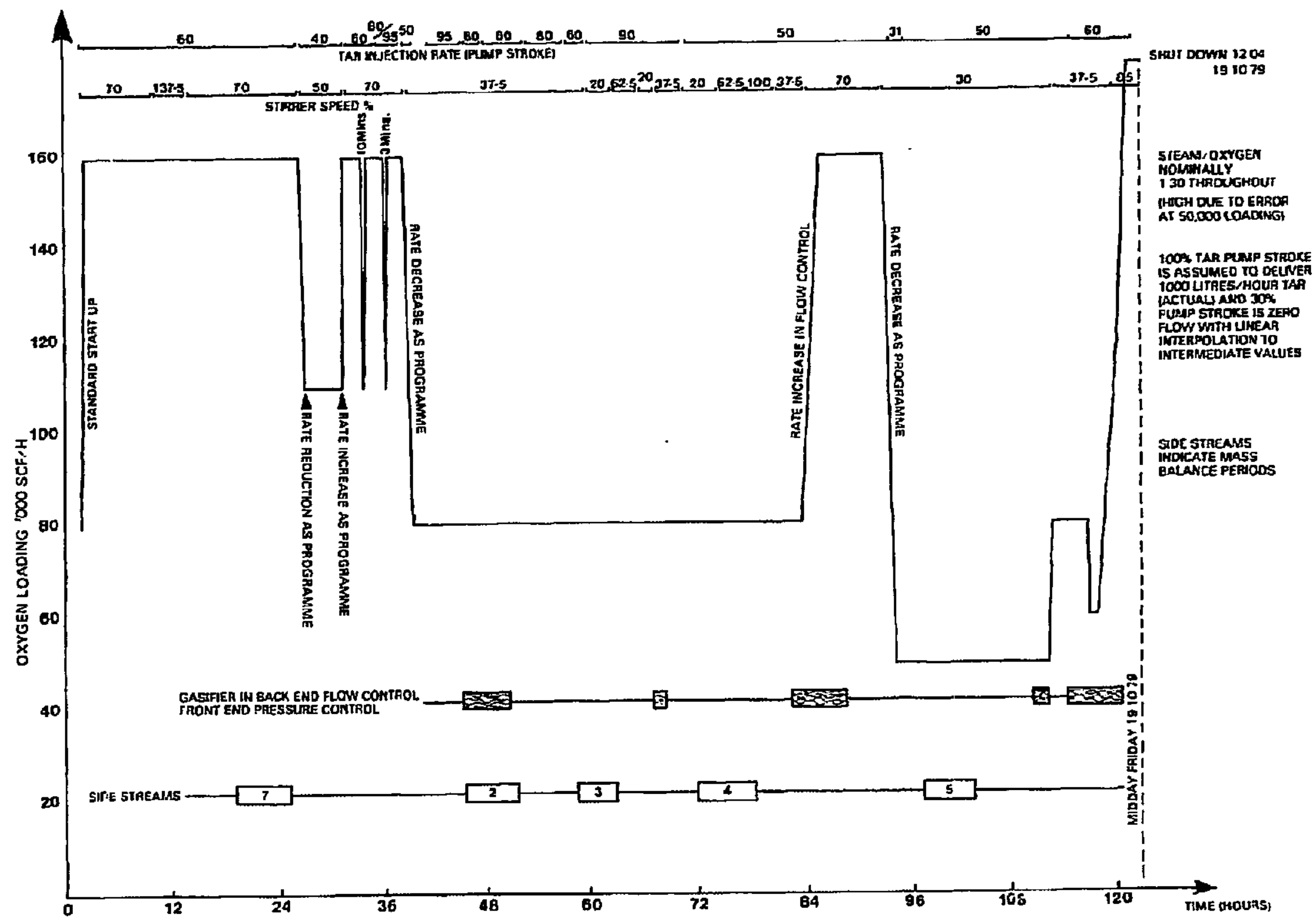

Figure 10-1. Schematic of EPRI - 01

TABLE 10-1. DESCRIBING LOAD CHANGE PERIODS - EPRI - 01

Change No.	Load 1 before Change (1000 SCFH O_2 Blast)	Load 2 after Change (1000 SCFH O_2 Blast)	Time Taken (Mins.)	Change in 1000 SCFH	Coal Gasified	Control Mode
1	160	110	12	50	Rossington	P
2	110	160	14	50	Rossington	P
3	160	110	4½	50	Rossington	P
4	110	160	4	50	Rossington	P
5	160	110	1	50	Rossington	P
6	110	160	1	50	Rossington	P
7	160	80	34½	80	Rossington	P
8	80	160	92	80	Rossington	F
9	160	50	120	110	Rossington	P
10	50	80	18	30	Rossington	P
11	80	60	19	20	Rossington	F
12	60	180	135	120	Rossington	F

RUN DIARY

The run was initiated at 08:45 on 14 October, 1979. At 11:12 locking on Rossington washed singles (¼" - 1¼") commenced. Coal locking was intially on raw gas. By 11:45 the gasifier was under standard conditions of 160,000 SCF/H oxygen loading, 1.30 steam/oxygen ratio and 340 psig. It was decided to hold pressure slightly below the normally used 350 psig in order to avoid any problems with pressure fluctuations lifting relief valves when the various control modes were tried later on in the run.

A good start up was obtained, with automatic slag tapping being commissioned and tar injection to the gasifier top being brought on at 60% pump stroke.

Early on 15 October the gasifier was settled down and running well on Rossington coal at standard reference conditions and a performance test period was started, with a sidestream being put on at 04:17. Good gasifier performance continued throughout this mass balance period. At 11.00 rate reduction to 100,000 SCF/H oxygen commenced, with the gasifier being brought down from 160,000 SCF/H oxygen in ten minutes. Tar injection was brought down to 50% pump stroke. This reduction in rates was carried out smoothly and produced no discernable transient effects upon the gasifier.

At 15:43, after four hours of steady running at 110,000 SCF/H oxygen, the rates were brought back up to 160,000 SCF/H in ten minutes, again, with no problems and again, without the gasifier exhibiting any significant transient phenomena. After allowing two hours for the gasifier to settle down at standard loading, the rates were dropped to 110,000 SCF/H oxygen and then brought back up again, all in the space of ten minutes. This produced no effect upon the steady performance of the gasifier, nor did it when the exercise was repeated again at 20:15, this time over three minutes. Again, no upsets were created.

At 22:07 gasification rates were starting to be lowered towards 80,000 SCF/H oxygen loading, this loading being reached at 22:30. Again, a good transition was obtained, with no obvious major transient phenomena, and the gasifier settled down to work well at this half load, with good performance in all areas; Tar injection to the gasifier top was proving difficult to establish and at 03:42 both pumps were turned on at high stroke. This appeared to upset the bed slightly with a high bed DP and offtakes temperatures. Tar injection was then settled on to one pump at 60% stroke and good gasifier performance continued. Performance data was gathered at this loading and a sidestream was run this period ending at 11:37 when tar injection

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was raised to 80% pump stroke Between 14:00 and 17:00 experiments were carried out on gasifier control, which ended with successful front end pressure control, back end flow control being achieved. Tar injection was then brought to as high as possible, at 90% pump stroke, which represented a tar rate of about 12% of the DAF coal feed, and further performance data gathered under these conditions, with another sidestream being run. At this high tar load the effect of three stirrer speeds were investigated. Gasifier control in back end flow was restored and further adjustments and improvements made to the control system. Gasifier performance continued to be good and at 06:15 tar injection was reduced down to 50% pump stroke, so that further performance data could be gathered at half load, this time with a low tar injection rate to the gasifier top. Sidestream number 4 was put on. Distributor revolutions were changed at 13:15 and again at 16:33, which represented the last test at 80,000 SCF/H loading.

At 19:33 the gasifier was put in back end flow control and rate increased achieved by moving up the flow control set point at the back end of the plant. In this way standard running conditions of 160,000 SCF/H oxygen, H_2O/O_2 1.30 were reached by 21:05, and early on 18 October the gasifier was restored to back end pressure control. Gasifier performance continued to be satisfactory at the high loads and at 04:15 rate reduction to 50,000 SCF/H oxygen loading was commenced as per programme, this being finally reached at 06:15. Due to a calibration error on the steam flow indicator, the steam/oxygen ratio at this loading was subsequently discovered to be 1.88. However, this ratio still allowed acceptable hearth conditions and good bed behaviour.

For 50,000 SCF/H running, tar was reduced to 31% pump stroke. A good transition was made to this loading. Performance data was gathered under these conditions with a sidestream being run, after which the gasifier was put in flow control and the rates raised. Early on 19 October, while this was occurring, gasifier conditions deteriorated due to ad hoc manipulation of the steam/oxygen flows because of uncertainty about the steam/oxygen ratio. However, the problem was idendified, and this allowed rate increase at 1.30 steam/oxygen ratio to take place. The rates were successfully brought to 180,000 SCF/H oxygen at which load steady running was obtained for a brief period before the run was terminated with a controlled shutdown at 12:04.

Figure 10-1 shows the run schematically with the run periods in Table 10-2.

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Table 10-2

SUMMARY OF RUN PERIODS

Date	Time	Period	
14.10.79	08:45	1	Start up phase.
14.10.79	11:12	1	Steam/oxygen on. Locking on Rossington coal and 180lb/lock $CACO_3$ flux.
14.10.79	11:45	1	Rates now 160,000 SCF/H, steam/oxygen 1.3/1.
14.10.79	13:00	1	Tar injection on at 60%
15.10.79	04:05-10:10	1	Sidestream No. 1 completed.
15.10.79	11:00	1	Start of rate reduction to 110,000 SCF/H. Distributor revs down 70-60-50%. Tar injection down to 50%.
15.10.79	11:26	2	Rates now 110,000 SCF/H. Tar injection 40%.
15.10.79	15:10	2	Increased proportional band to 100% to steady flow.
15.10.79	15:43	2	Starting to increase rates.
15.10.79	15:53	3	Rates 160,000 SCF/H. Tar injection up to 60%.
15.10.79	18:15	3	Dropping rates in steps of 10,000 SCF/H.
15.10.79	18:20	3	Rates now 100,000 SCF/H.
15.10.79	18:25	3	Back up to 160,000 SCF/H.
15.10.79	19:15	3	Tar injection up to 80%
15.10.79	20:15	3	Dropping rates to 110,000 SCF/H over 90 seconds then increased to 160,000 SCF/H over 80 seconds.
15.10.79	20:27	3	Tar injection now 95%.
15.10.79	20:37	3	Tar injection rate now 50%.

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Date	Time	Period	
15.10.79	22:07	3	Cutting rates.
15.10.79	22:39	4	Rates now 80,000 SCF/H.
16.10.79	02:50	4	Tar injection now 100%.
16.10.79	02:57	4	First tar injection pump on also at 95%.
16.10.79	03:52	4	Second tar pump shut down. First one still at 95%.
16.10.79	04:31	4	Tar injection pump turned down to 80%.
16.10.79	05:00	4	Sidestream No. 2 on.
16.10.79	06:15	4	Proportional band on gasifier pressure controller is now 60%.
16.10.79	06:35	4	Tar injection down to 60% pump stroke.
16.10.79	11:10	4	Sidestream No. 2 off.
16.10.79	11:37	4	Tar injection up to 80%.
16.10.79	13:40	4	Gasifier No. 4 valve in manual control.
16.10.79	14:37	4	Steam into manual control.
16.10.79	15:17	5	Into front end automatic control.
16.10.79	15:22	5	Into flow control on CV 201.
16.10.79	15:37	5	Gasifier pressure dropped to 323 psig.
16.10.79	15:56	5	Tar injection dropped to 60%.
16.10.9	16:30	5	Switched from local to remote control on FC.100.D.
16.10.79	16:45	5	Gasifier pressure up to 334 psig.
16.10.79	17:15	5	Tar injection pump up to 90%.
16.10.79	18:15	5	Sidestream No. 3 on.
16.10.79	18:20	5	Distributor down to 20%.
16.10.79	20:15	6	Gasifier back on pressure control.

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Date	Time	Period	
16.10.79	22:00	6	Tar injection stopped.
16.10.79	22:19	6	Tar injection back on line.
16.10.79	22:20	6	Sidestream No. 3 off.
17.10.79	01:10	6	Tar injection pump stopped.
17.10.79	02:05	6	Tar injection back on line.
17.10.79	03:40	7	Automatic flow control established.
17.10.79	04:40	7	Back to pressure control conditions.
17.10.79	06:15	7	Tar injection down to 50%.
17.10.79	08:10	7	Sidestream No. 4 on.
17.10.79	14:00	7	Sidestream No. 4 off.
17.10.79	18:37	7	Flow control established in automatic.
17.10.79	19157	8	Increased rates to 100,000 SCF/H.
17.10.79	20:12	8	Proportional band down from 60–50%.
17.10.79	10:15	8	Rates 115,000 SCF/H.
17.10.79	20:30	8	Rates 130,000 SCF/H.
17.10.79	21:05	8	Rates 160,000 SCF/H.
18.10.79	04:15	8	Starting to reduce rates.
18.10.79	05:23	8	Rates now 80,000 SCF/H.
18.10.79	06:15	9	Rates down to 50,000 SCF/H. Proportional band set at 700%.
18.10.79	06:30	9	Tar injection set at 31%.
18.10.79	06:45	9	Proportional band set at 40%.
18.10.79	07:25	9	Increased tar injection to 40%.
18.10.79	07:45	9	Tar injection 50%.
18.10.79	11:10	9	Sidestream No. 5 on.

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Date	Tima	Period	
18.10.79	17:20	9	Sidestream No. 5 off.
19.10.79	00:00	10	Changing to flow control at back end.
19.10.79	01:50	10	Double loads of flux added; i.e., 360 lbs.
19.10.79	02:06	10	Back to pressure control. Rates at 60,000 SCF/H.
19.10.79	02:17	10	Rates at 80,000 SCF/H.
19.10.79	02:57	10	Changing ratio to 1.1/1.
19.10.79	03:39	10	Ratio back to 1.3/1.
19.10.79	04:25	10	Going over to flow control. Tar injection up to 60%.
19.10.79	07:40	10	Reducing rates in flow control to 60,000 SCF/H.
19.10.79	08:40	10	Increasing rates.
19.10.79	09:32	10	Rates now 110,000 SCF/H.
19.10.79	09:45	10	Rates up to 120,000 SCF/H.
19.10.79	10:24	10	Rates now 155,000 SCF/H.
19.19.79	11:00	10	Rates 180,000 SCF/H, ratio 1.25/1. Tar injection off.
19.10.79	11:56	10	Into pressure control.
19.10.79	12:04	10	Shut down.

(End of Table 10-2)

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SLAG TAPPING AND SLAG REMOVAL

Slag tapping and slag removal were good throughout run EPRI - 01, although when changes were made to gasifier loading it was noticeable that the hearth took far longer than any other gasifier area to completely respond to these changes.

FLUXING SYSTEMS

Fluxing during EPRI - 01 was on limestone flux.

Table 10-3 gives the flux/coal ash-slag balances for the various run periods. Some of the errors observed can be attributed to the fact that the periods analysed were short and time to attain equilibrium was not achieved. Errors in iron balance may be attributed to difficulties in sampling and analysing the slag for fixed and free iron.

Fluxing systems worked well during EPRI - 01, yielding good hearth conditions with a free flowing slag.

BED BEHAVIOUR

Bed behaviour was generally good throughout EPRI - 01. The steady run periods of the gasifier allowed the following periods to be analysed in some depth:

1A: 160,000 SCF/H, 1.30 H_2O/O_2; distributor 1.37% tar injection 60% pump stroke.

1B: As above with distributor at 70%

3: 160,000 SCF/H, 1.30 H_2O/O_2; distributor 70% various tar injection rates. This period includes two brief excursions down to 110,000 SCF/H oxygen.

6A: 80,000 SCF/H, 1.30 H_2O/O_2; distributor 20% tar high (90% pump stroke).

6B: As above with distributor 65.5%.

6C: As above with distributor 100%.

6D: As above with distributor 37.5%.

7A: 80,000 SCF/H, 1.30 H_2O/O_2; distributor tar injection low (50% pump stroke).

7B: As above with distributor 62.5%.

7C: As above with distributor 100%

7D: As above with distributor 37.5%.

TABLE 10-3. FLUX/COAL ASH-SLAG BALANCES FOR EPRI - 01

Calculated figures refer to those calculated from known inputs of coal ash and flux and their known analyses.

Period	1		2		3		4		5	
Flux:Ash	0.42		0.46		0.47		0.49		0.48	
% x	Actual	Calc.	* Actual	* Calc.	* Actual	* Calc.	* Actual	* Calc.	* Actual	* Calc.
SiO_2	36.15	33.02	36.15	32.50	35.02	31.97	35.02	31.71	33.89	31.44
Al_2O_3	24.38	19.90	24.38	19.57	23.77	20.13	23.77	19.97	23.15	20.68
Fe_2O_3	7.21	12.10	7.21	11.89	5.52	10.74	5.52	10.65	3.83	9.59
CaO	21.78	21.69	21.78	22.90	23.60	23.86	23.60	24.45	25.42	24.82
MgO	1.96	1.82	1.96	1.87	1.92	1.88	1.92	1.91	1.87	1.89
Na_2O	1.98	1.80	1.98	1.76	1.87	1.80	1.87	1.79	1.76	1.84
K_2O	1.56	1.66	1.56	1.64	1.60	1.68	1.60	1.66	1.63	1.72
Silica Ratio	51.17	45.80	51.17	44.79	50.37	44.45	50.37	43.95	49.55	44.10

x Other minor constituents make up to 100%

TABLE 10-3 (Con't)

Period	6		7		8		9		10	
Flux:Ash	0.49		0.52		0.52		0.51		0.61	
	Actual	Calc.	Actual	Calc.	Actual*	Calc.*	Actual	Calc.	Actual	Calc.*
SiO_2	33.89	31.32	32.98	32.54	33.02	32.07	31.73	35.73	35.73	30.57
Al_2O_3	23.15	20.60	22.65	20.31	22.64	20.26	22.62	20.28	26.10	19.49
Fe_2O_3	3.83	9.55	4.98	10.94	5.22	10.76	5.47	10.62	6.53	10.21
CaO	25.42	25.11	29.38	22.70	30.02	23.21	30.67	23.43	20.89	26.05
MgO	1.87	1.90	1.93	1.73	2.00	1.78	2.06	1.82	1.77	1.91
Na_2O	1.76	1.83	1.71	2.03	1.74	2.06	1.77	2.11	2.79	2.02
K_2O	1.63	1.71	1.31	1.62	1.28	1.60	1.26	1.58	1.13	1.52
Silica Ratio	49.55	43.85	45.62	45.47	45.06	44.87	44.49	44.51	51.90	42.29

* Average Values

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8: 160,000 SCF/H at 1.30 H_2O/O_2; gasifier in flow control; distributor 70% tar injection 60% pump stroke.

9: 50,000 SCF/H oxygen loading at 1.88 H_2O/O_2. Distributor at 30% tar injection 40% pump stroke.

The offtake temperature analysis for the periods above shows a trend to higher mea offtake temperatures at higher loads as per Table 10-4.

Table 10-4

OFFTAKE TEMPERATURES

Test Period	Mean Temp. °F	Median Temp. °F
1A	889	886
1B	983	977
3	943	940
6A	965	964
6B	901	898
6C	920	913
6D	918	919
7A	919	913
7B	896	903
7C	894	899
7D	888	880
8 (All)	949	944
8 (last 2 hrs)	956	945
8 (last 5 hrs)	947	948
9	880	883

The effect of tar injection and stirrer speed on offtake temperature was also examined at 80,000 SCF/H oxygen loading at 1.30 during periods 6 and 7. The conclusions were:

- Tar injection leads to less than perfect bed conditions.
- There is an optimum stirrer speed for the best bed conditions.

The CO_2 trace is more ragged at high tar injection (Figure 10-2) than at low tar injection (Figure 10-3). "High" tar injection rate refers to a recycle rate of about 900 lbs/hr of wet dusty tar to the bed top; "low" refers to a rate of about 450 lbs/hr.

As was stated earlier, bed behaviour during the run was good. There appeared to be little effect of transients and any effect of a change in loading produced an immediate effect upon those factors which did change, such as bed DP. The mean offtake temperature adjusts quickly to the new conditions, there was no significant change in average offtake temperature.

DUST IN FLARE GAS

An Isokinetic gas probe was installed at the high pressure side of number 4 flare valve for EPRI - 01 and the gas passed at pressure under flow control through a Whatman GF/B paper filter. The filter was weighed before installation, then dried at 100°C for 24 hours, and weighed again after it was removed. The assembly was heated to prevent ammonium carbonate entrainment on the filter. The results obtained are summarised in Table 10-5. The results indicate that dust levels in the flare gas are low, of the order of 0.1 ppm (W/W).

PLANT BEHAVIOUR IN FLOW CONTROL

The controller settings were altered from time to time during the run because of the various throughputs and control methods used. These settings are given in Table 10-6. The steam/oxygen ratio was nominally 1.3 to 1 throughout except for the 50,000 SCF/H oxygen loading when it was higher at 1.88 to 1. The steam/oxygen flow characteristics and the gasifier pressure plus flare gas flow for the periods referred to in Table 10-5 are given in Figure 10-4, which is made up of sections of recorder chart taken during a typical period.

These figures show, as expected, variable gas flow in pressure control and variable steam/oxygen flow in flow control.

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Figure 10-2. Offtake CO_2 Analysis at 80,000 Loading at 100% and 35.5% with High Tar Recycle
Chart Speed: 2 Min./CM

Figure 10-3. Offtake CO_2 Analysis at 80,000 SCF/H Oxygen Loading at 100% and 37.5% with Low Tar Recycle
Chart Speed: 2 Min./CM

TABLE 10-5. SUMMARY OF DUST IN GAS RESULTS FROM RUN EPRI - 01

Day of Test	Time	Test Period	Test Length (Hrs.)	Gasifier Oxygen Flow (SCFH)	Gas Flow Through Probe (SCFH)	Flow Meter Setting	Filter Status	Weighings	Wt. of Dust (mg/SCF) Gas	Gas Temp. after Orifice Plate (^oC)	Sent to LRS.
No Tests											
14.10.79											
Test 2											
15.10.79	17.00	3/4	9.5	160,000	5,000	7 ± 0.2	Filter and wire mesh lost in pipe.	-	-	24/27.5^oF	-
16.10.79	02.30										
Test 3											
16.10.79	02.30	4	9.5	80,000	*5,000	7 ± 0.2	Filter intact, stuck to wire mesh. Dried 24 Hrs at 100^oC (after drying)	Filter + Dust = 0.5578g. Filter = 0.4629g Wt. of Dust = 0.0949g.	2×10^{-3}	19/23^oC	Yes
Test 4											
16.10.79	21.30	6/7	12.0	80,000	2,857	4 ± 0.2	Erosion at filter edge. Wire mesh bowed (not dried)	Not weighed		21/25^oC	Yes
17.10.79	09.40										
Test 5											
17.10.79	10.00 20.30	7/8	10.5	80,000	2,857	4 ± 0.2	Filter intact. Dried 24 Hrs. at 400^oC.	Filter + Dust = 0.4851g. Filter - 0.4649 Wt. of Dust = 0.0202g.	6.7×10^{-4}	22/25^oF	Yes

TABLE 10-5 (Con't)

Day of Test	Time	Test Period	Test Length (Hrs.)	Gasifier Oxygen Flow (SCFH)	Gas Flow Through Probe (SCFH)	Flow Meter Setting	Filter Status	Weighings	Wt. of Dust (mg/SCF) Gas	Gas Temp. after Orifice Plate (°C)	Sent to LRS.
Test 6											
17.10.79	21.35	8	6.5	160,000	5,000	7 ± 0.2	Erosion of filter edge (not dried)	Not weighed	-	27/33°C	Yes
Test 7											
18.10.79	11.30	9	9.5	50,000	2,141	3	Filter intact.	Filter + Dust = 0.4834g. Filter = 0.4625g. Dust = 0.0209g.	1.02×10^{-3}	35°C	Yes

* Not isokinetic at 80,000 SCFH O_2 Flow.

