EPA U.S. Environmental Protection Agency
Office of Research and Development

Industrial Environmental Research Laboratory
Research Triangle Park, North Carolina 27711

EPA-600/7-77-027
March 1977

FIRST TRIALS OF CAFB PILOT PLANT ON COAL

Interagency
Energy-Environment
Research and Development
Program Report

RESEARCH REPORTING SERIES

Research reports of the Office of Research and Development, U.S. Environmental Protection Agency, have been grouped into seven series. These seven broad categories were established to facilitate further development and application of environmental technology. Elimination of traditional grouping was consciously planned to foster technology transfer and a maximum interface in related fields. The seven series are:

1. Environmental Health Effects Research
2. Environmental Protection Technology
3. Ecological Research
4. Environmental Monitoring
5. Socioeconomic Environmental Studies
6. Scientific and Technical Assessment Reports (STAR)
7. Interagency Energy-Environment Research and Development

This report has been assigned to the INTERAGENCY ENERGY-ENVIRONMENT RESEARCH AND DEVELOPMENT series. Reports in this series result from the effort funded under the 17-agency Federal Energy/Environment Research and Development Program. These studies relate to EPA's mission to protect the public health and welfare from adverse effects of pollutants associated with energy systems. The goal of the Program is to assure the rapid development of domestic energy supplies in an environmentally-compatible manner by providing the necessary environmental data and control technology. Investigations include analyses of the transport of energy-related pollutants and their health and ecological effects; assessments of, and development of, control technologies for energy systems; and integrated assessments of a wide range of energy-related environmental issues.

REVIEW NOTICE

EPA-600/7-77-027
March 1977

FIRST TRIALS OF CAFB PILOT PLANT ON COAL

by

D. Lyon

Esso Research Centre
Abingdon, Oxfordshire OX13 6AE
England

Contract No. 68-02-2159
Program Element No. EHE623A

EPA Project Officer: S.L. Rakes

Industrial Environmental Research Laboratory
Office of Energy, Minerals, and Industry
Research Triangle Park, N. C. 27711

Prepared for

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, D.C. 20460

FIRST TRIALS OF 3/4 MWe

C.A.F.B. PILOT PLANT ON COAL

Author:- D. Lyon

Work Done By:- D. Lyon, A.W. Ramsden, A.W. Brimble, O.R. Priestnall, A. Jennings, D. Storms, (see acknowledgement)

November 1976

CONTENTS

CONTENTS (Cont.d)

I SUMMARY

The information in this report relates to a mini-run carried out on the 3/4 MWe continuous CAFB pilot plant during July-August 1976. These tests represent the first step in a 3 year research and development programme funded by E.P.A. to extend the CAFB process to operate on coal. This research effort is seen as a necessary back-up to a 20 MWe demonstration plant to be constructed in 1977 at San Benito, Texas.

The entire report is based on about 8½ hours of gasification on Texas lignite and Illinois No.6 coal. Nevertheless the results do represent the best information available to date on coal gasification using the CAFB process.

No major barriers have yet been identified in conversion of the process from oil to coal. The quality of the gas produced is similar and the desulphurising efficiency on coal can be expected to match or exceed that on oil. The target of 88% lignite utilization set by Foster Wheeler Corp Ltd for the Texas demonstration plant is seen as realistic and was approached in the mini-run under conditions which can be considered far from optimised. Because of the need for more air to gasify coal the CAFB converted boiler dimensioned for fuel oil will probably be downrated by 30% on coal.

Satisfactory regeneration while gasifying on lignite was demonstrated by slowing the bed transfer rate compared with operation on fuel oil. This mode of operation does not in any way impair gasifier performance but may result in the need to control regenerator temperature for example by steam injection or flue gas recycle.

The new fines return system appears to have reduced the burden of lime into the boiler by a factor of ten but did little to reduce the 'fly-ash' level in the boiler during coal gasification. Further work in this area is obviously needed.

Successful operation of the pilot plant on an intermittent basis as opposed to round-the-clock operation was clearly demonstrated. This opens up the technical possibility of using a CAFB converted boiler on a peak shaving rather than a base load basis but obviously the economics of conversion will largely determine the extent this is practicable.

An extensive programme of work on lignite gasification is scheduled to begin in Mid November 1976.

II INTRODUCTION

Considerable background of knowledge of CAFB operation on heavy fuel oil

A 3/4 MWe pilot plant was built in 1970 to study continuous operation of the CAFB process using high sulphur fuel oil. Essentially this process gasifies and desulphurises heavy fuel oil producing a clean, virtually sulphur free gas which can be combusted in a conventional gas-fired boiler. Since 1970, nine runs totalling approximately 2,700 hours of operation on fuel oil have been carried out. In November 1975, a new gasifier/regenerator unit was built and a run of approximately 300 hours duration (Run 10) was carried out by the end of that year. A report on the findings of Run 10 is currently being written while reports covering earlier work are available (EPA-650/2-74-109; EPA-650/2-75-027b). The success of the pilot plant studies has led to the first commerical demonstration of the process being built under licence by Foster Wheeler Development Corp in Texas. This involves conversion of a 20 MWe boiler currently fired on natural gas and is scheduled to be completed by mid 1977.

Only limited information is available on extending CAFB process to run on coal

A prime condition imposed by Central Power and Light owners of the Texas power plant for conversion of their boiler to CAFB was that the process should be capable of operating on coal. Prior to the mini-run, information on the operation of the CAFB process on coal was limited to batch studies carried out on a 7" diameter reactor. Experimental work is continuing on this unit.

Recently completed Min-Run represents first trials of 3/4 MWe continuous unit on coal

A mini-test run on the 3/4 MWe continuous unit has recently been completed in August 1976. This run is the forerunner of a major study of coal gasification (Run 11) scheduled to start November 1976.

The main objectives of the mini-test were:-

(1) Test all new subsystem equipment incorporated since Run 10 under actual running conditions.

(2) Gasify and regenerate with Illinois No.6 coal and also if possible lignite selected by EPA for Run 11.

(3) Identify problem areas associated with coal gasification which may necessitate major structural changes prior to Run 11 e.g. need for two cell regenerator.

(4) Carry out tests specifically requested by Foster Wheeler Corporation i.e.

- Test Commercial SO_2 sample probe
- Test silicon carbide coal injection needles.

The generation of quantitative information was not a prime objective of the mini-run but during the final week the unit was operating sufficiently trouble free for this to be obtained. While it is not claimed that the system was in any way optimised the results presented in this report represent the best information to date on coal gasification using the CAFB process.

COOLING CURVE FOR GASIFIER BED AFTER SHUTDOWN

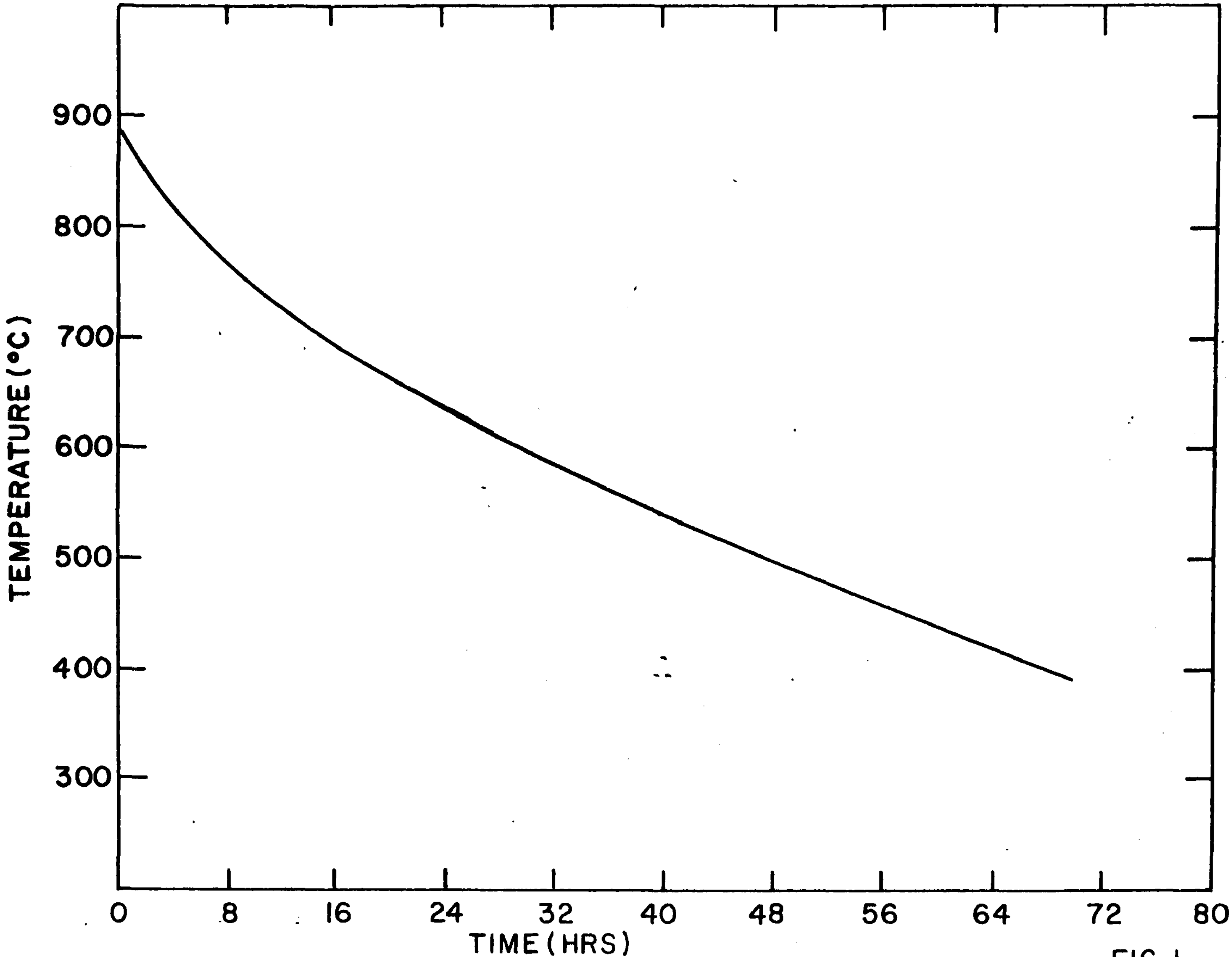

FIG. I.

