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CHEMICALLY ACTIVE FLUID-BED PROCESS FOR SULPHUR REMOVAL DURING GASIFICATION OF HEAVY FUEL OIL - SECOND PHASE

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CHEMICALLY ACTIVE FLUID-BED PROCESS FOR SULPHUR REMOVAL DURING GASIFICATION OF HEAVY FUEL OIL - SECOND PHASE

by

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ABSTRACT

This report describes the second phase of studies on the CAFB process for desulphurising gasification of heavy fuel oil in a bed of hot lime.

The first test of the continuous pilot plant with U.S. limestone BCR 1691 was hampered by local stone sintering and severe production of a sticky dust during start up conditions. Batch tests confirmed that BCR 1691 produced more dust than either of the higher purity Denbighshire or U.S. BCR 1359 stones. With BCR 1691, dust production rate was tenfold higher during kerosene combustion at 870 deg. C than during gasification/regeneration cycles.

Modifications were made to the continuous pilot plant to improve operability and three more runs were made using BCR 1359, BCR 1691 and Denbighshire stone totalling 1167 hrs. In the final run 211 hours of uninterrupted gasification were achieved.

Improvements in gas analysis techniques allowed good material balances on process streams, including sulphur. Maximum sulphur removal efficiency under lined out conditions was 84%. This was limited by a maximum attainable bed depth of 61 cm (24 inches). Results indicate improved sulphur removal efficiency with deeper beds.

An engineering scoping study estimates that total CAFB development through a large demonstration test will take about 6-7 years and require $3,320,000 in engineering effort.

This report was submitted as a requirement of Contract No. 68-02-0300 by Esso Research Centre, England under the sponsorship of the Environmental Protection Agency.

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ACKNOWLEDGMENTS

The support of Esso Petroleum Company Limited is acknowledged for the construction of a 2930 kW (10 million BTU/hr) Chemically Active Fluid Bed Gasifier facility at the Esso Research Centre, Abingdon, Berkshire, England, and for its use in the generation of continuous gasification data for this project.

The authors would also like to thank Mr. O.R. Priestnall, Mr. J. Buzzacott, Mr. D. Storms, Mr. J. Cocker and Mr. A. Brimble for their strenuous efforts in generating the experimental data on which this report is based.

Finally, they acknowledge the assistance given by other Esso Petroleum Company personnel in operating the pilot plant, in maintenance of equipment, and in chemical analysis of test samples.

SECTION 1

CONCLUSIONS

TASK I

1. Stability and quality of continuous pilot plant operations were greatly improved by modifications which included uninterrupted limestone addition, regenerator back pressure control, flue gas recycle scrubbing, nitrogen quench for regenerator over temperature protection, and a positive pressure pilot flame for the main burner.

2. There are indications that prior to Run 6 the SO_2 content of the flue gas was dependent on its dust content. This effect was eliminated in Runs 6 and 7.

3. No significant difference was found between the sulphur removing abilities of BCR 1691 and Denbighshire stone at comparable operating conditions in the continuous pilot plant. There were however indications that BCR 1359 gave a better performance. Pure stones such as Denbighshire and BCR 1359 are preferred to impure stones such as BCR 1691 because they make less dust and are less likely to sinter.

4. The effect of lime replacement rate on sulphur removal efficiency diminishes rapidly as the Ca/S ratio is increased above 1.0 and even when the Ca/S ratio is reduced to 0.5 an S.R.E. in the region of 70% is obtainable at a bed depth of 53 cm.

5. Increasing the gasifier bed depth does appear to improve S.R.E. and the indications are that S.R.E's better than 85% should be obtainable at bed depths greater than 70 cms.

6. S.R.E. appears to depend on the sulphur content of the bed material which for good results should be less than 4% by weight in the gasifier bed.

7. On the limited evidence of Run 7, stone size appears to have very little effect on S.R.E. On this evidence there is no clear advantage in using a finer material than is required to ensure adequate fluidisation at the optimum superficial gas velocity within the gasifier bed.

8. The temperature of the gasifier bed appears to have a relatively minor effect on S.R.E. within the range 870 - 920°C.

9. Although poor sulphur balances were obtained during Run 5, Runs 6 and 7 gave good sulphur balances which were within the calculated margin of error.

10. Regenerator performance in the continuous pilot plant is sensitive to slight changes in reactor geometry. Subsequent to Run 5 the distributor of the regenerator was lowered by 10 cms in order to allow fluidising air to be heated before meeting the incoming stone. This resulted in a considerable improvement in regenerator selectivity and an increase in SO_2 concentration in the regenerator off-gas.

11. During Runs 6 and 7 the air/fuel ratio at which the gasifier was operated was always within the range 20 - 23% of stoichiometric. Within this range variations in air/fuel ratio have no obvious effects on S.R.E.

12. Combustion of CAFB gasifier product in the pilot plant burner produces less nitrogen oxides (166 cm^3/m^3 (p.p.m.) average) than direct combustion of fuel oil (263 cm^3/m^3 average).

13. During Run 5, 36% of the sodium in the fuel, 75% of the nickel and virtually 100% of the vanadium was bound by the bed material.

14. Self bonded silicon carbide is quite unaffected by the conditions within the gasifier and gives very satisfactory service. It has been used for the construction of the gas outlet pipes from the gasifier cyclones.

15. Coke laydown at the cyclone entrances and within the cyclone barrels is the factor which limits both the duration of a continuous run of the unit under gasifying conditions and the extent to which bed material is retained by the cyclones. The coke can be removed by a simple burn-out procedure and the longest period of operation between such burn-outs, in the runs covered by this report, was of 211 hours duration. In a large scale multi-cyclone unit it should be possible to burn-out the gas ducts in rotation whilst the unit is on stream.

16. Stainless steel cyclone liners are not satisfactory as a means for providing a smooth cyclone surface. In the pilot plant test they failed under decoking conditions and provided a surface for increased carbon deposition.

17. The heat release by fuel partial combustion in the gasifier is approximately 7211 J/kg (3,100 BTU/lb) at 20% of stoichiometric air based on the fraction of carbon and hydrogen oxidised and the amount of CO produced.

TASK II

1. Lime from BCR 1691 stone produces much more dust under comparable CAFB fluidisation conditions than either of the higher purity limes: BCR 1359 and Denbighshire.

2. Dust production with BCR 1691 lime is particularly severe during combustion with kerosene at 870 deg. C, the normal CAFB pilot plant start-up condition.

3. Kerosene combustion at 1050 deg. C causes less dust production than combustion at 870 deg. C both with BCR 1691 and with Denbighshire lime.

4. The fine dust produced during 870 deg. C combustion with BCR 1691 lime is sticky in nature and clings to pipe walls and cyclone internals unless mechanical force such as rapping is employed to dislodge it. Under gasification conditions the dust is not sticky, and is produced at a lower rate.

5. Pfizer calcite decrepitates during gasification/ regeneration cycling. This results in unacceptably high dust losses.

6. There is little difference in the desulphurising performance of all four stones which were tested under batch operating conditions although Denbighshire stone was marginally the best.

7. Carbon deposition on the stone appears to be related to the Conradson carbon quality of the fuel oil. The heavier fuel oils which were tested had Conradson carbon values of 17% and 33% and gave higher rates of carbon deposition than Amuay fuel oil (11% Conradson carbon) which was used for the bulk of the work.

8. When the 33% Conradson Carbon fuel was used it was not possible to bring down the carbon content of the bed material to a level which would allow satisfactory continuous regeneration even with air fuel ratios as

high as 32% of stoichiometric. It follows that such fuels will require a steam/air mixture for satisfactory gasification, unless a large gasifier fitted with heat exchangers is used.

TASK III

An engineering scoping study by Esso Engineering indicates that total CAFB development through a 100 + MW demonstration test period is expected to take about 6½ years and require $3,320,000 in engineering effort. Optimistically the development time might be reduced to 4-½ years with a cost of $2,520,000, but risks associated with the large unit would be correspondingly increased.

SECTION II

RECOMMENDATIONS

1. Further work is required in order to clearly establish the effects of the most important variables which have emerged from this study.

   1, Bed Depth
   2, Bed Sulphur Content.

2. Only a limited range of superficial gas velocities has been used in the work reported here and the effect of raising the gas velocity to 1.83 m/sec (6 ft/sec) should be explored.

3. A satisfactory cyclone fines drain and return system should be installed prior to further tests.

4. Methods should be developed for the control of carbon deposition in the gas ducts or alternatively for the removal of deposited carbon under running conditions.

5. In all C.A.F.B. installations particular care must be taken with the sampling of the boiler flue gas in order to avoid errors in the measurement of SO_2 concentration. A high velocity hot flue gas system using a hot cyclone and filter has given satisfactory results and is recommended.

6. It is important to measure the dust producing characteristics of candidate stones, especially under start-up and hot standby conditions.

7. Attention should be paid to the possibility of minimising stone consumption.

8. Attention should be paid to the effect of stone replacement rate on the amount and composition of stack dust emissions.

9. More evidence is required concerning the effect of stone size on S.R.E. and it is recommended that future tests should be planned to provide this information.

10. An emergency regenerator quench system should be included in CAFB installations to prevent sintering and agglomeration by temperature upsets.

11. Regenerator operation should be tested at lower air rates to confirm if reduced CaS conversion level will improve selectivity of CaS oxidation to CaO and reduce the quantity of $CaSO_4$ returned to the gasifier.

12. Tests should be made to establish the effectiveness of steam in gasifying carbon laid down on the gasifier bed material by heavy fuel oil.

SECTION III

INTRODUCTION

GENERAL

The Chemically Active Fluid Bed process is a means of avoiding sulphur oxide pollution while using heavy fuel oil for production of power. The process uses a fluidised bed of lime particles to convert the oil into a hot, low sulphur gas ready for combustion in an adjacent boiler. Sulphur from the fuel is absorbed by the lime which can be regenerated for reuse. During lime regeneration the sulphur is liberated as a concentrated stream of SO_2 which may be converted to acid or elemental sulphur.

Exploratory work on the CAFB began at the Esso Research Centre, Abingdon (ERCA) in 1966. In 1969 a six-phase programme of work was prepared to take the CAFB process from the laboratory stage through to a demonstration of the process on a 50 to 100 megawatt power generation boiler located in the United States. A summary of this six phase programme is shown in Figure 1. Phase I studies at Esso Research Centre were funded under Contract CPA 70-46 in June 1970, and consisted of batch reactor fuel and limestone screening studies, a variable study with U.S. limestone BCR 1691, and initial operation of a pilot plant incorporating continuous gasification and regeneration. The results of these studies were described in the final report (Reference 1) for that contract, dated June 1972.

This report covers work on the second phase of studies carried out in the period July 1, 1972 through May, 1974.

GASIFIER CHEMISTRY

When heavy fuel oil is injected into a bed of fluidised lime under reducing conditions at about 900 deg. C, it vaporises, cracks, and forms a series of compounds ranging from H_2 and CH_4 through heavy hydrocarbons to coke. The sulphur contained in the oil forms compounds such as H_2S, COS and CS_2 with H_2S predominating. The sulphur compounds react with CaO to form CaS and gaseous oxides.

For example:

$$CaO + H_2S \longrightarrow CaS + H_2O$$

OVERALL PROGRAMME OF WORK TO ACHIEVE CONVERSION OF A 50 TO 100 MW POWER GENERATION BOILER TO C A F B OPERATION

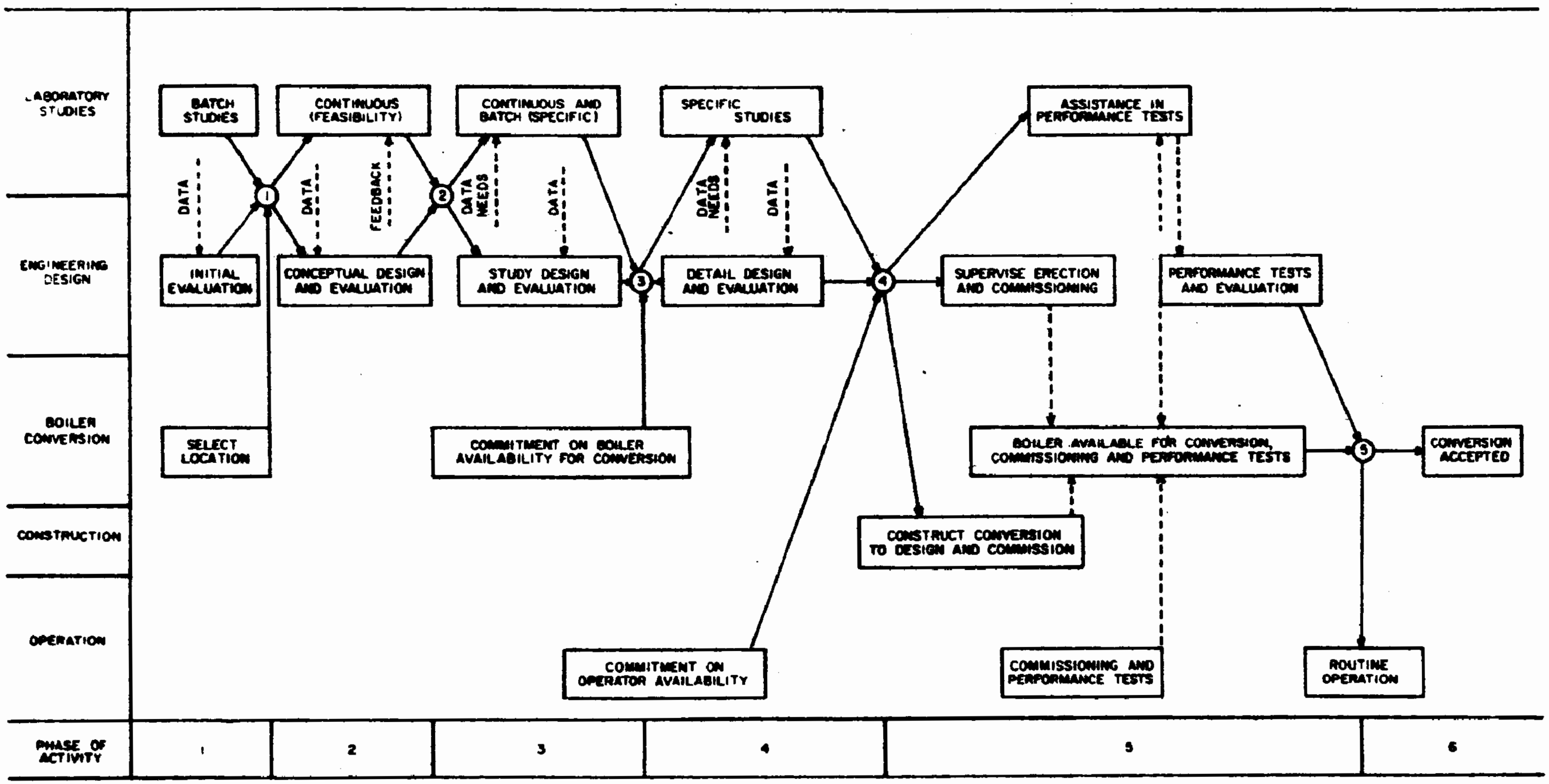

FIG. 1

The equilibrium for this reaction is far to the right. With a fuel containing 4% sulphur the equilibrium permits a desulphurising efficiency greater than 90% up to 1100 deg. C. Other factors however limit gasification temperature to the range of about 850 to 900 deg. C where the equilibrium sulphur removal would be about 99% (see Reference 1).

In the shallow fluidised bed of the gasifier there is a rapid circulation of lime between top and bottom. Indications are that coke is laid down on the lime in the upper portion of the fluid bed by oil cracking and coking reactions and that this coke burns off in the lower portion where oxygen is supplied by the air distributor.

Gasification conditions of temperature and air-fuel ratio must be chosen to maintain a balance between the rate of coke and carbon deposition and the rate of carbon burnoff. Broadly, this balance is met at gasification temperatures in the range of 850 to 900 deg. C and air-fuel ratios around 20% of stoichiometric. Lower air fuel ratios are operable at the upper end of the temperature range, and higher air-fuel ratios are needed as temperature is reduced.

Much of the oxygen entering the gasifier is consumed in oxidising coke to CO and CO_2 near the distributor. Of course, some enters other regions of the bed where it reacts with H_2 and hydrocarbons to form water and more carbon oxides. The final product from the gasifier is a hot combustible gas containing H_2 hydrocarbons CO, CO_2, H_2O, and N_2. Most of the energy released by partial combustion of the fuel is retained by this gas as sensible heat.

Only a portion of the CaO in the lime is reacted on each pass of solids through the gasifier. Good sulphur absorption reactivity has been obtained with up to 20% of calcium reacted in single cycle batch reactor tests, but in the continuous unit, the average extent of calcium conversion to sulphide is held to less than 10%.

When a single batch of lime is cycled between gasification and regeneration conditions it gradually loses activity for sulphur absorption. The activity of the bed can be maintained if some of the lime is purged each cycle and replaced by fresh material. Reactivity of the bed is therefore a function of the lime replacement rate. The replacement lime is usually added to the gasifier as limestone which calcines in situ.

Vanadium from the fuel oil deposits on the lime during gasification. Experimental evidence is that practically all of the fuel vanadium can remain fixed with the lime.

REGENERATOR CHEMISTRY

Calcium sulphide is regenerated to Calcium oxide by air oxidation.

$$CaS + 3/2\ O_2 \longrightarrow CaO + SO_2$$

$$\Delta H = -458.1\ kJ/mol$$

A competing reaction also consumes oxygen and forms calcium sulphate.

$$CaS + 2\ O_2 \longrightarrow CaSO_4$$

$$\Delta H = -921.3\ kJ/mol$$

Both reactions are strongly exothermic. A third reaction between the solid species is also possible.

$$CaS + 3CaSO_4 \longrightarrow 4CaO + 4SO_2$$

$$\Delta H = 926.8\ kJ/mol$$

This reaction is strongly endothermic.

The equilibrium constants (Appendix A, Reference 1) for these reactions determine the maximum partial pressure of SO_2 which can exist in equilibrium with mixtures of CaS, CaO, and $CaSO_4$ at any given temperature. These equilibria also determine a relationship between regenerator temperature and the maximum theoretical selectivity of oxidation of CaS to CaO.

At low oxidation temperature, the equilibrium SO_2 partial pressure is too low to permit all the oxygen supplied to leave in the form of SO_2. The excess oxygen then goes to form $CaSO_4$. Experimental oxidation selectivities are lower than the theoretical maximum, probably because of contacting and kinetic factors.

Since each sulphided lime particle passes through a range of temperatures and oxygen concentrations during its transit through the regenerator, it is exposed, on average, to less favourable selectivity conditions than those at the top of the bed.

Calcium sulphide oxidation selectivities to CaO of 70 to 80% and regenerator SO_2 concentrations of 8 to 10% have been achieved in pilot plant operations at regenerator temperatures in the range of 1040 to 1070 deg. C.

During the conversion of CaS to CaO and $CaSO_4$ there is evidence for existence of a transient liquid state (Reference 2). If air is passed through a hot static bed containing CaS, some of the particles will agglomerate into lumps during the regeneration reaction. Agglomeration does not occur if the bed is vigorously fluidised.

PREVIOUS EXPERIMENTAL WORK

A basis for the CAFB process had been established by experiments in 17.8 cm (7-inch) i.d. batch reactors at ERCA prior to 1970.

During 1970, two new batch reactor units were constructed for the OAP contract. Work in these batch units established the suitability for CAFB of a Venezuelan fuel oil available in the U.S., and selected the better of two U.S. limestones suggested by OAP. Both of the U.S. stones, BCR 1690 and BCR 1691, were lower in CaO content than the U.K. stones tested previously. In cycle tests the BCR 1691 stone gave sulphur removal activity comparable to that of the higher purity U.K. stones at equal Ca/S ratios. The BCR 1690 stone was found to be unsuitable in three respects. It gave lower sulphur recovery at equal Ca/S ratio; it attrited badly; and it sintered and agglomerated during regeneration. The BCR 1691 stone therefore was selected for further study. During 1971 an intensive study of gasification variables was conducted in the batch reactors with this stone. Tests with fresh beds screened the effects of major variables including air fuel ratio, gasification temperature, bed depth, lime particle size, and gas velocity in the bed. The variables of lime replacement rate and extent of calcium reaction between regenerations were probed in cyclic tests where the lime was cycled between gasifying and regeneration conditions in the batch reactors.

These studies provided the basis for a number of guidelines and process correlations. The effects of bed depth (25 to 51 cm) (10 to 20 inches) and fluidisation velocity (1.22 to 2.44 m/sec) (4 to 8 ft/sec) were correlated as gas residence time in the bed, giving an approximately first order sulphur removal rate expression. Sulphur differential, the quantity of sulphur to which the lime was exposed in each gasification cycle, emerged as an important variable. As an approximation lime reactivity varied inversely with the square of this differential and increased directly with lime to sulphur replacement ratio.

In parallel with the batch unit experiments, ERCA constructed a pilot plant in which the gasification and regeneration reactions could be studied under continuous operating conditions. A 2930 kW (10 million Btu/hr) water cooled boiler was included in the system to burn the gasifier product and dispose of the heat. Three tests designated CAFB Runs 1, 2 and 3 were made in this pilot plant during 1971. Run 3 lasted 230 hours of which 204 were at gasifying conditions. Denbighshire limestone (UK) was used throughout these runs together with Venezuelan fuel oil containing 2.5% sulphur. The pilot plant successfully demonstrated many features of the process including sulphur removal, lime regeneration, temperature control, start up, shut down, solids circulation, and release of the sulphur as a rich (8-10%) stream of SO_2. It also pinpointed areas for improvement which included reduction or elimination of carbonaceous deposits in cyclones and gas transfer ducts, minimisation of fines production and losses into the boiler, and improvement of regenerator oxidation selectivity.

WORK OBJECTIVES

Work on this contract constituted the second phase of a six phase programme to demonstrate and evaluate the CAFB gasifier on a commercial scale with a power plant boiler. Work on this phase consisted of three tasks.

Task I was operation of the CAFB pilot plant with the following set of objectives.

a. Verify that continuous gasification, sulphur and metal removal and lime regeneration results are as batch studies have indicated and evaluate the effects of bed depth, velocity, fuel/air ratio, lime make-up and fuel rate.

b. Determine minimum excess air requirements for operation of a continuous regenerator with good temperature control and maintenance of a low residual concentration of sulphur on the lime bed.

c. Demonstrate operability of the process over a prolonged period of time to show that accumulation of fines, agglomerates, carbon or other deposits do not interfere with continuous operation.

d. Demonstrate means of preventing or removing deleterious accumulations of tar or carbon from gasifier and transfer duct internals.

e. Determine effects of number and location of fuel injectors on gasification, sulphur removal, and carbon content of gasifier lime. Include operation with single oil injector passing through the air distributor.

f. Test and demonstrate means of process start-up, shut-down, turndown and control. Determine maximum turndown ratio with independent control of gasifier and regenerator variables.

g. Determine effect of regeneration temperature on the maintenance of lime activity.

h. Study the existing burner operation with CAFB gasifier product. Establish operability with high gas velocity, measure flame characteristics, efficiency of combustion, production of NO_x, and flame stability.

i. Under conditions of lined out operation with equilibrated lime, measure SO_2 removal in the regenerator and determine rate of lime attrition and particle size distribution of solids carried over from the gasifier and regenerator. Determine engineering properties of equilibrium solids such as fluidized bed density, minimum fluidization velocity and particle size distribution.

Four pilot plant runs were planned to accomplish these objectives. To continue the numbering system begun in Phase I, these runs are designated runs 4, 5, 6 and 7.

Task II was an evaluation of additional limestones with two new fuel oils in batch reactor experiments conducted between continuous unit runs. One of the test oils was a high sulphur residue by-product of gas oil desulphurisation, the other a high sulphur pitch. Originally, four new stones were to be studied. Because of factors uncovered during the first pilot plant run, the batch unit programme was modified to include measurements of dust production tendencies of the stones under combustion conditions and to reduce the number of stones to be investigated to three.

Task III was a definition and assessment of the scope of engineering effort required to move the CAFB from the pilot plant stage through the development stage including the demonstration unit.

Tasks I and II were completed at the Esso Research Centre Abingdon, England. Task III was conducted by the Esso Research and Engineering Co., Florham Park, N.J., U.S.A.

REPORTING AND DISCUSSION OF RESULTS

During the performance of Task I the objectives for Task II were modified as a result of information generated during the first continuous gasifier run under Task I. Consequently, the reporting and discussion of results of Tasks I and II is set out in Sections V and VI in chronological order, so that the sequential logic of changes to objectives, equipment and techniques can be readily followed.

SECTION IV

DESIGN AND CONSTRUCTION OF EQUIPMENT

GENERAL

The experimental equipment used in this study consists of two batch reactor units and the continuous CAFB pilot plant. These units have been described (Reference 1) previously in detail. For the current work, the batch units remain essentially unchanged. However several modifications were made to the pilot plant on the basis of experience gained in the first three runs.

BATCH UNITS

Each batch unit contains a reactor, air and fuel systems, flare for product gas disposal, and gas sampling and analysis system as shown in the flow plan, Figure 2.

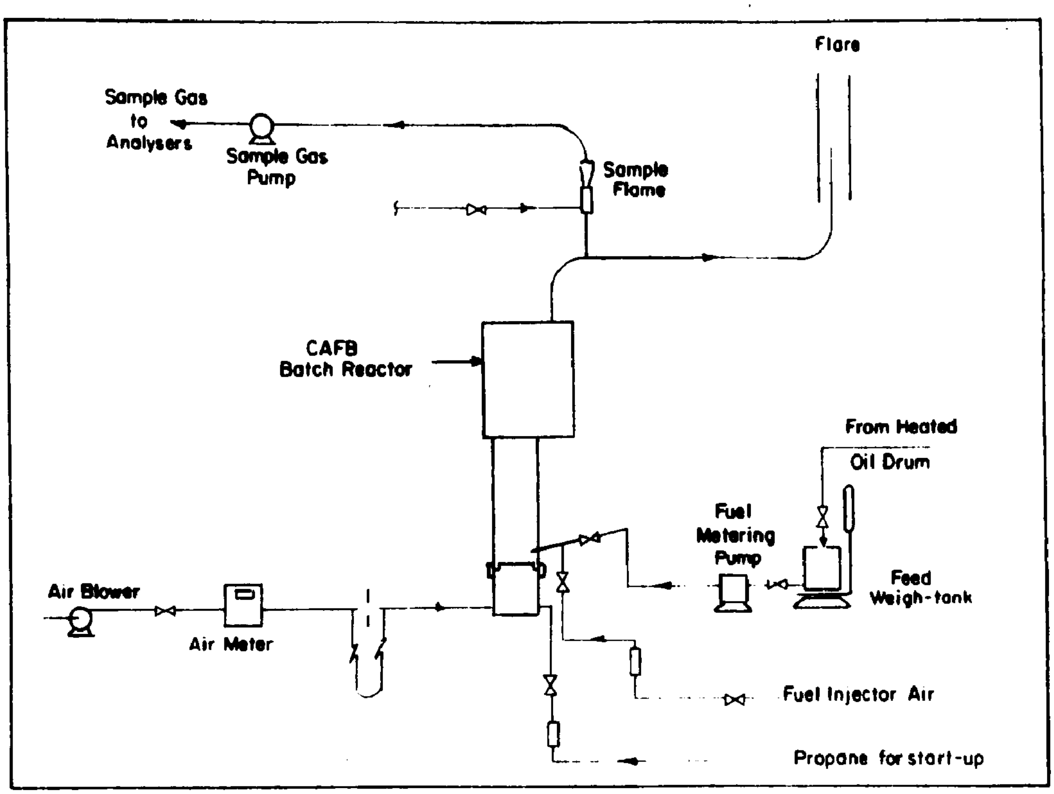

FIGURE 2 Batch Unit Flow Plan

A reactor is illustrated in Figure 3.

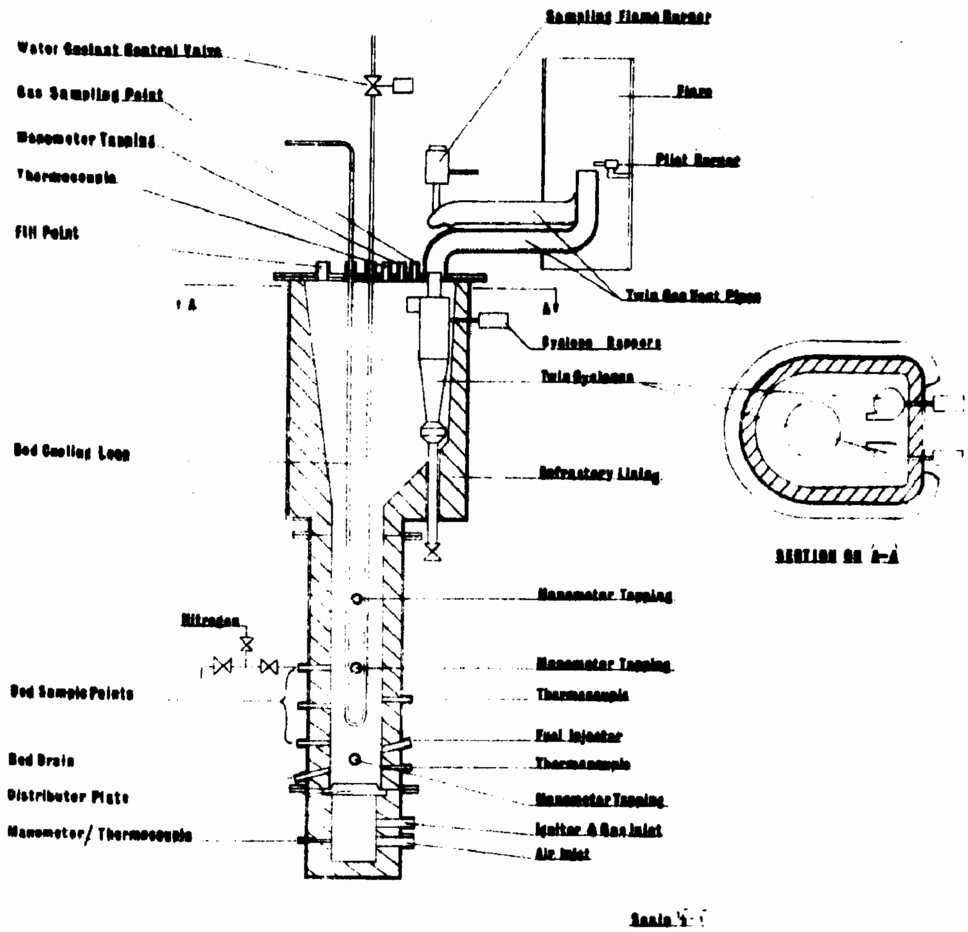

FIGURE 3 Batch Unit Reactor

The reactor is of refractory lined carbon steel construction. The lower section, which contains the fluid bed, is 17.8 cm (7 inches) i.d. by 83.8 cm (33 inches) high. The upper section is expanded to reduce gas velocity and is non symmetrical to permit internal cyclones to drain externally.

The plenum beneath the gas distributor is also refractory lined to serve as a combustion chamber for propane-air mixtures used during unit start up. The distributor is a "top hat" shape cast refractory design. The central raised section, 12.7 cm (five inches) in diameter, contains 16 horizontal holes around its circumference.

For the current work, a rapper was installed on one cyclone in each reactor to prevent fine particles sticking to the cyclone walls. The pneumatic activator of the rapper is located outside the reactor and drives a striker rod through a gland to tap on the cyclone wall.

The gas analysis equipment used in this study is the same as used in Phase I, and is summarised in Table 1. Full details are given in Appendix H.

Table 1

Batch Unit Gas Analysis Equipment

Analyser	Type	Manu-facturer	Model	Response	Range
SO_2	Infra-red	Maihak	Unor 6	Continuous	0-1, 000 cm^3/m^3 (ppm)
SO_2	Infra-red	Maihak	Unor 6	"	0-20% by vol.
SO_2	Conductimetric	Wostoff	-	"	0-1000 cm^3/m^3 (ppm)
CO_2	Infra-red	Maihak	Unor 6	"	0-20% by vol.
CO	Infra-red	Maihak	Unor 6	"	0-20% by vol.
O_2	Paramagnetic	Servomex	OA 137	"	0-25% by vol.

CONTINUOUS PILOT PLANT

Process Flow Plan

Figure 4 is a process flow plan of the continuous pilot plant. The heart of the system is the gasifier-regenerator unit cast of refractory concrete contained in an internally insulated steel shell. The product gas of the gasifier fires a 2930 kW (10 million Btu/hr) pressurised water boiler. The hot water is heat exchanged with a secondary water circuit which loses its heat through a forced convection cooling tower. The rest of the system consists of the necessary blowers, pumps and instruments to operate the gasifier, regenerator, burner and solids circulating system.

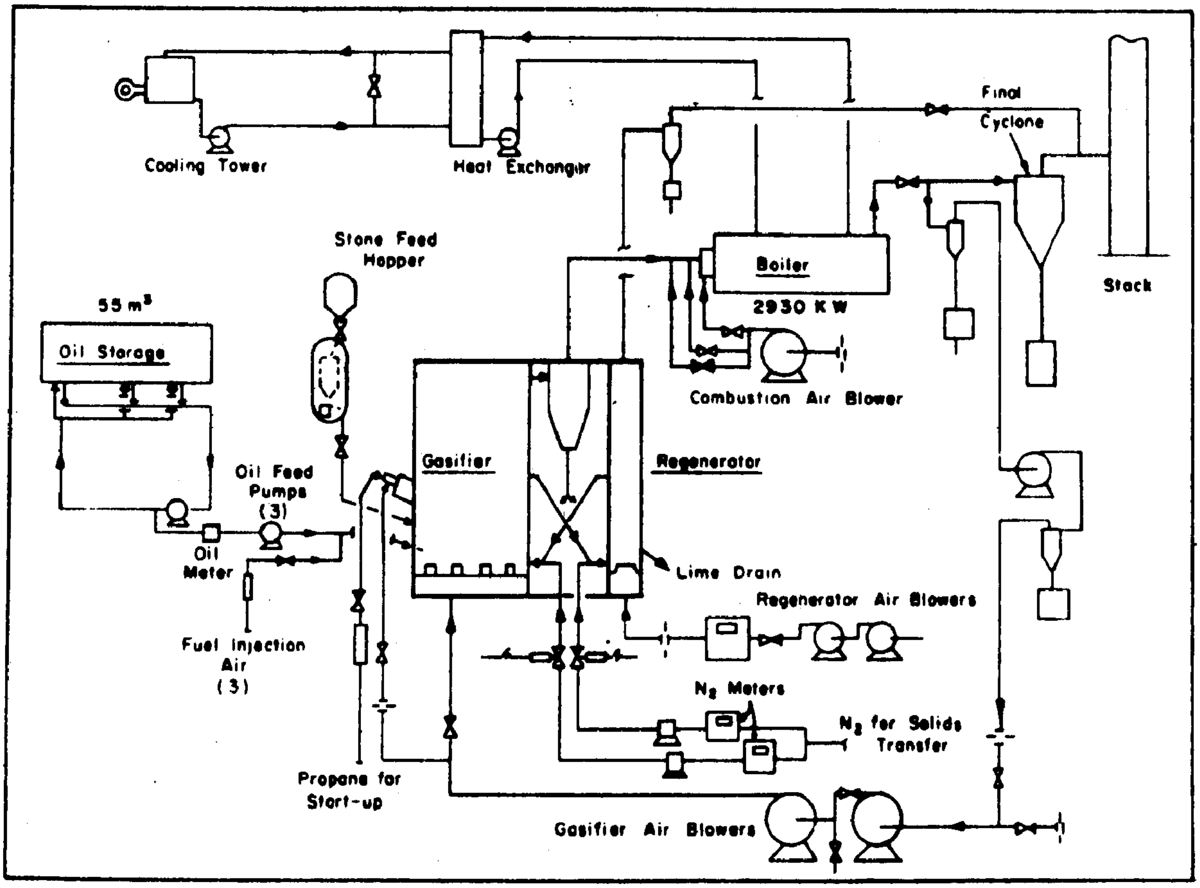

FIGURE 4 CAFB Pilot Plant Flow Plan

The gasifier itself sits within a pit to permit alignment of the gasifier outlet duct with the burner inlet. Fuel pumps, flow meters, and start up burner controls are mounted on a mechanical equipment console. Electrical instrumentation and manometers are mounted on a separate control console. Gasifier blowers are located in a separate blower house outside the main building, and the cooling tower is mounted on the roof.

The gasifier and regenerator reactors are cavities in a single refractory concrete block. The block contains other cavities which make up the gasifier outlet cyclones, the gas transfer ducts, and the transfer lines through which solids circulate between gasifier and regenerator. The gasifier cavity is rectangular in cross section, tapering from 44.5 x 94 cm (17.5 x 37 inches) at the distributor level to 49.5 x 99 cm (19.5 x 39 inches) at the 53 cm (21 inch) level. The upper portion has parallel sides. The regenerator tapers from 17.8 cm (7 inches) diameter at the bottom to 20.3 cm (8 inches) diameter 55.9 cm (22 inches) above distributor datum and remains parallel thereafter. A full description of the unit is given in Reference (1), pages 20-28.

PILOT PLANT MODIFICATIONS

The operation of the pilot plant in Runs 1, 2 and 3 showed the need for improvement in some areas. Changes were made to various sections between test runs as a result of experience and other changes were made to achieve specific objectives.

Modifications prior to Run 4

A new system of feeding limestone to the gasifier was constructed to provide continuous monitoring of the limestone feedrate and also enable the feed hopper to be refilled from an upper lock hopper without disturbing the stone feed into the gasifier.

An additional blower was installed to boost the flue gas recycle supply to the main blowers for the gasifier and a cyclone was installed in the main flue from the boiler to the stack.

The earlier test runs had illustrated the importance of pressure balancing the regenerator and gasifier pressures and an automatic pressure balancing valve was installed into the regenerator offgas line.

The gasifier was modified to include silicon carbide cyclone outlet tubes in place of the double wall stainless steel tubes with steam cooling used on Run 3. The regenerator distributor design was modified from a refractory construction which had a tendency to crack, to one with stainless steel nozzles protruding through a layer of refractory. This principle of distributor design had been proven on the gasifier distributor although the operating temperature was lower than the regenerator application.

The stainless steel nozzles in the gasifier distributor were modified to provide a low gas exit velocity to minimise damage to the bed material. The design utilised the original nozzle but included an additional outer ring to provide a staggered path for the outlet air before emerging through large holes at a lower velocity. This design still maintained the original nozzle pressure drop characteristic because of the retention of the original small diameter holes.

The bifurcated duct connecting the cyclone outlets to the burner was rebuilt with swept bends at the changes in duct direction to minimise the deposition of lime and carbon shown in the earlier runs.

Modifications prior to Run 5

Before Run 5 a number of modifications were made to the pilot plant to permit improved operations with a dusty limestone. The major changes were:

- External cyclone drainage
- Non obstructing regenerator pressure control system
- Flue gas recycle scrubber
- Regenerator overtemperature quench
- Flue gas stack scrubber
- Cyclone liners

Other minor changes were made to the unit to improve pilot flame stability and to assist in diagnosing boiler performance. The revised flow plan is shown in Figure 5.

Cyclone External Drain System

Pressure balance calculations on the gasifier cyclone return system indicated that there would be insufficient height of leg available to return fines to the gasifier through the internal passages if bed depth were increased to the levels desired for high sulphur recovery with low replacement rates of BCR 1691 stone. This problem increases in severity when fouling increases the pressure drop across the cyclone inlet. By using external cyclone drains, the pressure at the cyclone drains could be made independent of the gasifier bed pressure. It was not sufficient however just to drain the cyclones

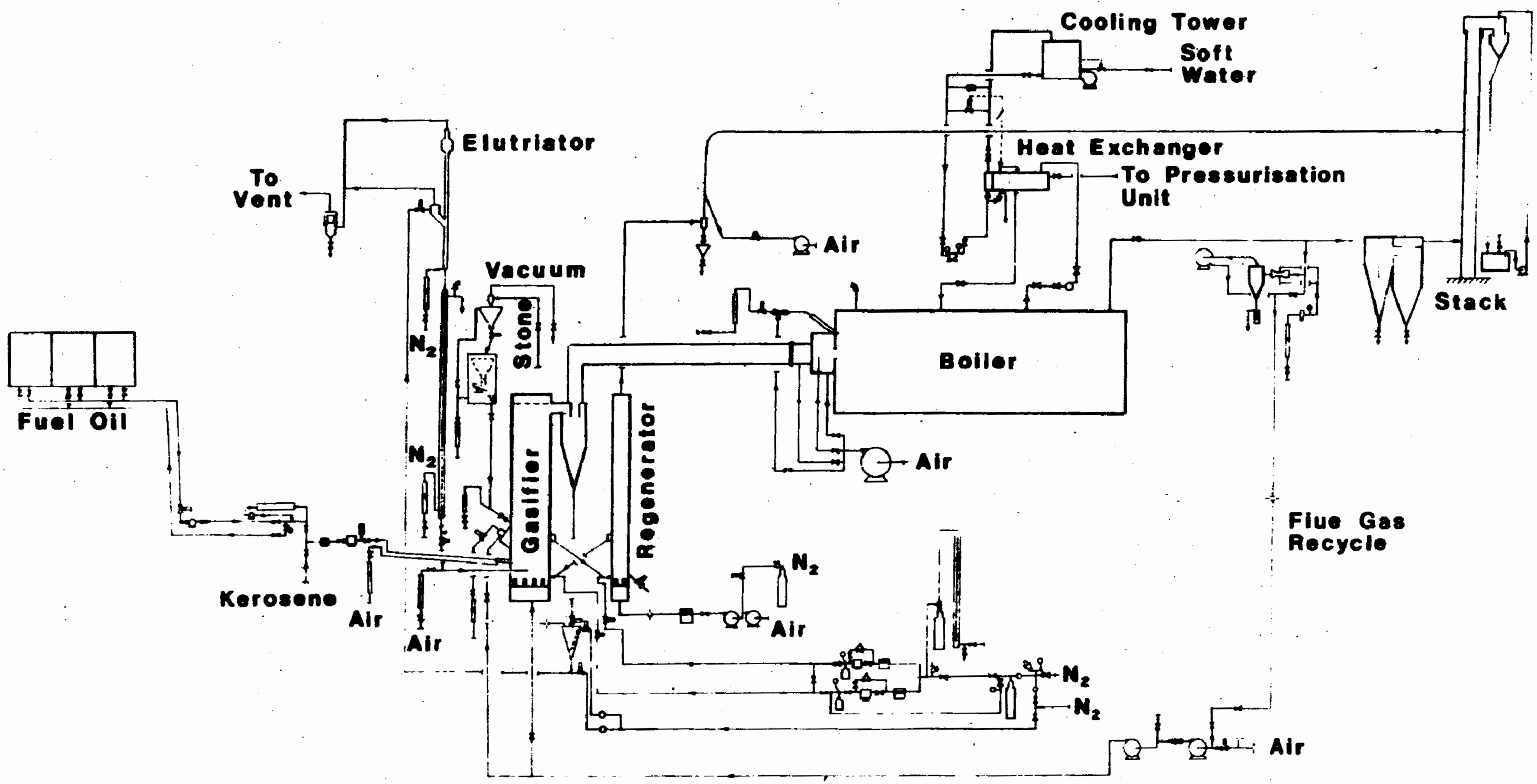

FIGURE 5 CAFB Revised Pilot Plant Flow Plan (Run 5)

externally. With deep beds the rate of entrainment into the cyclones could be high and stone losses severe unless the coarse fraction were returned to the gasifier. Therefore an external system to both drain the cyclones and return the fines was needed.

After consideration of several designs, a system was selected which met the constraints of available space, pressure, and gas consumption. Details of the system appear in Appendix G. In summary the system for each cyclone consists of a conical bottom pot receiver mounted beneath each cyclone drain, a butterfly valve to isolate the pot from the cyclone when the pot is being emptied and a Warren Springs Laboratory pulsed flow powder pump to transfer solids from the conical pot to an overhead receiver, common to both cyclones. An elutriator removes the very fine fraction from the cyclone solids, and a pneumatic injector returns the larger size fraction to the gasifier bed. Nitrogen is the operating gas for the transfer system. Most of the time the butterfly valve beneath the cyclone remains open draining solids to the conical pot. At timed intervals the butterfly valve closes and the pot is pumped out to the overhead receiver.

Regenerator Pressure Control

To avoid a repetition of the regenerator off gas line blockage which terminated Run 4, the pressure control valve was removed from the outlet line. This valve had become blocked by fine solids in the gas stream during that run. In order to regulate regenerator pressure a blower was fitted which injected air downstream of the cyclone outlet. Air flow from this blower is requlated by control loop which senses the difference between gasifier and regenerator pressure and adjusts a valve in the air line to achieve the desired pressure difference.

Flue Gas Recycle Scrubber

Run 4 demonstrated that simple cyclones were unable to provide sufficient cleaning of the recycle flue gas stream to prevent gradual blockage of the gasifier air distributor nozzles. A venturi scrubber system was designed to provide greater clean-up. The scrubber was designed to handle 340 m^3/hr (200 CFM) of gas at a pressure drop of 3.48 kPa (14" w.g.) Water is sprayed into the gas at the throat of a venturi. A knockout vessel at the venturi outlet removes the water and entrained dust. The venturi was placed on the suction side of the recycle blower to protect the blower from dust, and a recycle line was provided to permit a high gas circulation rate through the venturi even at low rates of flue gas flow to the gasifier.

Regenerator Quench

The circulation of fresh solids from the gasifier to the regenerator controls regenerator temperature. Upsets in the pressure balance between gasifier and regenerator or temporary obstructions in one of the solids transfer lines can sometimes interrupt this solids circulation and allow regenerator temperature to increase.

If regenerator temperature gets too high there is danger of sintering the lime particles and forming agglomerates. An emergency quench system was installed to prevent this occurrence. The lower regenerator bed thermocouple was connected to a controller which admits a flow of nitrogen to the intake of the regenerator air blower when bed temperature reaches the alarm point. The alarm was set to operate at 1100 deg.C. Nitrogen fed to the blower dilutes the regenerator air supply and reduces the rate of oxidation to prevent over temperature. The circuit is fitted with a manual switch so that the process operator can inject nitrogen at will in the event of other forms of upset.

Stack Top Gas Scrubber

To avoid particulate emissions to the atmosphere during periods of high lime losses from the gasifier cyclones, a final stage of water scrubbing was added to the pilot plant flue gas stack. Experience in Run 4 had indicated that the flue gas cyclone was not completely effective in recovering lime fines produced from BCR 1691 under combustion conditions.

The new scrubber consists of a section of ductwork shaped like an inverted "U" mounted on top of the stack. The down leg directs the gases into the top of a funnel shaped receiver which causes another reversal of gas direction upward to the atmosphere. Water is sprayed into the down leg and collected by the funnel. This water which picks up limedust by passage through the flue gas is conducted to a ground level settling vessel. Overflow from this vessel is circulated back to the scrubber nozzle by a centrifugal pump. The system is designed to circulate water to the scrubber at a rate of approximately 27 m^3/hr (100 gallons/min).

Modification prior to Run 6

The poor performance of the regenerator in Run 5 may have been partly caused by the absence of fines which previously had been returned from the right hand cyclone into the regenerator. It was decided therefor in Run 6 to reinstate this internal transfer line for the right hand cyclone.

The pressure vessel transfer system which had been used on the right hand cyclone was modified to act as a hot limestone ejector to reduce carbon and lime deposits in the cyclone entries. Figure 6 shows the modified flow plan and it will be seen that hot bed material could be drained from the gasifier and then ejected at a controllable frequency into either left or right hand cyclone entries. In addition to this change, other modifications were made to improve operability of the transfer system by installing perforated stainless steel plates within the conical transfer vessels to retain the flakes of carbon and lime which earlier runs had shown to choke the transfer pipes.

The regenerator distributor position was lowered with respect to the transfer port ducting material from the gasifier so that fresh material entering the regenerator would enter into a hotter zone with a possible improvement in selectivity. The distributor was lowered by inserting a silicon carbide ring into the regenerator plenum.

The fuel injection system was extended to provide a further injector through the gasifier distributor with one single outlet hole set to discharge fuel horizontally into the bed. The injector could be retracted into the distributor when not in use.

The piping was arranged so that the total fuel supply would be fed into the unit either totally or partially through the bottom injector and side injectors. The gasifier plenum was sub-divided into two sections in the ratio of 1:2 and by individually controlling the air to the two plenums it was possible to produce different velocities in the bed area and induce more rapid mixing across the width of the bed.

Modifications prior to Run 7

The major changes made prior to Run 7 were associated with the two distributors and the regenerator cyclone fines system. The gasifier distributor was modified to include two direct heat transfer water cooling tubes for bed temperature control instead of flue gas recycle. In addition the fuel injector through the centre of the distributor was modified to include six outlet holes around its periphery instead of one large outlet used in Run 6 in an attempt to improve single injector performance.

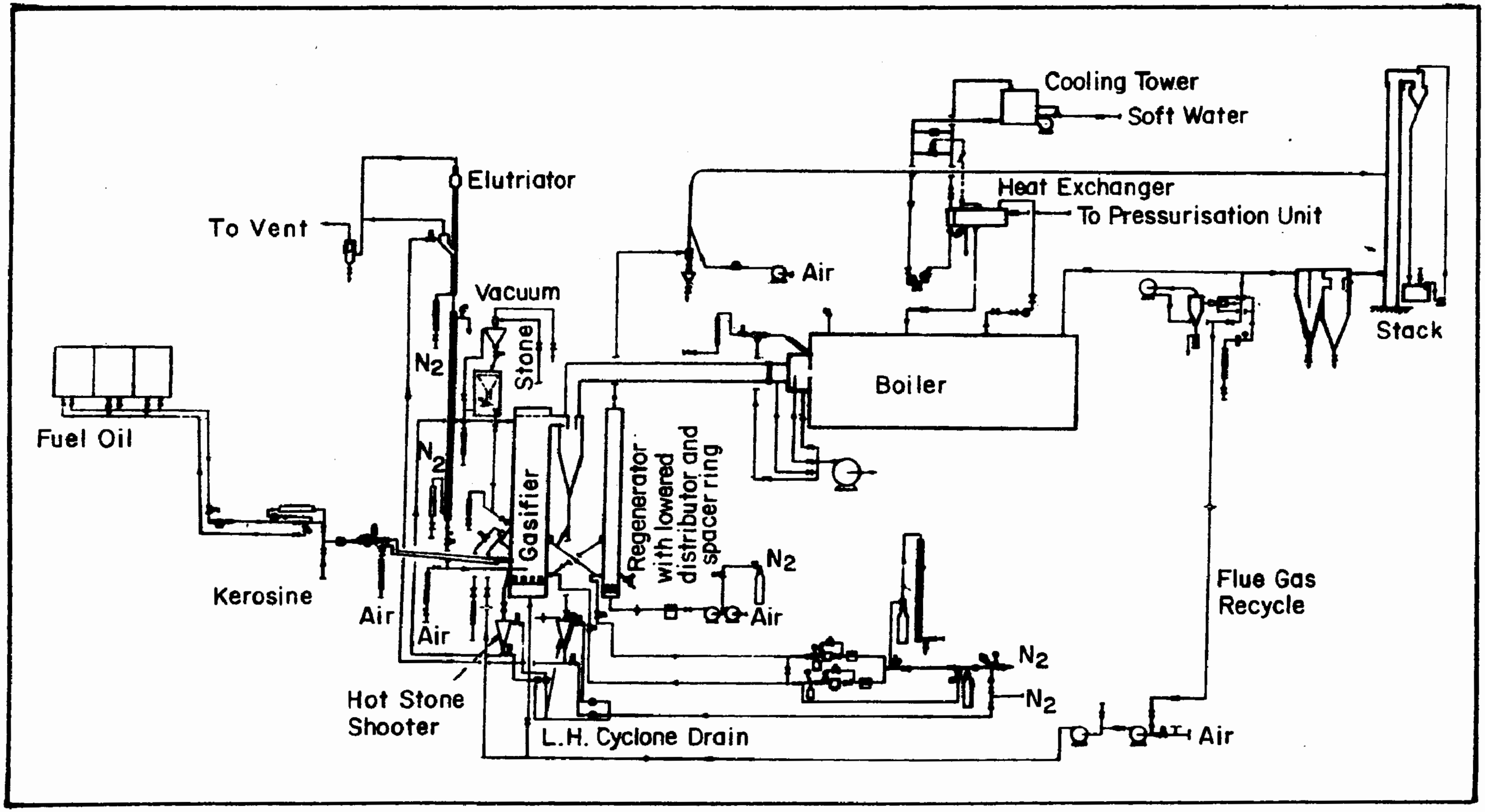

FIGURE 6 CAFB Revised Pilot Plant Flow Plan (Run 6)

Also the gasifier nozzle design was changed to the original straight path high velocity exit type because the staggered path low velocity design did not provide any evident improvement.

The regenerator distributor was changed to the top hat design used in the Batch unit test programme and Runs 1, 2 and 3 of the continuous unit programme. After the good performance of stainless steel in distributor designs, it was decided to make this distributor in stainless steel and so eliminate the unreliable performance of this component in refractory.

The fines collected in the regenerator cyclone had in all earlier runs been drained externally and discarded. It was considered useful if the unit could be modified to provide the facility to return these fines into the gasifier via the elutriator and air injection system which handled the fines from the external cyclone drainage system. The modified flow plan is shown in Figure 7.

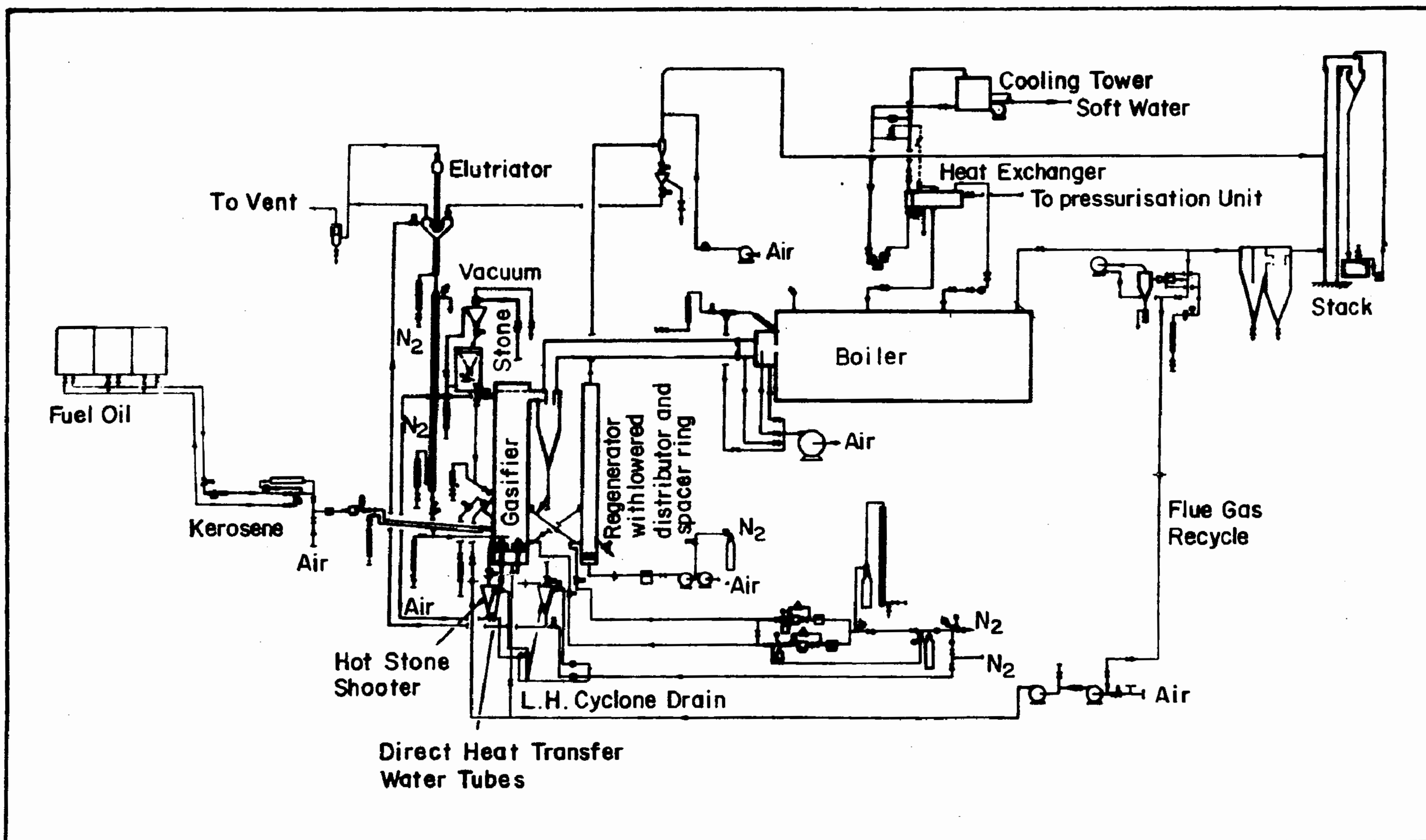

FIGURE 7 CAFB Revised Pilot Plant Flow Plan (Run 7)

SECTION V

PROGRAMME OF WORK

GENERAL

The programme of work consisted of three tasks - Task I, operation of the continuous CAFB gasifier in four runs (numbered four to seven), Task II, study of additional limestones and fuel in batch gasification units, and Task III, scoping of the engineering effort needed to take the CAFB process through to a 100 megawatt (electrical) scale conversion of a commercial power generation boiler.

As originally envisaged these three tasks were entirely separate, i.e. there was no intention of selecting limestones or fuels tested as part of Task II for use in Task I of this phase of studies. However, the first run in Task I, Run 4, identified severe operational problems in the form of gas line plugging by very sticky fines, and as a result Task II was modified to allow for an examination of the problem of fines formation. The final programme of work carried out under each task is set out below.

TASK I

Four runs were carried out in the 2930 kW (10 million BTU/hr) continuous CAFB gasifier. In each of these runs not only the work objectives were different, but also the configuration of the gasifier itself, since at this stage in a development project the pilot plant serves as a means for gathering data under realistic operating conditions and also provides an opportunity for evaluating new equipment, configuration and operating methods.

Run 4

The experimental plan for Run 4 called for use of limestone BCR 1691 at a series of pilot plant conditions to test correlations based on batch unit studies. Limestone replacement rate, gasifier bed depth, and gasifier bed temperature were the major variables to be examined. A brief test of the effect of regenerator excess oxygen content was also included in the experimental plan.

A major goal of the study was to find if increasing gasifier bed depth would have the beneficial effect that batch studies had indicated to be possible. As events developed, properties of limestone BCR 1691 prevented accomplishment of these test goals and focussed attention on development of means to start up and operate with a stone which produces a great deal of dust under certain conditions.

The dust forming characteristics of this stone under full combustion conditions were found to be much worse than had been encountered with Denbighshire stone. This high rate of fines production caused a number of operating problems, extended the start up period, and eventually caused termination of the run after nine hours of gasification. The start-up and operational problems and data are listed in Appendix A, and results are discussed in Section VI.

The configuration of the unit during Run 4 is fully described in Section IV, but essentially the gasifier was set up in the same manner as for Run 3, with minor changes to equipment such as improved stability of control over stone feed and withdrawal from the unit and pressure balance between the gasifier and regenerator, improved gas flow in the ducts to the burner, more durable cyclone outer tubes and modified air distributors in both gasifier and regenerator. The ventilation and dust extraction facilities were also improved in view of the potentially hazardous nature of the BCR 1691 limestone, which contains a significant amount of silica.

Run 5

Objectives of Run 5 were to measure sulphur removal efficiency at different bed levels and lime replacement rates, to test feasibility of desulphurising gasification at temperatures above 900 deg. C approaching adiabatic conditions, and to determine effectiveness of pilot plant modifications in solving problems met in Run 4.

The test programme was to start up with precalcined lime from Denbighshire stone, operate for a test period with Denbighshire stone, and then switch to BCR 1691 stone. Provision was allowed for returning to Denbighshire stone if the BCR 1691 proved inoperable.

In actual fact, Denbighshire stone was used during the first two days and the final week of operation. BCR 1691 was used for the rest of the run.

Table 2 summarises the various periods of operation during the run. For purposes of computer identification of data, each run hour is designated by a decimal number with its whole part signifying run day and its fractional part the hour. February 6 was run day 1.

For example, 4.1630, represents 4.30 p.m. on February 9. The same system is employed on the abscissa representing time in the data graphs of Appendix B. These graphs show the variation with time of major operating variables during the run. Tables of detailed operating data also appear in Appendix B, together with a log of the run and the post-run inspection. Results are discussed in Section VI.

The configuration of the unit for Run 5 is fully described in Section IV. Between Runs 4 and 5 a number of modifications were made to the pilot plant to permit improved operations under conditions like those encountered in Run 4. The experience of Run 4 and data gained in the batch unit programme revealed that operation is more difficult with limestone BCR 1691 than with Denbighshire stone. The batch work indicated that BCR 1359, another high purity stone, should behave more like Denbighshire stone. However there are many locations where high purity limestone will be considerably more expensive than lower purity stones available locally. It therefore was desirable to assess more fully the consequences of operating with a lower purity stone. Although BCR 1691 is not necessarily typical of all low purity stone, it has deficiencies which, if overcome, would assure that the CAFB gasifier could operate with stones of a wide quality range. Five major changes were made to assist operations with the dusty and more easily agglomerated stone.

- External cyclone drainage
- Non obstructing regenerator pressure control system
- Flue gas recycle scrubber
- Regenerator overtemperature quench
- Flue gas stack scrubber
- Cyclone liners

Other minor changes were made to the unit to improve pilot flame stability and to assist in diagnosing boiler performance.

Table 2

Summary of Run 5 Operating Periods

Dates	Run Days	Limestone Feed	Number of Test Periods	Gasification Hours	Cause of Termination
Feb 6 - Feb 9	1.2000-4.0500	Denbighshire	2	109	Pressure drop through boiler.
Feb 9 - Feb 11	4.0500-6.1000	BCR 1691	2		Change to BCR 1691 caused no interruption of gasification.
Feb 13 - Feb 14	8.1920-9.0630	BCR 1691	0	11	Regenerator Defluidisation
Feb 16 - Feb 18	11.1215-13.1000	BCR 1691	4	45	Fuel Injector Failure
Feb 21 - Feb 22	16.0600-17.2200	BCR 1691	2	40	Regenerator Defluidisation
Feb 25 - Feb 27	20.1800-22.1900	Denbighshire	4	49	Cyclone Inlet Pressure Drop
Feb 28 - March 3	23.1300-26.1900	Denbighshire	5	78	Voluntary - End of Run

Of the changes made the most significant was the introduction of an external cyclone fines drainage and return system. Previously the cyclone fines had been returned partly into a stream of regenerated stone flowing from the regenerator to the gasifier, and partly into a stream of stone flowing from the gasifier into the regenerator. No data is available on the mass flows for the cyclone fines via each route, but it is reasonable to assume them to be equal. Thus half the gasifier fines trapped by the gasifier cyclones were fed directly into the regenerator.

Under the revised configuration all cyclone fines were fed directly into the gasifier itself, hence it was recognised that there was a possibility that these fines would be stripped from the bed before reaching the catch pocket for the gasifier to regenerator bed circulation line, and consequently these fines would not enter the regenerator at any time, but would recirculate around the loop: gasifier bed-gasifier cyclones-fines return to gasifier bed. In time attrition would result in these fines being reduced to such a small size that they would enter the boiler - their only route for escape from the loop. To avoid the possibility an elutriator was installed in the fines return line to allow selective removal of fines.

Run 6

The experimental plan for Run 6 was intended to provide information in the following areas:

- Performance of Limestone BCR 1359
- Fuel injector number and location
- Regeneration
- Reduction of solids deposits in gasifier cyclones and ducts.

To obtain this information, the programme of test conditions shown in Table 3 was proposed. The first four tests were aimed specifically at finding the best set of regeneration conditions for the subsequent gasifier tests. The effects of excess oxygen level and temperature were to be tested with constant gasification conditions. Also by comparing overall results with those of Run 5 it was intended to determine if returning half the gasifier fines to the regenerator produced a more effective regenerator operation.

Table 3

Test Programme for Run 6

Test No	Air/Fuel Ratio (% of Stoich)	Gas Velocity (Gasifier) (m/sec)	Gasifier Temp (°C)	Limestone Particle size (μ)	Gasifier Bed depth (cm)	Make-up Rate (mol CaO/mol S)	Regenerator Temperature (°C)	Regenerator Excess O_2 (% by vol)	Number of Fuel Fuel Injectors and Position
1	20	1.22	870	600 - 3000	69	1.0	1050	0	3 (normal/side)
2	"	"	"	"	69	1.0	1050	1.0	" " "
3	"	"	"	"	69	1.0	1090	1.0	" " "
4	"	"	"	"	69	1.0	1090	0	" " "
5	"	"	"	"	51	1.0	(a)	(b)	" " "
6	"	"	"	"	69	1.5	"	"	" " "
7	"	"	"	"	51	1.5	"	"	" " "
8	"	"	"	"	69	0.5	"	"	" " "
9	"	"	"	"	69	1.0	"	"	" " "
10	"	"	"	"	69	1.0	"	"	1 (high/side)
11	"	"	"	"	69	1.0	"	"	1 (plenum)

(values for (a) and (b) to be decided after Test 4)

The next set of five tests was to provide additional information on the effects of bed depth and limestone replacement rate on sulphur removal efficiency. Two depths and three lime replacement rates were to be used. Regeneration conditions for these tests were to be selected on the basis of the results of the first four tests.

The final two tests consisted of a study of the use of a single fuel injector. Both a high level, side entering injector and a bottom entering, variable level injector were to be tested.

The investigation of a means to reduce cyclone inlet fouling was to continue throughout the run. The method under trial was the injection of gasifier bed solids into the cyclone inlet. At first, only the left cyclone was to be treated with the right cyclone untreated as a control. If the method proved successful in keeping the left cyclone inlet clear while the right one fouled, the solids injector would be transferred to the right side to see if it would clean a fouled inlet.

Precalcined Denbighshire lime and used bed removed from Run 5 was to be used to establish the initial fluidised bed during startup. Stone feed would then be switched to BCR 1359 for the remainder of the tests.

The configuration of the unit for Run 6 is fully described in Section IV. Modifications from the Run 5 configuration were made in the areas of fuel injection, plenum air distribution, cyclone fines return, boiler flue gas sampling, regenerator distributor location, and cyclone lining.

To permit tests of fuel injector location two new fuel injectors were installed and provision was made for piping the total fuel input to one, two or three injectors. One of the new injectors entered the bed vertically through the distributor. The other new injector entered through the side at a higher elevation than the original three injectors.

The air distributor plenum was subdivided to permit a variation in air supply rate in two different sectors. This arrangement was to permit a test of the effect of increasing solids circulation rate in the vicinity of the fuel injector when a single injector was employed.

The right hand gasifier cyclone was changed back to the original configuration so that fines were drained to the regenerator. The other cyclone remained as it was in Run 5 returning its fines to the gasifier by an external transfer system. A system to circulate solids from the gasifier bed to the left hand cyclone entry at a controlled rate was installed. This equipment was to test the effect of coarse solids injection on reduction of fouling in the cyclone inlet.

A new sampling system was installed at the rear of the boiler to increase reliability of the flue gas analysis. A large flue gas sample was to be withdrawn from the boiler through a hot cyclone to remove most of the dust. A smaller sample then would be drawn through a hot filter to remove residual dust before passage through a condenser to the analysers.

A high density silicon carbide liner was fitted to permit mounting the regenerator air distributor four inches lower than its former position. This arrangement was intended to increase regenerator residence time and provide a greater vertical separation between the air entry and solids entry points. The opportunity this provided for the air to be heated before meeting fresh sulphided solids was intended to improve selectivity of CaS oxidation to CaO and SO_2.

Stainless steel cyclone wall liners which had been tested in Run 5 proved to be unsatisfactory and were removed. The cyclone walls were treated by application of a layer of castable refractory to reduce surface roughness.

Run 7

The experimental programme for Run 7 was aimed at clarifying the differences between sulphur removal efficiencies observed in Runs 3 and 6. It was not possible to use data from Runs 4 and 5 to resolve these differences because of operational problems with Run 4 and data inconsistencies in Run 5.

The first test condition in Run 7 was selected to match conditions studied in Run 3, and the different test programmes were planned to take account of the two possible results: test sulphur removal efficiencies in Runs 3 and 7 were the same, or that Run 7 gave a lower efficiency than Run 3. The programmes differed only in the actual levels set for the major variables of bed depth and stone replacements rate. These were to be varied, together with air/fuel stoichiometry and bed temperature, in a factorial experiment as set out in Table 4.

Table 4

Factorial Plan, Run 7

Test	Bed Depth	Stoichiometry	Bed Temp.	Feed Rate
1	H	H	L	H
2	H	H	L	L
3	H	H	H	H
4	H	H	H	L
5	H	L	L	H
6	H	L	L	L
7	H	L	H	H
8	H	L	H	L
9	L	L	L	H
10	L	L	L	L
11	L	L	H	H
12	L	L	H	L
13	L	H	L	H
14	L	H	L	L
15	L	H	H	H
16	L	H	H	L

Other objectives of the run were to test a new single fuel injector and indirect cooling of the bed by means of immersed water cooled tubes.

The configuration of the unit for Run 7 is described in Section IV. Modifications from the Run 6 arrangements were minor and consisted of:

a. Gasifier Distributor

The nozzle configuration of the distributor was restored to that used in Run 3. Two heat exchanger tubes of portal frame configuration were installed in the new distributor together with a six-way central fuel injector. Both the heat exchangers and the fuel nozzles were rectractable and were installed in the retracted position. These components were to be used towards the end of Run 7.

b. <u>Regenerator Distributor</u>

The original top-hat design of distributor was installed in the regenerator but it was lowered **10 cm (4 inches)** by means of the silicon carbide ring used in Run 6. Provision was made for changing this distributor during the course of Run 7 should this prove to be necessary.

c. <u>Regenerator Cyclone Drain</u>

During Run 6 considerable amounts of fine bed material were drained from the regenerator cyclone and lost from the system. Provision was made in Run 7 for the elutriation of this stream and the re-injection of the coarser fraction into the gasifier bed.

d. <u>Bed Transfer System</u>

During Run 6 it was found expedient to rely on the manual setting of the pulser in the regenerator to gasifier transfer line for coarse temperature control, the pulser on the gasifier to regenerator line being used mainly to ensure that the R.H. cyclone drain functioned properly. For Run 7 the temperature controller was wired to the regenerator to gasifier line and the gasifier to regenerator line was to be operated manually.

e. <u>Flue Gas Recycle Scrubber</u>

During Run 6 there were several occasions when the water drain from the flue gas recycle scrubber plugged and water was entrained by the flue gas recycle stream. A larger diameter drain was fitted to the demister for Run 7.

TASK II

In the original programme of work, batch unit studies were to determine the suitability of additional fuel-limestone combinations for CAFB applications. However since Run 4 operations revealed a serious problem with BCR 1691 stone which had not appeared in earlier batch unit tests, the batch programme was revised to include an investigation of dust forming tendencies under a variety of operating conditions.

In earlier batch work there had been a comparison of dust losses between stones during gasification-regeneration cycles. No such comparison had been made under fully combusting conditions. In the normal batch unit test procedure there had been little exposure of the solids to combustion conditions in the absence of sulphur except during calcination. Consequently the conditions employed during Run 4 start up produced an entirely unexpected result in that the BCR 1691 stone formed copious quantities of a dust with a very sticky nature.

The batch unit test programme was revised to accomodate the following objectives.

- Determine if continuous unit conditions which produced large quantities of sticky dust could be duplicated in batch units.
- Compare dust producing tendencies of Denbighshire and BCR 1691 stones under different conditions.
- Provide a quantitative measurement of dust production to be expected under start up and operating conditions with Denbighshire, BCR 1691, BCR 1359, and two additional stones to be provided by New England Electric System (NEES).
- Measure sulphur absorption performance of BCR 1359 and the two NEES stones.
- Conduct tests of the feasibility of operating CAFB with very heavy refinery streams, specifically, vacuum pipe still bottoms.

In the event through activities pursued by Esso outside the EPA programme a second heavy refinery stream was tested in the batch units, and by agreement with Foster Wheeler Corporation the results have been included in this report.

The procedure for operating the batch units are given in Appendix M, and data are listed in Appendix N. Batch results are discussed in Section VI.

Batch test equipment is essentially the same as those used in Phase I studies and is described in Section IV. The only modification incorporated for those studies was the addition of mechanical rappers to aid cyclone drainage when producing the sticky fines characteristic of combusting conditions with limestone BCR 1691.

TASK III

As part of this project, Esso Engineering, Florham Park, New Jersey, USA, was requested to scope the engineering effort which might be required to carry CAFB from its present stage of development through the construction, startup, and testing of a large scale demonstration unit. A 100 MW scale unit was assumed as a basis. This scoping study is summarised in Section VI. Detailed results have been supplied to the Environmental Protection Agency in a separate memorandum.

SECTION VI

DISCUSSION OF RESULTS

TASK I - STUDIES IN CONTINUOUS GASIFIER

Four runs in Task I are discussed below in terms of equipment performance and process performance. Prime attention is given to Runs 6 and 7 since these have given the most self-consistent and reliable data so far. Run 4 was prematurely terminated by problems of dust formation, and data from Run 5 cannot yet be made to balance on a self-consistent basis. Further detailed examination of Run 5 data will be undertaken in Phase III studies, which make provision for more extensive data work-up and mathematical modelling based on results obtained during the execution of this task.

Run 4

Equipment Performance is described in Appendix A. Major problems were encountered during the start-up of the continuous unit, in the following areas

(a) Blockage in solids transfer line
(b) Plugging in regenerator gas outlet system
(c) Dust emissions to boiler from gasifier
(d) Dust in flue gas recycle stream
(e) Dust emissions to atmosphere
(f) Regenerator Agglomerates

All of the problems were related to differences in the characteristics of stone BCR 1691 from those of the Denbighshire stone used in the continuous unit during Phase I studies (Reference 1). The major differences are lower fusion temperature, the cause of problems (a) and (f) above, and production of a higher proportion of very fine dust in a fluidised bed under fully combusting coditions, the cause of problems (b) through (e). The dust produced from BCR 1691 is more difficult to retain in collection equipment than that originating from Denbighshire stone. It also clings to surfaces of pipes, cyclones, control valves etc, and is difficult to dislodge without application of direct mechanical force. It does not drain from hoppers, or even vertical pipes, without continuous rapping.

Despite these problems the unit was eventually started up and gasification continued for a total of nine hours. At this time the control valve downstream of the regenerator cyclone plugged again and the run was terminated.

Examination of the gasifier after shutdown, coupled with the results of batch tests on various limestones to assess dust forming tendencies showed the need for several modifications to the gasifier before attempting another run with a limestone like BCR 1691. The modifications adopted are described in Section IV. In retrospect it seems likely that if the control valve had been cleaned one more time, and gasifying conditions had then been maintained, a longer period of gasification could have been achieved. However, it is also probable control over gasifier conditions would have been difficult, and achievement of steady lined out performance unlikely.

Process Performance cannot be discussed in detail since the 9 hour period of gasification was far too short to achieve lined out conditions. Table 5 summarises gasification conditions and results obtained. Initially a high lime replacement rate of 2.1 mole CaS was employed to build gasifier bed level. A slight reduction to 1.7 mole CaS was used during the final 4 hours. Sulphur removal efficiency of nearly 98% at the higher rate declined to about 93% when stone rate was reduced. However the gasification period was too short to consider these results to represent lined out conditions.

Table I Appendix A lists the distribution of particle sizes in the solids from gasifier and regenerator beds and in solids recovered from the boiler fire tube and regenerator cyclone during gasification, and the elutriation effect of the gasifier bed in removing particles smaller than the 355-600 micron fraction is apparent. We would expect particles smaller than about 500 microns to be entrained at Run 4 test conditions. It is evident that little of the entrained material was returned to the gasifier by the cyclone. The presence of a wide spectrum of particle sizes in solids from the boiler fire tube also indicates poor cyclone performance. However the gasifier cyclone which drained back to the regenerator evidently was operating as there was an appreciable fraction of 150-250 micron solids in the regenerator bed.

Table 5

Gasification Summary - Run 4

Day	Hour	Temperature deg. C Gasifier	Temperature deg. C Regenerator	Superficial Air Rate m/sec.	Fluidised Bed Depth cm.	Lime Replacement Mol CaO/Mol S	Air/Fuel % Stoich	Sulphur Removal %
1	2030	870	-	1.13	49.5	2.1	23.1	-
1	2130	882	1035	1.16	50.8	2.1	24.0	-
1	2230	872	1040	1.13	55.6	2.1	23.6	-
1	2330	870	1080	1.31	59.2	2.1	24.3	95.7
2	0030	875	1100	1.25	59.9	2.1	24.1	97.9
2	0130	881	1110	1.25	60.5	1.7	24.1	93.3
2	0230	875	1015	1.22	57.7	1.7	23.7	93.3
2	0330	872	1068	1.25	56.4	1.7	24.0	87.3
2	0430	872	1060	1.25	58.4	1.7	24.2	92.6

The bulk density of the bed solids was higher than has been observed in earlier studies. Batch unit tests with BCR 1691 had given settled bed densities of about 0.83 g/cc compared with values over 1.0 observed here. A change in density of the fluidised bed had also been noted during the start up period of Run 4. This density increased from about 0.8 to nearly 1.1 during the start up. It is possible that a selective loss of lower density particles contributed to this increase in bed density.

The chemical analyses of bed samples listed in Appendix A - Table II show that silica content of the beds, and indeed all solids samples, increased over those of the raw limestones. This change indicates that minerals other than SiO_2 were preferentially lost from the system, probably as very small particles.

The difference between gasifier and regenerator bed sulphur contents was 2.8% on stone indicating a good level of regeneration. A very high fraction, 99%, of the regenerator sulphur appeared as sulphate. This represents a considerably higher degree of sulphide oxidation than achieved in earlier runs and may indicate some oxidation of the sample during its collection. The regenerator cyclone fines show a slightly higher sulphur content than the gasifier bed sample. They also show a high content of sulphide which indicates that the fines passed through the regenerator without undergoing much reaction.

Run 5

Equipment Performance is described fully in the run log and post-run inspection in Appendix B.

Performance was greatly improved over that experienced in previous runs, particularly that of Run 4. A number of the new features operated well, but some continued to be troublesome throughout the run.

Stone Feeder -

The stone feed system which used a vibrator in a pressurised shell to feed from a weighed hopper proved reliable throughout the run. Stone feed rates were usually quite steady and easily measured.

Regenerator Drain Valve -

The gasifier bed level control system which used a pressure switch in the gasifier to activate a drain valve in the regenerator proved to be reliable and to give good control of gasifier bed depth.

Regenerator Pressure Control -

No blockages were encountered in the regenerator off gas line during Run 5. This line and its control system had plugged continually during Run 4 start up. In Run 5 the line remained clear and showed no sign of pressure build up. The system used controlled introduction of excess air into the outlet line downstream of the cyclone and avoided restrictions in this line. It was not possible to operate the system in automatic mode due to the large pressure pulses introduced by the solids circulating system, but manual control of the pneumatic valve position proved satisfactory for control of the pressure difference between regenerator and gasifier. Pressure difference was regulated to within 0.25 to 0.5 kPa (one or two inches water gauge). Normally the regenerator pressure was adjusted to be 0.75 to 1.25 kPa (3 to 5 in w.g.) below gasifier gas space pressure although higher differences were sometimes used.

Cleanliness of the regenerator gas line was also aided by continuous use of a pneumatic rapper on the regenerator gas cyclone. This rapper ensured drainage of solids from the cyclone walls. Also, the conditions that produced the very sticky fines, kerosene combustion in a bed of BCR 1691 stone, were avoided as much as possible.

Flue Gas Recycle Scrubber -

The venturi scrubber on the flue gas recycle stream removed a great deal of lime fines from the gas, but was not completely effective. Some particles passed the scrubber, and some fouling of the recycle gas line, control valve, and gasifier distributor was encountered. The rate of gasifier distributor pressure rise in Run 5 was much less than in prior runs. In the initial stages of Run 5 there was frequent plugging of the inlet of the venturi throat itself with lime deposits. Increasing the gas flow through the venturi to the maximum rate available (estimated at 340 m^3/hr) (200 CFM) by using maximum recirculation eliminated plugging at this point.

The drain line from the water separator occasionally blocked and required cleaning.

On at least three occasions blockage of this discharge caused water carryover to the gasifier plenum itself with a consequent sharp decrease in gasifier temperature.

Regenerator Over Temperature Protection -

The new regenerator emergency quench system dilutes the inlet air with nitrogen when regenerator temperature reaches the set point of 1100 deg.C. This system proved to be quite valuable and avoided excess temperature several times when malfunction of the solids circulation system reduced lime flow rate through the regenerator. In only one case did regenerator temperature seriously exceed 1100 deg. C, and that was due to emptying of the quench N_2 supply bottle before normal conditions were restored. This quench system is believed to be responsible for avoiding regenerator blockages by agglomerated solids which occurred in previous runs.