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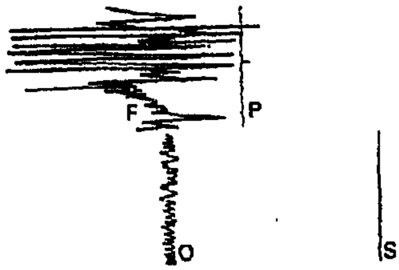
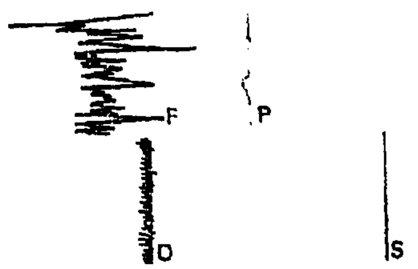
TABLE 10-6. GASIFIER CONTROLLER SETTINGS FOR EPRI - 01

Condition	P.C. Gasifier Pressure P.B.	P.C. Gasifier Pressure Reset	Flare Gas Flow P.B.	Flare Gas Flow Reset	Oxygen Flow P.B.	Oxygen Flow Reset	Steam Flow P.B.	Steam Flow Reset	Approx. Oxygen Blast SCFH	Control Type	Day and Time of Setting Changes
1	10%	10	1000 %	1	300%	10	200%	0	160,000	Pressure	-
2	20%	0	1000 %	1	300%	10	200%	0	160,000	Pressure	14th 14.19
3	20%	0	1000 %	1	300%	10	200%	0	110,000	Pressure	-
4	40%	0	1000 %	1	300%	10	200%	0	110,000 / 160,000	Pressure	15th 15.10
5	40%	0	1000 %	1	300%	10	200%	0	150,000	Pressure	-
6	60%	0	1000 %	1	300%	10	200%	0	80,000	Flow	15th 22.30
7	60%	0	1000 %	1	300%	10	200%	0	80,000	Pressure	-
8	60%	0	1000 %	1	300%	10	200%	0	Inc. from 80,000	Flow	-
9	30%	0	1000 %	1	300%	10	200%	0	160,000	Flow	17th 20.12
10	30%	0	1000 %	1	300%	10	200%	0	160,000	Pressure	-
11	30%	0	1000 %	1	300%	10	200%	0	160,000 / 50,000	Pressure	-
12	40%	0	1000 %	1	700%	10	200%	0	50,000	Pressure	18th 06.15
13	40%	0	1000 %	1	700%	10	200%	2	50,000	Pressure	18th 16.40
14	40%	0	1000 %	1	700%	10	200%	2	50,000	Flow	-
15	40%	0	1000 %	1	700%	10	200%	2	60,000	Pressure	-
16	40%	0	1000 %	1	700%	10	200%	2	80,000	Flow	-
17	40%	0	1000 %	1	700%	10	200%	2	Inc. from 60,000	Flow	-
18	40%	5	1000 %	1	700%	10	200%	2	180,000	Flow	19th 11.10

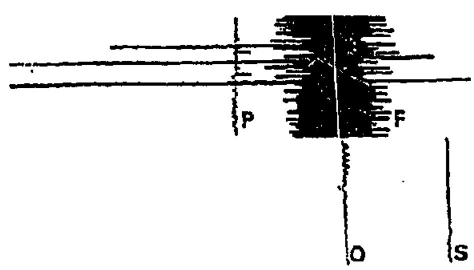
p



CONDITION 1. PRESSURE CONTROL. LOW P.B. AND HIGH RESET ON P.C. CAUSED FLUCTUATIONS ON GAS MAKE.

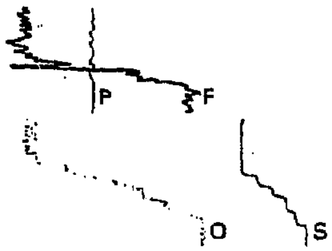


CONDITION 2. PRESSURE CONTROL. LOW P.B. AND OVERSIZE CONTROL VALVE ON P.C. SYSTEM CAUSED THE GAS FLOW TO VARY IN ORDER TO CONTROL THE GASIFIER PRESSURE.

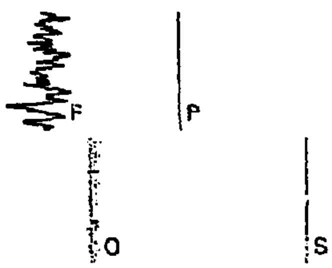


CONDITION 3. PRESSURE CONTROL. LOW P.B. AND AN OVERSIZE CONTROL VALVE ON THE P.C. SYSTEM CAUSED THE GAS FLOW TO FLUCTUATE.

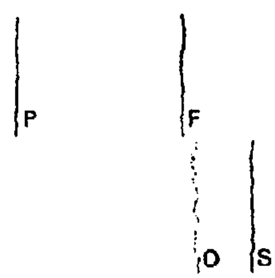
Figure 10-4. Recorder Charts Showing Flow and Pressure Characteristics at Several Controller Settings



CONDITION 4. PRESSURE CONTROL. INCREASING THE P.B. ON THE P.C. STEADIED OUT THE GAS FLOW AND THE ADDITIONAL PRESSURE VARIATION WAS NOT EXCESSIVE.

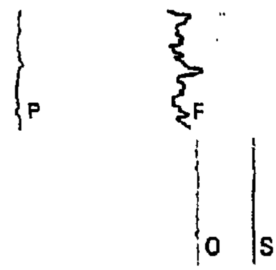


CONDITION 5. PRESSURE CONTROL. AT STANDARD RATES THE STEADY STATE CONTROL WAS QUITE GOOD.

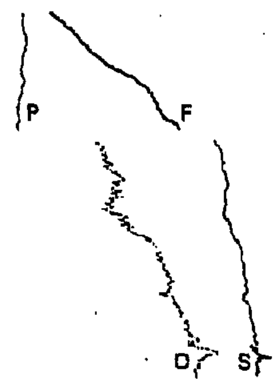


CONDITION 6. FLOW CONTROL. THE GAS FLOW WAS CONTROLLED AND THE STEAM AND OXYGEN FLOWS VARIED TO MAINTAIN A STEADY PRESSURE.

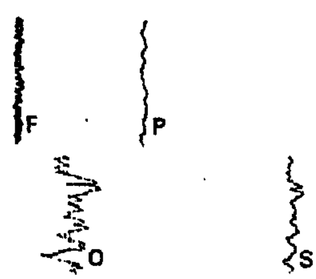
Figure 10-4 (Con't)



CONDITION 7. PRESSURE CONTROL. THE CONTROLLER SETTINGS ARE THE SAME AS FOR CONDITION 6 BUT THE GASIFIER IS IN PRESSURE CONTROL.

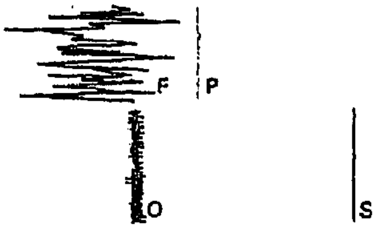


CONDITION 8. FLOW CONTROL.

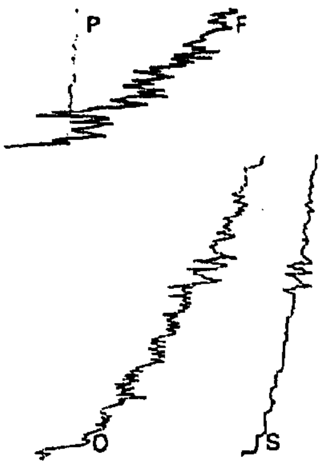


CONDITION 9. FLOW CONTROL. THE GAS FLOW IS STEADY. THE STEAM AND OXYGEN FLOWS ARE NOT VARYING EXCESSIVELY. THE GASIFIER PRESSURE IS UNSTEADY BUT AGAIN THESE SMALL FLUCTUATIONS ARE ACCEPTABLE.

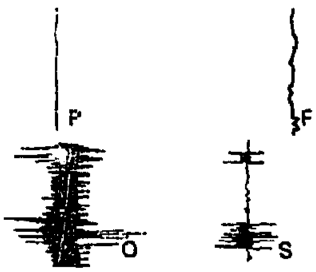
Figure 10-4 (Con't)



CONDITION 10. PRESSURE CONTROL. THESE CHARTS SHOULD BE COMPARED TO THOSE UNDER CONDITION 5. IT CAN BE SEEN THAT A 30% P.B. IS PROBABLY TOO LOW WHEN USING THE 6" CONTROL VALVE AS IT TENDS TO PRODUCE FLOW FLUCTUATIONS.

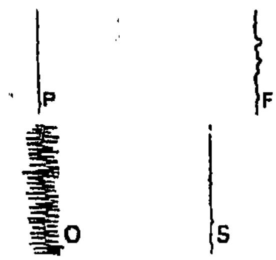


CONDITION 11. PRESSURE CONTROL.

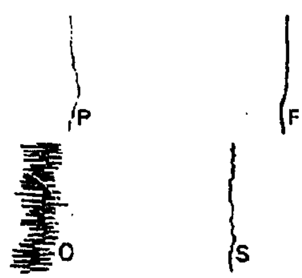


CONDITION 12. PRESSURE CONTROL. THERE WAS A FLOW VARIATION ON THE STEAM SYSTEM WHICH WAS BEING TRANSFERRED TO THE OXYGEN BY THE CASCADE CONTROL SYSTEM. THE LOW RANGE INDICATORS NEED DIFFERENT CONTROL SETTINGS TO MAINTAIN STABLE CONTROL.

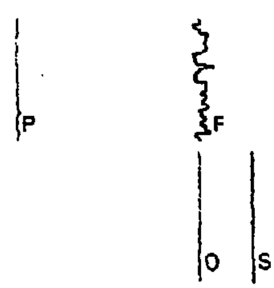
Figure 10-4 (Con't)



CONDITION 13. PRESSURE CONTROL. THE STEAM FLOW CONTROLLER RESET HAS BEEN INCREASED FROM ZERO TO 2.



CONDITION 14. FLOW CONTROL. SETTINGS AS FOR 13 BUT NOW IN FLOW CONTROL.



CONDITION 15. PRESSURE CONTROL. SETTINGS AGAIN AS FOR 13 BUT AT 80,000 RATHER THAN 50,000 SCFH OXYGEN BLAST EQUIVALENT.

Figure 10-4 (Con't)

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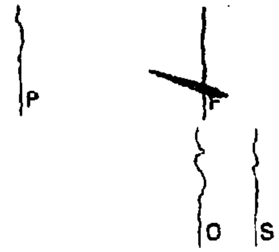
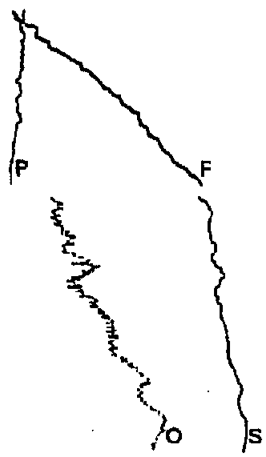
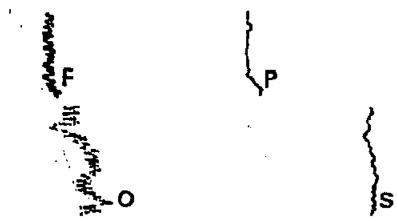


CONDITION 16. FLOW CONTROL. SETTINGS AS FOR 15.



CONDITION 17. FLOW CONTROL. CHANGING RATES IN FLOW CONTROL OVER AN EXTENDED PERIOD CAN BE CARRIED OUT WITHOUT ANY DIFFICULTY.



CONDITION 18. FLOW CONTROL. THE GASIFIER PRESSURE CHART SHOWS THE INCREASE CAUSED BY SETTING A RESET OF 5 ON THE PRESSURE CONTROLLER. WHEN THE RESET WAS ZERO THERE WAS AN OFFSET ON THE P.C.

Figure 10-4 (Con't)

Although gasifier bed behaviour was steady and very satisfactory throughout the run, the bed appeared to be more steady when operating in flow control, with the front end pressure control quickly recognising the slight pressure increase from a high offtake temperature and dropping the rates slightly. The 6" flare gas control valve was almost shut at the lower gasifier rates and a small valve would clearly be beneficial for control.

The zero reset levels recorded in Table 10-5 were selected by accident because the reset dials on the instruments were not continuous. Condition 18 on Figure 10-4 shows how the gasifier pressure increased to its set point when a reset level of 5 was selected.

The experience of this run showed that the gasifier can be operated satisfactorily in the flow control mode at both low and high flare gas production rates, although the controller settings are not at their optimum values. Comparison of periods 5 and 10 shows that in the gasifier pressure control mode, the gas flow rate, and also the gasifier pressure, is considerably more stable with a 40% proportional band value on the gasifier pressure controller. In the gasifier flow control mode the steam/oxygen flows are still rather variable at high rates, although no problems were experienced as long as sufficiently high vent rates on oxygen and steam ring mains were maintained.

STEAM/OXYGEN RATIO AT 50,000 SCF/H OXYGEN LOADING

The carbon dioxide level in the make gas was considerably higher during EPRI - 01 at 50,000 SCF/H oxygen loading than would be expected for the nominal set steam/oxygen ratio of 1.3, so the steam and oxygen sets were checked and calibrated immediately after the run.

The high steam and oxygen flow indicators and the low range oxygen indicator were found to be accurate, but the low steam range flow instrument was giving a much lower indicated flow than the actual value.

The effect of this was that the steam/oxygen ratio, instead of being 1.34 was actually 1.88 to 1. This ratio would give a considerably higher carbon dioxide level in the flare gas.

The limestone fluxing on EPRI - 01 will add about 1/3% CO_2, so that at a steam/oxygen ratio of 1.88 during this run the CO_2 level should be about 6.9%, which is close to the value recorded by the mass spectrometer during this period of EPRI - 01, confirming the error and correction.

BEHAVIOUR OF GASIFIER DURING TRANSIENTS

During run EPRI - 01 the gasifier was subjected to several programmed load changes which are represented generally in Table 10-1. Most of these changes were done in pressure control, that is, there was tight control at the gasifier front end on the steam/oxygen flows, with consequent floating and variation in pressure and flow in particular at the back end. Apart from start up, the transients can be listed as follows:

- Rate reduction from 160,000 SCF/H oxygen (H_2O/O_2 = 1.3) to 160,000 SCF/H in ten minutes, then settling down to steady running at 160,000 SCF/H oxygen.
- Rate increase from 110,000 SCF/H oxygen (H_2O/O_2 = 1.3) to 160,000 SCF/H in ten minutes, then settling down to steady running at 160,000 SCF/H oxygen.
- Rate reduction from 160,000 SCF/H oxygen (1.30 H_2O/O_2) to 110,000 SCF/H oxygen and back in ten minutes.
- Rate reduction from 160,000 SCF/H oxygen (2.30 H_2O/O_2) to 110,000 SCF/H oxygen and back in three minutes.
- Rate reduction from 160,000 SCF/H oxygen (1.30 H_2O/O_2) to 80,000 SCF/H oxygen in 35 minutes, then settling down to a steady running 80,000 SCF/H oxygen.
- Rate increase to 160,000 SCF/H oxygen (1.30), from 80,000 SCF/H oxygen in 35 minutes then settling down to a steady running at 80,000 SCF/H oxygen.
- Rate decrease to 50,000 SCF/H oxygen from 160,000 (1.30 H_2O/O_2). This also involved an alteration in H_2O/O_2 to 1.88.
- Rate increase from 50,000 SCF/H oxygen to 180,000 SCF/H oxygen, steam/oxygen also changing from 1.88 to 1.30, followed by a steady running at 180,000 SCF/H just prior to shutdown. This change was performed in flow control and involved a stop at 80,000 for four hours before going back to 50,000 SCF/H to check the sets and then bringing the gasifier up over about three hours to 180,000 SCF/H.

The above changes were carried out without causing any gasifier upsets. No problems were encountered that could be attributed to dust carry over, and manual operational requirements were untroubled.

A qualitative picture of the sharp spike down to 100,000 SCF/H from 160,000 SCF/H oxygen is given in Figure 10-5. This appeared to have no signficant effect upon gas analysis, as is seen in Figures 10-6 to 10-8.

Similar lack of significant change was observable at the rest of the transients. The only consistent discernable effect was the greater contribution of nitrogen to the total gas composition at lower loads, an expected phenomenon as the amount of nitrogen to the gasifier was roughly constant throughout the run.

POST RUN INSPECTION

The gasifier was shut down a few minutes after the bottom cone of the coal lock had cleared of coal, with the distributor being stopped at the same time as the tuyeres were switched off. Tar injection to the distributor had been stopped an hour previously.

After cooldown, the gasifier was opened and the contents inspected. There was some caked coal at the top of the bed, which had been broken down into reasonable sized lumps before the coal left the influence of the stirrer, and the rest of the bed was full of good char and showed no inhomogeneities or large caked agglomerations.

CONCLUSIONS

Run EPRI - 01 successfully met its objectives. During steady state running, the gasifier showed steady performance at 160,000 SCF/H oxygen (100% loading), at 110,000, at 80,000 and at 50,000 SCF/H oxygen. At all of the above loadings the gasifier showed that it was capable of sustained running, with the possibilities of even lower turndowns than the 30% reached on this run.

Change from one loading to another was carried out smoothly according to schedule, and produced no gasifier upsets and minimal transient phenomena such as varying gas analysis. The gasifier did not object to being controlled in the back end, flow control mode, although further work is needed to establish best operating conditions under this latter regime.

The available evidence suggests that turnup and turndown in load can be achieved very quickly on the gasifier itself when working on the medium caking and swelling Rossington coal.

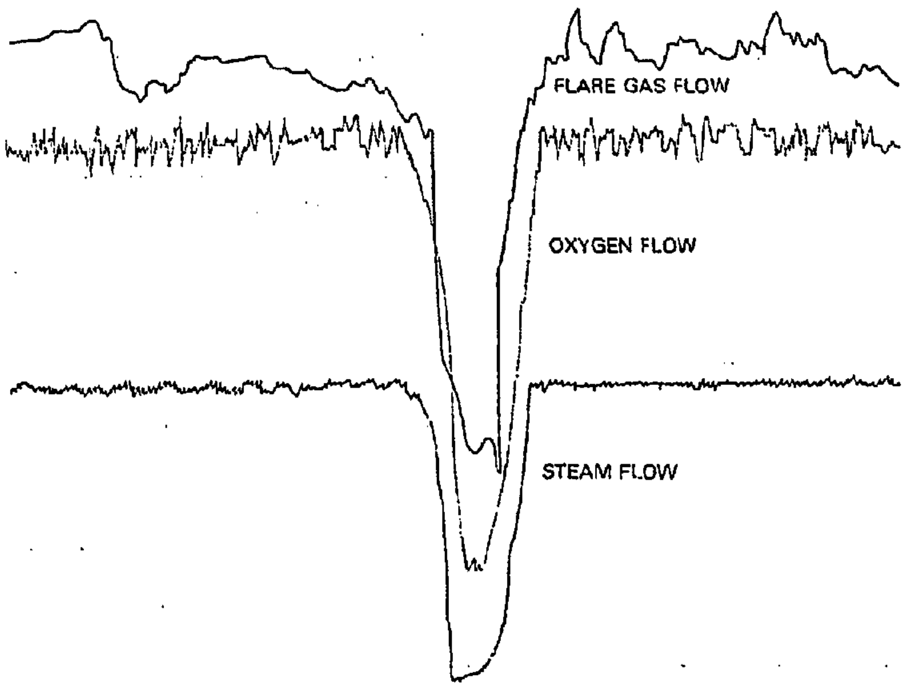


Figure 10-5. Spike Down to 110,000 SCF/H from 160,000 SCF/H
Recorder Speed: 2 Min./CM

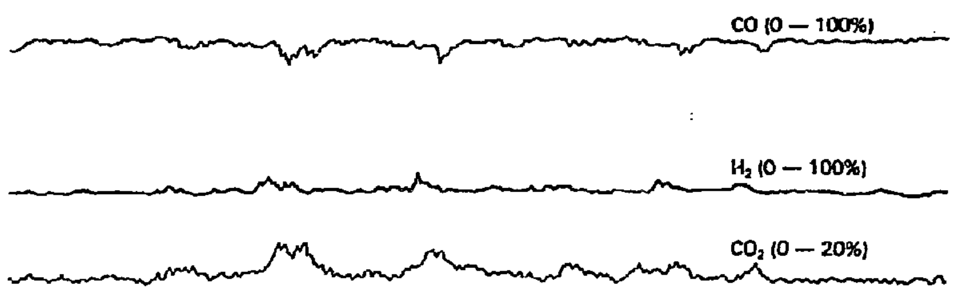


Figure 10-6. Major Gas Composition at Flare Across Sharp Spike
Chart Speed: 2 Min./CM

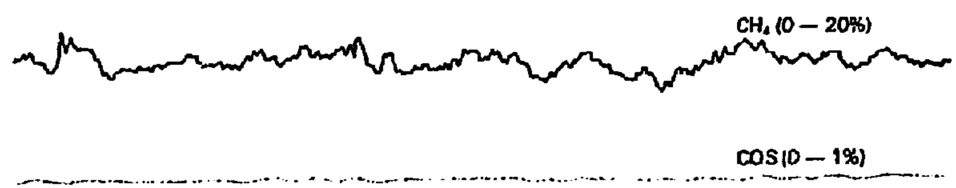


Figure 10-7. Minor Gas Composition at Flare Across Sharp Spike
Chart Speed: 2 Min./CM

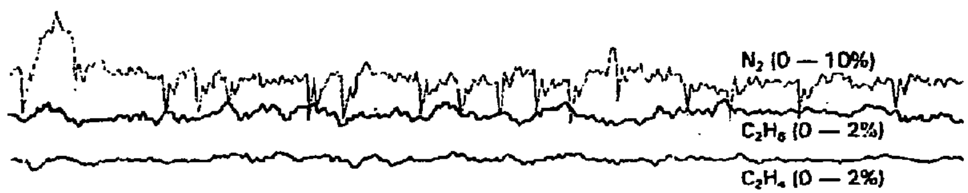


Figure 10-8. Minor Gas Composition Across Sharp Spike
Chart Speed: 2 Min./CM

All gasifier systems worked well during the run, and post run inspection revealed no significant wear or damage to gasifier internals.

Section 11

REPORT ON TEST RUN EPRI - 02

SUMMARY

EPRI - 02 was planned to essentially repeat run EPRI - 01, but to use Pittsburgh 8 coal instead of Rossington. Fluxing on Pittsburgh 8 coal would be on BFS and not limestone, as no experience had been obtained on the latter flux with Pittsburgh 8 coal.

The run was done in two parts because the first attempt (EPRI - 02A) was frustrated by an incurable leak at a tuyere flange after only 36 hours on line which necessitated shutdown for repairs. The run was then restarted as EPRI - 02B.

EPRI - 02A started early on 7 November, and after a standard startup on Rossington coal, the loading was reduced to 80,000 SCF/H oxygen, and Pittsburgh 8 coal introduced to the gasifier. Fluxing was with limestone and not with Blast Furnaces Slag (BFS) as originally planned, as there were problems with the BFS weigh hopper. On limestone there were problems with black tuyeres caused by underfluxing, these problems not being satisfactorily solved until BFS fluxing was restored at 130,000 SCF/H loading at 06:25 on 8 November. However, shortly after this, an incurable leak at the back end of a tuyere developed, enforcing a shutdown at 15:32.