COAL HANDLING SYSTEM

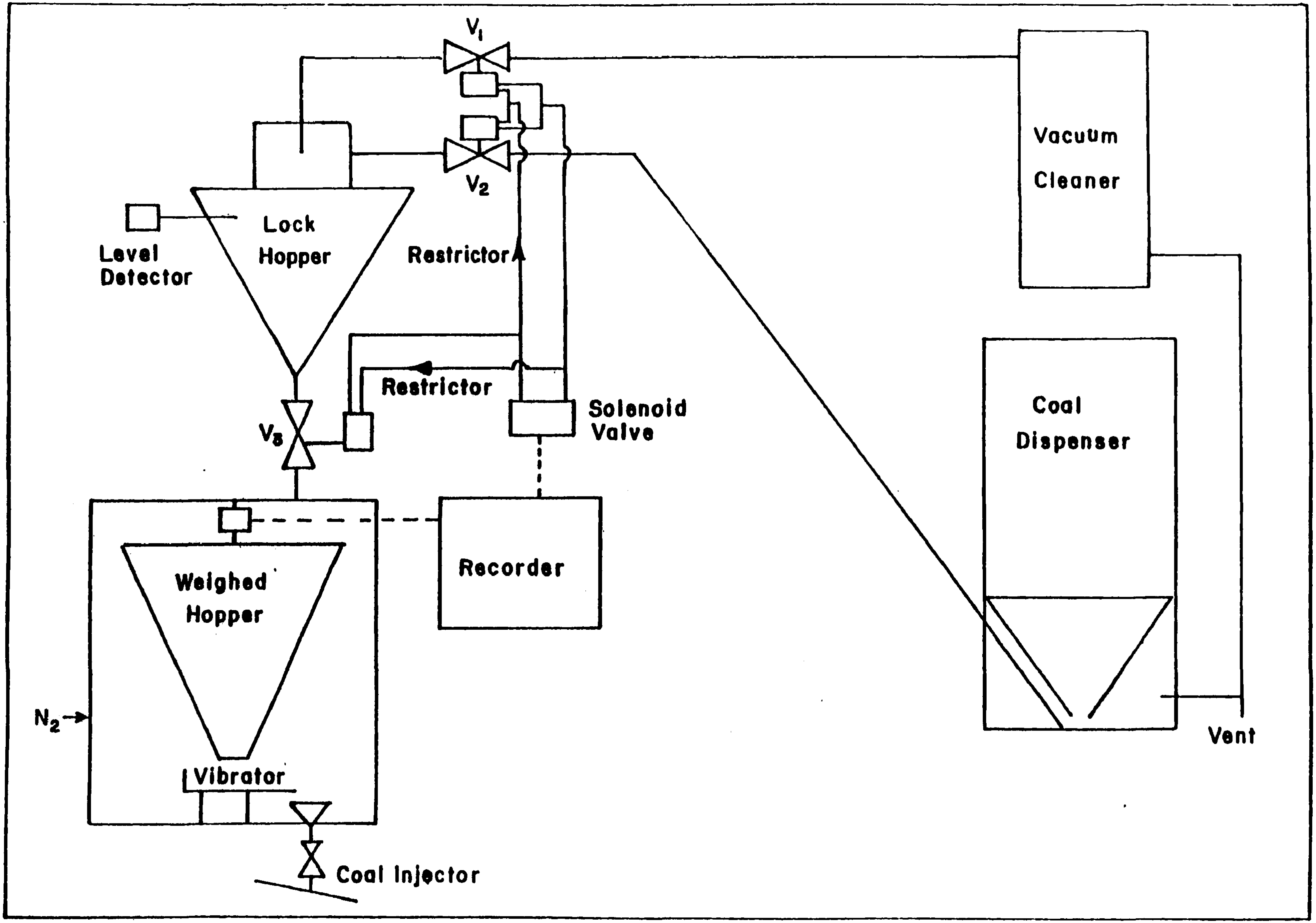

FIG.2.

III EXPERIMENTAL

Day-time running as opposed to round-the-clock operation was successfully demonstrated on the Mini-run

Previous runs on the continuous pilot plant involved round-the-clock operation of the unit carried out on a shift basis. However, during the mini-test the unit was run during the day, shutdown overnight and cooling of the unit was sufficiently slow to allow it to be restarted the next day without any problems. Figure 1 shows an example of the cooling curve obtained. Initially the unit was restarted on heavy fuel oil and the temperature of the bed material was not allowed to drop below 650°C. Later, however, the unit was successfully restarted at 390°C on coal and this meant that the unit could be left shutdown and unattended at weekends and still be successfully restarted.

The flexibility day-time running of the unit allows should result in manpower savings and more efficient tackling of any technical problems which may arise during Run 11. The need for continuous round-the-clock running of the unit to finally prove the process on coal will still remain but probably for shorter periods than in earlier runs.

Perhaps the most important point about the method of operating the unit during the mini-run is its implications for the Texas demonstration plant. The thermal inertia of the bigger unit in Texas will undoubtably be greater resulting in a longer time lag between shutdown and the need to restart. This opens up the possibility of using a CAFB converted boiler on a peak shaving rather than base load basis. Further work to explore the flexibility of the unit in this area is planned for Run 11.

New subsystem equipment incorporated since Run 10 greatly improved operational reliability

Several new features were incorporated since Run 10 to overcome problems of equipment reliability during that run. An important objective of the mini-run was to test these systems under actual running conditions and identify areas where improvements could be made.

Coal feed system

After Run 10 was completed on fuel oil the limiting factor on testing the unit on coal was the absence of a coal feed system. To overcome this the existing limestone handling system was modified to handle coal. To accommodate the higher feed rates this entailed, several improvements had to be made and Figure 2 shows a schematic diagram of the system which was evolved.

The system was fully automatic, the sequence being controlled by two micro-switches fitted to the weigh-cell recorder. When the low limit switch was actuated it operated a solenoid valve which admitted compressed air to the actuators of valves V_1, V_2 and V_3.

BROKEN SILICON CARBIDE COAL INJECTOR NEEDLE

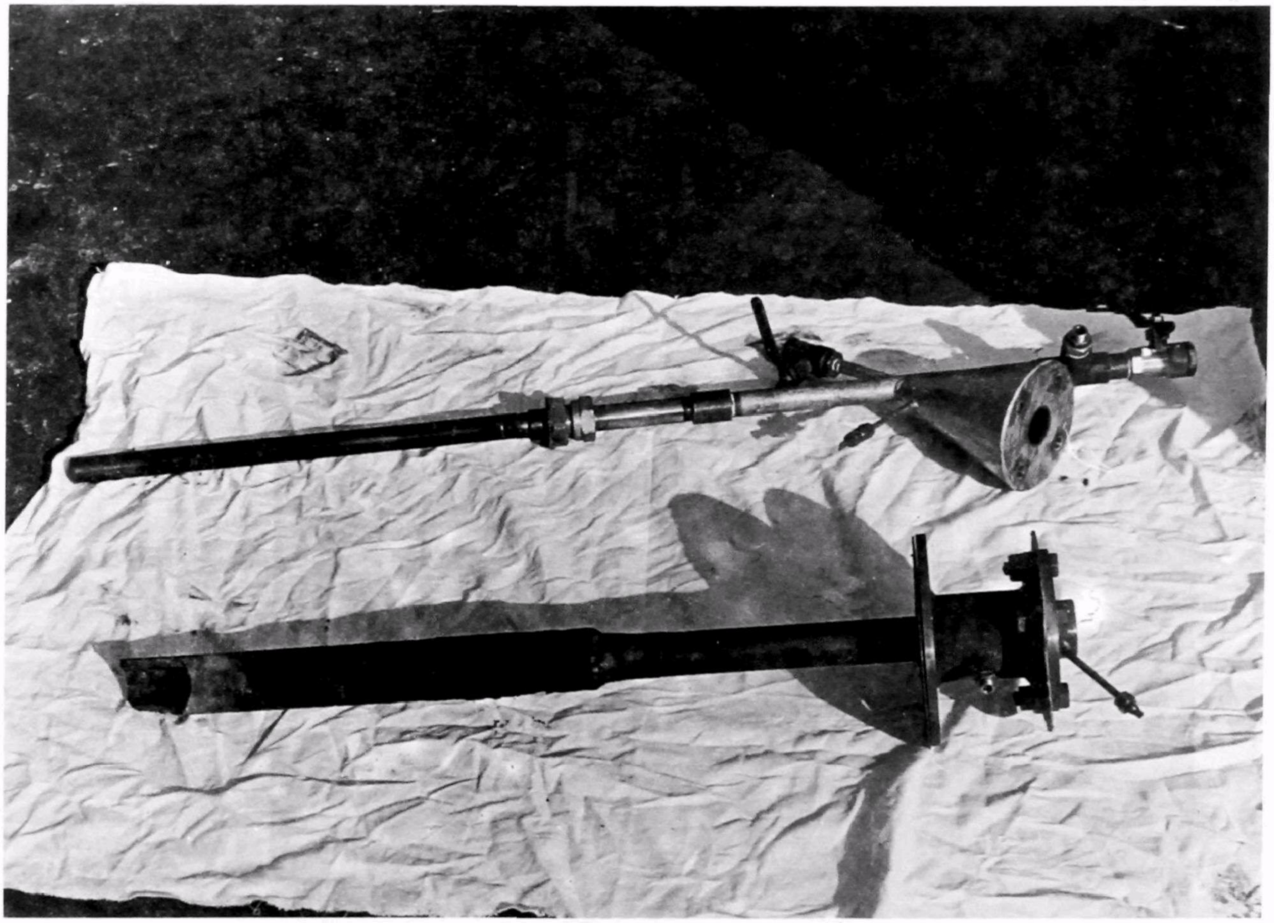

FIG. 3.

VIEW OF GASIFIER DISTRIBUTOR SHOWING PIECES OF SILICON TUBE

FIG. 4.