Boiler Pilot Flame -

The new boiler pilot burner and its gas and air system were quite effective in providing a stable and reliable pilot. No difficulty was met in lighting the pilot over a wide range of conditions nor in keeping it lit.

Stack Top Washer -

The water spray scrubber installed to prevent discharge of dust to the atmosphere was operated during part of the run when it appeared that some dust was passing the external flue cyclone.

When operated the scrubber appeared to be effective in avoiding dust emissions. Because of the diffuse upward discharge of gas from the system, it was not possible to obtain a quantitative measure of the actual dust content.

Corrosion of the water recycle piping in this scrubber system was severe because of SO_2 from the regenerator which was remixed with the boiler gas in the stack.

Cyclone Fines Return System -

The gasifier cyclone fines return system operated well during much of the run in spite of several deficiencies.

Operation was trouble-free for most of the first 109 hours of gasification and for the final 127 hours. During other periods there were a number of upsets. Two problems were encountered in the first period:-

(1) Dust worked its way back to the pneumatic control system through a pressure measurement line and caused stoppage. This problem did not recur after installation of a fine filter and additional N_2 bleed in the pressure line.

(2) Residual pressure remained in the conical dust receiver after a transfer of solids when the butterfly valve to the cyclone drain opened to begin refilling, this pressure caused a surge of gas back up the cyclone leg and upset the cyclone operation. The result was a burst of fines into the boiler after each transfer operation. These puffs were evident in the peaks observed in boiler SO_2 emission. This difficulty was removed by installation of a delay device which allowed pressure to discharge down stream from the conical receiver before the valve to the cyclone could open.

Most of the problems met during the mid run period were caused by chips, flakes, and chunks which found their way into the cyclones and transfer system after each temporary shutdown and decoking operation. It was necessary to disconnect vessels and lines on several occasions to remove these flakes and chunks. In other cases these solids prevented good operation of the butterfly valves. When the butterfly valves failed to seat properly before a transfer, N_2 gas again blew back through the cyclone and sent dust to the boiler.

Installation of a chunk trap in the elutriator drain during the run improved operation a great deal. Installation of additional chunk traps in the conical vessels were planned as a further aid.

Efficiency of the cyclones themselves deteriorated during the run. In the initial period efficiency was fairly high, and only a small amount of very fine solids entered the boiler. In later stages of the run the efficiency deteriorated, and a considerable quantity of quite coarse solids entered the boiler.

Inspection of the cyclones at the end of the run revealed them to be nearly completely choked with a mixed deposit of lime and carbon in the annular space between walls and gas outlet tube. The steel liners which had been installed before Run 5 were severely burned and distorted. The silicon carbide gas outlet tubes were strong, smooth, and intact. It is evident that demonstration of an effective way to maintain cyclone efficiency must remain an important problem area of this work.

In view of performance of the process, it is clear now that recycle of cyclones fines to the gasifier without a means to achieve their regeneration is not desirable. The fines, with their large surface area, pick up a considerable load of sulphur and make a number of cycles through the gasifier and cyclones without entering the regenerator. If any of these fines escape the cyclone, they enter the boiler and cause loss of sulphur removal efficiency. A means of providing preferential regeneration of the fines is a desirable process feature.

Regenerator Operation -

Defluidisation of the regenerator bed occurred twice during gasification in Run 5 and once under combustion conditions. The cause of this behaviour has not been established, but gas by-passing in some manner is suspected. The effects of such by-passing would be aggravated by lack of fines in the regenerator solids which would increase minimum fluidisation velocity. Two of the ways in which by-passing could occur are leakage of gas through the solids circulation passages and leakage through cracks. It is possible for air to enter a crack in the refractory near the bottom of the bed, travel upward through the crack, and return to the vessel higher up. A small crack was observed in the regenerator wall and patched during the run, but it did not appear large enough to account for the troubles observed.

The fact that the tendency to defluidise became more severe with time during the gasification periods involved suggests that it may have been related to another time dependent factor such as the increase in gas space pressure which took place as gasifier outlet passages gradually fouled. Leakage of air back into the gasifier-to-regenerator solids transfer line and up the unused portion of the cyclone fines return leg could follow such a course. Such leakage would have to pass into the cyclone past the steel sleeve insert which had been dropped into the cyclone leg as a seal. However, distortion of the steel sleeve by heat following gasifier decoking is a distinct possibility. On the other hand, such a loss of air near the regenerator bottom does not accord with the apparent low SO_2 concentration measured in the regenerator gas. Indeed, sulphur material balance considerations imply that the gas flow through the regenerator was higher than that supplied by the regenerator air blower. It is probable that further tests will be needed to establish the cause of this unusual regenerator behaviour.

Boiler deposits -

Boiler deposits found within the boiler were of two types:

(1) Loose accumulations of dust or coarser particles in the soot trap areas at the boiler ends.

(2) Agglomerated deposits formed from very fine particles which build up at the inlets to the first pass of small fire tubes.

The loose accumulations of particles do not appear to present a long term problem. They would be subject to easy removal by normal soot blowing techniques.

The agglomerated fines represent a potential problem area which requires additional study to define its severity in large scale equipment. Certainly it is an inconvenience in our pilot plant equipment. However it must be stressed that the fire tube boiler used for our pilot plant tests is in no way typical of a water tube power generation boiler and the problems we have experienced may be typical only of the particular boiler we are using. From the point of view of deposit buildup the pilot plant boiler is far from ideal. Deposits are very local in nature; being found only at the inlets to the first set of water cooled fire tubes, where the gases change direction by 180 deg. in a downward direction.

They did not form after the first few cm of tube length nor were they found to any significant degree on a test probe inserted radially into the gas stream at the end of the main fire tube as a simulation of a superheater tube in a water tube boiler. These deposits evidently form from fine particles which are in a sticky state following their passage through the flame. The growth of the deposits was faster during the first 109 hours of Run 5 gasification than during the final 127 hours. Whether this difference was due to the presence of BCR 1691 stone in the first period or to the eroding effect of a higher concentration of coarse particles during the final period remains to be established. It is possible that deliberate injection of a small amount of coarse stone into the boiler could prevent deposit formation. No deposits were found at the tube inlets during the 111 hours of gasification in Run 2 during which a high concentration of solids passed through the boiler.

Process Performance is summarised in Table 6 which lists values of operating conditions and results for the various test periods of Run 5. Each value is the average for four hours operation. In most test periods a set of solids samples was collected for analysis.

Sulphur Removal -

The degree of sulphur removal in the pilot plant is calculated from the measured SO_2 and CO_2 contents of the boiler flue gas compared with sulphur and carbon contents of the boiler fuel. The carbon content of the pilot burner propane is considered in this calculation, as is the CO_2 released by calcination of limestone makeup.

During Run 5, sulphur removal efficiency (SRE) varied from 60 to 99%. Appendix figure B.15 shows the hourly levels of SRE along with other important variables. Figure 8 shows the effect of lime replacement rate on SRE for individual test periods. It is apparent that lime replacement has an effect on SRE.

The effect of other variables are less clearly defined. In particular, increasing gasifier bed depth did not produce the expected improvement in SRE over results obtained in earlier runs with shallower beds.

It appeared in this run that the beneficial effect of increased bed depth was offset by increasing fines loss rate. The loss of fines hurts sulphur removal efficiency in two ways. Firstly it removes the high surface area fraction of the lime which has greatest potential for sulphur pickup, and secondly, when sulphur laden fines enter the boiler they partly regenerate to contribute SO_2 to the flue gas.

The effect of bed depth was also obscured by the fact that deterioration in cyclone performance made it difficult to operate with deep beds at very low lime replacement rates.

Table 6

Operating Conditions During Run 5 Test Periods

Day Time	Temperature deg. C Gasifier	Temperature deg. C Regenerator	Lime Replacement mol Ca/mol S	Gasifier Depth cm water	Fuel Rate kg/hr	Sulphur Removal %	Regenerator SO_2 Conc. Vol%	Regenerator SO_2 % of S Fed	Stone
2.2130	883	1047	.62	48.8	181	77.5	4.1	32.7	Denbighshire
3.0530	886	1063	.64	52.3	180	72.8	4.4	43.3	"
3.1530	884	1061	.48	54.6	179	74.8	3.0	28.0	"
3.2130	894	1066	.62	45.7	183	60.2	2.6	23.1	"
5.1030	852	1053	1.76	45.7	181	95.0	5.1	44.7	BCR 1691
6.0730	856	1055	1.19	44.7	174	94.5	4.6	36.9	"
12.0830	877	1050	.81	53.6	181	79.1	4.1	48.3	"
12.1530	870	1036	1.14	56.6	180	84.6	3.7	43.9	"
12.1930	871	1055	1.07	55.4	179	85.5	4.1	46.5	"
13.0330	868	1031	1.01	46.7	178	83.4	4.0	47.0	"
13.0630	876	1024	.87	46.2	177	87.2	3.9	48.5	"
17.1230	863	1052	.90	63.5	181	84.0	4.7	46.1	"
17.1730	861	1047	1.55	64.3	182	76.9	3.8	38.4	"
21.0930	903	1062	1.16	55.9	213	93.3	3.8	32.8	Denbighshire
21.1830	889	1065	1.40	57.7	213	89.1	4.4	35.8	"
22.0730	878	1070	.96	58.4	187	91.2	4.0	38.4	"
22.1830	873	1069	1.03	61	186	85.0	3.7	34.6	"
24.0730	862	1057	2.01	62.5	186	99.3	3.7	34.4	"
25.0530	878	1060	1.48	64.3	184	93.7	3.6	35.4	"
25.1530	875	1060	1.36	65	183	90.3	3.6	34.9	"
26.0530	880	1060	1.04	63.5	184	90.8	3.2	32.3	"
26.1130	866	1060	1.32	63.2	186	92.0	3.4	34.5	"
26.1830	872	1060	1.41	64.5	186	86.1	3.4	33.3	"

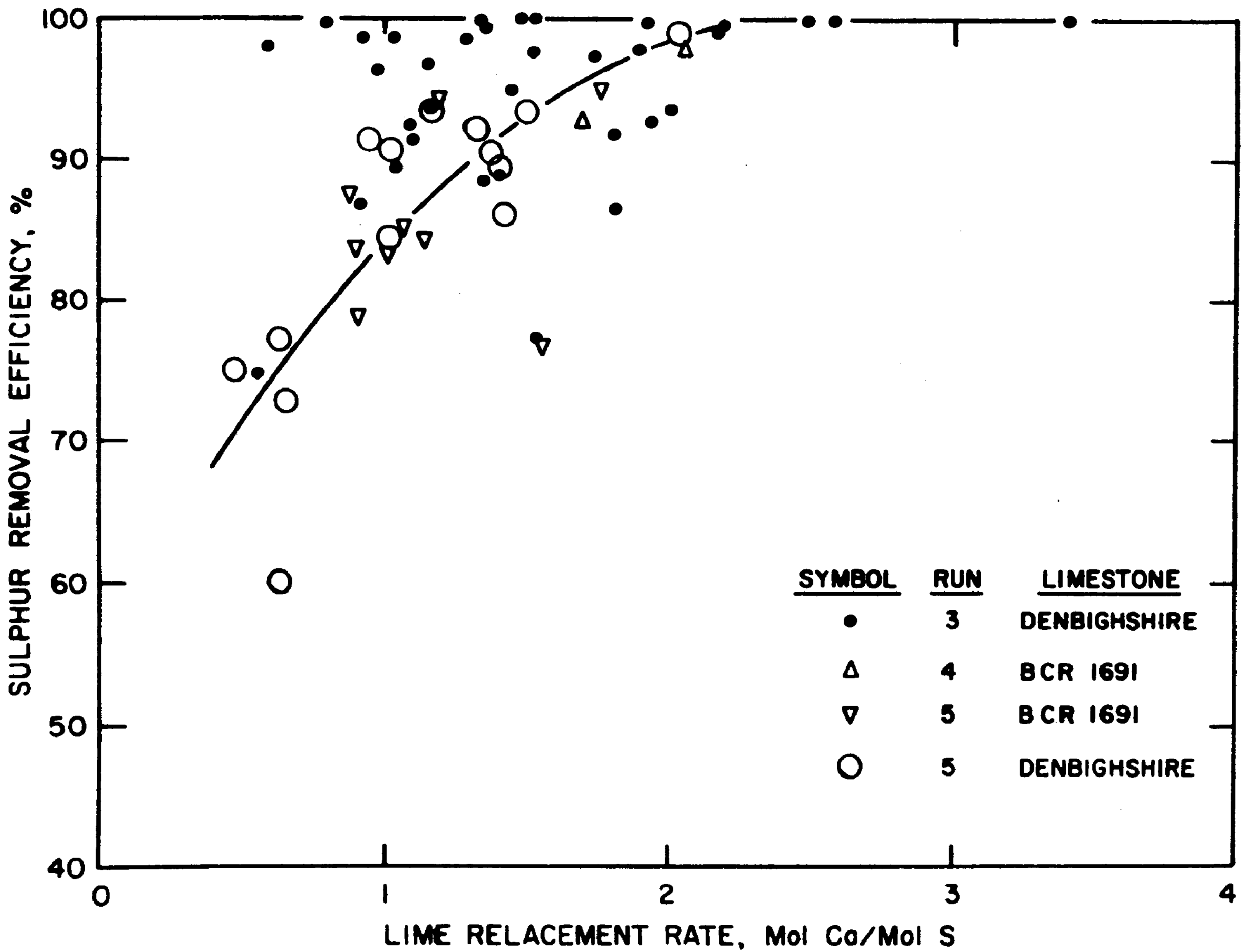

Figure 8 CAFB Pilot Plant Sulphur Removal Efficiency

Results are summarised in Table 7. The water scrubber on the stack top was able to catch the worst of the material which escaped the cyclone, but since the quantity recovered by the scrubber was not measured, it is included with the stack losses.

Table 7

Summary of Solids Loss - Run 5

Time Day.Hour	Limestone Feed	Make-up Rate (kg/hr)	Gasifier Fluid Bed Depth (cm)	Loss Rate (kg/hr) Gasifier	Stack	Bed SiO_2/CaO Ratio
3.1530	Denbighshire	3.3	69.9	0	0	0.006
3.2030	"	3.3	57.2	2.4	1.2	0.006
5.1030	BCR 1691	27.2	76.2	13.6	10.6	0.203
6.0730	" "	12.3	67.8	8.3	6.9	0.236
12.1730	" "	11.8	80.0	14.1	10.1	0.224
13.0330	" "	11.8	70.1	9.8	6.7	0.217
17.1130	" "	9.7	90.7	8.3	4.4	0.199
17.1730	" "	19.5	94.2	15.1	11.7	0.185
21.1830	Denbighshire	10.0	78.0	8.6	5.4	0.053
22.0630	"	10.0	72.9	7.8	4.7	0.053
22.1730	"	7.7	76.2	6.0	2.9	0.038
25.0530	"	11.3	90.7	7.4	0.1	0.023
25.1430	"	11.3	92.5	7.7	2.0	0.016
26.0430	"	7.9	90.7	6.0	0.5	-
26.1730	"	7.9	99.6	5.4	0	-

The test at 900 deg. C gasifier temperature on day 21 produced sulphur removal efficiency over 90% and demonstrated the feasibility of operating at this temperature with a low air/fuel ratio. The fuel rate in this test was 213 kg/hr (469 pounds per hour), the highest yet used in the pilot plant. Pressure drop in the boiler prevented increasing fuel rate still further to test completely adiabatic gasifier operation without flue gas recycle.

Metals Retention -

Comparing the metals content of spent lime with the metals content of fresh limestone and fuel oil, it is possible to estimate the degree of metal retention by the solids. Figs. 9, 10 and 11 compare the retained weights of vanadium, sodium, and nickel with quantities of these metals fed during various operating periods of the pilot plant.

Vanadium retention is essentially complete, in agreement with predictions based on batch unit tests. Sodium retention was 36%, somewhat higher than the 20% level obtained in batch tests. Nickel retention, which was not studied before, averaged 75%.

There appeared to be no significant difference between metals pick up efficiencies with Denbighshire and BCR 1691 stones. However because of differences in stone loss rates, there is a difference in absolute metal retention levels in the unit. With Denbighshire stone there was practically no lime loss from the system. That lime which escaped the gasifier cyclones was caught either in the boiler or by the external flue gas cyclones. However with BCR 1691, losses amounted to as high as 40% of the lime replacement rate and of course any associated metals were lost as well. While not affecting the ability of the CAFB gasifier to remove vanadium as a source of high temperature corrosion of boiler superheater tubes, the loss of metals on lime particles would be a pollution factor which could be reduced even further by increasing the efficiency of the particulate removal portion of the system.

Solids Losses -

To obtain a more comprehensive picture of solids losses during Run 5, both the amount of material emitted by the gasifier into boiler and the amount escaping the external cyclone into the stack have been computed.

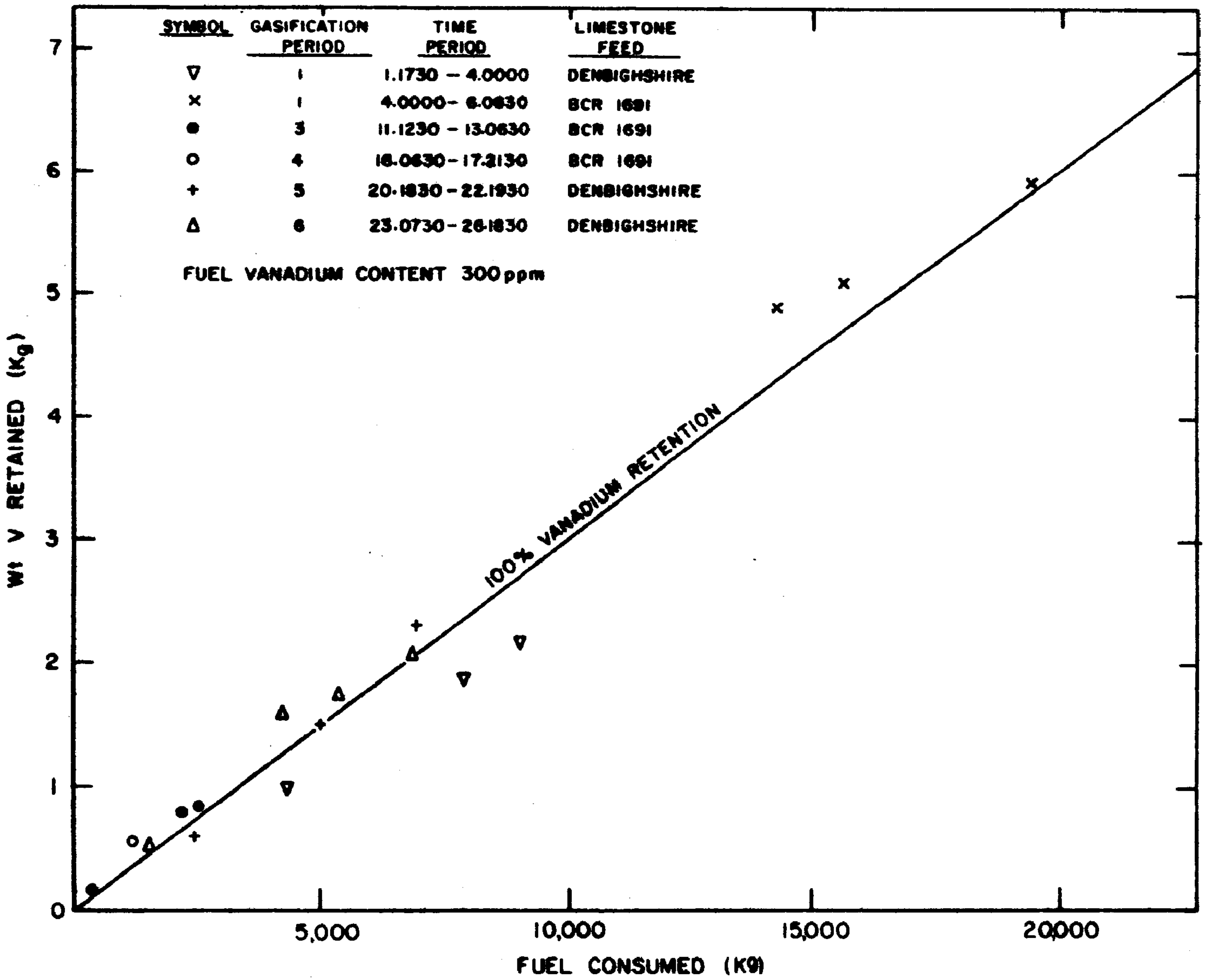

Figure 9 Vanadium Retention (Run 5)

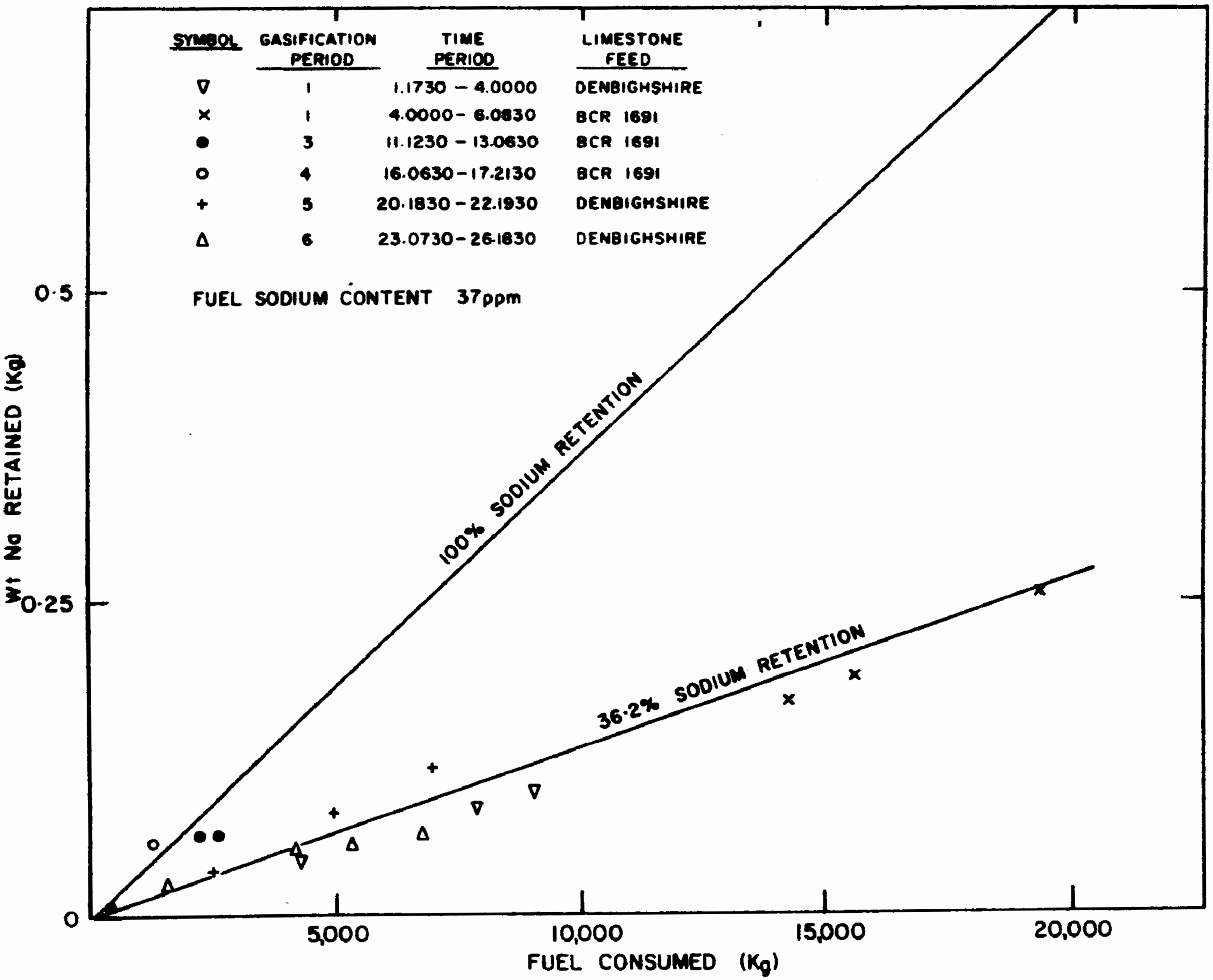

Figure 10 Sodium Retention (Run 5)

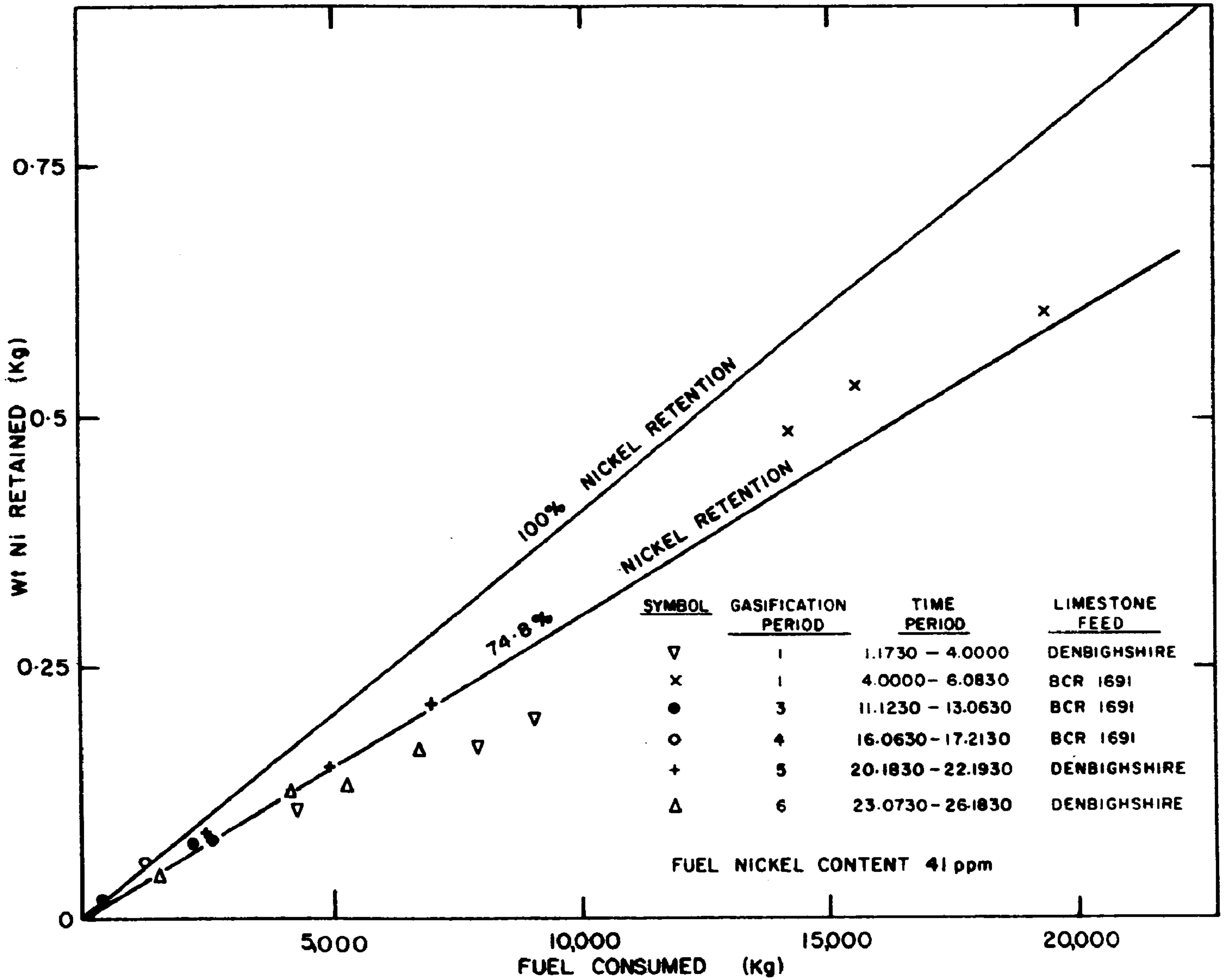

Figure 11 Nickel Retention (Run 5)

Taking gasifier losses first, it is apparent that these varied considerably during the run. Previously, in batch units it had been found that gasifier loss rate was dependant on make-up rate and bed depth (Reference 1). In Run 5, however, the situation was more complicated since cyclone performance was known to have deteriorated sometime during the run. This was evident from the after-run inspection which showed the cyclones to be in poor condition. Statistical analysis of gasifier loss rates showed an inconsistancy between the first and second data points (3.1530 and 3.2030). Also all later variations for both stones could be explained by changes in bed depth and make-up rate in a single correlation.

From this analysis, the following was deduced. Firstly, cyclone performance deteriorated between the 3.1530 and 3.2030 data points and did not alter appreciably thereafter. The reason for the deterioration is not as yet clear. Secondly, gasifier loss rates at constant cyclone performance were shown to depend on make-up rate and bed depth. Whilst the effect of make-up rate was similar to that observed in batch units, bed depth appeared to be less significant. Thirdly, gasifier loss rate was independent of limestone type in this instance. This would not always be the case. Here, it would appear that the cut-off point for the reduced performance cyclones and the attrition patterns for the two limestones is combined to cause this phenomenon.

For stack losses, a very different picture emerged when these results were examined in detail. The two limestones behaved differently. Under all conditions examined, stack losses were small when the bed was composed mainly of Denbighshire limestone. With predominantly BCR 1691 (5.1030 to 17.1730), however, they were appreciable. They were also higher when of the order of 20% of the bed as estimated from SiO_2/CaO ratios was BCR 1691 (21.1830 to 22.0630). Statistical analysis also showed that stack losses from a BCR 1691 bed correlated with make-up rate and bed depth.

Since the performance of the stack cyclone did not change during the run, these variations in stack losses can only be explained if it is accepted that BCR 1691 produces a fraction of material of much smaller particle size than Denbighshire. This has been indicated from Run 4 and batch test data.

Since the stack cylcone was designed to have the same efficiency as the gasifier cyclones, the results from Run 5 indicate that a gasifier following the same principles for solids handling as the pilot plant and with gasifier cyclones fully operational would give negligible gasifier losses with a limestone of Denbighshire type and gasifier losses of the order of stack losses with a limestone of BCR 1691 type.

Bed Homogeneity -

Both Denbighshire and BCR 1691 limestones were tested in Run 5. A measure of bed homogeneity with respect to limestone type was obtained by analysing bed samples for silica and calcium oxide and comparing the ratio of these two compounds. The silica to calcium oxide ratio for the high purity Denbighshire stone is much lower at 0.006 than BCR 1691 at 0.28. Results are summarised in Table 8.

For the first two data times, the system was completely homogeneous since only Denbighshire stone had been added. During the subsequent BCR 1691 test period, the bed always had a SiO_2/CaO ratio below that of the raw stone. Also no persistent increase in the ratio was observed.

During the final Denbighshire test period, a reduction in SiO_2/CaO ratio took place and the ratio of the raw Denbighshire stone was approached. Material from the stack cyclone gave similar results to that from the regenerator cyclone throughout. During the BCR 1691 test period, the SiO_2/CaO ratio in these fines was generally higher than that for the raw stone.

The significance of the Si/Ca ratios during the BCR 1691 tests is somewhat obscured by the fraction of Denbighshire stone which remained in the bed following changeover to BCR 1691.

We believe that the low SiO_2/CaO ratios in the bed during BCR 1691 tests was due to residual Denbighshire stone. The absence of any increase in ratio over the period can be attributed to the addition of Denbighshire stone during burn-out and maintenance periods between 6.0730 and 12.1630 and also 13.0600 and 17.1130.

We discount the possibility of a preferential loss of silica from the beds being the cause of the low results even though higher SiO_2/CaO ratios were observed in cyclone fines. The reason being that beds composed only of BCR 1691 in batch tests and Run 4 showed the opposite effect in that silica was concentrated in the bed. In those instances the fines which

Table 8

Silica/Calcium Oxide Ratios (Run 5)

Time Day Hour	Limestone Feed	Gasifier Bed SiO_2/CaO	Regenerator SiO_2/CaO	Stack Cyclone SiO_2/CaO	Regenerator Cyclone SiO_2/CaO
3.1530	Denbighshire	0.006	0.006	0.006	0.006
3.2030	"	0.006	0.006	0.006	0.006
5.1045	BCR 1691	0.203	0.215	0.283	0.305
6.0730	"	0.236	0.218	0.261	0.300
12.1630	"	0.228	0.258	0.301	0.283
12.1800	"	0.224	0.225	0.316	0.286
13.0400	"	0.217	-	0.313	-
13.0600	"	0.242	0.225	0.312	0.304
17.1130	"	0.199	0.106	0.303	0.266
17.1800	"	0.185	0.214	0.301	0.273
21.0700	Denbighshire	0.056	-	-	0.009
21.1800	"	0.053	-	0.015	-
22.0715	"	0.053	-	0.007	-
22.1745	"	0.038	-	0.010	-
25.0530	"	0.023	0.020	0.006	0.006
25.1430	"	0.016	-	-	-

were trapped also showed a higher ratio than the raw stone indicating that the material lost completely from the system was rich in calcium. Assuming the same to have happened here with the BCR 1691 fraction of the bed, then the higher SiO_2/CaO ratio in the fines can be explained by BCR 1691 being lost preferentially.

After the final change to Denbighshire feed, the SiO_2/CaO ratios indicated that some BCR 1691 was present in the bed to the end of Run 5, albeit in ever decreasing amounts.

Particle size of solids -

The particle size distribution of solids in the reactor beds depends on size of the feed, particle attrition, and effectiveness of the cyclones in returning fines. Sieve analysis of a number of gasifier and regenerator samples are presented in Appendix B. Histograms of the average stone feed and two sets of gasifier and regenerator bed samples are given in Figure 12.

The two sets of bed samples illustrated represent extremes of the samples taken. Performance of the fines return system was poor at 13.0400, and fines were being lost from the unit. The cyclones and fines return system were operating relatively well at 21.1600 as shown by the larger fraction of small particles.

These figures show that the fraction of particles in the 250 to 1400 micron size range was increased in the unit at the expense of both larger and smaller sizes. The fraction below 250 microns in the gasifier was lost altogether while the quantity of material above 1400 microns was reduced.

Poor performance of the fines return system, as at 13.0400 causes significant loss of particles as large as the 600-850 micron range. Regenerator and gasifier beds were quite similar in size distribution with the regenerator showing a slightly higher fraction of fines.

Figure 13 shows the variation, during the run, of the fraction of bed in the size range below 600 microns.

This figure indicates a rapid deterioration of fines return effectiveness during the early part of the run. It indicates that fines return was restored during decoking before day 21 startup but again declined toward the end of run.

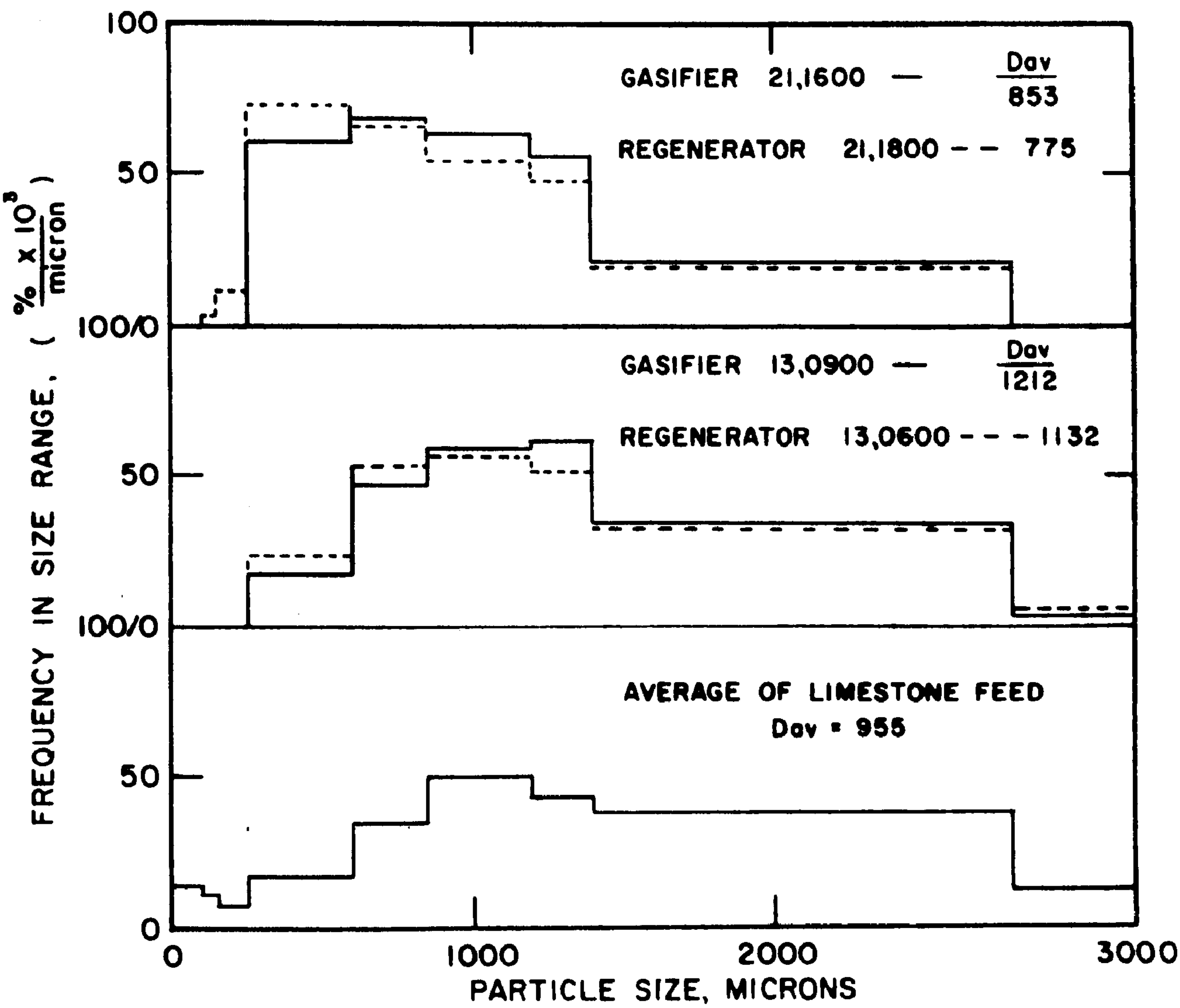

Figure 12 Size Distribution of Fluid Beds
(Run 5)

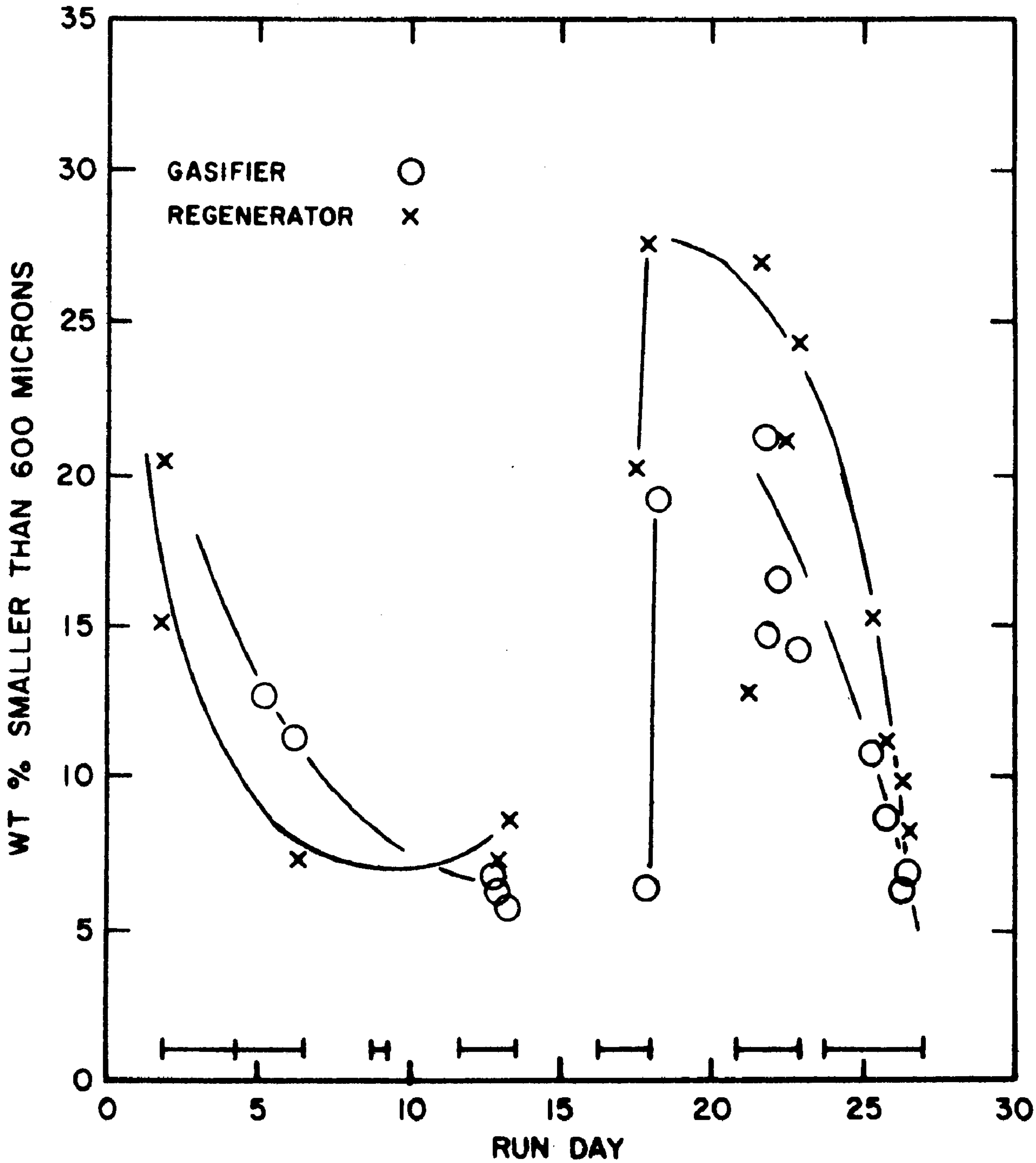

Figure 13 Gasifier and Regenerator Fines below 600 microns (Run 5)

A similar pattern is shown in Figure 14 where average particle size of the bed is plotted against run time. The particle size here is calculated from sieve analysis, using the relationship:

$$d_{AV} = \frac{W}{W/d}$$

which gives a surface area mean particle size.

Nitrogen Oxides -

The concentrations of nitrogen oxides measured in the boiler flue gas during Run 5 are compared in Table 9 with values measured in previous tests. All samples were taken from the boiler flue gas by means of gas sample bags and analysed off line in the laboratory. A chemiluminescence method was used for Run 5 samples.

The ASTM D1608 phenol-disulphonic acid method was used for the other samples. Laboratory cross checks have indicated that the two methods are in agreement.

Run 5 results agree generally with earlier gasification test results. All of the gasification results have lower NO_X concentrations than the tests with the oil burner. This improvement during CAFB gasification is probably due to a combination of the effects of two stage combustion and the use of flue gas recycle. It is possible that nitrogen compounds in the fuel are converted to a harmless form during gasification, and that flue gas recycle reduces maximum flame temperature which reduces equilibrium NO_X concentration.

Even with the original oil burner operation the NO_X concentrations were low when compared with concentrations in the flue gas of large power station boilers. This difference in overall level is believed to be caused by the close proximity of the flame to the large water cooled surface in the fire tube boiler used in the pilot plant. Therefore, although the reduction in NO_X level caused by CAFB gasification is believed to be realistic, the absolute low level achieved probably would not be reached in large power generation boilers where flame temperatures are much higher.

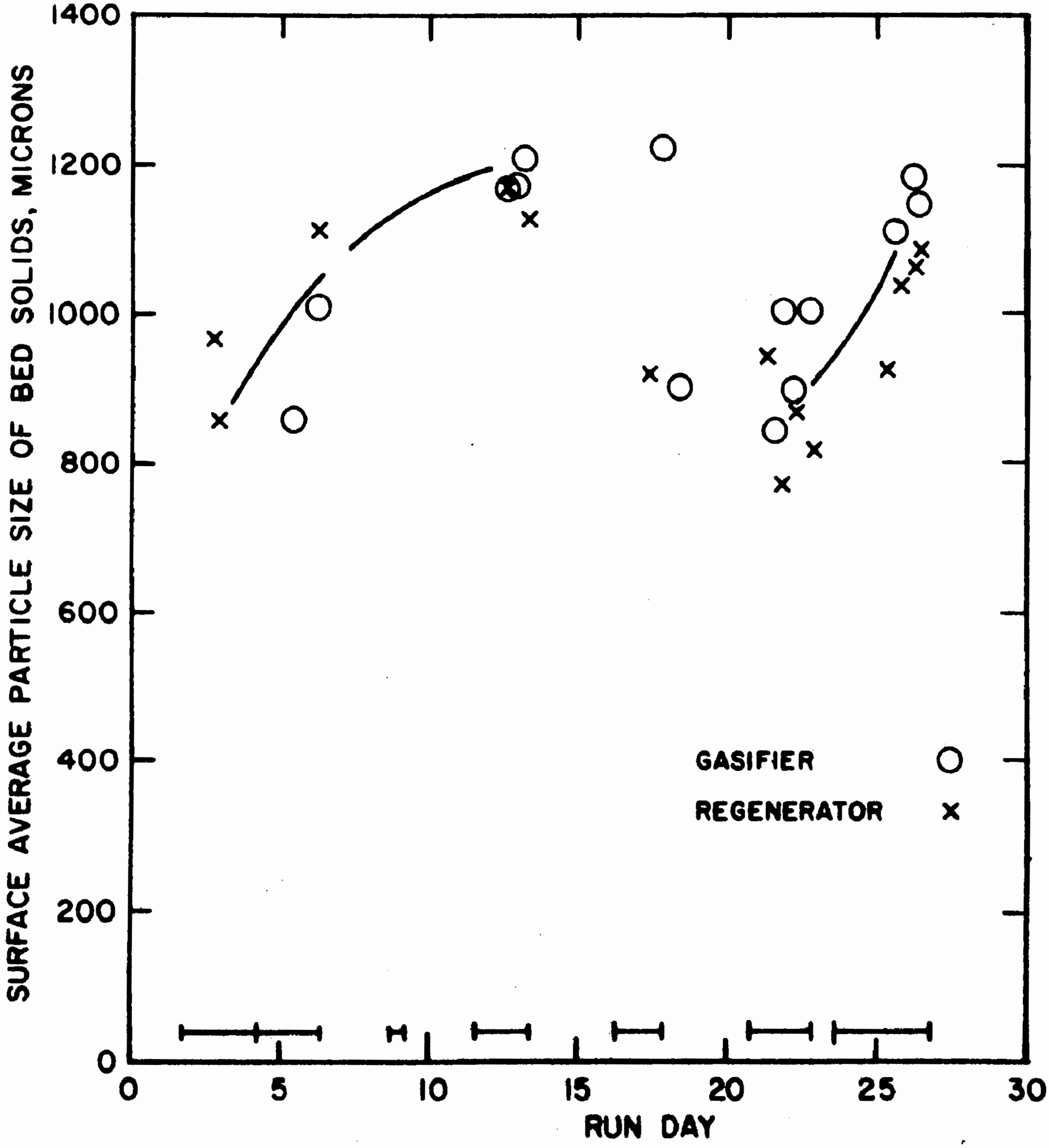

Figure 14 Average size of bed Particles (Run 5)

Table 9

Nitrogen Oxides in CAFB Boiler Flue Gas

Operating Mode		Sample Date-Time		Method	NO_x cm^3/m^3	Flue Gas O_2 Vol%	Oil Rate kg/hr
Oil Burner-low fire		May 1971		ASTM D1608	256	3	-
" "		" "		"	249	"	-
Oil Burner-High fire		" "		"	280	"	-
" "		" "		"	266	"	-
Gasification	Run 1	August 1971		"	179	-	-
"	"	" "		"	155	-	-
"	"	" "		"	172	-	-
"	"	" "		"	200	-	-
"	Run 3	1 Dec 1971	15:55	"	126	2.9 - 3	164
"	"		16:00	"	120	"	"
"	"		16:05	"	130	"	"
"	"	3 Dec 1971	09:50	"	163	2.4 - 2.3	175
"	"	" "	09:55	"	173	"	"
"	"	" "	10:05	"	169	"	"
"	"	" "	10:10	"	163	"	"
"	"	6 Dec 1971	13:45	"	237	1.0	145
"	"	" "	13:50	"	186	"	"
"	"	" "	14:00	"	181	"	"
"	"	" "	14:05	"	166	"	"
"	Run 5	9 Feb 1973	15:00	Chemiluminescence[a] 101, 108,	110	3.0	179
"	"	17 Feb 1973	00:00	"	180	3.7	183
"	"	22 Feb 1973	16:10	"	185	2.9	181

(a) Thermo Electron Corporation NO_x Analyser Model 10 (a)

Thermal Behaviour -

Most of the heat released by partial combustion of fuel in the gasifier is retained as sensible heat in the gas going to the burner. The heat release in the gasifier has been estimated from thermal equations for the masses and heat contents of the various streams entering and leaving the gasifier. Table 10 lists results for Run 5 test conditions. The equations used are explained in Appendix I.

Heat losses from the pilot plant gasifier bed, based on reasonable values of thermal conductivity and heat transfer coefficients, amount to about 1% of the heat released in the gasifier. Depending on lime replacement rate, about 2 to 4% of the heat released goes to calcine stone and raise the lime to gasifier temperature. At the air/fuel ratio employed, which did not deviate much from 20% of stoichiometric, the heat release per pound of fuel was estimated to be approximately 7211 kJ/kg (3100 BTU/lb) corresponding to 360,530 kJ/kmol (155,000 BTU/mole) of oxygen.

Figure 15 shows the observed variation of fuel heat release with air/fuel ratio, and compares the measured values with a line calculated for a release of 360,530 kJ/kmol (155,000 BTU/mole) of oxygen. Values from all the pilot plant runs to date fall along this line. This value of 360,530 kJ/kmol O_2 agrees well with heat release calculated from the fraction of carbon and hydrogen oxidised and the CO/CO_2 ratio formed in the gasifier. Details of this calculation also appear in Appendix I.

Product Gas Composition -

Four samples of the gasifier product vapour were collected during the run and analysed by gas chromotograph. This analysis gives dry gas composition on a water and liquid hydrocarbon free basis. Results are listed in Table 11.

Table 10

Heat Release in CAFB Gasifier

Run 5

Run.Time	Air/Fuel % of Stoichiometric	Heat Lost % of Heat Release	Calcination and Stone Heating % of Heat Release	Heat From Regenerator % of Heat Release	Gasifier Heat Release kJ	Oil Heat Release kJ/kg oil	O_2 Heat Release kJ/kmol
2.2130	19.8	.73	1.49	4.10	369	7378	370902
3.0530	20.4	.78	1.58	5.33	360	7215	352294
3.1530	20.6	.71	1.16	5.14	369	7436	359281
3.2130	20.3	.66	1.49	4.64	378	7432	363000
5.1030	21.3	.83	4.42	4.88	384	7620	355338
6.0730	20.4	.77	3.02	4.31	363	7529	367334
12.1530	20.6	.94	3.06	5.66	360	7222	347809
12.1930	20.6	.97	2.93	7.57	349	7036	339075
13.0330	20.7	.80	2.60	5.71	369	7478	359055
12.0830	20.6	.86	2.15	5.39	369	7362	354394
17.1230	21.3	1.09	2.38	5.31	366	7276	338761
17.1730	21.9	1.04	4.03	5.56	375	7411	335653
21.0930	19.7	.80	3.05	4.96	407	6901	348456
21.1830	19.0	.82	3.67	4.07	410	6925	362649
22.0730	19.6	.86	2.43	5.02	372	7132	362491
22.1830	19.8	.88	2.58	5.14	369	7159	358509
24.0730	19.2	.97	4.92	5.30	378	7315	378980
25.0530	19.2	1.06	3.66	5.89	372	7280	376956
25.1530	19.1	1.07	3.40	5.62	366	7173	373311
26.0530	18.9	1.07	2.63	6.06	366	7090	372893
26.1830	18.7	1.16	3.57	5.74	366	7085	375721
13.0630	20.8	.80	2.22	5.77	369	7501	357925
26.1130	18.1	1.10	3.37	5.98	363	7034	386546

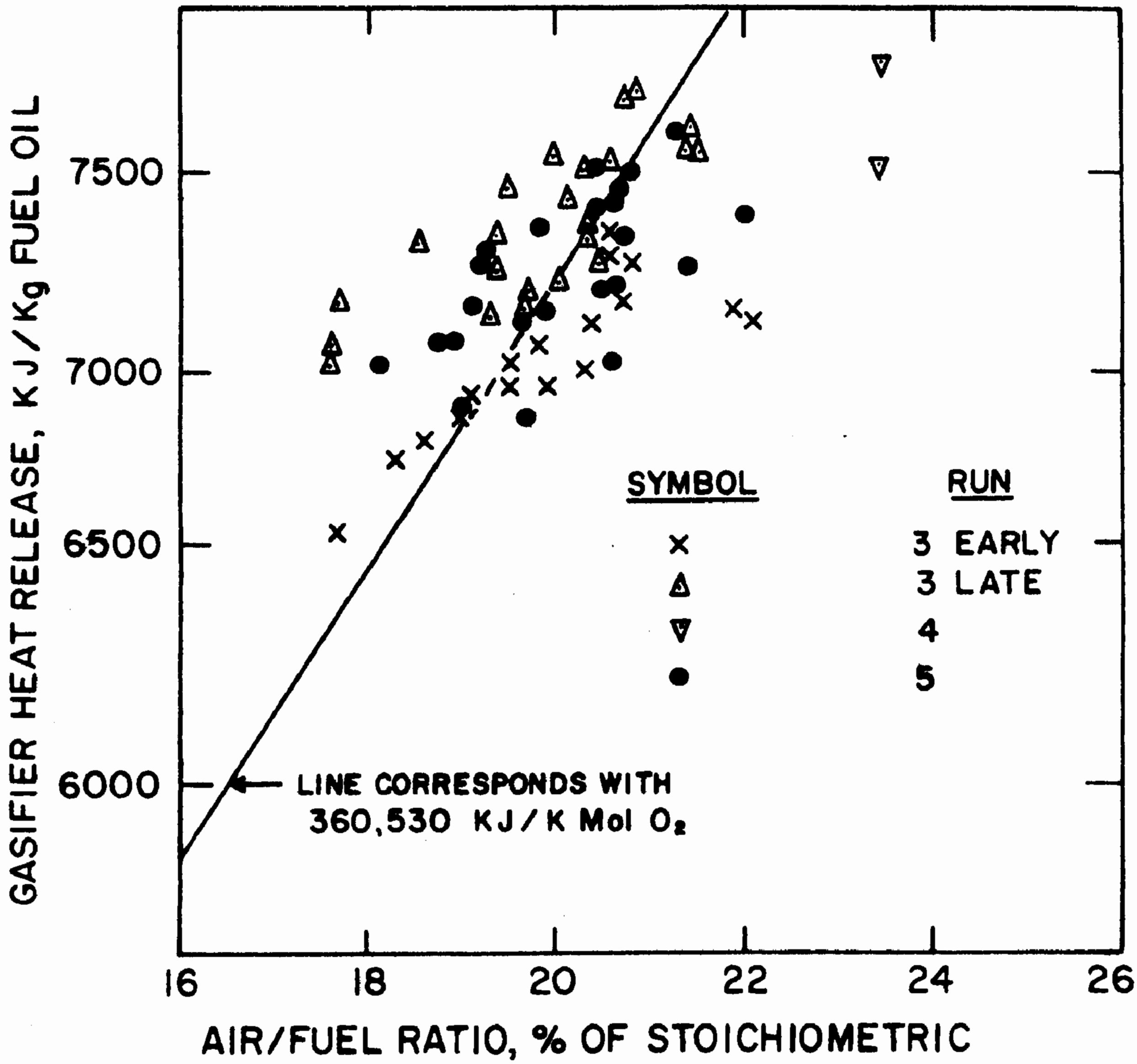

Figure 15 Heat Release of Air/Fuel Ratio
During Gasification

Table 11

Product Gas Composition

Run 5

Sample Time	22.1030	22.1745	26.0400	26.1800
Composition, Vol % (Air Free Basis)				
N_2	61.7	62.5	63.4	64.7
CO_2	10.82	10.89	10.88	10.17
CO	8.36	7.52	9.36	8.97
H_2	6.86	7.75	6.42	5.80
CH_4	6.75	6.29	6.09	6.02
C_2H_4	5.36	4.83	3.81	4.38
C_3H_6	-	.11	-	-

A material balance calculation on the product gas composition and unit feed rates permits an estimate of the quantity and composition of that portion of the product vapour which is missed by the gas chromatograph because of condensation in lines or in the sample container.

In this calculation nitrogen and oxygen balance are forced to 100% and the amount of hydrogen and carbon not accounted for is assumed to make up the liquid fraction. The flue gas recycle stream was assumed to contain the same H_2O/CO_2 ratio as the boiler products of combustion. Table 12 summarises results of this calculation. Details appear in Appendix Table L-1.

The results are fairly consistent in indicating the fractions of carbon and hydrogen oxidised and the quantity of carbon which goes either to coke or heavy hydrocarbons. The results on hydrogen disappearance show greater variability however.

Table 12

Summary of Gasifier Component Distributions

(Calculated from Flows and Product Gas Composition)

Sample Time	22.1030	22.1745	26.0400	26.1800
Oxygen In, % of Total				
With Air	67.7	68.4	66.9	66.4
With Flue Gas Components	23.7	23.6	23.0	23.6
From Solids Reactions	8.6	7.9	10.0	10.0
Oxygen Out, % of Total				
As Sulphate on lime	1.7	1.6	1.3	1.5
As Carbon Oxides	83.7	81.6	84.4	77.9
As H_2O (by Difference)	14.5	16.7	14.4	20.5
Hydrogen In, % of Total				
In Oil Feed	94.1	94.2	94.3	94.3
In Flue Gas H_2O	5.9	5.8	5.7	5.7
Hydrogen Out, % of Total				
As Dry Gas Components	82.2	79.6	67.1	49.8
As H_2O	13.8	15.8	13.6	18.7
As Heavy Components (By Difference)	4.0	4.6	19.3	31.4
Carbon To Gasifier, % of Total				
In Oil Feed	93.2	93.5	93.5	93.1
In Flue Gas, C oxides	5.8	5.6	5.5	5.5
In Stone	1.0	0.9	1.0	1.3
Carbon From Gasifier, % of Total				
As Oxides from Gasifier	40.6	38.7	41.5	37.1
As CO_2 from Regenerator	1.0	0.4	0.2	0.3
As Hydrocarbon in Dry Gas	37.0	33.5	28.1	28.6
As heavy Components (By Difference)	21.5	27.4	30.3	34.0
C oxidised, % of Feed	35.8	34.0	37.0	31.9
H oxidesed, % of Feed	14.6	15.8	14.4	19.9
C in Heavy Components, % of Feed	23.9	30.0	33.1	37.4
H in Heavy Components, % of Feed	4.3	4.8	20.4	33.3
CO/CO_2 in fresh oxides	1.102	.97	1.22	1.36
H/C in Heavy Components	.29	.26	.99	1.43
Air/Fuel Ratio, % of Stoichiometric	19.5	19.8	19.3	18.3

This variability is probably due to the method of calculation which depends on finding small differences between relatively large numbers.

In particular, the hydrogen/carbon ratios of .29 and .26 calculated for the heavy components on day 22 appear to be unreasonably low. It is unlikely that the true H/C ratios could be much less than 1.0. These results show the desirability of obtaining accurate samples of the total gasifier product, including light and heavy components. However, collection of such a sample is quite difficult in practice, and will require development of a suitable quench and recycle system to collect the liquid fraction without plugging.

Regenerator -

Performance of the regenerator during Run 5 appeared to be less satisfactory than in earlier tests and was somewhat inconsistent. It was disappointing in that sulphur concentration in the off gas and selectivity of calcium sulphide oxidation to calcium oxide plus SO_2 appeared to be much lower than the levels which earlier runs had shown to be possible. Results were inconsistent in that the apparent sulphur production rate of the regenerator could account for only about half the sulphur being absorbed in the gasifier.

Furthermore, SO_2 release based on gas analysis did not agree with SO_2 based on solids analysis. Obviously this matter requires additional study to locate the cause of the discrepancy.

The run data shown in Appendix Figure B-16 are the gas analysis based values of SO_2 concentration and selectivity.

Table 13 compares regenerator sulphur emission figures based on gas analysis with values calculated from solids analysis. Solids compositions were those of the gasifier and regenerator beds. It is apparent that sulphur production values based on the solids analysis are much higher than the gas analysis based figures. Similarly, the calculated values for oxidation selectivity are much higher when based on solids analysis. The solids based selectivities for this run are in good agreement with the gas analysis based selectivities of Run 3 (Figure 38 of Reference 1).

Table 13

Summary of Regenerator Performance

Run 5

Time	S in Solids wt.% Gasifier	S in Solids wt.% Regenerator	Selectivity % CaS to CaO Gas Analysis	Selectivity % CaS to CaO Solids Analysis	Regenerator Sulphur Output By Gas Analysis kg/Hr.	Regenerator Sulphur Output By Gas Analysis % of Fed	Regenerator Sulphur Output By Solids Analysis kg/Hr	Regenerator Sulphur Output By Solids Analysis % of Fed	Gasifier SRE %
3.1530	2.89	1.82	27.8	50.5	1.30	30.8	2.52	59.4	72.8
3.2230	2.72	1.81	37.6	52.9	1.73	39.7	2.46	56.6	60.5
5.1030	3.66	1.59	40.9	70.5	1.94	44.3	3.55	81.1	93.8
6.0730	3.72	1.84	40.5	61.0	1.63	38.9	2.65	63.1	96.6
12.1630	2.88	1.85	30.4	59.2	1.59	36.5	3.36	77.1	84.6
12.1830	2.80	1.80	32.8	53.5	1.76	40.7	3.03	70.2	86.1
13.0430	2.76	2.06	37.2	40.7	2.06	47.8	2.29	53.4	87.9
13.0630	2.53	1.83	34.2	40.7	1.92	45.2	2.33	54.8	88.2
17.1130	3.11	1.00	44.0	78.1	2.26	51.7	4.36	99.5	80.5
21.0730	2.69	1.78	28.2	50.0	1.38	26.9	2.62	50.9	92.9
21.1830	2.85	1.75	50.0	64.3	2.08	40.7	2.81	54.9	87.3
22.0730	2.90	2.21	32.5	46.0	1.51	34.1	2.19	49.6	90.9
22.1730	2.91	1.91	40.0	56.2	1.84	40.9	2.70	59.9	85.2
25.0530	2.74	1.88	32.0	43.7	1.66	37.4	2.34	52.3	95.7
25.1430	2.52	1.72	26.8	53.3	1.22	27.3	2.63	59.1	90.6
26.0430	2.74	2.17	27.5	34.5	1.43	31.8	1.85	40.2	88.9
26.1030	2.70	1.96	34.1	44.8	1.77	39.2	2.40	53.0	90.8
26.1800	2.87	1.96	31.3	50.6	1.52	34.0	2.64	58.9	85.4

In Run 5 the level of sulphur in the bed was relatively low compared with the sulphur content of the cyclone fines. With the external cyclone fines return system these fines had little chance to be regenerated. In Run 3 however, fines from one cyclone drained to the regenerator. This arrangement must have decreased the sulphur content of the circulating fines and reduced the effect of fines loss to the boiler on sulphur removal efficiency.

Sulphur Balance -

Use of the gas analysis figures for regenerator sulphur output leads to very low sulphur material balances for the unit. Assuming that measured values of sulphur removal efficiency are correct, it is not possible to account for the missing sulphur by assuming that it left with the lime purge stream of fines losses. A fault in the regenerator off gas analyser could explain the discrepancy, but checks and calibrations made on the instrument during the run indicated that it was functioning properly. Similarly, a larger gas flow through the regenerator than measured would cause a low estimate of sulphur production.

Such a large error in gas measurement is unlikely because air to the regenerator is measured both by orifice and gas meter, nitrogen to the solids transfer system is metered, and nitrogen to instrument bleeds is negligibly small. The only other possibility is a major leakage from gasifier to regenerator, and again this is believed unlikely. To help solve this mystery, additional analyses and measurements were planned for future runs, including checking of gas flow rate out of the regenerator by a helium tracer method, and use of a gas chromatograph to check regenerator gas composition for SO_2 and other sulphur compounds.

Plans were also made to modify the boiler flue gas sampling system to reduce further the possibility of losing sulphur in the sample lines and filters. It is possible that SO_2 is absorbed on lime dust which enters the sample line, and such absorption would produce an optimistic estimate of sulphur removal efficiency. Such errors are believed to be small however, as the lines are frequently cleaned, and spot checks with Draeger tubes (direct reading SO_2 colour change tubes) made directly through the boiler door sample point agree with the continuous reading instrument.

Run 6

Equipment Performance is described in Appendix C, Run Log and Post-run Inspection. Overall performance was improved over previous runs and the pilot plant gasifier was more easily held at steady conditions to record lined-out data.

The single fuel injector was tested during Run 6 and mechanically performed well. No symptoms of defluidisation were noted, but desulphurisation efficiency fell sharply. The divided gasifier plenum was used during the test of the single fuel injector to assess whether an induced bed circulation would assist desulphurisation. No effect was noted.

Restoration of the right hand gasifier cyclone drain into the gasifier to regenerator bed circulation line in itself posed no new problems, and none were expected as this was the original design for fines return from this cyclone. The purpose of this modification from Run 5 was intended to direct more fines into the regenerator, and thus to remove their sulphur burden and prevent SO_2 release into the boiler. However, as discussed below, it appeared from the post run inspection that the right hand cyclone drain was completely blocked for the latter part of the run, but no corresponding decrease in sulphur removal efficiency was observed.

During Run 6 the longest uninterrupted period of gasification was 193 hours - a substantial improvement over Run 5. Run 6 process performance is discussed below in conjunction with Run 7.

Process Performance, Runs 6 and 7 -

General Considerations. The Sulphur Removal Efficiencies measured during Runs 6 & 7 agreed reasonably well with those measured during previous runs at lime replacement rates less than 1 mol Ca/mol S but were considerably lower than was anticipated at higher lime replacement rates. The reason for this divergence seems to be that the SO_2 concentration in the flue gas samples prior to Runs 6 and 7 was dependent on the amount of lime dust in the flue gas, which in turn depended on the stone replacement rate. Modification to the flue gas sampling system made prior to Run 6 and changes in the monitoring procedure eliminated this source of error and the running sulphur balances for Runs 6 and 7 (Appendix Tables C.II and D.VI) are well within the calculated margin of experimental error (Appendix J).

Because the S.R.E.s measured in Runs 6 and 7 are considered to be the most reliable, they have been used in order to deduce which are the major factors affecting the desulphurising performance of the gasifier. Comparisons between the results obtained during these two runs must however take account of changes in reactor geometry, process flow plan and bed material which were made between the runs. These changes were undoubtedly important since, as will be shown later, the results obtained during Run 6 were significantly better than those obtained during Run 7, despite the fact that the mean superficial gas velocities during the two sets of test periods were respectively 1.29 m/sec and 1.15 m/sec.

A major geometrical change arose from the installation of heat exchanger tubes in the gasifier bed for Run 7. These tubes were made in the form of portal frames and were retractable. In order to accomodate them the two stage low efflux velocity nozzles of the air distributor which were installed for Run 5 had to be discarded in favour of the smaller diameter single stage nozzles which were used prior to Run 5. It was incidentally the installation of the heat exchanger tubes in the gasifier bed which resulted in the reduced gas velocities observed in Run 7 at fuel throughputs comparable with those for Run 6 and at marginally leaner air/fuel ratios. The reason for this is that the requirement for recycled flue gas for temperature control was reduced in Run 7 by the operation of the heat exchanger. So far as process flow plan is concerned, in Run 6 the fines collected by the regenerator cyclone were discarded, whereas in Run 7 they were reinjected into the gasifier bed. The object of this measure was to improve fines retention.

In both runs the left hand cyclone was drained externally but the right hand cyclone was drained into the gasifier to regenerator transfer fine. Considerable trouble was experienced with the cyclone fines return systems in both runs and in both cases there were extended periods of operation during which fines were not returned to the gasifier bed, without any obvious ill effects in terms of S.R.E.

Stone BCR 1359 was used during all of the lined out test periods reported for Run 6 whereas most of the results reported for Run 7 were obtained using Denbighshire stone. The three sets of results reported after day 12 of Run 7 were obtained using BCR 1359 feed and these are perhaps marginally better than the rest of the results in this test series.

Variables of Major Importance. The test results for Runs 6 and 7 are reported in Tables 14 and 15. All of these results were obtained by averaging sequences of ten hourly sets of data during a more extended period of stable operation. The test periods may be located in Appendix Figures C15 and D25 by the times listed for the first of each set of observations in Tables 14 and 15. The first number in the time sequence relates to the day, and the subsequent four figures give the time on the 24 hour clock.

Gross effects have been detected by plotting individual independent variables against sulphur removal efficiency. Thus in the case of Run 6 it may be seen (Fig.16) that there is a trend for S.R.E. to improve as the gasifier bed is deepened. There is however an ever stronger indication (fig. 17) that gasifier performance improves as the sulphur content of the bed material is reduced. The circled figures against each point in (fig. 17) relate to bed depth and it will be seen that there is no obvious correlation between bed depth and bed sulphur content in this set of results. This indicates that the two effects are independent of each other. The uncircled figures against each point in (fig. 17) relate to stone replacement rate (Ca/S ratio). In this case there is a trend for low stone replacement rates to be associated with high sulphur contents of the bed material, this therefore casts some doubt concerning which of these two variables caused the observed effect on S.R.E. This question can however be resolved by refering to the independent set of results obtained for Run 7. In this case (or as can be seen in figs. 18 and 19), there is a tendency for performance to improve as the bed is deepened and as its sulphur content falls. In this set of results however there is a tendency for deeper beds to have lower sulphur contents so that if these two effects had not been shown to be independent in Run 6 there would have been some doubt as to which was important. The figures shown against the plotted points in (fig. 19) again relate to stone replacement rate. In this case there is no obvious relationship between stone replacement rate and bed sulphur content and consequently S.R.E.

Taking the two sets of results together the indications are that the variables of major importance are bed depth and bed sulphur content. This doesn't mean that the other variables under consideration such as stone replacement rate, bed temperature and air/fuel ratio have negligible effects, but these effects do seem to be of secondary importance.

Table 14. AVERAGED RESULTS FOR SELECTED 10 HR PERIODS

RUN 6

Time of First Reading	S.R.E. %	Bed Depth cm	Ca/S Mol Ratio	Bed Temp °C	Superficial Gas Vel. m/sec	Air/Fuel Ratio % stoic	Stone	Bed Sulphur %wt	Bed Carbon %wt	Regen Selectivity %
2-0830	75.5	55	0.9	874	1.29	20.5	BCR 1359	5.1	0.19	81.9
3-0430	80.0	60	1.1	914	1.42	22.8	"	5.0	0.17	62.3
6-2230	80.0	53	1.4	889	1.34	22.1	"	4.7	0.10	74.3
8-0430	71.5	51	0.4	905	1.29	20.5	"	6.3	0.36	88.3
9-0030	71.5	53	0.6	905	1.16	21.0	"	5.5	0.45	70.7
11-0630	84.0	58	2.2	883	1.26	22.5	"	-	-	63.0
12-2030	82.0	50	1.4	880	1.30	21.6	"	4.5	0.02	55.2
15-1330	82.0	56	1.2	870	1.28	21.3	"	4.3	0.10	58.2
16-1930	71.5	42	1.5	880	1.27	21.0	"	-	-	50.1
19-1730	78.5	43	1.1	874	1.24	20.8	"	-	-	65.6

Table 15. AVERAGED RESULTS FOR SELECTED 10 HR PERIODS

RUN 7

Time of First Reading	S.R.E. %	Bed Depth cm	Ca/S Mol Ratio	Bed Temp °C	Superficial Gas Vel. m/sec	Air/Fuel Ratio % Stoic.	Stone	Bed Sulphur %wt	Bed Carbon %wt	Regen Selectivity %
4-0630	77.5	60	0.80	888	1.11	21.3	Denbighshire 600-3200 μ	3.4	0.53	75.8
5-0230	80.0	61	0.80	918	1.25	23.2	- " -	3.9	0.45	82.2
6-1830	67.5	53	0.50	914	1.20	22.2	- " -	5.5	0.85	74.3
7-1430	67.5	53	0.75	902	1.17	22.1	Denbighshire 300-2000 μ	5.6	0.85	68.4
9-2130	70.0	49	1.40	922	1.10	22.2	Denbighshire 600-3200 μ	5.1	0.08	68.0
11-0930	78.0	56	1.40	909	1.15	23.2	- " -	4.3	0.06	80.6
13-0030	81.0	60	2.40	914	1.12	25.0	BCR 1359	3.3	0.06	73.6
13-1530	80.0	55	1.20	924	1.14	23.3	- " -	3.6	0.14	73.7
14-1630	77.0	54	1.03	878	1.08	22.2	- " -	5.5	0.21	71.5

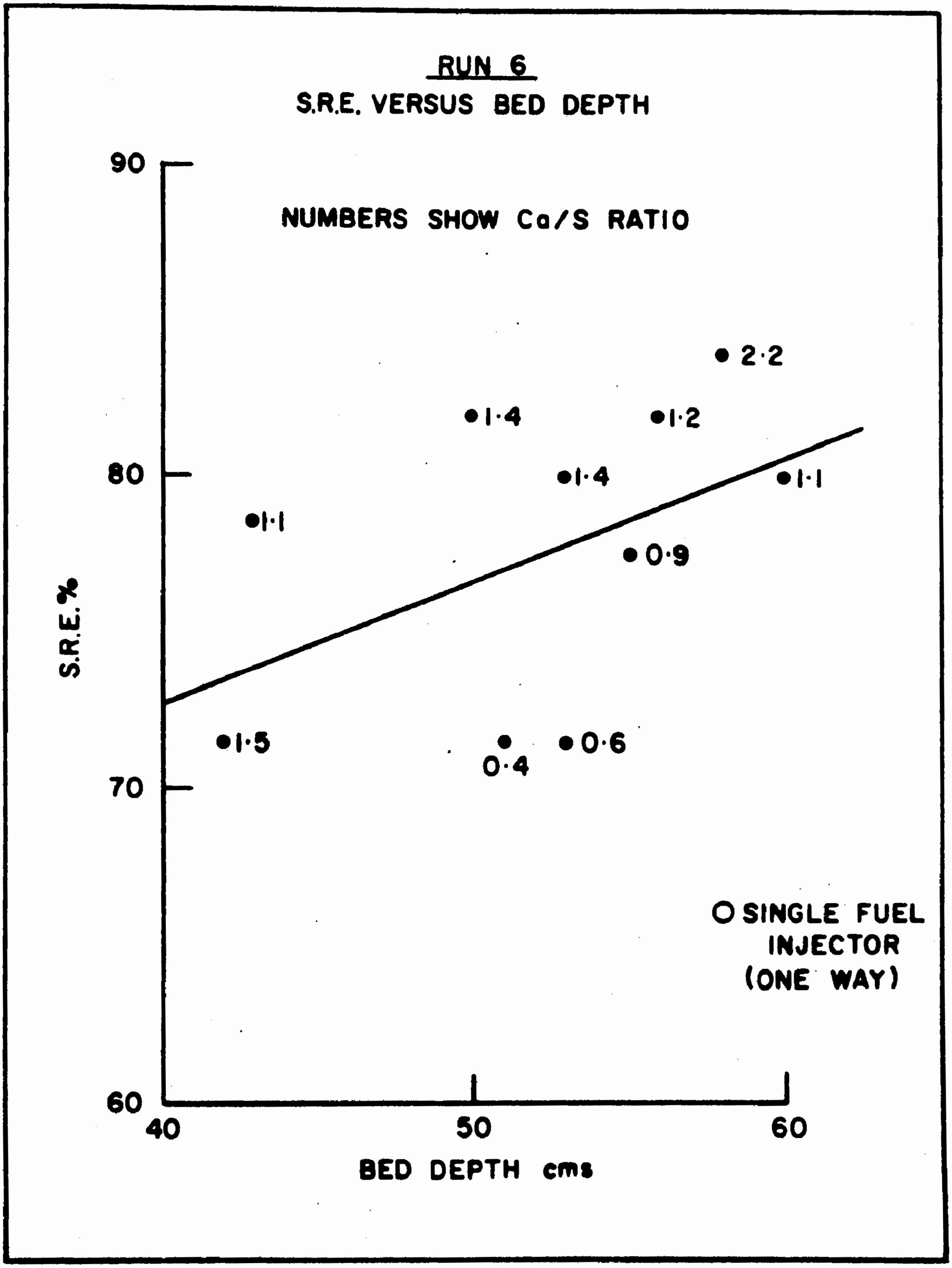

Figure 16 Sulphur Removal Efficiency vs Bed Depth (Run 6)

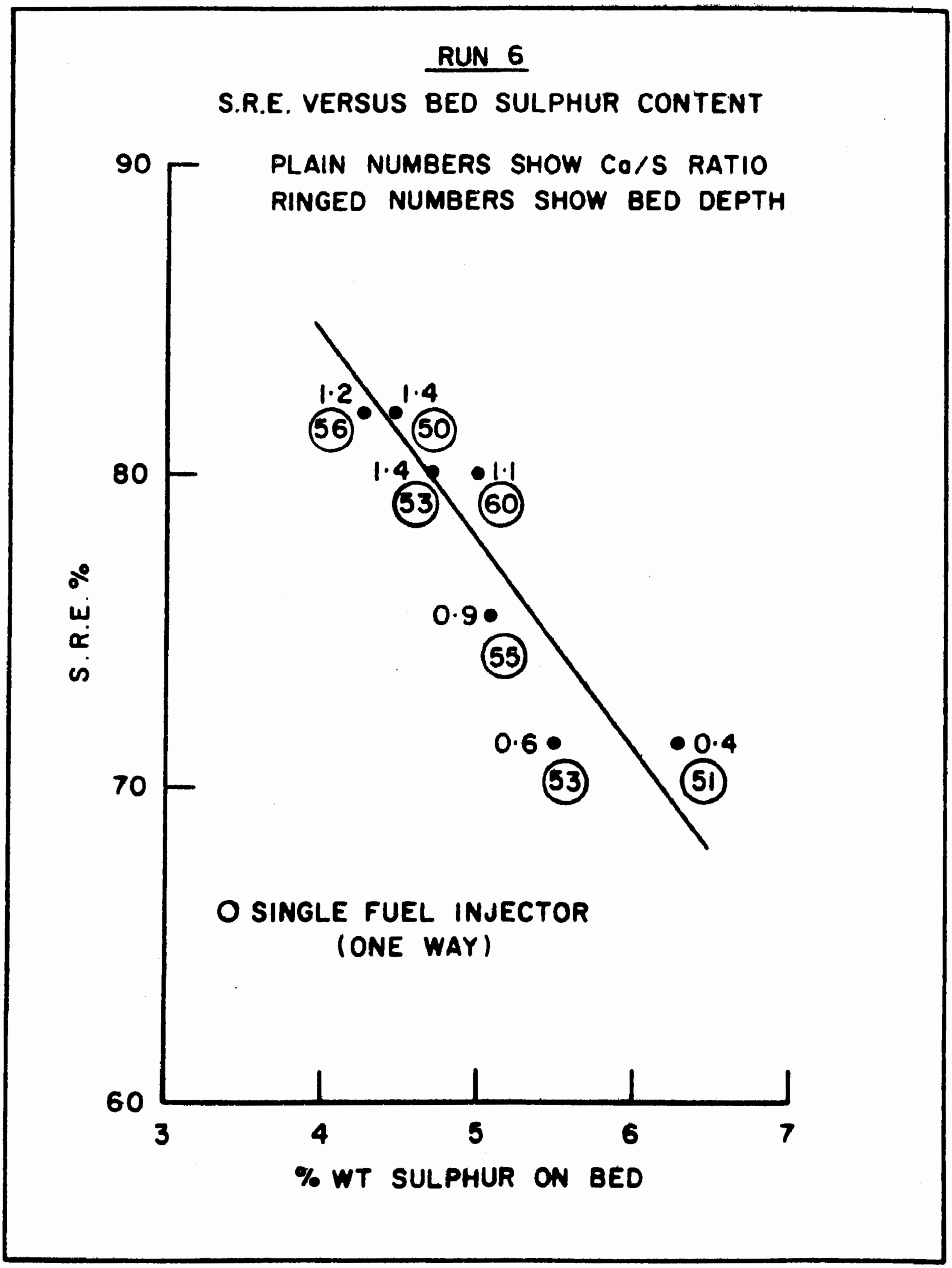

Figure 17 Sulphur Removal Efficiency vs Bed Sulphur (Run 6)

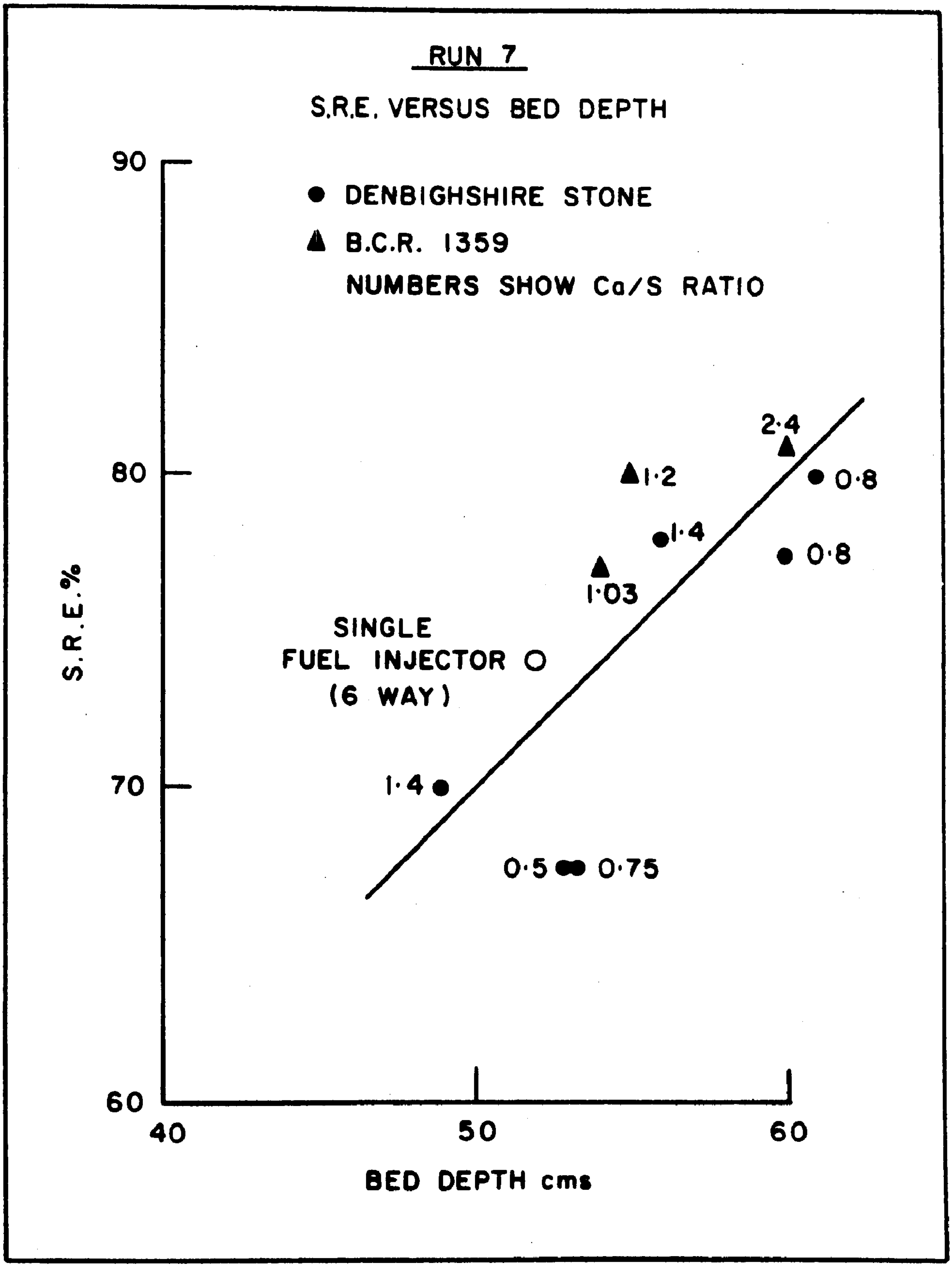

Figure 18 Sulphur Removal Efficiency vs Bed Depth (Run 7)

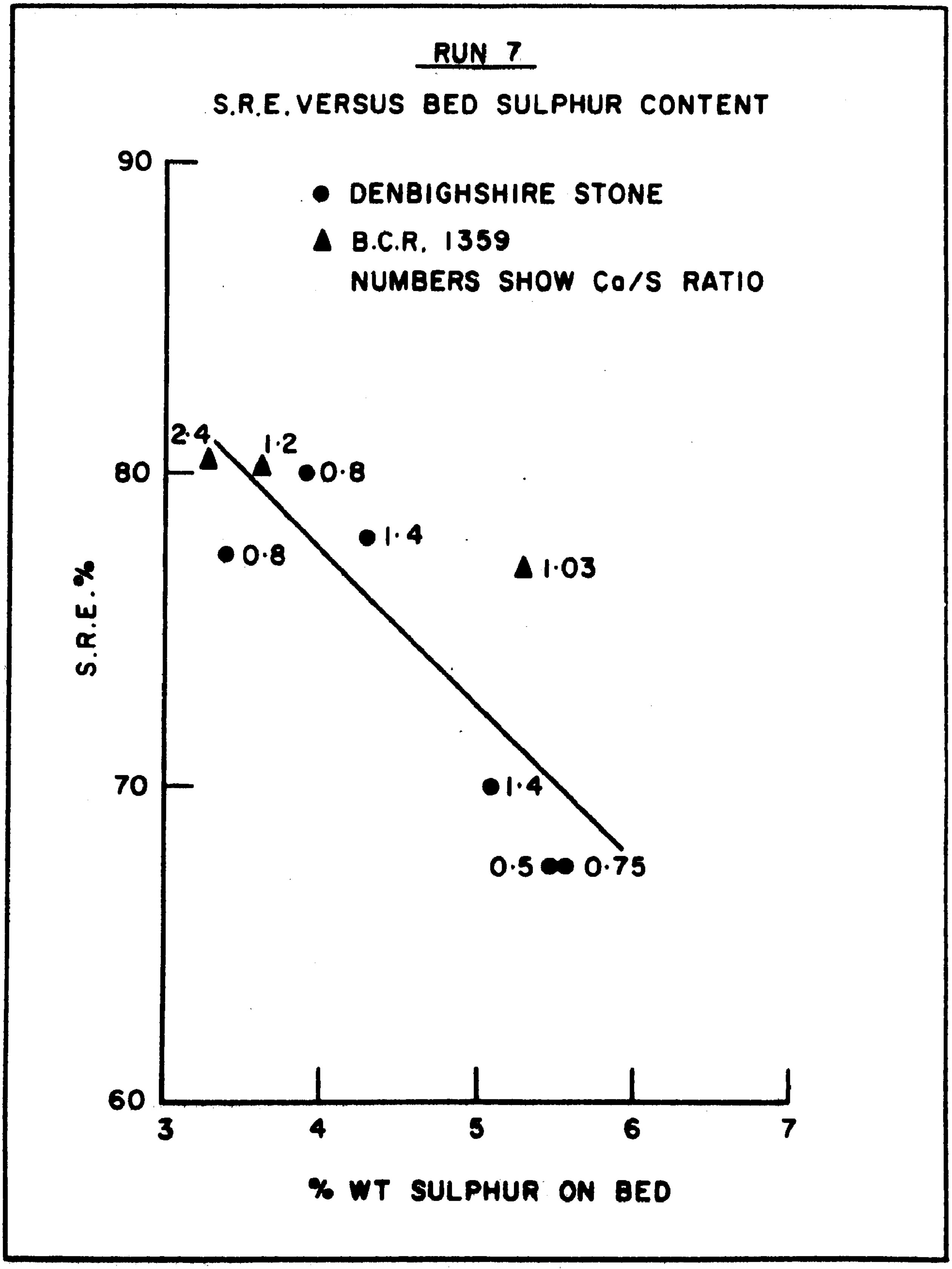

Figure 19 Sulphur Removal Efficiency vs Bed Sulphur (Run 7)

If a gross overall comparison is made between the sets of test results for Runs 6 and 7, it becomes obvious that Run 6 gave a significantly better performance than Run 7. In fig. 20 the two bed depth/S.R.E. relationships are plotted on common coordinates. Although the Run 6 results are very scattered, indicating a major effect for another variable, they are on the whole better than those obtained during Run 7.