The gasifier was unloaded after cool down, and the necessary repairs and checks made, and run EPRI - 02B was started on 21 November with steam/oxygen being introduced to the gasifier at 16:45. A good standard startup on Rossington coal fluxed with limestone was obtained and the rate was then dropped to 130,000 SCF/H in preparation for changeover to Pittsburgh 8 coal with BFS replacing limestone fluxing. Changeover to the latter fuel was started at 22:38 and a good transition obtained, except that a tuyere went black. This was cleared by 13:05 on 22 November, by which time the loading had been brought up to 160,000 SCF/H oxygen. Running on this loading was good, and performance tests were carried out and control in the flow mode tried. At 05:15 on 23 November the load was reduced to 110,000 SCF/H oxygen at 1.20 steam/oxygen, where it was held for 18 hours before being brought back up again to 160,000 SCF/H loading. At this loading the rates were spiked down to 110,000 SCF/H oxygen

in flow control without any gasifier upsets occuring. Early in the morning of the 24th there were problems in the quench chamber due to an instrument fault.

By 10:00 hours the problem had become severe, so it was decided to go on standby, cool the gasifier down and depressurise, thus allowing entry to remedy the problem. The gasifier was put on standby at 10:11 and cooled down and depressurised. The manway was taken off at 19:05 to reveal the problem which was rectified readily. The gasifier was then closed at 21:00 hours and restarted successfully on Rossington coal at 130,000 SCF/H oxygen loading, before being switched over to Pittsburgh 8 coal at 03:33 on 25 November. A good transition to this fuel was obtained, and at 06:38, with the gasifier in flow control, the load was brought down to 80,000 SCF/H oxygen. Performance data was gathered at 00:38 on 26 November, with running being established at this loading before the rates were dropped to 50,000 SCF/H oxygen by 05:21 hours. Further data was collected at this loading before the rates were again brought up to 160,000 SCF/H oxygen loading, settled down and then dropped to 110,000 SCF/H oxygen. Two performance tests were done at this loading, one with no tar injection to the top of the bed and one with high tar (100% pump stroke). The run was then terminated by standard procedures at 12:20 on 27 November, after nearly six days continuous running, including the cold standby period.

Post run inspection revealed satisfactory conditions. A schematic of the run is given in Figure 11-1. Table 11-1 summarises the load changes made during the run.

EQUIPMENT, INSTRUMENTS AND CONTROLS

No changes were made to gasifier configurations as compared to EPRI - 01, although continued improvement of instrumentation and data logging systems were carried out between runs EPRI - 01 and EPRI - 02.

The control system was as for run - 01 except that the size of the flare gas control valve was reduced from six inches to four inches in diameter.

RUN DIARY

Start up phase for EPRI - 02A began at 01:25 on 7 November, 1979. Steam/oxygen at start-up rates was admitted down the tuyeres at 03:03 and Rossington coal charging commenced, with limestone fluxing. A good startup was achieved, and the gasifier

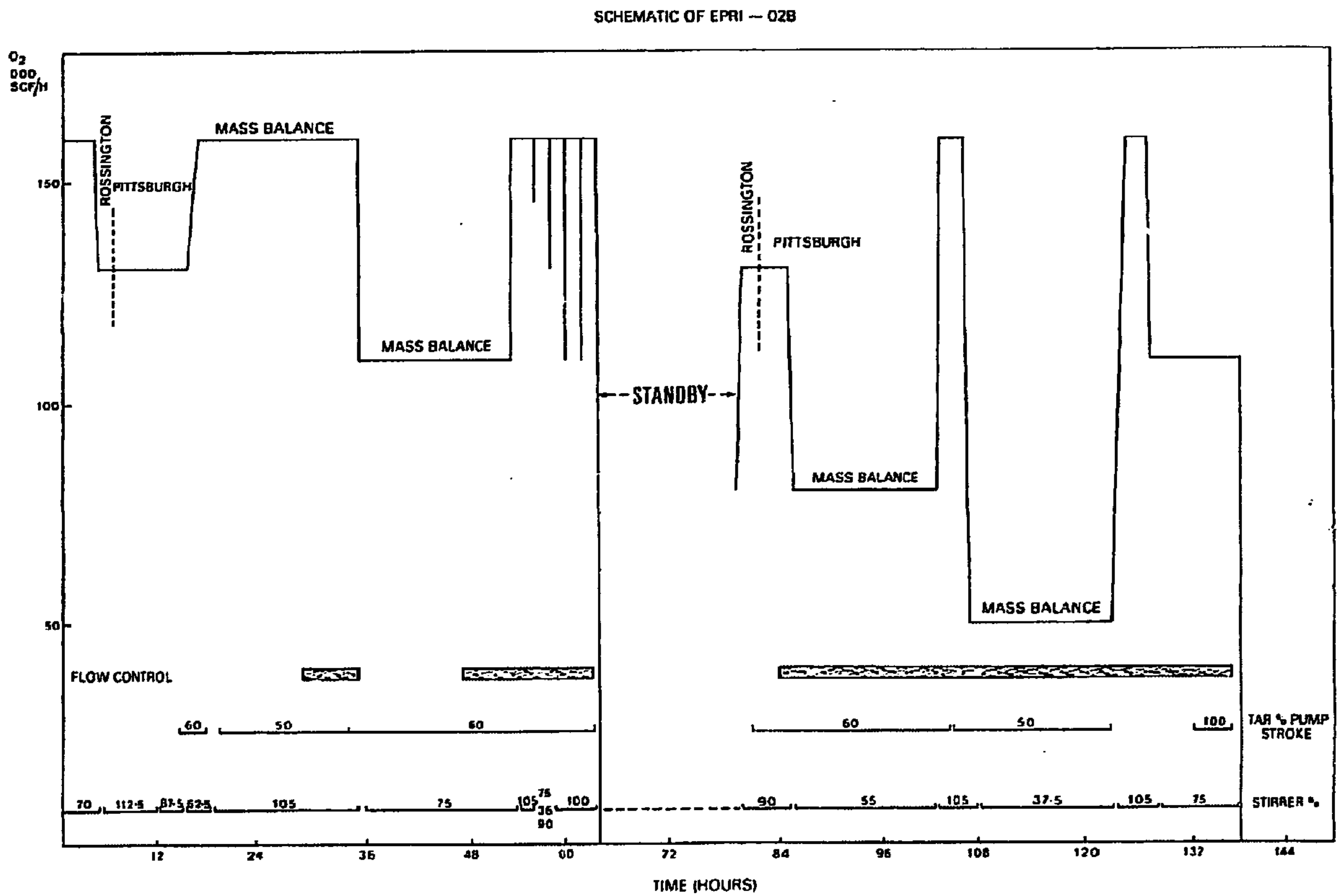


Figure 11-1. Schematic of EPRI - 02B

TABLE 11-1. DESCRIBING LOAD CHANGE PERIODS FOR EPRI - 02A AND - 02B

Change No.	Load 1 before Change (1000 SCFH O_2 blast)	Load 2 after Change (1000 SCFH O_2 blast)	Time Taken (mins.)	Change in 1000 SCFH	Coal Gasified	Control Mode
1	160	80	25	80	Rossington	P
2	80	130	48	50	Pitts. 8	P
End of EPRI 2A						
3	160	130	9	30	Rossington	P
4	(130	150	4	20	Pitts. 8	P
	(150	160	2	10	Pitts. 8	P
5	160	110	5	50	Pitts. 8	P
6	110	160	8	50	Pitts. 8	P
7	160	140	5	20	Pitts. 8	F
8	140	160	6	20	Pitts. 8	F
9	160	130	3	30	Pitts. 8	F
10	130	160	5	30	Pitts. 8	F
11	160	110	7½	50	Pitts. 8	F
12	110	160	6	50	Pitts. 8	F
13	160	110	4	50	Pitts. 8	F
14	110	160	3	50	Pitts. 8	F
15	130	80	30	50	Pitts. 8	F
16	80	160	35	80	Pitts. 8	F
17	160	50	80	110	Pitts. 8	F
18	50	160	72	110	Pitts. 8	F
19	160	110	12	50	Pitts. 8	F

was settled down under standard conditions of 160,000 oxygen loading, steam/oxygen at 1:30. Gasifier pressure was kept slightly low at 335 psig in anticipation of a fluctuating pressure when flow control was engaged; thus the lifting of a relief valve at WHB could be avoided.

Running under standard conditions was satisfactory, with all tuyeres bright, good slag tapping in automatic, and steady bed behaviour. At 08:45 the rates were reduced to 110,000 SCF/H oxygen followed by a reduction to 80,000 SCF/H. Running under these conditions was satisfactory and at 19:03 preparations were started to change over to Pittsburgh 8 coal. The south flux weighing hopper, which works on BFS, was found to be faulty, so the decision was taken to flow the Pittsburgh 8 coal with limestone, initially (at a flux/ash ratio of about 0.6). The first lock of Pittsburgh 8, fluxed in this way, was charged to the gasifier at 19:37, and a good transition to this fuel was achieved at 80,000 SCF/H oxygen loading, with steam/oxygen 1.30. Slag tapping was good with all tuyeres bright, although bed behaviour was more unsteady as compared to Rossington coal at this loading. However, conditions deteriorated overnight with tuyeres going dim and black so it was decided to purge out the gasifier with Rossington coal to establish good conditions. On switching back to Pittsburgh coal the tuyeres again went black so Rossington coal was again supplied as feed. The fluxing rate with limestone for Pittsburgh coal was clearly too low and needed further investigation.

There was also a problem with a leak to a tuyere flange and emergency repairs to this joint were carried out. By 10:54 the gasifier was working satisfactorily on Rossington coal, and as an emergency system was now available for BFS charging, it was decided to restore Pittsburth 8 feed with BFS fluxing and this was done at 13:08, with loading 130,000 SCF/H oxygen and steam/oxygen 1.30. A good transition was obtained and running was satisfactory under these conditions, but the leak to the tuyere flange became worse and with no possibility of on line repairs being carried out a controlled shut down was carried out at 15:32 hours.

The gasifier was cooled down and unloaded. The leak was repaired and other tuyeres checked and preparations made for restart and continuation of EPRI - 02 (EPRI - 02B). These were completed by 21 November at 11:30. Standard start up procedures were slightly delayed due to a problem with the coal lock hydraulics, but by 16:45, steam/oxygen were introduced to the gasifier at start up rates. Rossington coal charging was started with fluxing with limestone.

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Rossington coal was established under standard conditions with no problems and the rate was then brought down to 130,000 SCF/H oxygen in preparation for accepting Pittsburgh 8 coal. Fluxing was changed to BFS and at 11:38 the first lock of Pittsburgh 8 coal was charged to the gasifier. A good transition to this fuel was obtained. The steam/oxygen ratio was decreased to 1.10 and the fluxing rate increased slightly.

At 10:00 tar injection was brought on to the gasifier top and the loading was brought to 160,000 SCF/H. This increase in loading produced no problems, and the steam/oxygen ratio was brought to 1.20 at 13:05. Slag tapping remained good with all tuyeres bright. Performance data was collected at this loading, with a side stream being started at 21:20 and the gasifier put into the flow control mode at 23:25.

Running at this loading continued into 23 November and at 05:24 the rates were brought down to 110,000 SCF/H oxygen in pressure control. Steady running at this load was achieved with no sign of any transients due to the change in rates. A mass balance period was started at 11:30.

The gasifier was put into flow control at 17:00 and running continued to be satisfactory.

At 23:53 the rate was increased to 160,000 SCF/H, again in pressure control, with the gasifier put back in the flow control mode at 00:25. A good transition was obtained with no problems being encountered. During the early morning of 24 November some sharp spikes down to 110,000 SCF/H and back were done in flow control, with the controller settings being tuned to eventually allow the complete exercise to be carried out in seven minutes without causing any process problems. By 09:00 it was apparent there was a problem in the quench chamber so the decision was taken to do a cold standby, remove the manway door and investigate the problem. At 10:11 the preparations for standby were complete. By 18:15 the gasifier was nominally at zero pressure. Inspection of quench chamber was carried out from the manway door. The fault was rectified.

The manway door was replaced and it was decided to come back on line with Rossington coal. Steam/oxygen at start up rates was admitted to the gasifier at 01:28, with the first lock of Rossington coal, fluxed with limestone, being charged at 01:35. At 01:57 the gasifier had reached its planned loading of 130,000 SCF/H oxygen, with steam oxygen 1.20 and pressure 335 psig. All tuyeres were coming bright and a good restart had been obtained with no problems. At 03:06 the steam/oxygen ratio was trimmed to 1.30 and at 03:33 Pittsburgh 8 was charged to the gasifier with fluxing being switched to BFS. An excellent transition to Pittsburgh 8 was obtained and at 06:38 the gasifier was reduced to 80,000 SCF/H oxygen loading in flow control. The gasifier was settled down at this loading and performance data collected, with a side stream being run. Running was good at this loading and at 00:38 on 26 November the load was brought up to 160,000 SCF/H oxygen loading, without any upsets occurring, over half an hour.

With steady conditions established at the above loading, the rates were dropped to 50,000 SCF/H oxygen at 05:21. Tar injection to the gasifier top set at 50% pump stroke. Good running was obtained and performance data was collected at this loading with a side stream being run. At the end of this period, at 23:35, the gasifier loading was brought up again to 160,000 SCF/H oxygen.

Both the change down in loading to 50,000 SCF/H oxygen, then later back again to full loading proved to be satisfactory operations with the changes themselves producing no observable transients in gasifier behaviour. Good running at full load was had into 27 November and at 01:40 the steam/oxygen flow was reduced again, this time to 110,000 SCF/H oxygen at 1.30 H_2O/O_2. This loading was chosen to investigate the effect of tar injection at the bed top. Initially the pump was turned off and then, at 07:00 brought on at a maximum stroke. This latter process alteration brought about a slight, but significant effect upon bed behaviour, with offtake temperature, stirrer torque, CO_2 at offtake and bed DPs being more variable at high tar loadings to the top of the gasifier.

At 11:30 preparations were made for shut down with the tar injection to the gasifier top being turned off, and this was carried out in a standard manner at 12:20.

The run is shown sehcmatically in Figure 11-1. A breakdown of the major run periods for EPRI - 02 is given in Table 11-2.

SLAG TAPPING AND SLAG REMOVAL

EPRI - 02A started by fluxing the Pittsburgh 8 coal with limestone and the relatively low levels of this flux used led to poor hearth conditions, particularly with regard to tuyeres. When a further attempt was made to re-introduce Pittsburgh 8, again fluxed with limestone the same problems recurred, and it was not until Pittsburgh 8 was fluxed with BFS that completely satisfactory tapping was obtained,

TABLE 11-2. SIMPLIFIED RUN DIARY OF EPRI - 02

Date	Time	Period No.	Comments
7 Nov 79	03.03	-	Steam/oxygen in
(EPRI - 02A)	03.35		Rates 160,000 $SCFH^{-1}$ oxygen ratio 1.2/1.
	05.00		Tar injection on at 60%. Ratio 1.3/1
	07.58		Limestone and BFS in. Rates 110,000.
	08.50	2	Rates down to 80,000 $SCFH^{-1}$.
	09.15		PB changed from 40% to 80%.
	10.30		PB - 400%. RESET 1.
	14.15		Tar injection cut to 50%.
	14.20		Tar injection cut to 40%.
	15.45		PB - 400%. RESET - 0.2.
	16.30		PB - 400%. RESET - 0.1.
8 Nov 79	00.58		Reducing steam/oxygen to 1.1/1
	01.23		Steam/oxygen to 1.3/1.
	02.17		All tuyeres black.
	02.25	3	Rates now 100,000 $SCFH^{-1}$.
	02.38		Rates increased to 120,000 $SCFH^{-1}$.
	02.58		Rates to 130,000 $SCFH^{-1}$. Steam/oxygen to 1.1/1.
(EPRI - 02A)	03.44		Locked in Rossington.
	03.57		Tar injection on at 40%.
	04.48		Steam/oxygen to 1.3/1.
	06.10		PB - 100%.
	06.25		Pittsburgh in. Tar injection 60%.
	07.15		Tar injection 100%.
	07.30		Tar injection 60%.
	08.50		Tar injection off.
	09.00		Rossington and BFS.

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Date	Time	Period No.	Comments
	09.50		Steam/oxygen to 1.1/1.
	11.48		Steam/oxygen to 1.3/1.
	13.08		Pittsburgh + BFS.
	15.32		Shut down - leak in tuyere.
21 Nov 79			
(EPRI - 02B)	16.45		Steam/oxygen in.
	17.45		Rates 160,000 $SCFH^{-1}$ oxygen.
	18.45		Tar injection on at 100%.
	19.33		Tar injection 60%.
	20.15		Rates now 130,000 $SCFH^{-1}$ oxygen.
	21.44		Rossington + BFS.
	22.38		Pittsburgh + BFS.
22 Nov 79	00.00		Steam/oxygen to 1.2/1.
	00.47		Flux increased.
	01.00		Steam/oxygen to 1.1/1.
22 Nov 79	02.06	5	Flux up.
(EPRI - 02B)	07.05		Flux down.
	07.50		Flux up.
	10.20		Increasing rates.
	11.35	6	Rates 160,000 $SCFH^{-1}$.
	12.42		Tar injection off.
	13.05		Steam/oxygen to 1.2/1.
	14.18		Tar injection at 50%.
	17.20		Start of mass balance period.
	19.10		PB - 200%. RESET - 2.
	21.30		PB - 600%. RESET - 1.
	23.30		PB - 40%. RESET - 0.1.
	23.35		Into flow control.
23 Nov 79	00.15		Reverted to pressure control.
	00.18		Back to flow control.

Date	Time	Period No.	Comments
	01.47		PB - 30%. RESET - 1.
	02.35		In pressure control.
	02.42		Back to flow control.
	03.15		PB - 100%. RESET - 2.
	04.00		Tar injection 60%.
	05.04		PB - 200%. RESET - 2.
	05.10	7	Into pressure control.
	05.15		Reducing rates.
	05.24	A	Rates 110,000 SCFH^{-1} oxygen.
	08.15	Pressure	Tar injection 50%.
	11.30	Control	Start of mass balance period.
	17.00		PB - 100%. RESET - 2.
		B	
		Flow	Into flow control.
23 Nov 79	23.25	Control	Into flow control.
	23.31	8	PB - 200%. RESET - 2.
(EPRI - 02B)	23.42		Increasing rates.
	23.53		Rates 160,000 SCFH^{-1} oxygen.
24 Nov 79	00.25	C	Into flow control.
	01.03	160,000	PB - 100%. RESET - 2.
	02.50	in Flow	Spiking down.
	02.55		Only down to 145,000 SCFH^{-1}.
	03.29	Control	PB - 40%. RESET - 2.
	03.50		PB - 40%. RESET - 0.1.
	04.30	(mainly)	Reducing flow to 110,000 SCFH^{-1}.
	04.44	with	Rates increased to 160,000.
	08.07	Spikes	PB - 30%. RESET - 0.1.
	08.57		10 minute spike down to 110,000.
	10.11		Cold standby.
25 Nov 79	01.28		Steam/oxygen in. Rates to 130,000.
	01.35		Rossington + $CaCO_3$ locked in.

Date	Time	Period No.	Comments
	02.26	D	Tar inj on on at 60%.
	02.38	Restart	Steam/oxygen to 1.1/1.
	03.06	on	Steam/oxygen to 1.3/1.
	03.30	Rossington	Pittsburgh + BFS.
25 Nov 79	04.55	E	PB - 100%. RESET -2.
	05.30	Restart on	Into flow control.
	06.00	Pittsburgh	PB - 30%. RESET - 0.1.
	06.38	130,000	Reducing rates.
	06.45	9	Tar injection 50%.
	07.13	10	Rates 80,000 SCFH^{-1} oxygen.
		F	
		80,000	
	11.00	Pittsburgh	Start of mass balance period.
		in	
		Flow Control	
	23.02	11	Into pressure control.
	23.25		Into flow control.
26 Nov 79	00.38		Increasing rates.
	00.58		PB - 30%. RESET - 0.1.
	01.12		Rates 160,000 SCFH^{-1} oxygen.
		Period at	
	01.58	160,000	PB - 30%. RESET - 0.1.
		12	
	04.05	G	Starting to reduce rates.
	04.54		
	05.21		Rates 50,000 SCFH^{-1} oxygen.
		55%	
		50,000	
		37.5%.	Tar injection 40%.
	09.42	H_2	Tar injection 50%.
	17.36		PB - 30%. RESET - 0.2.

Date	Time	Period No.	Comments
	21.33	13	PB - 30%. RESET - 0.5. Starting to increase rates.
	23.35	I 160,000	Rates 160,000 $SCFH^{-1}$ oxygen.
27 Nov 79	01.00		Tar injection off.
	01.30	14a	Starting rate reduction.
(EPRI - 02B)	01.40	J NO TAR 110,000	Rates 110,000 $SCFH^{-1}$ oxygen. PB - 30%. RESET - 0.1.
	07.00	14b K	Tar injection 100%.
	11.30	7.00-11.30 HIGH TAR 110,000	Tar injection off. Into pressure control.
	12.20		Shut down.

and by then the run had to be terminated because of a bad leak at a tuyere main flange.

For the continuation of the run, EPRI - 02B, it was decided to flux with BFS at a flux ash ratio, on Pittsburgh 8, of about 1.60:1, and this high fluxing rate led to satisfactory slag tapping performance; with high slag discharge rates under all conditions. Return from standby was followed immediately by good hearth conditions.

FLUXING SYSTEMS

Fluxing on Pittsburgh 8 was planned to be on BFS throughout EPRI - 02, using the south bunker and the weigh cell system which enabled the weight of slag charged to each coal lock to be accurately estimated. Unfortunately, there was a problem with the systems early in EPRI - 02A and it was decided to go over to the north bunker and flux the Pittsburgh 8 coal with limestone.

Fluxing on limestone was initially at a flux to ash ratio of about 0.5, but this gave poor hearth conditions, particularly with respect to black tuyeres, and the fluxing load was later increased to flux ash ratio of 0.6. This again proved to be too low a fluxing level. At this stage in the run BFS flux was again available so this flux was switched to a flux/ash ratio of 1.6. This flux and fluxing level proved satisfactory.

For EPRI - 02B, fluxing with blast furnace slag flux was used throughout. This was brought up to a level corresponding to a flux to ash ratio of 1.6.

The flux/coal ash-slag balance is given in Table 11-3. However, no attempt was made to optimise the fluxing. Table 11-3 shows good agreement between the calculated and experimental slag composition for all the run periods analysed. The discrepancy in iron figures shows that more than the assumed 20% of iron total is reduced to free iron.

BED BEHAVIOUR

The bed behaviour throughout run EPRI - 02 on Pittsburgh 8 coal was satisfactory, causing no operational upsets, although it was generally less stable than for run EPRI - 01 on Rossington coal.

TABLE 11-3. EXPERIMENTAL AND CALCULATED SLAG COMPOSITIONS FOR PITTSBURGH 8 FLUXED WITH BFS: EPRI - 02B

Period	5		6		A		B		C		F		H		J		K	
Flux Rate per Lock	1278.7 lbs		1305 lbs		1305 lbs		1305 lbs		1305 lbs		1305 lbs		1305 lbs		1305 lbs		1305 lbs	
Flux: Ash	1.61		1.62		1.64		1.63		1.59		1.69		1.56		1.47		1.49	
	Actual*	Calculated †	Actual	Calculated	Actual*	Calculated †	Actual	Calculated	Actual*	Calculated	Actual	Calculated	Actual	Calculated	Actual*	Calculated	Actual*	Calculated
SiO	40.17	40.49	40.17	40.47	40.54	40.43	40.91	40.47	40.68	40.57	40.45	40.97	39.30	41.14	39.30	41.44	39.30	41.38
Al O	16.94	17.62	16.94	17.60	17.02	17.49	17.09	17.44	17.05	17.52	17.03	17.57	16.23	17.81	16.23	17.67	16.23	17.63
Fe O	4.43	4.82	4.43	5.07	3.55	4.88	2.67	5.00	3.34	5.11	3.42	4.93	2.19	5.18	2.19	4.80	2.19	4.77
CaO	25.09	23.83	25.09	23.88	26.04	24.02	27.00	24.00	26.05	23.79	23.60	24.28	26.58	23.56	26.58	23.34	26.58	23.46
MgO	6.89	6.56	6.89	6.67	6.95	6.57	7.00	6.54	6.76	6.48	6.47	6.77	7.32	6.56	7.32	6.47	7.32	6.50
Na O	0.46	0.50	0.46	0.51	0.44	0.48	0.43	0.45	0.44	0.45	0.41	0.46	0.42	0.52	0.42	0.52	0.42	0.52
K O	1.12	1.26	1.12	1.26	1.12	1.18	1.12	1.11	1.14	1.12	1.14	1.16	1.12	1.27	1.12	1.23	1.12	1.23
Silica Ratio	51.39	52.27	51.39	52.04	51.55	52.13	51.70	52.17	51.88	52.33	53.58	52.14	51.09	52.59	51.09	53.26	51.09	53.15

† Iron Separation assumed rate 20% throughout

* Average Values i.e. period under study outwith time interval of sampling

The steady run periods which were analysed in some depth are detailed in Table 11-4. The average offtake temperatures for these periods are given in Table 11-5. For normal behaviour on Rossington at standard rates the frequency of above temperature excursions is general of the order of 1.0 or less per hour. Figure 11-2 plots the frequency of high temperature excursions as a function of oxygen loading for EPRI - 02B shows, but not unexpectedly, a marked trend to less stable behaviour at high loads. This effect also shows up in the general bed DP behaviour, the stirrer torque and the CO_2 analysis at the offtake, all of which become more ragged at high loads.