Valve V_3 opened whilst valves V_1 and V_2 were shut but a flow restrictor in the line to V_3 ensured that V_1 and V_2 closed before V_3 opened. This enabled the pressure in the lock hopper to rise and prevented a sudden fall in pressure within the coal metering vessel while it was being filled with coal. Such a surge was found during testing to cause blockages in the coal injector because the pressure differential between the feed system and the gasifier was momentarily reversed. When enough coal had entered the weigh hopper to activate the high limit switch on the recorder this deactivated the solenoid valve and valve V_3 shut whilst valves V_1 and V_2 opened. In this case a flow restrictor was used to delay the opening of V_1 and V_2 until V_3 was shut. A third micro-switch controlled the power to the vacuum cleaner and was made by the closing of valve V_3. This ensured that at no time was the vacuum cleaner switched on when valve V_3 was open. A level detector switched off the vacuum cleaner when the lock hopper was filled with coal. Should the level detector not be actuated within a set time after the actuation of the high limit switch then an alarm indicated that the coal dispenser was empty and this was replaced.

During the mini-run the automatic coal handling system worked well but some problem was experienced with the coal injector. In response to Foster Wheelers wish to test self-bonded silicon carbide as an injector material a sample tube supplied by Carborundum Ltd. was used. This tube penetrated the gasifier wall at a position 20" above the distributor and was angled downwards so that it extended into the middle of the gasifier a few inches above the distributor. Unfortunately, during the course of the mini-run tests the brittle silicon carbide tube snapped off at the gasifier wall. Figure 3 shows the remains of the silicon carbide tube whereas figure 4 shows pieces of the broken tube found in the distributor pit.

The information on coal gasification found in this report relates to a period of operation after the coal injector was broken. Consequently the feeding of the coal at a position 20" above the distributor and flush with the wall of the gasifier cannot be considered optimum conditions and hence the results presented should be viewed in such a light.

In the forthcoming Run 11 tests a KT-silicon carbide injector supplied by Foster Wheeler will be tried and also injection of the coal into the distributor pit via a duct in the distributor. In order to be able to feed both limestone and coal simultaneously to the unit a second coal handling system similar to the existing design but incorporating a screw feeder instead of a vibrator is to be constructed before Run 11.

Fines Return System

A major source of problems during Run 10 and previous test runs has been the drainage of fines from the gasifier cyclones and their subsequent re-injection into the gasifier bed. During Run 10 a lock-hopper arrangement was used, the valves being operated on a cyclic sequence. Trouble was experienced with these valves,

FLOW DIAGRAM FOR FINES RETURN SYSTEM

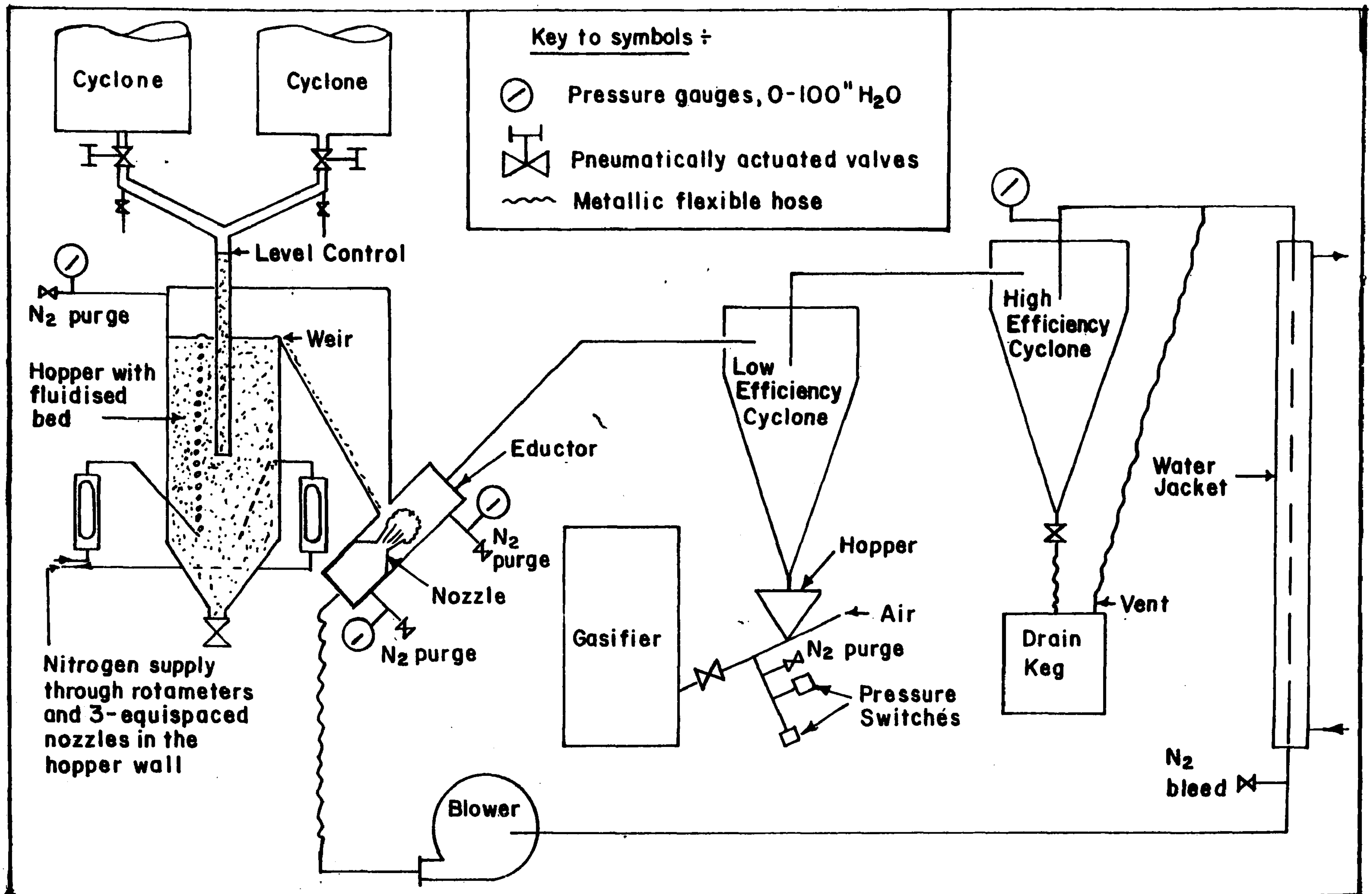

FIG. 5.

which ran at a high temperature and were consequently poorly lubricated. Also all of the fines were re-injected into the gasifier and continued use of the system may have led to problems of ash build-up while operating the unit on coal.

With these points in mind and also the need, due to restricted height, to positively inject the fines in the Texas unit a valveless system was designed with the following attributes

(a) Provision for the removal of large chunks from the system which may break loose from the cyclone walls during burnout.

(b) Provision for fractionating the fines so that a coarse fraction may be re-injected whilst a finer fraction may be discarded.

(c) Provision for re-injecting the coarse fraction of fines at a position remote from the gasifier cyclones and near the bottom of the gasifier bed.

A schematic layout of the new system is shown in figure 5. Fines from the two gasifier cyclones enter a Y-shaped duct which drains into a pot the contents of which may be fluidised by means of radially disposed gas nozzles. When the pot is not fluidised the fines accumulate in the Y-shaped duct but when fluidising gas is provided the contents of the duct drain into the pot and overflow a weir into an eductor, being conveyed to two small cyclones in series, by a circulating stream of gas supplied by a small blower. The fluidisation of the drain-pot is intermittent and is cut out when the height of solids held in the Y-shaped duct falls below a set level. In this way a seal is maintained and it is possible to recirculate fines at a chosen pressure.

The new fines return system operated successfully throughout the mini-run with only a few minor teething problems. Some improvements to the system were made during the run and a few more are to be incorporated before Run 11. Essentially this new system represents a considerable improvement in terms of reliability compared with previous systems.

Fractionation of the fines is made possible by having the 1st fines return system cyclone relatively inefficient compared with the second and thus it allows finer material to pass through and be collected in the second. No attempt was made during this mini-run to explore and optimise the partition between the two cyclones but this is obviously necessary and should be investigated in Run 11.

An analysis of the performance of the fines return system in reducing the dust losses to the boiler can be found in the discussion section of this report.

Flue Gas Recycle System

The bag filter which was incorporated in the flue gas recycle system for Run 10 was replaced by a high efficiency cyclone and the dust content of the incoming flue gas was reduced by moving the take-off point downstream of the stack cyclone. The recycled flue gas was fed directly into the fluidised bed via four 1 11/16" diameter firebird blue tubes.

This system was tested during the mini-run for several hours while gasifing on fuel oil and no problems were experienced. No change in the system is proposed for Run 11.

Bed Transfer System

To avoid problems with poor bed transfer experienced in Run 10 two improvements were carried out:-

(a) The rodding ports were enlarged

(b) A N_2 purge was introduced into the regenerator and gasifier during shutdown with a sulphided bed to avoid agglomeration of bed material contacted with air.

No problems with the bed transfer system were experienced during the whole of the mini-run. Any sluggishness experienced when transferring relatively cold beds was counteracted by increasing the transfer nitrogen pressure. To avoid free-wheeling of the transfer system the horizontal pulser injectors were progressively moved into the transfer pockets. Good control was obtained with the injectors protruding approximately 1½" into the pockets.

Minor Modifications

Several minor modifications to the unit were made which also proved successful during the mini-run:-

(a) New main gasifier air blower.

(b) Tapered nozzles were fitted to the gasifier and regenerator pressure tappings to increase the bleed velocity and inhibit blockage.

(c) All manometers were replaced by differential pressure gauges.

(d) Elimination of kerosene combustion from warm-up procedure.

New analytical system still has some problems to be ironed out before Run 11

With the exception of the Wostoff analyser which was retained as a back-up low SO_2 meter, the entire analytical system was replaced by new equipment supplied by Hartman and Braun. In addition a new stack gas sampling line was installed downstream of the point at which the regenerator tail gas is vented to the stack, at a position where the SO_2 content of the stack gas should be uniform over the cross sectional area of the stack. This was incorporated to enable a better sulphur balance to be obtained. A schematic diagram of the analytical system is shown in Figure 6.