The trend lines for the two groups of results tend to converge as the beds became deeper. This probably results from the correlation between bed depth and stone sulphur content in Run 7 which exagerates the effect of bed depth. Even if the trend for Run 6 is more realistic however, the indications are that substantial approvements in S.R.E. should be obtainable with gasifier beds more than 60 cm deep.

In fig.21 the two bed sulphur content/S.R.E. relationships are plotted on common coordinates. In this case there is clear evidence that Run 6 gave a better result for any given sulphur content than Run 7, and there is a strong indication that bed sulphur contents of less than 4% by weight will prove to be advantageous.

The most obvious explanation for the gross difference in performance between Runs 6 and 7 is a difference in the reactivities of the stones which were used. The Run 6 results listed in Table 14 relate entirely to BCR 1359, whereas in Run 7 Denbighshire stone was used up to May 12 and only then was the feed switched to BCR 1359. It will of course take a considerable time for a change in stone feed to have an appreciable effect on bed composition and it was during the period starting 14-1630 in Run 7 that an anomalous result was obtained for a high bed sulphur content which is typical of results obtained during Run 6. This explanation is necessarily very tentative in view of the paucity of the evidence, but it does account for a very atypical result obtained during Run 7.

Variables of Minor Importance. The stone feed rate does appear to have an effect on S.R.E. but rather less of an effect than was anticipated, bearing in mind the results of the batch tests. The magnitude of the effect may be gauged from Figs. 16 and 18. In both these cases the numbers against the plotted points relate to stone replacement rate and in both cases the values below the trend line are lower than those above the trend line. Unfortunately, in both cases the lowest stone replacement rates are associated with the highest stone sulphur contents so that

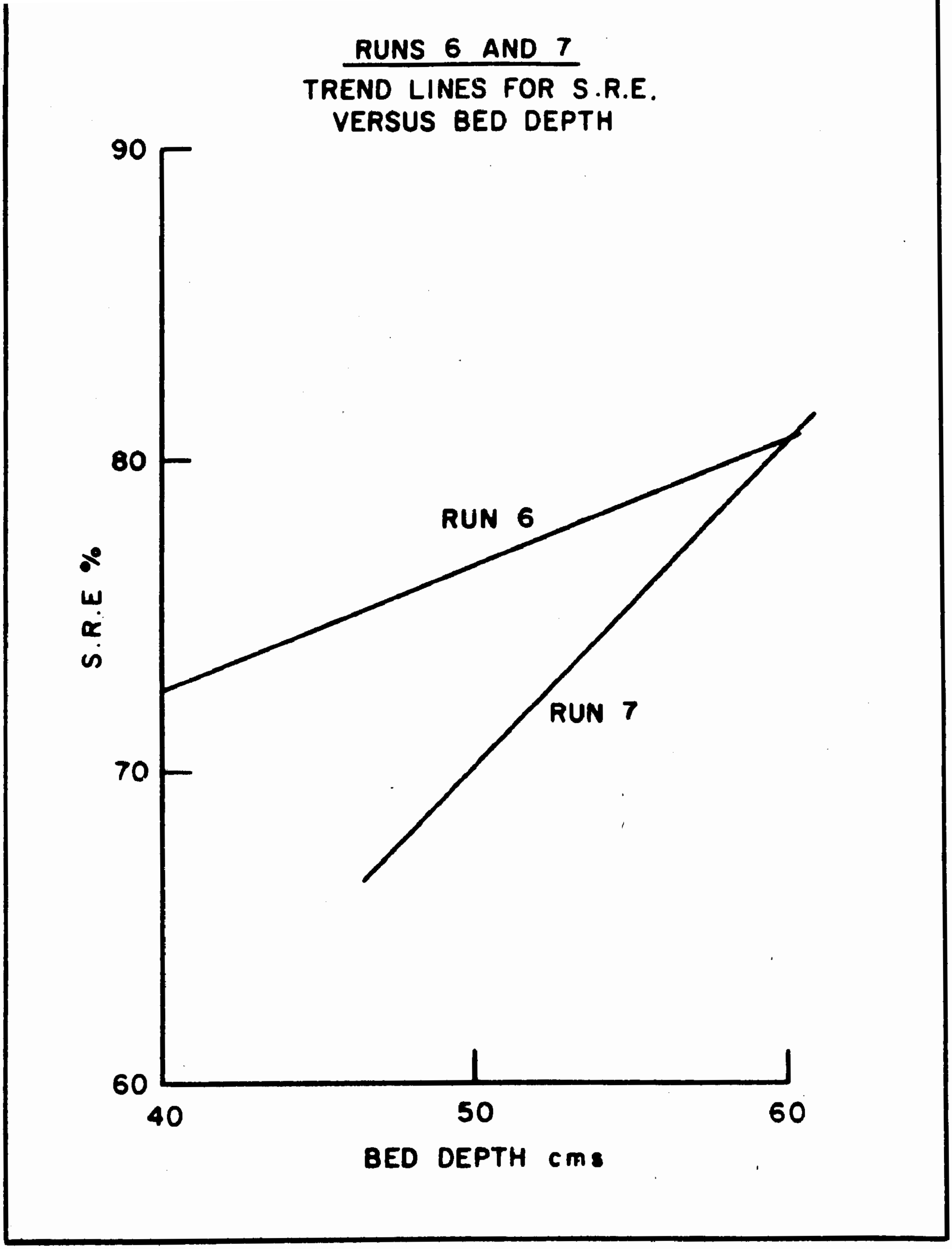

Figure 20 Sulphur Removal Efficiency vs Bed Depth
(Runs 6 and 7)

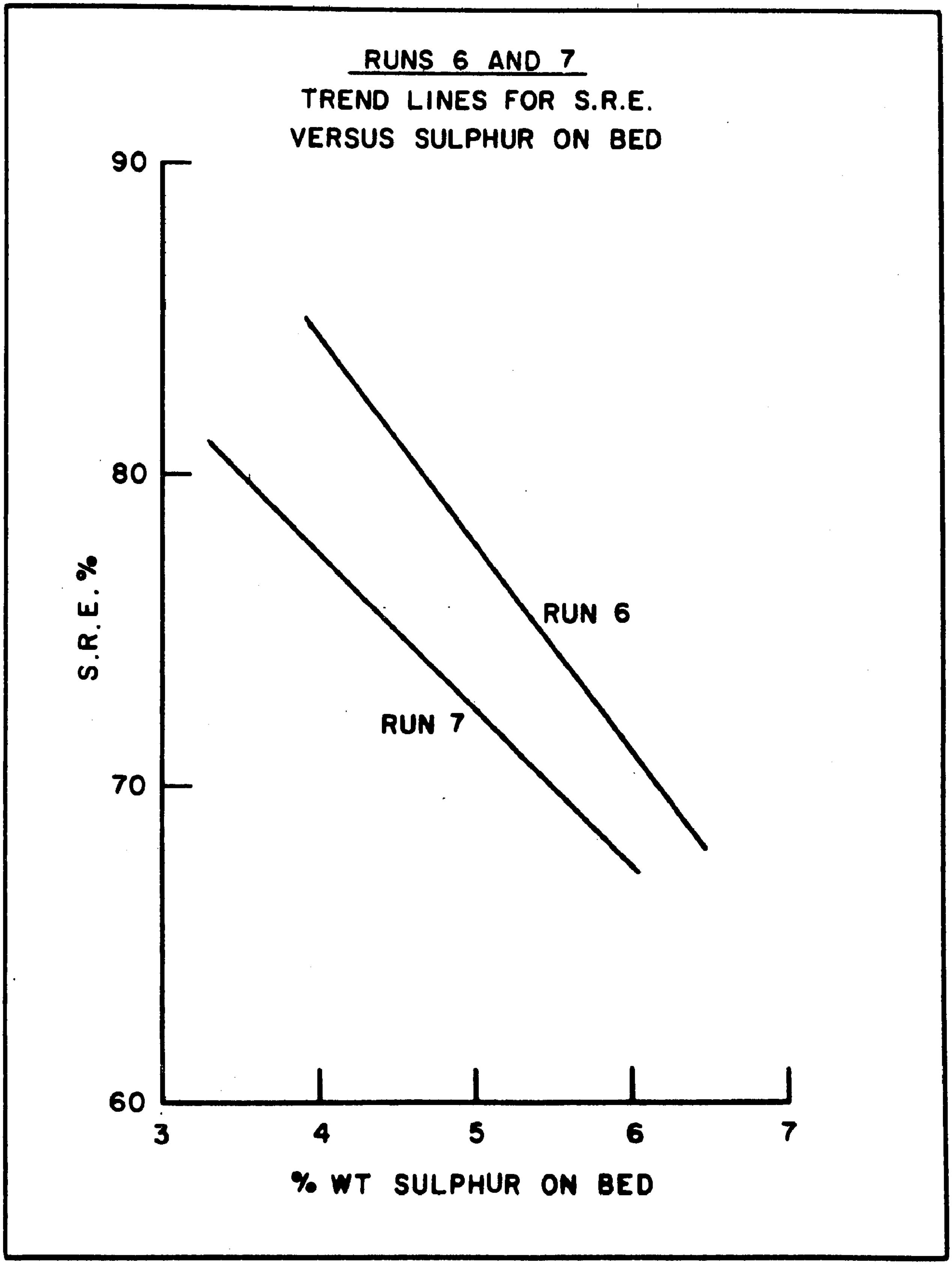

Figure 21 Sulphur Removal Efficiency vs Bed Sulphur
(Runs 6 and 7)

the effect of stone replacement rate as shown in figs 16 and 18 may well be somewhat exaggerated. From a purely practical point of view it is advantageous to minimise stone consumption and the indications are that stone consumptions less than stoichiometric may be anticipated.

The bed temperatures listed in Tables 14 and 15 do not appear to relate strongly with S.R.E. In the case of the Run 7 results there are two pairs of virtually identical bed depths and stone sulphur contents. These occur in periods 4-0630 and 5-0230 and periods 6-1830 and 7-1430. In the first pair of periods the stone feed rate is also constant, the only significant variables being bed temperature and air/fuel ratio, which tend of course to be related. A comparison of the S.R.E. values for these two test periods shows an apparent improvement in performance when the gasifier temperature is increased from 880°C to 918°C. In the case of the second pair however no improvement is seen when the gasifier temperature is raised from 902°C to 914°C but a slightly higher stone feed rate is associated with the lower temperature.

Such evidence as there is from Run 6 tends to be equally contradictory. If on the one hand period 6-2230 is compared with period 12-2030 then raising the bed temperature from 880°C to 889°C appears to have an adverse effect. If on the other hand period 12-2030 is compared with period 15-1330 then the effect of an increase in temperature from 870°C to 880°C seems to have balanced out the deterioration in performance which would otherwise have resulted from a reduction in bed depth from 56 cms to 50 cms. A possible explanation of these apparent contradictions may be that these are two favourable zones of operating temperature, one peaking at about 880°C the other peaking at about 908°C. It seems reasonable however to conclude that the S.R.E. of the gasifier is not unduly sensitive to variations in operating temperature in the range 870 - 920°C.

Effects of variations in fuel injection on S.R.E. Prior to Run 6 fuel oil had been injected into the gasifier through three downward sloping side nozzles. The same fuel injection system was used throughout most of Runs 6 and 7 but in both of these runs the gasifier was provided with an additional retractable fuel injector protruding through the distributor. In Run 6 this retractable injector was used during the period 13-1800 to 15-0300 whilst in Run 7 it was used during the period 16-1715 to the end of the run at 17-2300.

The retractable nozzle used in Run 7 differed in geometry from that used in Run 6. In both cases the fuel was injected horizontally with the assistance of an air blast but whereas in Run 6 all of the fuel entered via a single orifice, in Run 7 the retractable nozzle was provided with six radial holes. The positions of the retractable nozzles also differed. In Run 7 the nozzle was fitted in the centre of the air distributor whereas in Run 6 the nozzle was offset to a position about ¼ of the length of the gasifier bed from its L.H. end. In Run 6 the single orifice was aligned along the axis of symmetry of the gasifier bed so that the direction of fuel injection was from left to right.

The S.R.E. during the Run 6 test period 14-1330, Table 16 which was run whilst the single orifice was being used, was only 66% despite a bed depth of 58 cms and a bed sulphur content of only 3.4%. This may be compared with an S.R.E. of 82% for the test period 15-1330 when the bed was 56 cms deep and its sulphur content was 4.3% and another S.R.E. of 82% for the test period 12-2030 when the bed was only 50 cm deep and its sulphur content was 4.5%. It can be seen in fig. C.15 that there waS an immediate reduction in S.R.E. when the single nozzle was brought into use and that there was a rapid recovery in S.R.E. when the single nozzle was replaced by the three injectors originally used. During this period the divided gasifier plenum was used to induce "gulf streaming", and flow patterns in the bed. No effect was seen.

The effect of using the single injector during Run 7 was much less pronounced than that which was seen during Run 6 though as is shown in Fig. D.25 there does appear to have been a slight drop in S.R.E. after the change to a single injector was made at 16-2040. During the ten hour period commencing at 17-0430 (table 16) the S.R.E. was 74%, the bed depth being 54 cms and the bed sulphur content being 6%. This may be compared with an S.R.E. of 77% for the test period 14-1630 when the bed depth was again 54 cms and the bed sulphur content was 5.5%. The difference in S.R.E. in these two cases might well be accounted for by the slight difference in the bed sulphur contents.

In both Run 6 and Run 7, the fuel was initially injected from the single injector 11 cms above the plane bisecting the nozzles of the air distributor, this being the height of the three side fuel injectors. During Run 6 this height was not changed but during Run 7 an attempt was made to improve

Table 16. RESULTS WITH SINGLE FUEL INJECTORS

RUNS 6 & 7

RUN NO	6	7
Time	14-1330	17-0430
S.R.E.	66.0	74.0
Bed Depth cm	58	54
Ca/S Mol Ratio	1.1	0.9
Bed Temp °C	862	897
Superficial Gas Vel. m/sec.	1.29	1.01
Air/Fuel Ratio % Stoic	20.4	20.1
Bed Sulphur %wt	3.4	-

the performance of the gasifier by lowering the fuel injector 5 cms in two steps of 2.5 cms each. This had little if any effect on S.R.E. but at the lower level the temperature of the gasifier bed tended to fluctuate in an irregular fashion. An attempt was made to raise the injector above its original height but unfortunately it jammed in its sleeve and couldn't be shifted.

In view of the important influence of bed depth on sulphur removal efficiency it is desirable to establish whether bed depth and fuel injection level are interchangeable. If sulphur is mainly lost due to internal reflux within the bed, resulting from the oxidation of sulphide at the distributor, then a decrease in bed depth should have a greater effect than raising the fuel injector an equivalent height. If on the other hand sulphur is lost from the bed surface above the fuel injector point then the effects of varying bed depth and fuel injector height should be equivalent.

The effect of stone size on S.R.E. During Run 7 a specially sized batch of Denbighshire stone was substituted for the normal stone feed for a period of about 24 hours. This batch of stone was sized in the range 300-2000 microns as against the 600-3200 microns normally used and the purpose of the experiment was to determine whether S.R.E. is affected by the size of the bed material.

The results of this experiment may be assessed by comparing the data for test periods 6-1830 and 7-1430 (Table 15). It will be seen that the depth of the bed and the sulphur content of the bed material were equal during the two test periods and that the air/fuel ratio was also unchanged. The stone feed rates were not equal, that for the fine stone feed being 0.75 Ca/S whilst the coarse stone was fed at a Ca/S ratio of only 0.5. In consequence of the differing stone feed rates, the bed temperature when the coarse stone was used was slightly higher than when the fine stone was used. 914°C against 902°C. On the face of things it was to be expected that the combination of a higher stone feed rate and a smaller stone size would improve the S.R.E. In fact however the two results were identical at 67.5%. It is therefore reasonable to conclude that if the size of the stone feed does have an effect, it is a very small one.

Further independent evidence of the relatively minor effect of stone particle size on S.R.E. was obtained during Run 6 when three sets of gasifier and regenerator bed material samples were sieved into six fractions and these fractions were analysed for sulphur content. The results which were obtained are shown in Table 17 and it will be seen that, again contrary to expectations, the coarser stone fractions contained more sulphur than the finer ones although the average differences in sulphur content between the gasifier and regenerator samples were evidently independent of particle size.

Possible explanations for these observations are that the external surfaces of the particles take up a substantial proportion of the sulphur, and that finer particles are deactivated more rapidly than coarser particles. Since it is easier to retain coarse particles than fine particles there seems to be little incentive to use a finer bed material than is necessary to ensure good fluidisation at the optimum superficial gas velocity within the gasifier bed. In view of the importance of this finding however it is considered desirable to obtain confirmatory evidence in future runs.

Run 7

Equipment Performance is fully described in Appendix D Operational Log and Inspection. Overall performance was again improved over Run 6, and the longest period of uninterrupted gasification reached a new peak of 211 hours in the second part of the run. The first part consisting of a single gasification period of 165 hours.

The modified gasifier distributor, with its single multi-port fuel injector, and two water cooled heat transfer tubes. performed fairly well. The test of fuel injector height had to be abandoned when the injector jammed in position, but the effect of one injector on process performance was negligible. The water cooled tubes performed as expected, but because of a slight upward displacement of the front tube the cooling effect on the bed throughout the run was too great to allow the planned series of tests to take place. For considerable periods no flue gas recycle was used for temperature control.

Cyclone performance was again poor, as shown in the post-run inspection, and improvement of cyclone performance was identified as of prime importance in future runs.

The regenerator distributor worked well, and no agglomerated material was found on it in the post-run inspection. Despite the modification to the flue gas recycle scrubber, plugging of the drain occurred. An alarm will be installed for future runs.

Table 17. Sulphur Distribution in Bed Material Size Fractions

Size Range Microns	Bed Sample 1 S % by wt.	Regen Sample 1 S % by wt.	Bed Sample 2 S % by wt.	Regen Sample 2 S % by wt.	Bed Sample 3 S % by wt.	Regen Sample 3 S % by wt.	Average ΔS % by wt.
>1400	5.74	4.70	6.19	3.65	4.65	3.61	1.45
1180-1400	4.76	3.70	4.87	3.95	4.18	3.62	0.85
850-1180	4.51	3.14	4.42	2.94	3.62	2.91	1.19
600-850	4.14	2.91	3.85	2.30	3.69	2.91	1.19
250-600	3.51	2.04	3.14	1.79	2.93	2.46	1.10
<250	3.47	1.77	3.15	1.67	3.95	3.43	1.20

TASK II - BATCH STUDIES

In the original programme of work for this contract, batch units were to be used to determine the suitability of additional fuel - limestone combinations for CAFB applications. It was envisaged that four new limestones would each be studied with two fuels, one of which would be a vacuum pipestill bottoms. Neither Denbighshire nor BCR 1691 stones would be tested during this contract but earlier results on these stones would be used as a basis for comparison. Two of the stones were selected by EPA. These were BCR 1359 limestone and Tymochtee Dolomite. The other two were selected by New England Electric System (NEES) and were Pfizer Calcite and Pfizer Dolomite.

However, the following developments during the contract dictated that the programme be altered. Firstly, Run 4 operations revealed a serious attrition problem with BCR 1691 during start-up which did not appear in earlier batch unit tests. Previously, comparisons of dust losses between stones had been confined to gasification - regeneration cycles. No such comparisons had been made under fully combusting conditions. In the normal batch unit test procedure there had been little exposure of solids to combustion conditions in the absence of sulphur except during calcination. Consequently, the conditions employed during Run 4 start-up produced an entirely unexpected result in that BCR 1691 stone formed copious quantities of a dust with a very sticky nature. Secondly, considerable interest arose in operating CAFB with even heavier feedstocks than vacuum pipestill bottoms. The actual material proposed was a high sulphur petroleum pitch.

In order to accomodate the extra work resulting from these developments without extending the programme, testing of one of the EPA stones (Tymochtee Dolomite) and some of the stone/fuel combinations was dropped. The objectives for the revised batch unit test programme are listed below:

- Determine if continuous unit conditions which produced large quantities of sticky dust could be duplicated in batch units.

- Compare dust producing tendencies of Denbighshire and BCR 1691 stones under a variety of conditions.

- Provide a quantitative measurement of dust production to be expected under start up and operating conditions with Denbighshire, BCR 1691 BCR 1359, Pfizer Calcite and Pfizer Dolomite.

- Measure sulphur absorption performance of BCR 1359, Pfizer Calcite and Pfizer Dolomite.

- Conduct tests of the feasibility of operating CAFB with vacuum pipestill bottoms and High Sulphur Pitch.

The actual test programme which was carried out (Table 18) did not, in fact, include any tests on Pfizer Dolomite as Westinghouse had advised us that this stone decrepitated so badly on calcination that it was not worth testing. Additional tests on the heavy feedstocks were carried out instead.

The investigation into fines production rates and properties was based very much on our experience in Run 4. There, fines produced from BCR 1691 under fully combusting conditions did not drain freely from cyclones, whereas those produced under gasifying conditions did.

It was suspected that the higher resistance to flow of the combustion fines was due to their containing a much higher proportion of very fine particles. Why less of the very fine particles should be produced during continuous gasification and regeneration was not clear. It was considered possible that the presence of sulphur on the stone, the higher regenerator bed temperature, an ageing effect or a combination of these could be the answer. Any of these changes could have altered the particle surfaces in such a way as to make them less susceptible to decrepitation.

In the tests on the heavier fuels, with their higher Conradson Carbon contents, particular attention was paid to the increased rate of carbon deposited on bed particles. Fresh bed tests were carried out to determine the effects of air/fuel ratio, oxygen enrichment and bed temperature on this. Test 5-D was included for comparison purposes.

<u>Stone Comparison Tests</u> -

Four stones, Denbighshire, BCR 1691, BCR 1359 limestones and Pfizer Calcite (Appendix O) were compared in terms of sulphur removal efficiencies (SRE), fines production rates and fines

Table 18. Test Programme carried out in CAFB Batch Units

Test	Stone	Fuel	Conditions	Objective
1 - A	BCR 1691	Kerosene	Calcination and prolonged combustion at 870 deg. C	Measure fines loss rate and properties of dust produced with BCR 1691.
1 - B	"	"	Same at 1050 deg. C	Same
1 - C	"	Amuay 2.5% S fuel oil	Calcination and combustion at 870 deg. C	Same
1 - D	"	"	Calcination and gasification at 870 deg. C	Same
1 - E	"	"	Gasification - regeneration cycles at 0.9 moles CaO/mole S.	Measure lined out sulphur removal efficiency and fines loss rate
1 - F	"	Kerosene	Prolonged combustion using cycled bed from Test 1-E.	Measure fines loss rate during kerosene combustion in conditioned lime bed.
2 - A through 2 - F	Denbighshire	Kerosene and Amuay 2.5% S fuel oil	Entire programme same as in Test 1 through step F.	Measure fines loss rate under different sets of conditions with Denbighshire stone.
3 - A	BCR 1359	Kerosene	Calcination and prolonged combustion at 870 deg. C	Measure fines loss rate and properties of dust produced with BCR 1359.
3 - B	"	Amuay 2.5% S fuel oil	Calcination and prolonged gasification at 870 deg. C	Same
3 - C	"	"	Gasification - regeneration cycles at 0.9 moles CaO/mole S.	Measure lined out sulphur removal efficiency and fines loss rate
3 - D	"	"	Gasification - regeneration cycles at 1.2 moles CaO/mole S.	Investigate effect of make-up with BCR 1359.
3 - E	"	"	Gasification - regeneration cycles at 1.5 moles CaO/mole S	Same
4 - A	Pfizer Calcite	Kerosene	Calcination and prologned combustion at 870 deg. C	Measure fines loss rate and properties of dust produced with Pfizer Calcite.
4 - B	" "	Amuay 2.5% S fuel oil	Same	Same
4 - C	" "	"	Gasification - regeneration cycles at 0.9 moles CaO/mole S.	Measure lined out sulphur removal efficiency and fines loss rate.
5 - A	BCR 1359	Amuay 3% S Vacuum Resid.	Prolonged gasification at 870 deg. C	Measure sulphur removal efficiency and carbon deposition.
5 - B	"	"	Prolonged gasification with 25% excess oxygen at 900 deg. C.	Same
5 - C	"	"	Prolonged gasification with 25% excess oxygen at 950 deg. C.	Same
5 - D	"	Amuay 2.5% S Fuel oil	Prolonged gasification at 870 deg. C.	Same
5 - E	"	Amuay 3% S Vacuum Resid.	Gasification - regeneration cycles at 0.9 moles CaO/mole S.	Measure lined out sulphur removal efficiency and carbon deposition.
6 - A	"	High Sulphur Pitch	Prolonged gasification at 25% stoichiometric air	Measure sulphur removal efficiency and carbon deposition.
6 - B	"	"	Prolonged gasification at 30% stoichiometric air	Same
6 - C	"	"	Gasification - regeneration cycles at 0.9 moles CaO/mole S.	Measure lined out sulphur removal efficiency and carbon deposition.

properties. Since fines production tests were carried out under a variety of CAFB operating conditions, information on the effects of operating conditions on fines production was also obtained.

Sulphur Removal Efficiencies for the four stones were compared by cycle tests. It was originally intended that all stones would be tested at the same target conditions listed below which include a bed replacement rate less than 1 mole CaO/mole S.

Air/Fuel Ratio	(% of stoichiometric)	25
Gasification Temperature	(deg.C)	870
Bed Replacement Rate	(mole CaO/mole S)	0.9
Bed Depth	(cm w.g.)	38
Gas Velocity	(m/sec)	1.83
Potential Sulphur Differential	(% wt)	2
Limestone Particle Size	(microns)	600-3175

These were chosen to give SREs significantly less than 100%.

When successive cycles of gasification and regeneration are carried out under such conditions, SRE falls for several cycles and then lines out. Comparison values of SRE are measured at the lined out level. However, with Pfizer Calcite, bed losses were so high that this stone could not be tested at bed replacement rates less than 1.5 moles CaO/mole S.

In addition to direct comparison of stones at a single set of conditions, the effect of bed replacement rate on the performance of BCR 1359 was also examined and compared with previous data on BCR 1691. Detailed results of the tests are listed in Appendix N. Results are summarised in Table 19.

Measured SREs were similar for BCR 1691, Denbighshire and BCR 1359 stones at the lower bed replacement rate and for BCR 1359 and Pfizer Calcite at the higher rate. However, actual conditions did vary slightly from target conditions as shown in the table and, therefore, for comparison, lined out SREs were calculated for each of the test conditions for each stone from the equation derived for BCR 1691 (Reference 1). From the ratio of measured to calculated SRE, it appears that the Denbighshire stone is slightly more active than the other three which have similar activities.

Table 19

Summary of Stone Comparison SRE Results

Limestone	Residence Time (sec)	P.S.D.≠ (wt %)	Make-up Rate CaO/S wt.	Make-up Rate CaO/S Mole	SRE (Measured) %	SRE * (Calculated) %	Ratio of SRE (Meas'd) to SRE (cal)
BCR 1691	0.20	1.96	1.59	.91	75	78	0.96
Denbighshire	0.17	2.18	1.49	.85	76	65	1.17
BCR 1359	0.15	1.87	1.61	.92	76	76	1.0
BCR 1359	0.18	2.00	2.05	1.17	79	83	0.95
BCR 1359	0.19	1.76	2.83	1.62	89	95	0.94
Pfizer Calcite	0.16	1.81	2.71	1.57	89	93	0.96

* Calculated from equation for predicting SRE's for BCR 1691.
≠ Projected Sulphur Differential

The tests on bed replacement rate indicated that this variable has a significant effect on SRE of BCR 1359 although comparison with calculated SREs for BCR 1691 show that the effect may be less marked than with BCR 1691. These conclusions must be somewhat tentative, however, as other variables also changed slightly and the magnitudes of the effect of each variable may very well be different with BCR 1359.

Fines Production Rate measurements were based on bed losses. Table 20 summarises the bed loss results for the four limestones under a variety of CAFB conditions. Detailed results are given in Appendix N.

With BCR 1691, the highest bed loss rate occurred during kerosene combustion with a fresh bed at 870°C. Loss rates during fuel oil combustion and gasification where sulphur was being absorbed by the stone were lower, those during gasification where sulphur absorption was more rapid being less than during combustion.

The sulphided and aged stone from the cycle tests also showed a relatively low loss rate during kerosene combustion at 870 deg. C. The lowest loss rate measured by tests of the fresh bed type, however, was obtained during kerosene combustion at 1050 deg. C, the temperature level used in regeneration. This was only slightly above the lined out rate for the cycle tests in which the rate fell from 19.8 g/min in the first cycle to a stable level of 4.5 g/min after 5 cycles. We can conclude, therefore that raising bed temperature or introducing sulphur into the bed decreases BCR 1691 loss rate. This explains the reduction in losses observed when gasification was commenced in Run 4. Sulphur had been absorbed by the bed which was also being subjected to the high temperatures of regeneration. Prior to this, no sulphur had been introduced to the bed since kerosene combustion had been employed, and the regenerator temperature was low since no reaction was taking place.

A bed ageing effect is also indicated. Hourly losses during fresh bed tests decreased as the tests proceeded whilst loss rate during cycle tests had fallen from 19.8 g/min to 4.5 g/min by 5 cycles. It is important to note, however, that ageing during fresh bed tests is distorted to some extent due to the fact that bed depth was decreasing during each test. That bed depth has an important effect on loss rate was established in the Phase 1 work (Reference 1).

Table 20

Summary of Batch Unit Fines Loss

Test	Test Condition	Limestone	Temperature deg. C	Total Solids Loss from Bed, grams 1 hour	2 hours	3 hours	4 hours	Average Loss Rate over 4 hours (g/ min)
1-A	Kerosene Combustion	BCR 1691	870	2900	4220	5000	5430	22.6
1-B	Kerosene Combustion	"	1050	555	960	1220	1360	5.7
1-C	Fuel Oil Combustion	"	870	1110	1910	2690	3250	13.5
1-D	Fuel Oil Gasification	"	870	680	1190	1560	1860	7.8
1-E	Gasification-Regeneration Cycles	"	850 - 1050	-	-	-	-	19.8/4.5 *
1-F	Kerosene Combustion Sulphided Aged Bed	"	870	760	1300	1700	2040	8.5
2-A	Kerosene Combustion	Denbighshire	870	500	760	930	1020	4.3
2-B	Kerosene Combustion	"	1050	520	680	750	790	3.3
2-C	Fuel Oil Combustion	"	870	1160	1600	1920	2160	9.0
2-D	Fuel Oil Gasification	"	870	1460	1930	2160	2330	9.7
2-E	Gasification-Regeneration Cycles	"	850 - 1050	-	-	-	-	6.6/1.4 *
2-F	Kerosene Combustion Sulphided Aged Bed	"	870	48	96	140	188	0.8
3-A	Kerosene Combustion	BCR 1359	870	200	340	440	510	2.1
3-B	Fuel Oil Gasification	"	870	480	630	750	810	3.6
3-C	Gasification-Regeneration Cycles 0.9 moles CaO/Mole S	"	850 - 1050	-	-	-	-	6.4/1.9 *
3-D	Gasification-Regeneration Cycles 1.2 moles CaO/Mole S	"	850 - 1050	-	-	-	-	4.2/-*
3-E	Gasification-Regeneration Cycles 1.5 moles CaO/mole S	"	850 - 1050	-	-	-	-	-
4-A	Kerosene Combustion	Pfizer Calcite	870	600	1060	1380	1640	6.8
4-B	Fuel Oil Combustion	"	870	850	1320	1620	1800	7.5
4-C	Gasification-Regeneration Cycles 0.9 moles CaO/mole S	"	850 - 1050	-	-	-	-	10.0/6.3*

* First value is loss rate during 1st cycle, second is lined out loss rate after 5 cycles.

The relative importance of all variables effecting bed loss rate of BCR 1691 is summarised in the equation below. This was derived from further analysis of the fresh bed test results.

$$L = \frac{29.54 \times D^{2.17}}{A^{0.44} \times (T-750)^{1.80} \times S^{0.41}}$$

L = Loss rate (g/min)
D = Bed Depth (cm)
A = Bed Age (hours)
T = Bed Temperature (deg. C) (750 deg. C is taken as $CaCO_3$ decomposition temperature).
S = Bed Sulphur Content including inherent sulphur (% by weight)

This equation shows the approximately square relationship between losses and bed depth as observed previously. The loss rate of 3.1 g/min calculated from the above equation for cycle test conditions is in fair agreement with the measured rate of 4.5 g/min.

With Denbighshire stone fresh bed tests showed a decrease in loss rate with increased temperature and an increase when fuel was used instead of kerosene. The stable rate during cycle tests was lower than for fresh bed tests. The lowest rate, however was measured with aged, sulphided stone under kerosene combustion.

In light of our experience with the two previous stones, only kerosene combustion, fuel oil gasification and cycle tests were studied with BCR 1359. The gasification fresh bed test gave a higher loss rate and cycle tests a lower rate than the fresh bed test with sulphur free kerosene combustion.

With Pfizer Calcite only kerosene combustion, fuel oil combustion and cycle tests were examined. Fuel oil combustion gave a higher loss rate and cycle tests only a slightly lower rate than the fresh bed test with kerosene combustion.

The bed ageing effect which decreased losses with BCR 1691 was also evident with Denbighshire, BCR 1359 and to a lesser degree with Pfizer Calcite. For example, within 5 cycles, loss rate dropped from 6.6 g/min to 1.4 g/min with

Denbighshire, from 6.4 g/min to 1.9 g/min with BCR 1359 and from 10.0 g/min to 6.3 g/min with Pfizer Calcite. These results reveal several interesting comparisons between the stones. These are summarised in Table 21.

None of the higher purity stones gave the very high loss rate found with BCR 1691 during fluid bed combustion with kerosene at 870 deg. C. Both Denbighshire stone and BCR 1691 gave lower loss rates during kerosene combustion when temperature was increased from 870 deg. C to 1050 deg. C. Although the temperature effect was less dramatic for Denbighshire, its loss rate remained below that of BCR 1691. No high temperature tests were made with BCR 1359 or Pfizer Calcite.

All four stones exhibited a decreasing loss rate with age during the initial test periods. The change became less significant at long exposure times. This age effect may be due to a strengthening of particles by a sintering process, to elimination of particles of lower initial strength, or to a combination of these factors. The significant decrease in loss rate at the higher temperature appears to be a consequence of the more severe sintering which would be expected under those conditions.

The effect of changing from kerosene to fuel oil differed between the low purity and high purity stones. Whereas with BCR 1691, fresh bed loss rate decreased when fuel oil replaced kerosene combustion and decreased again on going to fuel oil gasification, opposite directional results were found with Denbighshire stone. Results from the shorter test programme on BCR 1359 and Pfizer Calcite indicate that their behaviour is similar to that of Denbighshire. In spite of its high loss rate during fresh bed gasification the Denbighshire stone gave the lowest rate of the three during gasification - regeneration cycles.

We believe that loss rate differences between kerosene and fuel oil operation are due to sulphur in the oil. However, the mechanism of the sulphur effect must be complex to increase losses with the pure stones whilst decreasing losses with the lower purity BCR 1691.

Although Pfizer Calcite behaved in a directionally similar manner to changes in operating conditions as the other pure stones the magnitude of its responses was considerably less. We consider that the lower level of response results from

Table 21

Summary of Batch Unit Loss Rates

Loss Rate (g/min) *

Conditions:	Kerosene** Combustion 870 deg.C	Kerosene** Combustion 1070 deg.C	Fuel Oil** Combustion 870 deg.C	Fuel Oil** Gasification 870 deg.C	Kerosene Combustion on Sulphided Aged Bed 870 deg. C	Gasification Regeneration Cycles
Limestone						
BCR 1691	22.6	5.7	13.5	7.8	8.5	4.5
Denbighshire	4.3	3.3	9.0	9.7	0.8	1.4
BCR 1359	2.1	-	-	3.6	-	1.9
Pfizer Calcite	6.8	-	7.5	-	-	6.3

* At all conditions except gasification/regeneration cycles, loss rate has been calculated over the first four hours of the test. The cycle loss rate is the stable loss rate.

** Fresh Bed Tests

the major attrition mechanism being different with this stone. As a result of its much larger crystallites, attrition results principally from the fracture of these whereas with the other stones it probably results mainly from the separation of crystallites. In line with this the much smaller reduction in loss rate with Pfizer Calcite during gasification/regeneration cycles could very well be due to the inability of the large crystallites to withstand the thermal shock associated with changes between gasification and regeneration conditions. Finally, on the basis of these tests, we have concluded that Denbighshire and BCR 1359 stones are suitable for the CAFB process, that BCR 1691 is unsuitable principally because of its high fines loss rate under combustion conditions which would be used for hot standby in commercial applications, and that Pfizer Calcite is unsuitable because of its high fines loss rate during gasification/regeneration cycles. We postulate that the best stones for the process have high purity but not large crystallites.

In addition to the results already discussed, bed losses were also measured during calcination and are summarised in Table 22.

Table 22

Summary of Batch Unit Calcination Losses

Stone	Low Sulphur Fuel Kerosene and propane *	High Sulphur Fuel 2.3% S Oil *
BCR 1691	18	15
Denbighshire	16	24
BCR 1359	6	-
Pfizer Calcite	18	21

* Losses as % of calcined stone

The low calcination loss rate from stone BCR 1359 makes it particularly attractive. The effects of sulphur on loss rate in calcination of Denbighshire, Pfizer Calcite and BCR 1691 stones are in the same direction as observed in the fresh bed tests with these stones.

Properties of Fines were deduced from material collected in the cyclone and deposited in pipes downstream of the cyclone.

Throughout the tests the cyclone operated satisfactorily with the aid of a mechanical rapper, and regular samples of fines were obtained. Table 23 contains the results of a microscopic examination of these fines together with cyclone efficiencies. With respect to physical appearance, the fines have been separated into five groups. Examples from each group are shown in Figures 22 to 26. The only fines represented by Figure 22 are those collected during kerosene combustion at 870 deg. C in BCR 1691. Most of the particles appear to be less than 50 μ. The fine particles seem to be adhering to each other to form loose agglomerates and to the surface of the few larger particles that are around. This stickiness was also observed during kerosene combustion in Run 4. Figure 23 shows the type of particles obtained when BCR 1691 was subjected to kerosene combustion at 1050 deg. C, to fuel oil combustion at 870 deg. C, and to gasification - regeneration cycles followed by kerosene combustion at 870 deg. C. There appears to be some larger particles in this group and there is less evidence of the stickiness. In Figure 24, the type of fines collected in all instances where gasification or gasification/regeneration cycles were carried out are illustrated. Due to the presence of carbon, it is difficult to estimate the particle size range present. However, it would appear that the great majority of particles are less than 50 μ. The fines collected under combustion conditions with Denbighshire, Pfizer Calcite and BCR 1359 stones are represented in Figure 25.

In this case there is a discrete mixture of particles with a maximum size around 1000 μ. Finally, Figure 26 illustrates the type of particles collected during kerosene combustion at 870 deg. C of the sulphided, aged, Denbighshire bed. Here, there is a discrete mixture of particles which are mainly in the size range 50-1000 μ.

These results indicate an approximate correlation between fines product rate and particle size of the fines collected by the cyclone. The large cyclone particles were collected at the lowest production rate i.e. during kerosene combustion of sulphided, aged, Denbighshire stone. Also it would appear that kerosene combustion at 870 deg. C in BCR 1691 which gave the highest fines production rate resulted in the smallest particles being collected in the cyclone.

Table 23

Nature of Cyclone Fines from Batch Unit Studies

Test Condition	Limestone	Appearance	Size	Cyclone Efficiency ,
Kero combustion 870 deg.C	BCR 1691	Loose agglomeration of particles less than 50 µ	99% <50 µ	80
Kero combustion 1050 deg.C	"	Mainly discrete particles less than 50 µ	90% <50 µ	100
Fuel oil combustion 870 deg.C	"	Mainly discrete particles less than 50 µ	90% < 50 µ	32.6
Fuel oil gasification 870 deg. C	"	Mixture of carbon & Lime	-	49.1
Gasification/Regeneration Cycles	"	Mixture of carbon & Lime	-	83.9
Kero combustion-sulphided aged bed 870 deg. C	"	Mainly discrete particles less than 50 µ	90% <50 µ	59.5
Kero combustion 870 deg.C	Denbighshire	Discrete mixture of particles up to 1000 µ	-1000 µ	79.4
Kero combustion 1050 deg.C	"	"	-1000 µ	62.0
Fuel oil combustion 870 deg. C	"	"	-1000 µ	49.8
Fuel oil gasification 870 deg. C	"	Mixture of carbon & Lime	-	56.6
Gasification/Regeneration Cycles	"	Mixture of carbon & Lime	-	100
Kero combustion - sulphided aged bed 870 deg. C	"	Mainly particles greater than 50 µ	90% > 50 µ	44.4
Kero combustion 870 deg. C	BCR 1359	Discrete mixture of particles up to 1000 µ	-1000 µ	86.3
Fuel oil gasification 870 deg. C	"	Mixture of carbon & Lime	-	83.0
Gasification/Regeneration Cycles	"	Mixture of carbon & Lime	-	100
Kero Combustion 870 deg. C	Pfizer Calcite	Discrete mixture of particles up to 1000µ	-1000µ	96.6
Fuel oil combustion 870 deg. C	"	" " "	-1000µ	62.0
Gasification/Regeneration	"	Mixture of carbon & lime	-	-

Figure 22 Cyclone Fines - BCR 1691 (Kerosene Combustion 870 deg. C.)

Figure 23 Cyclone Fines - BCR 1691 (Kerosene Combustion 1050 deg. C. Fuel Oil Combustion 870 deg. C., Kerosene Combustion 870 deg. C on aged bed).

Figure 24 Cyclone Fines - BCR 1691, BCR 1359 Pfizer Calcite and Denbighshire (Gasification - Regeneration Cycles)

Figure 25 Cyclone Fines - BCR 1359 Pfizer Calcite and Denbighshire (All Combustion Conditions)

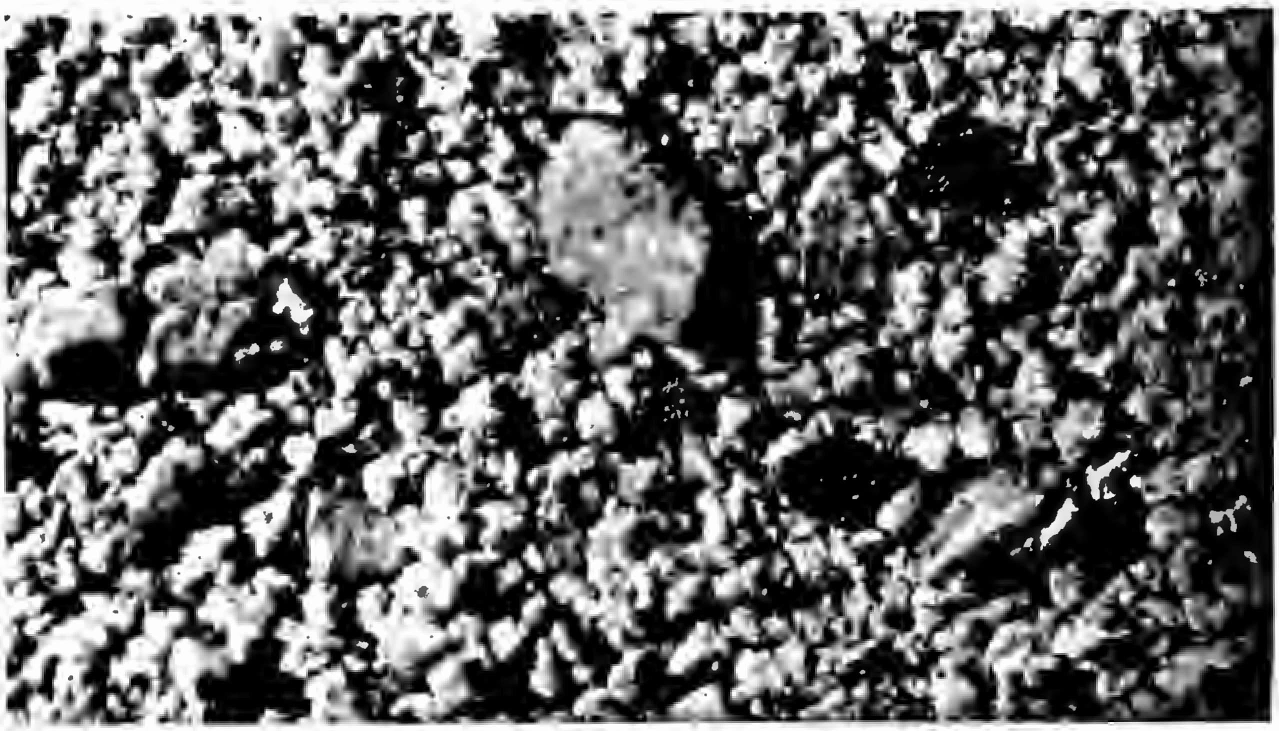

Figure 26 Cyclone Fines - Denbighshire (Kerosene Combustion 870 deg. C. on conditioned bed)

It is not possible, however, to extend this argument and correlate fines production rate with particle size of total fines produced because of differences in cyclone efficiency. One of the lowest cyclone efficiencies was recorded for the large cyclone particle sizes and one of the highest for the smallest cyclone particle size. If particle size of cyclone material was a true reflection of particle size of fines produced by the bed, then one would have expected cyclone efficiencies to have been highest with the largest particle size. Why cyclone efficiency should act in this way is indeed puzzling. However, it does mean that this information of fines size and appearance cannot be used with respect to fines production mechanisms.

Further tests on the fines collected showed that those collected during gasification flowed best and equally well for all four stones. Under all other conditions those from BCR 1359, Pfizer Calcite and Denbighshire flowed more freely than those from BCR 1691. The superior flow characteristics of fines from gasification are attributed to their high carbon content which could be as much as 45% by weight. The superior flow characteristics of fines from Denbighshire, Pfizer Calcite and BCR 1359 stones in relation to BCR 1691 under all the other CAFB conditions are attributed to the presence of fewer of the very fine particles. Although these differences were encountered in the flow properties of the fines, we never encountered batch unit fines with the severe stickiness of pilot plant fines made during Run 4 startup with BCR 1691.

Batch unit fines and bed samples from BCR 1691 were analysed for calcium and silicon. The results showed that the batch unit conditions also caused the preferential loss of calcium from the bed observed in pilot plant Run 4. It appeared to begin during calcination and continue through the tests. As in the pilot plant, calcium lost from the bed did not appear in recovered fines but was lost from the system. The extent of calcium loss in the batch unit tests was not as great as found in the pilot plant where the SiO_2/CaO ratio increased to 0.41 compared with an initial value of 0.27. In batch unit cycle tests, the SiO_2/CaO ratio lined out at approximately 0.33.

During each test unit pressures were monitored to determine if any blockages were occurring downstream of the bed and after each test exit gas lines were dismantled and examined. Only one blockage of any significance was ever encountered. This occurred with BCR 1691 under kerosene combustion conditions at 870 deg. C and was traced to a period of cyclone rapper malfunction. This emphasises the importance of fines concentration in the gas stream in relation to blockages and shows that even with BCR 1691's high rate of sticky fines production, proper draining of cyclones with the aid of rappers where necessary prevents blockages.

Heavy Fuel Tests -

The object of these tests was to investigate the feasibility of operating the process with heavy fuels such as Amuay Vacuum Pipestill Bottoms and High Sulphur Pitch. If the CAFB process could be made to operate satisfactorily with these types of fuel it would provide one of the few viable means of utilising these and similar products of fuel and desulphurisation processes.

It was recognised that the heavy fuels with their higher Conradson Carbon levels (Table 24) would very likely deposit carbon on the bed at a higher rate as Conradson Carbon relates well with carbon deposition during thermal cracking. Tests were largely designed to study means of controlling this within acceptable limits, air/fuel ratio, oxygen enrichment and gasification temperature being examined. In addition, a set of cycle tests were carried out with each fuel.

Table 24

Comparison of Fuel Conradson Carbon Levels

Fuel	Conradson Carbon (%wt)
Amuay Resid.	11.6
Amuay Vacuum Pipestill Bottoms	17.4
High Sulphur Pitch	33.0

Amuay Vacuum Pipestill Bottoms used in fresh bed tests gave the results shown in Figures 27 and 28. Operating conditions for these tests are summarised in Table 25. Full details are given in Appendix N.

SRE's DURING FRESH BED TESTS ON AMUAY VACUUM PIPESTILL BOTTOMS

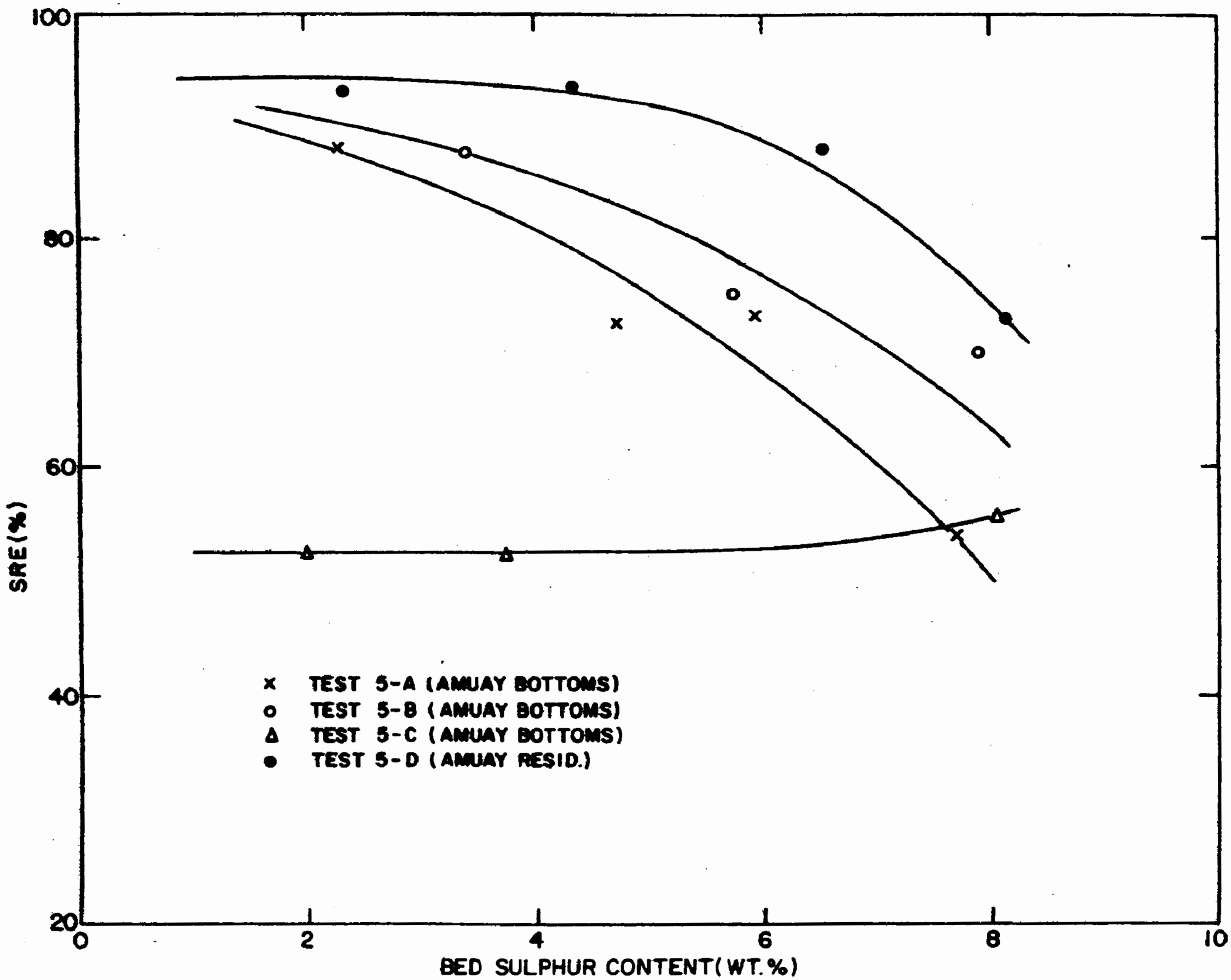

Figure 27

CARBON DEPOSITION ON BED DURING FRESH BED TESTS ON AMUAY VACUUM PIPESTILL BOTTOMS

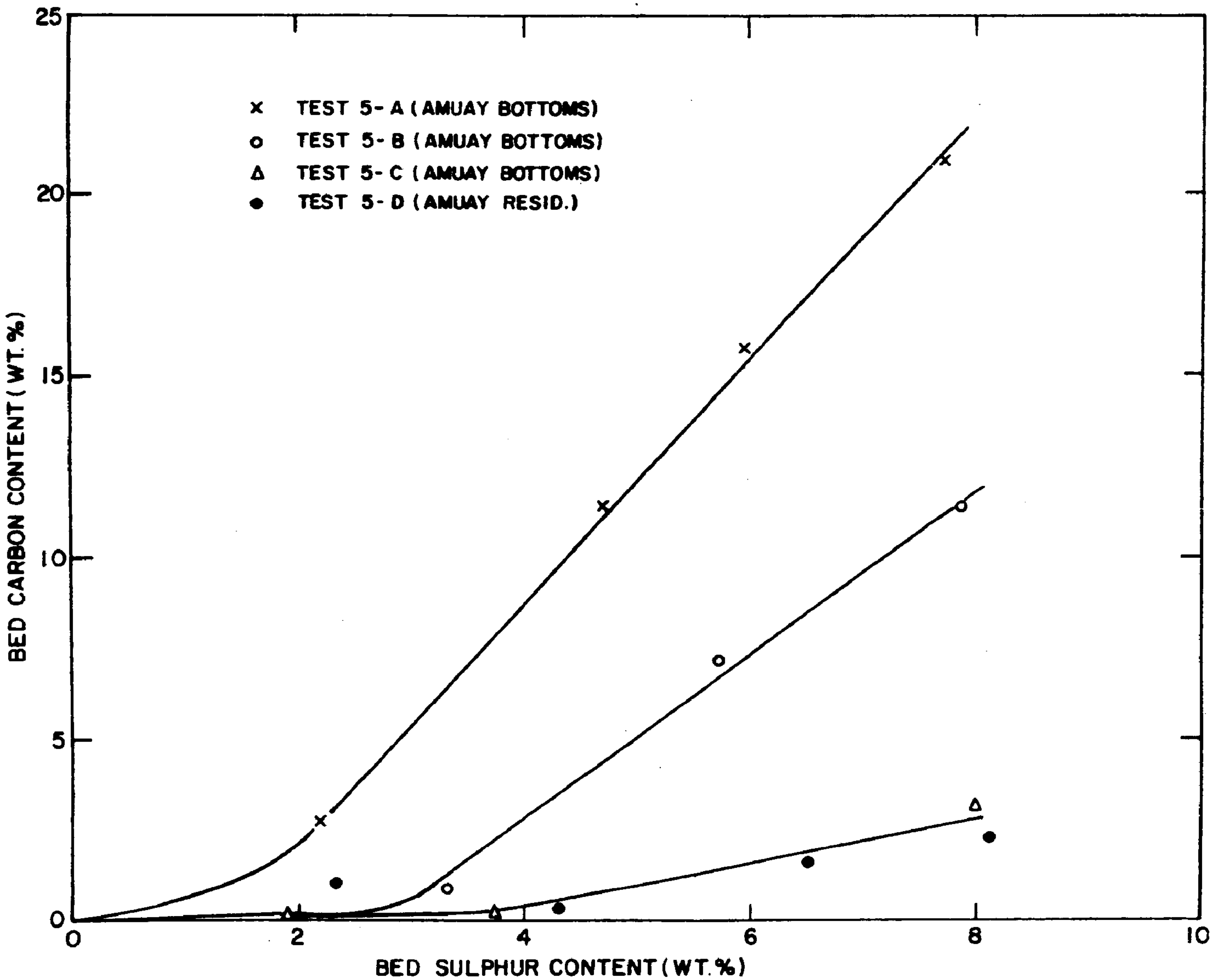

Figure 28

Table 25

Conditions for Tests Plotted in Figures 27 and 28

Test	Symbol	Oxygen/Fuel Ratio (% of stoich)	Oxygen Enrichment (% excess O_2)	Gasification Temperature (°C)	Superficial Gas Velocity (m/sec)	Initial Bed Depth (cm.)	Fuel	Limestone
5-A	X	23.0	0	855	1.77	39	Amuay Bottoms	BCR 1359
5-B	O	26.7	25	915	1.77	42	"	"
5-C	Δ	28.1	25	945	1.95	43	"	"
5-D	●	27.1	0	870	1.86	38	Amuay Resid.	"

In figure 27, SRE is plotted against bed sulphur content and in Figure 28, the rate of carbon deposition is plotted against bed sulphur content.

In Test 5-A the pipestill bottoms were studied under conditions which had been found most suitable for the Amuay Resid., the fuel which had been used to a large extent in the past. It can be seen in Figure 27 that SRE dropped quicker with increasing bed sulphur content than with the resid (Test 5-D). The likely reason for this is indicated in Figure 28 where the rate of carbon deposition on the bed was much higher with the pipestill bottoms. In Test 5-B, therefore, conditions were altered to reduce the rate of carbon deposition. Gasification temperature was raised as this has previously been shown to increase the rate of carbon burn-off in the bed (Reference 1) and the fluidising air was enriched with 25% excess oxygen. These changes together with the increase in oxygen/fuel ratio which accompanied them reduced the rate of carbon deposition (Figure 28) and gave a consequent improvement in SRE.

In order to achieve an even greater improvement and obtain a result similar to that with the Amuay Resid., gasification temperature was raised again whilst the same level of oxygen enrichment was continued. Although these conditions had the desired effect with regard to carbon deposition which was reduced to a rate comparable with that obtained with Amuay

Resid., (Figure 28) they gave lower SREs (figure 27). The likely reason for this is that the increase in temperature from 915 deg. C to 950 deg. C took the process outside its optimum temperature range for this low rate of carbon deposition (Reference 1).

In addition to these fresh bed tests, a series of 19 gasification/regeneration cycles was carried out. Conditions were similar to those normally employed with Amuay Resid. i.e. an air/fuel ratio of 27%, gasification temperature of 870 deg. C and no oxygen enrichment. Details are given in Appendix N. A lined out sulphur removal efficiency of 69% was obtained which was similar to that which would have been predicted if Amuay Resid., had been used under identical conditions. In this case, the build up of carbon to a level where it inhibits sulphur absorption did not take place as the carbon was burned off during the regeneration stage of each cycle.

High Sulphur Pitch is a brittle solid at ambient temperatures and needs to be heated to about 200 deg. C for handling as a liquid. At this temperature, however, it presents no problems as far as pumping and introduction into the fluid bed is concerned. Results obtained during fresh bed tests with High Sulphur Pitch are shown in Figures 29 and 30. Operating conditions for these tests are summarised in Table 26. Full details are given in Appendix N.

Table 26

Conditions for Tests Plotted in Figures 29 and 30

Test	Symbol	Air/Fuel Ratio (% of stoich)	Gasification Temperature (°C)	Superficial Gas Velocity (m/sec)	Initial Bed Depth (cm)	Fuel	Limestone
6-A	X	24.5	395	1.72	45	High Sulphur Pitch	
6-B	O	31.6	885	1.76	45	"	"
5-D	●	27.1	870	1.86	38	Amuay Resid.	"

SRE's DURING FRESH BED TESTS ON HIGH SULPHUR PITCH

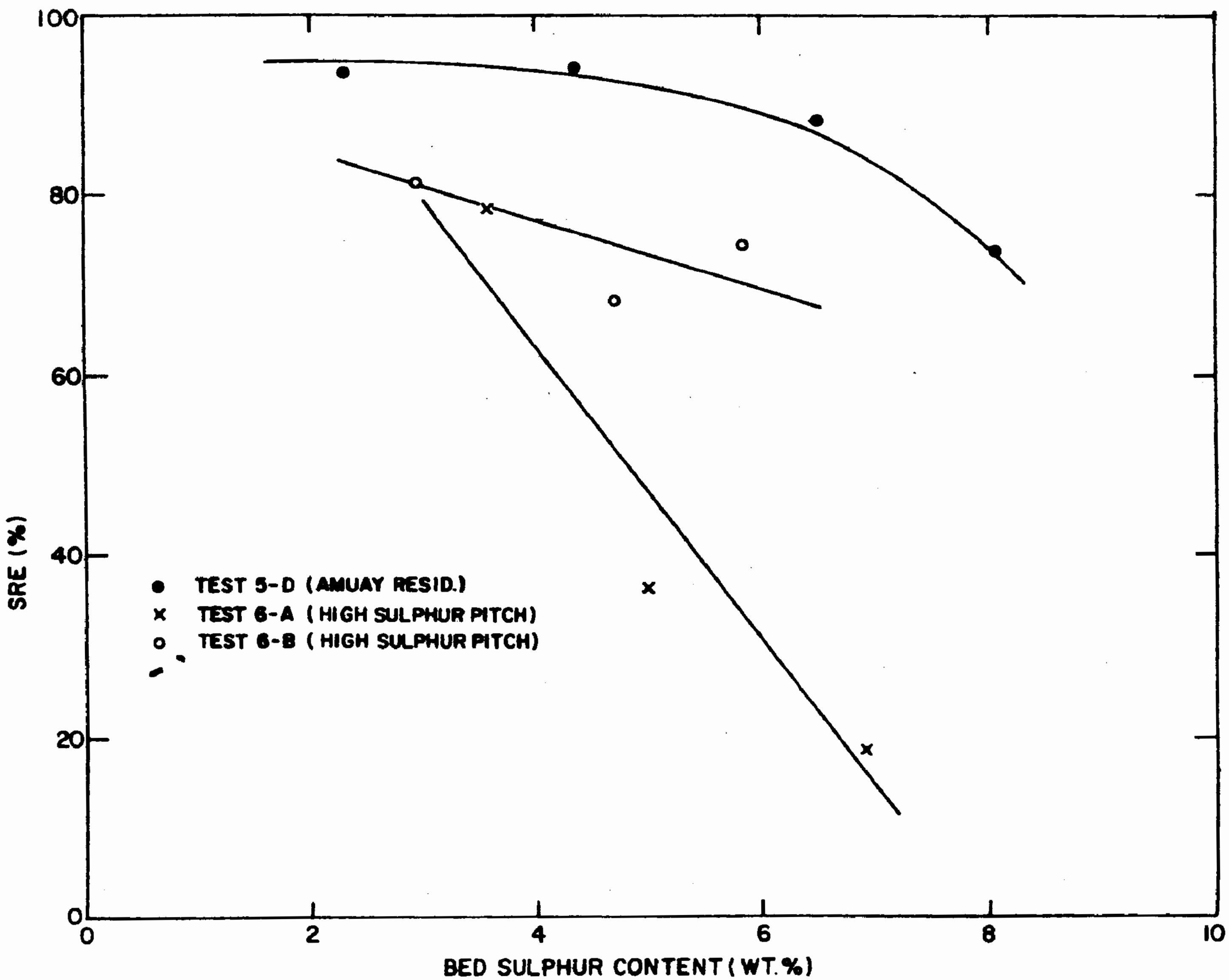

Figure 29

CARBON DEPOSITION ON BED DURING FRESH BED TESTS ON HIGH SULPHUR PITCH

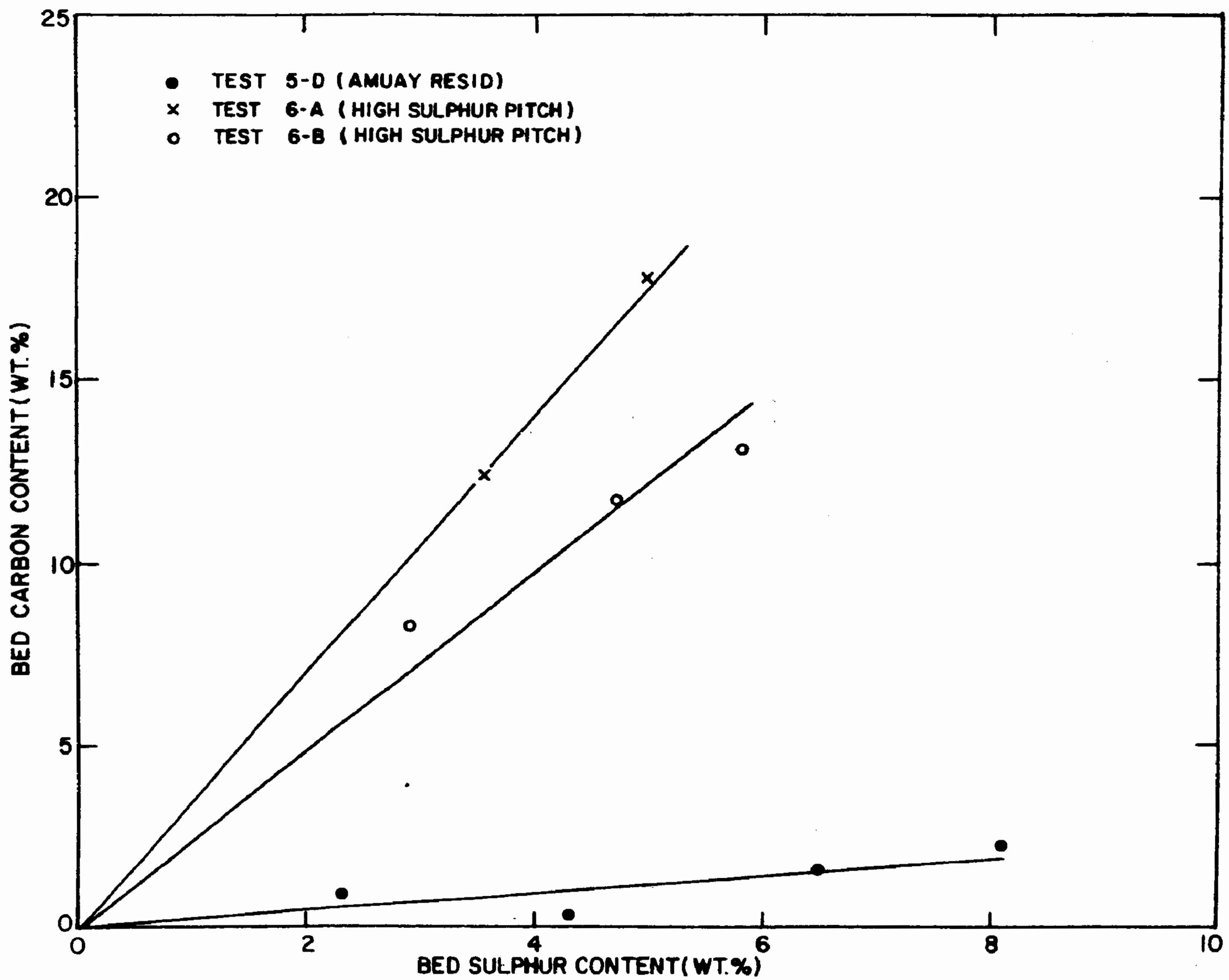

Figure 30

In Figure 29, SRE is again plotted against bed sulphur content and in Figure 30, the ratio of carbon deposition against bed sulphur content.

As with the pipestill bottoms, SRE is related to the rate of carbon deposition on the bed. At the higher rate of carbon deposition of Test 6-A, SRE fell more rapidly than at the lower rate of carbon deposition obtained at the higher air/fuel ratio of Test 6-B. Also, as expected the High Sulphur Pitch with its higher level of Conradson Carbon exhibited a higher rate of carbon deposition then the pipestill bottoms under similar operating conditions.

Comparison of SREs obtained in the tests with the heavier fuels shows that this is primarily dependent on stone carbon and sulphur levels. Similar sulphur and carbon levels giving similar SREs.

In addition to the fresh bed tests, a series of 20 gasification/regeneration cycles was carried out. Details of the test conditions which were similar to those normally employed are given in Appendix N. The lined out SRE of 79% which was obtained was similar to that expected from Amuay Resid., under similar conditions. As with the pipestill bottoms, the build up of carbon to a level where it inhibits sulphur absorption did not take place as the carbon was burned off during the regeneration stage of each cycle. In this case, the average carbon content of the stone at the end of each absorption test was 1.7% by weight.

The results on these two heavy fuels are considered to be sufficiently encouraging to warrant further tests on heavy fuels with steam injection to provide further information on a means of controlling carbon deposition without adversely affecting SRE. These tests should include a more detailed investigation into the balance of gasification temperature and steam injection required to give SREs comparable to those which can be obtained with Amuay Resid.

TASK III - SCOPING OF ENGINEERING EFFORT

The total development of CAFB through a 100+ MW demonstration test period is expected to take about 6-¼ years and require $3,320,000 in engineering effort. Of this total, $570,000 is for developmental engineering and pilot plant guidance, the remaining $2,750,000 is associated with the demonstration project. Optimistically, the total development might be accomplished in 4-¼ years with $2,520,000 of engineering effort. Conversely, greater costs and times could be experienced.

The approximately 1 MW pilot plant at Abingdon can be scaled to 100+ MW without an intermediate pilot plant but with some risk. To reduce the risk, large scale mock-up studies of the fluidisation system and special engineering development of critical equipment should be carried out. This scope includes the engineering manpower for these studies but does not include money to build any large scale test rigs. These are assumed to be included in the laboratory programme.

The approach used in preparing this study was to assess the state of development and the complexity of the process and then determine a reasonable schedule to carry the development to completion consistent with the criteria of reasonable risk. Previous experience in process and project developments similar to this was used as a guide. An estimate was then made of the various types of engineering activities required during the development and the extent of the effort required for each activity. This estimate excludes engineers who are directly associated on a full-time basis with research activities or the operation of pilot plants or demonstration plants.

The total programme has been divided into four activities, and a "most probably" and an "optimistic" development schedule has been estimated for each activity. Of course, there is also the possibility of a longer and more costly development programme if developmental problems prove extensive or if significant modifications are required to the demonstration plant due to start up difficulties.

The four activities in the development programme are:-

(1) Process Development
(2) Process Design
(3) Detailed Engineering, Procurement, Erection
(4) Startup and Test

The engineering effort is summarised in Table 27. The costs shown there are grouped to distinguish the engineering guidance costs during pilot plant development work from the engineering costs associated with design, erection, startup, and test of the large demonstration project. The project work is divided into two categories; basic engineering and the prime contractor's effort. Basic engineering includes the basic process design, owner's interest protection and follow-up during the detailed engineering and erection stage, and the engineering associated with start-up and testing. The prime contractor's effort involves the engineering required for mechanical design and erection of the demonstration plant.

In this schedule it was assumed that a client for the demonstration plant would be obtained near the end of the small scale development phase at which point a site would be selected. Certainly the earlier the client is located, the sooner it will be possible to direct both engineering and pilot plant activities towards a specific project with improved chances for shortening the time and cost of the development work.

The schedule and cost of the development activity is a major uncertainty in this type of effort. The schedule depends to a large extent on the degree of effort expended and the degree of risk which might be considered acceptable when starting the demonstration plant design. The estimate of 18 months for additional small scale development assumes minimum future pilot plant problems, very little process optimisation, and a higher risk in proceeding with the demonstration plant than if more extensive (experimentation and engineering) development work were undertaken.

Table 27

CAFB Development Programme

Summary of Engineering Effort[1]

	Most Likely		Optimistic	
	$k	Dates	$k	Dates
DEVELOPMENT	570	1/73-7/75	360	1/73-7/74
PROJECT				
● Basic Engineering				
- Process Design	420	10/74-10/75	270	1/74-8/74
- Owners Interests	580	10/75-7/77	390	6/74-9/75
- Startup and Test	550	7/77-4/79	500	9/75-4/77
	1,550		1,160	
● Contractors Design and Erection	1,200	10/75-7/77	1,000	6/74-9/75
	3,320			

(1) Cost of Engineering work only - costs of experimentation, pilot plant work, construction, and demonstration plant operation are excluded.

SECTION VII

REFERENCES

1. Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulphur Oxide Emissions. Final Report, OAP Contract CPA 70-46, Esso Research Centre, Abingdon, Berkshire, June, 1972.

2. Curran, G.P., Fink, C.E. and E. Gorin. Phase II Bench-Scale Research on CSG Process, R&D Report No. 16. Report to Office of Coal Research, Contract No. 14-01-0001-415, Consolidation Coal Co. July 1st, 1969.

SECTION VIII

INVENTIONS

1. UK 50014/72 Moss, Craig, Taylor and Tisdall

   Preventing agglomeration during regeneration of sulphides by passing stone from the gasifier into a region of the regenerator which is separated by a layer of fluidised stone from the regenerator distributor.

2. UK 24739/72 G. Moss

   Production of a highly sulphated lime from CAFB regenerator off-gas, to avoid the need for reduction of SO_2 to sulphur or production of sulphuric acid.

3. UK 29513/72 Moss and Taylor

   Reduction of attrition in fluidised beds by a two stage nozzle, the first stage being a high pressure drop orifice, and the second stage providing dissipation of kinetic energy and a non-attriting gas velocity into the fluid bed.

SECTION IX

GLOSSARY

TERMINOLOGY

Sulphur Differential	- The difference in total sulphur level on the fluid bed between start and finish of a batch gasification run, or between the inlet and outlet streams from gasifier to regenerator in a continuous unit.
Superficial Velocity	- The velocity of the fluidising gases (air plus flue gas recycle, but excluding gas and vapour formed from the fuel) in the empty gasifier or regenerator bed, at the temperature of the bed.
Fluidised Bed Depth (cm)	- (Fluid head from above distributor to gas space above the bed) ÷(Fluid head per cm of bed)
Lime Replacement	- Fresh limestone added to the gasifier, expressed as weights of CaO in the limestone added over a given period per unit weight of sulphur in the fuel gasified during the same period. Alternatively expressed as a ratio of moles CaO added per mole S in the fuel gasified.
Sulphur Removal Efficiency (SRE)	- $\left(1-\dfrac{SO_2 \text{ observed in flue gas}}{SO2 \text{ if none absorbed}}\right) \times 100\%$
Calcination	- Removal of CO_2 from limestone by heating above approximately 750 deg.C.
Adiabatic Gasification	- Operation at low air to fuel ratios in the gasifier such that heat released by partial oxidation of the fuel just serves to maintain the gasifier temperature at the required level. (Air supply about 14% of that needed to fully combust the fuel).