During periods A, C and E, bed behaviour was generally less stable than in the other periods studied. A comparison of periods A and B shows a marked improvement in gasifier behaviour obtained in going from pressure control to flow control. This latter effect was generally observable throughout the run.

The offtake temperature profile shows a slowly undulating pattern with reduced bed stability on Pittsburgh 8 coal compared with Rossington.

No significant trend to higher mean offtake temperatures with increasing oxygen loading was obtained from this run, in contrast with the trend observed on EPRI - 01.

Tar injection to the bed top was almost constant throughout the run and insufficient data were obtained to estimate the effect of this parameter on bed behaviour.

PLANT BEHAVIOUR IN FLOW CONTROL

Experience obtained during run - 01 indicated that very satisfactory steady flow control operation could be obtained when running on Rossington coal. Run - 02 was indended to compare the plant behaviour on run - 01 with that obtained on Pittsburgh coal.

The control system was as for run - 02 except that the size of the flare gas control valve was reduced from six inches to four inches in diameter.

The control characteristic settings were changed at various times throughout the run to obtain a number of different periods. The actual settings and their effects on the steam/oxygen flows, gasifier pressure and flare gas flows are shown in Table 11-6 and Figure 11-3. PB is the proportional band and the reset is shown as repeats per minute.

TABLE 11-4. TEST PERIODS ANALYSED: EPRI - 02B

Period	Coal	BFS	Flux Limestone	Oxygen Loading (SCFH)	$Steam/O_2$ Ratio	Stirrer Speed %	% Tar Injection	
							Tuyeres	Distributor
A	Pitts. 8	Yes	0	110,000	1.2	85.75	0	60,50
B	Pitts. 8	Yes	0	110,000	1.2	75	0	50
C	Pitts. 8	Yes	0	160,000	1.2	Variable	0	50
D	Ross.	0	Yes	130,000	1.1,1.3	90	0	60
E	Pitts. 8	Yes	0	130,000	1.3	90	0	60
F	Pitts. 8	Yes	0	80,000	1.3	55	0	50
G	Pitts. 8	Yes	0	160,000	1.3	55	0	50
H1	Pitts. 8	Yes	0	50,000	1.3	55	0	50
H2	Pitts. 8	Yes	0	50,000	1.3	37.5	0	40,50
I	Pitts. 8	Yes	0	160,000	1.3	100.5	0	50,0
J	Pitts. 8	Yes	0	110,000	1.3	75	0	0
K	Pitts. 8	Yes	0	110,000	1.3	75	0	100,0

TABLE 11-5. OFFTAKE TEMPERATURES EPRI - 02B

Test Period	No. 3 Offtake Mean °F	No. 4 Offtake Mean °F
A	933	933
B	945	927
C	928	929
D	873	873
E	964	965
F	920	912
G	908	914
H1	796	794
H2	890	884
I	914	923
J	894	888
K	938	931

p

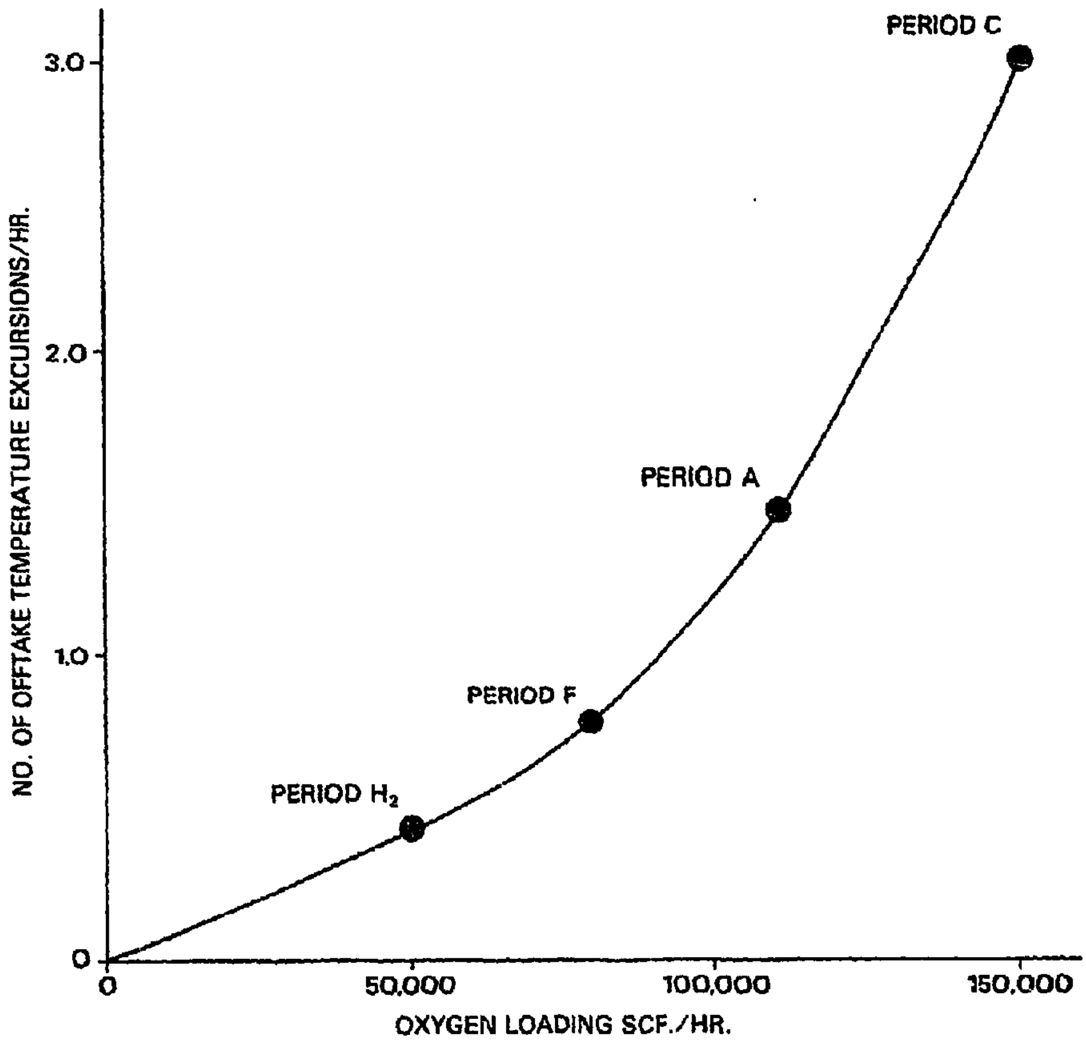


Figure 11-2. Graph of Offtake Temperature Excursions VO_2 Loading

TABLE 11-6. CONTROLLER SETTINGS DURING EPRI - 02

Con No.	Day and Time of Setting Up This Condition	Coal Type	(PC) Gasifier Pressure		Flare Gas Flow (GFC)		Oxygen Flow (OFC)		Steam Flow (SFC)		Approx. Oxygen Blast	Gasifier Control Mode	Pressure Tapping Point
			PB	Reset	PB	Reset	PB	Reset	PB	Reset			
1	7.11.79 - 04.00	Rossington	40	5	1000	1	700	10	200	20	160,000	Pressure	100 H
2	- 08.50	Rossington	40	5	1000	1	700	10	200	20	80,000	Pressure	100 H
3	- 09.15	Rossington	80	5	1000	1	700	10	200	20	80,000	Pressure	100 H
4	- 10.30	Rossington	400	10	1000	1	700	10	200	20	80,000	Pressure	100 H
5	- 12.20	Rossington	400	10	1000	1	700	10	200	20	80,000	Pressure	201
6	- 15.45	Rossington	400	2	1000	1	700	10	200	20	80,000	Pressure	201
7	- 16.30	Rossington	400	1	1000	1	700	10	200	20	80,000	Pressure	201
8	- 19.37	Pittsburgh	400	1	1000	1	700	10	200	20	80,000	Pressure	201
9	8.11.79 - 02.58	Pittsburgh	400	1	1000	1	700	10	200	20	130,000	Pressure	201
10	- 03.44	Rossington	400	1	1000	1	700	10	200	20	130,000	Pressure	201
11	- 06.10	Pittsburgh	100	0.1	1000	1	700	10	200	20	130,000	Pressure	201
12	- 09.00	Rossington	100	0.1	1000	1	700	10	200	20	130,000	Pressure	201
13	21.11.79 - 17.45	Rossington	40	5	1000	1	300	5	200	5	160,000	Pressure	202 A
14	- 19.45	Rossington	40	2	1000	1	300	5	200	5	160,000	Pressure	202 A
15	- 20.15	Rossington	40	2	1000	1	300	5	200	5	130,000	Pressure	202 A
16	- 22.38	Pittsburgh	40	2	1000	1	300	5	200	5	130,000	Pressure	202 A
17	22.11.79 - 11.35	Pittsburgh	40	2	1000	1	300	5	200	5	160,000	Pressure	202 A
18	- 19.10	Pittsburgh	200	20	1000	1	300	5	200	5	160,000	Pressure	202 A
19	- 21.30	Pittsburgh	600	10	1000	1	300	5	200	5	160,000	Pressure	202 A

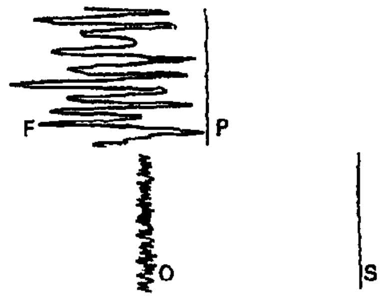
TABLE 11-6 (Con't)

Con No.	Day and Time of Setting Up This Condition	Coal Type	(PC) Gasifier Pressure		Flare Gas Flow (GFG)		Oxygen Flow (OFC)		Steam Flow (SFC)		Approx. Oxygen Blast	Gasifier Control Mode	Pressure Tapping Point
			PB	Reset	PB	Reset	PB	Reset	PB	Reset			
21	22.11.79 - 23.33	Pittsburgh	40	0.1	1000	1	300	5	200	5	160,000	Flow	202 A
22	23.11.79 - 01.47	Pittsburgh	30	1	1000	1	300	5	200	5	160,000	Flow	202 A
23	- 03.15	Pittsburgh	100	2	1000	1	300	5	200	5	160,000	Flow	202 A
24	- 05.24	Pittsburgh	200	20	1000	1	300	5	200	5	110,000	Pressure	202 A
25	- 06.42	Pittsburgh	200	20	1000	1	300	5	200	5	110,00	Pressure	201
26	- 17.00	Pittsburgh	100	2	1000	1	300	5	200	5	110,000	Flow	202 A
27	- 23.32	Pittsburgh	200	20	1000	1	300	5	200	5	110,000	Pressure	202 A
28	- 23.53	Pittsburgh	200	20	1000	1	300	5	200	5	160,000	Pressure	202 A
29	24.11.79 - 00.25	Pittsburgh	200	20	1000	1	300	5	200	5	160,000	Flow	202 A
30	- 01.03	Pittsburgh	100	2	1000	1	300	5	200	5	160,000	Flow	202 A
31	- 03.29	Pittsburgh	40	2	1000	1	300	5	200	5	160,000	Flow	202 A
32	- 03.50	Pittsburgh	40	0.1	1000	1	300	5	200	5	160,000	Flow	202 A
33	- 08.07	Pittsburgh	30	0.1	1000	1	300	5	200	5	160,000	Flow	202 A
34	25.11.79 - 01.35	Rossington	30	0.1	1000	1	300	5	200	5	130,000	Pressure	100 H
35	- 03.30	Pittsburgh	30	0.1	1000	1	300	5	200	5	130,000	Pressure	201
36	- 04.55	Pittsburgh	100	2	1000	1	300	5	200	5	130,000	Pressure	201
37	- 05.30	Pittsburgh	100	2	1000	1	300	5	200	5	130,000	Flow	202 A
38	- 06.00	Pittsburgh	30	0.1	500	1	300	5	200	5	80,000	Flow	202 A
39	- 07.13	Pittsburgh	30	0.1	500	1	300	5	200	5	80,000	Flow	202 A
40	- 17.00	Pittsburgh	30	0.1	500	1	700	* 2 LR	500	* 0.5 LR	80,000	Flow	202 A

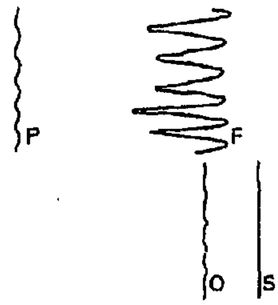
TABLE 11-6 (Con't)

Con No.	Day and Time of Setting Up This Condition	Coal Type	(PC) Gasifier Pressure		Flare Gas Flow (GFG)		Oxygen Flow		Steam Flow		Approx. Oxygen Blast	Gasifier Control Mode	Pressure Tapping Point
			PB	Reset	PB	Reset	PB	Reset	PB	Reset			
41	25.11.79 - 23.02	Pittsburgh	30	0.1	500	1	700	2	500	0.5	80,000	Pressure	202 A
42	- 23.25	Pittsburgh	30	0.1	500	1	700	2	500	0.5	80,000	Flow	202 A
43	26.11.79 - 01.12	Pittsburgh	30	1.0	500	1	700	2	500	0.5	160,000	Flow	202 A
44	- 01.58	Pittsburgh	30	0.1	500	1	700	2	500	0.5	160,000	Flow	202 A
45	- 05.21	Pittsburgh	30	0.1	500	1	700	2 LR*	500	0.5 LR*	50,000	Flow	202 A
46	- 17.36	Pittsburgh	30	0.2	500	1	700	2 LR*	500	0.5 LR*	50,000	Flow	202 A
47	- 21.33	Pittsburgh	30	0.5	500	1	700	2	500	0.5	50,000	Flow	202 A
48	- 23.35	Pittsburgh	30	0.5	500	1	700	2	500	0.5	160,000	Flow	202 A
49	27.11.79 - 01.40	Pittsburgh	30	0.1	500	1	700	2	500	0.5	110,000	Flow	202 A
50	- 11.30	Pittsburgh	30	0.1	500	1	700	2	500	0.5	110,000	Pressure	202 A

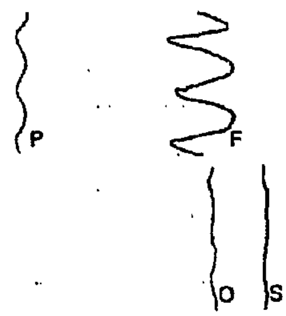
* *LR indicates steam oxygen flow signals being taken from the low range orifice instruments.



CONDITION 1 IN PRESSURE CONTROL. GAS FLOW IS VARIABLE DUE TO HIGH RESET VALUE ON P.C. THE PRESSURE IS BEING CONTROLLED QUITE WELL.

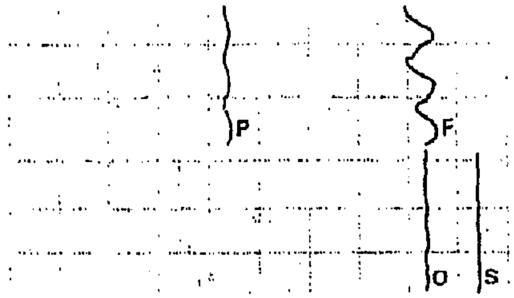


CONDITION 3 IN PRESSURE CONTROL. GAS FLOW PRESSURE AND OXYGEN FLOW ARE VARIABLE. THIS IS PROBABLY DUE TO THE REDUCED THROUGHPUT, THE INCREASED P.B. ON THE P.C. AND THE HIGH P.B. ON THE O.F.C.

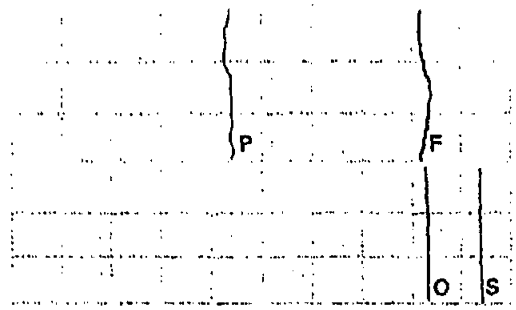


CONDITION 4 IN PRESSURE CONTROL. GAS FLOW, PRESSURE AND OXYGEN FLOW ARE VARIABLE. THIS IS PROBABLY DUE TO THE INCREASED P.B. AND INCREASED RESET ON THE P.C.

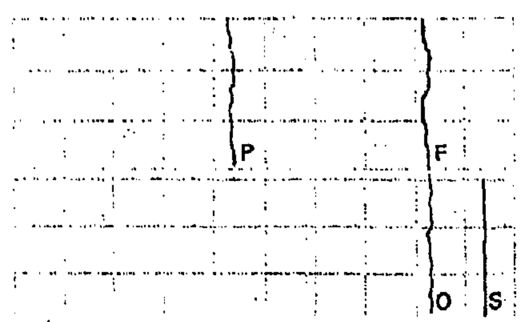
Figure 11-3. Recorder Charts Showing Flow and Pressure at Several Controller Settings



CONDITION 5 IN PRESSURE CONTROL. THE VARIABILITY HAS BEEN REDUCED BY SWITCHING TO THE PRESSURE TAPPING 201 WHICH IS CLOSE TO THE PRESSURE CONTROL VALVE.

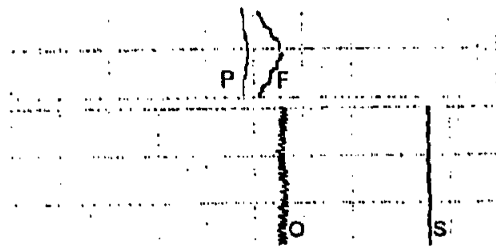


CONDITION 7 IN PRESSURE CONTROL. THE VARIABILITY HAS BEEN REDUCED AGAIN BY REDUCING THE RESET VALUE ON P.C. THE GAS FLOW IS NOW FAIRLY STEADY.

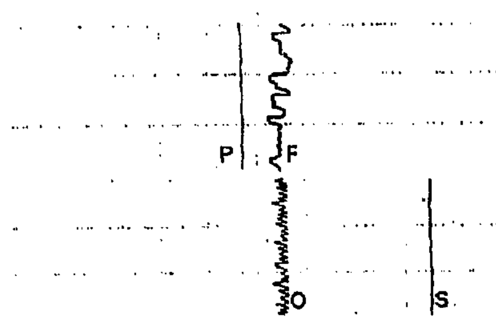


CONDITION 8 IN PRESSURE CONTROL. THIS SHOWS THAT CHANGING COALS DOES NOT MAKE ANY MAJOR DIFFERENCE TO THE CONTROL CHARACTERISTICS.

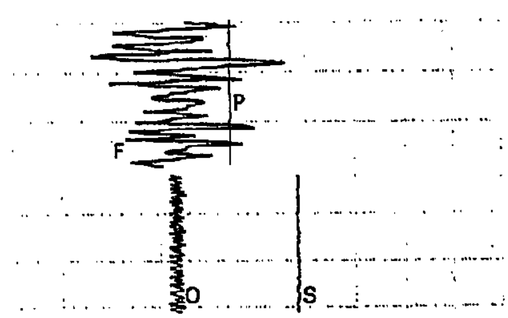
Figure 11-3 (Con't)



CONDITION 10 IN PRESSURE CONTROL. THE GAS FLOW HAS BECOME SLIGHTLY MORE VARIABLE AT THE HIGHER THROUGHPUT.

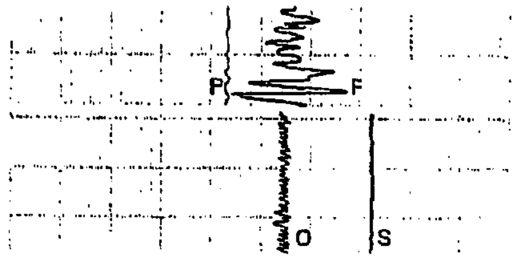


CONDITION 11 IN PRESSURE CONTROL. THE PRESSURE CONTROL HAS IMPROVED AND THE GAS FLOW IS STEADIER, DUE TO THE REDUCED PROPORTIONAL BAND AND REDUCED RESET ON THE P.C.

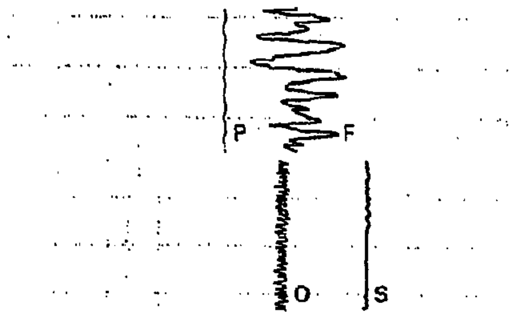


CONDITION 13 IN PRESSURE CONTROL. THE P.C. SETTINGS ARE THE SAME AS CONDITION 1, BUT THE P.B. AND RESET ARE REDUCED ON THE O.F.C. AND THE RESET IS REDUCED ON THE S.F.C.
THERE ARE AGAIN CONSIDERABLE VARIATIONS ON THE GAS FLOW. TAPPING POINT 202A IS IN USE.

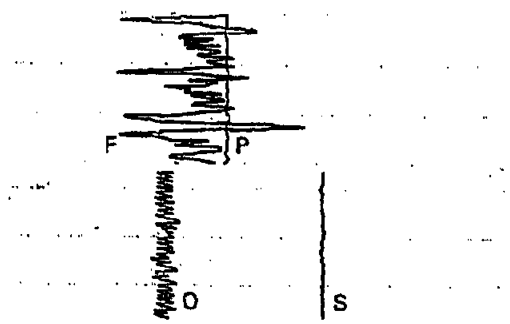
Figure 11-3 (Con't)



CONDITION 15 IN PRESSURE CONTROL. REDUCTION OF THE RESET ON THE P.C. HAS REDUCED THE VARIABILITY OF THE GAS FLOW SLIGHTLY.

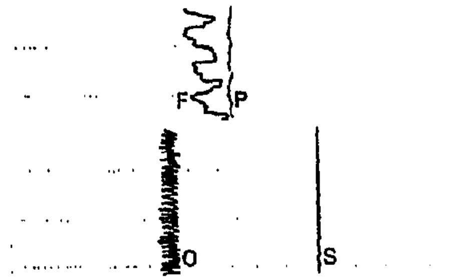


CONDITION 16 IN PRESSURE CONTROL. CHANGING TO PITTSBURGH COAL HAS AGAIN NOT MADE ANY MAJOR DIFFERENCE TO THE CONTROL CHARACTERISTICS. THERE IS A REDUCTION IN THE GAS MAKE AT THE SAME STEAM/ OXYGEN FLOW.