ANALYTICAL SYSTEM (SCHEMATIC DIAGRAM)

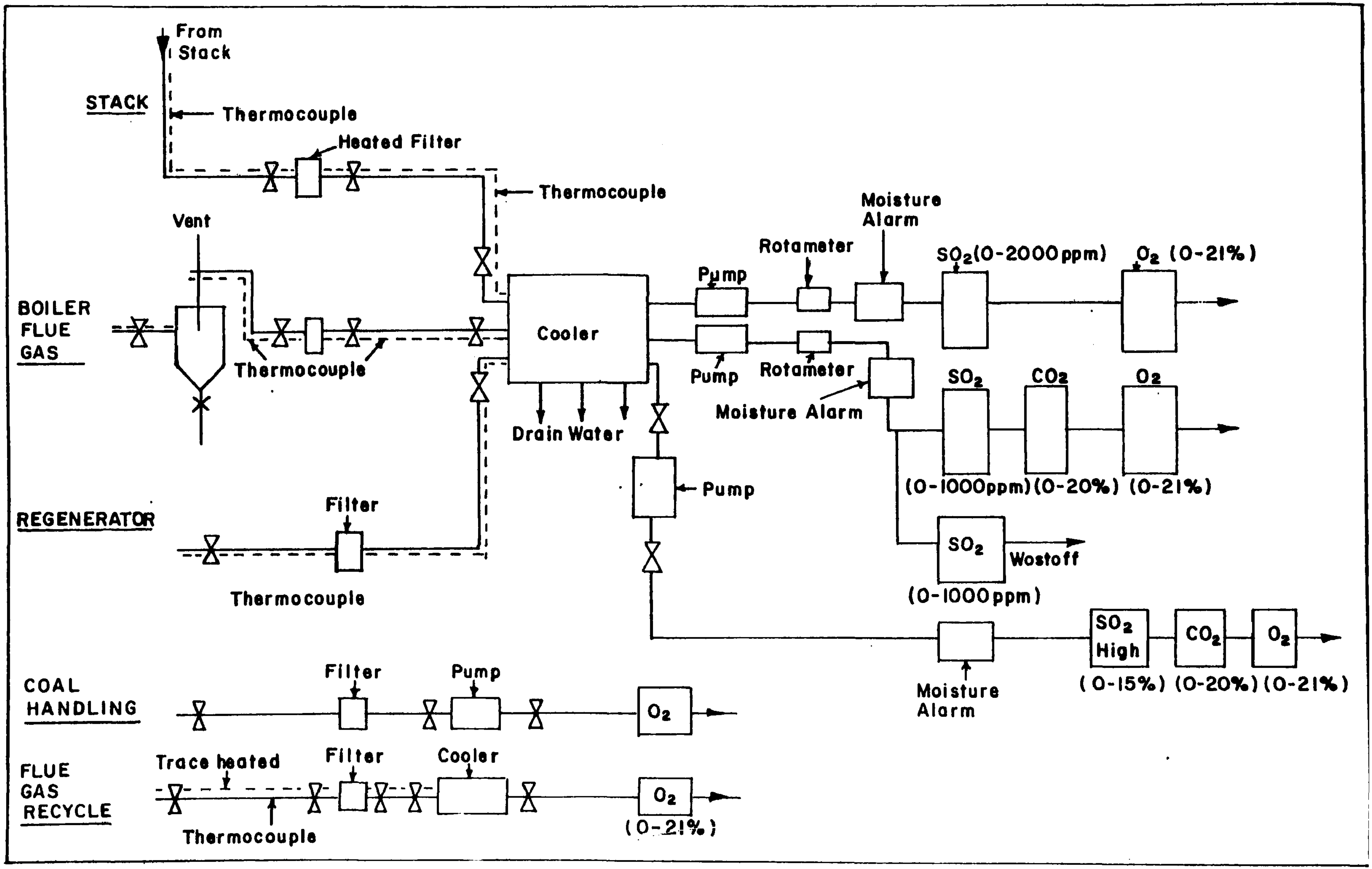

FIG. 6.

Both the Wostoff analyser and the Hartman and Braun infrared analyser simultaneously monitored the SO_2 emission from the boiler during the mini-run. Table 1 compares the averaged daily results obtained from both instruments. This shows the Wostoff consistently read higher than the Hartman and Braun analyser by approximately 10%. This would give a reasonably small difference in the measured % SRE of about 3% and therefore is acceptable. Part of the discrepancy between the two instruments may be due to the slow response of the Wostoff and for day running when readings were taken at short intervals of time the fast response infrared instrument is preferred.

TABLE 1

Comparison of Wostoff and Hartman and Braun SO_2 Analysers during mini-run

Run Day	No. of Readings	Average Hartman & Braun (ppm)	Average Wostoff (ppm)	Δ
1	5	367	392	25
2	10	309	363	54
3	9	287	316	29
MEAN	8	321	357	36

Boiler CO_2

Table 2 compares the analytical measurements of boiler CO_2 with the calculated values based on the fuel consumed. During days 2 and 3 the measured values were consistently lower than the calculated values both with oil and coal. In the case of coal the calculated values incorporate a correction factor for the percentage coal utilised (see page 20). Boiler CO_2 is a useful measurement particularly in the case of coal gasification since it provides a more direct route to measure utilization in the boiler than the calculation based on boiler O_2. However, because of the obvious inaccuracy of the measured values of boiler CO_2 they were not used in subsequent calculations. However, work is in hand in improve the reliability of this measurement before Run 11.

TABLE 2

Comparison of Measured and Calculated Values for Boiler CO_2

Run day	No. of Readings	Fuel	Average Measured CO_2 (Vol %)	Average Predicted CO_2 (Vol %)
1	2	Oil	11.7	10.2
2	5	Oil	6.6	9.0
2	4	Coal	6.9	10.3
3	5	Oil	7.3	9.0
3	4	Coal	5.9	8.4

Sulphur Balance (gaseous streams)

At equilibrium conditions with sulphur levels on both gasifier and regenerator bed material constant then the sulphur added to the system from the fuel should balance the sulphur found as SO_2 gas in the exit streams i.e. boiler flue and regenerator off-gas. In other words the sum of these two measurements expressed as a percentage of the feed sulphur should equal 100%. However, it is possible to operate the unit for short periods of time under conditions where sulphur is being stripped from the stone (i.e. giving values greater than 100%) or sulphur is being laid down on the stone (i.e. values lower than 100%).

During the mini-run this sulphur balance measurement could be calculated via three independent routes i.e.

(a) Stack SO_2 analysis (Method A)

(b) Boiler flue and regenerator off-gas analysis (Method B)

(c) Solids analysis for sulphur (Method C).

Table 3 compares the averaged sulphur balances obtained via all three routes. During days 1 and 2 when the regenerator was largely inoperative because of high carbon levels on the stone all three methods appear to give good agreement. However since the regenerator off-gas was low in SO_2 all this meant was that the analyses of the boiler flue and stack SO_2 were largely in agreement. During day 3, however, when the contribution to the sulphur balance by method B from the regenerator off-gas SO_2 analysis was high the agreement was poor particularly during coal gasification. Obviously the analytical measurement of regenerator SO_2 was spurious during the final day and this should be investigated before Run 11.

TABLE 3

Comparison of Sulphur Balances (in exit gases) by three Independent Routes

Run day	No. of Readings	Fuel	Sulphur Balance (% of Feed) Method A	Method B	Method C
1	3	Oil	45.5	42.5	-
2	5	Oil	46.0	49.8	-
2	3	Coal	56.6	60.3	66.6
3	4	Oil	146	85.3	105.8
3	4	Coal	291.6	142	150.4

IV COAL QUALITY AND SIZE

Lignite quality fed to Continuous Unit was fairly accurately defined but Illinois No.6 coal quality less so

Two different coals, Illinois No.6 and Texas Lignite were tested on the continuous unit. The Lignite chosen was that presently earmarked for use in the Texas Demonstration Plant and will also be studied exclusively on Run 11.

A major problem with any coal studies is the variability of coal quality. Consequently, in order to obtain representative coal analysis and hence realistic mass balances (see Appendix II) periodic sampling of the coal actually fed to the unit was carried out and sent for analysis. A complete listing of the sample analyses obtained is contained in Appendix I, Tables I and II while Table 4 below gives the average values for each coal.

TABLE 4

Analysis of Texas Lignite and Illinois No. 6 fed to continuous unit

	Illinois No. 6		Texas Lignite	
	ACTUAL %	DRY BASIS %	ACTUAL %	DRY BASIS %
Moisture	4.63	-	14.02	-
Ash	7.36	7.8	18.5	21.5
Carbon (corrected)	70.3	73.64	51.2	59.5
Hydrogen (corrected)	4.61	4.83	3.6	4.2
Sulphur (total)	1.78	1.9	0.82	0.95
Nitrogen	1.44	1.5	1.0	1.2
Oxygen + errors (by difference)	9.85	10.4	10.9	12.6
Gross calories/GM	6936	7260	5004	5817
Gross BTU/lb	12486	13068	9007	10470
CO_2 %	0.16	-	0.51	-

Compared with Illinois No.6, lignite has a higher moisture and ash content but a lower sulphur content. Actually the moisture content quoted (14 wt%) relates to the level fed to the unit while that analysed prior to grinding and some drying was 35%. Seven lignite samples and three Illinois No.6 coal samples were taken during the periods of operation and the standard deviation in the analysis results listed in Table 5 gives a measure of the uniformity in quality of the coals.

TABLE 5

Standard Deviation of Analysis Results from Lignite and Illinois No. 6

Measurement wt %	Lignite % Standard Deviation	Illinois No. 6 % Standard Deviation
Carbon	3.5	10.5
Hydrogen	4.1	8.7
Sulphur	15.5	20.0
Ash	5.3	51

In assessing the validity of any mass balances which are presented in this report the standard deviations presented in Table 5 are of paramount importance. Obviously in the case of lignite the quality is fairly accurately defined and reasonably precise mass balances should be possible. However, the numbers generated for gasification on Illinois No.6 should be assessed in the light of the low accuracy in defining the feed quality.

Several different coal feed size ranges have been tested on Mini-run

The gasification of coal introduces a new variable not relevant with fuel oil gasification namely feed particle size. Obviously, the more finely divided the coal the larger the surface area and hence the more rapid the gasification. On the other hand, small coal particles are likely to be elutriated more rapidly from the bed and hence the residence time available for gasification should be shorter. In practice, there may well be an optimum feed size for good gasification. Foster Wheeler have stated a preference for feeding coal ground to ½" down and one of the tasks in future coal runs will be to specify coal sizes acceptable for operation by the CAFB process. For the mini-run, coal was ground to 1/8" down and subsequently sieved into various size ranges. Table 6 lists the feed sizes used for both Lignite and Illinois No.6 coal. In future references to coal feed sizes in this report the general description given in Table 6 will be used.