Combustion	-	Operation at high excess air levels during combustion in the gasifier such that the gasifier temperature just remains at the required level (Air supply about 400% of that needed to fully combust the fuel).
Megawatt (MW)	-	Used in this report only for electrical power generation rate.

SYMBOLS USED IN TEXT

A	Bed age in hours (batch tests)
D	Bed Depth (cm)
d	Particle size, microns
d_{av}	Surface area mean particle size, microns
L	Loss rate from bed, g/min
S	Bed Sulphur Content (total), weight %
T	Bed temperature, deg.C.
W	Weight of fraction in sieve analysis
μ	Micron (10^{-6} metre.)

SECTION X

APPENDICES

APPENDIX A

STARTUP AND OPERATIONAL PROBLEMS

NATURE OF STARTUP PROBLEMS

During start up of the continuous pilot plant in Run 4, problems were encountered in the following areas:-

(a) Blockage in solids transfer line

(b) Plugging in regenerator gas outlet system

(c) Dust emissions to boiler from gasifier

(d) Dust in flue gas recycle stream

(e) Dust emissions to atmosphere

(f) Regenerator Agglomerates

All of the problems were related to differences in the characteristics of stone BCR 1691 from those of the Denbighshire stone used in the continuous unit during Phase I studies (Reference 1). The major differences are lower fusion temperature, the cause of problems (a) and (f) above and production of a higher proportion of very fine dust in a fluidised bed under fully combusting conditions, the cause of problems (b) through (e). The dust produced from BCR 1691 is more difficult to retain in collection equipment than that originating from Denbighshire stone. It also clings to surfaces of pipes, cyclones, control valves etc, and is difficult to dislodge without application of direct mechanical force. It does not drain from hoppers, or even vertical pipes, without continuous rapping.

Transfer Line Blockage

Heatup of the pilot plant for Run 4 started on July 28. Stone addition was begun the afternoon of August 1, and by early on Aug 2, a hot fluidised lime bed was established under kerosene combustion conditions. However, efforts to establish good solids circulation through gasifier and regenerator were unsuccessful. Attempts to rod out the transfer lines did not improve circulation very much. The unit therefore was shutdown on August 4, allowed to cool and opened for inspection on August 8. The blockage was found to consist of a fused mass of lime particles which obstructed most of the mixing pocket in the regenerator to gasifier (R to G) transfer line. The transfer line from this mixing pocket to the gasifier also contained a quantity of material with the appearance of foamed slag.

Reconstruction of the startup procedure indicated that the obstruction was caused by limestone particles and fines entering the R to G transfer line during the initial stages of stone addition while the pocket and transfer line were heated by direct gas flame. The geometry of the system is now such that stone enters the bed from the stone feeder at a point directly opposite the R to G line. The start up burner is between the stone feed point and the R to G line. During start up, gasifier pressure had been maintained well above regenerator pressure to drive hot gas into the regenerator to raise its temperature. It is evident now that flame had actually entered the transfer line along with stone. The silica content of the BCR 1691 stone lowers its melting point enough to allow fusion under these conditions, producing, in effect, flame spraying of fused stone directly into the R to G transfer line.

The solution to this problem is to adjust pressure balance during the early stages of stone addition to avoid overheating the R to G transfer system. This method was adopted for the second start up. As added precautions, a thermocouple was installed in the transfer pocket, and low silica Denbighshire stone was added initially to form a bed deep enough to cover the transfer line. The experience gained during the second bed addition on 15 August indicates that flow of hot gas and stone into the R to G line can be prevented by adjusting the pressure balance and that no trouble would have been experienced with the 1691 stone alone. The thermocouple in the transfer pocket is a valuable guide to temperature and flow conditions at that point, and its continued use is recommended. No blockage was encountered during the second stone addition, and once good fluidisation and combustion were established, high circulation rates between beds were easily obtained. Some difficulty was encountered in establishing initial fluidisation with 300 to 3200 micron stone. Indeed each startup had encountered some trouble during the initial period of stone addition because of the high heat load required for stone calcination and the high gas velocity required to fluidise uncalcined stone. In this startup, good fluidisation and operating bed temperature were achieved after addition of some precalcined stone removed from the unit in Run 3. Use of calcined lime is recommended for startup in future runs.

Regenerator Gas Outlet System

During the startup period of Run 4, dust in the regenerator off gas stream continually blocked the regenerator cyclones and the gas exit line and control valve down stream of the cyclone. It was possible to keep the cyclone functioning only by continuously rapping on its body. Without this rapping, solids failed to drain, quickly filled the cyclone interior, and went overhead to form restrictions further down stream. It was found that the overhead line remained relatively clean in straight sections and smooth bends but plugged at sharp bends and fittings.

The character of regenerator fines changed when gasification began on August 20. Almost immediately the regenerator cyclone became free draining and operative without rapping.

The colour of the fines also changed to a darker hue. Microscopic examination showed the dust from the combustion period to have a high proportion of very fine particles. Figures 1 and 2 are photomicrographs of regenerator solids obtained under combusting conditions (1) and gasifying conditions (2). The larger particle size and reduced agglomeration tendency of the gasifying samples is apparent.

Nevertheless, the regenerator outlet control valve eventually plugged after nine hours of gasification and the run was terminated. Inspection of the outlet system revealed no accumulation of fines except at the valve itself. The solids forming the plug appeared to be more like those formed during combustion than during gasification, and it is possible that they were a remnant of the pregasification period which had dislodged from the transfer line and moved down stream to the valve.

Dust Emissions to Boiler

A large quantity of dust passed from the gasifier into the boiler. Much was retained within the boiler, particularly at the back end of the main fire tube, where the flue gases change direction abruptly through 180 degrees, but some passed through to be caught in the external cyclone or to escape from the stack. There are indications that fines losses decreased during gasification, but the period was too short to confirm this observation.

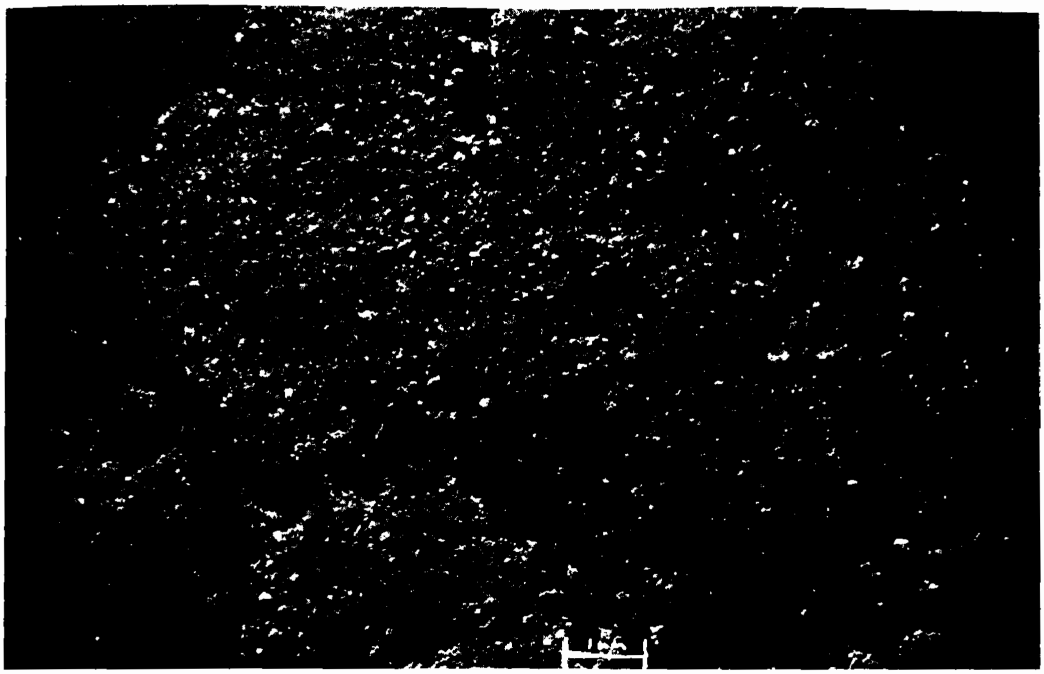

Fig. 1 Photomicrograph of Regenerator Fines under Combusting Conditions.

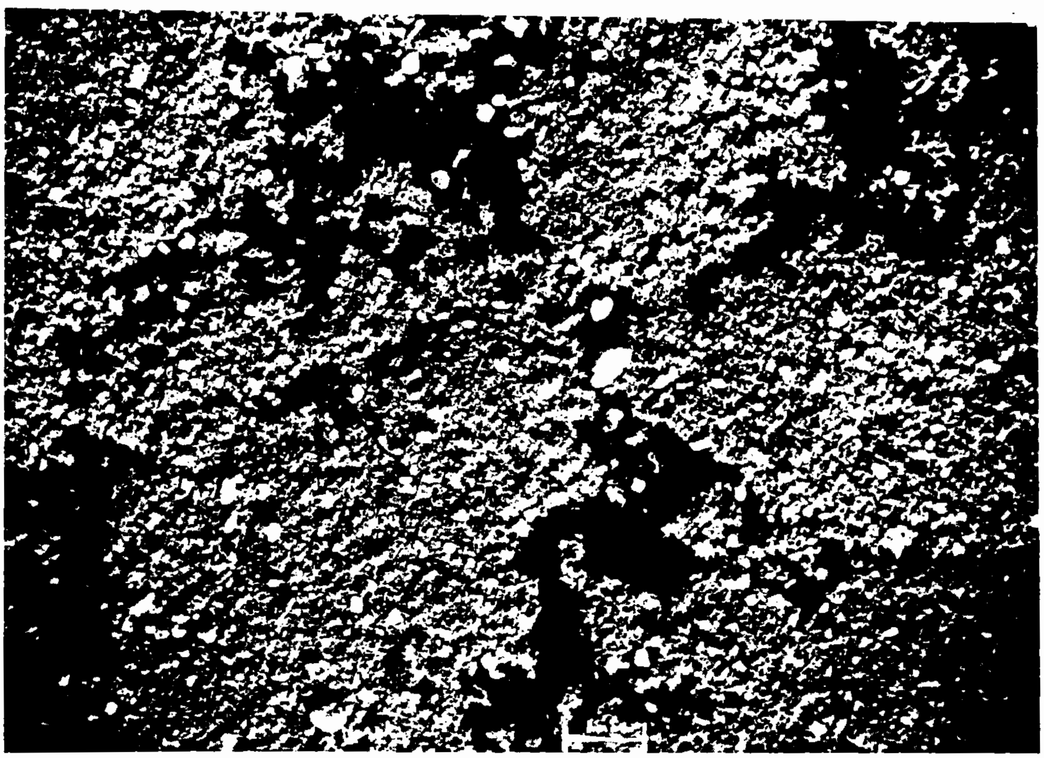

Fig. 2 Photomicrograph of Regenerator Fines under Gasifying Conditions.

Unlike previous operations where fines also entered the boiler, this time we were unsuccessful in withdrawing solids from the drain points at the boiler end. The solids failed to drain because of their steep angle of repose.

A contributing factor to high losses from the gasifier was the greater bed pressure drop used in the current run. In earlier runs, a maximum bed pressure drop of 4.73 kPa (19 inches water gauge) was used. In the current run this was increased to 6.47 kPa (26 in. w.g.).

This increase in pressure drop together with a reduced density of fluidised cyclone fines (due to lower average particle size) caused level of solids in the cyclone drain line to reach the cyclone itself and decrease cyclone efficiency.

Dust in Flue Gas Recycle

Two stages of cyclones in the flue gas recycle system failed to remove fines to a degree which would assure long term cleanliness of the gasifier distributor. No pressure drop increase in the distributor was observed during the short test period, but fines observed in the flue gas sample line filter (downstream of the cyclones) indicated that a problem eventually would have occurred.

External Dust Emissions

The external settling chamber and cyclone which proved adequate for final flue gas clean up in Run 3 was unable to cope with the fine dust produced under combustion conditions in the current run.

In part, poor performance of the external cyclone was also caused by the sticky nature of the fines. The interior of the cyclone was quickly coated with a layer of solids which impaired its efficiency.

Regenerator Agglomerates

Analysis of Run 4 temperature records and inspection of samples retrieved from the regenerator revealed an additional problem with BCR 1691 stone. During the early portion of the Run 4 gasification period, an upset in pressure balance interrupted solids circulation and allowed a brief temperature excursion. The temperature in the lower portion of the bed reached 1130 deg. C. After this upset, the temperatures in the upper and lower regions diverged with the lower temperature logging about 80 deg. C below the upper one. This condition indicates poor fluidisation. When the regenerator was opened for inspection a number of agglomerates were found. We believe that these lumps formed during the brief high temperature period in spite of bed fluidisation. In previous runs with Denbighshire stone there had been temperatures of over 1130 deg. C without encountering similar losses in fluidisation. This difference in behaviour is attributed to the lower fusion point of the lower purity BCR 1691 stone.

APPENDIX A - TABLE I

Particle Size Analysis Run 4

Sample Location	Gasifier Bed	Regenerator Bed	Solids from Boiler Fire Tube	Regenerator Cyclone Fines (gasification)
Particle Size, wt % in Fraction				
1400 Micron +	39.8	21.6	19.3	.27 *
1400 - 1180	11.5	8.8	6.3	)
1180 - 850	22.2	18.0	13.0	) .27 *
850 - 600	15.3	13.9	13.0	)
600 - 355	11.0	14.7	20.5	.27 *
355 - 250	0.5	5.8	7.1	.27 *
250 - 150	0.2	12.7	5.1	10.6
150 - 106	0.1	0.8	2.8	11.4
106 - Dust	0.5	3.7	13.0	76.9
Bulk Density g/cc	1.13	1.08	.85	1.03

* These particles had a white appearance, distinctly different from that of the finer particles.

APPENDIX A - TABLE II

Chemical Analysis of Lime Samples - Run 4

	Units	Gasifier (After Shutdown)	Regenerator (During Gasification)	Boiler (After Shutdown)	Regenerator (During Combustion)	Cyclone Fines (During Gasification)
CaO	wt%	53.9	58.8	57.0	68.1	59.9
MgO	"	3.95	4.7	4.35	4.1	4.2
Al_2O_3	"	2.35	3.0	2.85	4.1	3.6
SiO_2	"	22.1	24.1	18.6	21.2	21.6
Fe	"	0.73	0.87	1.02	0.97	0.87
Na	"	0.06	0.07	0.07	0.06	0.08
V	"	0.13	0.15	0.11	0.04	0.38
CO_2	"	1.84	0.73	0.36	0.21	0.12
S (total)	"	4.88	2.09	1.56	0.74	5.51
S as sulphate	"	4.39	2.06	1.28	0.72	2.58
Loss on Ignition	"	-	-	7.46	0.31	-
Gain on Ignition	"	0.21	0.16	-	-	4.09
SiO_2/CaO		0.41	0.41	0.33	0.31	0.36
MgO/CaO		0.074	0.080	0.076	0.060	0.070
Al_2O_3/CaO		0.043	0.051	0.050	0.060	0.060

SiO_2/CaO in Original Limestone				0.27
MgO/CaO	"	"	"	0.074
Al_2O_3/CaO	"	"	"	0.046

APPENDIX B

RUN 5

Operational Log, Inspection, and Data

APPENDIX B

CAFB RUN 5

OPERATIONAL LOG

2.2.73 to 6.2.73 (Unit warm up)

The warming up procedure started at 04.30 continuing at a steady rate with the propane burner. At 11.30 on 5.2.73 the temperature was sufficiently high to introduce kerosene using the three metering pumps. At 17.30 hours Denbighshire limestone was fed to the unit with the fuel rates adjusted to maintain an adequate bed temperature rise. Bed circulation and fluidisation was good.

6.2.73 (Day 1 of Gasification)

Preparations were made for gasification and at 15.00 the boiler door was shut and check out of the flue gas recycle system completed and gasification commenced. Soon after the start of gasification there was a series of automatic plant shut downs caused by a combination of boiler low pressure alarm actuation and cooling water high temperature alarms. Both these safety features are automatic in their operation and complete plant shut down cannot be overridden in these circumstances. The shut downs were started by the boiler water outlet temperature which began to climb from its normal level of 100°C to the alarm level of 120°C whilst maintaining a constant boiler water inlet temperature suggesting a fall off in water flow to the boiler. This situation had been seen briefly on Run 4 and instrumentation had been added to the pump on the boiler cooling water to monitor pump performance. At this period the pump discharge pressure had fallen, possibly due to an air lock in the pump suction line which was then bled releasing a considerable quantity of air from the line.

At the same time adjustments were made to the burner air distribution because observations had shown a very short intense flame in the boiler which could have caused some local boiling in the boiler. Either one or both of these actions re-established good water circulation which in turn cooled the boiler system sharply and the pressurisation unit pumps were unable to cope with the make up water required to maintain pressure and the low water pressure alarm operated and the plant shut down. Then followed a series of problems in which the pressurisation unit was unable to maintain pressure within its operating band and when the automatic cooling control valve cut in, the pressure dropped towards the low pressure shut down condition. Attempts to assist this transient by cutting the secondary side cooling pump slowed down the cooling rate but permitted local convection circuits to activate a high temperature alarm and shut down the plant again. After a series of shut downs due to either low pressure or high temperature, the control system on the pressurisation unit was reset with a wider operating band and some remaining air bled from the boiler circuit and this established a workable control system. At 20.45 gasification restarted and the unit settled out and gradually the regenerator came up to operating temperature by manual control of bed circulation.

7.2.73 (Day 2)

There was some difficulty in getting the regenerator temperature up to 1050°C and the gasifier temperature was increased to 950°C to help this problem. It was then apparent that bed transfer between gasifier and regenerator was improving and control was transferred to the automatic controller. The gasifier temperature was lowered to 870°C without any problems with the regenerator which was operating with an 8% SO_2 stream.

At 14.45 conditions were lined up for the first data point but steady regenerator conditions were now difficult to hold and at intervals the automatic temperature controller was unable to properly regulate the flow of material and high temperatures resulted in the regenerator.

The venturi scrubber drain on the flue gas recycle line became blocked and flooded the blower with water and the drain was rodded out to clear away the accumulation of caked lime dust which had been washed out in the scrubber.

8.2.73 (Day 3)

The regenerator to gasifier transfer line was partially blocked but was rodded out with some improvement in material transfer. The pressures on the transfer line nitrogen pulsers were reduced to investigate their effect upon bed transfer. At 15.15 bed material and dust samples were taken.

At 20.30 all the analytical equipment was checked and a water knock out pot placed in front of the boiler sample line cotton wool filter to reduce the water carried to this filter. At 23.30 further bed material and dust samples were collected.

The stone feed was temporarily stopped to determine the material loss rate and after 3 hours there was no appreciable change in either of the two bed levels.

9.2.73 (Day 4)

Fluctuations in the boiler burner throat temperature were observed which seemed to correspond with the operation of the fines return transfer system. The nitrogen pressure setting on the left hand vessel aerator was reduced to try and minimise the effect of these residual gas pulses which could pass up the cyclone drain after each operation of the fines transfer system. This action was almost immediately followed by a blockage in the transfer system which was subsequently cleared by raising the pressure back to its original value and knocking the pipe to encourage transfer.

Prior to adding BCR 1691 limestone the bed was lowered by draining 68 kgs (150 lbs) from the regenerator. It was observed that during this period with a low gasifier bed level the gasifier space pressure increased more sharply and this may have been because less material was thrown into the cyclones to scour the deposits.

At 04.55 BCR 1691 stone feed was started to build up the bed back to 64 cms (25 ins). The main air compressor on the site service supply developed a faulty valve which let most of the supply bleed to atmosphere and reduced the unit supply pressure to such an extent that the pneumatic controllers and rappers all started to malfunction. The fault was temporarily rectified and the unit then left to steady out after these disturbances.

The regenerator performance was not very good and at 08.30 the control temperature was lowered to encourage more circulation and hopefully better regeneration.

The limestone feed rate of 27 kgs/h (59.4 lbs/h) which was used to build up the bed height produced a larger quantity of material in the cyclone transfer system and also produced more problems in obstructions in the feed line to the elutriator.

At 15.00 three boiler gas samples was taken and gave 110 cms^3/m^3, 108 cms^3/m^3 and 101 cms^3/m^3 NO_x by the chemiluminescene method of analysis. During the afternoon trouble was experienced with the main flame failure alarm which repeatedly cut in but all other instruments were normal and it was assumed to be an electrial fault or dust masking the flame eye vision. The boiler sampling line became blocked in the boiler door and after cleaning this obstruction the Wostoff analyser showed higher SO_2 levels in the boiler.

At 23.30 some problems arose when the scrubber water separator drain blocked again and water was carried through the recycle blower and some reached the gasifier plenum from which water dripped for the following few hours.

10.2.73 (Day 5)

Some flame outs of the main flame pilot were encountered and the trouble was found to be caused by a blockage in the sighting tube for the pilot eye which was rodded out and this cleared the problem.

At 01.30 a sudden upward jump in the gasifier temperature accompanied by a drop in the pump delivery pressure showed that the fuel supply had run out due to some misunderstood directions about the supply situation. After some initial problems in reestablishing the oil supply the main flame was lit and the unit allowed to line out.

The gasifier bed density manometer showed a gradual lowering of the bed density during the preceeding hours of gasification suggesting the cyclones were retaining and the system returning a large proportion of the generated fines. At 08.15 the elutriator nitrogen rate was increased to remove some more fines because of the continued decrease in the bed density. The regenerator performance of 4-5% SO_2 was still well below previous results when values up to 10% were achieved.

Bed material and dust samples were taken at 10.45 before changing from this data point condition to one where the stone feed was reduced to about 15 kgs/h (33 lbs/h).

At 10.30 the boiler flue gas temperature began to rise at about 1°C per hour. About this period there were problems with an obstructed pressure tapping in the regenerator bed depth measurement and whilst this was drilled out the regenerator bed temperature rose to 1100°C but was prevented from going higher by the nitrogen quench system which had been installed for this overtemperature situation. These two problems were interlinked because when the bottom tapping blocked the bed height and hence the air rate could have changed so upsetting the balance of material flow and temperature too sharply for the automatic controller to follow.

At 17.00 the pneumatic controller on the left hand cyclone drain system stopped working because of some very fine dust which had passed through the filter and the purge in the pressure tapping on the vessel finally getting into the miniature control valve. This valve was removed and cleaned whilst a fine filter and longer purge line were installed. The systems were reconnected by midnight and normal stone feed resumed to re-establish conditions.

11.2.73 (Day 6)

The gasifier bed depth of 63 cms (24.6 ins) did not throw up much material into the cyclones and SO_2 in the boiler gradually decreased when the gasifier bed depth increased. The regenerator cyclone rapper stopped for a period and during this time there was some laydown of fines in the system because when the rapper was restarted the regenerator back-pressure suddenly decreased suggesting the removal of some blockage.

At 07.30 samples were taken of bed material and dust from the various collection points. The pressure in the gasifier space had reached 6.7 kPa (27 ins wg) due to carbon deposited in the cyclones and ducts which was removed by the standard operation of bed sulphation and a burn out procedure. At 10.30 the bed sulphation commenced and by 11.05 the process was complete. This was followed by burning out the carbon in the cyclones and ducts using the recycle system with a controlled oxygen content so keeping the duct temperatures below 1200°C. After the burn out was completed the unit was put onto combusting conditions with kerosene and maintenance completed on the venturi scrubber which was obstructed in the throat with lime deposits. The bed circulation under these combusting conditions was very bad and difficulty was experienced in getting material into the regenerator. The high regenerator distributor pressure drop suggested some obstruction in this area - possibly a fused lump on top of the distributor which was preventing fluidisation of the material. Troubles was also found with both the pneumatic control systems on the fines return from the cyclones. It was decided to shut down the plant temporarily to remove the distributor in the regenerator.

12.2.73 (Day 7)

The distributor was removed but the bore of the regenerator was completely clear. There was some obstruction in a few of the distributor holes which could have accounted for the high pressure drop. The gasifier to regenerator transfer line horizontal leg was rodded out and the distributor was replaced but bed circulation was still bad and all attempts at rodding, pressure adjustments and changing pulse settings did not improve matters significantly.

13.2.73 (Day 8)

Gradually bed transfer improved as the bed depth increased in the gasifier and the problems with the fines pneumatic control system were resolved. At 12.50 the boiler was cleaned out prior to preparation for gasification and at 19.20 gasification was restarted.

14.2.73 (Day 9)

The regenerator fluidisation was not satisfactory because below 1.22 m/s (4 ft/sec) the bed temperature at the bottom of the bed dropped sharply. The irregular behaviour of the regenerator bed temperature control thermocouple made automatic control very unreliable and manual control was resumed. At 04.30 the regenerator defluidised with the bottom temperature dropping 200°C and the upper temperature exceeded 1100°C. Further attempts at achieving good bed transfer were unreliable and once again the regenerator defluidised with a high top temperature excursion. At 05.35 the unit was sulphated and the ducts burnt out because continued operation could have produced a fused lump in the regenerator with repeated defluidisations and high temperatures.

The unit was set back on combusting conditions so that the problem associated with the regenerator and erratic fines transfer could be solved. The fines return system was stripped out and flakes of carbon and lime which were released into the system after the burn out were found obstructing the transfer lines. The right hand cyclone drain vessel was removed to permit access to the gasifier to regenerator transfer line drain plug. The regenerator distributor was removed and replaced with a temporary insulated mild steel pad inserted so that the unit could be restarted under combusting conditions whilst the distributor was investigated. The regenerator bore was obstructed with a considerable deposit of agglomerated material in the lower section.

15.2.73 (Day 10)

Further work was carried out on the fines control system and automatic feed of the fines back to the gasifier to free all obstructed areas and the system was reinstalled satisfactorily. The gasifier lower pressure tapping blocked and was drilled out. Examination of the regenerator distributor showed that the raised lip of refractory around the nozzles had broken away in one area and this was repaired by building up with refractory cement and refitted into the unit. Further trials on bed transfer were not very successful and persisting problems were encountered with the gasifier lower tapping which continued to plug up. Eventually the lower tapping was removed and replaced with a new one, having become totally obstructed with solids. The bed transfer lines were rodded to encourage circulation and after draining some material from the regenerator distributor drain, the flow rate improved and by 21.30 circulation was quite reasonable.

16.2.73 (Day 11)

The boiler was cleaned after the period of combusting conditions, door resealed and made ready for gasification. Further trouble was encountered with the fines return system becoming obstructed with flaky material which continued to fall from the cyclones. These flakes could only be removed by dismantling the transfer pipe from the pressure vessel and the cone feed control at the gasifier feedpoint.

At 12.15 gasification was restarted with some difficulties occurring in the fines transfer due to carbon-lime flakes which continued to drop from the cyclones. The scrubber also blocked at its entry point but was cleared by hammering the pipework. A stainless steel screen was fitted into the fines return line to the gasifier to arrest the carbon flakes before they obstructed the control valve.

17.2.73 (Day 12)

The regenerator continued to act in an erratic manner with uncontrolled temperature excursions to 1100°C and there were many periods of poor fines transfer mainly due to flakes falling from the cyclones. Some of the problems of poor fines return were eliminated by adjusting the outlet ball valve seats which had become worn and were not gastight. At 16.00 the unit was reasonably steady and conditions were held for 2 hours so that a set of samples could be taken at the beginning and end of the data period. After this was completed, the gasifier bed depth was reduced in preparation for the next data point. The boiler probe controller malfunctioned and permitted the temperatures of the probe to rise well above the 600°C control point.

18.2.73 (Day 13)

At 04.00 and 06.00 sets of bed material and dust samples were taken. The gasmeter measuring the nitrogen for the gasifier to regenerator transfer system was found to be leaking and was replaced.

The boiler SO_2 level appeared to be sensitive to regenerator performance and efforts were made to maximise the SO_2 output from the regenerator thus giving minimum boiler SO_2 levels. The gasifier pressure had gradually risen and at 09.00 there was gas leaking from the gasifier lid and a sulphation and burn out was necessary. Some of the bed material was removed before this procedure and by 12.30 the operation was completed. There was a fuel leakage from the unit showing as a distillate dripping out of the bottom with distillate fumes coming out of the top around the lid. A pressure check on the shell space showed that the inner refractory concrete was not withstanding any differential pressure which was contained by the steel casing.

The gasifier bed temperatures showed a spread of 100°C suggesting poor fluidisation and a possible contributory factor to the apparent incomplete fuel combustion. A bed sample was sieve analysed and 74% of the bed material was above 1400 microns showing that there could be fluidisation problems. This was supported by the poor regenerator bed behaviour.

19.2.73 (Day 14)

The fuel injectors were all rodded through but without much change in the unit behaviour. Some BCR 1691 bed material with 40% of the particles above 1400 microns was fed to the unit to lower the average particle size and at 09.30 a bed sieve analysis showed 51% of the bed was above 1400 microns.

Further rodding of No. 3 fuel injector with a high pressure nitrogen lance cleared an obstruction at the discharge end of the injector. The gasifier bed was slumped and the injector withdrawn whilst a purge of nitrogen was arranged to prevent the outflow of volatile product which was still leaking from this area. The end of the injector had burnt away leaving an unrestricted flow giving a very poor spray pattern and bad combustion conditions so permitting unburnt fuel to leak into and crack in the hot zones. The vapour passed up the expansion gaps around the gasifier refractory to emerge around the top of the unit with the remaining product leaking from the bottom of the gasifier.

<u>20.2.73 (Day 15)</u>

Attempts were made to break up any lumps in the bed by rodding out with stainless steel high pressure nitrogen lances through various access points. A temperature traverse was made through the bed above the distributor to investigate quality of fluidisation which looked reasonable. The permanent low value of the bottom bed thermocouple may have arisen from a build up of material around the couple shielding it from the correct bed temperature.

The remaining two fuel injectors were withdrawn and although apparently undamaged, new ones were fitted to eliminate any further difficulties from this source.

The gasifier bed fluidisation improved after this work and the unit was brought back to temperature and circulation checked out. At 21.30 the boiler was cleaned out and preparations were made to start gasification.

<u>21.2.73 (Day 16)</u>

Some problems were encountered with the regenerator, gasifier and fines return pressure tappings which had become blocked with fine dust during the long period of combustion conditions. There were additional problems with the butterfly valves on the cyclone drains which did not shut off tightly without manual assistance.

Gasification was restarted at 06.00 and at 10.00 the boiler SO_2 level was about 500 ppm and the regenerator output about 3% SO_2. The carbon flakes continued to occassionally block the cyclone fines return system and the scrubber became fouled again with lime deposits.

At 20.30 there was a main flame failure alarm which reset but during the following hour this recurred six times. The fuel oil supply tank was switched in case of starvation and the problem did not reappear.

The fines return to the gasifier was erratic and could not match the supply rate so that some of the material was externally drained and fed back into the unit with the stone feedstock while the problem was resolved.

22.2.73 (Day 17)

The right hand cyclone was erratic in its material return mainly caused by the coarseness of the material which does not move as well as fine material in this type of transfer system. Samples of bed material and dust were collected at 11.30 after a period of fairly steady operation.

Pneumatic delay valves was fitted to the operators of the cyclone fines outlet valves to ensure that they shut off after the opening of the butterfly valves at the foot of the cyclones. This prevented the partial depressurisations of the transfer vessel into the cyclone drain leg so releasing dust into the boiler. The perforated plate which was inserted into the cyclone fines return to the gasifier reduced the material flow rate back to the gasifier and was replaced by one with a larger open area to prevent the material building up in the system. Samples were taken of bed material and dust at 18.00.

The regenerator fluidisation was erratic with temperatures spread by 50°C but the situation was improved by periodic draining of material suggesting a build up of coarser particles which were close to defluidisation. During this period there were problems with the regenerator air rate which showed some supply limitation and coupled with the bad circulation of bed material, regenerator temperature control was very difficult. About 22.00 the regenerator temperature rose to almost 1100°C coupled with apparent defluidisation and it was necessary to sulphate the bed and burn out the ducts to prevent the formation of a solid plug in the regenerator.

23.2.73 (Day 18)

The bed sulphation and duct burn out was completed by 01.00 and the unit was put into combusting conditions. The bed level was lowered by removing 98 kgs (216 lbs) of bed material and preparing to feed Denbighshire limestone. The regenerator and gasifier beds were sampled and sieve analysis showed 75% and 68% respectively of material above 1400 microns indicating a large average particle size in the unit. The regenerator distributor was removed again to check the erratic performance but the bore was generally clean with only small areas of deposits on the joint between the distributor and the wall and some deposits between the nozzles. The cracks in the bore of the regenerator concrete were not obviously worse than before the run started but repairs were made to the lower section cracks using Sairset cement to prevent air bypassing the fluid bed.

<u>24.2.73 (Day 19)</u>

The regenerator distributor was repaired in those areas of the sealing face where the material had cracked away and the assembly installed with a soft high temperature insulation layer to act as a sealing gasket onto the refractory. The problems of achieving a good seal on this face during re-assembly were not helped by the high temperatures in the area. The transfer lines to and from the regenerator were rodded without meeting any obvious obstruction.

Fresh bed had not been added during the previous 15 hours and a sieve analysis of a gasifier lower bed sample showed that 47% of the material was greater than 1400 microns which was a fairly typical value from previous experience. This size range of material should have fluidised easily and did not explain the continual difficulties in obtaining good regenerator fluidisation and transfer to and from the regenerator.

<u>25.2.73 (Day 20)</u>

Better regenerator fluidisation was obtained by draining out a quantity of the bed and allowing it to refill with hot stone which increased the actual gas velocity in the bed without increasing the flow and pressure drop through the distributor.

At 08.45 regenerator transfer and fines return systems were both working well and preparations were made to start gasification by checking out instruments, cleaning drain lines and sample lines. At 18.00 gasification started without problems and conditions left to stabilise.

<u>26.2.73 (Day 21)</u>

The main feature of this data point was the high fuel rate of 215 kgs/h (474 lbs/h) with the minimum flue gas recycle rate. It was not possible to get to adiabatic conditions but the flue gas was reduced to 68 m^3/h (40 ft^3/m) with a superficial velocity in the gasifier about 1.35 m/sec (4.1 ft/sec.). Bed material and dust samples were taken at 07.00.

The interaction between the cyclone fines return into the gasifier and the boiler SO_2 concentration was observed again - adjustment was made to the transfer system pressure levels to reduce the gas leakage and blow back up the cyclone drain leg at each cycle operation. When the fines were diverted from returning into the gasifier bed, the boiler SO_2 level steadied indicating that the irregularities in the boiler SO_2 were caused by the fines injection. At 18.00 a further set of samples was taken and sieve analysis on two samples from the gasifier bed showed 36% and 30% above 1400 microns at 20.05 and 19.30 respectively.

27.2.73 (Day 22)

The unit ran steadily with the main problems occurring in the fines return transfer vessel operation. A set of samples was taken at 07.15 and gas samples were collected from the gasifier at 10.30.

The cyclone drains appeared to be partly obstructed because there was not very much material transferred at each operating cycle and the boiler dust collection systems were picking up more material than usual. The regenerator transfer system was erratic causing problems with temperature control, but the situation was improved by rodding out the lines with a nitrogen lance. A further set of samples was collected at 17.30 including two gas samples.

The gasifier space pressure had gradually built up to 7.0 kPa (28 ins w.g) and a sulphation and burn out was started at 19.00. After completion of this procedure the unit was set on combusting conditions with kerosine so that checks would be made on various systems before resuming gasification.

29.2.73 (Day 23)

The cyclone drain systems were apparently blocked because there was hardly any delivery of material to the elutriator. The valves above the cyclone outlets were used to pass a long nitrogen lance through to the cyclone drain legs and this displaced some lumps and finer material. After some further rodding of transfer lines the bed circulation was reasonable and preparations were made for gasification which was resumed at 13.00 after boiler cleaning. Soon after gasification started, lumps of material were still falling from the cyclones and obstructing the transfer system which was freed by removing the pipes on the vessel outlets and taking out the carbon lumps which bridged the bore of the pipe.

The analytical instruments were calibrated and the boiler SO_2 analyser showed considerable drifting about the calibration point and a new amplifier was installed. Corrections were made to previous data affected by this error assuming a linear build up of the error from the last calibration point.

1.3.73 (Day 24)

The stone feed rate of 30 kgs/h (55 lbs/h) over the first few hours had not built up the bed height significantly and the rate was increased to 36 kgs/h (79 lbs/h). The first period of this day was quite smooth without any significant mechanical problems.

At 12.15 the make up rate was reduced to 15.9 kgs/h (35 lbs/h) stone feed and conditions lined out with fairly smooth operation of the plant. At 22.00 there were further problems with the Wostoff boiler SO_2 analyser giving a higher value for the calibrant gas than normal. Back up analysis by Draeger on the boiler line gave higher values than the SO_2 analyser but the sample temperature was higher than ambient which would introduce some error in measurement.

2.3.73 (Day 25)

At 05.30 a complete set of samples were taken of the bed material and at the dust collection points. Further investigations were made into the effect of moisture in the boiler sample line by frequent replacement of filters and higher SO_2 values were subsequently recorded. At 14.00 a further set of samples was collected. The regenerator bottom tapping blocked periodically but it was cleared using the nitrogen lance.

3.3.73 (Day 26)

The gasifier gas space pressure was gradually rising towards the maximum recommended level and a set of samples was taken at 04.00 with a stoichometric stone feedrate before raising the stone feed rate to obtain two more data points before shut down. There was a pilot flame failure at 14.30 but it restarted after the flame eye was cleaned and replaced. Further calibration checks were made on the boiler SO_2 analyser which again had drifted away from its previous level. An appropriate correction factor was used in the analyser over the period since the last calibration was made assuming that a linear drift had occurred.

At 18.00 the last set of samples was taken prior to a controlled shut down at 19.00 completing Run 5. After the shut down all air entry points to the unit were closed and a small purge of nitrogen was introduced into the bed so that the carbon present in the unit would not be burnt away before an inspection could be made.

APPENDIX B

CAFB RUN 5

INSPECTION

Gasifier and Regenerator Unit

Gasifier Concrete

The walls of the gasifier were generally blackened with carbon and in the upper regions around the junction between flue lid and the walls the carbon was up to 6.3 mm (¼") thick. There were patches immediately above the cyclone inlet ducts where the carbon had burnt away suggesting that an air leak has occurred after the shut down. In the lower areas, the wall was glazed with a hard thin tenacious carbon deposit.

The cracks in the concrete hot faces which have been present from the first firing of the unit showed their customary fine black deposit of carbon about 25 mm (1") wide in the upper areas of the wall. The lower areas were clean because of the splashing action of the bed material. There was no sign of any deterioration in the concrete from this test run.

There were areas at the junctions between the distributor and the walls where a deposit of fine material had agglomerated together to form a small covering between the distributor and the wall, most likely caused by an area of poor fluidisation due to a blocked or partially blocked distributor nozzle.

Gasifier Penetrations

The thermocouples, fuel injectors and pressure tappings were in good order throughout although the left hand fuel injector had been replaced during the run.

The thermocouple in the lid had a considerable growth of carbon around its protruding end whilst those in the bed area were generally clean, apart from the one at the lowest point in the bed close to the fines return pipe. There was some agglomeration of lime and carbon bridging between this thermocouple, the distributor and the gasifier wall. This condition probably arose from poor local fluidisation due to an obstructed nozzle in the distributor together with the introduction of the fines from the return system into this poorly fluidised zone. This thermocouple did show a consistently low reading during the latter part of the run which was probably caused by this local static material.

The fuel injectors were layered with carbon on their protruding sections and the injector at the right hand side had a hollow cap of carbon and lime enclosing its end.

The two air injection tubes which pass through the lid to direct air into the cyclone inlets were both heavily scaled and burnt away at their protruding tips.

The stainless steel tubes for the stone feed and fines return were both in good order.

Cyclones

The cyclone bodies and their inlet sections which had both been lined with type 310 stainless steel to improve their surface finish and hence performance were heavily obstructed with a mixture of carbon and lime.

The inlet ducts are illustrated in figure B-1 and figure B-2 which show the black deposits around the inlets which gradually become lighter further into the cyclone.

The white area on the gaisifer wall immediately above the cyclone inlets can only be explained by some areas of carbon burning out after the unit was shut down possibly due to some local movements which may have broken any seals existing between the underside of the lid and the gasifier wall.

Figure B-1 also shows the white irregular deposits on the cyclone outlet tube which can be seen hanging down inside the cyclone body.

Figure B-3 shows a view into the right hand cyclone after removal of the outlet tube shown in figure B-4. The cyclone upper body was obstructed around its total circumference leaving the hole in the centre formed by the outlet tube. The material was laid down in a very irregular manner consisting of layers and folds of fine hard material with a tortuous gas path amongst this deposit. The deposit becomes less pronounced towards the bottom of the cyclone compared to the upper section, but still filled a considerable volume of the lower section. The stainless steel liner was badly scaled and distorted, in some areas it was completely burnt away due to the high surface temperatures when carbon was burnt off at the various burn out operations during the run. In the upper sections of the cyclone the liner was soft and easily broken into flakes but nearer the bottom of the cone the lining was generally stronger and some quite large pieces of steel were removed intact. The steel lining tube which sealed off the cyclone drain passage to the gasifier to regenerator transfer line was heavily scaled with a small area burnt away on the top retaining collar.

The deposit at the intersection of the rectangular inlet with the cylindrical cyclone body was layered with white fine material separated by thinner blacker layers, suggesting that the thicker white layers are laid down during some of the combusting periods. This is confirmed by the batch unit tests which showed that combusting conditions give rise to the most severe material deposition.

The silicon carbide outlet tube of the right hand cyclone shown in figure B-4 was removed with very little material adhering to the outer surface. Inside the tube there was an overall thin layer of carbon and lime about 1.6 mm (1/16") thick deposited around the bore and in some areas near the bottom of the tube there were thicker irregular deposits up to 15.9 mm (⅝") thick. These thicker layers could be removed fairly easily but the thinner layers were very firmly bonded to the tube.

The left hand cyclone was also severely blocked in the inlet section and in the body of the cyclone although it was not so severe as the right hand cyclone (figure B-5). The cyclone outlet tube was coated on its outside with a thick black flakey deposit (figure B-6) but the deposits did not completely bridge across the gas passage between the outlet tube and the cyclone wall.

The bore of the tube was coated with a layer of carbon and lime about 1.6 mm (1/16") thick deposited fairly evenly over the surface of the tube.

The stainless steel cyclone liner was severely scaled and locally distorted or burnt away in many locations near the top of the cyclone. Towards the lower end the lining was almost intact although still heavily scaled.

The stainless steel tube sealing off the drain to the regenerator to gasifier transfer line was heavily scaled and around the upper outside surface there were crystalline deposits of carbon which could have come up the transfer line from the gasifier bed. The upper retaining collar around the tube had burnt completely away.

Gasifier Distributor

The distributor was generally in good condition apart from some obstruction in the holes in the nozzles. The obstruction arose from two sources - one was from lime particles which entered from the gasifier bed and the other deposits have come from fine rust scale carried through from the flue gas recycle line which was carrying a saturated gas from the water scrubber.

The distributor nozzle design has three staggered holes in series with each other, the first smaller one to provide the pressure drop needed in a fluid bed distributor and the third larger one provides a low exit velocity to minimise damage to the stone in the bed. The middle hole forms a plenum between the inlet and outlet holes.

Generally, the outer holes were obstructed more with lime particles and the inner holes with deposits of rusty coloured material. There were considerable problems encountered in removing the distributor because it had become wedged with limestone and heavy fuel oil when one of the fuel injectors failed during the run. Some of the mechanical force used to remove the distributor may have dislodged some of the material found in the nozzles but examination did show that 42 of the 192 outlets were obstructed completely, 18 of the inlet holes were completely blocked, 109 partly obstructed and the remaining 65 holes were clear.

The stainless steel material used for the nozzles was in excellent condition and showed no sign of deterioration.

Gasifier Lid

The lid was heavily coated with carbon on its lower face and the protruding thermocouple had a considerable deposit of carbon around it. The refractory concrete was in good condition but the vermiculite and calcium silicate insulating slab was cracked in a number of places.

Gasifier Bed Material

The gasifier bed was slumped without sulphation at shut down and figure B-7 shows the typical bed material after half the material had been removed. There are two interesting features in this picture, one of them being the random area of completely white material which proved to be an area of fine particles free of carbon obviously formed after the bed was slumped which must then have had some oxygen in-leakage to burn off the carbon. This area of carbon free material existed in about the middle third of the bed depth. Another interesting and typical feature is the blackened area at the right hand side which corresponds to No.1 fuel injector location. After further bed removal this fuel injector was found to be encased with a hollow sphere about 10 cm (4") diameter of carbon and lime particles which would have restricted the throw of the injector.

The material was generally free from any large agglomerates apart from a few lumps of carbon and fine lime particles up to 3.8 cm (1½") across. There were some areas where there had been static zones in the fluid bed around the periphery of the distributor particularly near the fines return pipe from the elutriator shown in the right hand side of fig. B-8.

Regenerator

The regenerator material was free from agglomerates and the bore of the regenerator was generally clean apart from a few small deposits at the top outlet section and around the joint between the distributor and the wall. The refractory did not show any deterioration.

The distributor was in good condition with the nozzles clear and undamaged. The refractory lip around the distributor had cracked away during the run and the repair that was made had withstood the remainder of the run without any deterioration.

The transfer passages to and from the regenerator were free of any agglomerates and the refractory concrete around the transfer sections was in good condition.

Ductwork and Burner

Gasifier Outlet and Burner

The bifurcated duct had a uniform deposit of carbon on its inner surface about 1.6 mm (1/16") thick and at the junction between the two ducts there were thicker deposits where the two gas streams converged. The thermocouple in this area was heavily coated with a carbon and lime layer about 12 mm (½") thick.

The premix section upstream of the burner had a 1.6 mm (1/16") carbon layer on the stainless steel sections and the 3.2 mm (⅛") diameter stainless steel thermocouple in the burner throat was burnt away.

The main burner was in good shape with a layer of carbon about 1.6 mm (1/16") thick deposited on the internal gas ductwork. The outlet ring of the burner had a local build up mainly of lime on its outer face shown in fig. B-9 which was deposited from the turbulence in the gas streams arising from the stainless steel baffle plates located to shield the pilot flame from the main combustion air.

The pilot burner which had been very successfully modified to provide a forced gas and air premix system prior to the flame retention nozzle was in good condition apart from some local lime deposits around the nozzle end which protruded into the burner quarl in the boiler.

The stainless steel deflector plates in the boiler burner quarl were heavily deposited with lime.

Regenerator outlet

The outlet pipe from the regenerator top gas space was coated with long thin light purple deposits of material laid along in the direction of flow and projecting out from the wall of the duct (figure B-10). In one area there was one projection which extended out almost 25 mm (1") from the wall but generally they were 12 mm (½") or less and only about 3 mm (⅛") thick at the furthest tip. They were held fairly firmly to the wall and appeared to have passed through a molten phase having a fairly smooth outer face. Further downstream the pipe was coated with a much more uniform deposit about 3 mm (⅛") thick which was composed of fine particles rather than the liquid type of deposition in the hotter upstream section of the pipe (Figure B-11).

Downstream of the dust extraction cyclone the pipe was clean apart from a very thin white fine deposit.

Boiler and Stack

Boiler

The back end of the boiler had been cleaned periodically during the test run and figure B-12 shows the condition after shut down. There was some coarser material laying in the bottom of the boiler fire tube with some finer material deposited around the lower sides of the fire tube mainly on the left hand side when viewed from the rear, indicating that there was some swirl in the flame at the burner.

The entries to the first pass of fire tubes were deposited with rings of fine material although none of the tubes was totally blocked. Some of the deposits were smooth and rounded whilst others were spikey, shown clearly in figure B-13.

The return tube passes were generally clear apart from a few tubes which had some deposits at their ends. Some material had been deposited out of the gas stream and collected in the boiler space at the end of this second tube pass.

The refractory on the boiler rear door showed some signs of pitting but there has been a gradual deterioration not just associated with this particular run.

Boiler Probe

The boiler probe acquired some local light brown lumpy deposits which tended to be more concentrated around the "root" of tube figure B-14, in front of the entrance to the first tube pass. There was no indication of any deterioration of the tube which was cooled during the run to approximately 600°C.

Stack and Cyclone

There was a quantity of lime at the base of chimney which was otherwise clean and the external cyclone and knock out pot was also clear of any obstruction. Both these vessels were continually rapped during the run and this prevented any significant build up of material.

Flue Gas Scrubber and Recycle Line

Scrubber

There were considerable operational problems with the scrubber initially due to the wet fine slurry which was discharged and the deposits which built up at the scrubber entry. The problems were eased by running the scrubber with some recycle so reducing the dust burden concentration and increasing the velocity of the gas and hence effeciency of operation.

Examination afterwards showed that the knock out vessel had some hard fine deposits on its wall opposite the gas entry when the material would first impinge on the wall. The scrubber and its entry pipe was clear.

Recycle Line

The scrubber has cleaned up the solids content of the gas very well but introduced some problems due to the cold saturated gas leaving the scrubber which contained enough very fine particles to seize up the cycle line control butterfly valve and the control valve to the burner air valve which was in a static leg and probably contained a considerable quantity of condensation.

Figure B-1 R.H. Cyclone Inlet

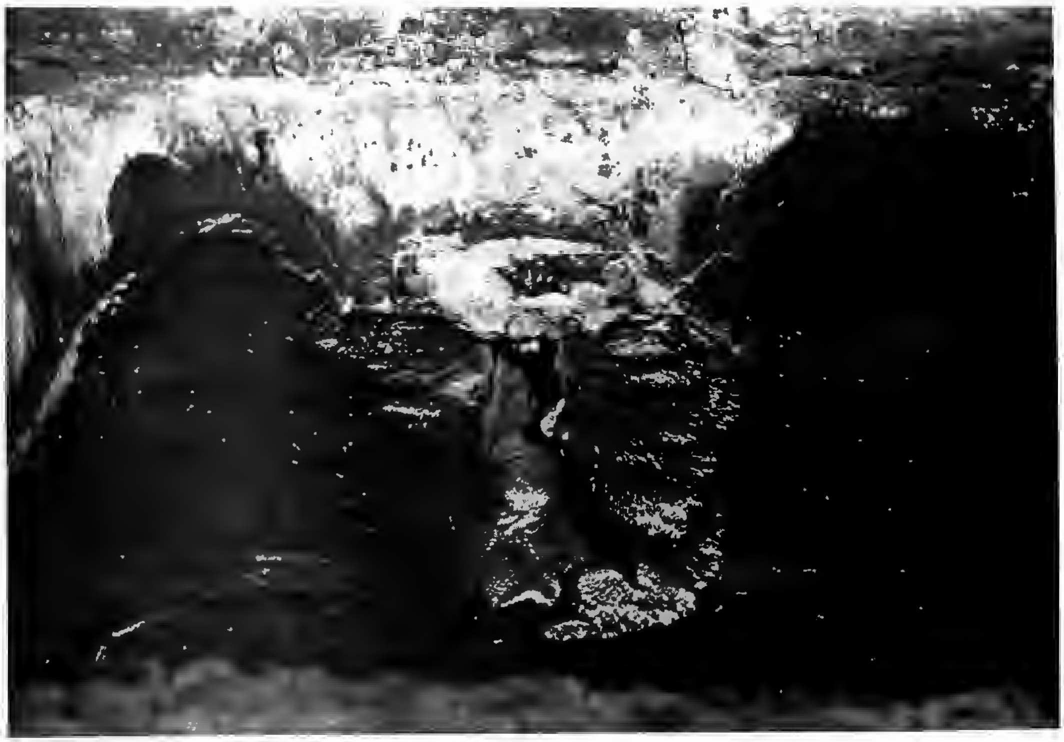

Figure B-2 L.H. Cyclone Inlet

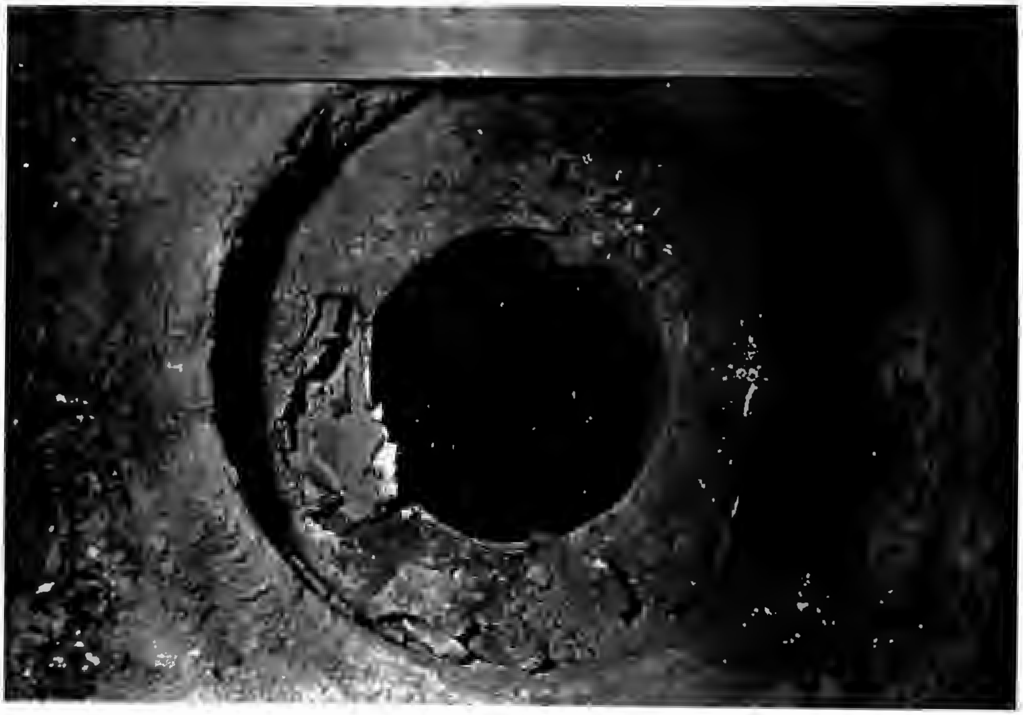

Figure B-3 R.H. Cyclone, Outlet Tube Removed

Figure B-4 R.H. Cyclone Outlet Tube

Figure B-5 -L.H. Cyclone, Outlet Tube Removed

Figure B-6 L.H. Cyclone Outlet Tube

Figure B-7 Gasifier Bed Half Removed

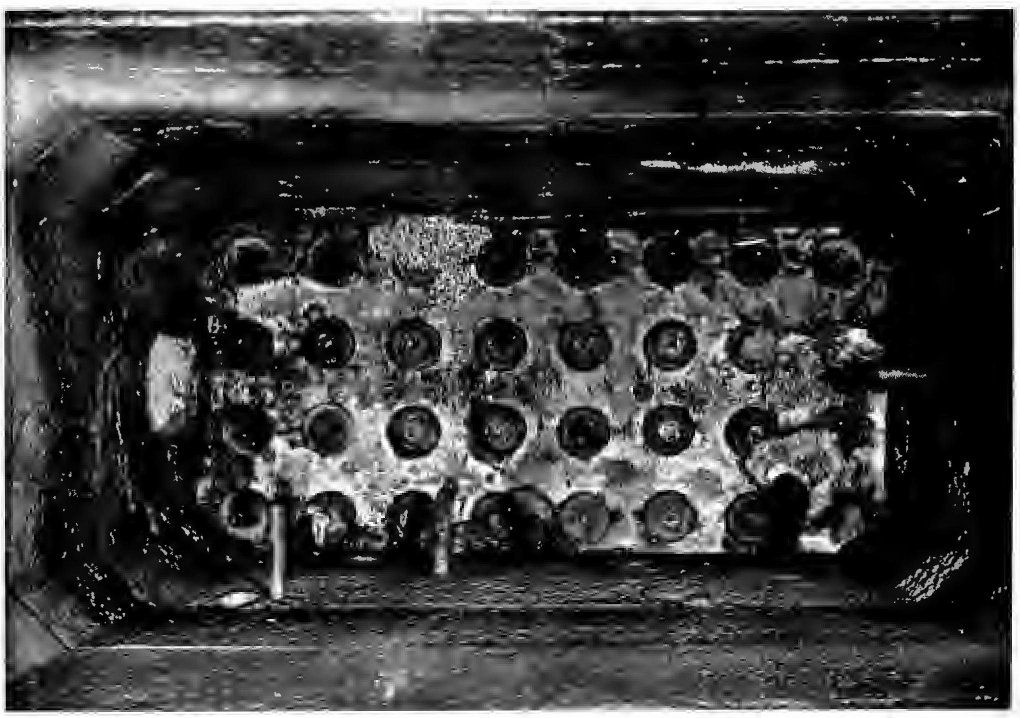

Figure B-8 Gasifier Bed Empty

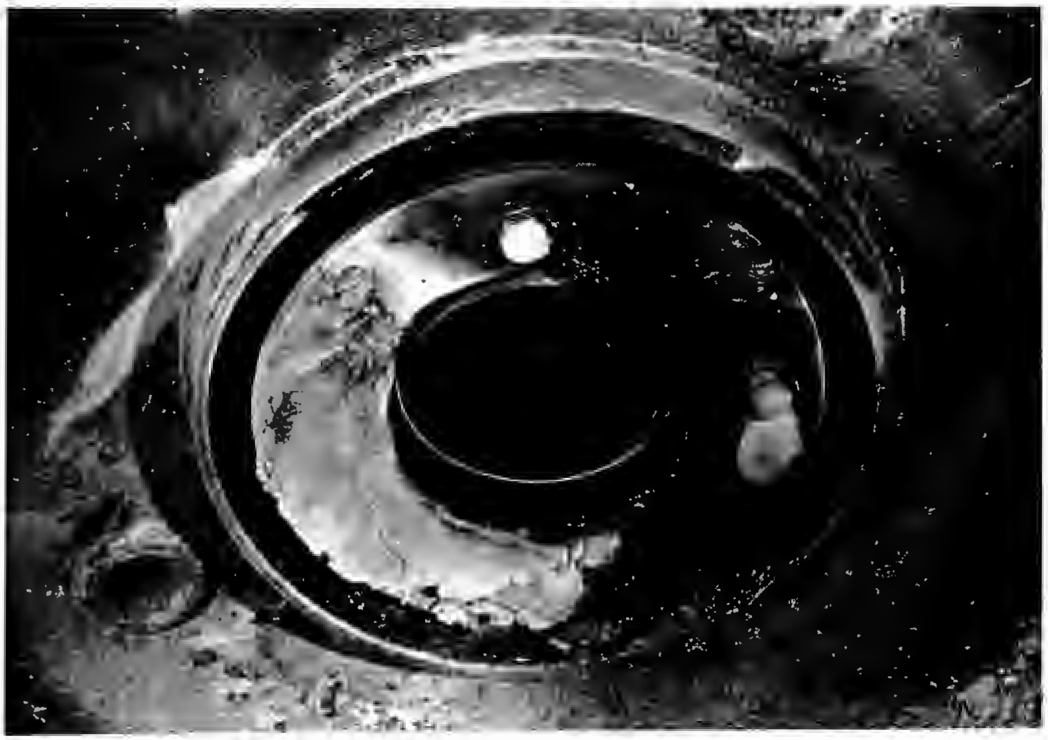

Figure B-9 Main Burner Outlet Ring

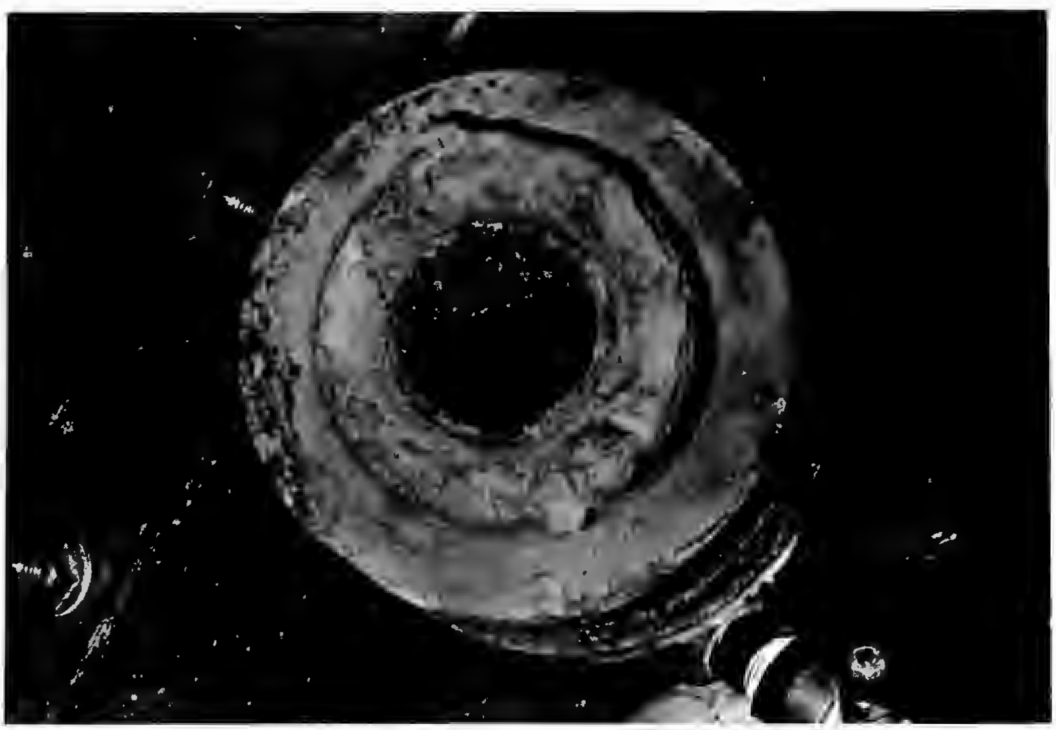

Figure B-10 Regenerator Outlet Pipe, Upstream End

Figure B-11 Regenerator Outlet Pipe, Downstream End

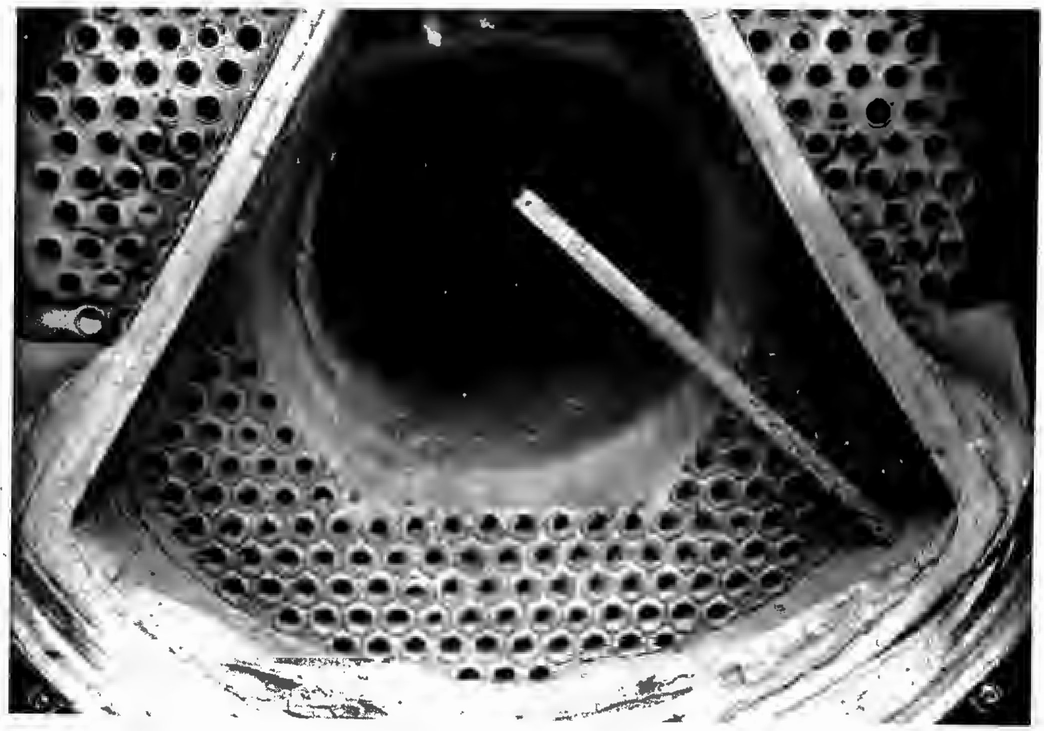

Figure B-12 Boiler Back End After Shut Down

Figure B-13 Fire Tube First Pass Inlets

Figure B-14 Boiler Probe

Fig. C.1 Gasifier cyclone inlets

Fig. C.2 L.H. cyclone inlet

APPENDIX B: TABLE I.

RUN 5: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
2.0130	933.	1067.	72.	182.6	34.5
2.0230	923.	1078.	72.	182.1	28.6
2.0330	939.	1069.	75.	175.1	10.0
2.0430	948.	1060.	76.	175.1	0.
2.0530	940.	1038.	76.	174.7	2.5
2.0630	940.	1053.	76.	174.7	2.5
2.0730	933.	1050.	77.	174.3	1.8
2.0830	MISSED DATA READING				
2.0930	940.	1060.	75.	177.2	3.6
2.1030	952.	1055.	75.	187.1	5.0
2.1130	882.	1030.	75.	187.1	5.9
2.1230	880.	1060.	72.	183.8	5.4
2.1330	885.	1050.	75.	182.1	6.4
2.1430	888.	1050.	75.	180.1	7.3
2.1530	882.	1050.	80.	183.0	7.7
2.1630	895.	1050.	80.	183.8	8.2
2.1730	892.	1050.	80.	180.9	7.7
2.1830	880.	1048.	75.	179.3	9.1
2.1930	880.	1048.	75.	179.3	9.1
2.2030	890.	1070.	80.	180.1	8.2
2.2130	882.	1023.	85.	180.1	8.2
2.2230	873.	1040.	95.	179.7	8.2
2.2330	880.	1048.	80.	180.5	9.1
3.0030	890.	1052.	78.	171.4	10.0
3.0130	MISSED DATA READING				
3.0230	882.	1061.	78.	182.1	9.1
3.0330	884.	1062.	80.	173.9	9.5
3.0430	889.	1065.	79.	176.8	8.6
3.0530	890.	1063.	80.	180.5	8.2
3.0630	888.	1057.	75.	181.7	7.3
3.0730	891.	1058.	75.	180.5	8.6
3.0830	899.	1060.	75.	181.3	8.6
3.0930	897.	1055.	70.	177.2	9.1
3.1030	891.	1041.	70.	179.3	6.8
3.1130	891.	1055.	70.	180.1	5.9
3.1230	892.	1055.	70.	177.6	5.9
3.1330	882.	1060.	70.	181.7	6.4
3.1430	882.	1070.	70.	179.7	7.5
3.1530	881.	1061.	70.	173.5	7.0
3.1630	886.	1065.	60.	173.5	6.4

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
3.1730	896.	1061.	70.	165.2	7.7
3.1830	890.	1062.	70.	190.0	9.5
3.1930	900.	1065.	75.	169.3	8.2
3.2030	895.	1068.	70.	198.3	8.6
3.2130	892.	1070.	80.	180.1	9.5
3.2230	902.	1065.	85.	179.7	5.9
3.2330	892.	1067.	83.	179.3	0.
4.0030	901.	1062.	83.	171.4	0.
		STONE CHANGE			
4.0130	876.	1062.	82.	182.6	0.
4.0230	831.	1055.	82.	182.1	0.
4.0330	888.	1060.	82.	173.9	0.
4.0430	910.	1065.	83.	176.8	0.
4.0530	850.	1067.	55.	180.5	45.8
4.0630	850.	1068.	65.	181.7	61.7
4.0730	853.	1066.	80.	180.5	49.9
4.0830	859.	1067.	80.	181.3	25.9
4.0930	870.	1050.	78.	177.2	8.6
4.1030	870.	1050.	78.	179.3	0.
4.1130	868.	1070.	78.	180.1	0.
4.1230	875.	1020.	78.	177.6	22.2
4.1330		MISSED DATA READING			
4.1430	861.	1050.	80.	177.2	28.6
4.1530	862.	1055.	80.	178.8	29.5
4.1630	852.	1049.	80.	176.4	28.1
4.1730	858.	1050.	80.	178.4	27.7
4.1830	860.	1050.	80.	178.0	35.4
4.1930	862.	1058.	80.	179.3	38.6
4.2030	860.	1052.	80.	180.1	35.4
4.2130	862.	1051.	86.	177.6	26.8
4.2230	865.	1053.	85.	177.6	27.2
4.2330	873.	1052.	60.	173.5	28.6
5.0030	880.	1055.	75.	186.7	33.1
5.0130		MISSED DATA READING			
5.0230	867.	1059.	80.	168.9	28.6
5.0330	848.	1050.	82.	174.7	29.0
5.0430	858.	1058.	80.	187.9	30.4
5.0530	858.	1052.	85.	180.9	29.0
5.0630	858.	1058.	82.	180.9	28.6
5.0730	859.	1058.	80.	180.1	22.7

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
5.0830	858.	1052.	80.	180.9	29.5
5.0930	845.	1050.	80.	180.9	38.1
5.1030	845.	1050.	80.	180.1	30.6
5.1130	851.	1050.	80.	180.1	29.0
5.1230	848.	1049.	80.	179.7	19.1
5.1330	MISSED DATA READING				
5.1430	870.	1055.	80.	175.1	13.6
5.1530	872.	1055.	80.	174.3	13.2
5.1630	888.	1054.	80.	174.3	15.9
SHUT DOWN AT 5.1630 FOR 10 HOURS					
6.0230	865.	1049.	85.	171.8	14.5
6.0330	862.	1050.	83.	171.8	22.2
6.0430	852.	1054.	80.	173.5	21.3
6.0530	868.	1050.	85.	172.2	20.9
6.0630	852.	1060.	83.	172.6	18.1
6.0730	850.	1055.	83.	172.2	17.7
6.0830	850.	1065.	82.	166.0	18.6
SHUT DOWN AT 6.0830 FOR 62 HOURS					
8.2230	895.	1005.	70.	172.2	24.5
8.2330	890.	1035.	70.	171.8	24.9
9.0030	872.	1062.	70.	173.1	32.2
9.0130	870.	1042.	70.	177.6	33.6
9.0230	875.	1042.	70.	177.2	29.9
9.0330	872.	1048.	72.	177.6	26.8
9.0430	880.	1060.	72.	176.8	19.5
9.0530	882.	1055.	72.	177.6	15.4
SHUT DOWN AT 9.0530 FOR 57 HOURS					
11.1430	871.	1060.	80.	177.2	39.5
11.1530	860.	1060.	80.	185.9	37.6
11.1630	860.	1062.	80.	177.6	15.9
11.1730	866.	1064.	80.	177.6	15.9

APPENDIX B: TABLE I.

RUN 5: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
11.1830	879.	1069.	74.	177.2	17.2
11.1930	882.	1068.	80.	177.2	17.2
11.2030	892.	1075.	80.	177.6	17.7
11.2130	892.	1067.	88.	176.8	20.0
11.2230	880.	1068.	85.	179.7	19.1
11.2330	867.	1068.	80.	179.7	20.9
12.0030	869.	1065.	83.	183.8	24.0
12.0130	MISSED DATA READING				
12.0230	875.	1065.	82.	184.6	17.7
12.0330	881.	1065.	80.	183.8	12.7
12.0430	880.	1050.	80.	190.0	10.9
12.0530	881.	1050.	80.	183.8	14.1
12.0630	879.	1050.	80.	183.8	15.4
12.0730	875.	1050.	75.	177.6	10.9
12.0830	872.	1050.	75.	173.5	15.4
12.0930	868.	1042.	75.	174.7	21.8
12.1030	865.	1042.	80.	178.0	20.4
12.1130	MISSED DATA READING				
12.1230	875.	1030.	80.	178.4	19.1
12.1330	868.	1040.	80.	178.8	18.6
12.1430	868.	1035.	80.	178.8	18.8
12.1530	870.	1042.	75.	178.4	21.3
12.1630	868.	1071.	83.	179.3	20.4
12.1730	865.	1049.	82.	178.4	15.4
12.1830	876.	1071.	85.	177.6	17.7
12.1930	878.	1028.	85.	178.0	19.1
12.2030	874.	1030.	85.	178.0	17.2
12.2130	871.	1012.	81.	178.0	18.8
12.2230	878.	1075.	80.	182.1	17.9
12.2330	875.	1020.	75.	179.7	15.9
13.0030	865.	1020.	75.	177.2	19.1
13.0130	862.	1025.	75.	175.9	18.1
13.0230	870.	1060.	75.	176.8	15.4
13.0330	875.	1020.	80.	177.2	15.4
13.0430	880.	1020.	80.	176.8	15.6
13.0530	880.	1035.	75.	176.4	12.7
13.0630	870.	1020.	75.	174.7	15.0

APPENDIX B: TABLE I.

RUN 5: TEMPERATURES AND FEED RATES PAGE 5 OF 9

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
SHUT DOWN AT 13.0630 FOR 73 HOURS					
16.0730	868.	990.	80.	174.7	17.7
16.0830	875.	1070.	81.	174.7	25.4
16.0930	878.	1068.	81.	175.5	24.9
16.1030	868.	1070.	80.	175.1	22.2
16.1130	881.	1029.	80.	174.7	20.4
16.1230	866.	1050.	79.	175.5	27.7
16.1330	869.	1051.	72.	173.5	28.6
16.1430	MISSED DATA READING				
16.1530	870.	1060.	72.	171.8	27.2
$16.1630	861.	1060.	71.	172.2	29.5
16.1730	865.	1060.	71.	172.2	31.8
16.1830	872.	1061.	71.	172.2	34.9
16.1930	865.	1060.	70.	172.2	34.0
16.2030	859.	1055.	70.	171.4	33.1
16.2130	855.	975.	70.	181.7	33.1
16.2230	848.	1012.	72.	181.7	33.6
16.2330	862.	1050.	72.	178.8	15.4
17.0030	861.	1061.	72.	179.3	16.3
17.0130	870.	1052.	71.	180.1	15.0
17.0230	872.	1039.	70.	179.3	15.4
17.0330	872.	1031.	70.	180.1	18.1
17.0430	872.	1038.	70.	188.8	13.6
17.0530	876.	1039.	62.	171.0	20.4
17.0630	870.	1020.	70.	180.1	18.1
17.0730	869.	1059.	70.	179.3	18.6
17.0830	MISSED DATA READING				
17.0930	868.	1053.	70.	180.1	6.9
17.1030	861.	1052.	70.	179.3	20.9
17.1130	860.	1051.	70.	180.1	17.1
17.1230	861.	1050.	70.	180.9	16.8
17.1330	862.	1048.	70.	180.1	20.0
17.1430	862.	1058.	70.	185.9	22.7
17.1530	865.	1041.	70.	178.8	31.3
17.1630	858.	1041.	70.	179.7	30.8
17.1730	860.	1049.	70.	179.3	22.2
17.1830	858.	1050.	70.	179.3	19.5

APPENDIX B: TABLE I.

RUN 5: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
17.1930	869.	1051.	65.	179.3	18.1
17.2030	862.	1050.	68.	179.3	16.3
		STONE CHANGE			

SHUT DOWN AT 17.2030 FOR 72 HOURS

DAY.HOUR	GASIFIER	REGEN.	RECYCLE	OIL	STONE
20.2030	918.	1012.	65.	193.3	39.9
20.2130	900.	1012.	68.	197.8	23.1
20.2230	888.	1055.	70.	205.3	26.8
20.2330	890.	1052.	70.	208.6	27.2
21.0030	915.	1055.	65.	208.6	25.9
21.0130	915.	1053.	65.	196.2	24.0
21.0230	915.	1060.	60.	211.1	25.4
21.0330	910.	1058.	65.	215.6	19.1
21.0430	905.	1060.	65.	212.7	21.8
21.0530	902.	1060.	65.	212.7	21.8
21.0630	914.	1060.	65.	211.9	19.1
21.0730	900.	1060.	65.	211.5	20.4
21.0830	898.	1060.	62.	215.6	20.4
21.0930	901.	1067.	65.	207.3	17.2
21.1030	900.	1064.	65.	211.5	19.5
21.1130	899.	1061.	65.	211.1	20.4
21.1230	900.	1062.	66.	211.5	20.9
21.1330	903.	1065.	63.	211.5	23.1
21.1430	901.	1068.	62.	211.1	21.8
21.1530	889.	1064.	64.	213.1	23.6
21.1630	895.	1063.	63.	213.9	23.1
21.1730	898.	1066.	62.	213.1	23.6
21.1830	875.	1066.	66.	210.2	23.1
21.1930	893.	1069.	60.	212.3	24.9
21.2030	892.	1069.	70.	196.6	15.9
21.2130	860.	1069.	70.	184.6	10.4
21.2230	856.	1068.	70.	188.3	13.6
21.2330	870.	1071.	70.	187.9	18.1
22.0030	871.	1068.	70.	185.4	21.8
22.0130	871.	1070.	70.	187.9	20.9
22.0230	878.	1068.	70.	184.2	14.1
22.0330	872.	1069.	70.	185.9	16.8
22.0430	874.	1070.	70.	190.0	13.6

APPENDIX B: TABLE I.

RUN 5: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
22.0530	875.	1069.	70.	191.2	13.6
22.0630	875.	1070.	70.	183.0	14.5
22.0730	888.	1070.	70.	181.7	14.5
22.0830	875.	1070.	70.	190.8	16.3
22.0930	872.	1071.	72.	186.3	16.3
22.1030	872.	1069.	72.	185.4	14.5
22.1130	871.	1064.	72.	185.0	13.2
22.1230	875.	1065.	72.	185.4	14.5
22.1330	873.	1069.	72.	189.6	15.4
22.1430	875.	1070.	72.	184.6	13.2
22.1530	872.	1069.	71.	184.6	13.6
22.1630	871.	1069.	72.	184.6	16.3
22.1730	877.	1071.	71.	185.4	13.6
22.1830	873.	1068.	70.	184.2	15.9

SHUT DOWN AT 22.1830 FOR 23 HOURS

23.1730	891.	1053.	75.	177.6	32.2
23.1830	869.	1055.	75.	182.6	33.1
23.1930	868.	1052.	75.	178.4	33.1
23.2030	869.	1061.	74.	178.0	30.4
23.2130	860.	1061.	75.	179.7	30.4
23.2230	858.	1060.	75.	178.8	31.8
23.2330	866.	1060.	75.	178.4	31.8
24.0030	862.	1060.	75.	179.7	31.8
24.0130	869.	1060.	75.	181.7	31.8
24.0230	862.	1060.	75.	183.0	32.2
24.0330	865.	1060.	75.	184.6	29.9
24.0430	862.	1055.	75.	184.6	29.0
24.0530	860.	1054.	74.	185.4	30.4
24.0630	861.	1059.	75.	185.0	28.6
24.0730	865.	1061.	74.	186.3	29.0
24.0830	878.	1061.	72.	185.4	35.4
24.0930	873.	1061.	72.	185.9	35.4
24.1030	873.	1061.	70.	184.6	31.3
24.1130	885.	1060.	72.	181.3	32.2
24.1230	879.	1061.	71.	184.6	27.2
24.1330	875.	1061.	72.	185.0	9.5
24.1430	870.	1060.	70.	184.6	13.6

APPENDIX B: TABLE I.

RUN 5: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
24.1530	870.	1060.	70.	185.4	17.2
24.1630	869.	1060.	75.	184.6	18.1
24.1730	869.	1059.	72.	184.2	21.3
24.1830	871.	1060.	75.	185.0	20.9
24.1930	865.	1060.	75.	183.8	23.6
24.2030	870.	1061.	75.	183.8	21.8
24.2130	876.	1060.	71.	183.4	21.3
24.2230	875.	1060.	70.	183.4	23.6
24.2330	883.	1060.	70.	183.4	20.4
25.0030	870.	1060.	70.	183.8	23.6
25.0130	878.	1062.	70.	183.4	24.5
25.0230	874.	1060.	70.	183.4	22.7
25.0330	880.	1060.	70.	183.4	23.6
25.0430	880.	1060.	70.	182.1	18.1
25.0530	880.	1060.	70.	182.6	20.4
25.0630	880.	1060.	70.	182.6	22.2
25.0730	880.	1060.	70.	181.3	19.1
25.0830	878.	1060.	70.	181.3	22.2
25.0930	880.	1060.	75.	182.1	20.4
25.1030	883.	1061.	75.	183.0	22.2
25.1130	890.	1060.	70.	181.7	17.7
25.1230	870.	1060.	73.	182.1	21.3
25.1330	878.	1059.	75.	181.7	20.4
25.1430	878.	1060.	75.	183.0	19.1
25.1530	873.	1060.	75.	182.1	16.8
25.1630	878.	1060.	75.	180.9	19.5
25.1730	871.	1060.	75.	186.3	13.6
25.1830	881.	1061.	75.	184.6	15.0
25.1930	879.	1061.	74.	185.4	14.5
25.2030	883.	1060.	74.	185.0	16.3
25.2130	882.	1060.	70.	185.9	14.1
25.2230	881.	1060.	70.	185.4	16.3
25.2330	876.	1058.	70.	184.6	16.3
26.0030	871.	1060.	70.	184.6	16.3
26.0130	872.	1060.	70.	185.0	14.5
26.0230	870.	1060.	70.	185.4	15.9
26.0330	878.	1060.	70.	184.6	14.5
26.0430	888.	1060.	70.	185.0	15.4
26.0530	885.	1060.	70.	185.0	14.5
26.0630	886.	1060.	70.	185.0	12.2

APPENDIX B: TABLE I.