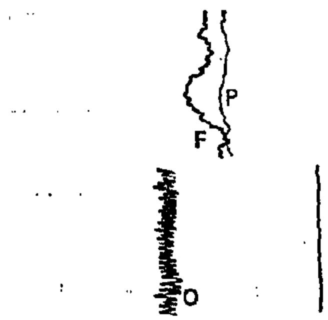


CONDITION 17 IN PRESSURE CONTROL. THE GAS FLOW IS STILL RATHER VARIABLE.

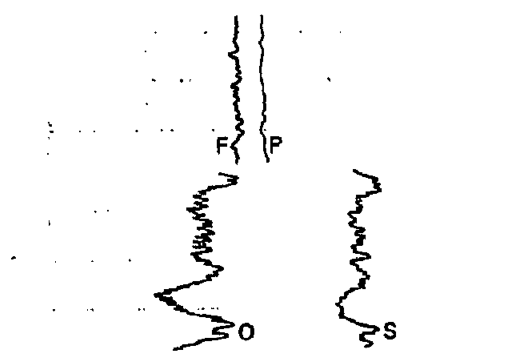
Figure 11-3 (Con't)



CONDITION 18 IN PRESSURE CONTROL. THE P.B. AND RESET ON THE P.C. HAVE BEEN INCREASED, WHICH HAS STEADIED OUT THE GAS FLOW. THE PRESSURE CONTROL HAS DETERIORATED BUT IS STILL ACCEPTABLE.



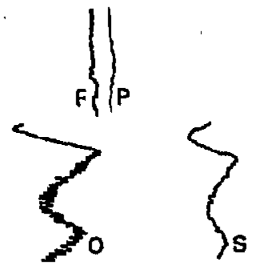
CONDITION 19 IN PRESSURE CONTROL. THE P.B. HAS BEEN INCREASED AND THE RESET DECREASED ON THE P.C. THIS HAS STEADIED OUT THE GAS FLOW EVEN MORE, BUT AGAIN THE PRESSURE IS SLIGHTLY MORE VARIABLE.



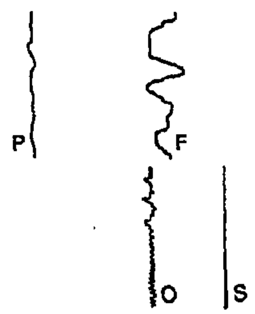
CONDITION 22 IN FLOW CONTROL. THE PRESSURE CONTROL IS REASONABLE. THE GAS FLOW AND THE STEAM/ OXYGEN FLOWS VARIABLE.

Figure 11-3 (Con't)

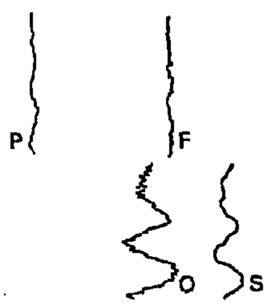
p



CONDITION 23 IN FLOW CONTROL. THE P.B. AND RESET HAVE BEEN INCREASED. ON THE P.C. THE PRESSURE CONTROL HAS THEREFORE DETERIORATED.

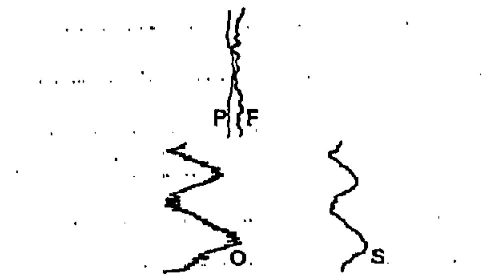


CONDITION 25 IN PRESSURE CONTROL.

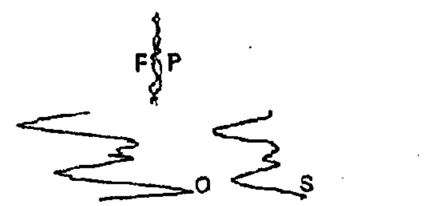


CONDITION 26 IN FLOW CONTROL. OXYGEN AND STEAM FLOW VARIABLE.

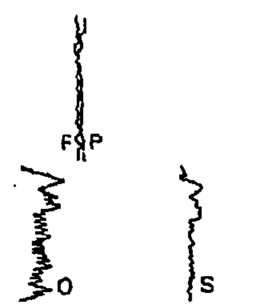
Figure 11-3 (Con't)



CONDITION 30 IN FLOW CONTROL. OXYGEN AND STEAM FLOW VARIABLE.



CONDITION 31 IN FLOW CONTROL. OXYGEN AND STEAM FLOWS MORE VARIABLE DUE TO THE REDUCTION OF THE P.B. ON THE P.C. BUT THE PRESSURE CONTROL HAS NOT IMPROVED.



CONDITION 32 IN FLOW CONTROL. THE OXYGEN AND STEAM FLOWS HAVE STABILISED AND THE PRESSURE HAS BECOME MUCH MORE STABLE. THIS WAS DUE TO THE REDUCTION IN THE RESET VALVE ON THE P.C. FROM 2 TO 0.1.

Figure 11-3 (Con't)

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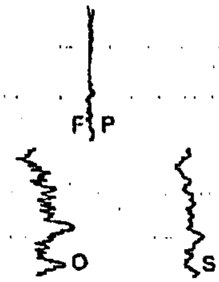
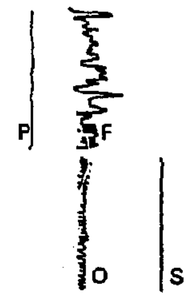
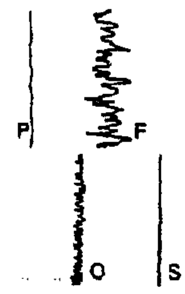


CONDITION 33 IN FLOW CONTROL. THIS CONDITION IS VERY SIMILAR TO CONDITION 32. THE SLIGHT REDUCTION OF THE P.B. ON THE P.C. SHOULD HAVE IMPROVED THE PRESSURE STABILITY SLIGHTLY, BUT DOES NOT APPEAR TO HAVE MADE MUCH DIFFERENCE.

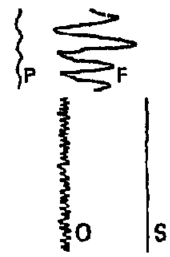


CONDITION 34 IN PRESSURE CONTROL. THIS GIVES VERY GOOD PRESSURE CONTROL WITHOUT EXCESSIVE SWINGS ON THE GAS FLOW RATES, I.E. A LOW P.B. WITH A LOW RESET VALUE ON THE P.C. GIVES THE BEST CONTROL.

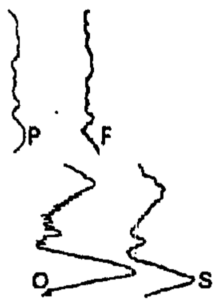


CONDITION 35 IN PRESSURE CONTROL. AGAIN A CHANGE OF COAL TO PITTSBURGH MAKES LITTLE DIFFERENCE, EXCEPT FOR A REDUCTION IN THE GAS MAKE.

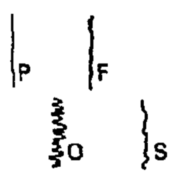
Figure 11-3 (Con't)



CONDITION 35 IN PRESSURE CONTROL. INCREASING THE P.B. AND RESET ON THE P.C. CAUSES A DETERIORATION OF THE PRESSURE CONTROL PLUS MORE VARIATION OF THE GAS FLOW.

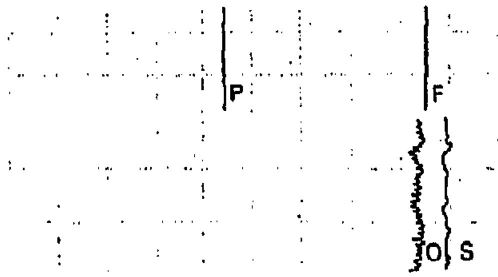


CONDITION 37 IN FLOW CONTROL. THE PRESSURE IS RATHER VARIABLE AS ARE THE OXYGEN AND STEAM FLOWS.

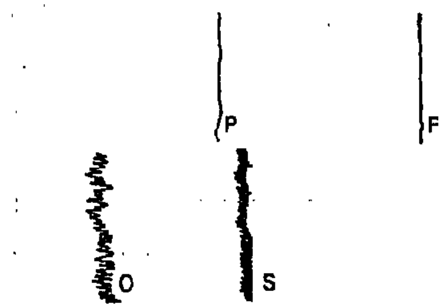


CONDITION 38 IN FLOW CONTROL. WITH P.C. SETTINGS SIMILAR TO CONDITION 35 EVERYTHING WAS FAIRLY STABLE AND THE PRESSURE CONTROL GOOD. (THIS WAS UNFORTUNATELY A RATHER SHORT PERIOD, THEREFORE THIS RESULT NEEDS CONFIRMATION.)

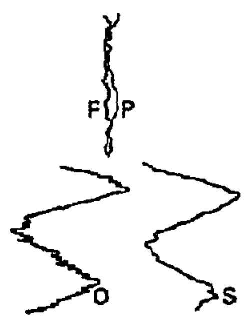
Figure 11-3 (Con't)



CONDITION 39 IN FLOW CONTROL. AGAIN WITH P.C. SETTINGS SIMILAR TO CONDITION 35 EVERYTHING WAS FAIRLY STABLE AND THE PRESSURE CONTROL GOOD. THE GASIFIER AT HALF THE STANDARD LOAD.



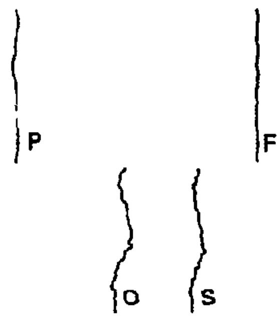
CONDITION 40 IN FLOW CONTROL. USING LOW RANGE ORIFICE SIGNALS, THE PROPORTIONAL BANDS HAD TO BE INCREASED AND RESETS DECREASED ON THE OFC AND SFC BECAUSE OF THE RANGE CHANGE.



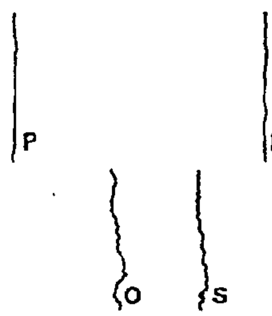
CONDITION 43 IN FLOW CONTROL. INCREASING THE RESET VALUE TO ONE ON THE P.C. CAUSED POOR PRESSURE CONTROL AND VARIABLE STEAM/OXYGEN FLOWS.

Figure 11-3 (Con't)

CONDITION 44 IN FLOW CONTROL. REDUCING THE RESET VALUE BACK TO 0.1 ON THE P.C. IMPROVED THE PRESSURE CONTROL AND REMOVED THE WIDE SWINGS ON THE STEAM AND OXYGEN FLOWS.

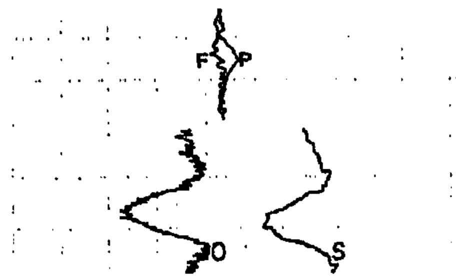


CONDITION 45 IN FLOW CONTROL. CONTROL WAS QUITE GOOD.



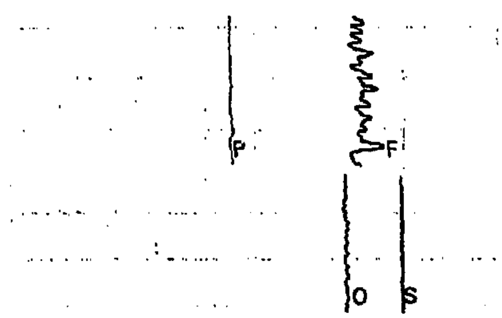
CONDITION 46 IN FLOW CONTROL. USING LOW RANGE ORIFICE SIGNALS. INCREASING THE RESET ON THE P.C. TO 0.2 MADE LITTLE DIFFERENCE.

Figure 11-3 (Con't)



CONDITION 48 IN FLOW CONTROL. INCREASING THE RESET ON THE P.C. TO 0.5 CAUSED THE PRESSURE TO BECOME VARIABLE AND THE STEAM AND OXYGEN FLOWS TO START SWINGING.

CONDITION 49 IN FLOW CONTROL. WITH THE PRESSURE CONTROLLER SETTINGS SIMILAR TO CONDITION 35, 38 AND 39, THE PRESSURE CONTROL WAS STEADY AND THE OXYGEN AND STEAM FLOWS NOT VARYING WILDLY.



CONDITION 50 IN PRESSURE CONTROL. THE PRESSURE CONTROL IS GOOD WITHOUT WILD SWINGS ON THE GAS FLOW.

Figure 11-3 (Con't)

Figure 11-3 consists of a number of chart sections showing the gas flow and gasifier pressure on the upper chart and the oxygen and steam flow on the lower chart. The flow is marked F, the Pressure P, the oxygen flow O and the steam flow S. A few conditions given in Table 11-6 are not shown in Figure 11-3 because either the period was too short or load changes were being carried out.

It may seem from conditions 35, 38, 39 and 50 that the best pressure control settings for the four inch control valve are 30% PB and 0.1 reset. These were the lowest values evaluated, therefore the optimum settings may be even lower. The smaller valve gives much better control than the six inch valve used on run - 01 which suffered from instability once the PB was reduced below 40%.

The steady state operation of the gasifier whether in pressure or flow control was at its best when the pressure controller settings were at the minimum values evaluated which were 30% PB and 0.1 reset in combination with the 4" pressure control valve. The optimum settings may be even lower.

The flare gas, steam and oxygen flow controllers were set up with wide proportional bands which gave acceptable control characteristics. An improvement will be achieved in the case of the steam/oxygen flows by commissioning the vortex flow meters so that one instrument range can be used to give the full turn down. The proportional band of the gas flow controller should be reduced during the next run such that more rapid changes of flow can be achieved in automatic control.

BEHAVIOR OF GASIFIER DURING TRANSIENTS

During run EPRI - 02, various programmed load changes were carried out as detailed on Table 11-1. No problems were encountered during the load changes.

A qualitative picture of the sharp spikes 1 and 2 down to 110,000 SCF/H from 160,000 SCF/H oxygen blast is given in Figures 11-4, 11-5, 11-6, 11-7, 11-8, 11-9, 11-10 and 11-11 for the two separate spikes down in rate. There was no significant change to the make gas during load changes except for the slightly greater contribution of the nitrogen input from the slag systems and purges did not vary with throughput. There was some modification of the gas composition approximately 15 minutes after Spike 1.

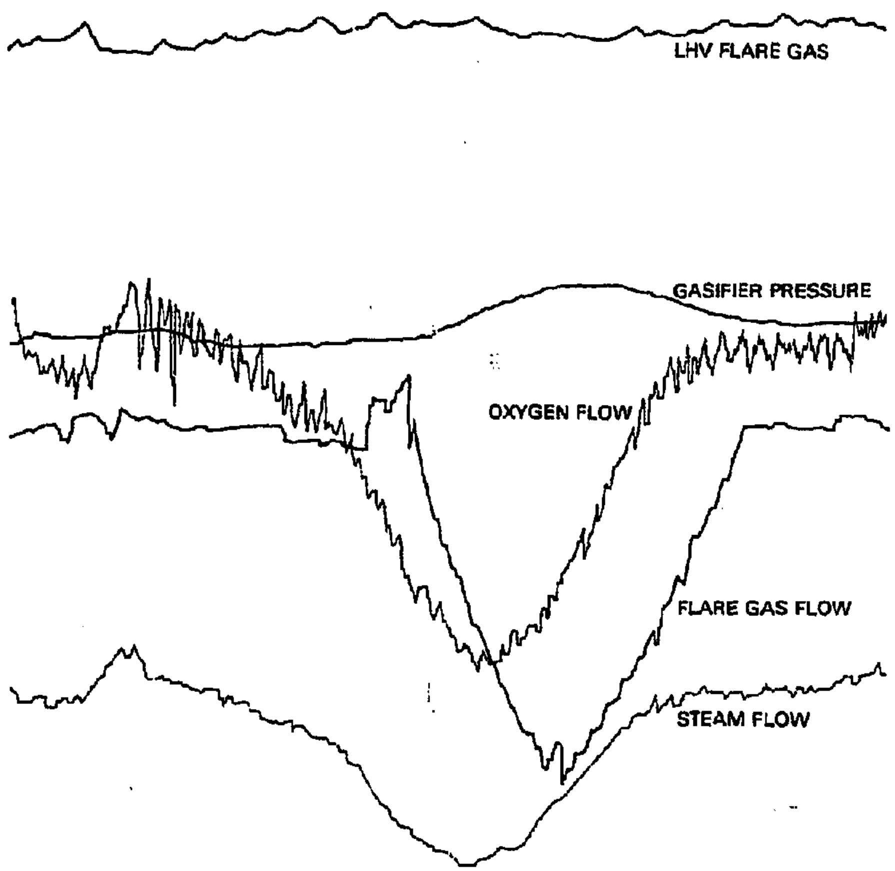


Figure 11-4. Major Gasifier Parameters Across Spike
EPRI - 02
Chart Speed: 2 Min./CM

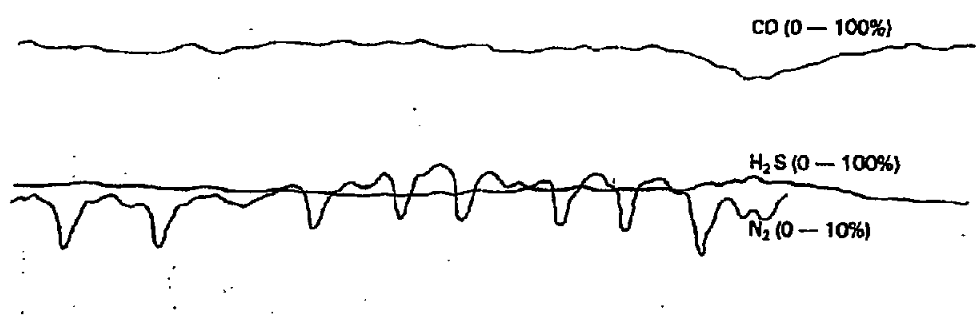


Figure 11-5. Offtake Gas Analysis Post Load Reduction Spike
EPRI - 02
Chart Speed: 2 Min./CM

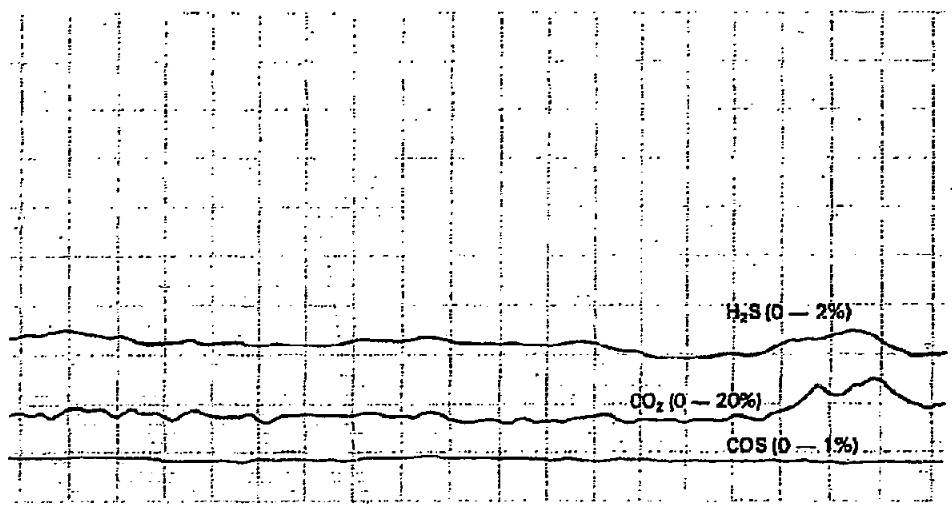


Figure 11-6. Offtake Gas Analysis Post Load Reduction Spike
EPRI - 02
Chart Speed: 2 Min./CM

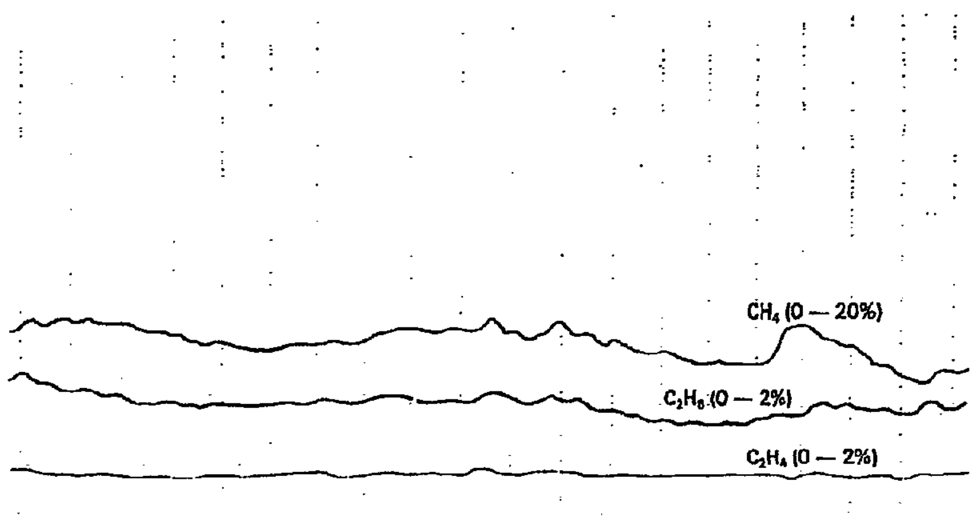


Figure 11-7. Offtake Gas Analysis Post Load Reduction Spike
EPRI - 02
Chart Speed: 2 Min./CM

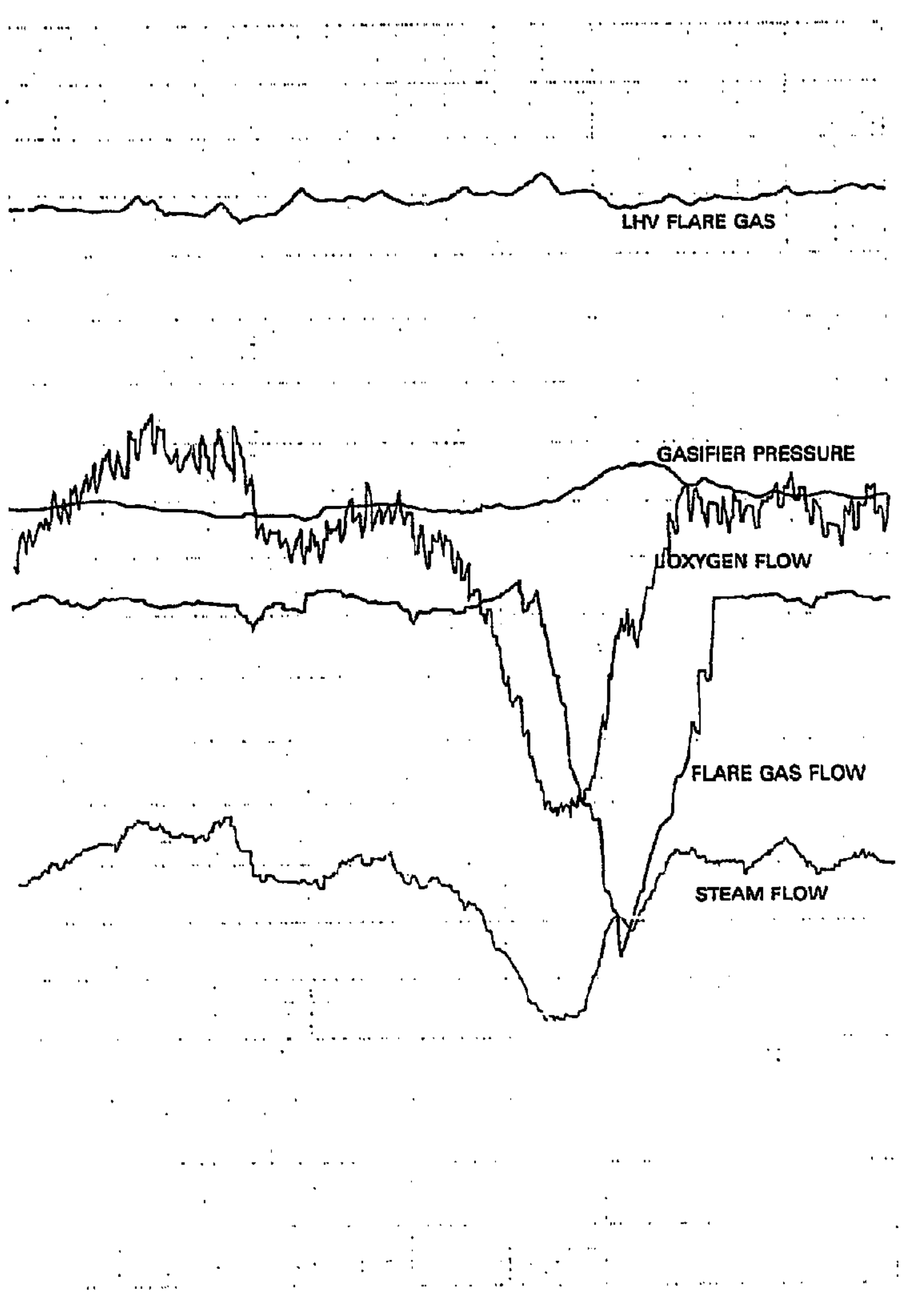


Figure 11-8. Major Gasifier Parameters During Spike Down in Rates
EPRI - 02
Chart Speed: 2 Min./CM