TABLE 6

Coal Feed Particle Size Distribution (wt %)

		Sieve Size in Microns								
Coal	General Description	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
Lignite	800μ down	0	0	0	0	5.5	37.7	32.6	4.4	19.7
Lignite	1405μ down	0	0	0.1	2.6	9.5	46.8	15.6	5.1	20.3
Lignite	1/8" down	3.9	4.3	4.9	9.8	10.5	25.5	11.3	5.0	25
Illinois No.6	1405μ down	0.0	0.8	0.8	1.3	10.7	31.8	15.0	5.9	33.7

VIEW OF STACK DURING
LIGNITE GASIFICATION

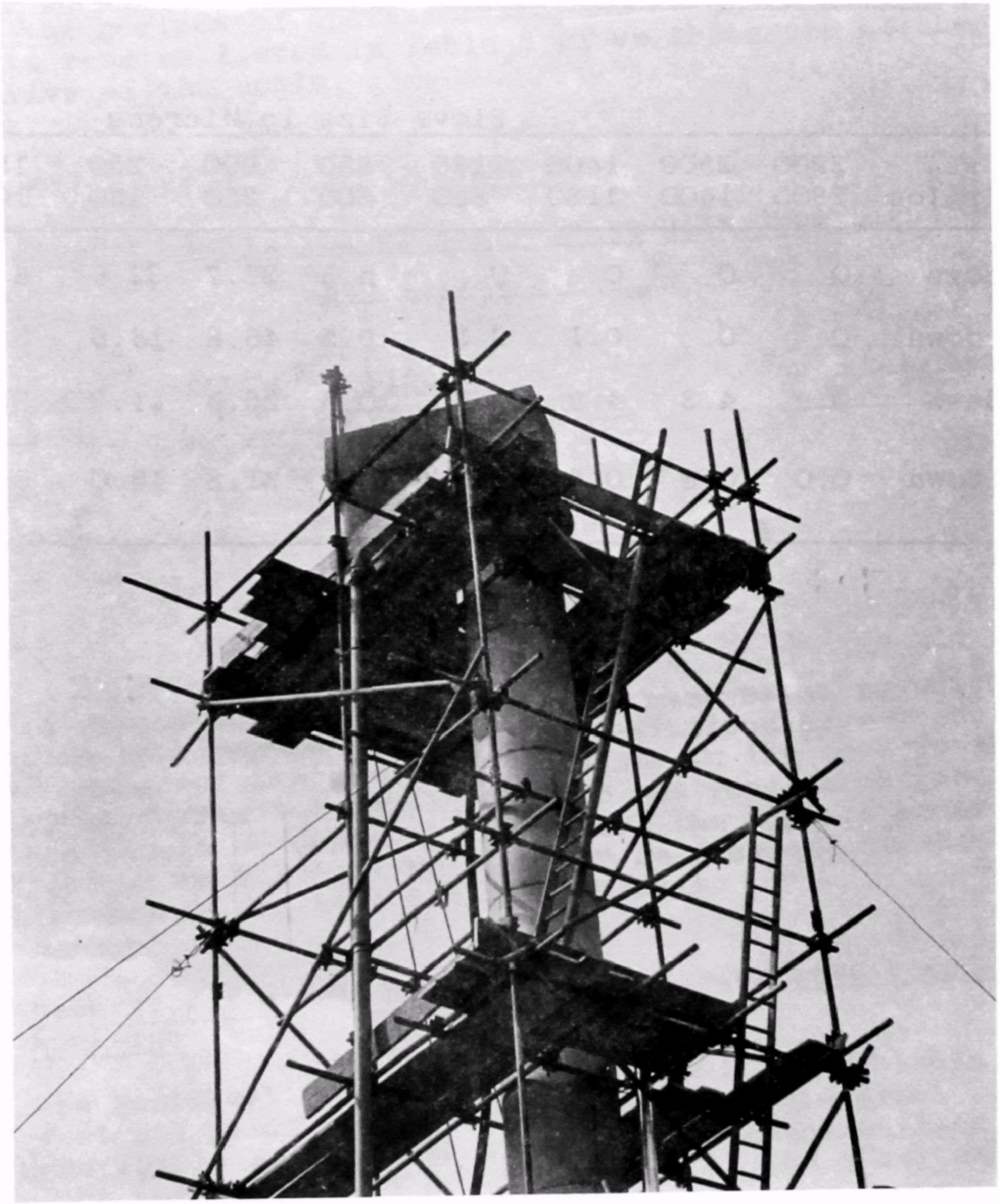

FIG. 7.

V RESULTS AND DISCUSSION

Both Texas Lignite and Illinois No.6 coal were successfully gasified on Continuous Unit

Two different coals, Illinois No.6 and Texas Lignite were gasified for the first time on the continuous unit during the mini-run. Initially the unit was gasified on heavy fuel oil and gradually changed over to coal gasification by slowly increasing the coal feed rate and backing off fuel oil. Subsequently, however, startup entirely on coal was found to be equally efficient and trouble-free and this method was used in later experiments.

Visual inspection of the flame through a viewing port at the back of the boiler showed a stable, yellowish, smoke-free flame indistinguishable from that produced from heavy fuel oil gasification. Figure 7 shows one of several photographs of the boiler stack taken during coal gasification which shows a clean flue gas.

Comparison of the gasifier gas quality obtained from heavy fuel oil and both of the coals tested are shown in Table 7. Apart from a tendency towards a higher proportion of simple organic molecules such as methane in the product gas from fuel oil and a lower quantity of hydrogen gas, the gases are remarkably similar.

TABLE 7

Comparison of Gasifier Gas Quality from Fuel Oil and Coal

Gasifier Fuel	Heavy Fuel Oil	Illinois No.6 Coal	Texas Lignite
Nitrogen + inerts	58.4	59.2	59.0
Carbon Monoxide	10.2	12.2	12.2
Carbon Dioxide	10.2	9.9	12.1
Methane	7.7	4.2	2.2
Ethylene	5.0	0.8	0.7
Ethane	0.1	0.1	0.1
Hydrogen	8.4	13.6	13.7

In total approximately 8½ hours of continuous gasification were carried out but quantitative information is confined to three periods of gasification totally slightly under five hours. In general readings were taken every 30 minutes and a complete listing of the important measurements is contained in Appendix III, Table I. In the present discussion the results of all three periods of gasification will be looked at together and discussed under four headings, gasification, desulphurisation, regeneration and combustion efficiency.

Gasifier conditions can be expected to be leaner and cooler on Lignite compared with fuel oil

Table 8 summarises the salient operating conditions for the three periods of gasification.

No attempt was made to explore different operating conditions and consequently the conditions for all three periods of gasification were relatively similar. The difference in gasifier temperature between Illinois No.6 and Lignite is due entirely to the cooling effect of the higher ash and moisture content for the Lignite feed. No fresh bed material was added during any of the periods of gasification which were carried out on separate days. The drop in bed depth shown in Table 8 is almost entirely due to bed material falling back through the gasifier distributor while the bed was slumped overnight and little or no change in bed depth was noted when gasifying on oil and a slight increase in bed depth on coal. The air/fuel ratio in the gasifier for both coals was similar and compares with typical figures for heavy fuel oil of 20-23% of stoichiometric combustion. This reflects the need in the case of coal for more oxidation and less cracking compared with fuel oil. While the effect of air/fuel ratio has still to be explored it is unlikely that the gasifier will be operated successfully on coal at air/fuel ratios significantly richer than 30% of stoichiometric combustion. This means that CAFB converted boiler dimensioned for fuel oil will be derated in the region of 70% on coal.

Desulphurising efficiency on coal can be expected to match that on fuel oil

Table 9 summarises the more detailed information on sulphur removal efficiency (% SRE) found in Appendix III, Table I. This compares the mean % SRE for the three gasification periods on coal with the expected result for heavy fuel oil under the same conditions. Since % SRE for fuel oil in the region 80-90% can be attained under specific operating conditions, the conditions the unit was run with coal were far from ideal from a desulphurising point of view. Nevertheless, the results do show that coal can be desulphurised as efficiently if not more efficiently than fuel oil under the same conditions.

Also included in Table 9 are the SO_2 emissions calculated in terms of lbs SO_2 per 10^6 BTU for each of the gasification periods. This shows the final period of gasification on 1/8" down Lignite was better than the New Source Performance Standard of 0.6 lbs SO_2 per 10^6 BTU (EPA).

Satisfactory regeneration while gasifying on Lignite was demonstrated by slowing bed transfer rate

In adapting the CAFB process to operate on coal the part of the process expected to cause most problems was the regeneration step. Two problems were anticipated:-

(1) At high carbon levels on the lime, regeneration does not take place because the oxygen in the air fed to the

TABLE 8

Summary of Gasifier Operating Conditions during Mini-run on Coal

Duration of Gasification (hours)	Coal (type & size)	Gasifier Temperature O°C	Gasifier bed depth (inches)	Gasifier bed velocity (ft/sec)	Fresh bed feed (lbs)	Air/Fuel Ratio (% Stoich)
1.0	Illinois No.6 (1405μ down)	970	42	3.7	0	30.3
2.25	Texas Lignite (800μ down & 1405 down)	896	39	3.5	0	32.2
1.5	Texas Lignite (1/8" down)	895	35	3.5	0	31.6

TABLE 9

Summary of Desulphurising Performance during Mini-run on Coal

Duration of Gasification (hours)	Coal (type & size)	Mean % S.R.E. on coal	Expected Result on oil (% SRE)	SO_2 Emission (lb SO_2 per 10^6 BTU)
1.0	Illinois No.6 (1405μ down)	74.8	73.8	0.73
2.25	Texas Lignite (800μ down & 1405μ down)	64.0	68.5	0.76
1.5	Texas Lignite (1/8" down)	82.4	67.2	0.42

regenerator preferentially attacks the carbon to the exclusion of the sulphur, resulting in an off-gas containing high concentrations of CO_2 and low SO_2.

(2) At the high operating temperatures necessary for regeneration (>1050°C) some fusion of ash particles from the coal feed may cause agglomeration to occur in the bed or alternatively coat the lime particles so as to make them inactive to regeneration.