RUN 5: TEMPERATURES AND FEED RATES PAGE 9 OF 9

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
26.0730	866.	1060.	70.	185.0	19.5
26.0830	870.	1060.	70.	185.0	20.0
26.0930	872.	1060.	70.	184.6	17.2
26.1030	862.	1060.	70.	185.9	19.5
26.1130	860.	1060.	70.	185.0	20.0
26.1230	864.	1060.	70.	185.4	20.4
26.1330	868.	1060.	74.	184.6	19.1
26.1430	865.	1062.	74.	185.0	20.4
26.1530	868.	1060.	73.	184.6	23.6
26.1630	870.	1060.	74.	185.4	19.5
26.1730	877.	1062.	74.	184.2	19.1
26.1830	872.	1059.	72.	184.2	19.5

APPENDIX B: TABLE II.
RUN 5: GAS FLOW RATES

DAY.HOUR	GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
	G A S		R A T E S	M3/HR		
2.0130	445.	149.	3.3	30.2	3.3	1.54
2.0230	445.	150.	3.3	31.7	3.3	1.63
2.0330	402.	201.	3.1	23.9	3.3	1.25
2.0430	419.	191.	3.1	25.6	3.3	1.32
2.0530	419.	191.	3.2	28.5	3.3	1.42
2.0630	419.	181.	3.2	25.1	3.3	1.29
2.0730	431.	181.	3.2	22.7	3.3	1.18
2.0830		MISSED DATA READING				
2.0930	456.	159.	3.2	25.2	3.3	1.30
2.1030	421.	210.	3.2	28.0	3.3	1.42
2.1130	404.	352.	3.2	30.3	3.3	1.50
2.1230	396.	229.	3.3	31.2	3.3	1.58
2.1330	404.	220.	3.3	29.7	3.3	1.50
2.1430	387.	220.	3.3	30.4	3.3	1.53
2.1530	397.	212.	3.3	30.5	3.3	1.53
2.1630	397.	212.	3.3	29.9	3.3	1.51
2.1730	388.	202.	3.3	29.6	3.3	1.49
2.1830	362.	179.	3.3	26.4	3.3	1.35
2.1930	362.	179.	3.3	26.4	3.3	1.35
2.2030	379.	164.	3.3	27.7	3.3	1.44
2.2130	380.	208.	3.3	22.5	3.3	1.15
2.2230	362.	193.	3.3	30.1	3.3	1.51
2.2330	380.	167.	3.3	32.6	3.3	1.64
3.0030	380.	162.	3.3	30.7	3.3	1.55
3.0130		MISSED DATA READING				
3.0230	380.	156.	3.3	31.7	3.3	1.61
3.0330	380.	152.	3.3	32.2	3.3	1.63
3.0430	380.	152.	3.3	33.4	3.3	1.70
3.0530	379.	148.	3.3	33.4	3.3	1.69
3.0630	388.	140.	3.3	33.5	3.3	1.69
3.0730	388.	140.	3.3	33.1	3.3	1.67
3.0830	388.	140.	3.3	32.9	3.3	1.66
3.0930	397.	135.	3.3	32.2	3.3	1.62
3.1030	379.	160.	3.3	30.8	3.3	1.54
3.1130	397.	177.	3.3	30.8	3.3	1.56
3.1230	397.	171.	3.3	30.3	3.3	1.53
3.1330	379.	177.	3.3	33.3	3.3	1.67
3.1430	379.	189.	3.3	30.7	3.3	1.57
3.1530	379.	189.	3.3	30.6	3.3	1.55
3.1630	379.	177.	3.1	30.4	3.3	1.55

APPENDIX B: TABLE II.

RUN 5: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
3.1730	379.	189.	3.3	25.0	3.3	1.29
3.1830	388.	180.	2.9	31.1	3.3	1.57
3.1930	388.	181.	3.2	28.7	3.3	1.47
3.2030	379.	180.	3.3	32.4	3.3	1.64
3.2130	384.	181.	3.3	22.8	3.3	1.20
3.2230	388.	186.	3.3	30.4	3.3	1.55
3.2330	381.	185.	3.3	33.0	3.3	1.67
4.0030	382.	185.	3.3	31.0	3.3	1.57
		STONE CHANGE				
4.0130	379.	158.	3.3	30.5	3.3	1.55
4.0230	380.	158.	3.3	32.1	3.3	1.61
4.0330	381.	154.	3.3	32.6	3.3	1.63
4.0430	380.	154.	3.3	33.8	3.3	1.70
4.0530	356.	189.	3.3	33.8	3.3	1.70
4.0630	382.	159.	3.3	33.8	3.3	1.70
4.0730	382.	153.	3.3	33.5	3.3	1.67
4.0830	382.	173.	3.3	33.3	3.3	1.67
4.0930	380.	162.	3.3	32.5	3.3	1.60
4.1030	363.	171.	3.3	31.1	3.3	1.54
4.1130	363.	190.	3.3	31.1	3.3	1.56
4.1230	363.	190.	3.3	30.7	3.3	1.49
4.1330		MISSED DATA READING				
4.1430	385.	205.	3.2	28.0	3.3	1.41
4.1530	398.	199.	3.2	30.4	3.3	1.52
4.1630	395.	205.	3.2	28.2	3.3	1.41
4.1730	395.	194.	3.3	30.5	3.3	1.52
4.1830	394.	194.	3.2	31.1	3.3	1.54
4.1930	394.	194.	3.2	27.3	3.3	1.38
4.2030	395.	200.	3.2	30.9	3.3	1.54
4.2130	394.	210.	3.5	30.7	3.3	1.53
4.2230	395.	208.	3.5	30.0	3.3	1.50
4.2330	361.	218.	3.5	31.1	3.3	1.55
5.0030	378.	189.	3.5	31.6	3.9	1.60
5.0130		MISSED DATA READING				
5.0230	378.	193.	3.5	31.8	4.6	1.65
5.0330	378.	195.	3.4	31.0	1.9	1.48
5.0430	378.	194.	3.5	31.0	3.7	1.57
5.0530	395.	197.	3.5	30.8	2.9	1.52
5.0630	395.	195.	3.5	30.8	0.7	1.42
5.0730	395.	193.	3.5	30.8	2.1	1.49

APPENDIX B: TABLE II.
RUN 5: GAS FLOW RATES PAGE 3 OF 9

DAY.HOUR	GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
	G A S R A T E S M3/HR					
5.0830	395.	193.	3.5	31.0	1.6	1.47
5.0930	393.	193.	3.3	30.9	2.3	1.49
5.1030	393.	193.	3.3	31.0	1.6	1.46
5.1130	385.	163.	3.2	30.9	2.1	1.49
5.1230	367.	173.	3.4	31.2	1.6	1.48
5.1330	MISSED DATA READING					
5.1430	378.	172.	3.4	30.2	2.0	1.45
5.1530	387.	168.	3.4	29.8	2.1	1.44
5.1630	378.	162.	3.4	27.5	1.5	1.31
SHUT DOWN AT 5.1630 FOR 10 HOURS						
6.0230	363.	218.	3.5	29.0	2.9	1.42
6.0330	363.	216.	3.5	28.0	3.7	1.41
6.0430	363.	203.	3.5	28.8	3.3	1.43
6.0530	362.	208.	3.5	25.2	1.7	1.20
6.0630	362.	185.	3.5	25.3	1.7	1.21
6.0730	380.	175.	3.5	26.8	1.6	1.26
6.0830	328.	174.	3.5	27.4	1.6	1.30
SHUT DOWN AT 6.0830 FOR 62 HOURS						
8.2230	388.	189.	3.3	36.1	2.8	1.72
8.2330	380.	189.	3.3	35.9	3.3	1.77
9.0030	398.	179.	3.3	38.1	3.3	1.91
9.0130	380.	173.	3.3	38.4	2.6	1.86
9.0230	389.	169.	3.3	39.8	3.6	1.97
9.0330	397.	169.	3.3	39.9	4.4	2.03
9.0430	397.	169.	3.3	43.5	4.8	2.23
9.0530	397.	159.	3.3	44.4	4.8	2.26
SHUT DOWN AT 9.0530 FOR 57 HOURS						
11.1430	387.	205.	3.5	34.4	6.7	1.88
11.1530	386.	174.	3.5	35.9	7.2	1.98
11.1630	386.	173.	3.3	36.7	6.2	1.97
11.1730	386.	173.	3.3	36.7	6.2	1.97

APPENDIX B: TABLE II.
RUN 5: GAS FLOW RATES PAGE 4 OF 9

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
11.1830	387.	171.	3.3	36.4	7.6	2.03
11.1930	404.	160.	3.3	36.1	6.7	1.97
11.2030	405.	149.	3.3	36.1	7.0	1.99
11.2130	404.	157.	3.3	37.9	7.0	2.07
11.2230	387.	176.	3.3	35.4	5.3	1.87
11.2330	387.	172.	3.3	31.6	5.6	1.71
12.0030	400.	174.	3.3	34.5	5.6	1.84
12.0130	MISSED DATA READING					
12.0230	396.	163.	3.2	33.9	7.7	1.91
12.0330	396.	162.	3.2	31.0	7.5	1.77
12.0430	396.	162.	3.2	38.0	7.8	2.08
12.0530	396.	162.	3.2	34.6	8.8	1.97
12.0630	379.	182.	3.2	34.8	7.6	1.93
12.0730	362.	190.	3.2	38.2	7.5	2.08
12.0830	362.	190.	3.2	35.1	7.0	1.91
12.0930	379.	160.	3.2	35.4	8.3	1.98
12.1030	380.	152.	3.2	36.1	7.2	1.96
12.1130	MISSED DATA READING					
12.1230	379.	152.	3.3	35.9	8.6	1.99
12.1330	379.	152.	3.3	35.7	7.1	1.94
12.1430	379.	152.	3.3	35.7	8.2	1.98
12.1530	378.	140.	3.3	36.4	6.0	1.92
12.1630	379.	143.	3.3	35.8	5.6	1.92
12.1730	379.	143.	3.3	35.7	4.9	1.85
12.1830	379.	144.	3.3	35.1	1.5	1.69
12.1930	379.	144.	3.3	35.8	4.7	1.81
12.2030	379.	155.	3.3	35.8	4.7	1.81
12.2130	379.	173.	3.3	35.8	4.7	1.79
12.2230	379.	191.	3.3	37.6	5.0	1.97
12.2330	362.	210.	3.3	36.9	7.5	1.97
13.0030	362.	209.	3.3	35.7	4.5	1.78
13.0130	361.	209.	3.3	35.7	6.8	1.89
13.0230	361.	209.	3.3	36.1	6.8	1.96
13.0330	361.	212.	1.7	36.9	7.2	1.95
13.0430	362.	212.	3.3	36.4	7.2	1.93
13.0530	362.	209.	3.3	36.8	7.1	1.97
13.0630	361.	210.	3.3	37.4	7.2	1.98

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	FLUE GAS	PILOT PROPANE	REGENERATOR AIR	NITROGEN	REGEN. VELOCITY M/SEC
SHUT DOWN AT 13.0630 FOR 73 HOURS						
16.0730	391.	0.	3.4	31.9	7.7	1.71
16.0830	387.	19.	3.4	31.1	7.7	1.79
16.0930	387.	193.	3.4	31.1	4.3	1.63
16.1030	387.	193.	3.4	30.5	4.6	1.62
16.1130	387.	186.	3.4	30.8	5.2	1.61
16.1230	387.	192.	3.4	34.2	6.2	1.84
16.1330	404.	179.	3.3	33.0	4.2	1.69
16.1430	MISSED DATA READING					
16.1530	404.	169.	3.3	32.7	4.4	1.70
16.1630	395.	148.	3.3	31.1	4.1	1.61
16.1730	404.	148.	3.3	31.7	4.7	1.67
16.1830	404.	158.	3.3	33.3	5.9	1.80
16.1930	395.	148.	3.2	33.0	6.0	1.78
16.2030	387.	148.	3.2	33.1	6.5	1.80
16.2130	395.	148.	3.4	31.6	6.2	1.62
16.2230	395.	149.	3.4	31.7	6.2	1.67
16.2330	396.	149.	3.4	33.1	6.3	1.79
17.0030	404.	149.	3.3	33.4	7.2	1.86
17.0130	405.	148.	3.3	33.0	4.9	1.72
17.0230	405.	148.	3.4	35.0	4.9	1.79
17.0330	405.	148.	3.3	33.3	4.7	1.70
17.0430	405.	148.	3.3	35.0	4.9	1.79
17.0530	405.	147.	3.3	59.2	4.9	2.87
17.0630	405.	148.	3.3	33.4	5.0	1.70
17.0730	405.	148.	3.3	33.4	4.9	1.75
17.0830	MISSED DATA READING					
17.0930	405.	147.	3.3	33.6	4.2	1.72
17.1030	405.	147.	3.4	33.2	3.9	1.68
17.1130	388.	147.	3.3	33.3	4.2	1.70
17.1230	405.	127.	3.4	32.7	2.9	1.61
17.1330	395.	127.	3.3	33.3	5.7	1.77
17.1430	421.	137.	3.3	32.7	3.8	1.66
17.1530	421.	137.	3.3	33.6	4.2	1.70
17.1630	403.	137.	3.2	35.2	4.2	1.77
17.1730	421.	127.	3.3	34.9	4.2	1.77
17.1830	412.	127.	3.3	34.0	4.2	1.73

APPENDIX B: TABLE II.
RUN 5: GAS FLOW RATES PAGE 6 OF 9

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
17.1930	403.	127.	3.3	34.3	3.9	1.73
17.2030	412.	127.	3.3	34.6	3.9	1.74
		STONE CHANGE				
SHUT DOWN AT 17.2030 FOR 72 HOURS						
20.2030	460.	85.	3.2	34.1	6.1	1.75
20.2130	443.	105.	3.2	36.3	4.9	1.79
20.2230	442.	105.	3.2	37.5	2.5	1.79
20.2330	460.	105.	3.2	37.1	2.1	1.75
21.0030	468.	65.	3.2	36.6	2.3	1.74
21.0130	467.	65.	3.2	33.9	2.1	1.61
21.0230	476.	52.	3.2	35.2	2.1	1.68
21.0330	459.	74.	3.2	35.8	2.1	1.70
21.0430	458.	70.	3.2	34.9	2.3	1.67
21.0530	458.	71.	3.2	34.9	2.2	1.67
21.0630	459.	64.	3.2	34.5	2.0	1.64
21.0730	441.	62.	3.2	34.2	2.0	1.62
21.0830	441.	59.	3.2	34.2	1.9	1.62
21.0930	441.	61.	3.3	34.1	1.9	1.62
21.1030	442.	61.	3.4	33.3	1.9	1.58
21.1130	435.	59.	3.4	33.2	1.9	1.57
21.1230	434.	59.	3.4	33.4	1.8	1.58
21.1330	436.	59.	3.4	33.4	2.0	1.59
21.1430	436.	57.	3.4	33.4	2.0	1.59
21.1530	436.	57.	3.4	33.2	1.9	1.58
21.1630	436.	56.	3.4	32.8	2.0	1.56
21.1730	436.	42.	3.4	33.4	2.0	1.58
21.1830	427.	96.	3.4	31.2	1.9	1.48
21.1930	435.	13.	3.4	32.0	1.9	1.52
21.2030	384.	136.	3.4	33.3	1.9	1.59
21.2130	367.	165.	3.4	34.7	2.0	1.66
21.2230	366.	165.	3.4	35.3	2.1	1.69
21.2330	382.	145.	3.4	34.7	2.2	1.67
22.0030	382.	144.	3.4	33.4	2.3	1.61
22.0130	382.	144.	3.3	34.0	2.3	1.64
22.0230	383.	144.	3.3	34.0	1.0	1.58
22.0330	383.	144.	3.4	33.7	1.7	1.59
22.0430	383.	144.	3.4	33.8	4.0	1.70

APPENDIX B: TABLE II.

RUN 5: GAS FLOW RATES

DAY.HOUR	GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
	GAS RATES M3/HR					
22.0530	383.	144.	3.4	34.2	2.3	1.64
22.0630	374.	144.	3.3	33.4	2.2	1.60
22.0730	383.	134.	3.3	33.2	2.2	1.59
22.0830	382.	148.	3.3	33.7	2.3	1.62
22.0930	382.	145.	3.4	33.1	2.4	1.60
22.1030	383.	145.	3.4	32.3	1.9	1.54
22.1130	383.	145.	3.4	33.3	2.2	1.59
22.1230	383.	145.	3.4	28.8	2.2	1.39
22.1330	382.	145.	3.4	32.7	2.2	1.57
22.1430	383.	145.	3.4	33.1	2.2	1.59
22.1530	384.	144.	3.4	33.4	2.2	1.60
22.1630	384.	145.	3.4	33.3	2.2	1.59
22.1730	384.	144.	3.4	33.8	2.2	1.62
22.1830	384.	144.	3.4	32.7	2.2	1.56
SHUT DOWN AT 22.1830 FOR 23 HOURS						
23.1730	329.	185.	3.5	34.6	2.5	1.67
23.1830	346.	185.	3.5	34.9	2.6	1.69
23.1930	346.	185.	3.5	34.9	2.3	1.67
23.2030	363.	185.	4.1	34.4	2.3	1.66
23.2130	329.	175.	3.5	33.9	2.3	1.64
23.2230	346.	179.	3.5	33.4	2.3	1.62
23.2330	347.	179.	3.5	32.5	2.3	1.57
24.0030	373.	175.	3.5	32.8	2.3	1.59
24.0130	373.	175.	3.5	33.9	2.3	1.63
24.0230	373.	165.	3.5	33.3	2.3	1.61
24.0330	373.	165.	3.5	32.7	2.3	1.58
24.0430	373.	165.	3.5	33.2	2.3	1.60
24.0530	381.	155.	3.5	33.9	2.3	1.63
24.0630	381.	155.	3.5	33.9	2.3	1.63
24.0730	355.	155.	3.5	33.4	2.3	1.61
24.0830	381.	135.	3.5	32.8	2.3	1.58
24.0930	389.	145.	3.5	32.5	2.5	1.57
24.1030	397.	144.	3.5	32.9	2.3	1.58
24.1130	379.	145.	3.5	33.6	2.5	1.62
24.1230	363.	154.	3.5	33.8	2.3	1.63
24.1330	373.	164.	3.5	34.1	2.3	1.65
24.1430	366.	164.	3.5	34.5	2.4	1.66

APPENDIX B: TABLE II.
RUN 5: GAS FLOW RATES PAGE 8 OF 9

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	FLUE GAS	PILOT PROPANE	REGENERATOR AIR	NITROGEN	REGEN. VELOCITY M/SEC
24.1530	366.	164.	3.5	34.2	2.4	1.65
24.1630	366.	165.	3.5	34.7	2.4	1.67
24.1730	383.	164.	3.5	34.2	2.4	1.65
24.1830	365.	155.	3.5	33.9	2.4	1.64
24.1930	365.	165.	3.5	33.8	2.6	1.64
24.2030	365.	165.	3.6	35.2	2.3	1.69
24.2130	366.	164.	3.6	34.8	2.3	1.67
24.2230	366.	164.	3.4	35.3	2.4	1.70
24.2330	365.	164.	3.4	35.2	2.4	1.69
25.0030	366.	164.	3.4	38.2	2.4	1.83
25.0130	366.	164.	3.2	35.5	2.4	1.71
25.0230	366.	164.	3.2	35.1	2.4	1.69
25.0330	366.	164.	3.2	35.3	2.5	1.70
25.0430	366.	164.	3.2	35.4	2.5	1.70
25.0530	366.	164.	3.2	35.4	2.5	1.70
25.0630	366.	164.	3.2	35.4	2.5	1.71
25.0730	366.	164.	3.2	35.5	2.5	1.71
25.0830	366.	164.	3.2	35.5	2.5	1.71
25.0930	349.	155.	3.2	35.4	2.5	1.71
25.1030	366.	155.	3.2	35.6	2.5	1.72
25.1130	348.	154.	3.2	35.2	2.5	1.70
25.1230	348.	155.	3.2	34.8	2.5	1.68
25.1330	365.	155.	3.2	34.4	2.3	1.65
25.1430	366.	155.	3.2	31.8	2.5	1.54
25.1530	366.	155.	3.2	35.0	2.5	1.69
25.1630	366.	155.	3.2	34.9	2.5	1.69
25.1730	366.	155.	3.2	35.3	2.4	1.70
25.1830	366.	165.	3.2	35.1	2.5	1.69
25.1930	365.	155.	3.2	34.1	2.5	1.64
25.2030	366.	155.	3.2	34.0	2.5	1.63
25.2130	374.	158.	3.2	33.9	2.5	1.63
25.2230	374.	158.	3.2	33.9	2.5	1.63
25.2330	349.	164.	3.2	33.5	2.4	1.61
26.0030	348.	164.	3.2	33.5	2.5	1.61
26.0130	366.	164.	3.2	34.4	2.4	1.65
26.0230	357.	164.	3.2	35.2	2.4	1.68
26.0330	374.	144.	3.2	35.8	2.5	1.71
26.0430	365.	154.	3.2	35.7	2.5	1.71
26.0530	366.	154.	3.3	35.8	2.5	1.71
26.0630	366.	154.	3.3	35.5	2.5	1.70

APPENDIX B: TABLE II.
RUN 5: GAS FLOW RATES PAGE 9 OF 9

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
26.0730	366.	164.	3.3	35.7	2.5	1.71
26.0830	357.	160.	3.3	35.9	2.5	1.72
26.0930	366.	158.	3.3	35.4	2.5	1.70
26.1030	339.	164.	3.3	36.0	2.4	1.71
26.1130	339.	164.	3.3	36.3	2.5	1.73
26.1230	348.	160.	3.3	36.0	2.3	1.71
26.1330	347.	155.	3.3	36.0	2.6	1.72
26.1430	401.	155.	3.3	35.8	2.5	1.71
26.1530	365.	151.	3.3	35.3	2.4	1.69
26.1630	366.	151.	3.3	35.1	2.5	1.68
26.1730	356.	145.	3.3	34.5	2.4	1.65
26.1830	365.	145.	3.3	34.7	2.5	1.66

APPENDIX B: TABLE III.
RUN 5: PRESSURES

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
2.0130	4.0	3.5	4.7	0.70	4.7
2.0230	3.7	3.5	5.0	0.70	5.7
2.0330	4.0	3.7	4.7	0.70	6.2
2.0430	4.2	4.0	4.6	0.70	6.7
2.0530	4.4	4.0	4.6	0.80	7.0
2.0630	4.4	4.0	4.6	0.75	7.5
2.0730	4.4	4.0	4.5	0.75	8.0
2.0830	MISSED DATA READING				
2.0930	4.5	4.0	4.6	0.75	8.0
2.1030	4.6	4.0	4.6	0.80	8.2
2.1130	4.0	4.0	4.7	0.70	8.0
2.1230	4.0	4.0	4.7	0.80	7.0
2.1330	4.0	4.0	4.7	0.80	8.3
2.1430	3.6	3.9	4.7	0.80	8.5
2.1530	4.0	3.9	4.7	0.70	7.0
2.1630	4.0	3.7	4.7	0.75	7.0
2.1730	4.0	3.7	4.9	0.75	7.0
2.1830	3.5	3.1	4.7	0.80	7.0
2.1930	3.5	3.1	4.7	0.80	7.0
2.2030	3.6	3.1	5.0	0.80	5.0
2.2130	3.9	3.5	4.7	0.75	5.0
2.2230	3.4	3.0	4.9	0.85	5.0
2.2330	3.4	3.1	4.9	0.85	5.0
3.0030	3.6	3.1	4.9	0.80	5.0
3.0130	MISSED DATA READING				
3.0230	3.5	3.5	5.0	0.80	5.0
3.0330	3.6	3.6	5.1	0.80	5.0
3.0430	3.5	3.5	5.2	0.85	5.0
3.0530	3.5	3.0	5.2	0.80	5.0
3.0630	3.5	3.0	5.2	0.85	5.1
3.0730	3.7	3.0	5.2	0.85	5.7
3.0830	3.7	3.0	5.2	0.85	5.7
3.0930	3.6	3.0	5.4	0.80	5.5
3.1030	3.6	3.2	5.4	0.85	5.2
3.1130	3.7	3.5	5.2	0.85	6.2
3.1230	4.0	3.0	5.4	0.85	6.7
3.1330	3.9	3.5	5.2	1.00	6.0
3.1430	3.9	3.5	5.4	1.00	6.5
3.1530	3.7	3.2	5.5	0.80	6.7
3.1630	4.0	3.2	5.0	0.75	6.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
3.1730	4.1	3.5	4.7	0.80	7.0
3.1830	4.1	3.5	4.5	0.70	6.5
3.1930	4.1	3.2	4.5	0.80	6.7
3.2030	4.1	3.4	4.5	0.80	7.0
3.2130	4.1	3.5	4.5	0.80	6.7
3.2230	4.2	3.4	4.5	0.80	6.7
3.2330	4.1	3.4	4.5	0.90	6.7
4.0030	4.2	3.4	4.5	0.80	6.7
		STONE CHANGE			
4.0130	4.1	3.2	4.5	0.80	7.0
4.0230	4.0	3.2	4.5	0.80	6.7
4.0330	4.4	3.4	4.4	0.80	6.8
4.0430	4.5	3.4	4.4	0.80	6.8
4.0530	4.5	3.4	4.2	0.75	7.0
4.0630	4.7	3.4	4.2	0.75	6.7
4.0730	4.9	3.4	4.7	0.75	6.7
4.0830	4.9	3.4	4.8	0.75	6.7
4.0930	5.5	3.7	4.5	0.80	4.5
4.1030	5.5	3.7	4.5	0.80	4.5
4.1130	5.5	3.7	4.5	0.80	4.5
4.1230	5.1	3.7	4.4	0.80	4.5
4.1330		MISSED DATA READING			
4.1430	5.2	4.0	4.5	0.70	4.5
4.1530	5.1	4.0	4.5	0.70	4.5
4.1630	5.2	4.0	4.5	0.70	4.5
4.1730	5.2	4.0	4.5	0.70	4.5
4.1830	5.2	4.0	4.5	0.70	4.5
4.1930	5.2	4.2	4.5	0.70	4.5
4.2030	5.2	4.2	4.5	0.70	4.5
4.2130	5.4	4.2	4.5	0.70	4.7
4.2230	5.2	4.2	4.5	0.70	5.0
4.2330	5.1	4.4	4.5	0.65	5.2
5.0030	5.1	3.9	4.5	0.65	5.0
5.0130		MISSED DATA READING			
5.0230	5.1	3.9	4.5	0.65	5.2
5.0330	5.0	3.7	4.5	0.65	5.2
5.0430	5.1	4.0	4.5	0.65	5.0
5.0530	5.2	4.7	4.5	0.60	4.7
5.0630	5.2	4.2	4.4	0.60	5.5
5.0730	5.4	4.4	4.5	0.60	5.2

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
5.0830	5.2	4.1	4.5	0.60	5.5
5.0930	5.1	4.0	4.5	0.60	5.2
5.1030	5.1	4.0	4.5	0.60	5.5
5.1130	5.0	4.2	4.5	0.60	5.7
5.1230	4.9	4.5	4.5	0.60	5.7
5.1330	MISSED DATA READING				
5.1430	5.1	3.9	4.4	0.60	5.7
5.1530	5.1	3.9	4.4	0.65	5.5
5.1630	5.2	3.9	4.4	0.65	5.5
SHUT DOWN AT 5.1630 FOR 10 HOURS					
6.0230	6.3	3.6	3.9	0.70	5.0
6.0330	6.0	3.7	4.1	0.65	5.5
6.0430	6.3	3.6	4.2	0.67	5.2
6.0530	6.2	3.6	4.4	0.67	5.7
6.0630	6.5	3.6	4.5	0.65	5.0
6.0730	7.0	3.5	4.5	0.70	5.7
6.0830	6.7	2.9	4.5	0.65	6.0
SHUT DOWN AT 6.0830 FOR 62 HOURS					
8.2230	3.2	4.0	4.6	0.85	0.
8.2330	3.4	4.2	4.6	0.75	0.
9.0030	3.3	4.2	4.7	0.75	0.
9.0130	3.2	4.2	4.9	0.75	0.
9.0230	3.2	4.2	5.0	0.75	0.
9.0330	3.2	4.2	5.0	0.70	0.
9.0430	3.2	4.2	5.0	0.80	0.
9.0530	3.2	4.2	5.0	0.75	0.
SHUT DOWN AT 9.0530 FOR 57 HOURS					
11.1430	3.9	0.	4.5	0.65	4.4
11.1530	3.9	0.	4.6	0.65	4.7
11.1630	3.7	0.	4.9	0.70	4.9
11.1730	3.7	0.	4.9	0.70	4.7

APPENDIX B: TABLE III.
RUN 5: PRESSURES PAGE 4 OF 9

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
11.1830	3.9	0.	5.4	0.70	5.7
11.1930	3.7	0.	5.5	0.70	5.5
11.2030	4.0	0.	5.5	0.70	5.7
11.2130	3.9	0.	5.5	0.70	5.5
11.2230	4.0	0.	5.4	0.70	5.0
11.2330	4.0	0.	5.4	0.70	5.0
12.0030	4.0	5.2	5.4	0.70	6.0
12.0130	MISSED DATA READING				
12.0230	4.1	5.2	5.2	0.70	5.5
12.0330	4.1	5.2	5.5	0.70	5.0
12.0430	4.1	5.2	5.2	0.70	5.0
12.0530	4.0	5.2	5.4	0.80	5.0
12.0630	4.0	5.0	5.2	0.70	5.0
12.0730	4.0	5.2	5.2	0.70	5.2
12.0830	4.1	5.2	5.2	0.70	5.5
12.0930	4.0	5.1	5.5	0.70	5.5
12.1030	3.7	4.6	5.5	0.70	5.5
12.1130	MISSED DATA READING				
12.1230	3.9	4.4	5.5	0.75	6.0
12.1330	3.7	4.6	5.5	0.70	6.0
12.1430	3.7	4.5	5.5	0.70	5.7
12.1530	3.7	4.7	5.7	0.70	6.2
12.1630	3.7	4.7	5.5	0.70	6.2
12.1730	3.7	4.7	5.5	0.70	6.0
12.1830	3.9	4.9	5.5	0.70	6.2
12.1930	3.9	5.0	5.2	0.70	6.0
12.2030	4.0	5.2	5.0	0.70	5.5
12.2130	4.0	5.4	7.6	0.70	5.7
12.2230	4.2	5.7	4.5	0.65	5.7
12.2330	4.2	5.7	4.5	0.65	4.7
13.0030	4.2	5.8	4.5	0.67	6.1
13.0130	4.2	5.6	4.6	0.67	5.4
13.0230	4.2	5.6	4.6	0.67	5.7
13.0330	4.2	5.6	4.6	0.67	5.5
13.0430	4.2	5.7	4.5	0.67	5.6
13.0530	4.2	5.6	4.5	0.67	5.5
13.0630	4.2	5.6	4.5	0.67	5.6

APPENDIX B: TABLE III.
RUN 5: PRESSURES

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
SHUT DOWN AT 13.0630 FOR 73 HOURS					
16.0730	4.2	5.1	4.2	0.75	6.5
16.0830	4.0	4.9	4.5	0.75	4.5
16.0930	4.0	5.0	4.9	0.75	3.7
16.1030	4.0	5.1	4.9	0.70	5.0
16.1130	4.0	5.5	5.0	0.70	4.5
16.1230	4.0	5.1	5.2	0.70	5.0
16.1330	4.0	5.0	5.2	0.70	5.2
16.1430	MISSED DATA READING				
16.1530	4.0	5.0	5.6	0.65	5.6
16.1630	4.0	4.7	5.7	0.70	5.5
16.1730	4.0	4.7	5.7	0.70	5.7
16.1830	4.0	4.9	5.5	0.70	5.2
16.1930	4.2	4.7	6.0	0.65	5.7
16.2030	4.2	5.0	6.1	0.60	5.1
16.2130	4.2	5.0	8.5	0.65	5.7
16.2230	4.2	5.0	6.2	0.65	5.7
16.2330	4.2	4.7	6.2	0.65	5.2
17.0030	4.4	5.0	6.0	0.65	5.0
17.0130	4.4	5.0	6.2	0.65	5.2
17.0230	4.4	5.0	6.0	0.65	5.5
17.0330	4.4	4.9	6.0	0.65	5.7
17.0430	4.4	5.0	6.0	0.65	5.7
17.0530	4.5	5.0	6.0	0.65	5.7
17.0630	4.4	4.9	6.2	0.65	6.0
17.0730	4.4	4.9	6.1	0.65	5.7
17.0830	MISSED DATA READING				
17.0930	4.4	5.0	6.2	0.65	6.0
17.1030	4.4	4.9	6.2	0.65	6.5
17.1130	4.4	4.9	6.2	0.70	6.2
17.1230	4.4	4.7	6.2	0.70	6.0
17.1330	4.4	4.7	6.2	0.70	6.0
17.1430	4.4	4.7	6.2	0.70	6.2
17.1530	4.4	4.7	6.2	0.70	5.2
17.1630	4.4	4.7	6.2	0.70	5.8
17.1730	4.4	4.5	6.5	0.70	5.8
17.1830	4.4	4.6	6.5	0.70	5.2

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
17.1930	4.4	4.6	6.5	0.70	5.7
17.2030	4.4	4.6	6.5	0.70	6.0
		STONE CHANGE			
	SHUT DOWN AT 17.2030 FOR 72 HOURS				
20.2030	5.5	4.8	4.8	0.73	5.8
20.2130	5.5	5.1	5.0	0.75	6.2
20.2230	5.6	4.7	5.1	0.70	6.2
20.2330	5.7	4.7	5.0	0.75	6.0
21.0030	5.8	4.5	5.1	0.70	6.0
21.0130	6.0	4.4	5.1	0.75	6.5
21.0230	6.0	4.4	5.2	0.70	6.5
21.0330	6.1	4.5	5.2	0.70	6.7
21.0430	6.1	4.5	5.5	0.75	6.5
21.0530	6.1	4.5	5.5	0.80	6.7
21.0630	6.2	4.5	5.5	0.75	6.7
21.0730	6.3	4.4	5.5	0.75	6.7
21.0830	6.2	4.2	5.5	0.80	6.7
21.0930	6.5	4.1	5.5	0.75	6.7
21.1030	6.3	4.2	5.6	0.75	6.7
21.1130	6.3	4.2	5.5	0.80	6.5
21.1230	6.2	4.1	5.6	0.75	6.2
21.1330	6.3	4.1	5.5	0.80	6.5
21.1430	6.3	4.1	5.5	0.80	5.7
21.1530	6.5	4.0	5.7	0.75	5.7
21.1630	6.6	4.1	5.6	0.75	6.2
21.1730	6.7	4.1	5.6	0.75	7.2
21.1830	6.8	4.2	5.7	0.75	6.0
21.1930	7.0	4.0	5.6	0.80	5.5
21.2030	6.1	4.4	5.4	0.75	6.2
21.2130	6.2	4.7	5.5	0.80	5.7
21.2230	6.1	4.7	5.5	0.75	5.7
21.2330	6.1	4.7	5.4	0.80	6.0
22.0030	6.2	4.7	5.5	0.75	6.7
22.0130	6.2	4.7	5.7	0.80	6.2
22.0230	6.3	4.7	5.6	0.75	6.2
22.0330	6.3	4.7	5.7	0.75	6.2
22.0430	6.5	4.7	5.7	0.80	6.0

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
22.0530	6.5	4.7	5.7	0.80	6.2
22.0630	6.5	4.7	5.7	0.80	6.2
22.0730	6.7	4.7	5.7	0.75	6.2
22.0830	6.6	4.9	5.7	0.80	6.2
22.0930	6.6	4.9	5.7	0.80	6.2
22.1030	6.5	4.9	5.7	0.75	6.5
22.1130	6.6	4.9	5.8	0.80	6.5
22.1230	6.6	4.9	5.8	0.80	6.5
22.1330	6.6	4.7	6.0	0.80	5.7
22.1430	6.6	4.7	6.0	0.80	6.0
22.1530	6.6	4.7	6.0	0.80	6.2
22.1630	6.6	4.7	6.0	0.80	6.2
22.1730	6.7	4.7	6.0	0.80	6.2
22.1830	6.8	4.7	6.0	0.80	5.5

SHUT DOWN AT 22.1830 FOR 23 HOURS

DAY.HOUR	GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
23.1730	5.2	6.0	4.6	0.70	5.0
23.1830	5.4	6.0	5.0	0.70	5.5
23.1930	5.2	6.0	5.2	0.70	5.7
23.2030	5.2	6.0	5.4	0.70	5.7
23.2130	5.1	5.8	5.5	0.70	5.5
23.2230	5.2	6.0	5.7	0.70	5.5
23.2330	5.4	6.1	5.7	0.70	6.0
24.0030	5.4	6.2	5.5	0.70	5.5
24.0130	5.5	6.2	5.7	0.70	6.0
24.0230	5.5	6.2	6.0	0.65	6.0
24.0330	5.6	6.2	6.0	0.70	6.0
24.0430	5.5	6.2	6.0	0.70	6.0
24.0530	5.5	6.1	6.0	0.70	6.0
24.0630	5.6	6.1	6.0	0.70	6.0
24.0730	5.5	6.1	6.0	0.70	6.0
24.0830	5.7	6.3	6.0	0.70	6.0
24.0930	6.0	6.2	6.1	0.70	6.0
24.1030	6.0	6.2	6.3	0.70	6.0
24.1130	5.7	6.2	6.3	0.75	6.0
24.1230	5.6	6.1	6.2	0.70	6.0
24.1330	5.5	6.3	6.3	0.75	6.0
24.1430	5.6	6.3	6.2	0.75	6.0

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
24.1530	5.5	6.3	6.1	0.75	6.0
24.1630	5.5	6.2	6.1	0.75	6.0
24.1730	5.6	6.2	6.0	0.70	6.0
24.1830	5.6	6.2	6.2	0.70	6.0
24.1930	5.7	6.5	6.2	0.70	6.0
24.2030	5.7	6.7	6.2	0.70	6.0
24.2130	5.7	6.7	6.2	0.70	6.0
24.2230	5.7	6.7	6.5	0.70	6.5
24.2330	5.8	6.7	6.2	0.70	6.7
25.0030	5.8	6.7	6.2	0.70	6.7
25.0130	5.7	6.7	6.2	0.70	6.7
25.0230	5.8	6.7	6.2	0.70	6.8
25.0330	5.8	6.7	6.3	0.70	7.0
25.0430	5.9	6.7	6.3	0.70	7.0
25.0530	6.0	6.7	6.2	0.70	7.0
25.0630	5.8	6.7	6.2	0.70	7.0
25.0730	5.8	6.7	6.2	0.70	7.0
25.0830	6.0	6.7	6.2	0.70	7.0
25.0930	6.0	6.7	6.3	0.70	7.0
25.1030	6.0	6.7	6.0	0.70	6.5
25.1130	5.8	6.7	6.0	0.70	6.5
25.1230	5.8	6.7	6.5	0.75	6.2
25.1330	5.8	6.7	6.3	0.70	6.2
25.1430	6.0	6.7	6.3	0.70	6.0
25.1530	5.8	6.7	6.3	0.70	6.0
25.1630	5.7	6.7	6.5	0.60	6.5
25.1730	5.7	6.7	6.5	0.70	6.5
25.1830	6.1	7.2	6.3	0.70	7.0
25.1930	6.2	7.2	6.2	0.65	6.2
25.2030	6.3	7.2	6.2	0.65	6.2
25.2130	6.2	7.3	6.1	0.65	6.2
25.2230	6.2	7.2	6.2	0.67	6.0
25.2330	6.2	7.3	6.2	0.67	6.2
26.0030	6.2	7.2	6.2	0.70	6.2
26.0130	6.2	7.2	6.2	0.70	6.2
26.0230	6.3	7.2	6.2	0.70	6.0
26.0330	6.2	7.2	6.2	0.70	6.0
26.0430	6.3	7.0	6.2	0.70	5.7
26.0530	6.3	7.1	6.2	0.70	5.7
26.0630	6.3	7.1	6.2	0.70	6.0

APPENDIX B: TABLE III.
RUN 5: PRESSURES PAGE 9 OF 9

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
26.0730	6.3	7.1	6.2	0.70	6.2
26.0830	6.3	7.1	6.2	0.70	6.2
26.0930	6.5	7.1	6.2	0.70	6.5
26.1030	6.5	7.1	6.2	0.65	6.6
26.1130	6.5	7.1	6.1	0.65	6.6
26.1230	6.5	7.0	6.2	0.65	6.6
26.1330	6.5	7.0	6.1	0.65	6.7
26.1430	6.5	7.0	6.2	0.65	7.0
26.1530	6.5	7.0	6.3	0.65	5.7
26.1630	6.7	7.0	6.3	0.65	6.2
26.1730	6.6	7.0	6.3	0.65	6.5
26.1830	6.6	7.0	6.2	0.65	6.5

APPENDIX B: TABLE IV.

RUN 5: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
2.0130	84.5	1.56	68.	23.2	2.46	24.4	25.5
2.0230	83.0	1.55	72.	23.3	2.04	32.5	36.4
2.0330	74.8	1.65	68.	22.2	0.74	46.4	42.0
2.0430	81.0	1.68	67.	22.9	0.	36.4	34.6
2.0530	95.9	1.66	58.	23.0	0.19	46.1	49.5
2.0630	97.8	1.63	62.	23.1	0.19	39.6	37.5
2.0730	69.1	1.66	60.	23.5	0.14	34.1	28.4
2.0830			MISSED DATA READING				
2.0930	46.9	1.64	62.	24.3	0.27	23.4	20.6
2.1030	45.1	1.74	58.	21.7	0.35	33.0	31.4
2.1130	63.9	2.06	68.	20.3	0.41	26.3	25.0
2.1230	64.5	1.62	60.	20.4	0.39	15.7	15.8
2.1330	58.2	1.65	60.	21.5	0.45	37.5	39.4
2.1430	63.9	1.62	60.	20.8	0.52	50.0	54.4
2.1530	66.7	1.64	68.	20.6	0.55	43.6	46.2
2.1630	70.3	1.66	64.	20.5	0.58	43.3	45.1
2.1730	75.0	1.60	66.	20.5	0.55	39.8	41.8
2.1830	74.3	1.42	60.	19.2	0.66	49.4	48.2
2.1930	76.2	1.42	60.	19.2	0.66	49.4	48.2
2.2030	75.6	1.45	63.	20.0	0.59	19.4	18.5
2.2130	76.2	1.64	64.	19.9	0.59	37.6	29.1
2.2230	69.9	1.67	58.	18.8	0.59	44.9	47.1
2.2330	69.9	1.45	58.	19.9	0.65	42.7	49.7
3.0030	70.5	1.43	61.	20.9	0.76	35.4	39.7
3.0130			MISSED DATA READING				
3.0230	69.6	1.41	63.	19.8	0.65	50.6	58.0
3.0330	70.8	1.41	65.	20.7	0.71	40.2	47.1
3.0430	71.4	1.41	62.	20.3	0.63	40.0	46.1
3.0530	70.7	1.40	66.	19.9	0.59	33.1	38.6
3.0630	73.0	1.36	62.	20.1	0.52	33.4	38.4
3.0730	71.7	1.36	62.	20.3	0.62	38.8	44.5
3.0830	77.5	1.37	62.	20.2	0.62	38.5	44.1
3.0930	79.5	1.36	68.	21.1	0.67	41.3	46.5
3.1030	72.7	1.38	64.	20.0	0.49	35.2	38.0
3.1130	75.9	1.48	62.	20.9	0.43	19.4	20.3
3.1230	75.0	1.45	64.	21.2	0.43	29.8	32.1
3.1330	77.5	1.42	53.	19.9	0.45	8.9	9.3
3.1430	68.5	1.45	54.	20.3	0.54	44.8	47.5
3.1530	72.5	1.45	69.	20.7	0.53	29.8	33.0
3.1630	74.4	1.38	67.	20.9	0.48	35.4	40.4

APPENDIX B: TABLE IV.
RUN 5: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
3.1730	66.6	1.47	60.	21.9	0.61	34.2	32.5
3.1830	65.5	1.45	65.	19.7	0.65	39.5	39.9
3.1930	58.7	1.50	57.	22.0	0.63	39.8	41.6
3.2030	56.4	1.44	57.	18.3	0.57	10.9	10.6
3.2130	57.6	1.52	57.	20.2	0.69	11.3	8.6
3.2230	59.7	1.60	57.	20.2	0.43	41.2	43.4
3.2330	52.2	1.55	50.	20.1	0.	12.7	14.4
4.0030	44.0	1.56	57.	21.0	0.	34.5	41.2
			STONE CHANGE				
4.0130	70.1	1.42	57.	19.6	0.	43.2	46.4
4.0230	69.6	1.37	57.	19.8	0.	50.5	58.6
4.0330	70.9	1.42	55.	20.7	0.	40.1	47.5
4.0430	71.4	1.46	55.	20.3	0.	40.0	46.6
4.0530	90.5	1.29	57.	19.2	2.73	-	-
4.0630	73.8	1.30	57.	20.3	3.65	25.7	26.9
4.0730	83.2	1.36	64.	20.1	2.97	51.0	57.1
4.0830	62.4	1.43	65.	20.4	1.53	45.9	53.1
4.0930	-	1.38	57.	20.6	0.52	37.8	45.9
4.1030	76.4	1.37	57.	19.4	0.	49.8	56.1
4.1130	-	1.43	57.	19.3	0.	53.4	63.7
4.1230	62.3	1.45	55.	19.7	1.34	32.3	32.1
4.1330			MISSED DATA READING				
4.1430	95.5	1.54	65.	21.0	1.73	53.6	54.2
4.1530	75.6	1.56	65.	21.2	1.77	41.5	45.4
4.1630	72.8	1.55	65.	21.8	1.71	35.3	36.1
4.1730	81.8	1.53	65.	21.5	1.67	25.6	27.4
4.1830	94.0	1.53	65.	21.2	2.14	52.7	58.6
4.1930	90.2	1.53	65.	21.0	2.31	51.0	48.4
4.2030	-	1.55	65.	21.2	2.11	48.0	53.5
4.2130	-	1.64	65.	21.2	1.62	55.7	62.7
4.2230	-	1.63	65.	21.3	1.65	52.2	57.4
4.2330	86.9	1.42	70.	21.2	1.77	46.2	53.8
5.0030	88.4	1.46	70.	19.7	1.91	55.0	61.8
5.0130			MISSED DATA READING				
5.0230	90.7	1.50	70.	21.5	1.82	16.2	17.9
5.0330	98.7	1.50	70.	20.7	1.79	10.3	10.8
5.0430	90.0	1.49	70.	19.3	1.74	42.4	45.0
5.0530	88.0	1.58	76.	20.8	1.72	35.0	37.9
5.0630	93.3	1.55	74.	20.9	1.70	41.8	45.1
5.0730	96.5	1.52	76.	21.2	1.35	40.3	44.5

APPENDIX B: TABLE IV.

RUN 5: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
5.0830	94.7	1.52	76.	20.9	1.75	44.8	47.4
5.0930	93.4	1.50	76.	20.8	2.26	53.6	55.8
5.1030	93.4	1.50	76.	20.9	1.83	48.2	51.2
5.1130	94.2	1.39	76.	20.3	1.73	51.9	54.3
5.1230	90.9	1.38	76.	19.4	1.14	45.8	51.7
5.1330	MISSED DATA READING						
5.1430	88.5	1.43	74.	20.4	0.83	50.9	56.6
5.1530	92.9	1.44	68.	21.0	0.81	48.6	55.8
5.1630	94.8	1.42	68.	20.6	0.98	44.2	45.5
SHUT DOWN AT 5.1630 FOR 10 HOURS							
6.0230	75.8	1.58	56.	20.1	0.91	38.0	38.7
6.0330	83.3	1.55	64.	19.9	1.39	44.8	42.5
6.0430	88.2	1.46	64.	19.7	1.32	31.8	31.1
6.0530	93.3	1.54	66.	20.1	1.30	43.7	38.5
6.0630	98.1	1.42	70.	19.8	1.13	52.1	48.2
6.0730	96.2	1.42	65.	20.7	1.10	48.4	46.0
6.0830	98.1	1.30	70.	18.7	1.20	59.9	58.6
SHUT DOWN AT 6.0830 FOR 62 HOURS							
8.2230	74.2	1.50	55.	21.8	1.53	-	-
8.2330	74.2	1.47	62.	21.4	1.56	5.7	7.0
9.0030	76.9	1.46	64.	22.0	2.00	-	-
9.0130	77.4	1.40	66.	20.7	2.03	14.7	18.7
9.0230	88.3	1.42	67.	21.0	1.82	15.0	20.0
9.0330	91.3	1.44	72.	21.5	1.62	-	-
9.0430	87.3	1.45	63.	21.6	1.19	27.0	40.3
9.0530	80.9	1.42	67.	21.2	0.93	33.2	51.9
SHUT DOWN AT 9.0530 FOR 57 HOURS							
11.1430	73.7	1.58	70.	21.1	2.39	25.2	29.2
11.1530	82.4	1.46	72.	20.0	2.18	12.3	13.6
11.1630	82.3	1.46	70.	21.0	0.96	32.3	41.0
11.1730	82.5	1.47	71.	21.0	0.96	20.5	25.8

APPENDIX B: TABLE IV.

RUN 5: DESULPHURISATION PERFORMANCE PAGE 4 OF 9

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
11.1830	93.1	1.44	78.	21.1	1.05	43.2	57.3
11.1930	95.8	1.49	79.	21.8	1.05	31.9	41.0
11.2030	91.5	1.46	79.	21.8	1.07	46.6	60.8
11.2130	80.3	1.55	79.	21.7	1.21	36.1	48.2
11.2230	83.8	1.54	78.	20.7	1.14	44.2	56.1
11.2330	74.7	1.47	78.	20.7	1.25	49.2	55.6
12.0030	72.0	1.53	78.	20.9	1.41	56.8	67.7
12.0130	MISSED DATA READING						
12.0230	81.2	1.48	76.	21.2	1.03	40.3	47.5
12.0330	78.8	1.47	79.	20.8	0.74	51.7	57.4
12.0430	80.7	1.47	76.	20.0	0.62	51.6	66.3
12.0530	81.4	1.47	68.	20.6	0.82	37.2	43.3
12.0630	71.4	1.50	76.	19.8	0.90	45.2	53.6
12.0730	77.3	1.44	76.	20.0	0.66	44.0	59.7
12.0830	78.8	1.43	76.	20.5	0.96	43.1	55.9
12.0930	81.2	1.38	79.	21.1	1.34	40.1	50.4
12.1030	82.0	1.38	79.	20.5	1.23	37.4	46.3
12.1130	MISSED DATA READING						
12.1230	80.8	1.39	74.	20.3	1.15	31.9	37.9
12.1330	83.8	1.38	79.	20.4	1.12	42.8	53.0
12.1430	84.2	1.38	79.	20.3	1.13	35.2	42.4
12.1530	83.1	1.32	83.	20.3	1.28	45.7	57.1
12.1630	83.5	1.38	79.	20.2	1.22	33.6	39.4
12.1730	83.7	1.36	79.	20.2	0.93	46.7	57.3
12.1830	84.7	1.40	79.	20.3	1.07	37.8	46.3
12.1930	83.9	1.40	76.	20.2	1.15	41.5	52.0
12.2030	82.3	1.44	72.	20.3	1.04	41.5	52.1
12.2130	80.0	1.46	110.	20.3	1.14	36.4	43.3
12.2230	76.4	1.52	70.	20.1	1.06	40.8	53.2
12.2330	77.7	1.50	70.	19.6	0.95	41.0	50.1
13.0030	81.0	1.48	68.	20.4	1.16	39.9	47.6
13.0130	80.2	1.48	70.	20.1	1.11	39.6	48.6
13.0230	84.3	1.49	70.	20.2	0.94	46.1	58.7
13.0330	80.2	1.55	70.	19.9	0.94	39.8	50.2
13.0430	87.1	1.55	68.	20.1	0.95	41.3	52.3
13.0530	89.4	1.50	68.	20.2	0.77	44.6	58.0
13.0630	87.3	1.49	68.	20.4	0.92	37.6	49.0

APPENDIX B: TABLE IV.

RUN 5: DESULPHURISATION PERFORMANCE PAGE 5 OF 9

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
SHUT DOWN AT 13.0630 FOR 73 HOURS							
16.0730	64.9	0.90	57.	20.6	1.09	-	-
16.0830	63.7	0.96	60.	20.5	1.56	12.9	13.3
16.0930	67.5	1.56	66.	21.1	1.53	-	-
16.1030	72.7	1.53	70.	21.2	1.36	38.3	41.5
16.1130	74.7	1.53	72.	21.2	1.26	31.8	34.7
16.1230	68.0	1.52	76.	21.3	1.69	31.4	38.3
16.1330	73.6	1.47	76.	22.4	1.77	43.3	52.1
16.1430	MISSED DATA READING						
16.1530	74.8	1.45	87.	22.9	1.70	30.4	35.7
16.1630	74.2	1.35	83.	22.2	1.84	39.9	45.7
16.1730	75.0	1.37	83.	22.4	1.98	48.4	57.7
16.1830	77.7	1.41	79.	22.5	2.18	51.0	62.6
16.1930	95.4	1.34	93.	22.0	2.12	42.9	51.3
16.2030	98.0	1.32	103.	21.6	2.08	52.9	66.5
16.2130	98.7	1.33	132.	20.8	1.96	15.8	15.1
16.2230	98.7	1.34	97.	20.7	1.98	42.1	43.8
16.2330	98.6	1.35	97.	21.2	0.93	34.2	38.0
17.0030	98.0	1.37	93.	21.5	0.98	37.6	43.6
17.0130	95.6	1.38	97.	23.3	0.89	45.1	52.7
17.0230	95.6	1.37	93.	23.4	0.92	41.1	51.0
17.0330	95.7	1.37	93.	23.3	1.08	-	-
17.0430	78.3	1.37	93.	22.3	0.77	41.5	48.6
17.0530	82.8	1.35	93.	24.9	1.28	39.7	84.7
17.0630	90.3	1.37	97.	23.3	1.08	39.7	45.0
17.0730	96.8	1.37	95.	23.4	1.11	41.7	49.0
17.0830	MISSED DATA READING						
17.0930	87.8	1.37	97.	21.5	0.41	42.3	48.7
17.1030	83.3	1.36	97.	21.6	1.25	42.4	47.3
17.1130	79.3	1.32	90.	20.6	1.02	50.3	58.4
17.1230	80.0	1.30	90.	21.3	1.00	47.2	52.9
17.1330	82.9	1.28	90.	20.8	1.19	47.5	54.4
17.1430	70.8	1.37	90.	21.5	1.31	48.1	49.2
17.1530	72.4	1.37	90.	22.4	1.88	41.9	47.7
17.1630	85.6	1.32	90.	21.4	1.84	22.9	27.5
17.1730	71.8	1.33	94.	22.3	1.33	36.8	43.2
17.1830	80.4	1.31	94.	21.8	1.17	53.8	58.4

APPENDIX B: TABLE IV.

RUN 5: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
17.1930	83.8	1.29	94.	21.3	1.09	41.5	48.7
17.2030	90.8	1.31	94.	21.8	0.98	46.0	50.2
			STONE CHANGE				
	SHUT DOWN AT 17.2030 FOR 72 HOURS						
20.2030	94.8	1.34	67.	22.2	2.69	26.0	27.6
20.2130	96.2	1.35	67.	20.9	1.52	16.5	17.9
20.2230	95.0	1.34	74.	20.1	1.70	32.6	33.7
20.2330	96.8	1.38	67.	20.6	1.70	37.0	38.2
21.0030	98.1	1.29	74.	20.9	1.61	33.9	37.3
21.0130	98.7	1.29	69.	22.2	1.59	36.3	38.8
21.0230	98.7	1.27	76.	21.0	1.57	26.4	26.6
21.0330	96.8	1.29	76.	19.8	1.15	28.5	29.7
21.0430	99.4	1.27	74.	20.1	1.33	37.2	37.0
21.0530	99.4	1.27	69.	20.1	1.33	45.7	44.5
21.0630	92.9	1.26	74.	20.2	1.17	36.8	36.4
21.0730	92.2	1.20	74.	19.5	1.26	30.9	29.4
21.0830	92.3	1.19	69.	19.1	1.23	45.0	39.6
21.0930	92.3	1.20	74.	19.8	1.08	36.8	36.7
21.1030	87.2	1.20	76.	19.5	1.20	33.7	31.8
21.1130	97.5	1.18	69.	19.1	1.26	39.3	37.0
21.1230	96.3	1.18	76.	19.0	1.28	48.1	39.9
21.1330	90.4	1.18	69.	19.2	1.42	38.5	38.5
21.1430	91.3	1.17	69.	19.2	1.34	43.8	39.5
21.1530	94.5	1.16	77.	19.0	1.44	33.2	31.5
21.1630	97.6	1.16	76.	18.9	1.41	65.2	47.1
21.1730	89.9	1.12	76.	19.0	1.44	43.2	36.1
21.1830	86.5	1.23	77.	18.9	1.43	57.4	46.0
21.1930	86.1	1.03	71.	19.0	1.53	60.0	48.5
21.2030	88.4	1.29	72.	18.4	1.05	38.9	38.2
21.2130	84.6	1.30	69.	18.9	0.73	35.6	36.6
21.2230	82.9	1.29	74.	18.4	0.94	36.5	35.0
21.2330	89.2	1.29	68.	19.2	1.26	36.5	38.0
22.0030	85.2	1.29	74.	19.5	1.53	36.5	34.0
22.0130	92.9	1.29	73.	19.2	1.44	41.6	43.5
22.0230	94.1	1.29	76.	19.7	0.99	34.3	36.5
22.0330	85.9	1.29	77.	19.5	1.17	51.9	46.1
22.0430	89.7	1.29	73.	19.1	0.93	43.6	39.8

APPENDIX B: TABLE IV.
RUN 5: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
22.0530	89.6	1.29	73.	19.0	0.93	42.6	44.6
22.0630	90.7	1.27	73.	19.5	1.03	47.0	49.7
22.0730	90.1	1.27	77.	20.0	1.04	36.3	37.8
22.0830	92.1	1.30	73.	19.1	1.11	54.2	49.5
22.0930	90.1	1.29	73.	19.5	1.14	33.4	36.3
22.1030	88.4	1.29	77.	19.5	1.02	26.7	27.8
22.1130	87.1	1.29	74.	19.5	0.92	35.2	36.9
22.1230	86.8	1.30	74.	19.6	1.02	46.8	36.7
22.1330	78.2	1.29	76.	19.0	1.06	50.5	50.7
22.1430	78.3	1.30	76.	19.7	0.93	47.3	45.5
22.1530	79.6	1.29	76.	19.8	0.96	44.1	45.3
22.1630	86.5	1.30	76.	19.7	1.15	33.5	33.4
22.1730	84.2	1.30	76.	19.7	0.95	45.6	45.8
22.1830	83.9	1.28	76.	19.7	1.12	29.9	29.0
SHUT DOWN AT 22.1830 FOR 23 HOURS							
23.1730	77.3	1.35	67.	17.7	2.36	15.2	17.6
23.1830	88.7	1.36	72.	18.2	2.36	27.8	31.2
23.1930	91.3	1.36	76.	18.6	2.41	37.7	44.1
23.2030	-	1.39	78.	19.6	2.22	10.8	12.3
23.2130	91.3	1.28	79.	17.6	2.20	24.4	26.4
23.2230	91.4	1.33	83.	18.5	2.31	31.4	34.4
23.2330	91.2	1.34	83.	18.7	2.31	12.6	14.0
24.0030	99.1	1.38	79.	19.9	2.30	39.5	45.2
24.0130	99.1	1.39	83.	19.6	2.27	31.9	36.1
24.0230	99.3	1.35	93.	19.4	2.29	37.2	41.2
24.0330	98.9	1.35	87.	19.3	2.11	38.5	40.4
24.0430	98.9	1.35	87.	19.1	2.04	35.1	37.1
24.0530	99.3	1.33	87.	19.4	2.13	32.0	32.1
24.0630	99.3	1.34	87.	19.5	2.01	32.6	35.6
24.0730	99.3	1.28	87.	18.1	2.03	47.5	49.3
24.0830	99.3	1.28	87.	19.4	2.48	43.1	44.5
24.0930	99.3	1.32	88.	19.8	2.48	26.3	28.3
24.1030	99.3	1.32	92.	20.8	2.20	26.6	28.7
24.1130	99.3	1.31	86.	19.7	2.31	39.0	42.5
24.1230	99.3	1.29	90.	18.7	1.92	36.2	38.0
24.1330	99.3	1.35	86.	19.2	0.67	43.7	46.4
24.1430	96.1	1.31	84.	18.9	0.96	38.9	42.7

APPENDIX B: TABLE IV.
RUN 5: DESULPHURISATION PERFORMANCE PAGE 8 OF 9

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
24.1530	96.1	1.31	82.	18.7	1.21	39.3	42.2
24.1630	97.4	1.34	82.	18.7	1.28	35.8	38.9
24.1730	91.6	1.36	87.	19.7	1.51	35.5	38.3
24.1830	-	1.31	90.	18.6	1.47	38.9	41.9
24.1930	-	1.33	90.	18.6	1.67	35.4	38.2
24.2030	-	1.34	90.	18.7	1.54	30.6	35.0
24.2130	-	1.32	90.	18.8	1.51	26.8	30.5
24.2230	-	1.32	94.	19.0	1.67	37.6	43.6
24.2330	96.0	1.32	90.	19.1	1.45	23.3	26.7
25.0030	91.5	1.31	90.	18.9	1.67	23.0	28.5
25.0130	94.2	1.32	90.	18.9	1.74	17.6	20.0
25.0230	89.3	1.31	90.	18.9	1.61	26.3	30.6
25.0330	92.8	1.32	92.	19.0	1.67	38.4	43.8
25.0430	94.7	1.32	92.	19.2	1.30	36.0	42.0
25.0530	95.4	1.32	90.	19.1	1.45	35.9	41.9
25.0630	92.7	1.32	90.	19.1	1.58	36.7	43.8
25.0730	94.3	1.32	90.	19.4	1.37	38.6	44.7
25.0830	90.3	1.32	90.	19.2	1.59	31.8	37.8
25.0930	95.4	1.28	92.	18.2	1.46	30.5	35.7
25.1030	-	1.32	87.	19.2	1.58	34.4	40.1
25.1130	98.6	1.26	87.	18.4	1.27	41.4	48.5
25.1230	85.7	1.26	88.	18.2	1.52	37.8	43.4
25.1330	92.9	1.32	92.	19.1	1.46	47.1	51.7
25.1430	89.9	1.32	92.	19.2	1.35	29.9	30.3
25.1530	89.9	1.31	92.	19.2	1.20	28.2	31.3
25.1630	87.3	1.32	110.	19.3	1.40	33.5	37.8
25.1730	90.3	1.31	94.	18.6	0.95	37.6	42.9
25.1830	90.0	1.35	92.	19.0	1.05	38.1	43.2
25.1930	86.4	1.31	97.	18.7	1.02	38.2	41.9
25.2030	88.4	1.31	97.	18.7	1.15	35.6	38.0
25.2130	89.6	1.32	95.	19.1	0.98	40.3	41.9
25.2230	91.4	1.32	94.	19.2	1.15	33.3	35.8
25.2330	90.3	1.27	94.	18.0	1.15	32.2	33.6
26.0030	90.8	1.26	90.	17.9	1.15	44.3	46.0
26.0130	92.9	1.31	90.	18.8	1.02	36.2	38.5
26.0230	91.3	1.28	90.	18.3	1.11	25.1	28.4
26.0330	90.0	1.28	90.	19.2	1.02	30.2	35.4
26.0430	88.5	1.29	90.	18.7	1.08	30.6	35.4
26.0530	91.0	1.29	90.	18.8	1.02	40.7	44.5
26.0630	89.1	1.29	90.	18.8	0.86	43.1	48.2

APPENDIX B: TABLE IV.
RUN 5: DESULPHURISATION PERFORMANCE PAGE 9 OF 9

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
26.0730	89.9	1.30	90.	18.7	1.37	20.0	22.3
26.0830	92.4	1.27	90.	18.2	1.40	32.3	35.9
26.0930	90.3	1.29	90.	18.8	1.21	33.1	37.4
26.1030	90.3	1.23	97.	17.4	1.36	38.3	43.9
26.1130	92.9	1.23	95.	17.5	1.40	34.4	36.7
26.1230	90.3	1.24	97.	17.8	1.43	46.9	49.1
26.1330	90.3	1.25	95.	17.8	1.34	35.2	40.2
26.1430	92.3	1.37	97.	20.5	1.43	31.6	35.6
26.1530	88.4	1.27	99.	18.7	1.66	33.7	37.3
26.1630	85.9	1.28	99.	18.7	1.37	32.1	35.0
26.1730	86.0	1.25	99.	18.3	1.34	35.3	36.9
26.1830	83.9	1.25	97.	18.8	1.38	35.1	38.8

APPENDIX B - Table V

RUN 5: GAS COMPOSITIONS PAGE 1 OF 6

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
2.0130	2.8	14.4	13.6	210.	0.	0.5	2.9	16.5	17.3	2.68	2.89
2.0230	3.0	14.4	13.4	228.	0.	0.6	3.9	16.5	17.4	2.60	2.89
2.0330	3.3	13.2	13.0	328.	0.	0.6	5.4	15.0	16.4	3.36	3.46
2.0430	2.7	13.8	13.5	255.	0.	0.5	4.2	15.0	16.6	3.27	3.29
2.0530	2.8	13.5	13.4	55.	0.	0.9	5.4	15.0	16.7	3.18	3.23
2.0630	3.5	12.7	12.9	27.	0.	0.5	4.6	15.0	17.0	3.10	2.91
2.0730	2.5	13.5	13.9	429.	0.	0.7	3.9	15.0	16.9	3.01	2.97
2.0830	MISSED DATA READING										
2.0930	2.1	14.4	14.1	757.	0.	0.6	2.7	15.5	17.3	2.76	2.84
2.1030	3.0	14.4	13.5	748.	0.	0.6	3.9	16.0	16.3	2.44	3.75
2.1130	0.8	15.6	15.0	547.	0.	1.7	2.9	14.0	13.3	3.63	5.96
2.1230	1.5	15.3	14.5	520.	0.40	0.8	1.8	14.0	15.0	3.54	4.67
2.1330	3.0	14.4	13.4	565.	0.	0.6	4.4	14.5	16.0	3.27	4.01
2.1430	2.8	15.0	13.6	492.	0.	1.3	5.8	14.0	15.7	3.45	4.33
2.1530	2.0	14.7	14.2	474.	0.	1.3	5.0	14.5	16.3	3.27	3.59
2.1630	2.1	14.4	14.1	420.	0.	1.1	5.0	14.5	16.4	3.10	3.52
2.1730	2.4	14.4	13.9	347.	0.	1.0	4.6	14.5	16.5	3.10	3.44
2.1830	2.0	14.7	14.2	365.	0.	0.8	5.8	14.8	16.1	2.93	3.79
2.1930	2.0	14.7	14.2	337.	0.	0.8	5.8	14.8	16.1	2.93	3.79
2.2030	2.5	14.1	13.9	337.	0.	0.6	2.2	15.0	17.1	2.76	2.93
2.2130	2.0	14.7	14.2	338.	0.	1.5	4.0	14.0	17.0	3.27	3.11
2.2230	2.3	14.4	14.1	420.	0.	1.7	5.0	14.5	19.4	3.01	1.19
2.2330	2.4	14.4	13.9	420.	0.	1.0	5.0	15.0	17.1	2.84	3.03
3.0030	2.0	14.4	14.2	420.	0.	1.2	4.0	15.0	16.9	2.76	3.11
3.0130	MISSED DATA READING										
3.0230	2.6	14.4	13.8	419.	0.	1.0	6.0	15.0	17.1	2.68	3.02
3.0330	2.6	14.1	13.8	401.	0.	1.3	4.6	15.0	17.4	2.60	2.77
3.0430	2.3	14.4	14.0	401.	0.	1.9	4.4	15.5	17.2	2.60	2.89

3.0530	2.3	14.1	14.0	410.	0.05	1.0	3.9	15.5	17.4	2.52	2.71
3.0630	2.1	14.7	14.2	383.	0.05	1.1	3.9	15.5	17.2	2.52	2.96
3.0730	2.1	14.4	14.2	401.	0.	1.4	4.4	15.5	17.2	2.52	2.90
3.0830	2.0	14.7	14.2	319.	0.	1.3	4.4	16.0	17.2	2.52	2.96
3.0930	2.0	14.4	14.2	292.	0.	1.7	4.6	16.0	17.0	2.44	3.00
3.1030	1.8	14.4	14.4	392.	0.	1.1	4.0	15.5	16.3	2.76	3.55
3.1130	1.8	14.4	14.4	347.	0.	0.8	2.2	15.0	16.1	2.76	3.69
3.1230	2.0	14.4	14.2	356.	0.	0.8	3.5	15.0	16.2	2.93	3.60
3.1330	2.0	14.7	14.2	319.	1.00	0.8	1.0	14.5	16.0	2.93	3.89
3.1430	2.5	13.8	13.8	438.	0.	1.7	5.0	14.5	15.8	3.10	3.85
3.1530	1.5	14.4	14.6	401.	0.	0.8	3.5	15.5	15.6	3.10	4.01
3.1630	2.0	14.4	14.2	365.	0.	0.4	4.2	15.5	15.4	2.93	4.20
3.1730	2.0	14.4	14.2	474.	0.	1.0	3.9	15.2	15.7	3.01	4.00
3.1830	2.8	13.5	13.6	474.	0.	1.6	4.4	15.5	16.2	3.01	3.57
3.1930	2.8	13.5	13.6	565.	0.	1.6	4.4	15.5	16.5	3.01	3.33
3.2030	2.2	14.4	14.1	620.	0.	0.8	1.2	15.5	15.9	2.93	3.86
3.2130	2.1	14.7	14.2	602.	0.	0.6	1.2	15.2	16.7	2.93	3.30
3.2230	1.7	14.4	14.5	584.	0.	1.6	4.6	15.2	17.2	2.76	2.79
3.2330	2.1	0.0	14.3	684.	0.	0.2	1.5	15.1	17.0	2.93	0.00
4.0030	2.1	0.0	14.3	803.	0.	0.0	4.2	15.2	17.0	2.93	0.00
			STONE CHANGE								
4.0130	2.3	14.4	14.1	420.	0.	1.1	5.0	15.0	17.4	2.76	2.73
4.0230	2.6	14.4	13.8	420.	0.	1.0	6.0	15.0	17.5	2.68	2.73
4.0330	2.6	14.1	13.8	401.	0.	1.3	4.6	15.0	17.6	2.60	2.62
4.0430	2.3	14.4	14.0	401.	0.	1.9	4.4	15.0	17.6	2.60	2.59
4.0530	2.5	13.8	13.8	137.	4.50	1.0	-	16.0	14.9	3.10	4.58
4.0630	3.0	14.4	13.4	374.	0.	1.6	2.9	16.0	16.3	2.84	3.73
4.0730	3.0	14.1	13.5	237.	0.	1.8	5.8	16.0	17.5	2.76	2.77
4.0830	4.0	13.0	12.7	492.	0.	1.1	5.4	16.5	17.3	2.60	2.82
4.0930	3.5	13.5	13.1	-	0.	0.2	4.6	16.5	17.2	2.44	2.93
4.1030	3.0	14.4	13.5	319.	0.	1.3	5.8	15.5	16.8	2.93	3.38
4.1130	2.5	13.8	13.9	-	0.	0.4	6.6	14.0	16.3	3.63	3.52
4.1230	3.0	13.8	13.5	520.	2.00	0.3	3.5	16.0	16.4	3.01	3.51
4.1330			MISSED DATA READING								
4.1430	3.5	13.2	13.1	59.	0.	1.3	6.2	15.0	16.8	3.10	3.21
4.1530	2.5	13.8	13.8	347.	0.	0.4	5.0	15.0	16.7	2.93	3.21

RUN 5: GAS COMPOSITIONS

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
4.1630	4.0	13.2	12.7	356.	0.	0.2	4.2	15.8	16.9	2.84	3.15
4.1730	4.0	12.7	12.7	237.	0.	0.2	3.1	15.5	17.1	2.76	2.91
4.1830	3.0	13.2	13.5	82.	0.	1.2	6.2	15.5	16.9	2.93	3.03
4.1930	2.8	13.8	13.6	137.	0.	1.4	5.8	15.5	16.8	3.01	3.16
4.2030	3.5	13.8	13.1	-	0.	0.6	5.8	15.8	16.9	2.76	3.23
4.2130	3.2	13.8	13.3	-	0.	1.1	6.6	15.5	17.5	2.93	2.72
4.2230	3.5	14.1	13.1	-	0.	1.0	6.2	16.0	17.4	2.84	2.88
4.2330	4.5	13.0	12.3	164.	0.	0.5	5.6	15.0	15.3	3.27	4.52
5.0030	3.5	13.0	13.1	155.	0.	0.6	6.6	17.0	16.5	2.44	3.36
5.0130	MISSED DATA READING										
5.0230	3.0	14.4	13.4	128.	0.	1.0	1.8	16.0	16.7	2.76	3.40
5.0330	3.0	14.4	13.4	18.	0.	0.6	1.2	16.0	16.9	2.93	3.23
5.0430	3.0	14.4	13.5	137.	0.	0.6	5.0	16.0	16.7	2.76	3.40
5.0530	3.0	14.4	13.5	164.	0.	0.4	4.2	16.0	17.5	2.84	2.81
5.0630	3.0	13.8	13.5	91.	0.	1.0	5.4	16.0	17.1	2.76	2.98
5.0730	3.5	13.2	13.1	46.	0.	0.6	5.0	16.0	17.0	2.76	3.02
5.0830	3.0	14.4	13.4	73.	0.	1.6	5.4	16.0	16.9	2.76	3.28
5.0930	3.0	14.1	13.4	91.	0.	2.0	6.2	16.0	16.9	2.93	3.21
5.1030	3.0	14.4	13.4	91.	0.	1.7	5.8	16.0	16.9	2.93	3.27
5.1130	2.5	14.4	13.8	82.	0.	2.1	6.0	16.5	17.2	2.60	2.92
5.1230	2.5	14.7	13.8	128.	0.	0.8	5.8	16.5	16.9	2.76	3.24
5.1330	MISSED DATA READING										
5.1430	2.0	14.4	14.2	164.	0.	1.3	6.2	16.5	16.9	2.60	3.11
5.1530	2.2	14.7	14.1	100.	0.	0.4	6.2	17.0	17.1	2.60	3.06
5.1630	2.5	14.1	13.8	73.	0.	0.7	5.6	17.0	17.2	2.60	2.90
SHUT DOWN AT 5.1630 FOR 10 HOURS											
6.0230	2.5	14.4	13.8	338.	0.	1.7	4.2	15.0	16.8	3.36	3.30
6.0330	1.5	15.0	14.5	246.	0.	2.5	4.6	15.0	16.2	3.45	3.67

6.0430	1.5	15.0	14.5	173.	0.	1.7	3.5	15.5	16.0	3.45	3.81
6.0530	3.0	14.4	13.4	91.	0.	2.0	5.0	16.0	17.0	2.93	3.17
6.0630	2.0	15.0	14.2	27.	0.	1.7	6.2	16.0	16.9	3.10	3.25
6.0730	1.5	15.6	14.6	55.	0.	2.2	5.6	16.0	17.1	2.93	3.11
6.0830	2.0	15.0	14.2	27.	0.	3.1	6.6	16.0	16.6	2.93	3.49

SHUT DOWN AT 6.0830 FOR 62 HOURS

8.2230	3.0	13.8	13.4	356.	0.20	0.8	-	15.7	16.1	4.02	3.77
8.2330	3.0	14.1	13.4	356.	0.20	0.5	0.7	15.7	16.0	3.82	3.92
9.0030	2.5	14.4	13.8	328.	0.	0.4	-	15.3	16.2	4.02	3.71
9.0130	3.2	14.1	13.3	310.	0.	0.8	1.8	15.8	16.3	3.63	3.68
9.0230	2.7	14.1	13.6	164.	0.	0.6	1.8	15.8	16.4	3.63	3.55
9.0330	3.0	14.1	13.4	119.	0.	0.9	-	16.0	16.7	3.45	3.40
9.0430	3.0	14.4	13.4	173.	0.	1.0	3.1	16.0	16.7	3.45	3.47
9.0530	2.2	14.4	14.0	274.	0.	0.9	3.9	16.0	16.7	3.45	3.31

SHUT DOWN AT 9.0530 FOR 57 HOURS

11.1430	3.5	13.5	13.0	356.	0.	0.8	2.7	16.0	16.7	4.42	3.34
11.1530	3.5	13.8	13.1	237.	0.	1.0	1.2	16.0	17.2	4.02	3.01
11.1630	4.0	13.5	12.7	228.	0.	1.0	3.5	16.5	17.3	4.02	2.94
11.1730	3.8	13.8	12.9	228.	0.	0.6	2.2	16.0	17.3	4.02	3.00
11.1830	3.5	13.2	13.1	91.	0.	0.7	4.6	16.5	16.7	3.82	3.23
11.1930	3.5	13.2	13.1	55.	0.	0.6	3.5	17.0	17.6	3.63	2.60
11.2030	4.0	13.2	12.7	109.	0.	1.0	5.0	17.0	17.8	3.45	2.46
11.2130	4.0	13.2	12.8	255.	0.	1.0	3.9	16.5	18.6	3.27	1.84
11.2230	4.0	13.5	12.7	210.	0.	0.8	5.0	16.5	17.8	4.02	2.51
11.2330	3.5	13.5	13.1	337.	0.	1.0	5.4	16.5	17.2	3.92	2.92
12.0030	4.0	13.8	12.7	365.	0.	1.5	6.2	16.8	17.7	2.76	2.64
12.0130	MISSED DATA READING										

DAY.HOUR	FLUE GAS O_2 %	FLUE GAS CO_2 VOL % ANAL	FLUE GAS CO_2 VOL % CALC	FLUE GAS SO_2 PPM	REGENERATOR GAS O_2 %	REGENERATOR GAS CO_2 %	REGENERATOR GAS SO_2 %	GASIFIER INLET GAS O_2 VOL % ANAL	GASIFIER INLET GAS O_2 VOL % CALC	GASIFIER INLET GAS CO_2 VOL % ANAL	GASIFIER INLET GAS CO_2 VOL % CALC
12.0330	4.0	13.2	12.9	274.	0.	0.8	5.4	16.5	17.6	2.60	2.62
12.0430	3.5	14.1	13.1	255.	0.	1.3	5.4	16.5	17.5	2.84	2.79
12.0530	3.5	14.1	13.1	246.	0.	1.1	3.7	16.5	17.5	2.76	2.79
12.0630	3.0	14.1	13.5	392.	0.	1.3	4.6	16.0	16.9	2.93	3.18
12.0730	4.0	13.2	12.7	292.	0.	1.2	4.6	16.0	16.4	3.10	3.57
12.0830	4.0	13.2	12.7	274.	0.	0.8	4.6	16.0	16.4	3.01	3.57
12.0930	4.2	13.5	12.3	237.	0.	1.1	4.0	16.7	17.2	2.76	3.09
12.1030	3.7	13.8	13.0	237.	0.	1.3	3.9	17.0	17.6	2.60	2.71
12.1130	MISSED DATA READING										
12.1230	3.5	13.8	13.1	255.	0.	1.6	3.1	17.0	17.6	2.60	2.70
12.1330	4.0	13.8	12.7	210.	0.	1.3	4.4	17.0	17.7	2.44	2.70
12.1430	3.5	13.8	13.1	210.	0.	1.4	3.5	17.0	17.6	2.44	2.71
12.1530	3.3	14.1	13.2	228.	0.	1.6	4.8	17.0	17.4	2.44	2.90
12.1630	3.6	13.8	13.0	219.	0.	1.9	3.5	17.0	18.0	2.44	2.34
12.1730	3.3	13.8	13.3	219.	0.	1.9	5.0	17.0	17.9	2.44	2.42
12.1830	3.7	14.1	13.0	201.	0.	1.3	4.6	17.2	18.3	2.44	2.22
12.1930	3.8	13.8	12.9	210.	0.	1.3	4.6	17.0	18.3	2.44	2.17
12.2030	4.0	13.2	12.7	228.	0.	1.3	4.6	17.0	18.2	2.44	2.21
12.2130	3.0	13.8	13.5	274.	0.	1.9	3.9	16.7	17.2	2.68	2.92
12.2230	3.7	13.5	12.9	310.	0.	1.0	4.6	16.7	16.9	2.68	3.17
12.2330	3.2	14.1	13.3	301.	0.	1.9	4.0	15.7	15.8	3.18	4.08
13.0030	4.6	14.1	12.2	237.	0.	2.0	4.2	15.7	16.3	3.18	4.07
13.0130	3.4	14.1	13.1	265.	0.	1.6	4.0	15.8	15.9	3.10	4.07
13.0230	4.1	13.2	12.6	201.	0.	1.4	4.8	16.0	16.1	3.01	3.83
13.0330	3.7	13.5	13.0	265.	0.	1.6	4.0	15.8	16.4	3.10	3.56
13.0430	4.2	13.2	12.6	164.	0.	1.4	4.2	16.0	16.6	2.93	3.49
13.0530	3.9	13.2	12.8	137.	0.	1.4	4.6	16.2	16.1	2.93	3.83
13.0630	3.9	13.5	12.8	164.	0.	1.5	3.9	15.9	16.0	3.01	3.91