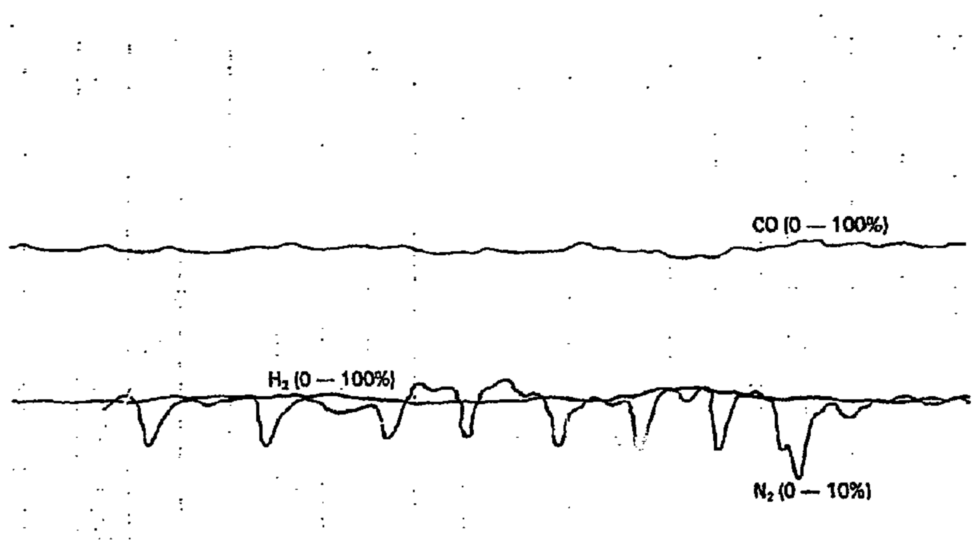


Figure 11-9. Offtake Gas Composition Post Spike in Rates
EPRI - 02
Chart Speed: 2 Min./CM

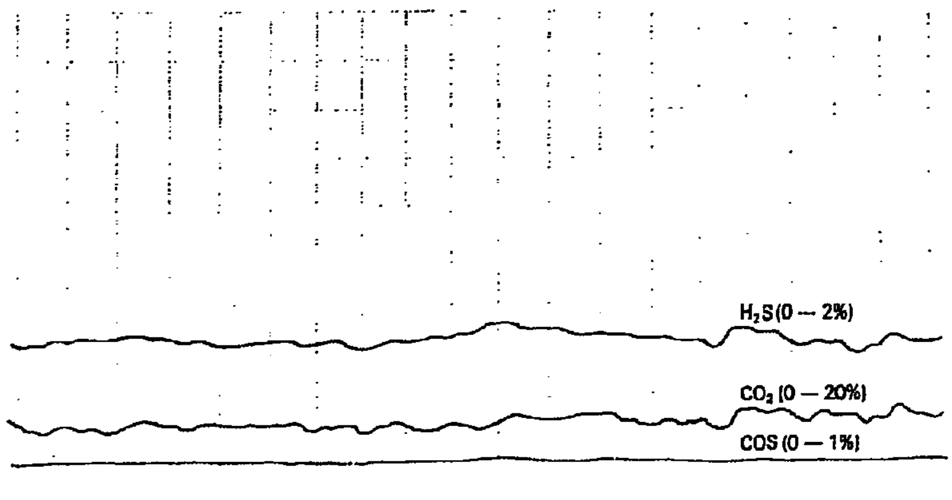


Figure 11-10. Offtake Gas Analysis Post Spike Rates
EPRI - 02
Chart Speed: 2 Min./CM

Figure 11-11. Offtake Gas Composition Post Spike Down in Rates
EPRI - 02
Chart Speed: 2 Min./CM

p

POST RUN INSPECTION

The gasifier was cooled down after EPRI - 02A but close inspection of the fuel bed and gasifier internals were not carried out. The necessary repairs were carried out before restarting the run as EPRI - 02B.

EPRI - 02B was shut down with the distributor almost empty of fuel. The coal was strongly caked in the region of the upper part of the stirrer, but there was transition to a good size distribution char by the time the bottom of the stirrer was reached. At the shaft top, there was a mass of what appeared to be partially carbonized fines/tar mixture.

The bed below the stirrer was full of good char but in the shaft centre it appeared to have a greater tendency to be stuck together than at the walls, perhaps indicating a more rapid flow of fuel down the centre of the shaft.

During EPRI - 02B the gasifier was put on to cold standby at 10:11 on 24 November due to a problem in the quench chamber. With the gasifier on a cold standby the quench chamber was opened. The problem was cured and the run restarted.

The major cause of the problem was due to an instrument error. Post standby, the conditions were corrected and there were no further problems.

CONCLUSIONS

Run EPRI - 02B successfully demonstrated that the gasifier can work well on Pittsburgh 8 coal and can respond favourably to rapid load changes. At standard load bed behaviour was less stable than on Rossington coal, with greater offtake temperature swings, but sustained running was always possible.

Shut down of EPRI - 02A was caused by a leak at a tuyere flange. This type of problem can be readily avoided. The standby of EPRI - 02B was a convincing demonstration of the gasifier's ability to come back, rapidly and reliably, on line after an upset.

It is clear that the Slagging Gasifier offers a reliable source of intermediate BTU gas for combined cycle power generation, and that changes in power demand can be rapidly accommodated by the fixed bed system, which is not upset by major changes and maintains a reliable supply of gas at a fixed CV. Tight gas flow control at the back end of the plant will lead to slight fluctuations in gasifier pressure and gasification medium flow, but these were shown to have no detrimental effect upon gasifier performance.

Section 12

REPORT ON TEST RUN EPRI - 03

SUMMARY

Run EPRI - 03 aimed to consolidate the results of EPRI - 02, which had been a successful gasifier operation on screened Pittsburgh 8 coal, and to carry out the three additional objectives:

- Running the Slagging Gasifier on Pittsburgh 8 coal with tar injection to the tuyeres.
- Running on Pittsburgh 8 coal as delivered, which contained about 25% material less than ¼", instead of on the screened fuel.
- Attempting to flux Pittsburgh 8 coal with limestone instead of BFS.

Run EPRI - 03 was planned to further investigate the behaviour of Pittsburgh 8 coal in the Slagging Gasifier and started on 13 December, 1979, with standard conditions being established on Rossington coal. Screened Pittsburgh 8 coal, fluxed with BFS was introduced at 19:18 at 130,000 SCF/H oxygen loading, 1.30 H_2O/O_2 and a performance test run under these conditions. Gasifier conditions were satisfactory and tar injection down tuyeres was brought on at 00:12 on 14 December, with satisfactory performance of the system. A performance test was done under these conditions and at 23:05 the tar rate to the tuyeres was increased and a further performance test run.

Satisfactory completion of the above major objective of the run enabled the second objective to be started, with as received Pittsburgh 8 coal, containing about 20% material less than ¼" being charged to the gasifier at 14:25 on 15 December. Performance on this fuel was satisfactory and at 04:23 on 16 December the rate was reduced from 130,000 SCF/H oxygen to 80,000 SCF/H, followed at 11:28 by a further reduction to 50,000 SCF/H. Both the step downs and subsequent running were satisfactory and at 18:34 the rate was returned 130,000 SCF/H oxygen. A spike down to 50,000 SCF/H loading over ten minutes was performed. A further run objective of fluxing Pittsburgh 8 coal with limestone was then started. Limestone was charged early on 17 December and no problems were encountered with this method of fluxing, and at 14:07 gasifier load was raised to 160,000 SCF/H oxygen with test periods at three different stirrer revolutions being initiated.

At 04:30 on 16 December the load was reduced rapidly from 160,000 SCF/H oxygen to 50,000 SCF/H and stabilised; it was then brought back to full load at 08:30. A further load spike down to 110,000 SCF/H was performed at 10:15. None of these rate changes caused any problems, and the gasifier was subjected to a controlled shutdown at 10:43 after 118 hours of continuous running.

Post run inspection revealed good conditions inside the gasifier. The schematic of EPRI - 03 is given in Figure 12-1. Table 12-1 shows the load changes carried out during the run.

EQUIPMENT, INSTRUMENTS AND CONTROLS

The gasifier system was modified to provide tar injection to tuyeres with the tar being pumped into the raceway.

Other gasifier systems were unchanged for the run except for continuing improvements in data acquisition and logging systems.

RUN DIARY

Preparations for EPRI - 03 were completed on 13 December and at 10:51 the run was initiated.

At 13:03 the steam/oxygen was brought on at startup rates. Rossington singles charging commenced, fluxed with limestone.

The startup was good and the gasifier was established at standard conditions of 160,000 SCF/H oxygen loading, with steam/oxygen 1.30 and pressure 335 psig. Control was initially in pressure control at number 4 flare, with the pressure being held slightly low, at 335 psig, in anticipation of the flow mode of control being engaged once satisfactory running was established.

Good conditions were obtained at the standard loadings with steady bed conditions and good slag tapping. At 17:09 the rate was reduced to 130,000 SCF/H oxygen over five minutes in preparation for accepting Pittsburgh 8 coal. This change went very smoothly with no discernable transients due to the load variation, and at 18:30 the gasifier was put in back end flow-front end pressure control.

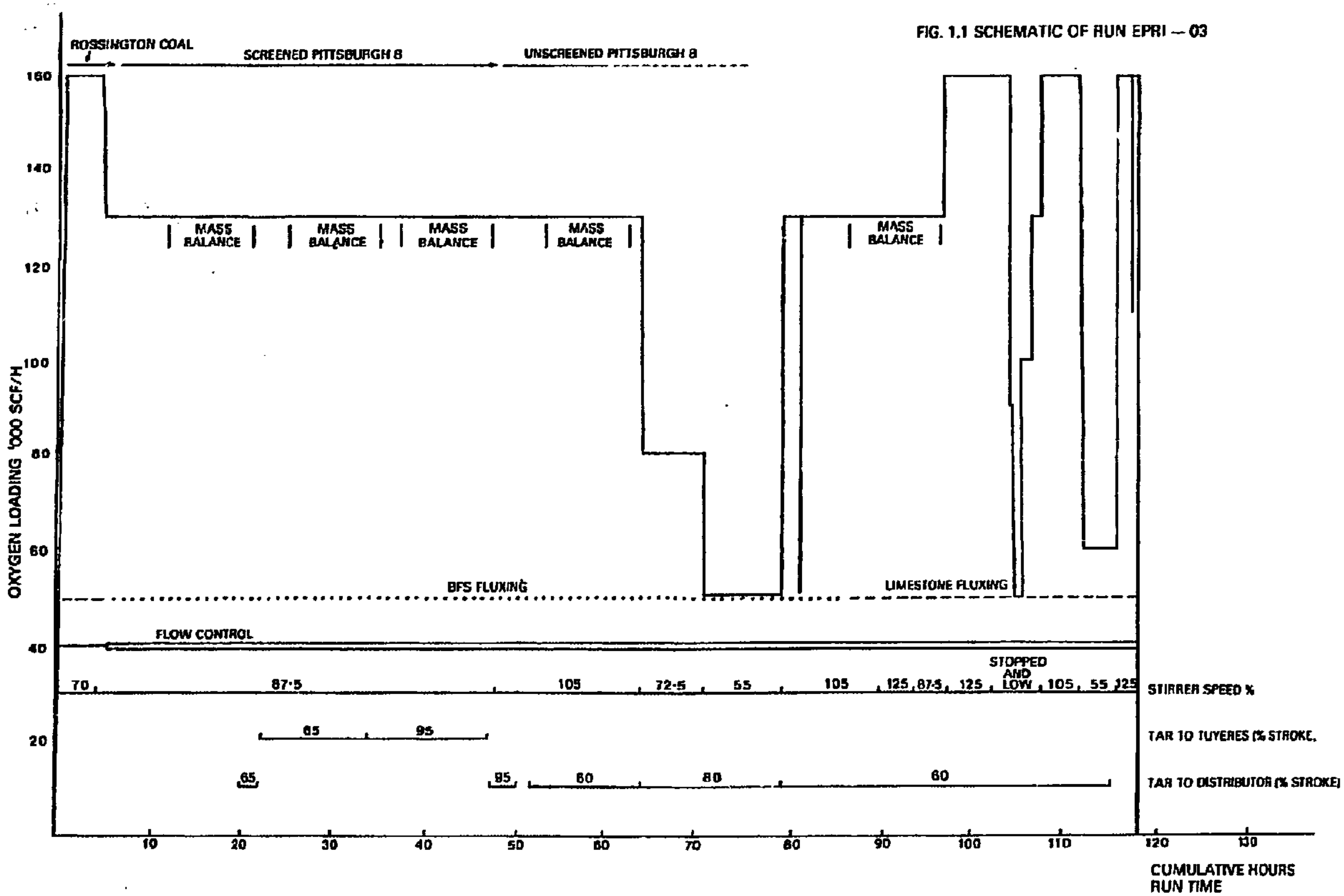


Figure 12-1. Schematic of EPRI - 03

TABLE 12.1 DESCRIBING LOAD CHANGE PERIODS EPRI - 03

Change No.	Load 1 before Change (1000 SCFH O_2 blast)	Load 2 after Change (1000 SCFH O_2 blast)	Time Taken (mins)	Change in 1000 SCFH	Coal Gasified	Control Mode
1	160	130	5	30	Rossington	P
2	130	80	4	50	Pitts. 8	F
3	80	50	11	30	Pitts. 8	F
4	50	130	3	80	Pitts. 8	F
5	130	50	5	80	Pitts. 8	F
6	50	130	3	80	Pitts. 8	F
7	130	160	6	30	Pitts. 8	F
8	160	90	6	70	Pitts. 8	F
9	(90	140	8	50	Pitts. 8	F
	(140	160	5	20	Pitts. 8	F
10	(160	90	7	70	Pitts. 8	F
	(90	50	8	40	Pitts. 8	F
11	50	90	10	40	Pitts. 8	F
12	90	130	12½	40	Pitts. 8	F
13	130	160	8	30	Pitts. 8	F
14	160	50	5	110	Pitts. 8	F
15	50	160	25	110	Pitts. 8	F
16	160	110	7	50	Pitts. 8	F
17	110	160	5	50	Pitts. 8	F

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There was a slight problem with flux flow, due to a blockage, and this was rectified; at 19:08 the first lock of Pittsburgh 8 coal was charged to the gasifier, with fluxing being switched to BFS. A good transition to this fuel was obtained with no bed problems.

With gasifier performance good, a mass balance period was started at the beginning of 14 December, with the side stream sampling system being put on. No tar was injected either to the gasifier top or down the tuyeres and gasifier performance was satisfactory in all areas. At 10:08, with the performance test on screened Pittsburgh 8 at 130,000 SCF/H oxygen loading completed, tar injection was put on to the gasifier top and this was switched to down tuyeres at 65% pump stroke, at 11:12. The system was successfully commissioned.

By 13:00 it was apparent that the estimated 1,000 lbs/hour of tar being injected into the gasifier raceway was having a significant effect upon gasifier performance. Coal locking frequency was down slightly, the methane content of the make gas had dropped by about 10% of the previous value and the average offtake temperature had risen appreciably. A performance test under these conditions was started.

Gasifier behaviour continued to be satisfactory. At 23:05 tar injection to the combustion zone was increased to 95% pump stroke, with no problems arising. A performance test period under these conditions was started at 01:00 on 15 December. The indications from average offtake temperature and methane concentration in the make gas suggested that not much more tar was being put into the raceway at 95% pump stroke as compared to 65% pump stroke, although there was a slight deterioration in slag tapping.

At 11:00 hours the performance test period was completed and tar injection was switched to the gasifier top with a small steam purge being left on the tuyere tar injection system. Preparations were being made for switching the screened Pittsburgh 8 feed to an unscreened fuel containing about 25% material less than ¼". By 14:25 unscreened Pittsburgh 8 coal was entering the gasifier, with tar injection turned off during this transition period, which produced little noticeable effect, although the bed conditions were slightly more ragged as compared to the screened fuel case.

Tar injection to the gasifier top was restored at 60% pump stroke at 16:15 and a performance test period was started at 17:00 on a coal with a fines content running at about 20% less than ¼". Gasifier performance under these conditions continued to be satisfactory into 16 December and at 00:30 hours coal locking, which had been on nitrogen, was switched to raw gas. With the gasifier in flow control, this depletion of gas down stream at intervals was easily accounted for by the control system, with no upsets being caused, and the gasifier was left in this mode for the rest of the run.

At 04:18 the gasifier rates were lowered to 80,000 SCF/H, from 130,000 SCF/H oxygen, across five minutes, the operation being carried out smoothly in flow control with no transient upsets occuring at the gasifier. Running at half loading proved to be satisfactory, and at 11:20 a further reduction in rates to 50,000 SCF/H was carried out successfully. Running at this loading was satisfactory. At 18:31, with running steady at 50,000 SCF/H oxygen loading, the rates were brought up to 130,000 SCF/H oxygen over three minutes.

With steady running established at these loadings on the unscreened coal, the gasifier load was spiked to 50,000 SCF/H oxygen and back again over ten minutes. This provided some noticeable transients in the gasifier.

Running continued into 17 December with these conditions and satisfactory performance was obtained. At 02:00 fluxing was changed over from BFS to limestone with the latter being charged at a flux/ash ratio of about 1.0. The changeover was accomplished without any problems with slag tapping remaining good and all tuyeres remaining bright, and a performance test period was started at 03:00. As part of this test period, stirrer revolutions were to be varied.

At 14:07 the testing at 130,000 SCF/H oxygen loading was completed and the rate was brought up to 160,000 SCF/H across three minutes, with the distributor revolutions then being brought up more slowly to 125%. At 21:21 rates were cut to 50,000 SCF/H and at 23:47 rates were increased to 130,000 SCF/H.

Four hours running were obtained at this loading without any incidents. At 04:30 the gasifier load was dropped very sharply down to 50,000 SCF/H over about two minutes and was held at this load until 08:15, when it was restored again to 160,000 SCF/H oxygen, this time over 25 minutes. Both the above load changes caused no gasifier upsets and generated no obvious transients.

With the loading established at 160,000 SCF/H oxygen, a sharp spike down to 110,000 SCF/H was performed and the load held briefly at the latter value for three minutes, before being rapidly brought back to full load. This load spike resulted in no gasifier problems and after re-stabilising at full load, preparations were made to carry out shutdown. This was carried out at 10:43 and was followed by standard cool down with nitrogen. Run EPRI - 03 lasted 118 hours. A schematic of the run is shown in Figure 12-1, and a run dairy is given in Table 12-1.

SLAG TAPPING AND SLAG REMOVAL

Slag tapping and slag removal was good throuout run EPRI - 03, with good performance being obtained across load changes and changes in fluxing agent. At the higher rate of tar injection to the tuyeres there was a subjective impression that slag tapping deteriorated, although any effect was small.

The transition from screened Pittsburgh 8 coal to the unscreened feed (containing 25% material less than ¼") had no effect upon slag tapping.

Slag quenching was good with dense black frit always being produced.

FLUXING SYSTEMS

Fluxing at the start of EPRI - 03 was set up to be on limestone fluxing while Rossington coal was being charged. In the initial stages a lump of concrete partially blocked the vibrator gate, resulting in underfluxing Rossington. This fault was soon corrected, and when the transition to Pittsburgh 8 coal was made the fluxing was switched to BFS from the South bunker.

Fluxing on BFS was at a rate giving a flux/ash ratio of about 1.6. Fluxing rate on BFS remained the same for the unscreened coal (which had a very similar ash content) and a good free flowing slag was obtained. The fluxing rate may have been slightly high but no attempt was made to optimise the flux/ash ratio.

During the later stages of the run, when operating on unscreened Pittsburgh 8, fluxing was switched to limestone, giving a flux/ash ratio of 1.20. Again, a free flowing slag was obtained, and again, no attempt was made to optimise the amount of flux required.

The flux balances across run EPRI - 03 are presented in Table 12-3. On BFS there was good agreement between the experimental and calculated ash composition. A good free flowing slag was obtained. Balance agreements with limestone fluxing were not as good, elements were out of balance, indicating that the hearth had not come to equilibrium on limestone fluxing following the long period on BFS fluxing. The main difference may be due to analytical error.

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TABLE 12-2. SIMPLIFIED RUN DIARY OF EPRI - 03

Date	Time	Period No.	Comments	Data Logger Period
13 Dec 79	10.51		Start up commenced.	
	13.03		Steam/oxygen down tuyeres.	
	13.45	1	Rates 160,000 $SCFH^{-1}$ oxygen. Steam/oxygen 1.3/1. Pressure 335 psig. In pressure control. On high range transmitters.	A
	17.09		Reducing rates.	
	17.15		Rates now 130,000 $SCFH^{-1}$ oxygen.	
	17.25		Distributor to 87.5%.	
	18.30		Into flow control.	
	19.08	2	Pittsburgh in with BFS.	
14 Dec 79	00.00	2	Start of Mass Balance 1	
	10.08		End of Mass Balance 1.	B
			Tar injection on to distributor - 60%	
	11.12		Tar injection on to tuyeres - 65%.	
	13.00	3	Start of Mass Balance 2.	
	19.30	3	Steam/oxygen ratio now 1.2/1	
	23.00	3	End of Mass Balance 2	C
	23.05		Tar injection set at 95%.	

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Date	Time	Period No.	Comments	Data Logger Period
15 Dec 79	01.00	4	Start of Mass Balance 3.	
	11.00	4	End of Mass Balance 3	D
	11.28		Tar injection back on to distributor	
	12.35		R.O.M. Pittsburgh starting.	
	14.30		Tar injection off.	
	16.15		Tar injection on at 60% to distributor.	
	17.00	5	Start of Mass Balance 4.	
16 Dec 79	03.00	5	End of Mass Balance 4.	E
	03.01		Tar injection up to 80% to distributor.	
	04.18		Dropping rates to 80,000 $SCFH^{-1}$ oxygen.	
			Tar injection down to 50%.	
	05.01	6	On low range transmitters.	
	09.00	6	Tar injection to 80% pump stroke.	F
	11.20	6	Starting to reduce rates.	
	11.28	7	Rates now 50,000 $SCFH^{-1}$ oxygen.	G
			Steam/oxygen ratio now on 1.3/1	
	14.26	7	Steam/oxygen ratio to 1.1/1.	