It was expected that both these problems might dictate the maximum coal feed particle size in the gasifier since carbon and ash levels in the bed would be expected to be higher with a larger particle feed.

Table 10 summarises the regenerator operating conditions during the periods of gasification on coal. During the first two periods of gasification little attempt was made to bring the regenerator into action and consequently the high carbon level on the stone rendered the regenerator virtually inactive. In order to counteract the high carbon levels on the stone the bed transfer rates were reduced during the final period of gasification and effective regeneration was made possible. This result is particularly encouraging since it was achieved during gasification with the largest particle size tested. It implies that Lignite sized at 1/8" down may well be eminently suitable for the CAFB process and suggests the need to explore larger coal feed sizes in Run 11 in order to define the upper limit on coal feed size which still avoids problems in the regenerator.

Two independent techniques were used to measure the bed transfer rate:-

	Transfer Rate (lbs/hr)
(a) From sulphur balance in Gasifier	114.2
(b) From carbon balance in Gasifier	135.6

This should be compared with figures of 400-600 lbs/hour normally encountered with oil gasification. One consequence of the lower transfer rate is the need to operate with high sulphur levels on the bed material. For example the sulphur level in the gasifier during the period under discussion was 9.3 wt%. Another result of the low bed transfer was a high temperature in the regenerator and the possible need to cool either by flue gas or steam injection. This aspect of the problem needs to be further explored in Run 11.

<u>Target of 88% Lignite utilization by CAFB process was approached under conditions far from optimised</u>

Perhaps the most important question to be answered with coal gasification is how much coal is actually gasified and burned in the boiler. This is not such an easy question to answer as it first appears particularly when dealing with comparatively short

TABLE 10

Summary of Regenerator Operating Conditions during Mini-run on Coal

Duration of Run (hours)	Coal Type fed to gasifier	Regenerator Temperature °C	Regenerator Bed Depth (inches)	Regenerator Bed Velocity (ft/sec)	Regenerator Output (% of Feed)
1.0	Illinois No.6 (1405μ down)	1070	52	6.1	2.7
2.25	Texas Lignite (800μ down & 1405μ down)	1075	37	7.1	24.3
1.5	Texas Lignite (1/8" down)	1092	44	6.8	124.4

periods of gasification when full equilibrium conditions have not been established.

Provided the following information is available the weight of fuel combusted in the boiler can be calculated:-

(a) Total air fed to the boiler.

(b) Oxygen concentration in flue gas.

(c) Carbon/hydrogen ratio of fuel.

An alternative method involves the measurement of boiler CO_2 but these readings were shown to be inaccurate (see page 9) during the mini-run. In order to obtain an accurate measurement of the boiler air the unit was calibrated on fuel oil for several hours and the air supply left untouched on transfer to coal gasification.

The carbon/hydrogen ratio of the gasifier product gas was assumed to be identical to the carbon/hydrogen ratio of the coal as a first approximation. However, there was some indication from the product gas analysis shown in Table 7 that some hydrogen enrichment occurs. Nevertheless this should not significantly alter the conclusions regarding coal utilization and hence for the purposes of this report was ignored.

Table I in the Appendix III lists the percentage of the coal combusted in the boiler calculated in this way. Table II summarises the results for the three periods of gasification. Carbon balances presented in Appendix II show that some of the coal not combusted in the boiler can be accounted for by an increase in bed carbon in the gasifier and fines return system. The final column in Table II corrects for this and gives the % coal utilised during each of the periods of gasification. Unfortunately, in the case of gasification with Illinois No.6 coal the change in bed carbon was not recorded.

The overall % coal utilised for the two periods of Lignite gasification achieved under conditions which can be considered far from optimum was 87% which compares favourably with the 88% coal utilisation target set by Foster Wheeler for the Texas Demonstration plant run on Lignite.

<u>New Fines Return System reduced burden of lime into boiler by factor of ten but did little to reduce the 'fly-ash' during coal gasification</u>

Prior to each period of coal gasification the unit was run for several hours on fuel oil and Appendix III, Table II gives the results of these tests. Analysis of solids during oil gasification was also carried out and this is listed together with the coal data in Appendix IV.

TABLE 11

Average % Combustion Efficiency in Boiler during Coal Gasification

Duration of Gasification (hours)	Coal (type & size)	Average % Combustion Efficiency	Average % Coal Utilised
1.0	Illinois No.6 (1405μ down)	101.3	-
2.25	Texas Lignite (800μ & 1405μ down)	88.8	89.0
1.5	Texas Lignite (1/8" down)	78.8	84.9

Table V, Appendix II gives a solids inventory for the period of oil gasification on day 3. Throughout this period the material collected at the various exit streams was high in ash and carbon despite the fact that no ash was being added to the system. Table 12 gives the proportions of ash, carbon and lime (by difference) found at the various sources.

The only explanation for this is that these 'fly-ash' particles were generated during the previous day's gasification on coal and were being slowly elutriated from the gasifier bed. This implies that the carbon on these particles was virtually inert to oxidation in the gasifier. At 3.1120 the carbon content of the bed totalled 24 lbs and this was probably made up of carbon coating the lime particles together with discrete ash/carbon particles. Since at the end of the oil period 20 lbs of carbon was recovered as 'fly-ash' in the exit streams from the gasifier the implication is that little of the carbon was present as carbon on lime particles but that the majority was associated with the ash in the bed. This can be easily confirmed by analysing the various sieve fractions of the bed material during this period of operation. Scrutiny of the stone analysis from the regenerator bed particularly for the period of coal gasification on day 3 indicates that these 'fly-ash' particles although virtually inert to oxidation in the gasifier cannot be so in the regenerator.

Another interesting aspect of this period of oil gasification is the lime losses from the system. Table V, Appendix II shows that during this period the lime losses totalled 12.9 lbs for the 3 1/3 hour period i.e. 4 lbs/hour. This compares favourably with a figure of approximately 6 lbs/hour for a period of 68 hours gasification during Run 10 with no limestone make-up. The significant difference in this case however is that 90% of the lime is captured in the 2nd cyclone of the new fines return system (see Table V, Appendix II). Neglecting for a moment the 'fly-ash' generated

TABLE 12

Analysis of Solids from Exit Streams of Gasifier
(Period 3.1120-3.1440)

Source	Time	Carbon (wt %)	Ash (wt %)	Lime (wt %)
Stack KO	3.1140	48	46	4.7
	3.1340	51	44	3.7
Stack cyclone	3.1140	46	50	3.3
	3.1340	38	53	7.1
Boiler Back)	3.1140	53	38	7.9
& Sides)	3.1340	44	38	16.1
Regenerator)	3.1140	52	48	1.8
Cyclones)	3.1340	34	52	11.3
Fines Return)	3.1120	45	42	11.2
(2nd cyclone)	3.1340	28	27	41.7
)	3.1420	27	28	41.3

from the coal this means that under purely oil gasification conditions the dust burden into the boiler can be reduced by a factor of ten by incorporation of the new fines return system.

During the periods of coal gasification few of the 'fly-ash' particles were recovered in the 2nd cyclone of the fines return system. This was probably because a large quantity were not in fact captured by the gasifier cyclones and hence were not available to the fines return system. Obviously a move towards larger coal feed particle size will improve this situation but it may also be necessary to improve the efficiency of the gasifier cyclones perhaps by operating the unit with one cyclone instead of two as at present.

VI CONCLUSIONS

The conclusions which are presented below are entirely based on about 8½ hours of gasification on Texas Lignite and Illinois No.6 coal carried out on the 3/4 MWe continuous pilot plant and therefore may well be subject to some modification when more in-depth studies are carried out. Nevertheless the results do represent the best information available to date on coal gasification during the CAFB process and must be viewed in that light.

- No major structural changes in the existing pilot plant are necessary before Run 11. A new coal feed system similar to the existing design is considered desirable to allow both coal and limestone to be fed simultaneously to the unit if and when this is necessary.

- Both Texas Lignite and Illinois No.6coal can be gasified using the CAFB process to produce a clean, virtually sulphur-free gas which can be fired in a conventional gas boiler.

- Target of 88% Lignite utilization set by Foster Wheeler for the Texas Demonstration plant appears to be realistic. This figure was approached in the mini-run under conditions far from optimised.

- The desulphurising efficiency of both coals can be expected to match or exceed the performance obtained on fuel oil under similar operating conditions. Flue gas SO_2 emission levels better than the New Source Performance Standards of 0.6 lbs SO_2 per 10^6 BTU (E.P.A.) were achieved with 1/8" down Lignite.

- A CAFB conversion boiler dimensioned for heavy fuel oil may well need to be downrated by approximately 30% on coal. This arises from the higher air/fuel ratios necessary in the gasifier when operating on coal.

- Satisfactory regeneration of bed material while gasifying on Lignite can be achieved provided the transfer rate between gasifier and regenerator is comparatively slow. This mode of operation does not adversely affect gasification or desulphurisation but may result in the need to cool the regenerator for example by steam or flue gas injection.

- An upper limit on coal feed particle size has not been determined but gasification and regeneration on 1/8" down was successfully demonstrated.

- Intermittent operation of the pilot plant as opposed to round-the-clock operation was successfully demonstrated. The unit was successfully restarted after a shutdown of 3 days by injecting coal into the bed fluidised at 390°C.

- The new subsystem equipment incorporated since Run 10, in particular the valve-less fines return system, greatly improved operational reliability.

- The removal of fines from the 2nd cyclone of the new fines return system should reduce the burden of lime fines into the boiler by a factor of ten but modification either to the gasifier cyclones to improve their efficiency or the move to larger coal feed particle sizes may be necessary to reduce 'fly-ash' in the boiler.

VII FUTURE WORK

The work presented in this report represents only the start of a major exercise to discover the important features of coal gasification with the CAFB process. Listed below are only a few of the many avenues which need to be explored before this work can move into a commercial situation with any degree of confidence.

- Explore the effect on gasification, desulphurisation and regeneration of operating variables such as bed depth, bed velocity, bed temperature, air/fuel ratio, added water etc.

- Explore the effect of coal feed particle size and if possible demonstrate the feasibility of operation with ½" down Lignite.

- Determine the constraints of intermittent operation of the unit.

- Explore the limitations on regenerator performance of various carbon and ash levels on the bed material.