SHUT DOWN AT 13.0630 FOR 73 HOURS

16.0730	4.0	13.8	12.8	456.	0.	0.6	-	21.0	21.0	0.02	0.
16.0830	4.0	14.4	12.8	474.	0.	1.0	1.2	20.5	20.5	0.34	0.42
16.0930	3.0	14.4	13.4	447.	0.	1.0	-	16.0	16.9	3.27	3.30
16.1030	3.0	14.4	13.4	374.	0.	1.0	4.2	15.5	16.8	3.45	3.39
16.1130	3.0	14.4	13.4	347.	0.	0.8	3.5	15.8	16.9	3.45	3.30
16.1230	3.5	14.4	13.1	429.	0.	0.4	3.5	16.0	16.8	3.10	3.47
16.1330	3.0	14.1	13.4	365.	0.	0.8	5.0	16.5	16.5	3.27	3.54
16.1430	MISSED DATA READING										
16.1530	4.0	13.5	12.7	328.	0.	0.6	3.5	16.8	16.9	2.84	3.25
16.1630	4.0	13.5	12.7	337.	0.	0.6	4.6	16.5	17.2	2.76	3.05
16.1730	3.0	14.4	13.4	346.	0.	0.6	5.6	16.8	17.0	2.68	3.18
16.1830	3.0	14.4	13.4	310.	0.	1.0	5.6	17.0	16.8	2.76	3.34
16.1930	3.0	14.4	13.4	64.	0.	1.0	4.6	17.0	16.9	2.60	3.28
16.2030	3.0	13.8	13.4	27.	0.	0.8	5.8	17.0	16.8	2.76	3.21
16.2130	3.0	14.4	13.4	18.	0.	1.9	1.5	17.0	16.9	2.60	3.28
16.2230	2.5	14.4	13.8	18.	0.	1.7	4.2	17.0	16.9	2.60	3.20
16.2330	3.0	13.8	13.5	18.	0.	1.6	3.5	17.0	17.0	2.60	3.07
17.0030	3.0	13.8	13.5	27.	0.	1.1	3.9	17.0	17.1	2.60	3.01
17.0130	-	-	8.3	-	0.	1.1	5.0	17.0	18.6	2.44	0.36
17.0230	-	-	8.3	-	0.	1.0	4.6	17.0	18.5	2.44	0.05
17.0330	-	-	8.3	-	17.50	0.0	-0.0	17.0	18.5	2.44	0.44
17.0430	-	-	8.3	-	0.	1.1	4.6	17.0	18.5	2.44	0.51
17.0530	-	-	8.3	-	0.	1.4	4.6	17.0	18.3	2.44	0.39
17.0630	-	-	8.3	-	0.	1.6	4.2	17.0	18.5	2.44	0.24
17.0730	-	-	8.3	-	0.	1.1	4.6	17.0	18.5	2.28	-0.67
17.0830	MISSED DATA READING										
17.0930	3.0	13.8	13.5	164.	0.	1.7	4.6	17.0	17.0	2.44	3.08
17.1030	3.0	14.1	13.4	228.	0.	1.9	4.6	17.0	17.0	2.44	3.14
17.1130	3.0	14.1	13.5	283.	0.	1.6	5.6	17.0	16.8	2.44	3.26
17.1230	3.0	14.1	13.5	274.	0.	1.7	5.4	17.5	17.4	2.20	2.80
17.1330	2.8	14.4	13.6	237.	0.	1.6	5.0	17.0	17.3	2.44	2.91
17.1430	2.7	14.1	13.7	410.	0.	2.6	5.0	17.2	17.3	2.44	2.87

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
17.1530	3.0	14.1	13.4	383.	0.	1.6	4.6	17.0	17.3	2.20	2.87
17.1630	2.8	14.4	13.6	201.	0.	0.4	2.7	17.0	17.2	2.44	3.04
17.1730	2.8	14.4	13.6	392.	0.	1.6	4.0	17.5	17.5	2.28	2.75
17.1830	2.8	14.4	13.8	274.	0.	3.1	5.4	17.5	17.4	2.12	2.81
17.1930	2.5	14.4	13.8	228.	0.	1.6	4.6	17.5	17.1	1.96	3.02
17.2030	2.7	14.4	13.7	128.	0.	3.1	4.6	17.7	17.3	2.12	2.87

STONE CHANGE

SHUT DOWN AT 17.2030 FOR 72 HOURS

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
20.2030	2.0	15.0	14.2	73.	0.20	1.3	2.7	18.0	18.4	1.65	2.04
20.2130	1.6	15.6	14.5	55.	0.20	1.3	1.8	17.5	17.8	2.20	2.53
20.2230	1.4	15.6	14.7	73.	0.20	2.8	3.5	17.7	17.9	2.12	2.47
20.2330	2.0	15.3	14.2	46.	0.20	2.6	4.0	18.0	18.1	1.81	2.33
21.0030	2.0	15.0	14.3	27.	0.20	1.1	4.0	19.0	19.0	1.33	1.57
21.0130	2.0	15.0	14.3	18.	0.20	1.5	4.2	19.0	19.0	1.24	1.57
21.0230	2.0	15.0	14.3	18.	0.40	1.3	3.1	19.3	19.3	1.08	1.34
21.0330	1.8	15.0	14.4	46.	0.20	0.8	3.5	18.7	18.7	1.41	1.79
21.0430	1.8	15.3	14.4	9.	0.20	1.9	4.2	18.8	18.8	1.33	1.75
21.0530	2.0	15.3	14.3	9.	0.20	2.6	5.0	18.8	18.8	1.33	1.77
21.0630	2.2	15.3	14.1	100.	0.20	1.8	4.2	19.0	19.0	1.33	1.62
21.0730	2.3	15.0	14.1	109.	0.30	2.0	3.5	19.0	19.0	1.33	1.60
21.0830	2.2	15.0	14.1	109.	0.20	3.7	4.6	19.0	19.0	1.33	1.59
21.0930	2.0	15.0	14.3	109.	0.20	1.9	4.2	19.0	19.0	1.33	1.57
21.1030	2.0	14.7	14.3	182.	0.20	1.9	3.9	19.0	19.0	1.24	1.54
21.1130	1.5	15.3	14.7	37.	0.10	2.3	4.4	19.0	19.0	1.33	1.56
21.1230	1.2	15.3	14.9	55.	0.10	4.7	4.6	19.0	19.0	1.33	1.54
21.1330	2.0	14.4	14.3	137.	0.10	1.3	4.6	19.0	19.0	1.24	1.51
21.1430	1.5	15.3	14.6	128.	0.10	3.3	4.6	19.0	19.0	1.24	1.56
21.1530	1.2	15.3	14.9	82.	0.10	1.7	3.9	19.0	19.0	1.24	1.54

21.1630	1.0	15.0	15.0	37.	0.10	6.7	5.4	19.0	19.0	1.24	1.50
21.1730	1.8	14.7	14.4	146.	0.10	4.3	4.2	19.5	19.5	1.16	1.14
21.1830	1.2	15.3	14.8	201.	0.10	4.6	5.6	18.0	17.9	1.81	2.39
21.1930	1.8	15.3	14.4	201.	0.10	4.9	5.8	20.5	20.5	0.23	0.40
21.2030	2.0	14.7	14.2	164.	0.10	2.8	4.2	17.0	16.9	2.60	3.19
21.2130	2.0	14.7	14.2	219.	0.20	3.5	3.7	15.8	16.0	3.27	3.85
21.2230	1.7	15.0	14.4	246.	0.20	4.5	3.5	15.7	15.9	3.27	3.92
21.2330	1.7	15.0	14.4	155.	0.20	3.1	3.9	16.4	16.6	2.93	3.43
22.0030	2.0	15.0	14.2	210.	0.	4.5	3.5	16.5	16.7	2.93	3.41
22.0130	2.1	15.0	14.1	100.	0.50	2.6	4.4	16.7	16.7	2.84	3.41
22.0230	2.2	14.7	14.1	82.	0.30	2.6	3.9	16.8	16.7	2.76	3.34
22.0330	2.0	15.1	14.2	201.	0.30	5.7	4.6	16.8	16.7	2.76	3.45
22.0430	2.1	15.0	14.1	146.	0.20	4.5	3.9	16.8	16.7	2.76	3.41
22.0530	2.2	15.0	14.1	146.	0.20	2.6	4.6	16.8	16.7	2.76	3.41
22.0630	2.6	15.0	13.7	128.	0.20	3.1	5.0	16.9	16.7	2.68	3.48
22.0730	2.5	14.7	13.8	137.	0.20	2.9	3.9	17.0	17.0	2.60	3.17
22.0830	2.5	14.7	13.8	109.	0.	5.2	5.0	16.5	16.7	1.81	3.41
22.0930	2.5	14.7	13.8	137.	0.10	1.6	3.9	16.5	16.9	2.76	3.26
22.1030	2.0	15.0	14.2	164.	0.10	1.7	3.1	16.5	16.8	2.76	3.32
22.1130	2.0	15.0	14.2	182.	0.	2.6	3.9	16.5	16.8	2.76	3.32
22.1230	2.5	14.4	13.8	182.	0.	5.2	4.2	16.5	16.9	2.60	3.19
22.1330	2.0	14.4	14.2	310.	0.10	3.1	5.4	16.5	16.8	2.68	3.19
22.1430	2.5	14.7	13.8	301.	0.10	4.3	4.6	16.5	16.9	2.68	3.26
22.1530	2.5	14.7	13.8	283.	0.10	3.3	4.6	16.5	16.8	2.60	3.30
22.1630	2.0	14.7	14.2	192.	0.10	3.3	3.5	16.5	16.8	2.60	3.26
22.1730	2.5	14.4	13.8	219.	0.10	3.8	4.6	16.5	16.8	2.60	3.23
22.1830	2.0	14.5	14.2	228.	0.10	3.3	3.1	16.5	16.7	2.60	3.31

SHUT DOWN AT 22.1830 FOR 23 HOURS

23.1730	2.0	14.7	14.1	319.	0.30	1.0	1.8	16.1	15.6	3.10	4.20
23.1830	2.5	14.7	13.8	155.	0.20	2.0	3.1	16.0	15.9	3.10	4.03

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
23.1930	2.5	14.7	13.8	119.	0.10	2.2	4.2	16.0	15.9	3.10	4.03
23.2030	2.5	14.7	13.8	-	0.20	1.1	1.2	16.0	16.0	3.10	3.94
23.2130	2.5	15.0	13.8	119.	0.50	2.0	2.7	16.0	15.9	2.76	4.12
23.2230	2.3	14.7	13.9	119.	0.40	2.0	3.5	16.0	16.0	3.10	3.94
23.2330	2.8	14.7	13.6	119.	21.00	0.5	1.5	16.0	16.1	2.93	3.94
24.0030	2.6	15.0	13.7	12.	0.10	1.5	4.6	16.0	16.4	2.93	3.73
24.0130	2.6	14.4	13.7	12.	0.10	1.7	3.7	16.5	16.4	2.93	3.58
24.0230	2.3	14.4	13.9	10.	0.10	1.9	4.2	16.5	16.5	2.76	3.43
24.0330	2.5	14.4	13.8	16.	0.	2.5	4.2	16.5	16.6	2.76	3.43
24.0430	2.0	15.0	14.2	16.	0.	2.5	3.9	16.5	16.5	2.76	3.57
24.0530	2.0	15.0	13.9	9.	0.	3.5	3.3	16.5	16.7	2.28	3.39
24.0630	2.0	15.0	14.2	9.	0.	2.2	3.7	16.5	16.8	2.76	3.34
24.0730	2.0	15.0	14.2	9.	0.	3.3	5.0	16.7	16.4	2.60	3.60
24.0830	2.0	15.0	14.2	9.	0.	3.0	4.6	17.0	17.0	2.44	3.14
24.0930	2.3	15.0	14.0	9.	0.	1.3	3.1	17.0	16.9	2.44	3.26
24.1030	4.0	14.7	12.7	9.	0.20	1.4	3.1	17.0	17.3	2.44	3.22
24.1130	1.8	14.7	14.3	9.	0.20	2.5	4.2	16.8	16.7	2.44	3.25
24.1230	2.0	14.7	14.2	9.	0.20	2.8	3.9	17.0	16.3	2.44	3.61
24.1330	2.1	14.7	14.1	9.	0.10	3.1	4.6	16.1	16.3	2.84	3.66
24.1430	2.1	14.7	14.1	55.	0.	2.6	4.2	16.0	16.1	2.93	3.82
24.1530	2.0	15.0	14.2	55.	0.	2.8	4.2	16.0	16.0	2.84	3.90
24.1630	1.8	15.0	14.3	37.	0.40	2.5	3.9	16.0	16.3	2.84	3.64
24.1730	2.0	15.0	14.2	119.	0.20	2.5	3.9	16.0	16.3	2.76	3.66
24.1830	1.8	15.0	14.3	-	0.	2.6	4.2	16.1	16.5	3.82	3.47
24.1930	1.2	15.3	14.8	-	0.	2.5	3.9	16.0	16.2	3.92	3.72
24.2030	1.5	15.0	14.6	-	0.	2.0	3.5	16.0	16.3	3.92	3.64
24.2130	1.6	15.0	14.5	-	0.	1.7	3.1	16.0	16.0	3.82	3.85
24.2230	2.0	13.2	14.2	-	0.	2.2	4.2	16.0	16.0	3.82	3.45

24.2330	2.5	14.4	13.8	55.	0.	1.7	2.7	16.0	16.2	3.82	3.74
25.0030	2.1	14.7	14.1	119.	0.	1.6	2.7	16.2	16.1	3.82	3.82
25.0130	2.0	15.0	14.2	82.	0.	1.7	2.0	16.3	16.0	3.92	3.90
25.0230	2.0	14.7	14.2	151.	0.	1.4	3.1	16.3	16.0	3.82	3.82
25.0330	2.3	14.4	14.0	100.	0.	2.5	4.2	16.3	16.1	3.73	3.74
25.0430	2.3	14.4	14.0	73.	0.	2.2	4.0	16.4	16.1	3.73	3.74
25.0530	2.4	14.4	13.9	64.	0.	2.1	4.0	16.4	16.2	3.73	3.74
25.0630	2.4	14.4	13.9	100.	0.	1.7	4.2	16.4	16.2	3.73	3.74
25.0730	3.0	13.8	13.4	77.	0.	2.5	4.2	16.6	16.3	3.54	3.60
25.0830	2.0	14.7	14.2	137.	0.	1.7	3.7	16.3	16.0	3.73	3.82
25.0930	2.1	14.4	14.1	64.	0.	1.9	3.5	16.3	16.4	3.73	3.47
25.1030	3.1	13.5	13.4	-	0.	2.2	3.9	16.6	16.8	3.54	3.13
25.1130	2.8	13.8	13.6	18.	0.	2.3	4.6	16.7	16.3	3.54	3.57
25.1230	2.2	13.8	14.1	201.	0.	2.2	4.2	16.7	16.3	3.54	3.44
25.1330	2.1	14.4	14.1	100.	0.	3.2	5.0	16.3	16.6	3.73	3.33
25.1430	2.9	14.1	13.5	137.	0.	2.3	3.3	16.5	16.8	3.54	3.27
25.1530	2.8	14.1	13.6	137.	0.	2.5	3.1	16.4	16.8	3.63	3.27
25.1630	2.7	14.1	13.7	173.	0.	2.5	3.7	16.3	16.8	3.63	3.27
25.1730	2.1	14.4	14.1	137.	0.	2.2	4.2	16.3	16.6	3.73	3.33
25.1830	2.6	13.8	13.8	137.	0.	2.3	4.2	16.2	16.5	3.82	3.35
25.1930	2.2	14.4	14.1	192.	0.	2.3	4.2	16.2	16.6	3.82	3.39
25.2030	2.1	14.4	14.1	164.	0.	2.6	3.9	16.2	16.5	3.82	3.39
25.2130	2.1	14.4	14.1	146.	0.	3.1	4.2	16.3	16.3	3.82	3.57
25.2230	2.0	14.4	14.2	121.	0.	2.4	3.7	16.4	16.3	3.73	3.57
25.2330	2.0	14.4	14.2	137.	0.	2.7	3.5	16.1	15.8	3.92	3.89
26.0030	1.9	14.7	14.3	131.	0.	3.3	4.6	15.9	15.8	4.11	3.97
26.0130	1.9	14.4	14.3	100.	0.	3.0	3.9	16.0	16.0	4.02	3.74
26.0230	2.0	14.1	14.2	123.	0.	1.7	2.9	16.0	15.9	4.02	3.74
26.0330	2.2	14.1	14.1	140.	0.	1.7	3.5	16.7	16.6	3.54	3.27
26.0430	1.8	14.4	14.4	164.	0.	1.9	3.5	16.5	16.2	3.63	3.58
26.0530	2.0	14.4	14.2	128.	0.	3.3	4.2	16.5	16.3	3.45	3.58
26.0630	2.0	14.1	14.2	155.	0.	3.0	4.6	16.3	16.3	3.63	3.50
26.0730	1.6	14.4	14.5	146.	0.	2.0	2.2	15.8	15.9	4.11	3.74
26.0830	1.7	14.1	14.4	110.	0.	2.8	3.5	16.1	16.0	3.92	3.67
26.0930	2.0	14.4	14.2	137.	0.	2.3	3.7	16.1	16.2	3.92	3.64
26.1030	2.0	14.4	14.2	137.	0.	2.5	4.2	15.8	15.7	4.11	3.97

RUN 5: GAS COMPOSITIONS

	FLUE GAS				REGENERATOR GAS			GASIFIER INLET GAS			
DAY.HOUR	O2	CO2 VOL %		SO2	O2	CO2	SO2	O2 VOL %		CO2 VOL %	
	%	ANAL	CALC	PPM	%	%	%	ANAL	CALC	ANAL	CALC
26.1130	2.0	14.4	14.2	100.	0.	3.7	3.5	15.8	15.7	4.11	3.97
26.1230	2.0	14.1	14.2	137.	0.	4.3	4.6	16.0	15.9	3.92	3.75
26.1330	2.0	14.1	14.2	137.	0.	2.5	3.9	15.8	16.3	3.82	3.46
26.1430	2.0	14.4	14.2	110.	0.	2.5	3.5	16.0	16.5	3.92	3.39
26.1530	2.0	14.1	14.2	164.	0.	2.6	3.7	16.3	16.5	3.63	3.30
26.1630	2.0	14.4	14.2	201.	0.20	2.5	3.5	16.3	16.6	3.73	3.32
26.1730	1.8	14.4	14.4	201.	0.20	3.1	3.7	16.2	16.6	3.63	3.29
26.1830	2.0	14.1	14.2	228.	0.	2.5	3.9	16.3	16.6	3.63	3.25

APPENDIX B - TABLE VI

RUN 5: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
2.0130	0.139	0.021	0.031	0.007	0.079	19.5	8.8	10.7
2.0230	0.277	0.045	0.080	0.008	0.144	35.7	10.8	24.8
2.0330	0.410	0.078	0.130	0.014	0.188	41.3	18.2	23.1
2.0430	0.543	0.103	0.175	0.023	0.242	41.3	28.5	12.8
2.0530	0.676	0.108	0.241	0.026	0.301	42.7	31.3	11.5
2.0630	0.808	0.111	0.290	0.027	0.380	44.2	32.7	11.4
2.0730	0.941	0.152	0.326	0.032	0.431	45.2	37.2	8.0
2.0830	MISSED DATA READING							
2.0930	1.077	0.223	0.352	0.035	0.466	47.2	40.3	6.9
2.1030	1.220	0.302	0.397	0.037	0.485	50.1	41.9	8.2
2.1130	1.365	0.353	0.432	0.039	0.541	53.4	43.5	9.9
2.1230	1.505	0.403	0.454	0.041	0.608	56.5	45.1	11.4
2.1330	1.645	0.460	0.508	0.043	0.633	60.1	47.1	13.0
2.1430	1.783	0.510	0.582	0.046	0.645	64.2	49.1	15.1
2.1530	1.922	0.556	0.646	0.050	0.671	68.5	52.8	15.7
2.1630	2.062	0.597	0.708	0.053	0.704	73.2	56.3	16.8
2.1730	2.200	0.631	0.766	0.056	0.747	77.5	59.4	18.2
2.1830	2.337	0.665	0.831	0.059	0.781	82.7	60.9	21.8
2.1930	2.473	0.698	0.897	0.069	0.810	87.8	74.6	13.2
2.2030	2.610	0.731	0.922	0.071	0.886	92.4	76.1	16.4
2.2130	2.747	0.763	0.962	0.073	0.950	97.0	77.0	20.0
2.2230	2.883	0.803	1.026	0.074	0.980	101.7	77.9	23.7
2.2330	3.020	0.844	1.094	0.077	1.006	106.8	80.4	26.4
3.0030	3.151	0.882	1.145	0.076	1.047	112.4	80.1	32.4
3.0130	MISSED DATA READING							
3.0230	3.290	0.924	1.226	0.076	1.064	117.6	79.9	37.7
3.0330	3.422	0.962	1.288	0.088	1.084	123.0	86.3	36.7
3.0430	3.556	1.000	1.349	0.093	1.113	127.9	88.7	39.1

3.0530	3.694	1.040	1.402	0.098	1.153	132.5	91.2	41.3
3.0630	3.832	1.077	1.455	0.103	1.196	136.6	93.6	43.0
3.0730	3.969	1.116	1.516	0.109	1.229	141.5	96.1	45.4
3.0830	4.107	1.146	1.576	0.113	1.271	146.3	98.3	48.0
3.0930	4.242	1.174	1.639	0.118	1.311	151.5	100.5	50.9
3.1030	4.378	1.211	1.690	0.123	1.354	155.3	102.8	52.6
3.1130	4.515	1.243	1.718	0.128	1.426	158.7	105.0	53.7
3.1230	4.651	1.277	1.761	0.136	1.477	162.0	108.3	53.7
3.1330	4.789	1.308	1.773	0.144	1.564	165.6	111.5	54.1
3.1430	4.926	1.350	1.838	0.160	1.577	169.8	128.8	41.1
3.1530	5.059	1.387	1.881	0.174	1.617	173.8	144.9	28.9
3.1630	5.191	1.420	1.934	0.186	1.651	177.4	158.5	18.9
3.1730	5.317	1.462	1.975	0.199	1.682	181.8	173.6	8.2
3.1830	5.463	1.511	2.032	0.204	1.714	187.2	175.8	11.4
3.1930	5.592	1.565	2.086	0.210	1.732	191.8	177.9	13.9
3.2030	5.744	1.630	2.102	0.214	1.798	196.7	179.3	17.3
3.2130	5.881	1.688	2.113	0.219	1.862	202.0	180.7	21.3
3.2230	6.018	1.742	2.172	0.223	1.881	205.4	182.1	23.2
3.2330	6.155	1.807	2.192	0.227	1.929	205.4	183.6	21.8
4.0030	6.287	1.880	2.245	0.233	1.928	205.4	185.7	19.6
			STONE CHANGE					
4.0130	6.425	1.921	2.309	0.277	1.918	205.4	253.4	-48.0
4.0230	6.564	1.963	2.390	0.283	1.929	205.4	255.6	-50.2
4.0330	6.696	2.001	2.452	0.289	1.954	205.4	257.8	-52.4
4.0430	6.831	2.039	2.515	0.294	1.983	205.4	260.0	-54.6
4.0530	-	-	-	-	-	234.8	262.3	-27.4
4.0630	6.978	2.077	2.554	0.300	2.047	274.5	264.5	10.0
4.0730	7.121	2.101	2.635	0.305	2.080	306.6	266.8	39.8
4.0830	7.263	2.154	2.710	0.311	2.089	323.2	269.0	54.2
4.0930	-	-	-	-	-	328.8	271.3	57.4
4.1030	7.399	2.186	2.786	0.332	2.096	328.8	291.5	37.3
4.1130	-	-	-	-	-	328.8	293.9	34.9
4.1230	7.538	2.237	2.829	0.347	2.124	343.0	310.7	32.3
4.1330			MISSED DATA READING					
4.1430	7.676	2.243	2.904	0.366	2.163	361.4	335.5	26.0
4.1530	7.816	2.277	2.967	0.377	2.195	380.4	348.6	31.8

RUN 5: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
4.1630	7.955	2.315	3.016	0.388	2.236	398.5	361.8	36.7
4.1730	8.094	2.340	3.053	0.398	2.303	416.3	374.6	41.7
4.1830	8.234	2.348	3.135	0.403	2.348	439.0	376.8	62.2
4.1930	8.375	2.361	3.202	0.416	2.395	463.8	392.3	71.5
4.2030	-	-	-	-	-	486.5	408.5	78.0
4.2130	-	-	-	-	-	503.8	425.9	77.9
4.2230	-	-	-	-	-	521.3	444.5	76.8
4.2330	8.511	2.379	3.275	0.423	2.434	539.6	448.5	91.2
5.0030	8.657	2.396	3.365	0.436	2.461	560.9	467.4	93.5
5.0130	MISSED DATA READING							
5.0230	8.789	2.408	3.389	0.434	2.558	579.3	466.3	113.0
5.0330	8.925	2.410	3.404	0.440	2.672	598.0	478.4	119.5
5.0430	9.072	2.424	3.468	0.467	2.713	617.5	497.9	119.6
5.0530	9.213	2.441	3.521	0.480	2.771	636.2	516.4	119.8
5.0630	9.354	2.450	3.584	0.494	2.826	654.5	534.5	120.0
5.0730	9.493	2.455	3.645	0.506	2.887	669.1	551.6	117.6
5.0830	9.634	2.462	3.711	0.520	2.940	688.1	568.8	119.3
5.0930	9.776	2.471	3.790	0.533	2.982	712.6	584.6	127.9
5.1030	9.917	2.481	3.862	0.549	3.026	732.3	604.7	127.6
5.1130	10.057	2.489	3.937	0.555	3.076	750.9	609.7	141.3
5.1230	10.196	2.501	4.009	0.559	3.128	763.2	613.4	149.8
5.1330	MISSED DATA READING							
5.1430	10.331	2.516	4.085	0.564	3.166	771.9	618.4	153.5
5.1530	10.465	2.526	4.159	0.569	3.211	780.4	623.5	156.9
5.1630	10.599	2.533	4.220	0.572	3.275	790.6	625.5	165.1
SHUT DOWN AT 5.1630 FOR 10 HOURS								
6.0230	10.732	2.565	4.271	0.572	3.324	799.9	625.7	174.3
6.0330	10.865	2.587	4.328	0.572	3.378	814.2	625.9	188.4

6.0430	11.000	2.602	4.370	0.573	3.455	827.9	626.1	201.9
6.0530	11.133	2.611	4.421	0.573	3.528	841.4	626.3	215.1
6.0630	11.266	2.614	4.485	0.573	3.594	853.0	626.5	226.6
6.0730	11.399	2.618	4.546	0.573	3.661	864.4	626.7	237.7
6.0830	11.527	2.621	4.621	0.574	3.711	876.3	626.9	249.5
	SHUT DOWN AT 6.0830 FOR 62 HOURS							
8.2230	-	-	-	-	-	892.1	630.0	262.1
8.2330	11.662	2.655	4.631	0.574	3.802	908.1	630.2	277.9
9.0030	-	-	-	-	-	928.8	630.4	298.4
9.0130	11.801	2.687	4.657	0.574	3.884	950.4	630.6	319.8
9.0230	11.940	2.703	4.684	0.574	3.979	969.7	630.8	338.9
9.0330	-	-	-	-	-	986.9	631.0	355.9
9.0430	12.077	2.720	4.739	0.574	4.043	999.4	631.2	368.2
9.0530	12.214	2.746	4.811	0.575	4.083	1009.3	631.4	377.9
	SHUT DOWN AT 9.0530 FOR 57 HOURS							
11.1430	12.354	2.783	4.852	0.575	4.145	1034.7	631.6	403.1
11.1530	12.501	2.808	4.871	0.575	4.246	1058.9	631.8	427.1
11.1630	12.638	2.832	4.928	0.575	4.303	1069.1	632.0	437.1
11.1730	12.775	2.856	4.963	0.575	4.380	1079.3	632.2	447.1
11.1830	12.911	2.865	5.041	0.575	4.429	1090.4	632.4	458.0
11.1930	13.048	2.871	5.097	0.576	4.504	1101.5	632.6	468.9
11.2030	13.185	2.882	5.180	0.574	4.548	1112.9	631.5	481.4
11.2130	13.321	2.909	5.246	0.574	4.592	1125.7	631.7	494.0
11.2230	13.460	2.931	5.324	0.575	4.630	1138.0	631.9	506.1
11.2330	13.600	2.966	5.402	0.580	4.652	1151.4	642.6	508.8
12.0030	13.743	3.006	5.498	0.581	4.658	1166.8	642.8	524.1
12.0130			MISSED DATA READING					

DAY.HOUR	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
12.0330	13.884	3.035	5.577	0.597	4.675	1175.0	689.2	485.8
12.0430	14.030	3.063	5.674	0.597	4.696	1182.0	689.4	492.6
12.0530	14.171	3.089	5.735	0.597	4.750	1191.1	689.7	501.4
12.0630	14.313	3.129	5.811	0.598	4.776	1201.0	689.9	511.1
12.0730	14.450	3.160	5.893	0.596	4.801	1208.0	688.8	519.1
12.0830	14.584	3.188	5.968	0.596	4.832	1217.9	689.1	528.8
12.0930	14.720	3.213	6.036	0.595	4.875	1231.9	688.0	543.9
12.1030	14.858	3.238	6.095	0.632	4.892	1245.0	740.6	504.4
12.1130	MISSED DATA READING							
12.1230	14.996	3.264	6.148	0.632	4.951	1257.3	741.0	516.2
12.1330	15.134	3.286	6.220	0.662	4.965	1269.2	764.2	505.0
12.1430	15.272	3.308	6.278	0.675	5.011	1281.3	773.9	507.5
12.1530	15.410	3.331	6.357	0.677	5.046	1295.0	775.8	519.2
12.1630	15.549	3.354	6.411	0.679	5.106	1308.2	777.7	530.4
12.1730	15.687	3.376	6.490	0.680	5.141	1318.1	779.7	538.4
12.1830	15.824	3.397	6.553	0.682	5.192	1329.4	781.6	547.9
12.1930	15.961	3.418	6.624	0.684	5.235	1341.7	783.5	558.2
12.2030	16.099	3.442	6.696	0.760	5.201	1352.8	850.7	502.1
12.2130	16.236	3.470	6.755	0.761	5.250	1364.9	852.6	512.2
12.2230	16.377	3.503	6.830	0.770	5.275	1376.4	866.6	509.8
12.2330	16.516	3.533	6.899	0.777	5.307	1386.6	876.2	510.4
13.0030	16.654	3.559	6.964	0.781	5.349	1398.9	882.0	516.9
13.0130	16.790	3.586	7.030	0.786	5.388	1410.5	887.8	522.7
13.0230	16.926	3.607	7.109	0.789	5.421	1420.4	890.8	529.7
13.0330	17.063	3.634	7.178	0.791	5.461	1430.4	893.8	536.6
13.0430	17.200	3.651	7.248	0.794	5.506	1440.4	897.2	543.2
13.0530	17.335	3.666	7.327	0.796	5.547	1448.6	899.7	548.9
13.0630	17.470	3.682	7.375	0.941	5.472	1458.2	1078.0	380.2

SHUT DOWN AT 13.0630 FOR 73 HOURS

16.0730	-	-	-	-	-	1469.6	1080.4	389.2
16.0830	17.606	3.731	7.393	0.943	5.538	1485.9	1083.3	402.6
16.0930	-	-	-	-	-	1502.0	1113.0	388.9
16.1030	17.742	3.768	7.449	0.961	5.564	1516.3	1142.7	373.5
16.1130	17.878	3.802	7.496	0.964	5.616	1529.4	1146.1	383.3
16.1230	18.015	3.846	7.548	0.968	5.653	1547.2	1152.8	394.4
16.1330	18.151	3.881	7.619	0.971	5.681	1565.5	1156.1	409.5
16.1430	MISSED DATA READING							
16.1530	18.286	3.915	7.667	0.973	5.732	1583.0	1159.3	423.7
16.1630	18.421	3.949	7.728	0.974	5.770	1602.0	1161.3	440.7
16.1730	18.557	3.983	7.806	0.976	5.792	1622.4	1164.5	457.9
16.1830	18.692	4.013	7.891	0.978	5.810	1644.9	1167.8	477.1
16.1930	18.827	4.019	7.960	0.981	5.867	1666.7	1171.9	494.8
16.2030	18.962	4.022	8.048	0.985	5.907	1688.0	1176.9	511.1
16.2130	19.104	4.024	8.069	0.988	6.023	1709.3	1181.9	527.5
16.2230	19.246	4.026	8.131	1.019	6.071	1730.9	1203.3	527.6
16.2330	19.383	4.027	8.182	1.023	6.151	1740.8	1209.2	531.6
17.0030	19.521	4.030	8.242	1.027	6.222	1751.3	1215.1	536.2
17.0130	-	-	-	-	-	1761.0	1221.0	539.9
17.0230	-	-	-	-	-	1770.9	1227.0	543.9
17.0330	-	-	-	-	-	1782.5	1233.1	549.4
17.0430	-	-	-	-	-	1791.3	1239.3	552.0
17.0530	-	-	-	-	-	1804.4	1250.4	554.0
17.0630	-	-	-	-	-	1816.1	1255.3	560.7
17.0730	-	-	-	-	-	1828.0	1260.3	567.7
17.0830	MISSED DATA READING							
17.0930	19.659	4.047	8.309	1.029	6.274	1832.5	1264.0	568.5
17.1030	19.798	4.070	8.374	1.033	6.321	1845.9	1268.9	577.0
17.1130	19.937	4.098	8.456	1.036	6.347	1856.9	1273.9	583.0
17.1230	20.077	4.126	8.529	1.039	6.383	1867.7	1278.1	589.6
17.1330	20.216	4.149	8.605	1.042	6.420	1880.5	1282.2	598.3
17.1430	20.361	4.191	8.676	1.045	6.449	1895.1	1286.4	608.7

DAY.HOUR	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
17.1530	20.501	4.230	8.742	1.048	6.481	1915.2	1290.5	624.7
17.1630	20.642	4.250	8.781	1.051	6.560	1935.0	1294.7	640.4
17.1730	20.781	4.289	8.840	1.064	6.588	1949.3	1318.2	631.2
17.1830	20.920	4.315	8.917	1.445	6.243	1961.9	1754.4	207.4
17.1930	21.058	4.338	8.984	1.449	6.288	1973.5	1759.2	214.3
17.2030	21.196	4.350	9.054	1.454	6.339	1984.0	1765.3	218.8
			STONE CHANGE					
	SHUT DOWN AT 17.2030 FOR 72 HOURS							
20.2030	21.343	4.358	9.094	1.459	6.432	2006.6	1771.4	235.3
20.2130	21.493	4.363	9.119	1.466	6.543	2019.7	1782.0	237.7
20.2230	21.648	4.371	9.171	1.480	6.627	2034.9	1794.7	240.2
20.2330	21.806	4.376	9.230	1.484	6.715	2050.3	1802.1	248.2
21.0030	21.964	4.379	9.288	1.496	6.801	2064.9	1813.6	251.3
21.0130	22.112	4.381	9.345	1.500	6.887	2078.5	1820.1	258.5
21.0230	22.272	4.383	9.386	1.504	6.999	2092.9	1826.5	266.4
21.0330	22.435	4.388	9.434	1.509	7.105	2103.7	1833.6	270.1
21.0430	22.596	4.389	9.492	1.513	7.202	2116.0	1840.6	275.4
21.0530	22.757	4.390	9.563	1.519	7.285	2128.3	1851.6	276.8
21.0630	22.917	4.401	9.621	1.522	7.373	2139.1	1855.9	283.2
21.0730	23.077	4.413	9.667	1.525	7.471	2150.7	1860.2	290.4
21.0830	23.240	4.426	9.731	1.529	7.554	2162.2	1867.1	295.2
21.0930	23.397	4.438	9.788	1.545	7.627	2172.0	1886.1	285.9
21.1030	23.557	4.458	9.838	1.550	7.711	2183.0	1893.8	289.2
21.1130	23.717	4.462	9.896	1.555	7.803	2194.6	1901.6	293.0
21.1230	23.877	4.468	9.960	1.557	7.892	2206.4	1904.4	302.0
21.1330	24.037	4.483	10.020	1.598	7.936	2219.5	1935.4	284.0
21.1430	24.197	4.497	10.083	1.600	8.017	2231.8	1938.2	293.6
21.1530	24.358	4.506	10.133	1.607	8.113	2245.1	1949.5	295.6

21.1630	24.520	4.510	10.209	1.609	8.192	2258.2	1953.2	305.0
21.1730	24.681	4.526	10.267	1.622	8.267	2271.6	1967.7	303.9
21.1830	24.841	4.547	10.339	1.625	8.329	2284.7	1971.1	313.6
21.1930	25.001	4.569	10.417	1.627	8.388	2298.8	1974.5	324.3
21.2030	25.150	4.586	10.473	1.630	8.461	2307.8	1977.9	329.8
21.2130	25.291	4.608	10.524	1.633	8.526	2313.7	1981.4	332.3
21.2230	25.434	4.632	10.574	1.635	8.593	2321.4	1984.8	336.6
21.2330	25.576	4.647	10.627	1.638	8.664	2331.7	1988.9	342.8
22.0030	25.717	4.667	10.674	1.642	8.733	2344.0	1993.4	350.6
22.0130	25.859	4.678	10.735	1.645	8.801	2355.8	1997.9	357.9
22.0230	25.999	4.686	10.786	1.649	8.879	2363.8	2002.4	361.3
22.0330	26.140	4.705	10.850	1.653	8.932	2373.3	2006.9	366.3
22.0430	26.284	4.720	10.907	1.656	9.001	2381.0	2011.8	369.2
22.0530	26.429	4.735	10.971	1.659	9.063	2388.7	2017.0	371.7
22.0630	26.568	4.748	11.040	1.663	9.118	2396.9	2022.1	374.7
22.0730	26.705	4.761	11.091	1.665	9.187	2405.1	2026.2	378.9
22.0830	26.850	4.773	11.162	1.668	9.247	2414.3	2029.0	385.3
22.0930	26.991	4.787	11.213	1.674	9.317	2423.6	2034.9	388.6
22.1030	27.132	4.803	11.252	1.678	9.399	2431.8	2042.2	389.6
22.1130	27.272	4.821	11.304	1.683	9.465	2439.2	2050.9	388.4
22.1230	27.413	4.839	11.355	1.686	9.533	2447.4	2054.7	392.7
22.1330	27.557	4.870	11.427	1.689	9.571	2456.2	2058.6	397.6
22.1430	27.697	4.900	11.490	1.692	9.615	2463.6	2062.5	401.2
22.1530	27.837	4.929	11.554	1.698	9.657	2471.3	2068.7	402.6
22.1630	27.978	4.947	11.600	1.701	9.730	2480.6	2072.6	408.0
22.1730	28.118	4.969	11.662	1.736	9.751	2488.3	2138.2	350.1
22.1830	28.258	4.992	11.702	1.739	9.825	2497.3	2141.0	356.3

SHUT DOWN AT 22.1830 FOR 23 HOURS

23.1730	28.393	5.022	11.726	1.742	9.904	2515.5	2143.7	371.7
23.1830	28.532	5.037	11.768	1.750	9.976	2534.2	2149.9	384.4

RUN 5: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
23.1930	28.667	5.049	11.827	1.758	10.033	2553.0	2156.0	397.0
23.2030	-	-	-	-	-	2570.2	2162.2	408.0
23.2130	28.803	5.061	11.862	1.766	10.115	2587.4	2168.4	419.0
23.2230	28.939	5.072	11.907	1.774	10.185	2605.3	2174.5	430.8
23.2330	29.074	5.084	11.930	1.783	10.277	2623.3	2182.3	441.1
24.0030	29.210	5.085	11.990	1.793	10.342	2641.3	2190.0	451.3
24.0130	29.348	5.086	12.037	1.803	10.421	2659.3	2197.7	461.6
24.0230	29.486	5.087	12.093	1.814	10.491	2677.5	2207.1	470.4
24.0330	29.626	5.089	12.148	1.826	10.562	2694.4	2216.5	477.9
24.0430	29.765	5.090	12.199	1.838	10.638	2710.9	2226.0	484.9
24.0530	29.906	5.091	12.243	1.848	10.723	2728.1	2233.9	494.2
24.0630	30.046	5.092	12.292	1.858	10.803	2744.2	2241.8	502.4
24.0730	30.187	5.093	12.361	1.868	10.865	2760.7	2249.7	510.9
24.0830	30.327	5.094	12.422	1.889	10.922	2780.7	2268.1	512.5
24.0930	30.467	5.095	12.460	1.911	11.002	2800.7	2286.5	514.2
24.1030	30.607	5.096	12.499	1.924	11.088	2818.4	2297.0	521.4
24.1130	30.744	5.097	12.556	1.937	11.154	2836.7	2307.5	529.2
24.1230	30.884	5.098	12.608	1.949	11.229	2852.1	2317.2	534.8
24.1330	31.024	5.099	12.672	1.960	11.293	2857.4	2325.8	531.6
24.1430	31.164	5.104	12.731	1.972	11.357	2865.2	2335.7	529.4
24.1530	31.304	5.109	12.787	2.022	11.386	2874.9	2375.9	499.1
24.1630	31.444	5.113	12.841	2.026	11.464	2885.2	2378.9	506.3
24.1730	31.583	5.125	12.894	2.029	11.535	2897.2	2381.9	515.3
24.1830	-	-	-	-	-	2909.1	2385.0	524.1
24.1930	-	-	-	-	-	2922.4	2388.0	534.4
24.2030	-	-	-	-	-	2934.7	2396.1	538.6
24.2130	-	-	-	-	-	2946.8	2406.4	540.4
24.2230	-	-	-	-	-	2960.1	2416.4	543.7

24.2330	31.722	5.130	12.930	2.041	11.620	2971.7	2426.0	545.7
25.0030	31.862	5.142	12.969	2.054	11.697	2985.0	2436.1	549.0
25.0130	32.001	5.150	12.996	2.067	11.788	2998.9	2446.2	552.7
25.0230	32.140	5.164	13.037	2.079	11.859	3011.7	2455.6	556.2
25.0330	32.279	5.174	13.097	2.091	11.917	3025.1	2464.9	560.1
25.0430	32.417	5.181	13.154	2.103	11.978	3035.4	2475.2	560.2
25.0530	32.555	5.188	13.211	2.116	12.040	3046.9	2485.0	561.9
25.0630	32.693	5.198	13.271	2.127	12.097	3059.5	2494.3	565.2
25.0730	32.831	5.205	13.331	2.139	12.155	3070.3	2503.5	566.7
25.0830	32.968	5.219	13.382	2.151	12.216	3082.9	2512.9	570.0
25.0930	33.106	5.225	13.430	2.163	12.288	3094.4	2523.0	571.5
25.1030	-	-	-	-	-	3107.0	2532.9	574.1
25.1130	33.243	5.227	13.496	2.176	12.345	3117.0	2542.8	574.2
25.1230	33.382	5.246	13.555	2.184	12.397	3129.1	2549.1	580.0
25.1330	33.519	5.256	13.625	2.199	12.439	3140.6	2561.8	578.8
25.1430	33.658	5.270	13.667	2.206	12.515	3151.4	2566.6	584.8
25.1530	33.796	5.284	13.709	2.219	12.584	3160.9	2576.8	584.1
25.1630	33.933	5.301	13.760	2.250	12.622	3171.9	2592.9	579.0
25.1730	34.075	5.314	13.820	2.258	12.682	3179.6	2598.3	581.4
25.1830	34.215	5.328	13.879	2.272	12.735	3188.1	2607.5	580.7
25.1930	34.355	5.347	13.937	2.281	12.790	3196.3	2613.1	583.3
25.2030	34.496	5.363	13.990	2.289	12.853	3205.6	2618.7	586.9
25.2130	34.637	5.378	14.048	2.298	12.913	3213.5	2624.3	589.3
25.2230	34.777	5.390	14.098	2.308	12.981	3222.8	2631.8	591.0
25.2330	34.917	5.403	14.144	2.319	13.051	3232.0	2639.5	592.5
26.0030	35.057	5.416	14.207	2.330	13.105	3241.3	2646.7	594.6
26.0130	35.197	5.426	14.260	2.340	13.172	3249.5	2653.8	595.7
26.0230	35.338	5.438	14.299	2.352	13.249	3258.5	2662.8	595.7
26.0330	35.478	5.451	14.347	2.364	13.315	3266.7	2671.5	595.1
26.0430	35.618	5.467	14.396	2.373	13.382	3275.4	2677.8	597.6
26.0530	35.759	5.480	14.458	2.381	13.440	3283.6	2684.1	599.5
26.0630	35.899	5.495	14.525	2.391	13.488	3290.6	2691.9	598.7
26.0730	36.039	5.509	14.555	2.401	13.574	3301.6	2699.7	601.9
26.0830	36.179	5.520	14.605	2.410	13.645	3312.9	2707.1	605.8
26.0930	36.319	5.533	14.656	2.421	13.710	3322.6	2716.1	606.5
26.1030	36.460	5.546	14.716	2.429	13.769	3333.7	2722.2	611.5

RUN 5: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
26.1130	36.601	5.556	14.767	2.437	13.841	3345.0	2728.2	616.8
26.1230	36.741	5.570	14.835	2.445	13.892	3356.5	2734.3	622.3
26.1330	36.881	5.583	14.890	2.455	13.954	3367.3	2741.4	625.9
26.1430	37.021	5.594	14.938	2.465	14.025	3378.9	2748.5	630.4
26.1530	37.161	5.610	14.989	2.475	14.088	3392.2	2755.5	636.7
26.1630	37.302	5.630	15.037	2.488	14.147	3403.3	2765.5	637.7
26.1730	37.442	5.649	15.088	2.503	14.203	3414.0	2775.5	638.5
26.1830	37.582	5.671	15.139	2.525	14.246	3425.1	2790.9	634.2

BYE

0116.38 CRU 0002.84 TCH 0078.97 KC

OFF AT 14:21 03/28/74

APPENDIX B - TABLE VII

ANALYSIS OF SOLIDS REMOVED DURING RUN 5

T O T A L S U L P H U R WT.%

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
3.1530	2.77	1.82	-	7.85	5.15	6.44	4.26
3.2200	2.72	1.81	8.11	7.17	4.99	8.04	8.39
5.1045	3.57	1.59	4.59	4.97	3.34	1.96	-
6.0730	3.72	1.84	4.69	2.34	3.38	1.74	5.24
12.1600	2.92	1.85	4.78	-	3.26	3.26	4.56
12.1800	3.05	1.80	4.59	4.41	3.70	2.96	3.72
13.0400	2.66	2.06	2.14	4.66	3.69	1.68	5.13
13.0600	2.55	1.83	4.21	4.03	3.50	1.75	3.57
17.1130	3.20	1.00	4.24	5.12	2.88	1.44	4.40
17.1800	3.64	-	5.11	4.81	4.03	1.56	4.92
21.0700	2.67	1.78	5.19	5.82	4.27	2.94	6.12
21.1800	2.91	1.75	5.82	5.57	4.52	3.59	7.28
22.0715	2.86	2.21	6.21	6.70	5.16	3.53	7.45
22.1745	2.92	1.91	6.62	6.30	4.62	3.36	7.09
25.0530	2.61	1.88	5.75	6.89	3.85	3.08	7.86
25.1400	2.52	1.72	5.79	6.94	4.08	3.14	6.21
26.0400	2.81	2.17	5.93	4.94	4.10	3.57	6.94
26.1000	2.79	1.96	5.88	2.35	4.30	3.54	6.68
26.1800	2.83	1.96	6.10	6.47	4.90	3.37	7.46

APPENDIX B - TABLE VIII

ANALYSIS OF SOLIDS REMOVED DURING RUN 5

S U L P H A T E   S U L P H U R   WT.%

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
3.1530	0.23	1.27	-	3.85	1.88	2.62	0.32
3.2200	0.38	1.19	4.63	0.25	3.20	1.65	0.17
5.1045	0.40	1.29	3.48	0.32	1.74	1.53	-
6.0730	0.28	1.48	3.40	0.46	2.11	1.57	0.31
12.1600	0.40	1.12	3.47	-	2.58	1.71	0.25
12.1800	0.35	1.18	2.91	0.30	2.93	1.51	0.23
13.0400	0.35	1.33	1.28	0.29	2.44	1.34	0.28
13.0600	0.35	1.28	2.71	0.37	1.96	1.34	0.25
17.1130	0.29	0.80	3.07	0.25	2.56	1.24	0.27
17.1800	0.22	-	3.17	0.29	1.41	1.26	0.28
21.0700	0.36	1.19	3.27	0.35	1.18	1.94	0.24
21.1800	0.19	0.80	2.99	0.36	1.37	1.79	0.21
22.0715	0.48	1.29	4.25	0.71	1.71	2.09	0.48
22.1745	0.43	1.26	4.55	0.66	1.66	2.26	0.44
25.0530	0.39	1.35	3.83	0.20	2.24	2.04	0.26
25.1400	0.44	1.12	3.68	0.55	1.40	2.30	0.41
26.0400	0.35	1.37	4.00	0.32	2.31	2.31	0.27
26.1000	0.39	1.26	3.83	0.36	1.47	2.22	0.32
26.1800	0.34	1.27	4.11	0.39	2.19	2.19	0.36

APPENDIX B - TABLE IX

ANALYSIS OF SOLIDS REMOVED DURING RUN 5

TOTAL CARBON WT.%

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
3.1530	0.	0.	-	19.97	0.32	8.83	1.41
3.2200	0.	0.02	2.81	26.05	0.07	9.09	9.47
5.1045	0.01	0.	0.	1.51	0.12	0.27	-
6.0730	0.05	0.	0.09	3.18	0.17	0.65	1.73
12.1600	0.12	0.04	0.46	-	0.14	0.50	0.92
12.1800	0.10	0.	0.08	2.49	0.	0.52	0.50
13.0400	0.09	0.	-	2.71	0.05	0.45	1.20
13.0600	0.06	0.	0.45	2.91	0.19	0.55	0.38
17.1130	0.08	0.	0.11	0.87	0.	0.15	0.36
17.1800	0.21	0.50	0.02	1.79	0.03	0.24	0.55
21.0700	0.	0.	0.	8.54	0.43	3.21	2.35
21.1800	0.14	0.	0.07	9.19	0.56	2.90	3.91
22.0715	0.	0.	0.20	5.56	0.13	1.61	2.82
22.1745	0.04	0.	0.05	5.78	0.29	0.65	2.67
25.0530	0.	0.	0.04	2.35	0.	0.41	1.49
25.1400	0.02	0.	0.	4.51	0.	0.50	1.26
26.0400	0.02	0.06	0.19	5.85	0.	1.66	1.99
26.1000	0.03	0.	0.07	5.91	0.19	0.	1.73
26.1800	0.02	0.	0.	5.14	0.12	0.89	2.37

APPENDIX B - Table X

Lime Metals Content (Run 5)

Time Day, Hour	Sampling Position	Vanadium ppm	Sodium ppm	Nickel ppm
3.1530	Gasifier Lower	5000	259	477
	Gasifier Upper	4800	296	454
	Regenerator	4200	375	451
3.2200	Gasifier Lower	4900	215	449
	Regenerator	6000	392	653
5.1045	Gasifier Lower	3200	494	454
	Gasifier Upper	4000	522	434
	Regenerator	3400	540	440
6.0730	Gasifier Lower	4400	530	552
	Regenerator	5000	541	622
12.1600	Gasifier Lower	2600	369	376
	Gasifier Upper	2500	361	363
	Regenerator	2000	330	301
12.1800	Gasifier Lower	2800	300	413
	Gasifier Upper	2700	317	337
	Regenerator	2800	317	351
13.0400	Gasifier Lower	3100	466	326
	Gasifier Upper	3200	469	360
	Regenerator	4200	510	470
13.0600	Gasifier Lower	3100	425	349
	Gasifier Upper	3100	492	354
	Regenerator	3600	487	363
17.1130	Gasifier Lower	1700	330	325
	Gasifier Upper	2000	415	280
	Regenerator	1300	300	265
17.1800	Gasifier Lower	2800	415	320
	Gasifier Upper	2300	430	320
	Regenerator	3000	505	360
21.0700	Gasifier Lower	2200	205	255
	Gasifier Upper	1900	185	266
	Regenerator	2500	195	315
21.1800	Gasifier Lower	2900	245	365
	Gasifier Upper	3300	215	420
	Regenerator	3600	255	406
22.0715	Gasifier Lower	4200	295	365
	Gasifier Upper	4700	295	405
	Regenerator	5300	280	470
22.1745	Gasifier Lower	6300	275	545
	Gasifier Upper	5700	320	495
	Regenerator	5300	280	480
25.0530	Gasifier Lower	5100	280	500
	Gasifier Upper	5000	245	470
	Regenerator	5700	235	625
25.1400	Gasifier Lower	5000	345	465
	Gasifier Upper	5800	230	495
	Regenerator	5400	245	475
26.0400	Gasifier Lower	6100	180	540
	Gasifier Upper	6300	155	500
	Regenerator	6500	240	620
26.1800	Gasifier Lower	5700	195	561
	Gasifier Upper	5800	185	520
	Regenerator	5600	245	580

APPENDIX B - TABLE XI

Calcium Oxide and Silica Contents of Bed during Run 5

Sampling Position:	Gasifier Upper		Gasifier Lower		Regenerator		Boiler		Stack Cyclone		Elutriator Fines		Elutriator Coarse		Regenerator Cyclone	
Time Day, Hour	CaO Wt %	SiO_2 Wt %	CaO Wt %	SiO_2 Wt %	CaO Wt %	SiO_2 Wt %	CaO Wt %	SiO_2 Wt %	CaO Wt %	SiO_2 Wt %	CaO Wt %	SiO_2 Wt %	CaO Wt %	SiO_2 Wt %	CaO Wt %	SiO_2 Wt %
5.1045	76.3	15.4	73.6	15.0	75.9	16.3	47.7	11.2	74.6	21.1	63.8	22.4	-	-	62.7	19.1
6.0730	-	-	74.5	17.6	75.4	16.4	66.4	15.5	63.5	16.6	64.3	19.4	61.1	20.6	61.3	18.4
12.1630	75.1	17.2	75.3	17.1	72.8	18.8	69.6	17.6	60.8	18.3	63.5	20.2	70.1	18.5	64.3	18.2
12.1800	74.5	17.2	76.3	16.5	79.5	17.9	67.8	17.7	59.5	18.8	-	-	68.4	18.6	65.8	18.8
13.0400	73.6	16.1	75.5	16.2	-	-	66.0	18.2	63.2	19.8	62.5	20.8	66.0	20.1	-	-
13.0600	76.4	18.4	75.6	18.4	75.5	17.0	69.6	19.4	64.8	20.2	63.2	20.8	74.2	19.0	65.1	19.8
17.1130	74.5	16.1	79.7	14.5	87.2	9.2	71.0	13.9	63.8	19.3	67.6	17.4	72.6	16.2	65.6	17.4
17.1800	75.8	15.3	79.4	13.3	72.8	15.6	70.2	16.3	64.7	19.5	66.3	18.1	71.2	15.9	66.7	18.2
21.0700	86.7	5.0	89.7	4.8	-	-	-	-	-	-	82.5	1.0	-	-	84.7	0.8
21.1800	87.8	4.7	84.7	4.5	-	-	-	-	80.8	1.2	85.0	0.8	-	-	-	-
22.0715	87.4	4.2	85.8	4.9	-	-	-	-	86.8	0.6	-	-	-	-	-	-
22.1745	87.8	3.3	87.1	3.4	-	-	-	-	78.8	0.8	-	-	-	-	-	-
25.0530	-	-	96.3	2.3	96.3	1.9	96.1	1.3	89.6	0.5	94.4	0.8	95.0	1.0	87.2	0.5
25.1430	-	-	97.1	1.6	-	-	-	-	-	-	-	-	-	-	-	-

APPENDIX B - TABLE XII

Sieve Analyses of Gasifier Bed Run 5

Sample No Gasifier Bed	+2800 μ Wt.%	2800 1400 μ Wt.%	1400 1180 μ Wt.%	1880 850 μ Wt.%	850 600 μ Wt.%	600 250 μ Wt.%	250 150 μ Wt.%	150-100μ Wt.%	Less Than 100 μ	Total %	Day/Hr.
51225	2.53	6.33	22.15	33.55	22.87	12.61	0	0	0	100	5.1045
51233	0.44	41.28	13.02	19.87	14.13	11.04	0.22	0	0	100	6.0730
51236	1.37	45.21	15.07	20.55	10.96	6.85	0	0	0	100	12.1600
51245	1.36	45.92	13.27	20.41	12.93	6.11	0	0	0	100	12.1800
50001	1.70	48.02	13.35	19.60	11.65	5.68	0	0	0	100	13.0400
50029	3.53	49.17	13.07	17.63	10.17	6.22	0.21	0	0	100	17.1800
50047	0.28	34.27	11.89	18.88	15.66	19.02	0	0	0	100	18.1100
50048	0.22	28.40	12.12	20.99	16.95	21.32	0	0	0	100	21.1600
50055	0.68	41.91	11.78	18.08	12.74	14.66	0.14	0	0	100	21.1800
50061	0.24	38.59	12.86	18.20	13.59	16.28	0.24	0	0	100	22.0730
50073	0.72	41.17	11.69	18.38	13.96	14.08	0	0	0	100	22.1745
50114	0.66	42.90	13.20	19.64	12.87	10.40	0.17	0.17	0	100	25.0530
50123	0.64	44.16	13.25	19.87	13.56	8.52	0	0	0	100	25.1400
50133	0.63	47.79	14.11	18.95	12.21	6.32	0	0	0	100	26.0400
50149	0.95	44.62	14.24	19.94	13.29	6.96	0	0	0	100	26.1000

APPENDIX B - TABLE XIII

Sieve Analyses of Regenerator Bed (Run 5)

Sample No Regen Bed Run 5	+2800 µ Wt.%	2800 1400 Wt.%	1400 1180 Wt.%	1180 850 Wt.%	850 600 Wt.%	600 250 Wt.%	250 150 Wt.%	150 100 Wt.%	-100 µ Wt.%	Total Wt.%	Day/Hr.
51197/73	0.64	39.30	10.22	18.53	16.29	15.02	0	0	0	100	2.1730
51198/73	0.33	31.46	8.94	20.2	18.54	20.53	0	0	0	100	2.1930
51232/73	1.15	41.67	12.93	22.41	14.66	7.18	0	0	0	100	6.0730
51239/73	0.64	47.92	13.42	18.85	12.14	7.03	0	0	0	100	12.1600
50008/73	2.14	45.99	11.23	18.72	13.36	8.56	0	0	0	100	13.0600
50124/73	0.46	40.23	13.79	20.23	14.25	10.80	0.23	0	0	100	15.1400
50025/73	1.44	38.46	10.58	16.83	12.50	20.19	0	0	0	100	17.1130
50036/73	0.43	17.93	7.34	17.06	18.79	27.65	2.16	1.73	6.91	100	17.1800
50044/73	0.44	34.37	9.31	22.62	20.40	12.42	0.44	0	0	100	21.0700
50051/73	0.21	27.66	10.64	17.98	16.49	25.85	1.06	0.11	0	100	21.1800
50059/73	0.43	32.83	11.59	18.67	15.45	20.82	0.21	0	0	100	22.0730
50071/73	0.36	28.83	10.85	18.74	16.94	24.14	0.18	0	0	100	22.1745
50012/73	0.31	35.64	12.75	20.28	15.67	14.90	0.15	0.15	0.15	100	25.0530
50131/73	0.57	40.92	13.58	20.84	14.15	9.75	0.19	0	0	100	26.0400
50150/73	0.74	41.98	13.09	19.26	16.79	7.9	0.25	0	0	100	26.1000

APPENDIX B - TABLE XIV

Composition of CAFB Solids Run 5 (Ignition at 900°C)

Sampling Position:	Boiler		Stack Cyclone		Elutriator Coarse		Elutriator Fines		Regenerator Cyclone	
Time Day, Hour	Loss Wt %	Gain Wt %	Loss Wt %	Gain Wt %	Loss Wt %	Gain Wt %	Loss Wt %	Gain Wt %	Loss Wt %	Gain Wt %
3.1530	-	3.39	7.91	-	0	0	11.54	-	-	-
3.2200	1.03	-	20.46	-	-	5.53	21.2	-	-	0.80
5.1045	-	0.47	15.62	-	-	-	-	4.69	-	0.86
6.0730	0.24	-	3.56	-	-	5.67	-	5.45	-	1.24
12.1600	0.10	-	4.77	-	-	6.78	-	4.14	-	1.13
12.1800	-	0.47	7.20	-	-	5.97	-	-	-	1.65
13.0400	-	1.40	3.91	-	-	6.55	-	3.90	-	1.07
13.0600	-	1.90	1.75	-	-	5.34	-	2.82	-	1.46
17.1130	1.18	-	2.22	-	-	6.52	-	7.27	-	1.18
17.1800	-	3.07	2.89	-	-	5.97	-	6.10	-	1.87
21.0700	-	2.53	5.32	-	-	7.22	0.39	-	-	1.33
21.1800	-	1.68	4.12	-	-	5.13	0.88	-	-	2.26
22.0715	-	1.64	3.19	-	-	8.00	-	5.70	-	1.35
22.1745	-	0.75	3.57	-	-	7.97	-	3.50	-	1.46
25.0530	-	1.32	3.30	-	-	10.82	-	6.11	-	1.52
25.1400	-	2.31	3.46	-	-	8.39	-	6.72	-	1.53
26.0400	-	1.49	5.24	-	-	8.79	-	4.47	-	1.40
26.1000	-	2.18	4.95	-	-	7.78	-	3.14	-	1.44

APPENDIX B - TABLE XV
SOLIDS REMOVED DURING RUN 5, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
0.2220	-	-	4.31	-	-	6.94	-
0.2335	-	-	-	-	-	1.59	-
1.0130	-	-	0.91	-	-	3.18	-
1.0330	-	-	0.45	-	-	1.59	-
1.0530	-	-	0.45	-	-	1.47	-
1.0730	-	-	0.45	-	-	1.13	-
1.0930	-	-	0.45	-	-	0.91	-
1.1230	-	-	0.45	-	-	1.02	-
1.1400	-	-	0.45	-	-	4.54	-
1.1600	-	-	1.13	-	-	-	-
1.2100	-	-	3.40	-	-	7.26	-
1.2200	-	-	4.99	-	0.34	1.36	-
1.2230	-	-	-	-	-	1.13	28.58
1.2330	-	-	1.81	-	-	0.45	-
2.0030	-	-	3.86	-	-	-	-
2.0130	-	-	-	-	-	3.86	-
2.0230	-	-	3.63	-	-	0.45	-
2.0330	-	-	7.71	-	-	0.23	-
2.0430	-	-	1.13	-	-	9.98	-
2.0530	-	-	0.34	-	-	-	-
2.0630	-	-	0.34	-	-	5.44	-
2.0730	-	-	1.59	-	-	-	-
2.0830	-	-	0.91	-	-	3.63	-
2.0930	-	-	0.91	-	-	1.36	-
2.1230	-	-	1.70	-	-	2.72	-
2.1630	-	-	2.61	-	-	5.22	-
2.1730	-	1.81	-	-	-	-	-
2.1830	-	1.59	-	-	-	-	-
2.1930	-	1.59	-	-	-	-	-
2.2030	-	-	0.57	3.29	-	5.22	-
2.2130	-	12.70	-	-	-	-	-
2.2330	-	-	0.23	1.13	-	2.04	-
3.0230	-	2.49	0.34	-	-	-	-
3.0330	-	-	-	2.72	0.	4.76	-
3.0730	-	-	0.68	3.63	0.34	7.26	-
3.1130	-	-	0.45	2.83	1.81	5.62	-
3.1430	-	-	0.68	3.63	1.81	5.90	-
3.1630	-	14.51	-	-	-	-	-
3.1730	-	14.51	-	-	-	-	-
3.1830	-	11.79	-	-	-	-	-

SOLIDS REMOVED DURING RUN 5, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
3.1930	-	13.38	0.91	3.18	3.63	5.44	-
3.2330	-	-	0.57	4.08	0.11	2.61	-
4.0330	-	67.70	-	-	-	-	-
4.1030	-	-	2.27	10.21	0.45	16.78	-
4.1230	-	14.74	-	-	-	-	-
4.1330	-	14.97	-	-	-	-	-
4.1430	-	-	1.36	6.58	0.23	9.07	-
4.1530	-	13.61	-	-	-	-	-
4.1630	-	21.77	-	-	-	-	-
4.1730	-	11.34	-	-	-	-	-
4.1830	-	11.34	1.36	6.80	0.23	2.04	-
4.1930	-	10.89	-	-	-	-	-
4.2130	-	12.02	-	-	-	-	-
4.2230	-	12.70	-	-	-	-	-
4.2330	-	13.88	-	-	-	-	-
5.0030	-	15.20	1.81	9.07	0.23	15.31	-
5.0130	-	14.83	-	-	-	-	-
5.0330	-	16.33	-	-	-	-	-
5.0430	-	-	1.19	7.31	0.23	12.38	-
5.0530	-	13.15	-	-	-	-	-
5.0730	-	14.29	0.95	6.40	0.23	7.76	-
5.0830	-	13.83	-	-	-	-	-
5.0930	-	12.70	0.91	5.44	-	8.62	-
5.1030	-	11.34	-	-	-	-	-
5.1050	-	10.43	-	-	-	-	-
5.1130	-	9.07	-	-	-	-	-
5.1230	-	10.43	-	-	-	-	-
5.1530	-	-	1.36	12.70	0.45	17.24	-
5.2200	-	-	-	-	0.91	-	-
6.0030	-	-	-	7.48	-	-	-
7.0430	-	11.79	0.79	-	-	-	-
8.0100	-	-	0.45	4.31	-	-	-
8.0715	-	-	0.23	-	-	-	-
8.2030	-	-	-	0.68	0.11	6.58	-
8.2130	-	-	-	-	-	5.22	-
9.0030	-	-	-	-	-51.82	-	4.10
9.0430	-	-	-	-	-	6.35	-
9.0530	-	-	-	28.80	-	-	-
9.0700	-	-	0.23	6.58	-	1.93	-
10.1930	-	-	-	11.85	-	-	-

SOLIDS REMOVED DURING RUN 5, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
11.0930	-	1.13	-	-	-	7.31	-
11.2230	-	-	1.59	-	0.68	-	-
12.0030	-	-	-	-	-	18.14	-
12.0130	-	12.70	-	-	-	-	-
12.0930	-	-	-	4.31	-	-	-
12.1030	-	-	3.18	-	2.38	50.12	-
12.1230	-	-	-	-	-	1.47	-
12.1330	-	-	-	-	-	7.26	-
12.1430	-	-	1.47	3.06	2.61	-	-
12.1530	-	-	-	-	-	-	13.61
12.1630	-	-	-	-	-53.98	-	5.53
12.2230	66.90	-	-	-	-	-	-
12.2330	-	-	1.36	3.18	4.08	10.89	-
13.0030	-	12.47	-	-	-	-	-
13.0130	-	7.94	0.68	-	2.49	8.85	-
13.0400	-	-	0.91	0.68	0.91	5.67	-
13.0600	-	-	0.45	0.45	1.81	5.44	-
13.1230	-	-	1.36	2.27	2.49	19.96	-
13.1700	-	-	-	-	2.61	18.48	-
14.0630	-	-	2.27	2.38	3.63	27.22	-
14.1100	-	-	-	-	4.31	6.58	-
14.1630	-	-	1.36	-	2.04	7.26	-
14.1900	-	-	0.45	0.45	-	-	6.45
15.0330	-	-	0.91	0.45	7.26	13.61	-
15.0730	-	-	-	-	-	9.98	-
15.1230	-	-	-	-	-	10.43	-
15.1730	-	-	-	-	0.45	11.79	-
16.0005	-	-	4.99	0.23	0.11	3.29	-
16.0800	-	-	0.91	0.45	0.11	20.07	-
16.1140	-	27.22	-	-	-	-	-
16.1230	-	27.22	0.91	0.68	3.63	10.89	-
16.1900	-	-	0.91	2.72	0.45	18.14	-
16.2230	-	-	1.81	2.04	0.79	13.61	-
17.0010	-	-	-	-	-	-	17.24
17.0215	-	-	1.47	2.83	2.83	16.10	-
17.0530	-	-	0.91	1.70	3.29	15.03	-
17.0830	-	-	1.02	1.02	5.10	13.61	-
17.1130	-	-	-	-	-	11.79	-
17.1830	-	-	1.36	1.02	8.16	19.96	-
17.1930	-	19.96	-	-	-	-	-

SOLIDS REMOVED DURING RUN 5, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
17.2100	-	-	1.42	7.17	7.60	14.63	-
18.0100	97.98	22.00	-	21.32	-	-	-
18.0430	-	-	0.45	-	2.72	6.71	-
18.0840	-	-	-	-	-	-	12.16
18.1130	-	-	0.23	0.91	3.18	4.54	-
18.1530	-	-	-	0.23	1.81	6.35	-
18.2115	-	-	-	-	1.13	-	-
18.2345	-	-	-	-	-	1.59	-
19.0100	-	-	-	32.21	-	-	-
19.0815	-	-	-	1.13	2.04	0.68	-
19.1330	-	-	-	-	5.67	-	-
19.2030	-	-	-	0.11	2.27	3.18	-
20.0200	-	-	0.23	-	5.78	22.23	-
20.0500	-	-	0.11	0.45	3.86	7.03	-
20.0715	-	-	0.11	-	3.29	8.98	-
20.1400	-	-	-	-	180.53	76.20	-
20.1515	-	-	0.45	0.23	0.45	8.16	-
20.1730	-	-	0.09	0.11	0.06	1.81	-
20.1910	-	-	0.28	0.23	0.23	16.33	-
20.2030	-	-	-	-	-	9.53	-
20.2130	-	-	0.91	1.36	-	8.85	-
20.2330	-	-	1.36	0.45	-	13.72	-
21.0030	-	-	-	-	-	6.58	5.44
21.0230	-	-	2.04	0.68	-	11.57	4.54
21.0430	-	-	-	-	-	13.38	-
21.0530	-	-	1.81	-	-	10.66	-
21.0730	-	-	0.68	0.45	0.23	8.16	-
21.0930	-	-	1.36	0.23	1.13	11.57	-
21.1130	-	-	-	1.02	-	13.15	-
21.1155	12.47	-	-	-	-	-	-
21.1430	-	-	2.27	0.45	2.72	5.44	-
21.1530	-	-	-	-	-	10.32	19.05
21.1600	10.21	-	-	-	-	-	-
21.1630	-	-	0.68	1.36	-	-	-
21.1930	11.79	-	-	-	-	-	-
21.2000	11.34	-	-	-	-	-	-
21.2230	-	-	-	-	3.18	-	-
21.2300	-	-	2.72	1.81	-	18.60	-
22.0400	-	-	2.72	2.27	4.08	14.97	-
22.0700	-	-	-	-	-	12.70	-

SOLIDS REMOVED DURING RUN 5, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
22.0945	-	-	-	-	-	4.99	-
22.1030	-	-	-	1.13	-	-	-
22.1130	-	-	2.04	-	5.44	13.78	3.18
22.1730	-	-	2.49	1.36	4.54	16.10	2.49
23.0300	-	-	-	-	-	43.54	-
23.0730	-	-	-	-	-	-	19.50
23.0800	-	-	1.13	0.79	1.81	11.79	19.96
23.1730	-	-	0.68	-	1.02	11.11	-
23.1930	-	-	-	3.40	-	-	-
23.2230	-	-	3.40	5.35	3.18	17.46	-
24.0130	-	-	3.86	2.04	4.54	14.29	-
24.0430	-	-	2.83	5.90	3.29	18.14	-
24.0730	-	-	2.04	4.08	4.76	14.42	-
24.0930	-	-	2.38	4.42	3.86	28.58	-
24.1130	-	-	-	-	-	12.59	-
24.1230	-	-	2.27	4.99	7.26	-	-
24.1330	-	-	-	-	-	11.11	-
24.1530	-	-	1.47	4.65	4.76	13.83	-
24.1730	-	-	1.36	4.76	5.90	10.43	-
24.1930	-	-	-	-	4.08	12.59	-
24.2030	-	-	1.59	5.22	-	-	-
24.2130	-	-	-	-	-	10.89	-
24.2230	-	-	1.81	4.54	7.03	-	-
24.2330	-	-	-	-	-	10.43	-
25.0030	-	-	1.36	4.54	-	-	-
25.0130	-	-	-	-	5.90	11.79	-
25.0230	-	-	1.36	4.54	-	-	-
25.0330	-	-	-	-	4.54	9.53	-
25.0430	-	-	1.36	4.54	-	-	-
25.0530	-	-	-	-	4.99	10.89	-
25.0630	-	-	1.13	4.08	-	-	-
25.0730	-	-	-	-	4.99	9.53	-
25.0830	-	-	1.13	4.08	-	-	-
25.0930	-	-	0.91	2.49	2.27	12.47	-
25.1130	-	-	-	-	-	12.70	-
25.1230	-	-	2.04	5.90	4.76	2.49	-
25.1330	-	-	-	-	-	10.43	-
25.1430	-	-	-	-	-	2.04	-
25.1530	-	-	1.02	3.63	4.88	7.71	-
25.1730	-	-	-	-	0.45	6.35	-

SOLIDS REMOVED DURING RUN 5, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
25.1830	-	-	1.59	5.56	5.90	-	11.43
25.2130	-	-	-	-	7.26	6.35	-
25.2230	-	-	1.81	6.35	-	-	-
25.2330	-	-	-	-	5.90	6.35	-
26.0030	-	-	1.13	3.18	-	-	-
26.0130	-	-	-	-	4.08	7.26	-
26.0230	-	-	1.13	2.95	-	-	-
26.0330	-	-	-	-	4.54	10.89	-
26.0430	-	-	0.91	2.72	-	-	-
26.0530	-	-	-	-	3.18	6.80	-
26.0630	-	-	0.91	2.72	-	-	-
26.0730	-	-	-	-	4.08	9.07	-
26.0830	-	-	0.91	2.72	-	-	-
26.0930	-	-	0.68	2.72	2.04	10.43	-
26.1230	-	-	1.81	2.49	5.22	9.98	-
26.1530	-	-	2.04	3.63	5.67	11.57	-
26.1800	-	-	1.59	3.40	4.99	17.01	-
26.1830	-	-	1.13	1.81	-	5.81	-

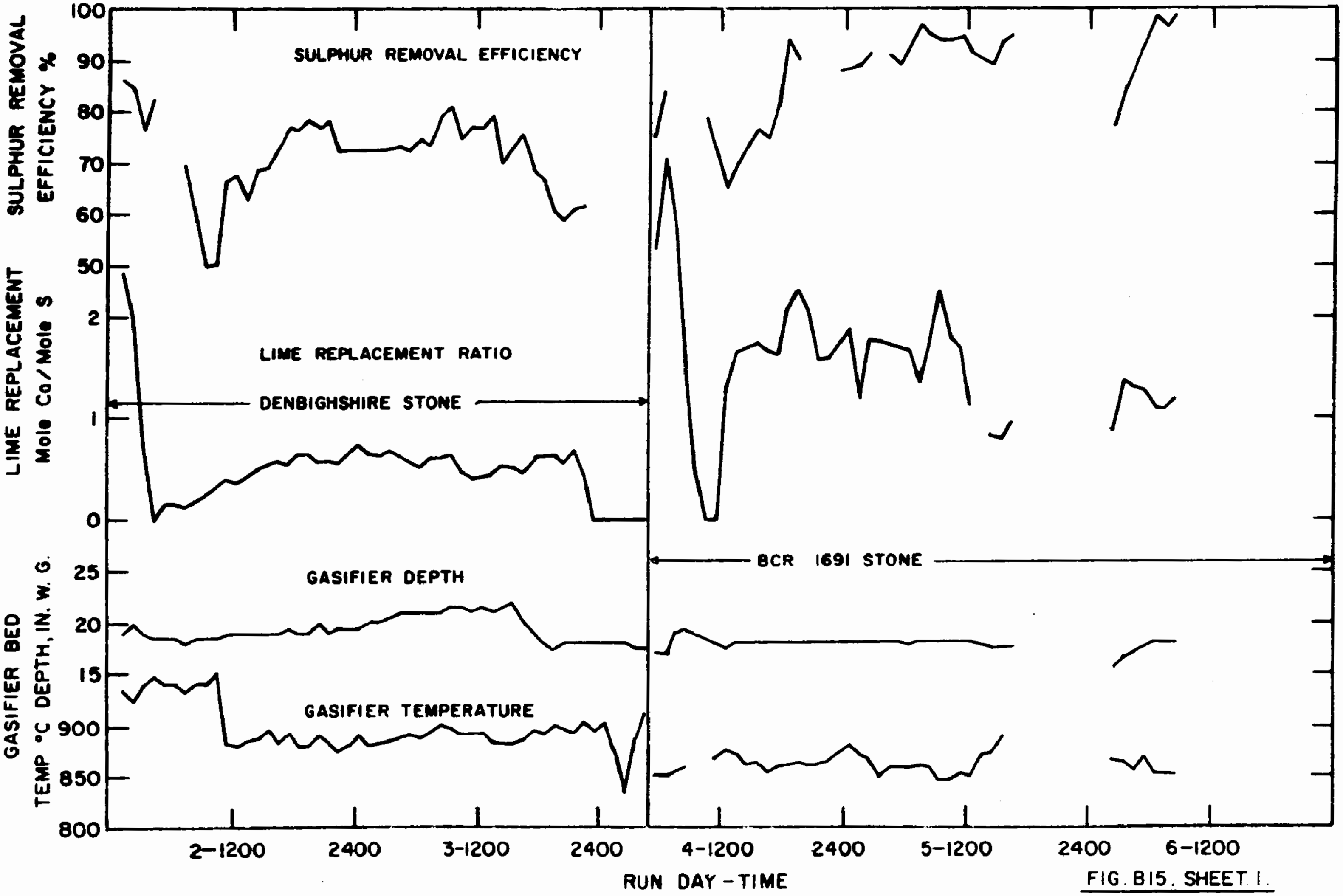

FIG. B15. SHEET 1.