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Date	Time	Period No.	Comments	Data Logger Period
	17.05	7	Tar injection down to 60%.	I
	17.54	7	On high range transmitters	
	18.31		Starting to increase rates.	
	18.35		Rates now 130,000 $SCFH^{-1}$ oxygen.	
	20.04		Spike down to 50,000 $SCFH^{-1}$ oxygen.	
	20.14		Spike up to 130,000 $SCFH^{-1}$ oxygen.	
	21.30		Tuyere shut off. Rates still 130,000 $SCFH^{-1}$ oxygen. Steam/oxygen 1.2/1.	
17 Dec 79	01.53		Pittsburgh and limestone.	J
	03.00	8	Start of Mass Balance 5.	
	08.15	8	Steam/oxygen ratio 1.3/1	L
	14.00	8	End of Mass Balance 5.	M
	14.05		Starting to increase rates.	
	14.10	9	Rates now 160,000 $SCFH^{-1}$ oxygen.	
	20.07		Using Vortex meters on flow controller.	
	20.42		Back on orifice meters.	
	21.21		Rates cut to 50,000 $SCFH^{-1}$ oxygen.	

Date	Time	Period No.	Comments	Logger Period
	23.47		Starting to increase rate.	
	23.55		Rates now 130,000 $SCFH^{-1}$ oxygen. Distributor at 85%.	
18 Dec 79	00.22		Increasing rates.	
	00.32	10	Rates now 160,000 $SCFH^{-1}$ oxygen.	
	04.30	10	Reducing rates.	O
	04.35		Rates cut to 50,000 $SCFH^{-1}$ oxygen.	
	07.03		Tar injection off.	
	07.59		Tar injection on.	P
	08.15		Starting to increase rates and distribution speed.	
	08.22		Tar injection off.	
	08.40	10	Rates at 160,000 $SCFH^{-1}$ oxygen.	Q
	10.15		Starting to spike down to 110,000.	
	10.26		Rates back to 160,000 $SCFH^{-1}$ oxygen.	
	10.43		Shut down.	

TABLE 12-3. FLUX BALANCES FOR EPRI - 03

Period	B		C		D		E		F		H I		J		K L M		N		O	
Flux rate per Lock	1327 lbs.		1327 lbs		1327 lbs		1327 lbs		1327 lbs		1327 lbs		1327 lbs		916 lbs		916 lbs		916 lbs	
Flux. Ash	1.50		1.58		1.77		1.59		1.58		1.58		1.65		1.21		1.18		1.20	
	Actual	Calc.	Actual	Calc.	Actual	Calc	Actual	Calc	Actual	Calc	Actual	Calc	Actual	Calc	Actual	Calc	Actual	Calc	Actual	Calc.
SiO_2	40.18	40.76	39.88	40.56	40.01	40.15	40.64	40.57	40.40	40.60	40.01	40.60	40.01	40.44	34.94	31.74	34.94	32.01	34.94	31.83
Al_2O_3	17.64	16.67	17.01	16.52	17.59	16.20	17.54	16.66	17.93	16.68	17.71	16.68	17.71	16.55	17.44	15.22	17.44	15.35	17.44	15.26
Fe O	4.85	+5.32	5.16	+5.17	4.97	+4.80	4.44	+4.82	4.13	+4.83	2.87	+4.83	2.87	+4.72	4.86	+7.19	4.86	+7.25	4.86	7.21
CaO	25.63	23.62	24.83	24.05	25.17	24.98	25.80	24.12	25.66	24.01	24.85	24.07	24.85	24.43	35.74	36.21	35.74	35.72	35.74	36.05
MgO	7.10	6.27	7.13	6.39	6.92	6.63	7.12	6.40	6.80	6.38	6.77	6.38	6.77	6.48	1.95	2.12	1.95	2.10	1.95	2.11
Na_2O	0.48	0.39	0.44	0.39	0.45	0.39	0.43	0.39	0.38	0.39	0.38	0.39	0.38	0.39	0.27	0.30	0.27	0.30	0.27	0.30
K_2O	1.16	1.15	1.13	1.13	1.15	1.08	1.18	1.14	1.16	1.14	1.15	1.14	1.15	1.13	1.36	1.37	1.36	1.38	1.36	1.37
Silica Ratio	50.60	52.59	50.78	52.21	50.86	51.42	51.05	52.30	51.45	52.45	52.62	52.45	52.62	52.12	44.16	40.21	44.16	40.64	44.16	40.36

←— Pittsburgh 8 and B.F.S —→ (B–D) R.O.M. Pittsburgh 8 and B.F.S. (E–F) ROM Pittsburgh 8 and B.F.S (H I–J) ←— ROM Pittsburgh 8 and Limestone —→ (K L M–O)

+ Iron Separation 20% throughput.

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BED BEHAVIOUR

Bed behaviour on Pittsburgh 8 coal during EPRI - 03 was generally less steady than that expected on Rossington coal during the test periods of significant duration. As with EPRI - 02B, many of the test periods were short, although it was apparent that there was a trend to greater bed unsteadiness at the higher gasifier loadings. However, at all loadings no significant operational problems were encountered, with the gasifer capable of running under sustained loading at all conditions.

Table 12-4 gives details of the steady run periods analysed for EPRI - 03. An examination of the charts for offtake temperatures, and offtake CO_2 analysis shows that periods C, D and E exhibit more instabilities, which correlates with the observations on offtake temperatures (Figure 12-2). Figure 12-3 compares the charts obtained for period B (reasonably well behaved) and period C (less steady). These charts indicate that tar injection to the tuyeres results in a less steady bed performance, although it should be emphasised that bed behaviour during this latter period was steady and allowed sustained running with no operational problems.

There was a trend to higher offtake temperature with increased oxygen loading.

The behaviour of the screened and unscreened Pittsburgh 8 coal in the gasifier cannot be directly compared as no two substantial run periods differed only in this factor. However, there is little evidence to suggest that there is any difference in performance between the two feeds. Thus, comparison of B, C, D periods with E and M reveal no significant effects of coal feed.

A similar situation arises when a comparison between the gasifier running on BFS and on limestone fluxing is attempted. At the changeover to limestone, at the end of period J into period K, the gasifier was in flow control, and steadier behaviour on limestone fluxing was apparent, as evidenced by less cycling of input/steam flows (which follow offtake swings caused by bed instabilities).

Comparison of behaviour in periods B, C, D and E with that of K, L and M suggests again that the limestone flux gives better bed behaviour than the BFS.

Bed behaviour during EPRI - 03 was thus usually satisfactory, with no operational problems. However, there were too many process changes to properly assess the effect of a multitude of operational parameters on gasifier behaviour. These

TABLE 12-4. PERIODS ANALYSED: EPRI - 03

Period	Coal	Oxygen Loading (SCFH)	Steam/O_2 Ratio	% Tar Injection	
				Tuyeres	Distributor
A	Ross.	160,000	1.3	0	0
B	Pitts. 8	130,000	1.3	0	0
C	Pitts. 8	130,000	1.3,1.2	65	0
D	Pitts. 8	130,000	1.2	95*	0
E	R.O.M. Pitts. 8	130,000	1.2	0	60
F	" "	80,000	1.2	0	50
G	" "	80,000	1.2	0	80
H	" "	50,000	1.3	0	80
I	" "	50,000	1.1	0	80
J	" "	130,000	1.2	0	60
K	" "	130,000	1.2	0	60
L	" "	130,000	1.2,1.3	0	60
M	" "	130,000	1.3	0	60
N	" "	160,000	1.3	0	60
O	" "	160,000	1.3	0	60
P	" "	50,000	1.3	0	60,0
Q	" "	160,000	1.3	0	0
MB1	Pitts. 8	130,000	1.3	0	0
MB2	Pitts. 8	130,000	1.3,1.2	65	0
MB3	Pitts. 8	130,000	1.2	95*	0
MB4	R.O.M. Pitts. 8	130,000	1.2	0	60
MB5	" "	130,000	1.2,1.3	0	60

* Estimate

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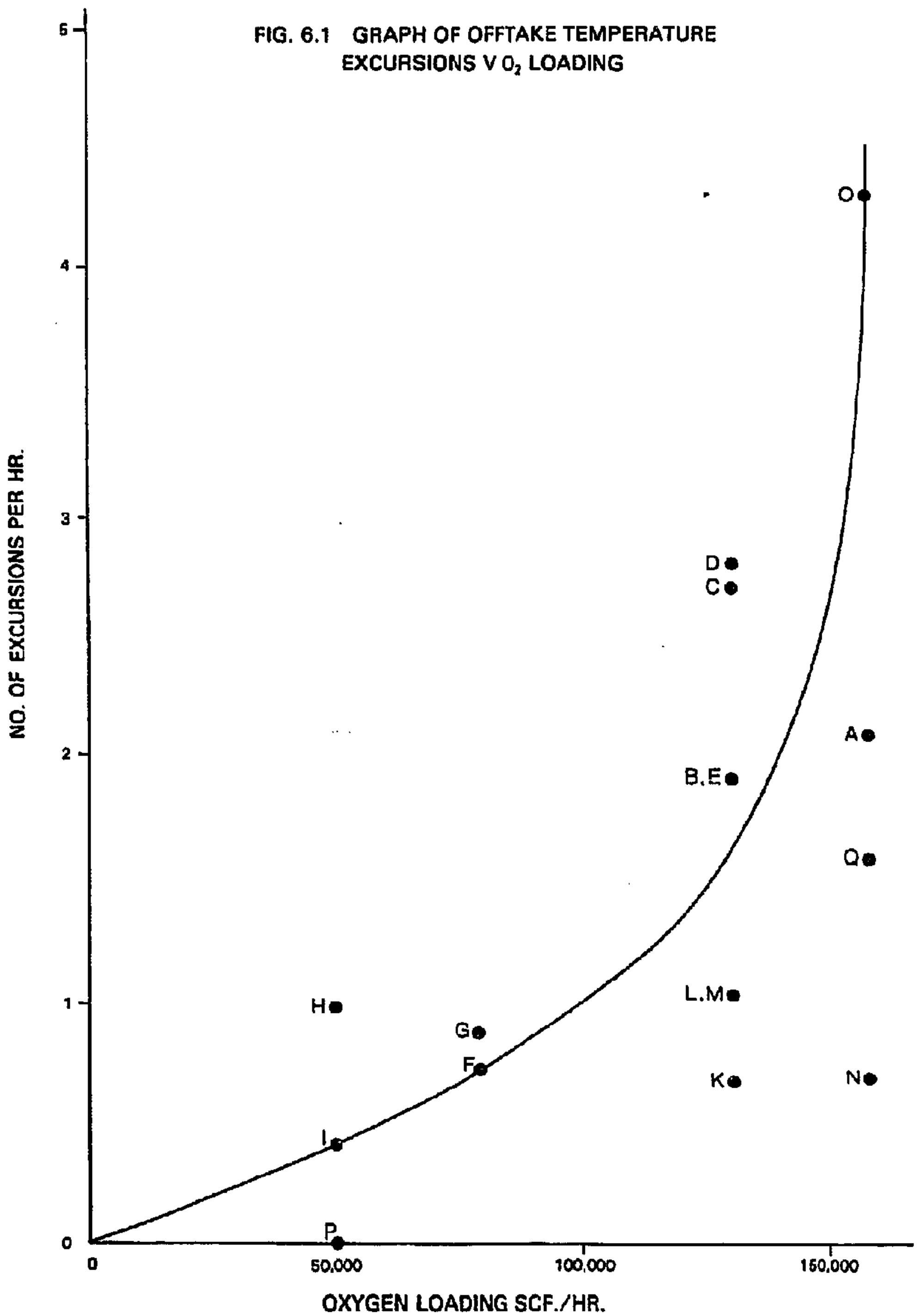


Figure 12-2. Graph of Offtake Temperature Excursions Vs. O_2 Loading

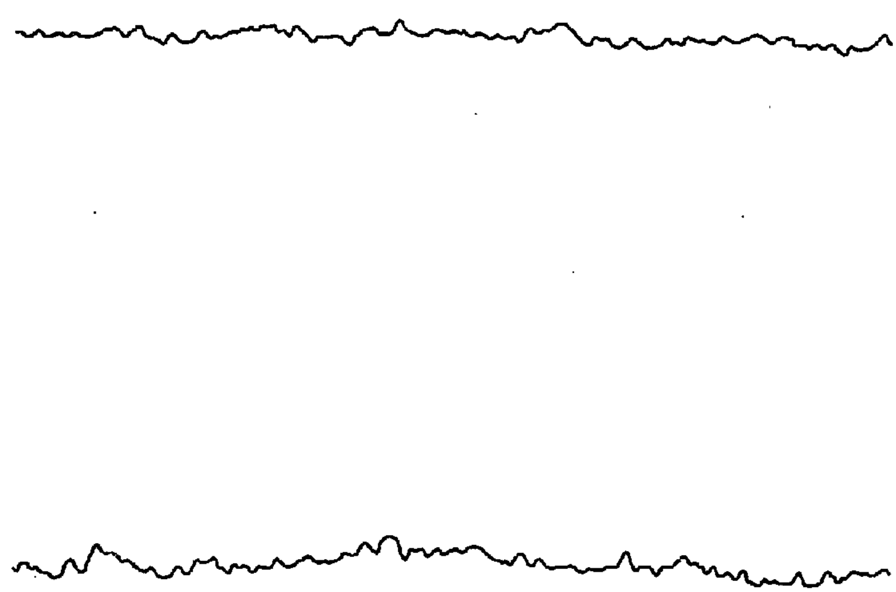


Figure 12-3. % CO_2 Charts. Top Chart: Period B. Bottom Chart: Period C
Chart Speed: 2 Min./CM

include tar injection at both the top and bottom of the gasifier, stirrer speed, effect of flux type, amount of fines in the coal and gasifier loading. All the above factors will contribute to bed conditions. Again, for the above reasons, it is difficult to compare the performance of Rossington in the gasifier to that of Pittsburgh 8.

PLANT BEHAVIOUR IN FLOW CONTROL

One of the objectives of run EPRI - 03 was to improve the gasifier's speed of response under flow control mode conditions. This was to be done in combination with trials on the tar injection to the tuyere region, gasification of Pittsburgh 8 coal containing a substantial amount of fines and fluxing with limestone. Control systems were as for EPRI - 02.

The controller settings for the various run periods are given in Table 12.5. Sections of the flow and pressure charts are given in Figure 12-4. During steady state operation, pressuring the coal lock on raw gas, which was practiced for most of the run, caused the oxygen and steam demand to fluctuate in order to provide the additional gas surge required.

The vortex meters gave satisfactory outputs throughout the run and should provide a better method of flow measurement because of their turn down capability. They did, however, also give a more variable flow signal at the lower rates.

The gas flow control valve was affected such that the change from low to standard throughputs caused it to fluctuate, and thus, the response to this valve was set rather slower than had been necessary on run EPRI - 02.

The experience of this run shows that it is possible to control the gasifier in the back end flow mode. The use of raw gas for pressurising the coal lock would mean that the oxygen/steam supply system would have to be fairly flexible with regard to throughput.

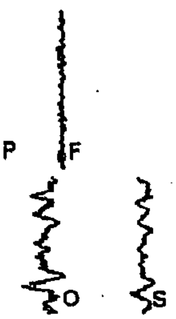
If the coal lock is pressurised with nitrogen, then fairly stable steam and oxygen flows will still give good control. Vortex meters, which were only used briefly during the run, are probably satisfactory for providing the flow signals to control the steam and oxygen throughputs.

TABLE 12-5. CONTROLLER SETTINGS FOR EPRI - 03

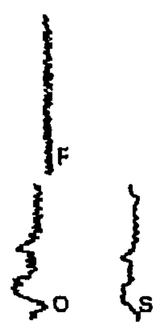
Condition	Day and Time of Setting up	Gasifier Pressure PC		Flare Gas Flow GFC		Oxygen Flow OFC		Steam Flow SFC		Approx Oxygen Blast	Control Type	Coal
		PB	Reset	PB	Reset	PB	Reset	PB	Reset	SCFH		
1	13 Dec - 19.15	30	0.1	500	2	200	10	200	5	130,000	Flow	Pittsburgh
2	14 Dec - 11.10	"	"	"	"	"	"	"	"	"	"	" + tar injection to tuyeres
3	15 Dec - 15.43	"	"	"	"	"	"	"	"	"	"	R.O.M. Pittsburgh
4	16 Dec - 05.20	25	0.5	"	"	300	5LR*	500	2LR*	80,000	"	R.O.M. Pittsburgh + coal locking on raw gas
5	16 Dec - 11.28	"	"	"	"	"	"LR*	"	2LR*	50,000	"	" " "
6	16 Dec - 18.15	"	"	"	"	200	10	200	5	130,000	"	" " "
7	17 Dec - 14.35	"	"	500	"	"	"	"	"	160,000	"	" " "
8	17 Dec - 20.07	"	"	"	"	"	"V*	"	"V*	"	"	" " "
9	18 Dec - 04.44	"	"	"	"	"	"	"	"	50,000	"	" " "
10	18 Dec - 09.30	"	"	700	"	"	"	"	"	160,000	"	" " "

LR* Low Range Transmitters
V* Vortex Meters

CONDITION 1 FLOW CONTROL. GASIFIER PRESSURE AND ALL FLOWS RELATIVELY STEADY.

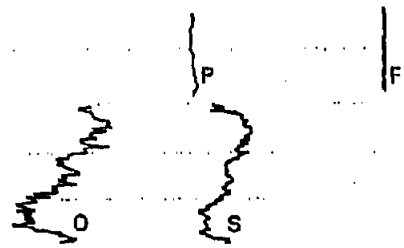


CONDITION 2 FLOW CONTROL. PRESSURE AND FLOWS ALL SLIGHTLY MORE VARIABLE DUE TO THE AFFECT OF TAR INJECTION DOWN TO THE TUYERES.

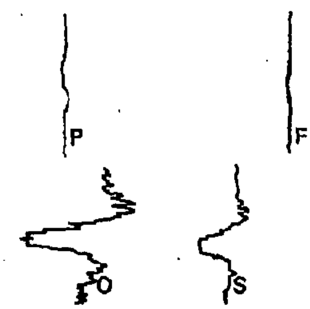


CONDITION 3 FLOW CONTROL. AGAIN EVERYTHING RELATIVELY STEADY. R.O.M. PITTSBURGH HAS LESS AFFECT THAN TAR INJECTION DOWN THE TUYERES.

Figure 12-4. Recorder Charts Showing Flow and Pressure Characteristics at Several Controller Settings



CONDITION 4 FLOW CONTROL. THE LOW RANGE TRANSMITTERS WERE IN USE THEREFORE THE O AND S VARIATIONS LOOK WORSE. THE CONTROL SETTINGS ALSO HAD TO BE ADJUSTED TO AVOID FLUCTUATIONS ON THE OXYGEN AND STEAM FLOW. LOCKING ON RAW GAS CAUSED PEAKS ON THE OXYGEN AND STEAM FLOWS.



CONDITION 5 FLOW CONTROL. THE LOW RANGE TRANSMITTERS WERE STILL IN USE. A TYPICAL FLOW AND PRESSURE VARIATION CAUSED BY A COAL LOCK IS SHOWN.

CONDITION 6 FLOW CONTROL. THE STEAM AND OXYGEN FLOWS ARE BACK ON THE HIGH RANGE TRANSMITTERS AND HAVE BEEN RESET TO THE NORMAL CONTROL SETTINGS. THE CONTROL ACHIEVED IS SIMILAR TO THAT IN CONDITION 1 EXCEPT FOR THE FLOW PEAKS CAUSED BY THE COAL LOCKS.

Figure 12-4 (Con't)

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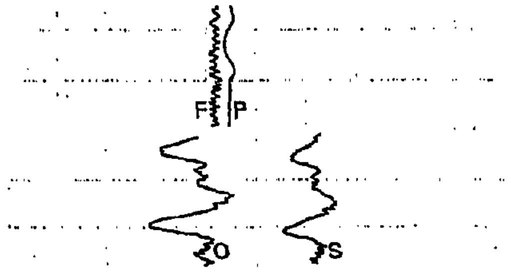
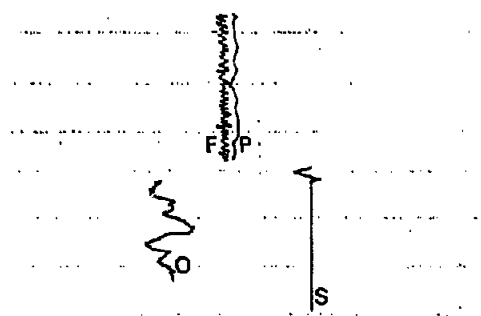
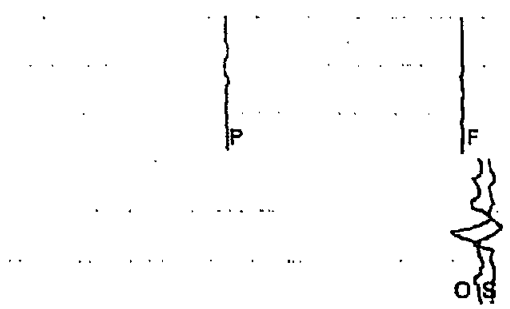


CONDITION 7 FLOW CONTROL. THE GAS FLOW CONTROL VALVE OSCILLATED WHEN THE FLOW WAS INCREASED TO THE EQUIVALENT OF 160,000 SCFH OXYGEN BLAST SO THE P.B. WAS INCREASED TO 500. THE GASIFIER PRESSURE, OXYGEN AND STEAM FLOWS ARE AGAIN VARIABLE BECAUSE OF THE FREQUENT COAL LOCKING ON RAW GAS.

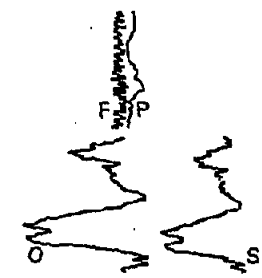


CONDITION 8 FLOW CONTROL. THE VORTEX SIGNAL IS LINEAR WITH FLOW. ONLY THE OXYGEN AND STEAM FLOWS WERE PARTIALLY CONTROLLED USING THE VORTEX SIGNALS. THEY MADE LITTLE DIFFERENCE TO THE FLOW VARIATIONS AT THIS POINT ON THE RANGE.



CONDITION 9 FLOW CONTROL. AT LOW RATES THE COAL LOCKS ARE RELATIVELY FAR APART. THE AFFECT OF COAL LOCK IS SHOWN. THE PRESSURE CONTROL IS VERY GOOD EXCEPT FOR THE DIP AND RECOVERY DURING A LOCK.

Figure 12-4 (Con't)



CONDITION 10 FLOW CONTROL. THE GAS FLOW CONTROL VALVE AGAIN OSCILLATED WHEN THE FLOW WAS INCREASED TO THE EQUIVALENT OF 160,000 SCFH OXYGEN BLAST SO THE P.B. WAS AGAIN INCREASED TO 700. THE GASIFIER PRESSURE, OXYGEN AND STEAM FLOWS ARE AGAIN VARIABLE BECAUSE OF THE FREQUENT COAL LOCKING ON RAW GAS, BUT THIS HAD BEEN MADE WORSE BY THE REDUCED CONTROL OF THE GAS FLOW.
THE VALVE OSCILLATION MAY BE LINKED TO THE CHANGES OF FLOW AS THE SYSTEM HAD WORKED PREVIOUSLY AT THESE SETTINGS (SEE CONDITION 7).

Figure 12-4 (Con't)

BEHAVIOUR OF THE GASIFIER DURING TRANSIENTS

During EPRI - 03 several transient load changes were carried out and the general description of these changes is given in Table 12-1.

With the control valves set up as for Condition 10, a load reduction spike was carried out reducing from 160,000 SCF/H oxygen blast equivalent to 110,000 SCF/H and back again. The change and its effect on the constituents of the make gas are showin in Figures 12-5 to 12-8.