- Determine optimum conditions for maximum Lignite utilization in terms of boiler efficiency.

- Compare coal needle injection with injection directly into the distributor pit. Assess the suitability of self-bonded silicon carbide as injector material.

- Explore distributor nozzle design to prevent bed material falling back under slumped bed conditions.

- Determine contribution of fines return system to combustion efficiency and coal utilization. Explore methods of improving 'fly-ash' capture by the 2nd cyclone of the fines return system.

- Explore the possibility of coarser gasifier beds and the need for a fines return system at all with oil gasification.

VIII ACKNOWLEDGEMENTS

Work on the CAFB pilot plant is of necessity a team effort and without the contribution of each and everyone of the CAFB section the pages in this report would be blank.

Those not mentioned on the front cover of this report but who also made their contribution to the team effort were contractors R. Barbour and F. Rolph. Also the work carried out by Service Division i.e. workshop, electrical, analytical, site contractors is also gratefully acknowledged.

APPENDIX I

TABLE I

Analysis of Illinois No 6 Coal Feed to Continuous Unit

	A.R.	DRY	A.R.	DRY	A.R.	DRY	
Moisture	7.12	-	2.14	-	4.64	-	
Ash	10.42	11.22	3.18	3.25	8.48	8.89	
Carbon (Corrected)	63.56	68.44	78.19	79.9	69.21	72.58	
Hydrogen (Corrected)	4.21	4.53	5.02	5.13	4.61	4.83	
Sulphur (Total)	2.03	2.19	1.37	1.40	1.95	2.05	
Nitrogen	1.20	1.29	1.71	1.75	1.40	1.47	
Oxygen & Errors (By Difference)	11.46	12.33	8.39	8.57	9.71	10.18	
Gross Calories/GM.	6155	6625	7780	7950	6875	7205	
Gross BTU/lb	11,080	11,925	14,005	14310	12,375	12,970	
CO_2%	2.4	-	0.32	-	0.16	-	

NOTE: A.R. = As received

TABLE II

Analysis of Texas Lignite Feed to Continuous Unit

	A.R.	DRY	A.R.	DRY	A.R.	DRY	A.R.	DRY	A.R.	DRY	A.R.	DRY	A.R.	DRY
Moisture	14.15	–	15.78	–	11.68	–	16.70	–	11.38	–	14.85	–	13.59	–
Ash	19.72	22.97	17.03	20.22	17.43	19.73	19.16	23.0	19.13	21.59	18.75	22.02	18.38	21.27
Carbon (Corrected)	49.66	57.87	50.38	59.82	53.57	60.65	48.64	58.39	52.52	59.26	50.68	59.51	52.6	60.87
Hydrogen (Corrected)	3.53	4.11	3.59	4.26	3.80	4.30	3.39	4.07	3.73	4.21	3.66	4.30	3.45	3.99
Sulphur(Total)	0.80	0.93	0.83	0.99	0.88	1.00	0.76	0.91	0.81	0.91	0.84	0.99	0.81	0.94
Nitrogen	0.94	1.10	1.04	1.23	0.91	1.03	1.15	1.38	1.06	1.20	1.04	1.22	1.08	1.25
Oxygen &Errors (by Difference)	11.18	13.02	11.35	13.48	11.73	13.29	10.2	12.25	11.37	12.83	10.18	11.96	10.09	11.68
Gross Calories / GM	4980	5800	4865	5775	5210	5900	4710	5650	5120	5775	5030	5905	5110	5915
Gross BTU/lb	8965	10440	8755	10395	9380	10620	8480	10170	9215	10395	9055	10630	9200	10645
CO_2%	0.5	–	0.78	–	0.29	–	0.5	–	0.41	–	0.55	–	0.53	–

NOTE A.R. = As received

APPENDIX II

Mass Balances for Carbon, Sulphur and Ash

Table I lists the carbon and sulphur inventory for a section of the first period of lignite gasification with 800μ down and 1405μ down feed. The information is presented in terms of the weight percent of the feed, identified at the various sources. Considering the errors involved (see table 2 main report) the carbon and sulphur balances are satisfactory.

TABLE I

Carbon and Sulphur Inventory for Lignite Gasification (Period 2.1440 ~~- 2.1640~~)

Source	Carbon (% of Feed)	Sulphur (% of Feed)
Solids Inventory	6.8	66.6
Boiler Flue Gas	88.8	36
Regenerator Gas	3.9	24.3 (1)
Total	99.5	126.9

(1) From Stack SO_2 measurement

Feed:- Carbon 514.6 lbs; sulphur 8.33 lbs

A detailed breakdown of the solids inventory for carbon, sulphur and ash is presented in table II. Poor recovery of the acid insoluble ash is evident with the major proportion being recovered at the stack cyclone after the boiler.

TABLE II

Solids Inventory for Lignite Gasification (Period 2.1440 - 2.1640)

Source	Carbon (% of Feed)	Sulphur (% of Feed)	Acid Insoluble Ash (% of Feed)
Stack Knockout	1.0	1.1	6.1
Stack Cyclone	4.2	3.0	25.9
Boiler Back & Side	0.6	1.7	1.81
Regenerator Cyclone	0.1	0.4	0.6
Fines Return System change	- 0.9	3.6	- 0.1
2nd Cyclone (Fines Return)	0.9	3.0	3.5
Gasifier bed change	1.1	68.9	7.2
Regenerator bed change	- 0.2	- 15.1	0.3
Total	6.8	66.6	45.3

Feed:- Carbon 514.61bs; sulphur 8.331bs; Acid insoluble ash 143.7 lbs

Table III lists the carbon and sulphur inventory for a section of the second period of lignite gasification with 1/8" down feed.

TABLE III

Carbon and Sulphur Inventory for Lignite Gasification (Period 3.1540-3.1640)

Source	Carbon (% of Feed)	Sulphur (% of Feed)
Solids inventory	14.6	- 50.4
Boiler Flue gas	78.8	15.5
Regenerator Gas	1.8	126.5 (1)
Total	95.2	91.6

(1) From Stack SO_2 measurement

Feed:- carbon 238.1 lbs
sulphur 3.86 lbs

Again the balance for carbon and sulphur is acceptable. Table IV gives a detailed breakdown of the solids inventory for carbon, sulphur and ash. The larger coal feed size has resulted in a better ash recovery and a higher proportion found in the gasifier.

Table IV

Solids Inventory for Lignite Gasification (Period 3.1540 -3.1640)

Source	Carbon (% of Feed)	Sulphur (% of Feed)	Acid Insoluble Ash (% of Feed)
Stack Knockout	0.8	1.3	3.5
Stack Cyclone	3.4	3.9	19.0
Boiler Back & Side	4.3	5.2	11.7
Gasifier Bed change	5.3	-28.2	35.3
Regen Bed Change	- 0.5	-25.9	1.8
Fines Return System Change	1.3	-6.7	5.3
Total	14.6	-50.4	76.6

Feed:- Carbon 238.1 lbs; sulphur 3.86lbs; acid insoluble ash 66.5 lbs

TABLE V

Solids Inventory for Heavy Fuel Oil Gasification (Period 3.1120 - 3.1440)

Source	Carbon (lbs)	Sulphur (lbs)	Ash (lbs)	Lime (lbs)
Stack K.O.	3.0	0.08	2.7	0.28
Stack Cyclone	0.84	0.03	1.02	0.02
Boiler Back & Side	3.4	0.1	2.7	0.84
Regenerator Cyclone	0.6	0.03	0.8	0.12
Fines Return (2nd cyclone)	12.2	1.09	11.8	11.6
Fines Return System Change	0.35	0.24	-0.35	0
Gasifier bed Change	-12.0	-4.0	-	-
Regenerator Bed change	-0.7	1.05	-	-
Total	7.7	-1.38	18.7	12.9

Feed:- Carbon 784.8 lbs; sulphur 23.9 lbs; Ash 0lbs, lime 0 lbs

Appendix III (Table I)

Experimental Results of Mini-Run on Lignite & Illinois No 6 Coal

Time	1.1530 1.1630	2.1500	2.1600	2.1630	2.1655	3.1530	3.1600	3.1630	3.1645
Coal Type & Size	Illinois No.6 -800 µ	Lignite -800 µ	Lignite -1405 µ	Lignite -1405 µ	Lignite -1405 µ	Lignite - 1/8"	Lignite -1/8"	Lignite -1/8"	Lignite -1/8"
Gasifier Temperature °C	970	890	885	890	920	915	900	885	880
Gasifier Bed Depth (inches)	42	38	39	39	39	34	35	36	36
Gasifier Bed Velocity (ft/sec)	3.7	3.5	3.5	3.5	3.6	3.6	3.5	3.5	3.5
Gasifier Air/Fuel (% Stoich)	30.3	29.9	32.2	33.9	32.7	31.3	29.6	31.7	33.7
Caps Feed Mole Ratio	0	0	0	0	0	0	0	0	0
Regen Temp °C	1070	1065	1075	1080	1080	1080	1090	1095	1110
Regen Velocity (Ft/sec)	6.1	7.6	6.9	7.0	6.9	6.7	6.5	6.7	7.2
Regen CO_2 (Vol %)	12.4	17.6	17.0	16.8	16.4	6.0	5.6	10.0	10.0
Regen SO_2 (Vol%)	0.15	0.4	0.9	0.3	0.4	7.5	6.8	7.3	7.8
Regen S Output (% of Feed)	2.7	16.0	35.5	12.6	16.0	270.8	216.9	275.4	341.1
Regen Bed Depth (inches)	52	36	36	38	40	44	44	42	44
% S.R.E.	74.8	64.2	56.5	73.7	61.5	95.6	86.9	78.5	76.8
% SRE Expected On Oil	73.8	68.4	69.1	68.9	67.5	65.8	66.8	67.9	67.9
Δ	+1.0	-4.2	-12.6	+4.9	- 6.0	+29.8	+20.1	+9.4	+8.9
Stack Sulphur (% of Feed)	14.2	62.4	78.7	39.9	39.7	163.6	124.2	128.8	150
Sulphur Balance (% of Feed)	27.8	51.8	79.0	38.9	54.5	275.2	230	296.8	364.3
% Combustion in Boiler	101.3	84.3	84.0	89.1	97.7	73.6	72.9	79.1	89.7

Appendix III (Table II)