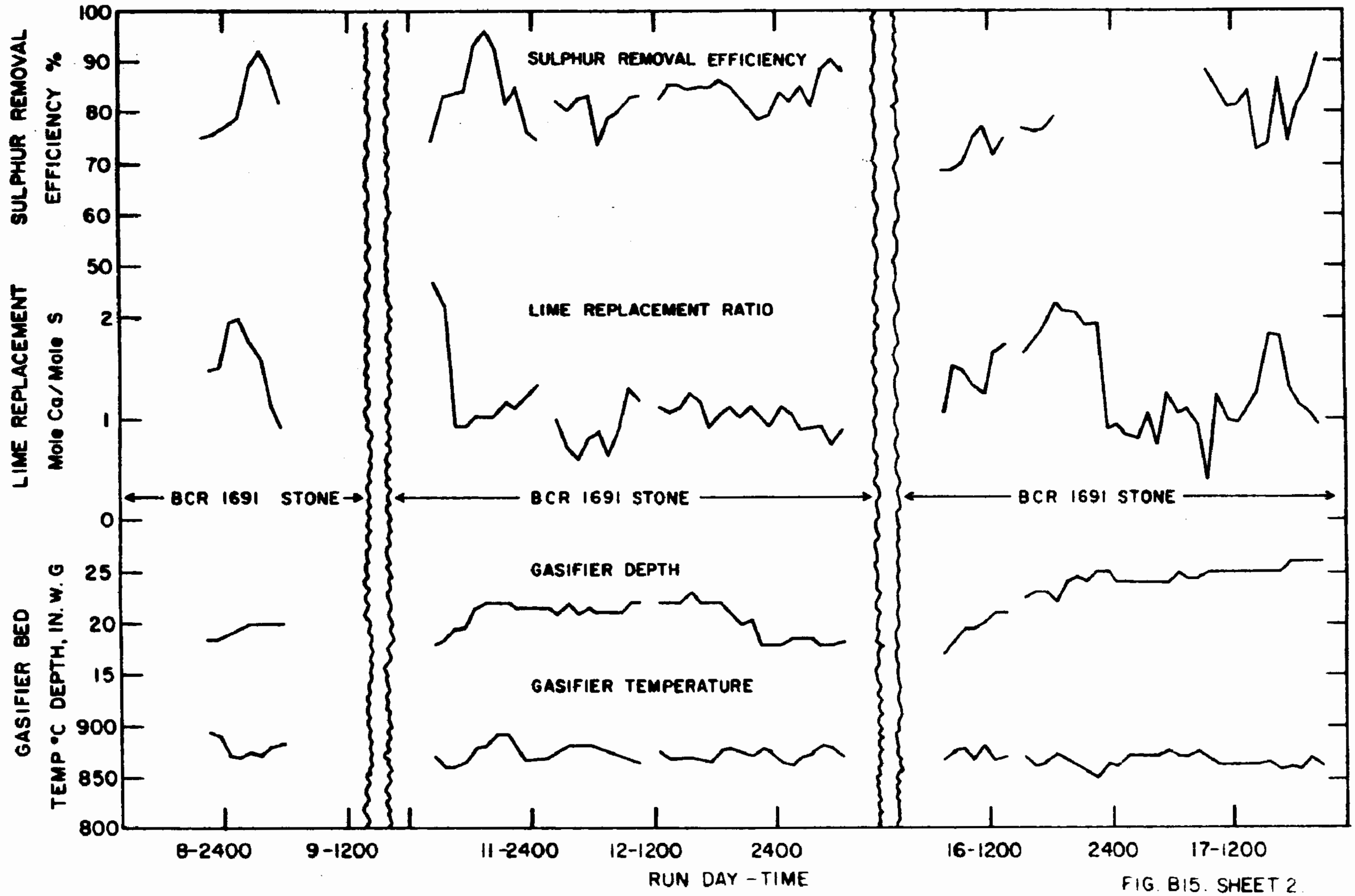

FIG. B15. SHEET 2

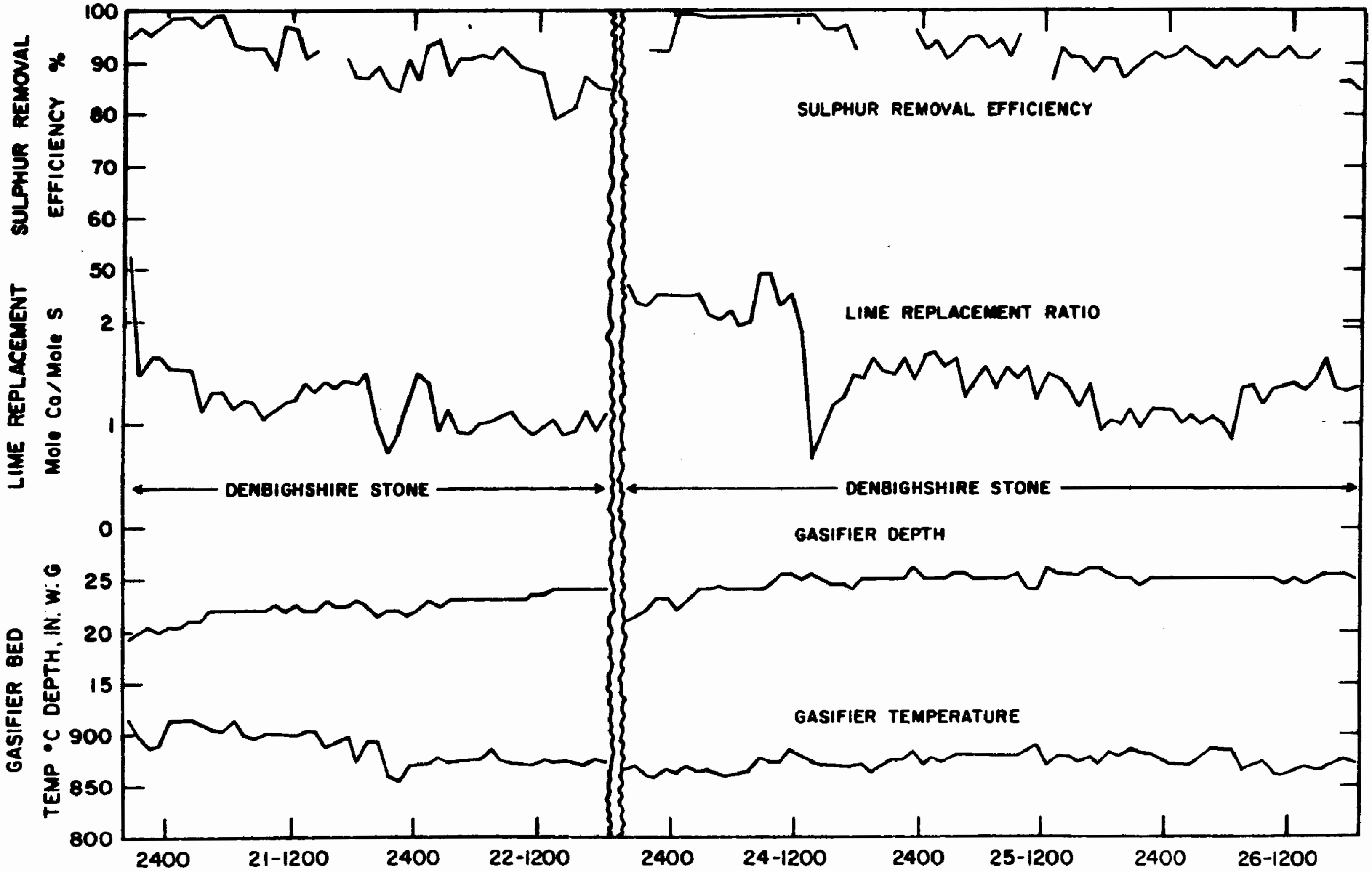

FIG. B15. SHEET.3.

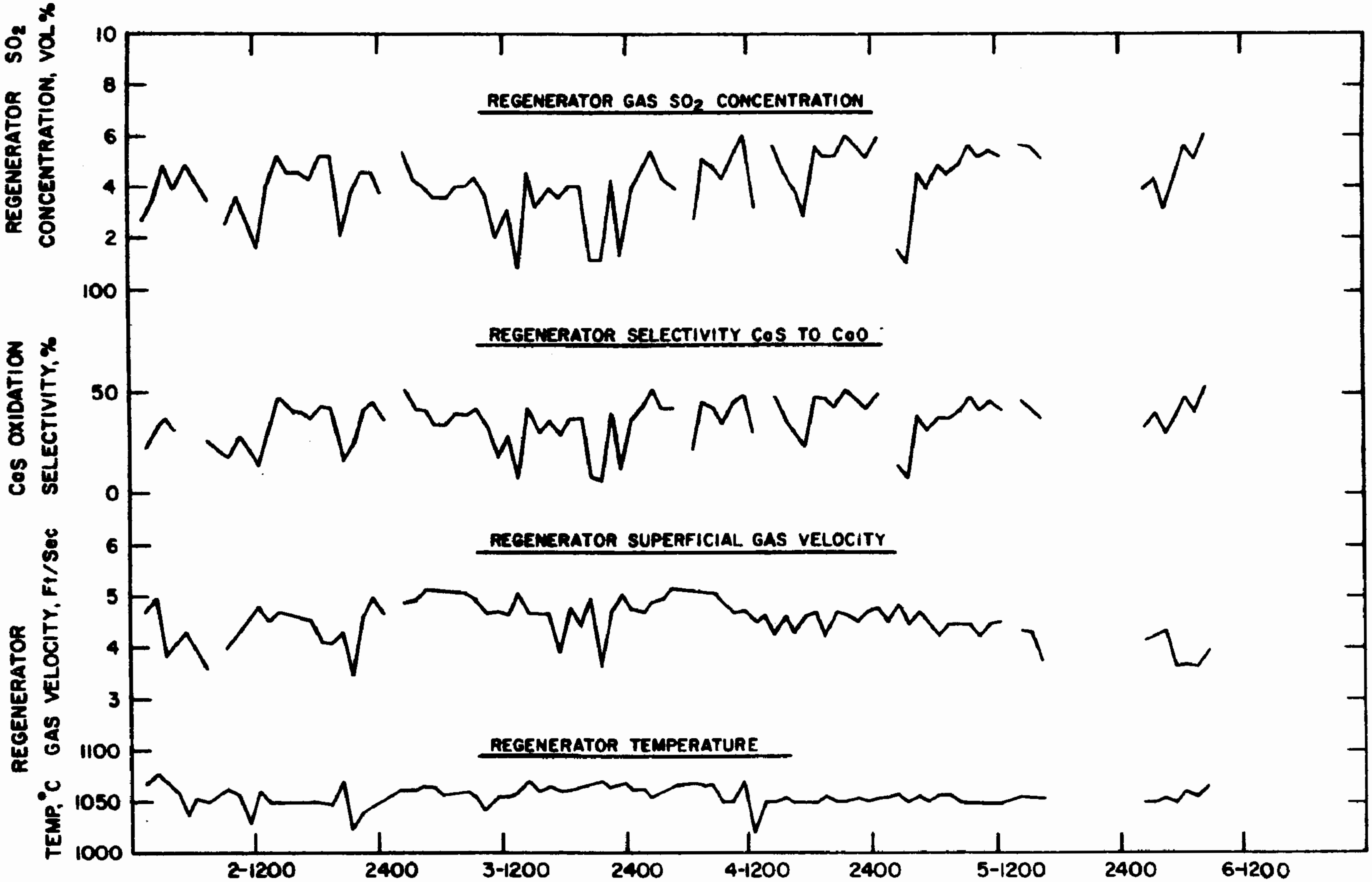

FIG. B16. SHEET. 1

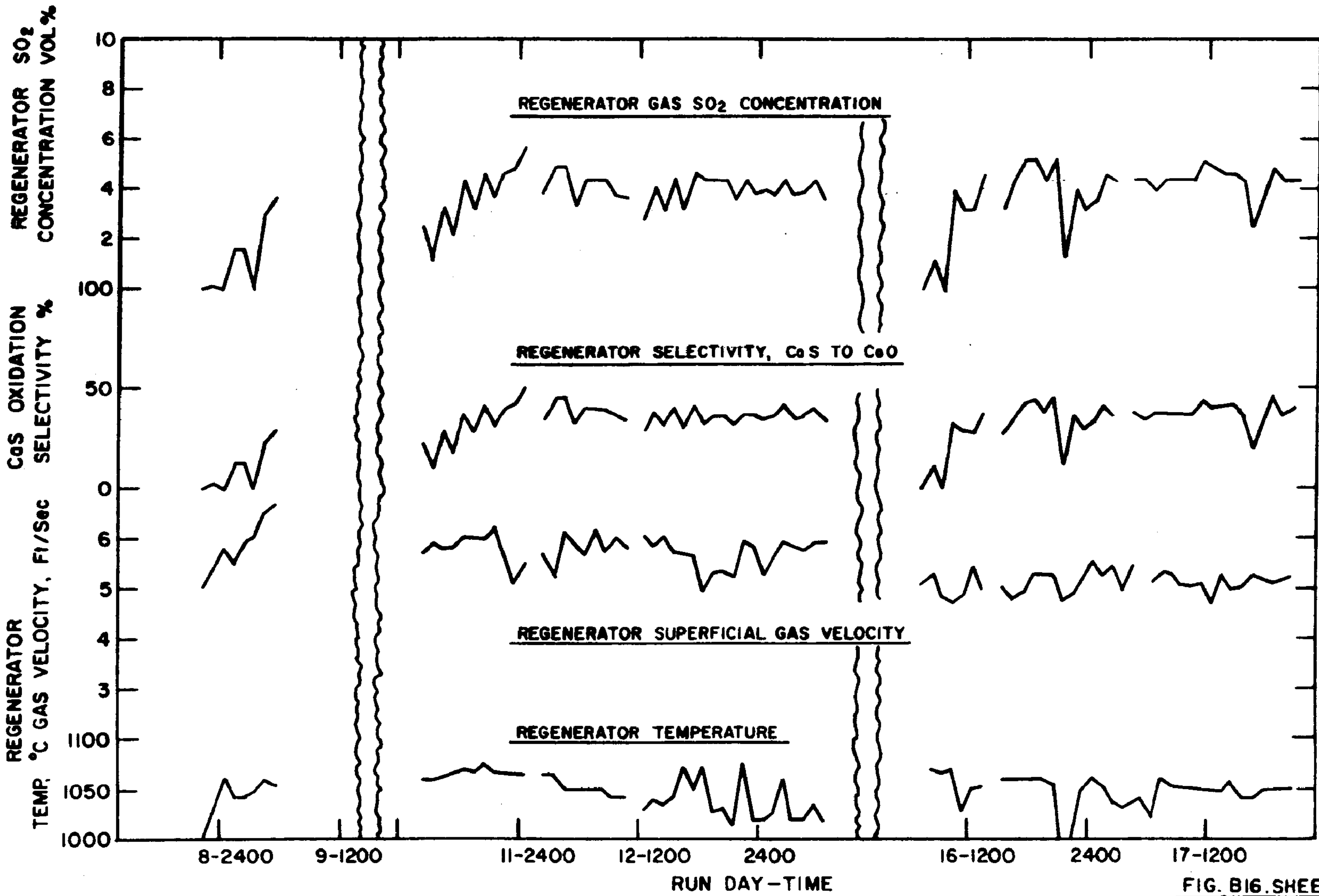

FIG. B16. SHEET. 2

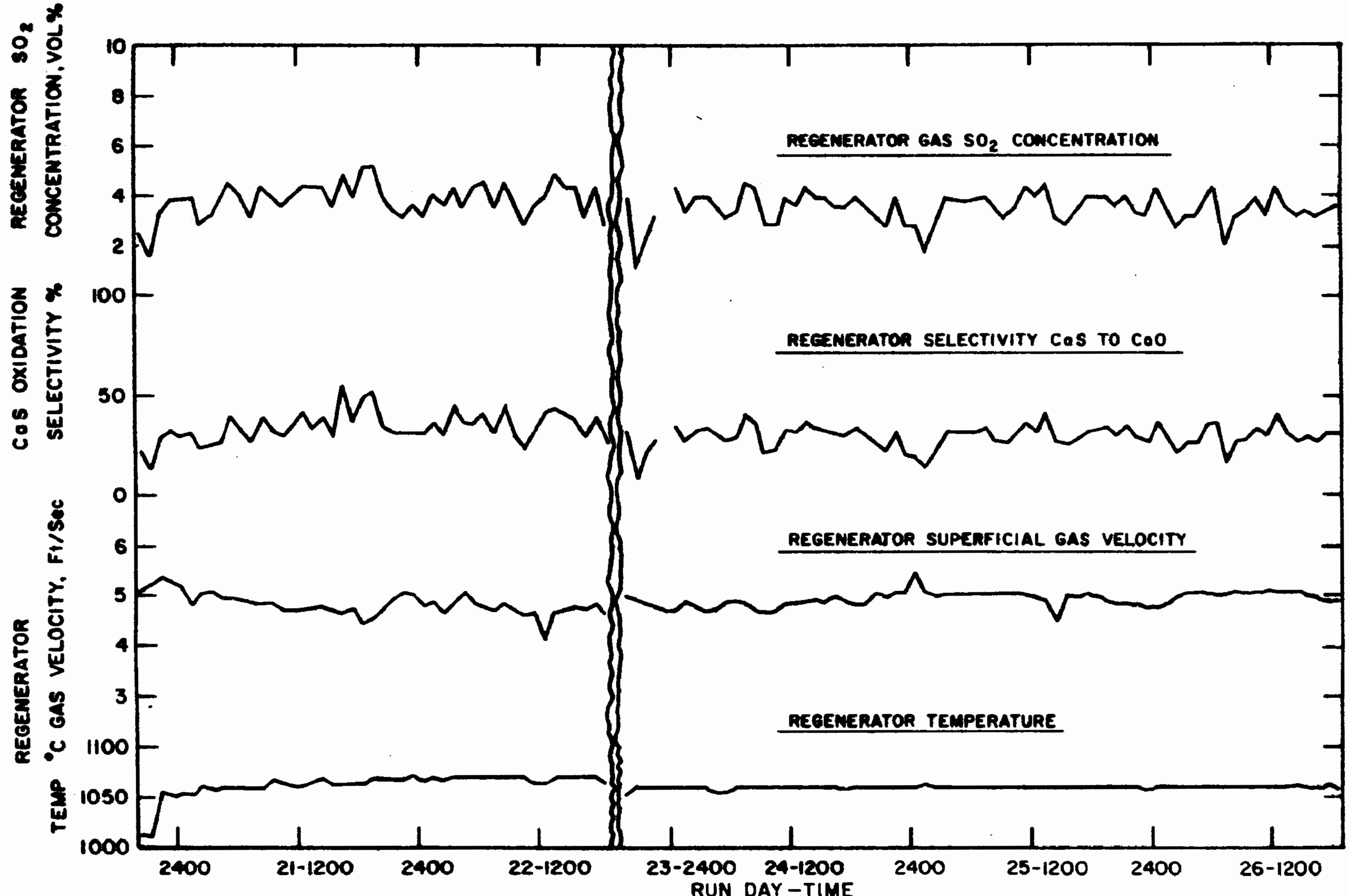

FIG. B16, SHEET 3

APPENDIX C

RUN 6

Operational Log, Inspection, and Data

APPENDIX C

CAFB RUN 6

OPERATIONAL LOG

3.6.73 to 5.6.73 (Unit warm up)

The scheduled start up was delayed a few hours due to difficulty in starting the gas burner which was traced to deposits in the gas pilot venturi which was stripped out, cleaned and replaced. Warm up started at 07.00 and continued uneventfully at 12°C per hour until a gas space temperature of 500°C was reached when after a brief holding period the rate was increased to 20°C per hour. Kerosene was added at 11.00 on 5.6.73 to bring the temperature near the target of 850°C and the joint around the gasifier lid was sealed with fireclay and asbestos. At 16.30 the stone feed was started using material retained from the end of Run 5 and apart from some gas pilot flame outs the feeding continued steadily. At 20.45 BCR 1359 limestone feed was started and the bed depth continued to rise slowly with a feed rate of 62 kgs/h (135 lbs/h).

6.6.73 (Day 1 of Gasification)

The regenerator was well fluidised and bed transfer was good but the gasifier bed temperatures were spread by 25°C and showed some general instability possibly due to erratic fuel input. The gasifier lid seal was leaking in some areas and was repacked with fireclay and asbestos rope. The automatic valve for controlling the fines return into the gasifier was not operating and the pressure tapping which controls this operation was found to be blocked with fine material.

At 13.30 the boiler door was shut, the test probe inserted and combusting conditions resumed to check out bed transfer and other features prior to gasification. Some problems were found with the flame detector sensors but apart from an intermittent fault in the pilot flame failure repeater light all the systems were working.

Gasification started at 21.10 and almost immediately the persistant lid leakage stopped due to the carbon deposition. The new boiler gas sampling system was installed using a small cyclone on the rear end of the boiler and sampling the gas from the stream through the cyclone. Initial SO_2 levels in the boiler were approximately 240 ppm with a regenerator SO_2 output of 8% at 1045°C.

7.6.73 (Day 2)

There were some problems with the main flame and pilot flame failure warnings which persisted even though both flames were well established. The unit ran very steadily with consistent boiler and regenerator SO_2 levels. A system of air injection into each leg of the bifurcated duct had been installed so that differences in gas flow in each duct might be measured by the differences in the temperature rise with a given air injection rate. Initial trials were made on the left hand duct and although sharp temperature increases of 10°C were observed with 18 m^3/h (10.6 ft^3/m) of air injection the results were not repeatable and it was apparent that the gas burnt in an irregular manner because one thermocouple further downstream registered a 60°C temperature rise for the same flow rate of air.

Bed transfer between the gasifier and regenerator was very good and it was necessary to reduce the bleed rate and pulse rate of the nitrogen injections to very low levels to maintain regenerator temperature. At 20.00 samples were taken of the bed material and dust from the various collection points.

8.6.73 (Day 3)

Some adjustments were made to the regenerator air rate to provide a 0.2% oxygen level in the off gas stream after samples of bed and dust material was collected at 01.30. It was not possible to maintain the trace oxygen level in the regenerator off gas even with high air rates suggesting excessive carbon on the stone which was burning off in the regenerator. The gasifier temperature was raised to 900°C in an attempt to burn off more of the carbon.

A further set of samples was taken at 13.00 before tests were made on the regenerator off gas flow rate by helium injection into the regenerator upper gas space and measurement of the downstream concentration.

The pump on the boiler water pressurisation system developed a problem at 19.30 and was unable to maintain the pressure in the system without the standby pump. The pump could not be examined without a total shut down and an emergency hand operated pump was connected up to provide additional back up facilities.

9.6.73 (Day 4)

Bed material and dust samples were collected at 02.00 before cutting off the bed feed to lower the bed height. Soon after this the scrubber knock out vessel blocked and water was drawn through the recycle blower before the supply could be turned off. When the system was cleared and restarted the gasifier distributor pressure drop had risen from 4 kPa (16 ins w.g) to 5.5 kPa (22 ins w.g) with the same total gas flow. Debris from the obstructed flue gas recycle line must have been carried through to the distributor nozzles.

At 06.30 some problems were encountered with the oil preheater which repeatedly cut out on overtemperature so reducing the oil temperature and upsetting the oil input to the gasifier. During this period without bed feed the loss in bed height was small but there was a steady decline in gasifier performance.

The regenerator off gas oxygen level had gradually risen to 0.5% and the gasifier temperature was lowered to 880°C which was reckoned to be a more efficient operating temperature but within the following few hours there was no observable improvement in SO_2 removal efficiency and the temperature increased to 910°C. At 14.00 further bed and dust samples were collected before the resumption of stone feed at a molar rate. It was observed that there was some slight smoke from the stack which disappeared when the flue oxygen was increased from 1.5 to 2.0%.

10.6.73 (Day 5)

The unit continued to run steadily with stoichiometric stone feed and efficiencies between 85% and 90% were measured. A problem arose with the liquid nitrogen supply due to a fault on the road tanker which prevented the refilling of the supply and the tanker did not return later in the day as arranged. This situation resulted in a critical period during which the nigrogen supply was obtained from compressed gas cylinders until the liquid supply tank was replenished. Some bed material and dust samples were taken at 10.30 in case the unit was shut down because of the nitrogen shortage. During this period further problems were encountered with the oil heater which continued to shut down at intervals due to the over temperature switch operation.

At 18.00 bed material and dust samples were collected and after this the sampling cyclone on the boiler door was removed so that the line could be rodded out. It was heavily obstructed with lime deposits in the boiler door. After this operation the boiler SO_2 level apparently rose from approximately 175 ppm to 350 ppm. Bed material and dust samples were collected at 23.59.

<u>11.6.73 (Day 6)</u>

The boiler sample line was rodded at regular intervals to remove any lime deposits and the SO_2 concentration in the boiler usually increased marginally after this operation indicating that absorbtion could take place in the sample tube. During this period the scheduled stone feed of 20 kgs/h (44 lbs/h) was difficult to hold steady and when a fresh bag of stone was added the variations were most marked possibly due to varying stone particle distribution.

At 10.20 the air cooling supply to the boiler probe failed and the standby compressor was switched in but in the interval the temperature of the probe rose considerably above the control point of 600°C. The regenerator lower bed tapping blocked but it was successfully cleared by rodding. Soon after this the regenerator temperature started to drop below the set point too fast for the controller to hold and the situation was controlled by switching off the transfer pulse system to prevent the flow of bed material and gradually the temperature recovered to its operating level. The boiler flue gas was sampled at 13.30 to determine the solids content. During the afternoon there were several instances of the regenerator temperature dropping due to high flow rates in the transfer line but after some adjustments the system recovered. Bed material, dust and product gas samples were taken at 19.00.

Various mechanical problems arose towards the end of the day due to blocked pressure tappings and seized valves but all these were overcome without difficulty. The butterfly valve in the left hand cyclone drain leg began to stick and it was discovered that the pneumatic valve operator was leaking around its seal so reducing the operating torque. The joints were tightened but still manual assistance was required at intervals to assist valve operation. Further problems with low regenerator temperatures and adequate bed transfer could be overcome by shutting off the pulse system and using the slow steady bleed of nitrogen to operate the material transfer.

12.6.73 (Day 7)

It appeared that some change had occurred in the fines return system because the material collection rate at the boiler and stack drain points increased significantly and the regenerator fines collection rate fell. Soon after this period the regenerator cyclone accumulated an abnormally large quantity of material at a rate of 5.2 kgs/h (11.5 lbs/h) and there may have been a hold up in the fines returning to the regenerator which cleared releasing a large quantity of material. During this period of operation the regenerator air rate had been reduced thus requiring a lower circulation rate to maintain the control temperature. This low rate was near the minimum rate obtainable by using the bleed gas only without nitrogen pulses for trimming control.

A sieve analysis on the bed feedstock at 03.00 showed one batch was very dusty with only 9.2% of the material greater than 1400 microns. The erratic fines collection rate may have arisen from the feedstock variation. At 08.00 bed material and dust samples were collected with the unit running at a steady level.

The stone feed rate was reduced to about ½ stoichmetric about 11.30 but further stone size variation made accurate metering difficult. Following this, the regenerator air rate was increased to achieve maximum SO_2 removal rate. This caused a regenerator high temperature condition due to the increased air rate liberating more heat and whilst the automatic controller demanded more bed flow for coolant the system was hampered by the rate of transfer from the regenerator which is manually controlled. During this period the gasifier drain was inadvertantly left open after bed material was drained off to lower the gasifier bed level by 12.5 mm (.5 ins).

The regenerator air rate had been gradually increased until a value of 30 m^3/h (17.6 ft^3/m) showed a maximum SO_2 removal rate under these conditions of unit operation. After six hours of good operation bed, gas and dust samples were collected at 20.00. After repeated malfunction of the oil preheater it was discovered that the electrical relay for the heat circulating pump was sticking and by selecting a manual override control the intermittent functioning was overcome and the circulating pump operated correctly. Samples were collected of bed material, dust and the gas product at 23.59.

13.6.73 (Day 8)

The continuous operation of the circulating pump increased the oil feed to the gasifier and this increase in fuel rate caused an increase in the gasifier space pressure but after reducing the fuel back to the correct value some increase in pressure still remained. Preparations were made to start shooting gasifier bed material into the left hand cyclone to remove some of the deposits in the entry which were probably causing the increased pressure drop. During this period the left hand cyclone stopped working and significant quantities of material were carried over into the boiler so lowering the gasifier bed height. The cyclone drain system was restarted and began working reasonably well until the butterfly valve actuator malfunctioned again requiring manual assistance at almost every operation to complete its cycle. The compressor for the boiler probe cooling air failed during the night with the result that the probe temperature rapidly rose above 1000°C before the standby compressor could be brought in.

At 05.00 some experiments were made by inducing different velocities in two sections of the gasifier bed promoting a gross circulation or gulf streaming in the bed. This was achieved by controlling the air flows to the split plenum of the distributor.

At 08.00 samples were taken of the bed material, dust and product gas. The cyclone fines return system continued to give trouble and a new actuator was obtained and fitted together with a water spray to keep the actuator mechanism cool. The stack top washer support legs buckled due to corrosion and a temporary repair was made. At 16.00 bed material and dust samples were collected.

14.6.73 (Day 9)

The unit continued to run steadily with some small problems with the regenerator bottom tapping and scrubber which was rectified. At 04.00 samples of bed material and dust were collected. A gas analysis was made on the regenerator off gas using gas chromotography with an average SO_2 concentration of 6.8% compared to 7.2% with the Maihak analyser. Some problems arose during the early morning with the cyclone drain system which stopped functioning because the butterfly valve was not sealing properly. During this period the shooter overfilled the cyclone and a large quantity of material was transported into the boiler and through to the scrubber blocking up the water drain and so flooding the recycle and gasifier blowers. At this period the gasifier gas space pressure was rising fairly rapidly towards the recommended maximum level and further trials were started to shoot material at both cyclone inlets in turn in an attempt to prolong the operational period before a burn out.

At 15.30 samples were collected of bed material and dust. The controlling air valve on the probe cooling system failed and went to a fully open position which overcooled the probe. At 21.30 the shooter was stopped to the right hand cyclone because the gasifier space pressure rise was not improved by its use. Soon afterwards there were high material losses through to the boiler probably caused by a gas flow up the right hand cyclone drain leg which was unable to sustain a sufficiently deep seal of fine material necessary to balance the high pressure drop across the cyclone entry. At this stage further unit operation was not very useful and sulphation and burn out was started at 22.45.

15.6.73 (Day 10)

The carbon burn out was prolonged and after six hours kerosene combustion was established although there was still some residual carbon in the ducts which then burnt off. The unit was shut down at 11.00 to clean the rear end of the boiler and restarted to recover the temperature before shutting down again to clean the boiler front soot box. Gasification was restarted at 19.55 without difficulty.

16.6.73 (Day 11)

The bed circulation system was erratic in the early part of the day and there was an apparent link between irregularities in the boiler SO_2 level and the operation of the gasifier to regenerator transfer pulse system. It was possible that there was a back flow of gas up the internal cyclone drain line so disturbing the cyclone performance. The remainder of this day was spent in settling the bed transfer system and regenerator performance both of which were rather erratic. The regenerator lower pressure tapping blocked repeatedly and proved an unreliable guide to bed height.

17.6.73 (Day 12)

The stone feed rate of 27 kgs/h (59.5 lbs/h) was maintained but was erratic probably due to inconsistent limestone feedstock size distribution. At 05.00 bed and dust samples were collected to determine the performance of the unit with the 2 X stoichmetric feed rate of limestone. During this period the regenerator performance was not as good as usual although the gasifier efficiency was high. During the early morning the fines recovered from the regenerator were averaging 6 kgs/h (13.2 lbs/h) over a typical 3 hour period and later in the afternoon this increased to 13.6 kgs/h (30 lbs/h) over a 5 hour period.

At 15.30 samples were collected of the bed material and dust from the various collection points. The hot stone shooting system was set to operate on the right hand cyclone which drains into the gasifier to regenerator transfer line. During some periods when the shooter operated frequently, the regenerator temperature dropped markedly and the automatic controller found difficulty in accomodating these slugs of colder stone which drained from the cyclone.

18.6.73 (Day 13)

The regenerator performance improved during the first few hours of this day in spite of a slight drop in the gasifier bed height. Stone and dust samples were collected at 01.00 before changing the stone feed rate to 1½ stoichmetric addition rate. The regenerator temperature control became erratic after the hot stone shooter operating rate was slowed down and this has been due to the colder stone which results from a slower operating rate.

Three gas samples were taken from the boiler flue at 16.30 and analysed for NO_x giving 130, 146 and 152 ppm, bed material and dust samples were taken at 17.00.

At 18.08 the left hand fuel injector was shut off and its oil supply routed to the single nozzle positioned through the distributor. At 19.12 the centre injector fuel supply was added into this single nozzle and at 20.06 the third fuel injector was included to provide all fuel entry through this single nozzle positioned in the distributor. The boiler SO_2 level increased very sharply after the total oil supply was fed to the bottom injector giving efficiencies of 51% minimum and at the same time there was an increase in the fines drained from the left hand cyclone suggesting a bigger material carry over due to the single injection fuel supply condition.

19.6.73 (Day 14)

The air supplied with the fuel was changed from 11.9 m^3/h (7 ft^3/m) to 17 m^3/h (10 ft^3/m) to determine the effect of this air rate but the boiler SO_2 level remained high at 600 ppm. A set of samples was collected at 05.00 followed by trials with the single fuel injector performance at gasifier temperatures of 900°C, 870°C and 840°C. Boiler SO_2 levels were higher at 900°C than at the two lower temperatures, both of which gave similar values around 66% sulphur removal efficiency and the unit was then returned to 870°C gasifier temperature. The regenerator off gas was analysed by gas chromatography showing 1.3% CO_2 and 7.1% SO_2 with the balance being nitrogen with a trace of water. This SO_2 concentration was higher than 6% average value obtained by the Maihak analyser which was the instrument used for continuous analysis.

At 21.00 some problems were encountered in maintaining good bed circulation between the gasifier and regenerator. The automatic controller was unable to control correctly and the system was switched to manual control with a high pulse rate consuming 4 times the normal nitrogen demand.

20.6.73 (Day 15)

Further trials were made with gasifier bed gulf streaming to determine if this would improve the fuel distribution from the single injector which at this point could have been the source of continuing poor bed transfer. Trials were made with 30% velocity differentials but there was not any marked effect upon transfer rate or boiler SO_2 level. The latter was masked by the irregularity in the left hand cyclone drain butterfly valve closure which permitted nitrogen to leak up the drain leg and blow fines into the boiler.

At 03.10 the fuel supply was returned to the three side injectors and by 05.30 bed transfer improved after rodding out the regenerator to gasifier transfer leg and the cyclone return system functioned without excessive leakage. At 06.10 the gulf streaming trials were stopped. Some of the erratic fines transfer problems were caused by lumps of carbon in the transfer line from the cyclone drain.

The unit gradually lined out after the unsettled period and at 18.00 bed material, dust and gas samples were collected before trials commenced by switching the fuel supply from the left hand sidewall injector to the injector through the distributor without any significant effects upon performance. At 19.50 the centre sidewall injector supply was transferred to the distributor injector leaving one sidewall injector on the right hand side. About 40 minutes later the boiler SO_2 line had risen about 20 ppm together with a small rise in the gasifier temperature.

21.6.73 (Day 16)

Some adjustments were made with the fuel supplied through the injector in the distributor but increased flow produced higher boiler SO_2 levels showing that 120 kgs/h was the maximum throughput with this single outlet fuel injector. Some problems were encountered with the regenerator air rate which for some reason could not be raised beyond 30 m^3/h (17.6 ft^3/m) with all the control valves open. The regenerator performance of less than 60% sulphur removal was caused by this limited air supply and the stone sulphur level was apparently building up due to the inability of the regenerator to adequately strip the stone. Samples of bed material and dust from the various collection points were taken at 10.15 before a retrial with the total fuel supply fed through the single outlet fuel injector confirmed the previous result of higher boiler SO_2 levels.

The behaviour of the unit was limited by the regenerator air supply and before burning out and investigating the regenerator, preparations were made to carry out tests at lower bed depths to provide further data and at the same time provide an increased regenerator air rate. Material was withdrawn until a bed depth of 40 cms (15.7 ins) was reached and the unit allowed to level out.

22.6.73 (Day 17)

Problems were encountered with the fines return system and the increase in material collected at the boiler suggested some cyclone drain obstruction. The fuel injector through the distributor was retracted without trouble so that preparations could be completed for the burn out which was started at 12.25 and by 15.25 most of the carbon had been burnt out. The unit was put onto combusting conditions to raise the temperature to 925°C so that a reasonably prolonged shut down could be tolerated whilst the regenerator distributor was removed. The bore of the regenerator was quite clear and there were some glazed 1.5 mm (.06 ins) thick deposits around the lower wall area. The distributor was generally clear apart from some local deposits between the nozzles but these would not have caused any fluidisation problems. The distributor was replaced without the silicon carbide spacer ring so returning to the distributor configuration of Run 5.

23.6.73 (Day 18)

During the early part of this day the flue gas recycle system was stripped and cleaned. The left hand cyclone leg was rodded out to remove any obstruction in the 50 mm (2 in) diameter drain leg and the fines return control system checked over. The pressurisation unit for maintaining the boiler water pressure was also checked over in an attempt to locate the problem which prevented the pump building pressure and the performance improved after cleaning the filters and non-return valves.

In the afternoon two pilot flame burners were installed on each cyclone outlet stream so that SO_2 samples could be taken from each stream thereby showing any difference between the concentrations in each cyclone outlet. This was arranged because there was some possibility of SO_2 passing from the regenerator to the gasifier by flowing up the bed transfer passage and in such an event the SO_2 level in the right hand cyclone would be greater due to the proximity of this transfer passage to this cyclone inlet.

At 17.00 the boiler rear door was opened and 175 kgs (385 lbs) of material removed from the back and 66 kgs (145 lbs) removed from the front soot box.

24.6.73 (Day 19)

At 01.30 gasification was restarted using a molar stone feed rate and the unit allowed to line out at 880°C gasifier temperature and 1020°C regenerator temperature whilst checking out the fines transfer system which although operating through the control cycle was not transferring much material. The hot lime shooter system was started up on the left hand cyclone entry but causes a rise of 100 ppm in the boiler SO_2 level and during a comparatively short operating period the bed level in the gasifier dropped by 51 mm (2 ins) suggesting that much of the material shot into the cyclone was carried into the boiler and not retained by the cyclone. Various tests were made to determine if the cyclone drain was obstructed but apparently material was draining into the transfer vessel and the slow rate of collection was probably due to the low gasifier bed depth of 38 cm (15 ins) which threw less material into the cyclone. The regenerator cyclone dust collection rate had become small indicating that either the right hand cyclone was not collecting much material or else the internal drain to the transfer line was obstructed. A high pressure nitrogen lance was inserted down the right hand cyclone drain leg and produced a severe disturbance in the temperature and gas composition in the regenerator showing that the cyclone drain leg to the regenerator was clear.

At 19.00 trials were made with the two pilot burners on the bifurcated duct but there were some initial difficulties in maintaining a steady flame due to irregularities in the air supply pressure. The regenerator performance improved as a result of the rodding with the nitrogen lance.

25.6.73 (Day 20)

The pilot flame trials continued but it was difficult to obtain steady conditions because of the carbon laydown in the burner ducts which disturbed the gas flow. The infra-red gas analyser did show some short spikes in the SO_2 level but the Wostoff analyser did not pick up these transients. The regenerator performance improved a little after the pressure of the nitrogen pulse transfer was reduced so creating less disturbance in the gasifier to regenerator transfer line at each operation.

At 22.40 there was a sharp increase in the regenerator circulation rate shown by a sudden temperature drop and at the same time the fines collection rate increased in the left hand cyclone transfer vessel. This change could have been caused by a change in limestone feed size distribution or the clearing of some obstruction in a transfer line. Various small problems arose during this day in the flue gas recycle system when blockages formed in the scrubber outlet chamber and the control valve together with obstructions in the gasifier pressure tappings.

26.6.73 (Day 21)

At 04.00 samples of bed material and dust were collected and the limestone feed temporarily stopped to determine the short term effect upon boiler SO_2 level. After 4 hours the level had risen from a previous average of 290 ppm to a new average value of 400 ppm whilst other conditions including bed depth remained reasonably constant. The stone feedstock was changed to Denbighshire at 08.40 with ½ molar feed rate. The regenerator controller experienced some difficulty in maintaining the temperature because the renewal of stone feed introduced fines into the bed apparently increasing the transfer rate between gasifier and regenerator.

At 13.00 the limestone feed rate was increased to stoichmetric and bed material was drained where necessary to maintain a gasifier bed of about 53 cm (21 ins). The regenerator performance was not good during the majority of this day possibly due to irregularities in the fines within the system. Some investigations were made into the poor regenerator performance by varying the air rate from 29 m^3/h (17 ft^3/m) to 25.5 m^3/h (15 ft^3/m). Initially the regenerator offgas SO_2 concentration remained unchanged but gradually increased although not sufficiently to give an overall improvement in performance.

<u>27.6.73 (Day 22)</u>

At 00.55 the boiler SO_2 level had risen by 40 ppm during the rise in the regenerator offgas SO_2 concentration. At 01.35 the boiler SO_2 was still higher with a further increase of 20 ppm. The regenerator air rate was increased to 30.6 m^3/h (18 ft^3/m) and the trend in boiler SO_2 and regenerator SO_2 concentration was reversed. Further trials were made with the two pilot flames when the unit conditions steadied around 04.00 and reasonably steady conditions prevailed during this period. Prior to shut down the regenerator bed was slumped so that the boiler SO_2 level could be measured without any possibility of SO_2 leaking up the material transfer line between the gasifier and regenerator, no change was observed.

The plant was shut down at 18.46 with nitrogen purges in the gasifier and regenerator to prevent the ingress of air during the cooling period which could burn out some of the carbon in the ductwork and cyclones.

APPENDIX C

CAFB RUN 6

INSPECTION

Gasifier and Regenerator Refractory

The gasifier refractory was in reasonable condition without significant deterioration. The walls were originally blackened by carbon deposits which were thin and shiny in the lower area but thicker in the upper section particularly near the lid (fig. C.1). The cracks in the upper concrete were again deposited with bands of thicker carbon about 3 cms wide. The gasifier lid hot face refractory slab was coated with carbon and the insulation behind this concrete was badly cracked accounting for the leakage problems experienced during part of the run.

The transfer passages to and from the regenerator were both clear and the refractory in excellent condition. There was some agglomerated material in the static corners of the gasifier transfer pocket but this would not have caused any circulation problems.

The regenerator bore was generally clean but there were some new cracks in the lower section and some grooves in the concrete where small pieces had fallen away. The upper walls were lightly deposited with a hard coating of material with an irregular needle like surface. The silicon carbide spacer ring which was used initially to lower the distributor was in excellent condition with only a few local areas of fine material firmly bonded to the inner bore.

Gasifier and Regenerator Penetrations

The thermocouples, fuel injectors and pressure tappings were all in good order with some deposits of carbon and lime on their exposed portions. The centre fuel injector was particularly deposited with a large build up of carbon which bridged across from the injector to the refractory wall.

Cyclones

The cyclone entries (fig. C.1) were coated with carbon and lime on all sides with a maximum thickness of 10 mm in some areas. The left hand cyclone entry (fig. C.2) shows the irregular surface with some fragmented deposits suggesting that the shooter did have some effect upon the deposits. In some areas the deposit had peeled away from the refractory although still firmly attached at one end.

The right hand cyclone entry (fig. C.3) showed similar deposits and in some areas there were upstanding ridges of carbon and lime deposits along the line of the duct. The bores of the cyclones were deposited with carbon and lime which tended to be thick in the upper sections.

The left hand cyclone (fig. C.4) drain leg to the external transfer system was quite clear but the right hand cyclone drain (fig. C.5) was blocked with fine material. This cyclone was connected by an internal duct to the gasifier to regenerator transfer line and this passageway was blocked with an agglomeration of fine particles.

The two silicon carbide cyclone dip pipes are shown in (fig. C.6) with the particularly flaky deposits on the external surface of the tubes. The right hand tube has an area on the right hand side corresponding to the gas impingement area from the cyclone entry mouth.

Gasifier and regenerator distributor

The gasifier distributor was in very good condition (fig. C.7) with some deposits within some of the nozzle outlets. The stainless steel was undamaged and the refractory on the distributor face was in excellent condition. (Fig. C.7) shows one of the carbon deposits broken away from a fuel injector lying on top of the distributor. The gasifier distributor drain was solidly obstructed with fine material and it had not been possible to use this drain during the test period.

The regenerator distributor was in good condition with the stainless steel nozzles lightly deposited on their top faces with fine lime particles firmly bonded to the metal (fig. C.8) The centre drain hole was obstructed in its lower portion in spite of the nitrogen purge maintained during the operational period. The holes in the distributor were generally clean with only a slight deposition of fine material in some of the holes.

Bed material

The unit was shut down without sulphation of the bed and (fig. C.9) shows the slumped gasifier bed after removal of the lid. The bed was generally free of agglomerates and was homogeneous throughout its depth. The regenerator bed was free flowing and again was free from any agglomerates.

Outlet Ducts

The bifurcated duct was coated with carbon and lime along the bore of the gas passages with larger deposits at the junction between the two ducts and around penetrations such as thermocouples and pressure tapping probes.

The regenerator outlet pipe was coated with a hard irregular deposit (fig. C.10) which was thickest in the duct when it joined with the main gasifier outlet. Further downstream the growth formation became thinner until it formed a light coating uniformly deposited within the pipe bore.

Premix section

The air premix section situated between the bifurcated duct outlet and the main burner provides the first stage of air admission to the hot gas. The central hot gas duct built from stainless steel was coated uniformly with a thin tenacious layer of carbon deposited over all the internal surface. The steel was in good condition without sign of scaling or cracking.

Burner section

The main burner was undamaged although deposited with carbon and lime in the hot gas duct. The stainless steel clad thermocouple placed in the burner throat had burnt away.

The pilot burner was in good order with some light deposits of lime around its flame holder.

Boiler and stack

The boiler rear end was deposited with a quantity of lime particles (fig. C.11) some of which were quite coarse indicating that the cyclones had not been very effective over part of the operational period. The entries into the first tube pass (fig. C.12) were coated with a hard crust built up around their peripheries but without much penetration into the tube bore.

The bottom corrugations of the main corrugated fire tube of the boiler were deposited with fairly coarse material and down the sides and top with local agglomerates of fine material (fig. C.13).

The stack was generally clean apart from a small quantity of material built up at the base of the stack. The boiler had been cleaned during the shut down during day 10 and (fig. C.14) shows the boiler rear end immediately before cleaning.

Boiler Probe

The boiler probe shown in (fig. C.13) was coated with an uneven deposit of lime leaving a rough surface. It is not possible to draw conclusions from the deposits on the boiler probe because the service conditions were not constant at 600°C because of some failures in the cooling compressors during the operating period.

Fig. C.1   Gasifier cyclone inlets

Fig. C.2   L.H. cyclone inlet

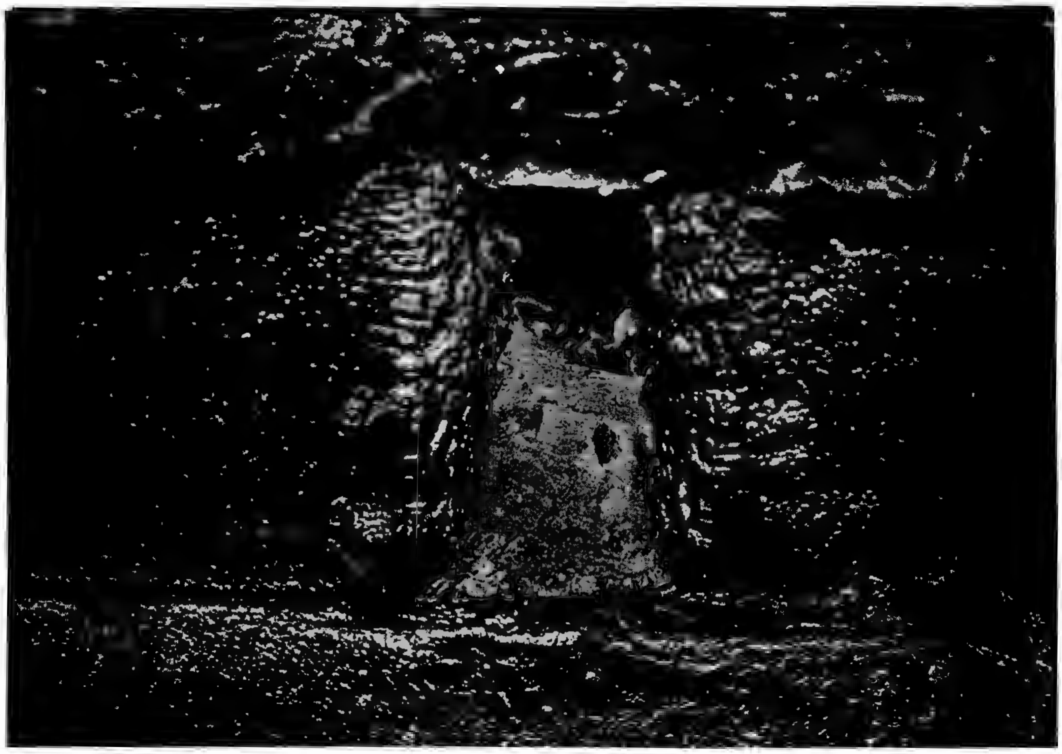

Fig. C.3   R.H. cyclone inlet

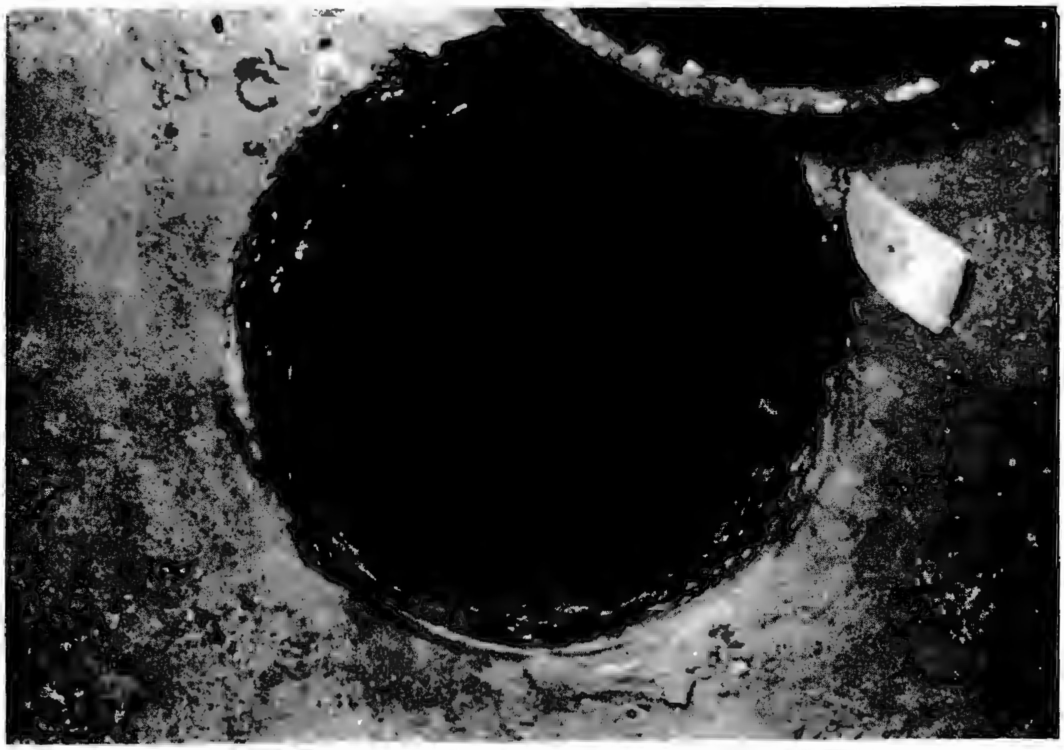

Fig. C.4   L.H. cyclone

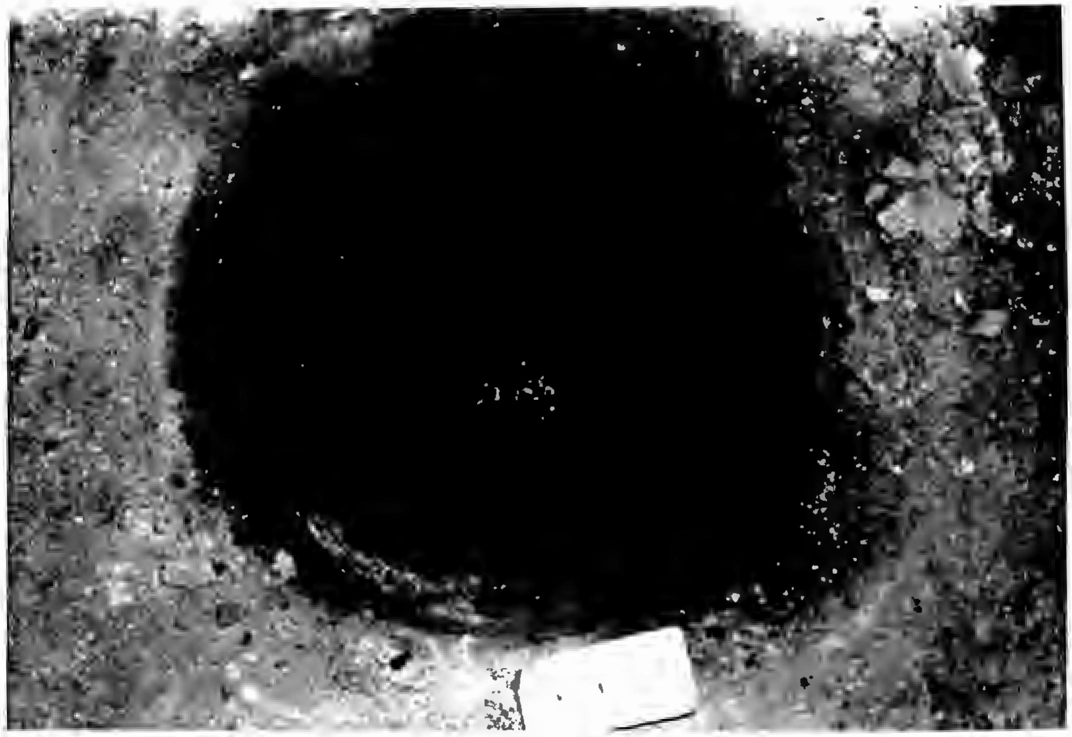

Fig. C.5 R.H. cyclone

Fig. C.6 Cyclone outlet pipe

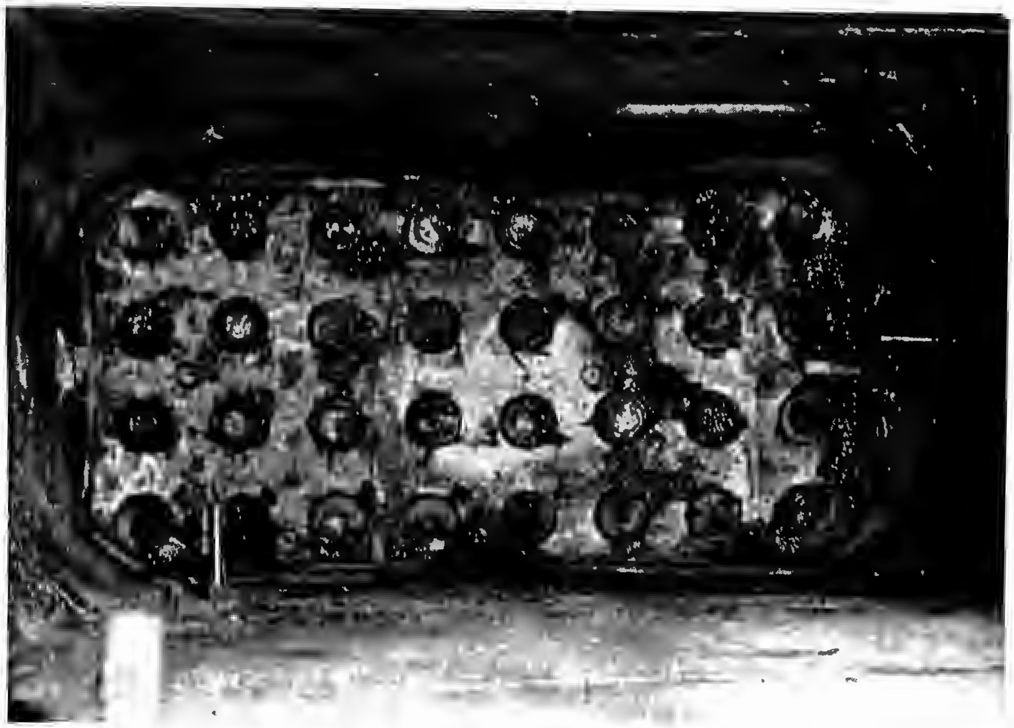

Fig. C.7   Gasifier distributor

Fig. C.8   Regenerator distributor

Fig. C.9 Slumped Gasifier Bed

Fig. C.10 Regenerator outlet pipe

Fig. C.11 Boiler back end

Fig. C.12 Boiler first tube pass

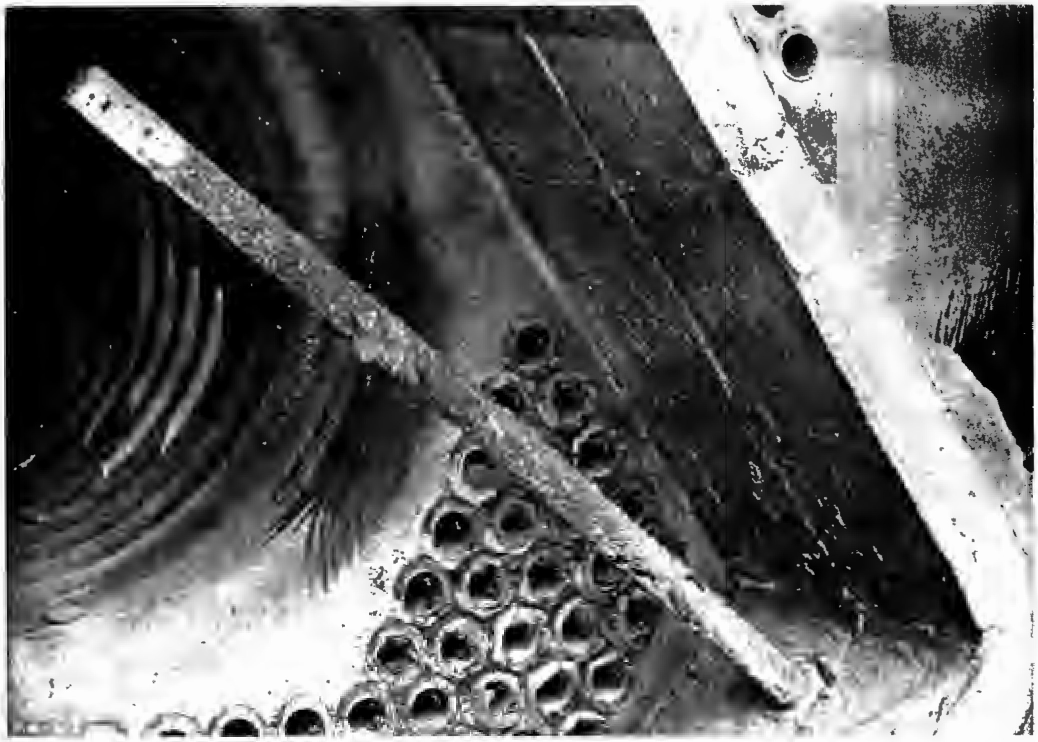

Fig. C.13 Boiler fire tube and probe

Fig. C.14 Boiler backend - day 10

APPENDIX C: TABLE I.

RUN 6: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
1.2230	862.	658.	70.	158.7	11.8
1.2330	955.	950.	70.	175.1	5.9
2.0030	886.	965.	70.	179.3	0.
2.0130	880.	1060.	70.	178.0	0.
2.0230	882.	1055.	70.	183.8	0.
2.0330	871.	1052.	70.	183.0	7.7
2.0430	860.	1052.	70.	185.5	16.3
2.0530	870.	1053.	70.	185.0	13.6
2.0630	868.	1052.	70.	186.7	13.2
2.0730	868.	1055.	70.	187.1	13.6
2.0830	870.	1045.	70.	185.9	14.5
2.0930	870.	1047.	70.	189.2	10.4
2.1030	872.	1049.	70.	189.6	9.5
2.1130	876.	1049.	70.	190.0	10.4
2.1230	868.	1050.	70.	186.7	11.8
2.1330	861.	1049.	70.	189.6	15.9
2.1430	868.	1049.	70.	190.4	11.3
2.1530	872.	1049.	70.	190.0	11.3
2.1630	869.	1050.	70.	188.3	15.0
2.1730	865.	1051.	70.	188.3	16.3
2.1830	880.	1052.	70.	188.7	17.7
2.1930	898.	1050.	70.	190.0	19.5
2.2030	901.	1051.	70.	189.2	15.4
2.2130	901.	1031.	70.	186.7	16.3
2.2230	892.	1021.	70.	189.6	18.1
2.2330	893.	1050.	70.	188.7	16.3
3.0030	898.	1051.	68.	188.7	18.1
3.0130	912.	1052.	65.	188.7	17.2
3.0230	900.	1058.	65.	187.5	16.8
3.0330	895.	1051.	65.	187.5	17.2
3.0430	898.	1050.	65.	188.7	19.1
3.0530	928.	1052.	65.	189.6	18.1
3.0630	918.	1060.	65.	188.3	15.9
3.0730	910.	1050.	65.	187.5	11.3
3.0830	910.	1050.	68.	187.9	16.8
3.0930	908.	1050.	69.	188.7	13.6
3.1030	910.	1051.	69.	188.3	17.2
3.1130	911.	1052.	69.	190.0	17.2
3.1230	920.	1051.	69.	192.5	12.2
3.1330	910.	1052.	69.	192.5	14.5

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
3.1430	908.	1061.	70.	192.9	14.1
3.1530	905.	1069.	70.	192.0	13.2
3.1630	905.	1078.	70.	192.0	14.5
3.1730	898.	1079.	70.	192.9	19.5
3.1830	902.	1088.	70.	192.0	24.9
3.1930	905.	1084.	69.	193.7	22.7
3.2030	910.	1085.	69.	192.9	21.3
3.2130	910.	1082.	70.	191.2	19.5
3.2230	908.	1085.	68.	194.5	21.3
3.2330	912.	1082.	70.	191.6	17.7
4.0030	910.	1081.	70.	193.3	22.2
4.0130	918.	1080.	70.	192.0	19.5
4.0230	891.	1085.	60.	192.5	6.8
4.0330	920.	1085.	60.	194.5	0.
4.0430	912.	1082.	70.	194.5	0.
4.0530	925.	1084.	70.	191.6	0.
4.0630	922.	1083.	70.	185.5	0.
4.0730	918.	1083.	70.	190.0	0.
4.0830	920.	1081.	70.	191.6	0.
4.0930	928.	1082.	68.	193.7	0.
4.1030	918.	1085.	68.	192.0	0.
4.1130	918.	1082.	68.	192.0	0.
4.1230	882.	1083.	70.	192.9	0.
4.1330	878.	1082.	69.	192.9	0.
4.1430	888.	1082.	69.	192.0	0.
4.1530	912.	1083.	70.	192.0	11.8
4.1630	905.	1082.	69.	191.2	14.5
4.1730	902.	1081.	69.	192.0	14.1
4.1830	921.	1083.	68.	191.6	14.1
4.1930	911.	1082.	65.	193.3	11.3
4.2030	913.	1082.	67.	191.6	13.6
4.2130	918.	1082.	70.	192.0	17.7
4.2230	913.	1082.	67.	187.1	13.6
4.2330	908.	1082.	65.	192.5	15.9
5.0030	910.	1082.	62.	199.5	18.6
5.0130	918.	1082.	62.	190.8	15.4
5.0230	921.	1085.	62.	190.8	11.3
5.0330	930.	1085.	62.	191.2	15.0
5.0430	928.	1082.	70.	192.0	14.5
5.0530	895.	1080.	77.	190.4	14.1

APPENDIX C: TABLE I.

RUN 6: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
5.0630	889.	1081.	78.	191.2	13.6
5.0730	896.	1082.	70.	191.2	14.1
5.0830	908.	1082.	70.	190.4	18.6
5.0930	899.	1081.	60.	189.6	14.5
5.1030	890.	1080.	70.	195.8	14.5
5.1130	898.	1081.	70.	192.9	16.3
5.1230	898.	1082.	70.	191.2	18.1
5.1330	910.	1083.	70.	189.6	18.1
5.1430	898.	1081.	69.	189.6	19.1
5.1530	900.	1081.	69.	191.6	17.2
5.1630	889.	1080.	68.	190.0	18.6
5.1730	891.	1082.	70.	190.8	16.3
5.1830	895.	1080.	70.	190.4	12.7
5.1930	891.	1081.	70.	190.8	12.7
5.2030	890.	1081.	70.	189.6	16.3
5.2130	888.	1081.	70.	189.6	15.0
5.2230	894.	1079.	70.	188.3	13.2
5.2330	890.	1080.	70.	190.4	13.2
6.0030	896.	1082.	70.	189.6	15.0
6.0130	901.	1081.	70.	189.6	11.8
6.0230	898.	1081.	70.	188.3	14.1
6.0330	897.	1080.	70.	187.5	15.4
6.0430	886.	1080.	70.	191.6	25.4
6.0530	897.	1080.	70.	189.6	18.6
6.0630	908.	1080.	70.	188.3	20.4
6.0730	904.	1080.	70.	189.6	23.6
6.0830	900.	1080.	70.	189.6	19.1
6.0930	902.	1080.	70.	189.6	26.8
6.1030	918.	1081.	70.	193.3	12.7
6.1130	900.	1079.	69.	187.9	22.7
6.1230	910.	1079.	70.	188.7	21.3
6.1330	911.	1080.	70.	189.2	16.3
6.1430	898.	1081.	70.	188.7	22.2
6.1530	898.	1081.	70.	190.4	26.8
6.1630	885.	1081.	70.	194.1	19.5
6.1730	895.	1084.	70.	190.8	26.8
6.1830	903.	1083.	70.	197.0	15.0
6.1930	916.	1081.	68.	193.7	24.5
6.2030	MISSED DATA READING				
6.2130	893.	1080.	70.	190.8	20.0

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
6.2230	897.	1080.	70.	192.9	20.0
6.2330	890.	1045.	70.	192.0	12.7
7.0030	881.	1056.	71.	194.1	24.9
7.0130	900.	1081.	68.	193.7	26.3
7.0230	898.	1079.	70.	191.6	18.1
7.0330	902.	1080.	70.	192.5	23.1
7.0430	907.	1070.	69.	192.9	19.1
7.0530	902.	1079.	68.	191.6	24.9
7.0630	911.	1072.	67.	191.2	18.1
7.0730	891.	1079.	64.	190.8	22.7
7.0830	900.	1081.	63.	191.2	20.9
7.0930	899.	1080.	64.	190.8	15.9
7.1030	898.	1075.	61.	192.0	15.0
7.1130	909.	1081.	60.	191.6	11.3
7.1230	910.	1079.	60.	192.5	5.9
7.1330	910.	1070.	60.	192.5	6.8
7.1430	912.	1079.	60.	193.3	7.3
7.1530	910.	1080.	60.	192.0	10.0
7.1630	914.	1079.	61.	192.0	10.9
7.1730	912.	1073.	61.	191.2	10.9
7.1830	910.	1072.	61.	193.7	5.9
7.1930	920.	1078.	61.	187.9	6.4
7.2030	915.	1081.	61.	191.2	5.4
7.2130	916.	1080.	65.	191.2	6.4
7.2230	915.	1080.	68.	189.6	5.4
7.2330	896.	1079.	68.	202.8	6.8
8.0030	904.	1080.	67.	203.2	5.4
8.0130	910.	1081.	65.	199.1	6.8
8.0230	894.	1072.	63.	192.5	7.3
8.0330	905.	1080.	62.	192.0	6.8
8.0430	905.	1080.	65.	192.5	6.4
8.0530	905.	1080.	68.	192.5	5.4
8.0630	905.	1081.	68.	192.0	5.9
8.0730	907.	1085.	68.	192.5	7.3
8.0830	908.	1081.	69.	192.0	7.7
8.0930	908.	1080.	70.	192.0	7.7
8.1030	900.	1082.	70.	192.5	5.9
8.1130	899.	1083.	70.	192.5	5.9
8.1230	910.	1084.	70.	192.5	5.9
8.1330	906.	1080.	70.	191.6	5.9

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
8.1430	910.	1085.	70.	191.6	4.5
8.1530	900.	1082.	70.	192.5	7.7
8.1630	910.	1079.	70.	192.9	7.7
8.1730	895.	1080.	70.	181.3	8.2
8.1830	885.	1080.	70.	173.1	9.5
8.1930	894.	1084.	66.	169.4	10.4
8.2030	902.	1079.	67.	169.8	5.4
8.2130	912.	1082.	66.	167.3	5.0
8.2230	911.	1082.	62.	171.9	6.4
8.2330	912.	1085.	70.	169.4	9.5
9.0030	912.	1086.	70.	169.8	6.8
9.0130	899.	1083.	70.	169.8	7.7
9.0230	899.	1083.	70.	170.6	5.4
9.0330	900.	1081.	70.	170.6	5.4
9.0430	903.	1083.	70.	170.6	4.5
9.0530	905.	1085.	70.	170.6	5.9
9.0630	908.	1088.	70.	171.0	6.4
9.0730	906.	1087.	70.	171.0	7.3
9.0830	898.	1088.	70.	171.0	7.7
9.0930	902.	1089.	70.	171.4	7.7
9.1030	898.	1086.	70.	163.2	7.3
9.1130	899.	1085.	70.	163.6	9.1
9.1230	900.	1096.	70.	164.4	9.1
9.1330	891.	1080.	70.	166.5	10.4
9.1430	888.	1084.	70.	169.0	9.5
9.1530	891.	1086.	70.	167.3	11.3
9.1630	898.	1088.	70.	166.5	9.1
9.1730	880.	1079.	70.	176.4	12.2
9.1830	905.	1070.	70.	175.1	13.2
9.1930	910.	1069.	70.	156.6	14.5
9.2030	920.	1070.	70.	164.4	11.3
9.2130	906.	1062.	69.	165.7	9.1
9.2230	891.	1063.	64.	165.7	13.6
SHUT DOWN AT 9.2230 FOR 23 HOURS					
10.2130	892.	1057.	70.	170.6	23.1
10.2230	899.	1070.	69.	177.2	26.3
10.2330	898.	1070.	70.	178.9	20.9

APPENDIX C: TABLE I.

RUN 6: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
11.0030	898.	1074.	70.	178.4	17.7
11.0130	898.	1080.	70.	175.6	23.1
11.0230	892.	1079.	70.	172.7	27.2
11.0330	889.	1080.	69.	171.9	26.8
11.0430	889.	1078.	69.	178.4	20.9
11.0530	901.	1079.	69.	175.6	12.7
11.0630	890.	1080.	68.	171.9	29.9
11.0730	904.	1080.	67.	169.8	29.5
11.0830	890.	1078.	65.	171.0	26.8
11.0930	882.	1080.	70.	171.4	30.8
11.1030	880.	1072.	68.	173.5	29.0
11.1130	880.	1072.	68.	173.1	29.0
11.1230	880.	1072.	65.	174.3	29.0
11.1330	890.	1052.	65.	173.1	28.6
11.1430	890.	1070.	69.	172.7	31.3
11.1530	881.	1080.	68.	172.7	29.5
11.1630	886.	1060.	68.	172.7	29.9
11.1730	888.	1056.	68.	173.1	31.8
11.1830	900.	1080.	68.	170.6	27.2
11.1930	924.	1080.	67.	173.1	21.3
11.2030	910.	1070.	70.	174.3	0.
11.2130	885.	1080.	70.	170.6	15.9
11.2230	865.	1065.	70.	171.0	20.9
11.2330	870.	1066.	70.	169.8	29.9
12.0030	878.	1061.	70.	169.4	21.8
12.0130	883.	1080.	70.	170.2	25.4
12.0230	899.	1080.	70.	169.8	12.2
12.0330	892.	1078.	70.	169.4	23.1
12.0430	898.	1064.	70.	173.1	30.4
12.0530	890.	1079.	70.	173.5	29.5
12.0630	893.	1079.	70.	169.0	24.0
12.0730	889.	1070.	69.	170.2	24.5
12.0830	887.	1070.	69.	169.4	28.1
12.0930	878.	1086.	70.	169.8	29.9
12.1030	901.	1085.	70.	170.2	20.0
12.1130	899.	1067.	70.	171.0	18.1
12.1230	900.	1064.	71.	171.9	29.9
12.1330	899.	1065.	70.	170.2	28.1
12.1430	905.	1065.	70.	171.0	28.6
12.1530	900.	1048.	70.	171.4	24.5

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
12.1630	883.	1076.	71.	170.2	23.6
12.1730	861.	1068.	72.	170.6	18.1
12.1830	871.	1075.	71.	170.6	22.2
12.1930	870.	1078.	72.	171.0	17.7
12.2030	870.	1078.	72.	170.2	17.2
12.2130	871.	1076.	72.	171.4	20.4
12.2230	881.	1076.	72.	169.0	16.3
12.2330	882.	1076.	71.	169.0	19.5
13.0030	888.	1072.	71.	168.6	21.3
13.0130	881.	1070.	71.	170.2	16.3
13.0230	880.	1071.	71.	170.6	24.5
13.0330	875.	1049.	71.	170.6	18.6
13.0430	883.	1073.	71.	166.1	21.3
13.0530	889.	1068.	70.	169.4	20.0
13.0630	876.	1073.	71.	169.8	20.4
13.0730	882.	1061.	71.	165.7	30.4
13.0830	872.	1065.	72.	172.7	21.8
13.0930	870.	1079.	71.	169.8	25.9
13.1030	870.	1080.	72.	170.2	19.1
13.1130	864.	1079.	72.	170.2	22.7
13.1230	867.	1079.	72.	170.6	23.1
13.1330	879.	1082.	70.	170.6	22.7
13.1430	892.	1078.	72.	170.2	16.3
13.1530	890.	1075.	72.	170.6	11.3
13.1630	894.	1072.	72.	171.0	16.8
13.1730	902.	1074.	72.	170.2	15.9
13.1830	900.	1078.	72.	170.2	19.1
13.1930	904.	1070.	72.	170.2	19.1
13.2030	919.	1068.	71.	165.7	17.7
13.2130	902.	1070.	72.	167.7	16.8
13.2230	902.	1069.	72.	171.0	20.4
13.2330	881.	1070.	72.	171.0	18.6
14.0030	880.	1070.	71.	173.5	15.9
14.0130	881.	1069.	71.	174.3	16.8
14.0230	888.	1070.	71.	174.3	15.4
14.0330	881.	1068.	72.	173.1	10.4
14.0430	878.	1070.	72.	173.9	15.4
14.0530	881.	1070.	70.	176.4	18.1
14.0630	895.	1070.	71.	175.6	16.8
14.0730	900.	1072.	71.	173.5	20.0

APPENDIX C: TABLE I.

RUN 6: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
14.0830	901.	1072.	70.	174.3	21.3
14.0930	888.	1068.	70.	167.3	13.2
14.1030	872.	1075.	70.	176.8	15.4
14.1130	872.	1059.	70.	176.4	15.0
14.1230	869.	1062.	70.	176.8	13.2
14.1330	872.	1069.	70.	176.8	14.5
14.1430	873.	1062.	70.	176.8	14.1
14.1530	869.	1063.	70.	177.6	12.7
14.1630	856.	1055.	71.	177.2	11.8
14.1730	859.	1060.	70.	177.6	16.8
14.1830	850.	1061.	70.	177.2	20.9
14.1930	862.	1060.	72.	177.6	16.3
14.2030	869.	1053.	71.	177.6	15.0
14.2130	865.	1070.	70.	178.0	16.8
14.2230	860.	1051.	70.	174.7	15.0
14.2330	865.	1053.	70.	176.8	11.3
15.0030	878.	1081.	70.	179.3	15.9
15.0130	878.	1078.	70.	178.0	19.1
15.0230	878.	1080.	70.	180.9	16.8
15.0330	859.	1050.	70.	184.2	17.2
15.0430	862.	1058.	70.	188.7	16.3
15.0530	845.	1052.	70.	191.2	13.6
15.0630	855.	1080.	70.	188.3	13.2
15.0730	866.	1088.	70.	188.3	16.8
15.0830	872.	1070.	70.	189.6	16.3
15.0930	872.	1040.	71.	188.3	14.5
15.1030	879.	1038.	72.	182.2	16.3
15.1130	873.	1042.	70.	178.0	16.8
15.1230	867.	1045.	70.	177.2	14.1
15.1330	869.	1051.	72.	177.2	15.9
15.1430	872.	1051.	70.	173.5	13.2
15.1530	870.	1055.	70.	174.7	13.6
15.1630	870.	1050.	70.	175.6	12.2
15.1730	865.	1049.	70.	175.1	21.8
15.1830	865.	1060.	70.	174.7	19.5
15.1930	859.	1060.	70.	176.8	17.7
15.2030	880.	1059.	71.	176.0	18.6
15.2130	878.	1061.	71.	173.5	16.8
15.2230	871.	1050.	70.	172.3	18.6
15.2330	869.	1055.	72.	174.3	16.8

APPENDIX C: TABLE I.

RUN 6: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
16.0030	870.	1055.	72.	173.9	17.7
16.0130	874.	1057.	70.	173.5	16.3
16.0230	872.	1058.	70.	176.0	18.1
16.0330	890.	1062.	70.	168.1	15.9
16.0430	880.	1056.	70.	166.5	10.9
16.0530	878.	1058.	70.	171.0	12.2
16.0630	871.	1057.	70.	169.8	19.1
16.0730	880.	1057.	70.	169.8	14.1
16.0830	880.	1059.	70.	170.2	11.3
16.0930	882.	1057.	70.	169.4	10.9
16.1030	880.	1059.	70.	169.4	14.1
16.1130	871.	1060.	70.	166.5	15.0
16.1230	871.	1070.	70.	166.1	17.2
16.1330	871.	1069.	70.	165.3	22.2
16.1430	889.	1065.	70.	164.8	19.1
16.1530	866.	1061.	70.	164.8	19.1
16.1630	861.	1066.	70.	165.3	24.0
16.1730	879.	1068.	70.	165.3	10.9
16.1830	866.	1059.	70.	165.7	15.4
16.1930	871.	1062.	70.	165.3	23.1
16.2030	874.	1068.	70.	165.3	21.8
16.2130	868.	1070.	70.	164.8	25.9
16.2230	880.	1072.	70.	164.8	20.4
16.2330	884.	1072.	70.	164.8	17.2
17.0030	890.	1079.	70.	165.3	17.2
17.0130	890.	1052.	70.	166.9	15.4
17.0230	888.	1062.	70.	167.7	13.2
17.0330	878.	1062.	70.	165.3	24.0
17.0430	880.	1063.	70.	168.6	22.7
17.0530	883.	1061.	70.	168.6	20.9
17.0630	875.	1060.	70.	168.6	18.6
17.0730	880.	1055.	70.	169.0	15.0
17.0830	873.	1058.	70.	167.7	24.5
17.0930	883.	1055.	70.	166.1	18.6
17.1030	876.	1058.	70.	164.8	15.9
17.1130	872.	1052.	70.	164.8	18.1

SHUT DOWN AT 17.1130 FOR 40 HOURS

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
19.0430	882.	1079.	70.	166.9	22.2
19.0530	886.	1079.	70.	167.3	13.2
19.0630	880.	1080.	70.	166.9	15.9
19.0730	879.	1081.	70.	166.5	15.9
19.0830	881.	1079.	70.	167.7	14.1
19.0930	879.	1080.	70.	168.1	14.1
19.1030	882.	1068.	70.	168.1	16.3
19.1130	871.	1059.	70.	170.2	15.9
19.1230	868.	1060.	70.	171.0	15.9
19.1330	872.	1059.	70.	172.7	17.2
19.1430	875.	1060.	70.	173.5	17.2
19.1530	895.	1064.	70.	170.2	14.5
19.1630	890.	1060.	70.	165.3	15.4
19.1730	872.	1060.	70.	171.4	16.3
19.1830	866.	1059.	70.	169.0	13.6
19.1930	872.	1059.	70.	168.1	12.7
19.2030	870.	1059.	70.	168.1	18.1
19.2130	862.	1059.	70.	169.4	17.2
19.2230	866.	1056.	70.	170.6	14.1
19.2330	878.	1060.	70.	168.6	16.3
20.0030	880.	1060.	70.	168.1	13.2
20.0130	882.	1060.	70.	165.7	12.7
20.0230	880.	1056.	70.	171.0	11.3
20.0330	880.	1060.	70.	168.1	11.3
20.0430	880.	1060.	70.	168.1	13.2
20.0530	880.	1060.	70.	168.1	15.0
20.0630	889.	1060.	70.	168.1	11.3
20.0730	886.	1060.	70.	167.7	10.9
20.0830	879.	1060.	70.	166.9	21.3
20.0930	880.	1062.	70.	168.1	14.5
20.1030	882.	1060.	70.	168.1	10.9
20.1130	877.	1060.	70.	168.6	10.9
20.1230	866.	1060.	70.	168.1	15.4
20.1330	865.	1060.	70.	169.4	14.5
20.1430	872.	1061.	70.	168.6	8.6
20.1530	873.	1060.	70.	169.0	8.2
20.1630	870.	1065.	70.	168.6	16.3
20.1730	874.	1060.	70.	169.0	20.4
20.1830	884.	1080.	70.	169.0	20.9
20.1930	888.	1087.	70.	169.4	23.6

APPENDIX C: TABLE I.

RUN 6: TEMPERATURES AND FEED RATES PAGE 11 OF 12

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
20.2030	883.	1072.	70.	171.4	16.8
20.2130	888.	1072.	70.	176.0	14.5
20.2230	895.	1072.	70.	169.4	19.1
20.2330	898.	1070.	70.	169.8	18.6
21.0030	918.	1070.	70.	171.9	13.2
21.0130	902.	1069.	70.	171.4	20.4
21.0230	902.	1065.	70.	171.0	22.2
21.0330	895.	1062.	70.	171.9	13.6
21.0430	897.	1071.	70.	169.0	18.6
		STONE CHANGE			
21.0530	905.	1067.	70.	173.1	13.6
21.0630	905.	1070.	70.	171.0	0.
21.0730	900.	1060.	70.	170.2	0.
21.0830	889.	1049.	70.	176.0	0.
21.0930	885.	1063.	70.	171.9	13.2
21.1030	884.	1060.	70.	171.4	14.5
21.1130	890.	1058.	70.	172.7	14.5
21.1230	890.	1053.	70.	173.5	12.2
21.1330	890.	1080.	70.	171.9	18.6
21.1430	891.	1060.	70.	173.1	17.7
21.1530	886.	1050.	70.	173.9	7.7
21.1630	884.	1060.	70.	172.3	15.0
21.1730	880.	1062.	70.	172.7	16.3
21.1830	882.	1067.	70.	171.0	18.1
21.1930	872.	1068.	70.	170.6	17.2
21.2030	874.	1050.	70.	171.4	24.0
21.2130	871.	1061.	70.	172.7	15.0
21.2230	885.	1079.	70.	172.3	17.7
21.2330	881.	1078.	70.	171.9	16.3
22.0030	875.	1079.	70.	171.9	15.9
22.0130	871.	1079.	70.	171.4	15.4
22.0230	870.	1064.	70.	171.0	17.7
22.0330	872.	1064.	70.	171.9	17.2
22.0430	874.	1061.	70.	169.0	16.3
22.0530	882.	1062.	70.	173.1	14.5
22.0630	882.	1060.	70.	171.0	13.6
22.0730	880.	1060.	70.	170.2	15.0
22.0830	882.	1061.	73.	171.0	15.4
22.0930	880.	1062.	74.	172.3	13.2
22.1030	878.	1049.	74.	171.4	12.2

APPENDIX C: TABLE I.