The major step change of the run was from 160,000 SCF/H to below 50,000 SCF/H oxygen blast equivalent in about two minutes with the controllers set up as for Condition 7. This change and its effect on the constitutents of the make gas are shown in Figures 12-9 to 12-13. The most rapid increase in the gas make was from 50,000 SCF/H to 130,000 SCF/H oxygen blast equivalent in three minutes with the controllers set up as for Condition 6. This change and its effect on the constituents of the make gas are shown on Figures 12-14 to 12-17. It can be seen from these figures that, apart from the normal changes in nitrogen levels, there is a small temporary increase in the methane and ethane percentages of 2% and one tenth of a percent respectively after the rapid major load reduction and, similarly, there is also a similar temporary reduction after the rapid increase in throughput, although the latter is not so well defined. Care should be taken to keep the gas flow control valve in good condition so as to avoid fluctuations when making large increases in throughput.

Major changes in the gas make carried out in periods of two or three minutes can cause a slight temporary change in the methane and ethane levels in the product gas. Flow spikes from 160,000 to 110,000 SCF/H oxygen blast equivalent have no significant effect. The slight temporary change in methane and ethane levels may disappear by change in operational techniques.

TAR INJECTION TO THE TUYERES

With period B of the run completed (which was carried out at 130,000 SCF/H oxygen loading, 1.30 H_2O/O_2 and no tar injection either at the top or the bottom) tar injection was brought down tuyeres at a pump stroke setting of 65% to give an estimated tar flow rate of between 1,000 to 1,200 lbs/hr. The introduction of tar down the tuyeres produced no dramatic effects at the gasifier bottom, and the steam/oxygen ratio was moved from 1.30 to 1.20 in anticipation of cooler raceway conditions.

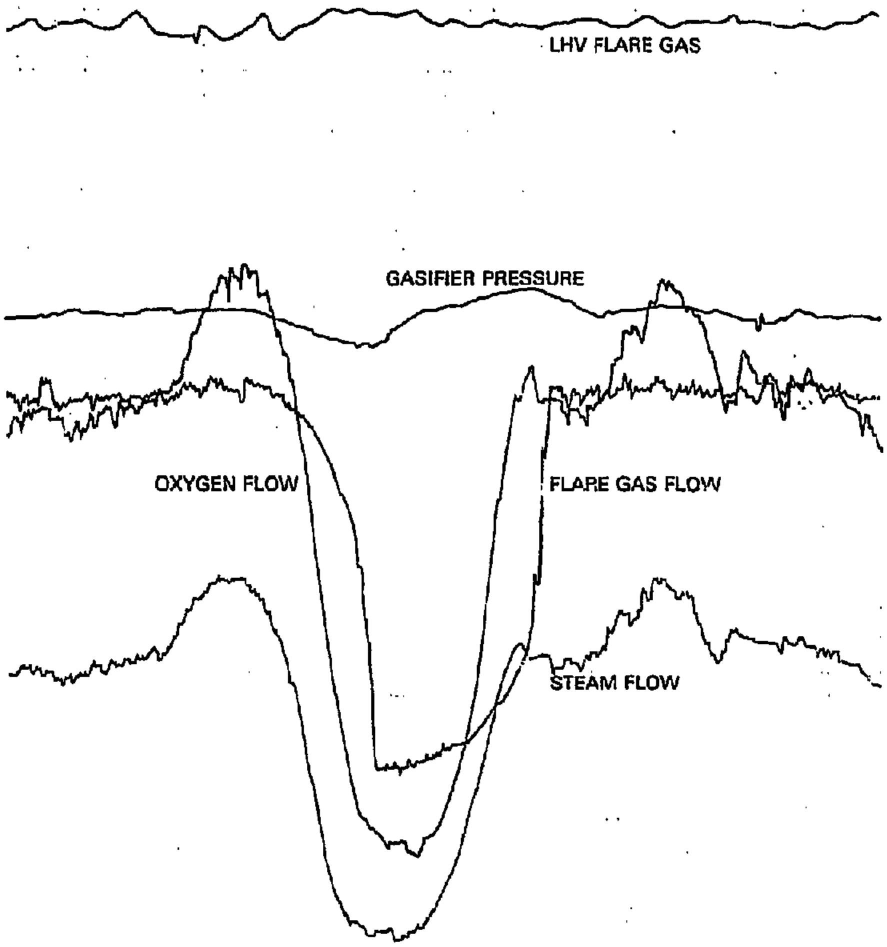


Figure 12-5. Major Gasifier Parameters Across a Load Change from 160,000 to 110,000 SCF/H Oxygen and back
EPRI - 03
Chart Speed: 2 Min./CM

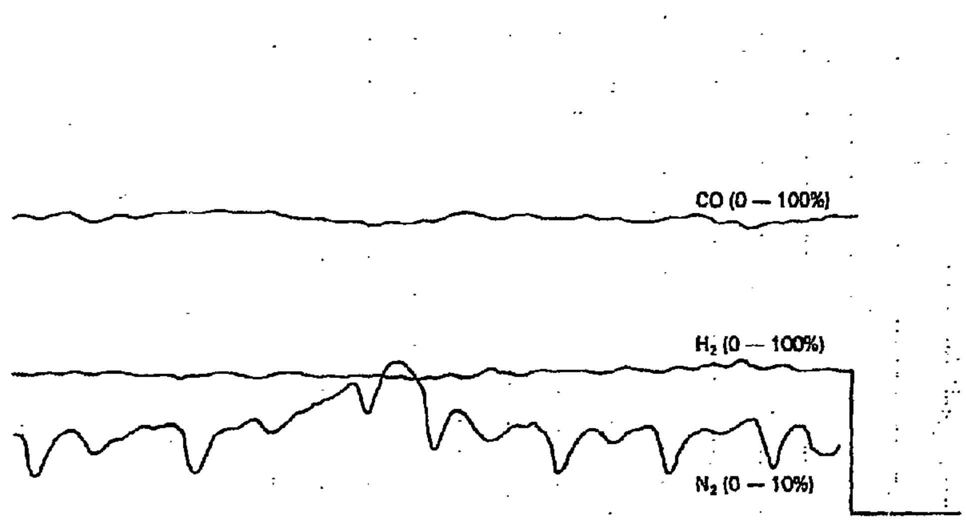


Figure 12-6. Offtake Gas Composition Across a Load Change from 160,000 to 110,000 SCF/H Oxygen and Back
EPRI - 03
Chart Speed: 2 Min./CM

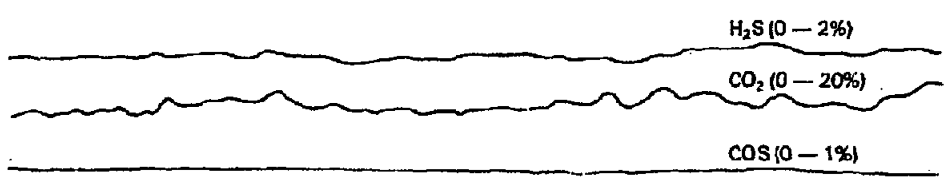


Figure 12-7. Offtake Composition Across a Load Change from 160,000 to 110,000 SCF/H and Back
EPRI - 03
Chart Speed: 2 Min./CM

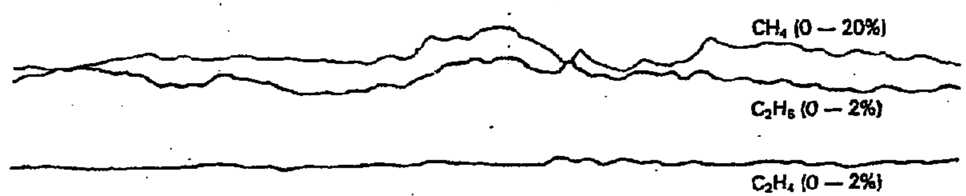


Figure 12-8. Offtake Gas Composition Across a Load Change from 160,000 to 110,000 SCF/H and Back
EPRI - 03
Chart Speed: 2 Min./CM

ρ

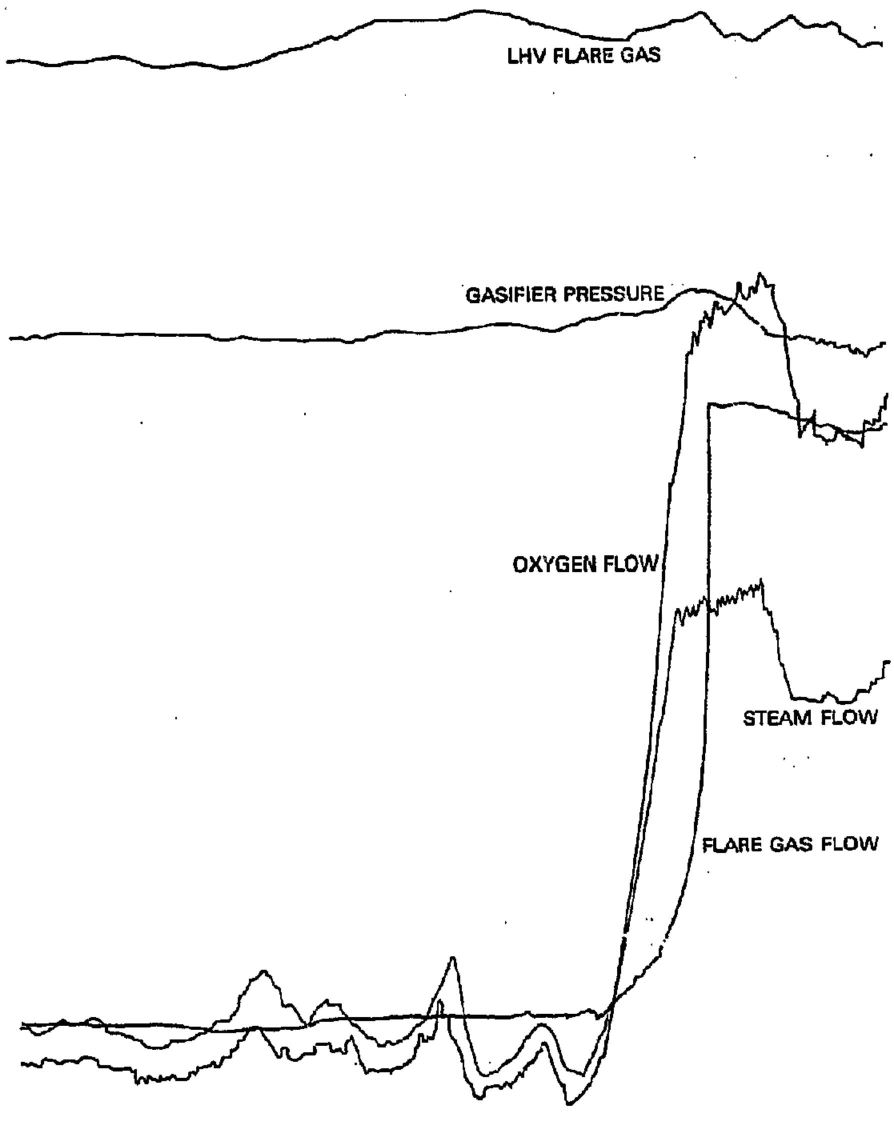


Figure 12-9. Major Gasifier Parameters Across a Load Reduction from 160,000 to 50,000 SCF/H
EPRI - 03
Chart Speed: 2 Min./CM

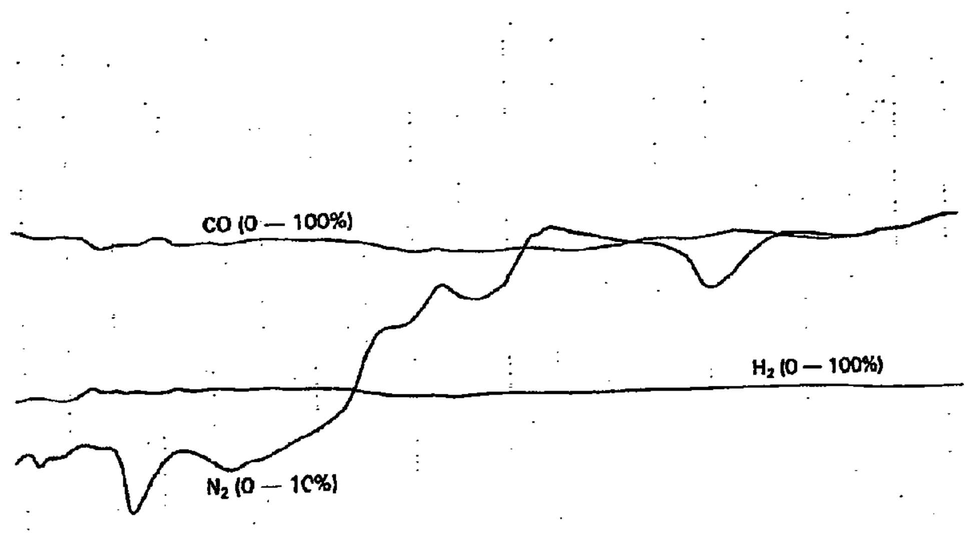


Figure 12-10. Offtake Gas Composition Across a Load Reduction from 160,000 to 50,000 SCF
EPRI - 03
Chart Speed: 2 Min./CM

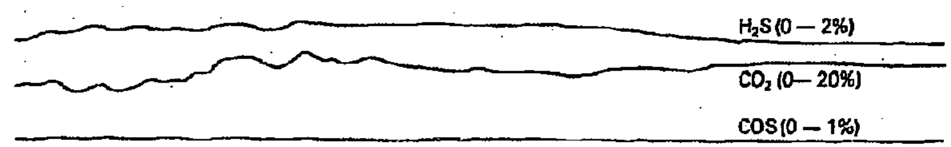


Figure 12-11. Offtake Gas Composition Across a Load Reduction from 160,000 to 50,000 SCF/H
EPRI - 03
Chart Speed: 2 Min./CM

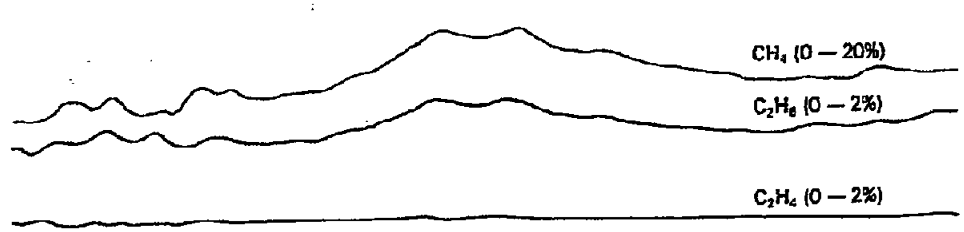


Figure 12-12. Offtake Gas Analysis Across a Load Reduction from 160,000 to 50,000 SCF/H
EPRI - 03
Chart Speed: 2 Min./CM

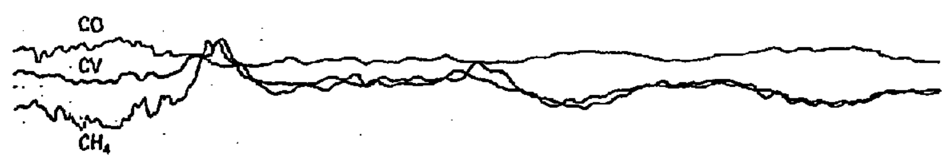


Figure 12-13. Make Gas Composition and CV Across a Load Reduction from 160,000 to 50,000 SCF/H
EPRI - 03
Chart Speed: 10 Min./CM

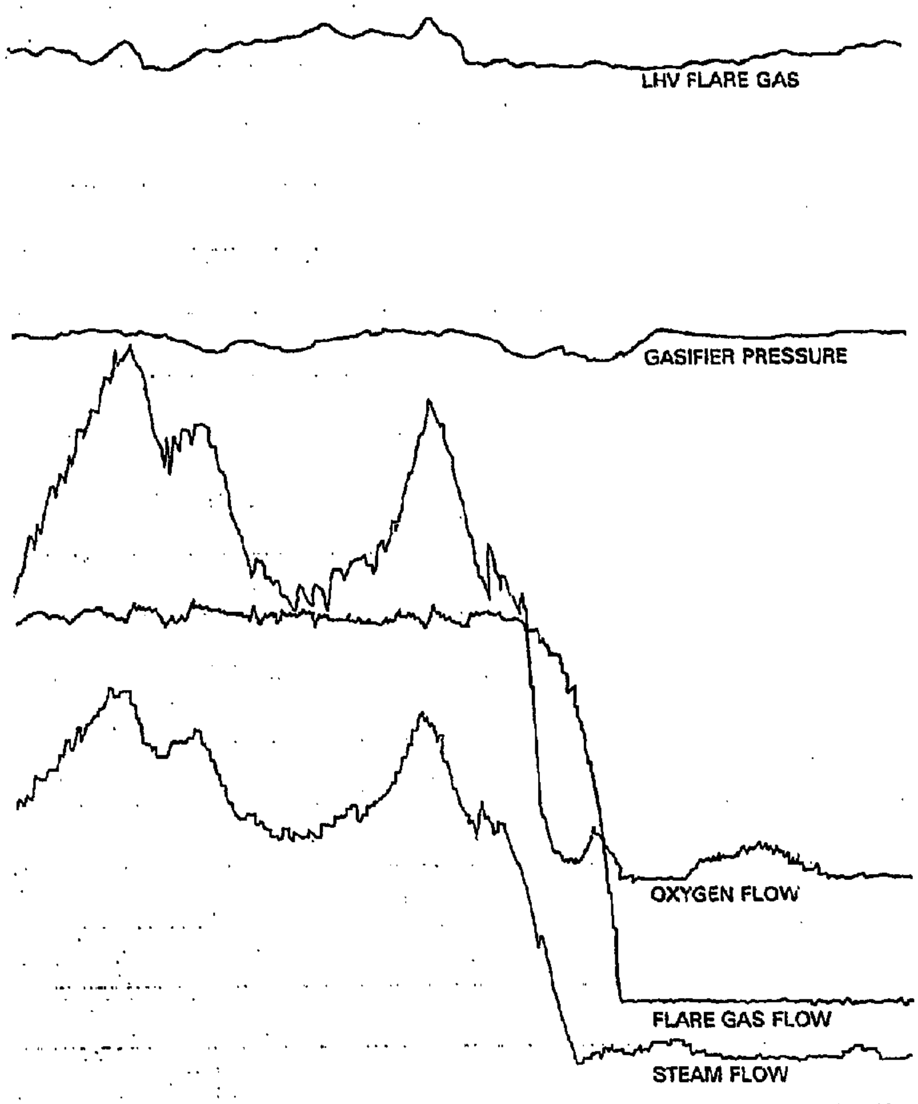


Figure 12-14. Major Gasifier Parameters Across a Load Increase from 50,000 to 130,000 SCF/H
EPRI - 03
Chart Speed: 2 Min./CM

Figure 12-15. Offtake Gas Composition Across a Load Increase from 50,000 to 130,000 SCF/H
Oxygen
EPRI - 03
Chart Speed: 2 Min./CM

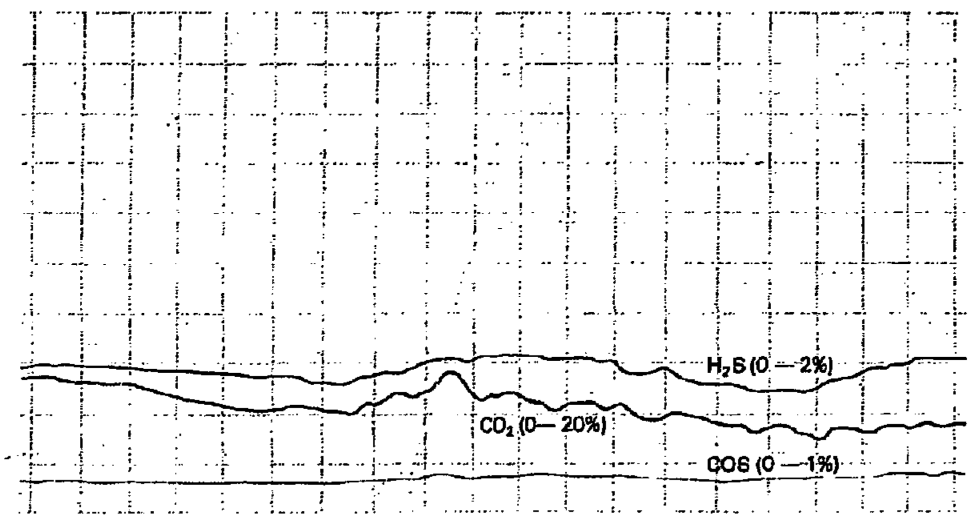


Figure 12-16. Offtake Gas Composition Across a Load Increase from 50,000 to 130,000 SCF/H Oxygen
EPRI - 03
Chart Speed: 2 Min./CM

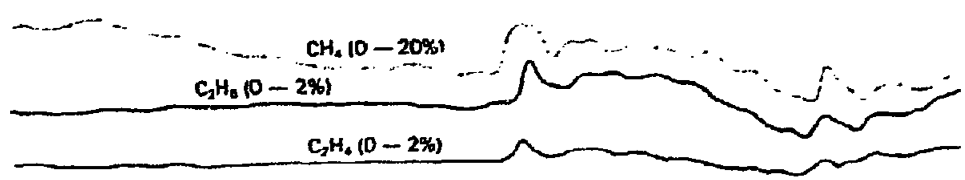


Figure 12-17. Offtake Gas Composition Across a Load Increase from 50,000 to 160,000 SCF/H
Oxygen
EPRI - 03
Chart Speed: 2 Min./CM

Later analysis (see Figure 12-3, Table 12-4, and related discussion) indicated that the bed was slightly less steady with tar injection to the raceway and that hearth conditions had deteriorated slightly, although both these factors did not interfere with continued satisfactory operating conditions. The average offtake temperature of the gasifier rose from 902 to 971°F, and the average methane concentration in the offtake gases dropped to 5.7% from 6.4%. The above represented the total of appreciable differences that could be observed on line, and, after a 12 hour period, the pump stroke was raised to 95%. During the ensuing period, at the supposedly higher injection rate, the offtake temperature average remained the same and the methane concentration in the make gas did not change, leading to the suspicion that the flow rate of tar had not changed, despite the alteration in pump stroke.

POST RUN INSPECTION

The gasifier was shut down while running at 160,000 SCF/H oxygen loading on unscreened Pittsburgh 8 at the moment the coal lock was empty. Standard procedures were used.

Post run inspection revealed that the distributor was nearly empty. There was some caked material at the top of the bed.

The distributor area was full of caked coal which was readily broken into smaller pieces, and below the distributor, there were discrete char lumps generally in the range ½" to 1½"; two exceptions were two discrete volumes of caked coal, which, again, were readily broken up into smaller pieces. There was little dust in the bed until the very bottom was reached, when there was a considerable amount of dust associated with the raceway char.

CONCLUSIONS

Run EPRI - 03 extended the experience of run EPRI - 02 on Pittsburgh 8 coal with further experience on load turn up and turn down, with particular emphasis on control systems; with running on limestone instead of BFS flux; with running on a coal feed containing a considerable amount of fine material, and with running with tar being injected down the tuyeres.

Load turn up and turn down again proved to be a reliable operation with little evidence of transient upsets occurring in the gasifier. The gasifier worked well in flow control, and it appeared possible to tune the control characteristics of the plant to obtain a minimum of fluctuation in the steam/oxygen flows and on gasifier pressure, although some significant variation in gasification medium is likely to occur if coal locking on raw gas is practiced. The gasifier operated at least as well on limestone flux as on BFS flux, although there was not time to optimise either fluxing agent. Up to 25% material less than ¼" in the coal feed appeared to make little difference to gasifier performance.

Tar injection to the tuyeres at a rate at least equal to the net gasifier make at the top end was successfully accomplished. When running with tar injection to the tuyeres, both hearth and bed behaviour appeared to deteriorate a little, although there was no indication of any process problems.

The three EPRI runs have therefore been able to demonstrate the potential and versatility of the Slagging Gasifier when working on Pittsburgh 8 coal, although there was little time to optimise gasifier operating conditions, and still less to optimise gasifier design for performance of this coal. Given this time, it is apparent that commercial operation in the combined cycle mode, for electric power generation, with all its advantages, is certain to be successful.