Experimental Results of Mini-test on Heavy Fuel Oil

Time	1.1220	1.1330	1.1435	2.1130	2.1200	2.1230	2.1330	2.1430
Gasifier Temperature (°C)	930	940	950	945	940	975	970	960
Gasifier Bed Depth (Inches)	41	41	42	39	38	40	38	38
Gasifier Bed Velocity (Ft/sec)	3.6	3.6	3.7	3.7	3.7	3.8	3.8	3.8
Gasifier Air/Fuel (% Stoich)	18.9	20.2	18.9	22.2	22.2	22.2	22.2	22.2
Ca/S Feed Mole Ratio	0	0	0	0	0	0	0	0
Regen Temperature (°C)	1040	1060	1055	1080	1079	1100	1070	1065
Regen Velocity (Ft/sec)	6.2	6.3	6.1	5.6	5.5	8.2	7.0	7.0
Regen CO_2 (Vol%)	18.0	15.2	12.6	18.0	17.4	18.0	14.6	16.2
Regen SO_2 (Vol%)	0.8	0.25	0.2	1.2	0.3	0.5	0.5	0.4
Regen S Output (% of Feed)	15	4.8	3.4	23.1	5.6	13.6	11.4	9.2
Regen Bed Depth (inches)	50	50	52	52	52	44	44	44
% S.R.E.	59.5	63.8	63.3	56.5	78.9	73.3	62.1	62.1
Predicted % S.R.E.	72.4	72.8	70.7	73.5	73.6	69.4	69.7	71.2
Δ	-12.9	-9.1	-7.4	-16.9	+3.7	+3.5	-7.6	- 9.0
Stack Sulphur (% of Feed)	52.9	39.7	35.0	74.0	47.4	37.4	49.8	40.5
Sulphur Balance (% of Feed)	55.5	41.0	40.0	66.6	26.8	40.3	49.2	47.0

Appendix III (Table II Cont'd)

Experimental Results of Mini-Test on Heavy Fuel Oil

Time	3.1030	3.1130	3.1230	3.1330	3.1430
Gasifier Temperature (°C)	955	950	930	940	940
Gasifier Bed Depth (Inches)	36	37	36	35	34
Gasifier Bed Velocity (Ft/sec)	3.7	3.6	5.2	5.2	5.4
Gasifier Air/Fuel (% Stoich)	20.5	20.5	24.1	23.9	24.5
Ca/S Feed Mole Ratio	0	0	0	0	0
Regen Temperature (°C)	1020	1075	1085	1080	1070
Regen Velocity (Ft/Sec)	6.6	7.0	6.9	6.9	6.6
Regen CO_2 (Vol%)	18.8	17.0	14.4	11.8	7.6
Regen SO_2 (Vol %)	0.8	3.0	3.8	5.7	6.5
Regen S Output (% of Feed)	18.2	68.3	84.5	125.3	132.7
Regen Bed Depth (Inches)	48	40	44	44	44
% S.R.E.	59.8	57.6	60.4	53.8	55.7
Predicted % SRE	69.4	70.4	75.2	73.8	73.8
Δ	-9.6	-12.8	-14.8	-19.9	-18.1
Stack Sulphur (% Feed)	44.6	68.8	75.2	87.9	109.1
Sulphur Balance (% of Feed	58.4	110.6	124	171.5	177.0

Appendix IV (Table I)

Stone Analyses of Gasifier Bed Material

Time	Sulphur (wt%)	Carbon (wt%)	Acid insoluble Ash (wt%)	Sulphur as SO_4 wt%
1.1440	4.22	5.6	3.2	< 0.1
1.1540	4.74	6.7	2.8	0.1
2.1400	6.86	2.6	4.1	0.1
2.1500	7.67	2.6	4.2	0.1
2.1600	8.29	2.5	5.6	0.1
2.1700	8.12	3.6	5.1	0.1
3.1120	9.1	2.8	3.6	< 0.1
3.1240	10.2	3.1	3.2	< 0.1
3.1340	9.5	2.5	3.3	0.1
3.1440	9.24	1.4	4.6	< 0.1
3.1540	9.36	1.9	3.7	< 0.1
3.1640	9.23	3.4	6.5	< 0.1

Appendix IV (Table II)

Stone Analyses of Regenerator Bed Material

Time	Sulphur (wt%)	Carbon (wt%)	Acid Insoluble Ash (Wt%)	Sulphur as SO_4 (wt%)
1.1440	4.28	1.1	3.4	0.1
1.1540	4.6	0.8	3.5	<0.1
2.1400	7.8	1.6	3.4	0.1
2.1500	9.05	1.8	3.9	0.1
2.1600	7.36	0.9	3.6	0.1
2.1700	7.42	0.9	4.3	0.1
3.1120	8.2	1.0	3.3	0.1
3.1240	7.9	0.8	5.4	0.3
3.1340	8.5	1.4	3.2	0.1
3.1440	7.31	0.4	3.9	0.3
3.1540	6.22	1.1	3.1	0.9
3.1640	5.45	0.2	4.0	0.9

Appendix IV (Table III)

Stone Analyses from 1st Cyclone Fines Return System

Time	Sulphur (wt%)	Carbon (wt%)	Acid Insoluble Ash (wt%)	Sulphur (as SO_4) wt %
1.1520	3.43	41	14	<0.1
1.1540	2.2	54	22	0.1
1.1620	1.32	37	45	0.1
2.1340	3.74	29	22	0.1
2.1420	1.48	56	35	0.1
2.1520	1.57	45	41	0.1
2.1540	1.19	61	39	<0.1
2.1620	4.35	23	28	0.1
2.1700	2.06	36	41	0.1
3.1120	1.75	46	38	<0.1
3.1340	4.41	26	19	<0.1
3.1420	3.12	48	36	<0.1
3.1500	3.82	27	27	0.1
3.1540	2.68	28	28	<0.1
3.1620	1.19	46	48	0.1

Appendix IV (Table IV)

Stone Analyses from 2nd Cyclone Fines Return System

Time	Sulphur Wt %	Carbon wt%	Acid Insoluble Ash wt%	Sulphur as SO_4 wt%
1.1520	2.22	49	21	0.1
1.1540	2.47	57	15	0.1
1.1640	2.18	61	25	0.1
2.1340	3.06	36	22	0.1
2.1420	2.03	45	34	0.2
2.1520	2.36	51	34	0.1
2.1540	1.68	46	38	<0.1
2.1620	2.09	32	47	0.1
2.1700	2.2	40	44	<0.1
3.1120	1.8	45	42	<0.1
3.1340	3.33	28	27	<0.1
3.1420	3.72	27	28	<0.1
3.1500	3.29	31	27	<0.1
3.1520	3.47	36	24	<0.1
3.1600	2.77	21	24	0.2

Appendix IV (Table V)

Stone Analyses from Stack Knockout, Stack Cyclone and Boiler Back & Sides

Source	Time	Sulphur wt %	Carbon wt%	Acid Insoluble Ash wt%	Sulphur as SO_4 wt %
Stack KO	1.1500	1.59	59	40	0.7
"	1.1700	1.09	64	36	0.2
"	2.1330	1.35	84	13	0.3
"	2.1640	0.65	38	62	0.1
"	3.1140	1.3	48	46	0.1
"	3.1340	1.35	51	44	0.5
"	3.1620	0.95	45	52	0.3
Stack Cyclone	1.1500	2.46	45	25	1.2
"	1.1700	1.09	57	40	0.3
"	2.1330	1.7	66	30	0.3
"	2.1640	0.4	36	64	0.2
"	3.1140	0.75	46	50	<0.1
"	3.1340	1.93	38	53	1.0
"	3.1620	0.71	39	60	0.2
Boiler Back	1.1500	1.93	41	31	0.5
& Sides	1.1700	0.98	53	41	0.1
"	2.1330	1.33	68	29	0.2
"	2.1640	1.9	43	37	0.3
"	3.1140	1.13	53	38	0.1
"	3.1340	1.94	44	38	0.3
"	3.1620	1.05	57	43	0.1

Appendix IV (Table VI)

Stone Analyses from Regenerator Cyclone

Time	Sulphur wt %	Carbon wt%	Acid Insoluble Ash (wt%)	Sulphur as SO_4 wt%
1.1500	3.98	37	22	0.6
1.1700	2.23	63	31	0.4
2.1330	3.51	43	32	-
2.1640	1.7	40	54	0.3
3.1140	1.78	52	48	0.1
3.1340	2.68	34	52	0.7
3.1620	2.36	20	60	1.4

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/7-77-027	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE First Trials of CAFB Pilot Plant on Coal		5. REPORT DATE March 1977
		6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) D. Lyon		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Esso Research Centre Abingdon, Oxfordshire 0X13 6AE England		10. PROGRAM ELEMENT NO. EHE623A
		11. CONTRACT/GRANT NO. 68-02-2159
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Task Final; 6-8/76
		14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES IERL-RTP Project Officer for this report is S. L. Rakes, Mail Drop 61, 919/549-8411 Ext 2825.

16. ABSTRACT The report gives results of a minirun, carried out on a 0.75-MWe continuous, chemically active fluidized-bed (CAFB) pilot plant during July-August 1976, as part of a program to extend the CAFB process to operate on coal. After 8.5 hours of gasification on Texas lignite and Illinois No. 6 coal, no major barriers were identified. The quality of the gas produced was similar to, and the desulfurizing efficiency on coal appeared to match or exceed, that for oil. The target of 88% lignite utilization was approached in the minirun under conditions which were far from optimum. Because of the need for more air to gasify coal, a CAFB unit dimensioned for fuel oil will probably have an energy capacity 30% lower when on coal. Satisfactory regeneration while gasifying on lignite was demonstrated, but control of regenerator temperature was more difficult. A new fines return system worked well, but did little to reduce the fly-ash level in the boiler during coal gasification. Operation of the pilot plant on an intermittent, as opposed to round-the-clock, basis was successfully demonstrated.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution	Regeneration (Engineering)	Air Pollution Control	13B
Combustion		Stationary Sources	21B
Coal	Fly Ash	CAFB	21D
Fluidized Bed Processing		Particulate	13H,07A
Gasification	Fines		
Desulfurization			07D

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS (This Report) Unclassified	21. NO. OF PAGES 44
	20. SECURITY CLASS (This page) Unclassified	22. PRICE

EPA Form 2220-1 (9-73)