RUN 6: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C.			FEED RATE KG/HR	
	GASIFIER	REGEN.	RECYCLE	OIL	STONE
22.1130	880.	1058.	72.	172.3	4.5
22.1230	891.	1060.	72.	170.6	0.
22.1330	888.	1056.	72.	171.9	0.
22.1430	893.	1052.	75.	171.9	0.
22.1530	877.	1050.	80.	172.7	10.0
22.1630	896.	1050.	75.	170.2	13.6
22.1730	912.	1042.	75.	169.4	15.9

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
1.2230	435.	153.	3.4	30.7	11.7	1.35
1.2330	409.	163.	3.4	31.9	10.3	1.76
2.0030	421.	163.	3.4	31.0	11.6	1.80
2.0130	421.	163.	3.4	31.0	4.9	1.63
2.0230	426.	163.	3.4	30.8	4.9	1.62
2.0330	413.	163.	3.4	33.6	3.3	1.67
2.0430	401.	173.	3.3	34.0	3.3	1.69
2.0530	401.	173.	3.3	34.0	3.0	1.68
2.0630	401.	153.	3.3	33.5	3.3	1.67
2.0730	402.	153.	3.3	32.7	3.6	1.65
2.0830	410.	147.	3.0	33.0	2.7	1.61
2.0930	410.	148.	3.0	33.0	3.3	1.64
2.1030	418.	124.	3.0	33.2	1.3	1.56
2.1130	419.	159.	3.0	32.9	1.4	1.55
2.1230	408.	144.	3.0	32.3	1.4	1.53
2.1330	400.	144.	3.0	31.9	1.3	1.50
2.1430	391.	153.	3.0	33.3	1.3	1.57
2.1530	400.	153.	3.0	34.3	1.5	1.62
2.1630	408.	148.	3.0	34.6	1.4	1.64
2.1730	409.	144.	3.0	35.6	1.5	1.68
2.1830	429.	144.	3.0	36.1	1.5	1.71
2.1930	439.	134.	3.0	35.6	1.6	1.68
2.2030	429.	134.	3.0	35.4	1.9	1.69
2.2130	437.	144.	3.0	35.4	1.6	1.65
2.2230	437.	144.	3.0	34.9	3.1	1.68
2.2330	436.	144.	3.0	35.0	1.8	1.66
3.0030	436.	143.	3.0	36.0	1.5	1.70
3.0130	437.	143.	3.0	36.9	1.0	1.72
3.0230	437.	143.	3.0	36.8	2.3	1.78
3.0330	438.	143.	3.0	37.3	1.7	1.77
3.0430	437.	143.	3.0	37.3	1.7	1.76
3.0530	472.	143.	3.0	37.7	2.1	1.80
3.0630	455.	143.	3.0	39.5	1.1	1.85
3.0730	454.	143.	3.1	38.2	1.2	1.78
3.0830	453.	143.	3.1	38.3	1.6	1.81
3.0930	454.	124.	3.1	38.4	1.7	1.82
3.1030	455.	124.	3.3	38.7	1.8	1.84
3.1130	453.	124.	3.3	38.4	1.6	1.82
3.1230	462.	124.	3.3	38.1	1.6	1.80
3.1330	463.	114.	3.3	37.9	1.4	1.79

APPENDIX C: TABLE II.
RUN 6: GAS FLOW RATES PAGE 2 OF 12

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
3.1430	456.	115.	3.3	38.3	1.3	1.81
3.1530	462.	105.	3.3	37.9	1.1	1.80
3.1630	463.	115.	3.3	37.9	1.1	1.81
3.1730	464.	115.	3.3	37.4	1.0	1.78
3.1830	472.	115.	3.0	37.2	1.1	1.78
3.1930	471.	114.	3.0	36.8	1.1	1.76
3.2030	472.	114.	3.0	37.0	1.0	1.77
3.2130	471.	105.	3.0	37.1	1.1	1.77
3.2230	471.	105.	3.0	36.7	0.9	1.75
3.2330	471.	109.	3.0	37.1	0.9	1.76
4.0030	471.	109.	3.0	36.9	1.0	1.76
4.0130	472.	109.	3.0	36.5	1.2	1.74
4.0230	455.	201.	3.0	36.0	1.1	1.72
4.0330	472.	29.	3.0	35.9	1.2	1.73
4.0430	473.	135.	3.0	35.9	1.2	1.72
4.0530	464.	125.	3.0	35.9	1.3	1.73
4.0630	455.	125.	3.0	36.1	2.1	1.77
4.0730	439.	125.	3.0	36.4	1.8	1.78
4.0830	438.	125.	3.0	36.4	1.3	1.75
4.0930	447.	125.	3.0	37.3	1.6	1.81
4.1030	446.	135.	3.0	37.1	1.4	1.80
4.1130	437.	135.	3.0	37.6	1.3	1.81
4.1230	402.	184.	3.0	37.9	1.3	1.82
4.1330	401.	175.	3.0	38.0	1.2	1.82
4.1430	410.	165.	3.0	38.0	1.3	1.83
4.1530	453.	125.	3.0	37.7	1.3	1.81
4.1630	445.	125.	3.0	37.0	1.3	1.78
4.1730	445.	125.	3.1	36.7	1.2	1.76
4.1830	454.	125.	3.0	37.0	1.2	1.77
4.1930	436.	125.	3.1	37.0	1.3	1.78
4.2030	446.	125.	3.1	37.2	1.3	1.79
4.2130	455.	125.	3.1	37.2	1.3	1.79
4.2230	455.	125.	3.1	37.2	1.3	1.79
4.2330	456.	125.	3.1	36.2	1.2	1.74
5.0030	474.	124.	3.1	33.4	1.0	1.59
5.0130	447.	124.	3.1	33.3	1.0	1.59
5.0230	446.	124.	3.0	33.0	1.1	1.58
5.0330	446.	124.	3.0	32.8	1.2	1.58
5.0430	455.	137.	3.1	34.9	1.1	1.66
5.0530	446.	147.	3.1	34.0	1.1	1.63

APPENDIX C: TABLE II.

RUN 6: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
5.0630	411.	147.	3.1	34.1	1.9	1.67
5.0730	422.	145.	3.1	33.5	0.9	1.59
5.0830	453.	145.	3.1	31.5	0.9	1.50
5.0930	467.	136.	3.4	31.5	0.9	1.50
5.1030	453.	135.	3.4	32.6	0.9	1.55
5.1130	468.	135.	3.4	32.6	0.9	1.56
5.1230	468.	135.	3.4	31.6	1.9	1.54
5.1330	468.	137.	3.4	36.4	1.2	1.75
5.1430	463.	139.	3.4	40.6	0.9	1.93
5.1530	468.	141.	3.4	37.5	0.9	1.78
5.1630	454.	135.	3.4	38.0	0.9	1.80
5.1730	454.	141.	3.4	37.7	1.2	1.81
5.1830	463.	141.	3.4	39.6	0.6	1.87
5.1930	454.	141.	3.4	36.7	0.9	1.75
5.2030	454.	141.	3.4	37.7	0.9	1.79
5.2130	446.	137.	3.4	38.0	1.1	1.81
5.2230	445.	141.	3.5	37.7	1.1	1.80
5.2330	445.	145.	3.5	38.0	0.9	1.81
6.0030	446.	141.	3.5	38.1	0.9	1.81
6.0130	445.	141.	3.5	37.5	0.9	1.78
6.0230	437.	135.	3.5	37.5	0.9	1.78
6.0330	438.	137.	3.5	37.5	0.9	1.78
6.0430	438.	141.	3.5	36.7	1.1	1.75
6.0530	439.	135.	3.5	37.6	1.1	1.80
6.0630	456.	135.	3.5	36.8	1.1	1.76
6.0730	456.	135.	3.5	37.1	1.1	1.77
6.0830	455.	133.	3.5	36.7	1.1	1.75
6.0930	455.	133.	3.5	36.7	1.1	1.75
6.1030	469.	131.	3.5	37.8	1.1	1.81
6.1130	463.	125.	3.5	38.1	1.1	1.82
6.1230	455.	116.	3.5	36.4	1.1	1.74
6.1330	455.	116.	3.5	37.3	1.1	1.78
6.1430	455.	119.	3.4	37.7	1.1	1.80
6.1530	455.	121.	3.4	36.0	1.1	1.72
6.1630	455.	121.	3.4	33.7	1.1	1.62
6.1730	455.	121.	3.5	34.1	0.8	1.62
6.1830	469.	109.	3.5	34.1	1.1	1.63
6.1930	468.	104.	3.5	33.4	1.1	1.60
6.2030		MISSED DATA READING				
6.2130	464.	110.	3.5	33.5	1.4	1.61

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
6.2230	438.	97.	3.5	33.5	1.3	1.61
6.2330	421.	121.	3.5	32.8	1.1	1.53
7.0030	448.	131.	3.5	29.2	1.3	1.39
7.0130	456.	101.	3.5	27.3	1.5	1.33
7.0230	448.	121.	3.5	28.8	1.5	1.40
7.0330	447.	121.	3.5	30.2	1.5	1.46
7.0430	465.	108.	3.5	30.4	1.3	1.45
7.0530	461.	115.	3.5	31.0	1.3	1.49
7.0630	443.	119.	3.4	31.0	1.2	1.48
7.0730	456.	117.	3.4	30.7	1.2	1.47
7.0830	461.	128.	3.4	30.2	1.2	1.45
7.0930	460.	115.	3.4	30.3	1.3	1.46
7.1030	460.	114.	3.3	30.4	1.5	1.47
7.1130	452.	114.	3.1	30.1	1.1	1.44
7.1230	452.	114.	3.1	31.4	1.8	1.53
7.1330	466.	114.	3.1	31.4	2.0	1.53
7.1430	460.	114.	3.1	33.7	2.7	1.68
7.1530	452.	116.	3.1	32.7	2.1	1.61
7.1630	452.	114.	3.1	33.3	1.8	1.62
7.1730	452.	114.	3.1	32.3	1.5	1.55
7.1830	452.	114.	3.1	32.0	1.5	1.54
7.1930	452.	114.	3.2	32.3	1.8	1.57
7.2030	452.	114.	3.2	31.7	1.5	1.53
7.2130	443.	115.	3.1	32.2	1.8	1.57
7.2230	443.	117.	3.2	32.0	1.8	1.56
7.2330	426.	115.	3.2	32.1	1.5	1.55
8.0030	434.	115.	3.2	30.8	1.7	1.50
8.0130	417.	138.	3.2	30.3	1.8	1.48
8.0230	414.	144.	3.2	30.1	1.8	1.46
8.0330	409.	140.	3.2	28.8	1.5	1.39
8.0430	418.	138.	3.2	29.1	2.1	1.44
8.0530	418.	140.	3.2	28.8	1.8	1.38
8.0630	417.	140.	3.2	28.6	1.8	1.40
8.0730	417.	138.	3.2	28.6	2.1	1.42
8.0830	417.	138.	3.2	28.3	1.9	1.39
8.0930	416.	135.	3.2	28.6	2.2	1.42
8.1030	415.	135.	3.2	30.6	2.0	1.51
8.1130	412.	135.	3.3	32.2	2.1	1.58
8.1230	410.	135.	3.3	32.2	2.0	1.58
8.1330	410.	135.	3.2	32.7	2.2	1.60

APPENDIX C: TABLE II.
RUN 6: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
8.1430	403.	135.	3.2	32.8	1.6	1.59
8.1530	403.	125.	3.2	33.3	2.1	1.63
8.1630	415.	125.	3.2	32.5	1.8	1.58
8.1730	364.	154.	3.2	29.8	1.7	1.45
8.1830	347.	105.	3.2	31.2	1.8	1.53
8.1930	352.	105.	3.2	30.4	1.9	1.50
8.2030	357.	105.	3.2	30.9	1.8	1.51
8.2130	383.	115.	3.2	30.5	1.6	1.49
8.2230	383.	114.	3.3	29.8	1.6	1.45
8.2330	383.	115.	3.2	31.6	1.7	1.54
9.0030	383.	125.	3.1	32.5	1.6	1.58
9.0130	366.	134.	3.2	32.8	1.6	1.58
9.0230	365.	144.	3.2	32.5	1.6	1.57
9.0330	366.	135.	3.2	32.1	1.6	1.55
9.0430	366.	135.	3.2	32.4	1.6	1.56
9.0530	374.	164.	3.2	32.3	1.6	1.56
9.0630	366.	135.	3.2	32.3	1.6	1.56
9.0730	365.	135.	3.2	33.1	1.6	1.59
9.0830	366.	135.	3.2	36.1	2.0	1.75
9.0930	371.	135.	3.3	29.2	1.7	1.42
9.1030	371.	125.	3.3	32.8	1.7	1.59
9.1130	366.	145.	3.3	31.8	1.7	1.54
9.1230	375.	145.	3.3	30.7	1.6	1.50
9.1330	374.	135.	3.2	30.2	1.7	1.46
9.1430	374.	135.	3.2	30.7	1.6	1.48
9.1530	374.	135.	3.2	30.8	1.6	1.48
9.1630	374.	135.	3.2	30.8	1.7	1.50
9.1730	400.	135.	3.2	29.4	1.8	1.42
9.1830	391.	135.	3.2	27.6	1.7	1.32
9.1930	400.	135.	3.3	28.2	1.3	1.33
9.2030	383.	135.	3.2	29.7	1.5	1.41
9.2130	375.	144.	3.2	30.2	1.8	1.43
9.2230	375.	143.	3.2	30.7	1.8	1.45
SHUT DOWN AT 9.2230 FOR 23 HOURS						
10.2130	403.	-	3.2	28.6	3.1	1.46
10.2230	412.	134.	3.2	32.6	2.0	1.60
10.2330	403.	144.	3.3	33.7	1.7	1.64

APPENDIX C: TABLE II.
RUN 6: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
11.0030	411.	135.	3.3	33.4	1.7	1.63
11.0130	412.	135.	3.3	33.1	1.9	1.64
11.0230	411.	135.	3.3	33.7	1.7	1.66
11.0330	412.	125.	3.3	35.4	1.6	1.73
11.0430	411.	125.	3.2	31.4	1.6	1.54
11.0530	411.	125.	3.2	33.0	1.3	1.61
11.0630	411.	125.	3.2	33.2	1.6	1.63
11.0730	420.	125.	3.2	32.7	1.6	1.61
11.0830	402.	115.	3.2	33.2	1.4	1.62
11.0930	401.	116.	3.2	33.3	1.3	1.62
11.1030	402.	115.	3.2	33.1	1.4	1.61
11.1130	407.	105.	3.2	33.1	1.4	1.61
11.1230	405.	109.	3.2	33.5	1.7	1.65
11.1330	384.	85.	3.2	31.3	1.8	1.52
11.1430	427.	135.	3.2	26.6	1.6	1.31
11.1530	427.	135.	3.2	26.0	1.5	1.28
11.1630	427.	139.	3.2	24.9	1.5	1.22
11.1730	437.	139.	3.2	29.0	1.4	1.40
11.1830	436.	139.	3.2	24.3	1.4	1.21
11.1930	427.	134.	3.2	28.2	1.5	1.39
11.2030	427.	135.	3.2	31.7	1.5	1.54
11.2130	419.	145.	3.2	33.5	1.5	1.64
11.2230	411.	161.	3.2	29.1	1.8	1.43
11.2330	402.	161.	3.2	29.5	1.2	1.42
12.0030	394.	164.	3.2	27.4	1.2	1.32
12.0130	416.	153.	3.2	27.0	2.5	1.38
12.0230	402.	155.	3.2	27.3	1.9	1.36
12.0330	419.	154.	3.2	28.6	2.9	1.47
12.0430	416.	156.	3.2	29.6	0.6	1.39
12.0530	419.	156.	3.2	29.9	0.6	1.42
12.0630	416.	154.	3.1	29.0	1.6	1.42
12.0730	402.	154.	3.1	29.3	1.2	1.41
12.0830	406.	154.	3.1	28.1	1.6	1.37
12.0930	409.	151.	3.2	28.3	1.5	1.40
12.1030	427.	135.	3.2	28.9	2.0	1.45
12.1130	417.	155.	3.2	30.2	1.4	1.46
12.1230	418.	145.	3.2	29.9	1.2	1.43
12.1330	418.	145.	3.2	30.8	1.2	1.48
12.1430	409.	145.	3.2	30.7	1.2	1.47
12.1530	410.	145.	3.2	31.3	1.4	1.49

APPENDIX C: TABLE II.
RUN 6: GAS FLOW RATES PAGE 7 OF 12

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
12.1630	392.	165.	3.2	30.3	1.5	1.48
12.1730	367.	179.	3.2	31.1	1.2	1.50
12.1830	375.	165.	3.2	30.2	1.4	1.47
12.1930	383.	165.	3.2	30.1	1.9	1.49
12.2030	375.	145.	3.2	29.3	1.6	1.44
12.2130	384.	165.	3.2	27.7	1.8	1.37
12.2230	376.	165.	3.2	30.9	1.6	1.51
12.2330	384.	174.	3.2	29.7	1.2	1.44
13.0030	381.	164.	3.4	29.0	1.6	1.42
13.0130	381.	164.	3.4	31.2	0.9	1.49
13.0230	376.	180.	3.4	31.2	1.6	1.52
13.0330	381.	180.	3.4	30.9	1.2	1.47
13.0430	385.	184.	3.4	30.9	1.6	1.51
13.0530	385.	184.	3.4	30.3	1.2	1.46
13.0630	376.	184.	3.4	30.3	1.2	1.46
13.0730	385.	174.	3.4	29.3	1.2	1.41
13.0830	383.	176.	3.4	29.7	1.2	1.43
13.0930	385.	174.	3.4	29.2	1.2	1.42
13.1030	384.	174.	3.4	28.9	1.7	1.43
13.1130	384.	174.	3.4	28.7	1.8	1.43
13.1230	384.	174.	3.4	28.2	1.7	1.40
13.1330	401.	164.	3.4	28.9	1.8	1.43
13.1430	400.	164.	3.4	29.8	1.3	1.45
13.1530	400.	169.	3.4	28.2	1.2	1.37
13.1630	410.	174.	3.4	31.8	1.2	1.53
13.1730	401.	165.	3.4	31.9	1.1	1.53
13.1830	410.	165.	3.4	31.9	1.2	1.54
13.1930	389.	174.	3.4	31.8	1.2	1.53
13.2030	388.	164.	3.4	30.1	1.9	1.48
13.2130	405.	172.	3.4	30.1	2.5	1.52
13.2230	378.	174.	3.4	30.0	2.2	1.49
13.2330	386.	184.	3.4	30.2	2.5	1.51
14.0030	371.	188.	3.4	30.7	2.2	1.52
14.0130	374.	191.	3.4	30.0	2.5	1.51
14.0230	394.	190.	3.4	28.2	2.2	1.41
14.0330	372.	192.	3.4	29.1	1.5	1.42
14.0430	337.	190.	3.4	29.3	2.5	1.47
14.0530	380.	189.	3.3	29.1	2.2	1.45
14.0630	389.	174.	3.3	29.3	2.2	1.46
14.0730	389.	145.	3.4	27.7	2.1	1.38

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
14.0830	389.	144.	3.4	27.7	2.1	1.38
14.0930	387.	154.	3.1	26.9	2.3	1.35
14.1030	388.	183.	3.1	29.0	1.9	1.44
14.1130	380.	184.	3.1	30.7	1.9	1.49
14.1230	380.	184.	3.2	31.6	1.7	1.54
14.1330	371.	184.	3.2	31.1	1.6	1.52
14.1430	371.	184.	3.2	31.2	1.8	1.52
14.1530	380.	174.	3.2	31.1	1.7	1.51
14.1630	379.	194.	3.2	31.0	1.7	1.50
14.1730	379.	194.	3.3	31.2	1.7	1.51
14.1830	380.	194.	3.3	31.9	1.5	1.54
14.1930	380.	185.	3.2	31.2	1.7	1.52
14.2030	380.	175.	3.2	31.2	1.7	1.51
14.2130	371.	175.	3.2	32.7	5.0	1.75
14.2230	352.	184.	3.3	33.7	8.0	1.91
14.2330	387.	184.	3.3	33.9	7.8	1.90
15.0030	385.	184.	3.3	33.0	6.5	1.84
15.0130	381.	174.	3.3	34.2	6.0	1.87
15.0230	342.	184.	3.3	32.0	6.5	1.80
15.0330	413.	184.	3.3	34.8	8.7	1.98
15.0430	413.	174.	3.3	35.4	8.6	2.01
15.0530	397.	174.	3.3	34.7	5.2	1.82
15.0630	397.	184.	3.3	34.0	4.8	1.81
15.0730	414.	164.	3.2	33.9	4.8	1.81
15.0830	414.	174.	3.2	34.2	5.3	1.82
15.0930	414.	175.	3.3	34.4	5.0	1.78
15.1030	413.	165.	3.3	33.8	4.2	1.72
15.1130	395.	175.	3.3	33.8	4.5	1.73
15.1230	395.	174.	2.6	33.7	4.6	1.73
15.1330	387.	175.	3.3	32.0	1.5	1.53
15.1430	386.	175.	3.3	31.7	1.6	1.52
15.1530	387.	175.	3.3	31.6	1.6	1.52
15.1630	388.	175.	3.3	31.5	1.8	1.52
15.1730	387.	175.	3.3	31.7	1.4	1.51
15.1830	387.	165.	3.2	30.8	1.6	1.49
15.1930	410.	155.	3.2	30.8	1.5	1.49
15.2030	388.	165.	3.2	30.8	1.6	1.49
15.2130	387.	165.	3.2	31.0	1.4	1.49
15.2230	377.	165.	3.2	31.0	1.4	1.48
15.2330	378.	155.	3.2	30.7	1.3	1.47

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
16.0030	378.	155.	3.2	30.6	1.4	1.47
16.0130	387.	165.	3.2	30.5	1.4	1.46
16.0230	395.	155.	3.2	30.6	1.4	1.47
16.0330	386.	155.	3.2	30.3	1.5	1.46
16.0430	379.	161.	3.2	30.1	1.4	1.44
16.0530	370.	174.	3.2	30.6	1.2	1.46
16.0630	378.	165.	3.2	30.4	1.6	1.47
16.0730	377.	165.	3.1	30.7	1.4	1.47
16.0830	378.	165.	3.1	30.7	1.4	1.48
16.0930	379.	164.	3.2	30.6	1.5	1.47
16.1030	388.	174.	3.2	30.8	1.4	1.47
16.1130	389.	185.	3.2	30.4	1.4	1.46
16.1230	389.	185.	3.2	30.4	1.4	1.47
16.1330	387.	185.	3.2	30.4	1.4	1.47
16.1430	386.	175.	3.2	30.4	1.4	1.46
16.1530	378.	185.	3.2	30.7	1.3	1.47
16.1630	377.	194.	3.2	31.4	1.3	1.51
16.1730	377.	194.	3.2	31.7	1.5	1.54
16.1830	376.	194.	3.2	32.7	1.6	1.58
16.1930	369.	185.	3.2	33.2	1.5	1.60
16.2030	369.	194.	3.2	33.5	1.6	1.62
16.2130	352.	185.	3.2	33.5	1.5	1.61
16.2230	361.	185.	3.2	33.5	1.4	1.61
16.2330	361.	185.	3.2	33.6	1.8	1.64
17.0030	362.	185.	3.2	33.7	2.0	1.66
17.0130	362.	68.	3.2	34.9	1.7	1.67
17.0230	363.	60.	3.2	35.4	1.6	1.70
17.0330	353.	49.	3.2	35.4	1.5	1.70
17.0430	352.	37.	3.2	35.5	1.6	1.70
17.0530	361.	31.	3.2	35.2	1.6	1.69
17.0630	358.	25.	3.2	35.2	1.6	1.69
17.0730	338.	24.	3.2	35.3	1.4	1.68
17.0830	393.	28.	3.2	34.8	1.5	1.66
17.0930	376.	18.	3.2	33.9	1.8	1.63
17.1030	375.	193.	3.2	33.9	1.8	1.63
17.1130	375.	185.	3.2	33.9	1.8	1.62

SHUT DOWN AT 17.1130 FOR 40 HOURS

APPENDIX C: TABLE II.
RUN 6: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
19.0430	346.	179.	3.4	31.9	1.0	1.53
19.0530	349.	189.	3.4	31.4	1.0	1.51
19.0630	363.	179.	3.4	30.8	1.4	1.50
19.0730	354.	194.	3.4	32.2	1.0	1.55
19.0830	354.	175.	3.4	31.6	1.0	1.51
19.0930	328.	177.	3.4	31.9	1.1	1.54
19.1030	363.	185.	3.4	31.3	1.3	1.50
19.1130	363.	185.	3.4	30.5	1.3	1.46
19.1230	368.	175.	3.4	31.7	1.5	1.52
19.1330	368.	165.	3.4	29.6	1.5	1.43
19.1430	380.	165.	3.4	30.5	1.5	1.47
19.1530	372.	161.	3.4	30.2	1.8	1.48
19.1630	372.	161.	3.4	31.7	1.8	1.54
19.1730	372.	185.	3.4	30.9	2.4	1.53
19.1830	351.	181.	3.4	31.1	1.8	1.51
19.1930	372.	165.	3.4	28.0	2.1	1.38
19.2030	368.	181.	3.4	27.7	1.5	1.34
19.2130	360.	177.	3.4	28.2	1.5	1.36
19.2230	360.	185.	3.4	28.1	1.6	1.36
19.2330	354.	175.	3.4	29.1	1.7	1.41
20.0030	358.	175.	3.4	31.1	1.8	1.51
20.0130	358.	175.	3.4	30.5	1.8	1.48
20.0230	358.	175.	3.4	31.8	1.6	1.53
20.0330	350.	175.	3.4	33.1	2.2	1.62
20.0430	350.	175.	3.4	30.9	2.0	1.51
20.0530	356.	175.	3.4	32.3	2.0	1.57
20.0630	350.	171.	3.4	33.1	1.7	1.59
20.0730	359.	171.	3.4	33.1	2.4	1.63
20.0830	359.	165.	3.4	32.8	2.6	1.63
20.0930	348.	165.	3.2	33.0	2.3	1.62
20.1030	339.	165.	3.2	34.1	1.9	1.65
20.1130	356.	165.	3.2	34.5	1.9	1.67
20.1230	348.	155.	3.2	32.5	4.2	1.69
20.1330	347.	155.	3.2	31.8	2.1	1.56
20.1430	346.	155.	3.2	32.0	1.9	1.56
20.1530	336.	155.	3.2	31.6	1.6	1.53
20.1630	354.	165.	3.3	33.0	2.1	1.62
20.1730	354.	175.	3.3	40.5	1.7	1.93
20.1830	388.	155.	3.3	40.2	1.8	1.95
20.1930	379.	155.	3.3	35.7	2.0	1.76

APPENDIX C: TABLE II.
RUN 6: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
20.2030	385.	155.	3.3	29.7	1.6	1.45
20.2130	390.	155.	3.4	35.1	1.6	1.70
20.2230	408.	155.	3.1	26.4	2.0	1.31
20.2330	392.	155.	3.3	29.5	1.3	1.42
21.0030	391.	155.	3.2	32.3	2.0	1.58
21.0130	393.	155.	3.2	29.9	1.3	1.44
21.0230	374.	165.	3.2	35.0	2.0	1.70
21.0330	375.	175.	3.2	31.8	1.3	1.52
21.0430	384.	175.	3.2	34.2	1.4	1.64
		STONE CHANGE				
21.0530	383.	175.	3.2	35.3	1.8	1.71
21.0630	374.	185.	3.2	34.7	1.5	1.67
21.0730	382.	175.	3.2	34.8	1.6	1.67
21.0830	356.	175.	3.2	31.2	1.5	1.49
21.0930	371.	175.	3.2	30.9	1.6	1.49
21.1030	380.	165.	3.2	33.6	2.1	1.63
21.1130	379.	155.	3.2	31.6	1.9	1.53
21.1230	388.	155.	3.2	30.2	1.8	1.46
21.1330	389.	155.	3.2	30.0	1.7	1.47
21.1430	389.	155.	3.2	35.5	2.0	1.72
21.1530	371.	165.	3.2	35.2	1.7	1.68
21.1630	355.	175.	3.2	35.2	1.6	1.68
21.1730	355.	175.	3.2	36.7	1.6	1.76
21.1830	356.	175.	3.2	36.9	1.9	1.79
21.1930	354.	175.	3.2	35.0	1.9	1.70
21.2030	364.	175.	3.2	32.1	1.8	1.54
21.2130	365.	175.	3.2	31.2	1.6	1.50
21.2230	374.	175.	3.2	30.7	1.6	1.50
21.2330	374.	175.	3.2	33.2	1.5	1.61
22.0030	374.	194.	3.2	32.3	1.4	1.56
22.0130	358.	185.	3.2	29.9	1.5	1.46
22.0230	374.	185.	3.2	35.0	1.4	1.67
22.0330	373.	185.	3.2	31.8	1.6	1.53
22.0430	356.	185.	3.2	32.2	1.5	1.54
22.0530	357.	185.	3.2	35.3	1.8	1.70
22.0630	357.	185.	3.2	34.7	1.7	1.67
22.0730	357.	185.	3.2	34.8	1.6	1.67
22.0830	366.	176.	3.2	34.8	1.3	1.65
22.0930	338.	176.	3.2	35.5	1.6	1.70
22.1030	338.	176.	3.2	34.5	1.4	1.63

APPENDIX C: TABLE II.

RUN 6: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
22.1130	354.	175.	3.2	29.4	2.2	1.45
22.1230	354.	175.	3.2	26.7	1.4	1.29
22.1330	337.	165.	3.2	31.7	1.8	1.53
22.1430	336.	166.	3.2	32.5	2.1	1.57
22.1530	344.	169.	3.2	34.6	1.6	1.64
22.1630	374.	166.	3.2	32.2	1.4	1.52
22.1730	375.	166.	3.2	31.6	1.3	1.48

APPENDIX C: TABLE III.

RUN 6: PRESSURES

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	GASIFIER P. KILOPASCALS DISTRIB. D.P.	GASIFIER P. KILOPASCALS BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
1.2230	4.5	4.0	5.5	0.90	6.0
1.2330	4.5	4.0	5.2	0.90	6.0
2.0030	4.5	4.0	5.2	0.90	6.0
2.0130	4.5	4.0	5.2	1.00	6.0
2.0230	4.5	4.0	5.1	0.95	6.0
2.0330	4.5	4.0	5.2	1.00	6.0
2.0430	4.5	4.0	5.1	0.90	6.0
2.0530	4.4	3.2	5.2	0.95	6.0
2.0630	4.5	3.6	5.5	1.00	6.0
2.0730	4.4	3.5	5.7	0.95	6.0
2.0830	4.4	3.5	5.4	1.00	6.0
2.0930	4.4	3.4	5.4	0.95	6.0
2.1030	4.4	3.5	5.5	0.95	6.2
2.1130	4.3	3.5	5.5	0.98	6.2
2.1230	4.2	3.5	5.5	1.00	6.2
2.1330	4.2	3.4	5.6	1.00	6.2
2.1430	4.2	3.4	5.6	1.00	6.0
2.1530	4.2	3.5	5.3	1.00	5.7
2.1630	3.6	3.5	5.1	1.00	5.5
2.1730	4.1	3.4	5.2	1.00	5.5
2.1830	4.4	3.6	5.1	0.95	6.0
2.1930	4.4	3.7	5.2	1.00	5.5
2.2030	4.5	3.7	5.1	1.00	5.7
2.2130	4.5	3.9	5.4	1.00	7.2
2.2230	4.5	3.7	5.5	1.00	7.5
2.2330	4.5	3.9	5.5	1.00	7.0
3.0030	4.5	3.9	5.5	1.00	6.5
3.0130	4.5	3.9	5.6	1.00	7.0
3.0230	4.5	3.9	5.6	1.00	6.7
3.0330	4.5	3.9	5.7	1.00	7.0
3.0430	4.6	3.9	5.8	1.00	7.2
3.0530	4.7	4.2	6.0	1.00	7.5
3.0630	4.6	4.0	5.6	1.00	7.5
3.0730	4.4	3.9	5.8	1.00	7.0
3.0830	4.4	3.9	6.1	1.00	7.0
3.0930	4.3	3.9	5.8	0.95	6.5
3.1030	4.2	3.9	6.0	1.00	6.5
3.1130	4.3	4.0	6.1	1.00	6.5
3.1230	4.4	4.0	6.2	1.00	6.5
3.1330	4.2	4.0	6.2	1.00	6.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
3.1430	4.2	3.9	6.2	1.00	6.2
3.1530	4.2	3.9	6.3	1.00	6.2
3.1630	4.2	3.9	6.3	1.00	6.3
3.1730	4.2	3.7	6.2	1.00	6.5
3.1830	4.2	4.0	6.2	1.00	6.7
3.1930	4.2	4.0	6.2	1.00	6.7
3.2030	4.4	4.0	6.2	1.00	6.5
3.2130	4.4	4.0	6.3	1.00	6.5
3.2230	4.4	4.0	6.5	1.00	6.5
3.2330	4.3	4.0	6.3	1.00	6.7
4.0030	4.4	4.0	6.3	1.00	6.7
4.0130	4.4	4.0	6.5	1.00	7.0
4.0230	4.2	4.0	6.4	1.00	6.7
4.0330	3.9	4.1	6.5	1.00	7.0
4.0430	4.4	6.5	6.5	1.10	7.0
4.0530	4.2	5.8	6.5	1.10	7.2
4.0630	4.2	5.5	6.3	1.00	7.0
4.0730	4.2	5.4	6.2	1.00	6.7
4.0830	4.1	5.5	6.2	1.00	7.0
4.0930	4.2	5.5	6.2	1.00	6.5
4.1030	4.2	5.5	6.2	1.00	6.5
4.1130	4.1	5.5	6.2	1.05	6.5
4.1230	4.1	5.4	6.2	1.10	6.0
4.1330	4.2	5.4	6.2	1.10	6.0
4.1430	4.1	5.5	6.0	1.10	6.2
4.1530	4.2	5.7	6.1	1.09	6.5
4.1630	4.2	5.6	6.2	1.05	6.7
4.1730	4.2	5.6	6.3	1.05	6.7
4.1830	4.4	5.7	6.2	1.05	6.7
4.1930	4.4	5.7	6.2	1.03	6.7
4.2030	4.4	5.6	6.2	1.05	6.7
4.2130	4.4	5.5	6.3	1.05	6.7
4.2230	4.4	5.5	6.5	1.10	6.7
4.2330	4.4	5.5	6.5	1.10	6.7
5.0030	4.4	5.7	6.6	1.00	7.2
5.0130	4.4	5.4	6.1	1.00	6.7
5.0230	4.4	5.5	6.1	1.00	7.2
5.0330	4.5	5.6	6.1	1.00	7.0
5.0430	4.7	6.2	5.8	1.00	7.2
5.0530	4.4	5.5	6.2	1.05	7.0

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
5.0630	4.4	5.5	6.3	1.05	7.0
5.0730	4.4	5.8	6.6	1.05	8.0
5.0830	4.4	6.2	6.6	1.10	9.0
5.0930	4.4	6.2	6.6	1.10	9.0
5.1030	4.4	6.2	6.6	1.10	9.0
5.1130	4.4	6.3	6.5	1.10	8.5
5.1230	5.2	5.8	6.5	1.00	7.0
5.1330	4.2	6.0	6.2	1.00	6.2
5.1430	4.2	6.2	6.0	1.00	6.7
5.1530	4.4	6.2	5.7	1.00	6.2
5.1630	4.4	6.2	6.0	1.00	6.2
5.1730	4.2	6.0	6.0	1.10	6.5
5.1830	4.2	6.1	6.0	1.05	6.0
5.1930	4.2	6.0	6.1	1.00	6.2
5.2030	4.2	6.0	6.0	1.05	6.0
5.2130	4.2	6.1	6.1	1.10	6.2
5.2230	4.2	6.1	6.0	1.05	6.2
5.2330	4.2	6.1	6.0	1.05	6.2
6.0030	4.2	6.0	6.0	1.05	7.5
6.0130	4.4	6.2	6.0	1.05	7.5
6.0230	4.2	6.1	6.0	1.05	7.0
6.0330	4.4	6.1	6.0	1.05	7.0
6.0430	4.2	6.1	6.2	1.05	7.0
6.0530	4.4	6.3	6.2	1.05	7.0
6.0630	4.5	6.2	6.2	1.00	7.0
6.0730	4.4	6.5	6.2	1.00	7.0
6.0830	4.4	6.2	6.2	1.05	7.0
6.0930	4.3	6.2	6.5	1.05	5.5
6.1030	4.2	6.2	6.0	1.05	7.0
6.1130	4.4	6.2	6.0	1.00	6.7
6.1230	4.4	6.2	5.7	1.05	7.0
6.1330	4.2	6.3	5.7	1.00	6.2
6.1430	4.2	6.3	6.0	1.10	6.2
6.1530	4.4	6.5	5.7	1.05	6.2
6.1630	4.2	6.2	5.6	1.10	5.7
6.1730	4.4	6.5	5.6	1.10	6.7
6.1830	4.4	6.5	5.7	1.00	6.2
6.1930	4.5	6.5	5.5	1.10	6.2
6.2030	MISSED DATA READING				
6.2130	4.6	6.2	5.5	1.10	6.0

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
6.2230	4.5	6.0	5.5	1.15	6.0
6.2330	4.5	6.1	6.0	1.20	6.5
7.0030	4.6	6.3	5.5	1.15	8.0
7.0130	4.6	6.3	5.5	1.10	8.0
7.0230	4.7	6.3	5.7	1.10	8.0
7.0330	4.6	6.3	5.7	1.05	7.5
7.0430	4.7	6.5	5.7	1.05	7.5
7.0530	4.7	6.5	5.7	1.05	7.5
7.0630	4.7	6.5	5.7	1.10	7.5
7.0730	4.7	6.5	5.7	1.10	7.5
7.0830	4.7	6.2	5.7	1.10	7.5
7.0930	4.7	6.5	5.7	1.10	8.0
7.1030	4.7	6.2	5.7	1.10	7.7
7.1130	4.7	6.5	5.7	1.10	7.7
7.1230	4.7	6.1	5.8	1.10	7.7
7.1330	4.7	6.2	5.6	0.90	6.5
7.1430	4.6	6.2	5.6	0.90	6.7
7.1530	4.7	6.0	5.6	0.90	6.7
7.1630	4.7	6.0	5.6	1.00	6.7
7.1730	4.6	6.0	5.7	1.00	6.7
7.1830	4.6	6.0	5.7	1.00	6.7
7.1930	4.6	6.0	5.7	0.90	6.7
7.2030	4.6	6.0	5.7	1.00	6.7
7.2130	4.7	6.0	5.7	1.00	6.7
7.2230	4.7	6.0	5.7	1.00	6.7
7.2330	5.0	6.0	5.7	1.00	6.7
8.0030	5.1	6.0	5.5	1.00	6.7
8.0130	4.9	6.2	5.2	0.95	6.7
8.0230	4.9	6.2	5.2	1.00	6.7
8.0330	5.0	6.2	5.2	1.00	6.7
8.0430	4.9	6.2	5.2	1.00	6.7
8.0530	7.3	6.2	5.2	1.00	7.0
8.0630	4.9	6.2	5.2	1.00	7.2
8.0730	5.0	6.2	5.2	1.00	7.2
8.0830	5.0	6.2	5.2	1.00	7.2
8.0930	5.0	6.3	5.2	1.00	7.2
8.1030	5.0	6.0	5.2	1.00	7.2
8.1130	5.0	5.8	5.0	1.00	7.2
8.1230	5.1	6.0	5.0	1.00	7.5
8.1330	5.0	6.0	5.2	1.00	7.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
8.1430	5.1	5.9	5.2	1.00	7.5
8.1530	5.0	5.7	5.2	1.00	8.5
8.1630	5.0	5.8	5.2	1.00	7.2
8.1730	4.7	5.6	5.2	1.00	6.5
8.1830	4.1	4.1	5.2	1.00	6.5
8.1930	4.2	4.2	5.2	1.00	6.7
8.2030	4.2	4.4	5.5	1.05	7.0
8.2130	4.6	5.1	5.5	1.05	6.7
8.2230	4.6	5.0	5.5	1.00	7.0
8.2330	5.0	4.9	5.6	1.05	7.0
9.0030	5.1	5.0	5.2	1.00	6.5
9.0130	5.1	5.1	5.2	1.00	6.5
9.0230	5.2	5.0	5.4	1.00	6.7
9.0330	5.2	5.0	5.4	1.00	6.7
9.0430	5.4	5.0	5.5	1.00	7.0
9.0530	5.5	5.0	5.2	1.00	7.0
9.0630	5.6	5.1	5.0	1.00	7.0
9.0730	5.7	5.1	5.0	1.05	6.7
9.0830	5.7	5.1	5.0	1.00	7.0
9.0930	5.5	5.4	5.1	1.00	6.2
9.1030	5.5	5.4	5.2	1.00	6.2
9.1130	5.5	5.2	5.2	1.00	6.5
9.1230	5.6	5.2	5.2	1.05	7.5
9.1330	5.6	5.4	5.4	1.00	7.5
9.1430	5.7	5.4	5.2	1.00	7.5
9.1530	5.7	5.4	5.0	1.05	7.7
9.1630	5.7	5.4	5.2	1.05	8.0
9.1730	6.2	5.7	5.5	1.05	8.0
9.1830	6.3	5.6	5.5	1.05	7.2
9.1930	6.5	5.6	5.5	1.00	7.5
9.2030	6.5	5.4	5.0	1.00	7.0
9.2130	6.6	5.5	4.7	1.05	7.0
9.2230	6.8	5.5	4.9	1.00	6.7
SHUT DOWN AT 9.2230 FOR 23 HOURS					
10.2130	3.2	6.0	5.0	1.00	6.5
10.2230	3.4	5.8	5.0	1.00	6.5
10.2330	3.2	5.8	5.1	1.05	6.5

APPENDIX C: TABLE III.
RUN 6: PRESSURES

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
11.0030	3.4	5.8	5.4	1.05	6.5
11.0130	3.2	5.8	5.5	1.05	6.7
11.0230	3.2	5.8	5.6	1.00	7.0
11.0330	3.2	5.7	5.8	1.05	7.0
11.0430	3.2	5.7	6.0	1.05	7.0
11.0530	3.1	5.7	5.8	1.05	7.0
11.0630	3.0	5.7	6.0	1.05	7.0
11.0730	3.1	5.8	6.0	1.05	7.0
11.0830	3.0	5.5	6.1	1.05	7.0
11.0930	3.0	5.5	6.0	1.05	6.7
11.1030	2.9	5.5	6.0	1.05	6.5
11.1130	2.9	5.5	6.0	1.05	7.0
11.1230	2.9	5.2	6.2	1.10	6.7
11.1330	2.9	5.1	6.1	1.05	7.0
11.1430	3.2	7.0	5.7	1.05	7.0
11.1530	3.2	6.8	6.0	1.03	7.0
11.1630	3.1	6.8	6.0	1.05	7.1
11.1730	3.1	7.0	6.0	1.05	6.7
11.1830	3.0	7.0	6.0	1.10	6.7
11.1930	3.5	7.0	6.0	1.14	8.2
11.2030	3.5	6.7	5.5	1.00	7.0
11.2130	3.4	7.0	6.0	1.10	6.6
11.2230	3.5	7.0	6.0	1.05	7.5
11.2330	3.4	7.0	6.0	1.05	7.5
12.0030	3.6	6.8	6.0	1.05	7.2
12.0130	3.5	6.8	6.0	1.00	7.2
12.0230	3.5	7.0	6.0	1.10	7.2
12.0330	3.7	7.2	5.7	1.00	7.2
12.0430	3.6	7.0	5.6	1.00	7.5
12.0530	3.6	7.2	5.6	1.10	7.0
12.0630	3.5	7.2	5.7	1.00	7.2
12.0730	3.6	7.2	5.6	1.10	7.7
12.0830	3.5	7.2	5.6	1.10	6.7
12.0930	3.5	7.1	6.0	1.00	7.5
12.1030	3.6	7.5	6.0	1.00	5.7
12.1130	3.6	7.7	5.7	1.00	5.5
12.1230	3.6	7.7	5.7	1.00	5.7
12.1330	3.6	7.7	5.5	1.00	6.7
12.1430	3.6	7.6	5.5	1.00	7.0
12.1530	3.6	7.7	5.5	1.00	7.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
12.1630	3.5	7.7	5.2	1.00	7.5
12.1730	3.1	7.2	5.2	1.00	6.5
12.1830	3.2	7.5	5.5	1.00	6.7
12.1930	3.2	7.2	5.5	1.00	6.7
12.2030	3.2	7.5	5.5	1.00	6.7
12.2130	3.5	7.5	5.5	1.00	6.7
12.2230	3.4	7.5	5.5	1.10	6.0
12.2330	3.4	7.5	5.2	1.10	6.0
13.0030	3.4	7.7	5.2	1.10	5.8
13.0130	3.5	7.7	5.1	1.10	5.7
13.0230	3.5	7.8	5.2	1.10	6.0
13.0330	3.5	7.7	5.2	1.00	5.7
13.0430	3.6	7.8	5.1	1.10	8.2
13.0530	3.5	8.0	5.0	1.05	8.2
13.0630	3.5	8.0	5.1	1.10	8.2
13.0730	3.5	8.0	5.2	1.10	8.5
13.0830	3.5	7.8	5.2	1.10	8.5
13.0930	3.4	7.8	5.5	1.05	8.5
13.1030	3.4	7.8	5.5	1.10	7.2
13.1130	3.2	7.8	5.5	1.10	7.5
13.1230	3.4	7.8	5.2	1.10	7.5
13.1330	3.4	8.1	5.5	1.10	7.5
13.1430	3.5	8.2	5.2	1.10	7.2
13.1530	3.5	8.2	5.2	1.10	7.0
13.1630	3.6	8.2	5.0	1.10	7.0
13.1730	3.6	8.2	4.7	1.10	7.0
13.1830	3.6	8.1	4.7	1.05	7.0
13.1930	3.5	8.0	4.9	1.05	7.0
13.2030	3.5	7.8	5.2	1.05	7.5
13.2130	3.4	7.8	5.2	1.00	7.5
13.2230	3.5	7.7	5.4	1.10	7.5
13.2330	3.5	7.5	5.5	1.05	8.7
14.0030	3.5	7.5	5.5	1.00	8.7
14.0130	3.5	7.5	5.5	1.05	8.7
14.0230	3.5	7.5	5.5	1.10	8.7
14.0330	3.6	7.2	5.5	1.10	7.7
14.0430	3.6	7.2	5.5	1.10	7.7
14.0530	3.6	7.5	5.5	1.10	6.6
14.0630	3.6	7.5	5.5	1.00	6.6
14.0730	3.6	7.5	5.6	1.00	6.7

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
14.0830	3.6	7.2	5.6	0.90	6.7
14.0930	3.6	7.2	5.6	1.00	6.7
14.1030	3.5	7.2	5.6	1.00	6.7
14.1130	3.6	7.2	5.6	1.00	8.0
14.1230	3.5	7.2	5.6	1.00	7.2
14.1330	3.5	7.2	5.7	0.95	7.5
14.1430	3.5	7.1	5.7	0.95	7.7
14.1530	3.4	7.1	5.7	0.95	8.0
14.1630	3.5	7.2	6.0	0.95	8.0
14.1730	3.5	7.2	6.0	0.95	8.0
14.1830	3.5	7.2	5.7	1.00	8.0
14.1930	3.6	7.2	5.5	1.00	8.0
14.2030	3.5	7.1	5.6	1.00	8.0
14.2130	3.5	6.7	5.7	1.00	8.0
14.2230	3.5	6.5	5.7	1.00	8.0
14.2330	3.7	8.0	5.5	1.00	8.0
15.0030	3.9	7.6	5.6	1.00	7.5
15.0130	3.7	6.8	5.5	1.00	7.5
15.0230	3.7	7.8	5.0	1.00	7.5
15.0330	3.9	8.5	4.9	1.00	7.5
15.0430	4.0	8.6	4.7	1.00	7.5
15.0530	4.0	8.6	5.0	1.00	7.5
15.0630	3.9	7.8	5.0	1.00	7.5
15.0730	3.9	7.6	5.0	1.00	7.5
15.0830	3.9	7.8	4.7	1.00	7.5
15.0930	3.8	7.6	5.0	1.00	7.5
15.1030	3.9	7.6	5.1	1.00	7.5
15.1130	3.9	7.5	5.1	1.00	7.5
15.1230	3.9	7.6	5.1	1.00	7.5
15.1330	3.9	7.7	5.2	1.00	7.5
15.1430	3.7	7.5	5.4	1.00	7.5
15.1530	3.7	7.5	5.5	1.00	7.5
15.1630	3.6	7.5	5.5	1.00	7.7
15.1730	3.7	7.6	5.5	1.00	7.7
15.1830	3.7	7.5	5.5	1.00	7.7
15.1930	3.6	7.5	5.6	1.00	7.5
15.2030	3.7	7.5	5.6	1.00	7.5
15.2130	3.6	7.3	5.5	1.00	7.5
15.2230	3.6	7.4	5.7	1.00	7.5
15.2330	3.6	7.4	5.7	1.00	7.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
16.0030	3.6	7.3	5.6	1.00	7.7
16.0130	3.6	7.3	5.6	1.00	7.5
16.0230	3.7	7.5	5.5	1.00	7.5
16.0330	3.7	7.5	5.6	1.00	7.5
16.0430	3.6	7.2	5.6	1.00	7.5
16.0530	3.6	7.5	5.6	1.00	7.5
16.0630	3.6	7.3	5.6	1.00	7.5
16.0730	3.6	7.5	5.6	1.00	7.5
16.0830	3.6	7.5	5.5	1.00	7.5
16.0930	3.6	7.6	5.5	1.00	7.5
16.1030	3.6	7.6	5.5	1.00	7.5
16.1130	3.6	8.1	5.5	1.00	7.5
16.1230	3.6	8.2	5.5	1.00	7.5
16.1330	3.6	8.2	5.5	1.00	7.5
16.1430	3.6	8.0	5.2	1.00	7.5
16.1530	3.6	7.7	5.1	1.00	7.5
16.1630	3.5	7.8	4.9	1.00	7.5
16.1730	3.6	8.0	4.5	1.00	7.5
16.1830	3.6	8.0	4.5	1.00	7.5
16.1930	3.6	8.1	4.1	1.00	7.5
16.2030	3.6	8.1	4.1	1.00	7.5
16.2130	3.9	8.1	4.2	1.00	7.5
16.2230	3.7	8.1	4.5	1.00	7.5
16.2330	3.7	8.2	4.5	1.00	7.5
17.0030	3.8	8.3	4.2	1.00	7.5
17.0130	3.8	8.3	4.2	1.00	7.5
17.0230	3.8	8.3	4.0	1.00	7.5
17.0330	3.8	8.3	4.2	1.00	7.5
17.0430	3.8	8.5	4.2	1.00	7.5
17.0530	3.9	8.6	4.2	1.00	7.5
17.0630	3.7	8.6	4.2	1.00	7.5
17.0730	3.8	8.6	4.4	1.00	7.5
17.0830	3.9	8.8	4.2	1.00	7.5
17.0930	4.0	8.7	4.5	0.90	7.5
17.1030	3.9	8.7	4.2	0.90	7.5
17.1130	4.0	8.7	4.2	0.90	7.5

SHUT DOWN AT 17.1130 FOR 40 HOURS

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
19.0430	3.7	6.8	4.0	1.00	7.5
19.0530	3.7	7.5	4.0	1.00	7.5
19.0630	3.7	7.5	4.1	1.00	7.5
19.0730	3.5	7.5	4.1	1.00	7.5
19.0830	3.7	7.6	3.7	1.00	7.5
19.0930	3.7	7.7	3.7	1.00	7.5
19.1030	3.7	8.0	3.9	1.00	7.5
19.1130	3.7	8.0	3.9	1.00	7.5
19.1230	3.7	8.0	3.9	0.95	7.5
19.1330	3.6	8.0	4.0	0.90	7.5
19.1430	3.7	8.0	4.0	1.00	7.5
19.1530	3.7	8.0	4.0	1.10	7.5
19.1630	3.7	8.2	4.0	1.00	7.5
19.1730	3.7	8.2	4.0	1.00	7.5
19.1830	3.7	8.3	4.0	1.00	7.5
19.1930	3.7	8.5	4.1	0.95	7.5
19.2030	3.7	8.5	4.2	1.00	7.5
19.2130	3.7	8.5	4.4	1.00	7.5
19.2230	3.7	8.5	4.5	1.00	7.5
19.2330	3.7	8.5	4.4	1.00	7.5
20.0030	3.7	7.5	4.4	1.00	7.5
20.0130	3.7	7.5	4.4	1.00	7.5
20.0230	3.7	7.5	4.4	1.00	7.5
20.0330	3.7	7.5	4.4	1.00	7.5
20.0430	3.9	7.5	4.2	1.00	7.5
20.0530	3.7	7.5	4.2	1.00	7.5
20.0630	3.9	7.5	4.2	1.05	7.5
20.0730	3.9	7.5	4.2	1.00	7.5
20.0830	3.9	7.5	4.2	1.00	7.5
20.0930	3.9	7.5	4.2	1.00	7.5
20.1030	3.9	7.5	4.4	1.00	7.5
20.1130	3.9	7.5	4.4	1.00	7.5
20.1230	3.7	7.5	4.7	1.05	7.5
20.1330	3.6	7.5	4.7	1.00	7.5
20.1430	3.6	7.5	4.7	1.00	7.5
20.1530	3.6	7.5	4.7	1.00	7.5
20.1630	3.9	7.5	4.7	1.00	7.5
20.1730	3.9	7.5	4.7	1.00	7.5
20.1830	3.9	7.5	4.7	1.00	7.5
20.1930	3.9	7.5	4.7	1.00	7.5

APPENDIX C: TABLE III.
RUN 6: PRESSURES PAGE 11 OF 12

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
20.2030	3.9	7.5	5.5	1.00	7.5
20.2130	3.9	7.5	5.5	1.00	7.5
20.2230	4.1	7.5	4.2	1.00	7.5
20.2330	4.0	7.5	4.5	0.95	7.5
21.0030	4.0	7.5	4.5	0.95	7.5
21.0130	4.0	7.5	5.0	0.90	7.5
21.0230	4.0	7.5	5.0	0.95	7.5
21.0330	4.0	7.5	5.0	0.95	7.5
21.0430	4.1	7.5	4.7	0.95	7.5
		STONE CHANGE			
21.0530	4.1	7.5	5.0	0.95	7.5
21.0630	4.0	7.5	4.7	0.95	7.5
21.0730	4.0	7.5	4.7	1.00	7.5
21.0830	4.0	7.5	4.7	1.05	7.5
21.0930	4.0	7.5	5.0	1.00	7.5
21.1030	4.0	7.5	5.0	1.00	7.5
21.1130	3.9	7.5	5.0	1.00	7.5
21.1230	4.0	7.5	4.7	1.00	7.5
21.1330	4.0	7.5	4.7	1.00	7.5
21.1430	4.0	7.5	4.7	1.05	7.5
21.1530	4.0	7.5	5.0	1.05	7.5
21.1630	4.0	7.5	5.0	1.05	7.5
21.1730	4.0	7.5	5.0	1.05	7.5
21.1830	4.0	7.5	5.2	1.05	7.5
21.1930	4.0	7.5	5.2	1.05	7.5
21.2030	4.0	7.5	5.1	1.05	7.5
21.2130	4.0	7.5	5.1	1.05	7.5
21.2230	4.1	7.5	4.7	1.03	7.5
21.2330	4.0	7.5	4.9	1.05	7.5
22.0030	4.1	7.5	4.9	1.00	7.5
22.0130	4.0	7.5	4.7	1.00	7.5
22.0230	4.0	7.5	4.7	1.05	7.5
22.0330	4.0	7.5	4.7	1.00	7.5
22.0430	4.0	7.5	4.9	1.05	7.5
22.0530	4.0	7.5	4.7	1.10	7.5
22.0630	4.1	7.5	4.7	1.05	7.5
22.0730	4.0	7.5	4.9	1.10	7.5
22.0830	4.1	7.5	4.7	1.05	7.5
22.0930	4.0	7.5	4.7	1.10	7.5
22.1030	4.0	7.5	4.9	1.10	7.5

APPENDIX C: TABLE III.
RUN 6: PRESSURES

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
22.1130	4.0	7.5	4.7	1.10	7.5
22.1230	4.0	7.5	4.6	1.10	7.5
22.1330	4.1	7.5	4.6	1.10	7.5
22.1430	4.2	7.5	4.6	1.10	7.5
22.1530	4.2	7.5	4.7	1.10	7.5
22.1630	4.2	7.5	4.7	0.90	7.5
22.1730	4.4	7.5	4.1	0.90	7.5

APPENDIX C: TABLE IV.
RUN 6: DESULPHURISATION PERFORMANCE PAGE 1 OF 12

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
1.2230	92.9	1.44	62.	26.8	0.97	-	0.
1.2330	85.9	1.53	59.	22.7	0.44	-	0.
2.0030	84.1	1.47	59.	22.7	0.	49.1	47.0
2.0130	83.0	1.46	53.	22.8	0.	66.3	71.7
2.0230	81.0	1.48	54.	22.1	0.	79.7	75.3
2.0330	81.1	1.43	53.	21.5	0.55	83.0	98.6
2.0430	84.0	1.42	57.	20.7	1.15	84.1	97.9
2.0530	81.1	1.44	56.	21.0	0.96	81.9	89.0
2.0630	79.9	1.37	55.	20.6	0.92	96.0	94.7
2.0730	78.9	1.38	61.	20.7	0.95	91.2	88.2
2.0830	76.9	1.38	54.	21.2	1.02	89.4	86.9
2.0930	76.5	1.38	57.	20.6	0.72	88.3	75.0
2.1030	76.4	1.33	58.	20.9	0.65	83.8	85.3
2.1130	75.1	1.45	57.	21.1	0.71	66.3	67.1
2.1230	75.4	1.36	55.	20.8	0.82	75.3	67.0
2.1330	75.2	1.34	57.	20.0	1.09	85.4	68.0
2.1430	74.8	1.35	57.	19.5	0.78	87.4	60.0
2.1530	74.8	1.38	54.	20.0	0.78	86.6	69.7
2.1630	75.3	1.39	52.	20.5	1.03	76.7	67.4
2.1730	75.6	1.36	53.	20.5	1.13	80.0	67.9
2.1830	76.5	1.43	54.	21.6	1.22	73.2	77.2
2.1930	78.1	1.44	53.	21.9	1.34	70.1	69.6
2.2030	78.8	1.42	52.	21.5	1.06	67.7	75.4
2.2130	77.8	1.47	54.	22.2	1.14	68.8	76.0
2.2230	79.2	1.46	55.	21.9	1.25	56.2	56.2
2.2330	80.2	1.46	55.	22.0	1.13	71.7	79.0
3.0030	79.9	1.45	56.	22.2	1.25	64.8	73.4
3.0130	80.3	1.46	57.	22.1	1.19	62.7	73.7
3.0230	80.7	1.45	57.	22.3	1.16	69.7	74.4
3.0330	80.7	1.44	58.	22.3	1.20	59.6	66.4
3.0430	81.0	1.44	59.	22.1	1.31	61.5	73.6
3.0530	80.6	1.56	60.	23.7	1.25	61.2	76.7
3.0630	81.3	1.51	57.	23.1	1.10	61.9	82.6
3.0730	81.8	1.50	59.	23.0	0.79	60.2	71.1
3.0830	81.1	1.51	62.	22.9	1.16	59.9	73.5
3.0930	80.5	1.46	62.	22.7	0.94	62.4	76.1
3.1030	80.4	1.46	60.	22.8	1.19	62.0	73.0
3.1130	79.9	1.46	62.	22.5	1.18	64.3	75.8
3.1230	78.8	1.49	63.	22.7	0.83	68.4	79.2
3.1330	78.8	1.45	62.	22.6	0.98	61.6	71.3

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
3.1430	80.1	1.44	63.	22.2	0.95	73.9	77.9
3.1530	81.6	1.42	64.	22.6	0.89	67.4	77.9
3.1630	81.7	1.45	64.	22.7	0.98	73.3	84.9
3.1730	82.6	1.44	63.	22.6	1.32	74.9	83.4
3.1830	81.7	1.47	63.	23.1	1.69	77.8	87.4
3.1930	82.6	1.47	63.	22.9	1.52	67.0	76.2
3.2030	83.7	1.47	63.	23.0	1.44	76.4	86.5
3.2130	83.8	1.44	64.	23.2	1.33	80.1	88.5
3.2230	84.1	1.44	66.	22.8	1.43	85.0	88.6
3.2330	84.3	1.46	64.	23.1	1.20	69.7	83.6
4.0030	85.2	1.46	64.	23.0	1.50	75.0	85.8
4.0130	85.0	1.47	66.	23.1	1.32	101.3	82.3
4.0230	83.3	1.63	65.	22.6	0.46	74.8	78.9
4.0330	78.0	1.22	66.	22.6	0.	61.6	74.0
4.0430	75.0	1.54	60.	22.9	0.	71.1	79.5
4.0530	76.6	1.51	60.	23.0	0.	78.7	88.4
4.0630	78.2	1.49	64.	23.8	0.	60.3	70.2
4.0730	78.0	1.44	63.	22.2	0.	69.9	78.1
4.0830	75.2	1.44	63.	22.0	0.	74.6	83.0
4.0930	73.7	1.47	63.	22.2	0.	72.9	80.5
4.1030	71.5	1.48	63.	21.8	0.	65.4	71.2
4.1130	68.0	1.46	60.	21.4	0.	69.8	76.4
4.1230	68.6	1.49	57.	19.7	0.	60.5	65.2
4.1330	66.6	1.45	57.	19.6	0.	63.7	65.8
4.1430	68.8	1.45	55.	20.1	0.	69.2	73.2
4.1530	68.9	1.47	57.	22.1	0.80	66.4	75.5
4.1630	72.6	1.44	60.	21.9	0.99	61.7	69.7
4.1730	75.7	1.43	61.	21.9	0.95	68.0	73.7
4.1830	75.5	1.47	60.	22.5	0.95	69.7	74.9
4.1930	79.4	1.41	61.	21.4	0.76	71.9	78.7
4.2030	82.4	1.44	60.	22.2	0.92	67.3	73.3
4.2130	84.3	1.48	61.	22.7	1.20	62.9	73.9
4.2230	87.2	1.46	60.	23.1	0.95	60.9	70.0
4.2330	88.7	1.45	60.	22.5	1.07	56.4	63.3
5.0030	88.5	1.49	67.	22.5	1.21	77.4	72.6
5.0130	87.7	1.43	62.	22.2	1.05	66.8	70.2
5.0230	85.9	1.43	62.	22.3	0.77	51.5	47.9
5.0330	85.6	1.44	62.	22.2	1.02	70.3	73.5
5.0430	83.7	1.52	59.	22.6	0.98	70.7	81.1
5.0530	80.4	1.53	60.	22.3	0.96	58.2	61.9

APPENDIX C: TABLE IV.

RUN 6: DESULPHURISATION PERFORMANCE PAGE 3 OF 12

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
5.0630	86.5	1.45	61.	20.4	0.93	58.7	58.6
5.0730	86.4	1.43	64.	20.9	0.96	65.2	65.1
5.0830	86.4	1.52	61.	22.6	1.27	75.7	71.3
5.0930	86.1	1.48	61.	23.4	1.00	73.4	68.2
5.1030	87.4	1.47	61.	21.8	0.96	83.7	73.5
5.1130	87.2	1.51	60.	22.9	1.10	80.8	70.6
5.1230	86.8	1.50	66.	23.2	1.23	55.6	53.7
5.1330	88.5	1.54	63.	23.2	1.25	66.6	75.7
5.1430	88.4	1.51	60.	23.2	1.31	79.7	84.2
5.1530	88.2	1.53	58.	23.1	1.17	67.7	75.0
5.1630	87.8	1.46	60.	22.7	1.27	66.4	72.4
5.1730	88.1	1.49	55.	22.5	1.11	66.6	72.2
5.1830	87.3	1.52	58.	23.1	0.87	69.0	79.4
5.1930	-	1.49	62.	22.7	0.87	67.4	70.3
5.2030	72.9	1.49	58.	22.7	1.12	66.1	69.9
5.2130	76.4	1.46	56.	22.3	1.03	82.5	89.9
5.2230	74.6	1.47	58.	22.5	0.91	69.1	73.3
5.2330	75.7	1.48	58.	22.1	0.90	68.1	72.6
6.0030	78.3	1.48	58.	22.3	1.03	73.4	78.0
6.0130	78.1	1.48	58.	22.3	0.81	70.0	74.3
6.0230	73.7	1.44	58.	22.0	0.97	66.8	72.1
6.0330	74.6	1.45	58.	22.2	1.07	69.6	74.8
6.0430	73.8	1.45	60.	21.8	1.73	68.8	69.7
6.0530	77.3	1.44	60.	22.0	1.28	68.3	72.5
6.0630	76.9	1.50	63.	23.2	1.41	68.6	73.3
6.0730	75.8	1.49	63.	22.9	1.62	67.3	73.0
6.0830	77.0	1.48	60.	22.8	1.31	69.1	72.7
6.0930	77.9	1.48	62.	22.7	1.84	69.6	72.5
6.1030	78.3	1.53	58.	22.9	0.86	70.1	73.9
6.1130	78.2	1.47	60.	23.5	1.57	72.4	74.7
6.1230	79.7	1.44	55.	22.7	1.47	67.9	70.4
6.1330	78.8	1.44	58.	22.7	1.12	75.7	69.6
6.1430	78.7	1.44	55.	23.0	1.53	62.1	64.2
6.1530	79.9	1.44	55.	22.6	1.83	66.5	65.1
6.1630	79.0	1.43	51.	22.1	1.31	72.7	54.2
6.1730	78.5	1.44	51.	22.5	1.83	73.2	68.9
6.1830	78.8	1.44	58.	22.4	0.99	70.4	67.3
6.1930	78.1	1.44	50.	22.7	1.65	72.0	67.2
6.2030			MISSED DATA READING				
6.2130	78.7	1.42	50.	23.0	1.36	68.8	65.0

APPENDIX C: TABLE IV.

RUN 6: DESULPHURISATION PERFORMANCE PAGE 4 OF 12

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
6.2230	78.8	1.33	48.	21.4	1.35	66.2	63.7
6.2330	78.2	1.35	50.	20.8	0.86	54.0	44.8
7.0030	79.8	1.43	48.	21.9	1.67	70.5	54.5
7.0130	80.4	1.38	50.	22.3	1.77	72.2	57.4
7.0230	81.6	1.42	53.	22.1	1.23	88.9	71.5
7.0330	79.4	1.43	55.	22.0	1.56	82.1	76.3
7.0430	78.0	1.43	55.	22.8	1.29	84.6	75.3
7.0530	78.8	1.43	55.	22.9	1.69	67.9	65.8
7.0630	80.7	1.41	53.	22.0	1.23	75.0	69.0
7.0730	81.2	1.40	53.	22.6	1.55	81.1	72.7
7.0830	82.2	1.45	53.	22.9	1.42	77.3	72.1
7.0930	82.5	1.41	53.	22.9	1.08	86.2	78.0
7.1030	83.1	1.40	53.	22.8	1.01	88.6	78.6
7.1130	83.7	1.39	53.	22.3	0.77	84.3	74.9
7.1230	81.7	1.40	54.	22.3	0.40	85.6	81.3
7.1330	80.7	1.43	63.	23.0	0.46	85.0	73.0
7.1430	79.4	1.42	63.	22.7	0.49	86.2	85.0
7.1530	80.4	1.40	63.	22.3	0.68	84.0	77.7
7.1630	79.2	1.40	57.	22.3	0.74	83.7	78.2
7.1730	78.4	1.40	58.	22.4	0.74	82.8	75.5
7.1830	78.5	1.40	58.	22.0	0.40	81.7	72.2
7.1930	81.0	1.41	64.	22.8	0.44	79.5	77.2
7.2030	79.4	1.41	58.	22.4	0.37	78.0	73.5
7.2130	77.3	1.39	58.	22.0	0.43	77.5	71.7
7.2230	75.3	1.41	58.	22.2	0.37	77.1	73.6
7.2330	74.2	1.34	58.	19.8	0.44	89.5	72.0
8.0030	74.6	1.36	55.	20.2	0.35	92.2	73.3
8.0130	73.4	1.39	56.	20.0	0.44	86.8	72.9
8.0230	75.8	1.38	53.	20.6	0.49	91.8	72.1
8.0330	76.5	1.36	53.	20.3	0.46	82.0	70.5
8.0430	74.2	1.39	53.	20.8	0.43	87.7	76.6
8.0530	74.6	1.37	53.	20.7	0.37	89.8	77.2
8.0630	72.8	1.40	53.	20.7	0.40	87.5	74.9
8.0730	71.8	1.40	53.	20.7	0.49	86.2	75.2
8.0830	71.6	1.40	53.	20.7	0.52	86.8	75.8
8.0930	69.9	1.39	53.	20.6	0.52	89.5	76.0
8.1030	70.7	1.38	53.	20.5	0.40	90.7	79.0
8.1130	70.6	1.37	50.	20.3	0.40	85.8	80.4
8.1230	71.4	1.38	50.	20.3	0.40	87.7	80.6
8.1330	71.4	1.38	53.	20.4	0.40	90.9	81.0

APPENDIX C: TABLE IV.

RUN 6: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
8.1430	71.3	1.36	53.	20.1	0.31	83.9	77.1
8.1530	71.4	1.33	53.	19.9	0.52	86.6	75.7
8.1630	69.9	1.37	53.	20.5	0.52	86.0	74.9
8.1730	68.4	1.32	53.	19.6	0.59	79.5	70.1
8.1830	71.2	1.13	53.	19.2	0.72	83.5	74.0
8.1930	73.4	1.14	53.	19.9	0.80	87.3	78.5
8.2030	73.4	1.16	53.	20.1	0.42	87.8	79.4
8.2130	73.1	1.25	53.	22.0	0.39	90.9	81.6
8.2230	72.8	1.24	55.	21.5	0.48	83.1	74.6
8.2330	74.0	1.26	54.	21.7	0.73	81.8	79.9
9.0030	73.2	1.29	53.	21.7	0.52	73.3	76.3
9.0130	70.9	1.27	53.	20.9	0.59	70.9	74.4
9.0230	71.1	1.29	54.	20.8	0.42	78.2	76.8
9.0330	71.7	1.26	54.	20.7	0.42	77.7	74.1
9.0430	70.8	1.27	55.	20.8	0.35	78.2	76.6
9.0530	65.9	1.38	53.	21.4	0.45	68.6	74.3
9.0630	66.1	1.27	50.	20.7	0.48	67.7	72.2
9.0730	65.9	1.26	48.	20.7	0.55	64.1	69.4
9.0830	65.7	1.26	50.	20.7	0.59	63.7	74.1
9.0930	70.7	1.28	52.	20.4	0.59	64.9	58.5
9.1030	71.2	1.24	53.	21.3	0.58	68.8	71.4
9.1130	71.3	1.29	53.	21.0	0.72	65.2	68.2
9.1230	72.5	1.31	50.	21.8	0.72	74.1	74.6
9.1330	74.8	1.27	54.	21.4	0.82	88.0	78.6
9.1430	76.4	1.27	53.	21.0	0.73	82.1	75.8
9.1530	76.9	1.27	48.	21.1	0.88	86.1	77.2
9.1630	72.3	1.28	50.	21.3	0.71	84.8	83.7
9.1730	72.7	1.31	53.	21.5	0.90	86.2	76.0
9.1830	75.9	1.32	53.	21.3	0.98	79.8	64.1
9.1930	77.6	1.34	55.	24.4	1.21	74.3	71.3
9.2030	79.1	1.31	50.	22.2	0.90	74.5	75.7
9.2130	69.2	1.30	45.	21.7	0.71	68.2	70.3
9.2230	71.0	1.26	49.	21.7	1.07	60.3	62.6
SHUT DOWN AT 9.2230 FOR 23 HOURS							
10.2130	71.6	0.94	50.	21.9	1.76	42.2	43.3
10.2230	74.7	1.39	50.	21.9	1.93	62.5	67.6
10.2330	76.3	1.40	49.	21.2	1.52	71.6	77.9

APPENDIX C: TABLE IV.
RUN 6: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
11.0030	76.4	1.39	52.	21.7	1.29	69.0	75.2
11.0130	77.3	1.40	53.	22.1	1.71	55.8	60.9
11.0230	79.1	1.39	57.	22.4	2.05	66.2	75.6
11.0330	81.1	1.35	56.	22.5	2.03	65.2	77.2
11.0430	81.0	1.35	58.	21.7	1.52	69.9	70.8
11.0530	81.7	1.37	56.	22.2	0.94	68.2	74.9
11.0630	83.1	1.35	58.	22.5	2.27	65.8	74.4
11.0730	83.7	1.38	58.	23.4	2.26	62.6	73.5
11.0830	85.1	1.29	59.	22.2	2.04	61.3	70.0
11.0930	85.4	1.30	58.	22.0	2.34	60.8	67.7
11.1030	84.5	1.29	58.	21.7	2.18	54.3	56.9
11.1130	82.7	1.27	58.	22.0	2.18	60.9	67.0
11.1230	82.8	1.27	57.	21.8	2.17	63.7	68.5
11.1330	83.9	1.16	59.	20.8	2.15	56.6	55.1
11.1430	84.7	1.42	55.	23.4	2.36	55.6	49.7
11.1530	86.0	1.40	59.	23.5	2.22	51.3	46.2
11.1630	85.9	1.42	58.	23.5	2.26	67.3	56.5
11.1730	85.9	1.45	58.	23.8	2.39	62.4	60.7
11.1830	84.7	1.46	55.	24.2	2.08	63.8	57.8
11.1930	82.7	1.45	53.	23.4	1.60	71.6	75.6
11.2030	82.7	1.44	55.	23.2	0.	72.8	79.8
11.2130	80.9	1.43	55.	23.3	1.21	61.6	75.1
11.2230	81.4	1.43	58.	22.9	1.59	65.3	63.3
11.2330	83.5	1.42	58.	22.7	2.29	49.5	47.5
12.0030	84.4	1.41	58.	22.4	1.67	49.4	47.5
12.0130	85.4	1.44	60.	23.3	1.94	54.4	52.5
12.0230	85.1	1.43	55.	22.7	0.94	58.7	59.3
12.0330	82.0	1.46	58.	23.7	1.78	58.0	60.5
12.0430	80.6	1.47	57.	23.1	2.28	52.2	55.1
12.0530	81.8	1.46	51.	23.3	2.21	43.7	45.1
12.0630	84.0	1.46	58.	23.7	1.85	45.5	47.0
12.0730	86.3	1.41	51.	22.6	1.87	51.4	53.8
12.0830	86.2	1.42	51.	22.8	2.16	52.4	52.6
12.0930	86.5	1.41	60.	22.9	2.29	39.3	38.7
12.1030	85.2	1.43	60.	23.8	1.53	61.6	66.4
12.1130	84.6	1.47	58.	23.3	1.38	55.8	59.7
12.1230	84.3	1.44	58.	23.1	2.27	51.2	54.0
12.1330	83.8	1.44	55.	23.4	2.15	53.3	58.3
12.1430	85.2	1.42	55.	22.7	2.17	51.8	55.8
12.1530	85.6	1.42	55.	22.7	1.86	49.7	55.2

APPENDIX C: TABLE IV.
RUN 6: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
12.1630	85.3	1.42	53.	22.0	1.80	42.0	43.2
12.1730	84.1	1.39	53.	20.4	1.38	47.9	53.1
12.1830	82.5	1.37	55.	21.0	1.70	53.9	56.1
12.1930	81.6	1.39	55.	21.3	1.35	49.4	52.6
12.2030	81.6	1.32	55.	21.0	1.32	53.3	56.5
12.2130	82.2	1.40	55.	21.5	1.55	52.8	51.9
12.2230	83.4	1.39	50.	21.3	1.26	55.1	61.9
12.2330	83.7	1.44	48.	21.9	1.50	57.4	62.5
13.0030	83.5	1.40	48.	21.9	1.65	60.2	65.8
13.0130	80.7	1.39	47.	21.5	1.25	41.8	45.4
13.0230	80.7	1.43	48.	21.2	1.87	61.2	66.2
13.0330	81.9	1.43	53.	21.7	1.42	58.7	58.2
13.0430	82.8	1.46	47.	22.5	1.67	54.2	62.6
13.0530	82.4	1.47	48.	22.1	1.53	53.6	59.5
13.0630	82.1	1.44	47.	21.5	1.56	53.9	57.8
13.0730	83.3	1.43	48.	22.5	2.39	54.0	57.2
13.0830	84.6	1.43	48.	21.4	1.64	51.9	53.8
13.0930	85.0	1.42	53.	21.7	1.98	38.7	39.3
13.1030	82.5	1.42	50.	21.7	1.46	47.0	48.6
13.1130	82.8	1.42	50.	21.6	1.73	51.0	52.2
13.1230	84.6	1.42	48.	21.5	1.76	53.6	54.5
13.1330	83.8	1.43	50.	22.5	1.73	57.2	61.2
13.1430	83.2	1.46	48.	22.4	1.25	73.2	82.9
13.1530	80.9	1.47	48.	22.4	0.86	67.7	70.6
13.1630	78.5	1.51	46.	22.8	1.28	62.6	72.3
13.1730	77.6	1.47	43.	22.5	1.21	60.8	72.2
13.1830	77.4	1.49	45.	23.0	1.46	57.0	66.1
13.1930	75.6	1.48	47.	22.1	1.46	53.1	61.8
13.2030	69.2	1.46	50.	22.4	1.39	58.5	66.2
13.2130	51.0	1.51	53.	23.1	1.30	42.4	46.5
13.2230	54.3	1.45	49.	21.1	1.55	35.9	37.6
13.2330	58.8	1.47	53.	21.6	1.42	43.5	47.4
14.0030	63.7	1.44	55.	20.5	1.19	44.2	48.8
14.0130	62.8	1.46	53.	20.6	1.25	50.8	55.6
14.0230	61.8	1.51	50.	21.7	1.15	47.0	48.3
14.0330	61.2	1.46	50.	20.6	0.78	42.9	45.5
14.0430	62.6	1.37	50.	18.7	1.15	49.7	51.4
14.0530	64.1	1.46	50.	20.7	1.34	46.8	46.2
14.0630	59.0	1.46	55.	21.3	1.24	47.6	49.9
14.0730	58.8	1.37	57.	21.3	1.50	47.4	44.9

APPENDIX C: TABLE IV.

RUN 6: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
14.0830	61.5	1.37	63.	21.3	1.59	46.9	44.6
14.0930	63.1	1.38	57.	22.1	1.02	48.5	46.0
14.1030	65.7	1.45	57.	21.0	1.14	39.5	40.0
14.1130	66.9	1.43	57.	20.7	1.10	50.3	55.1
14.1230	67.1	1.43	57.	20.6	0.97	50.0	56.2
14.1330	66.9	1.41	61.	20.1	1.07	50.7	56.9
14.1430	68.2	1.42	61.	20.1	1.04	47.2	51.9
14.1530	66.4	1.40	61.	20.5	0.93	52.6	57.2
14.1630	66.0	1.45	64.	20.4	0.87	47.6	51.3
14.1730	66.3	1.45	64.	20.5	1.23	51.2	56.9
14.1830	65.2	1.44	58.	20.5	1.53	48.6	52.5
14.1930	66.3	1.43	55.	20.5	1.20	48.5	53.4
14.2030	65.8	1.41	57.	20.5	1.10	48.0	53.3
14.2130	66.8	1.38	58.	20.0	1.23	34.8	39.0
14.2230	67.5	1.36	58.	19.4	1.11	50.6	61.3
14.2330	68.2	1.44	55.	21.0	0.83	50.0	60.6
15.0030	63.4	1.45	57.	20.6	1.15	41.5	47.3
15.0130	61.0	1.41	55.	20.5	1.39	45.6	54.9
15.0230	63.4	1.35	50.	18.2	1.21	47.4	52.3
15.0330	65.4	1.49	49.	21.2	1.22	45.0	49.5
15.0430	72.1	1.46	48.	20.7	1.13	59.4	63.8
15.0530	74.4	1.40	50.	19.6	0.93	59.2	66.5
15.0630	76.5	1.45	50.	20.0	0.91	62.2	70.0
15.0730	80.0	1.44	50.	20.7	1.16	59.2	67.3
15.0830	79.9	1.48	48.	20.5	1.12	61.2	70.5
15.0930	77.5	1.49	50.	20.7	1.00	55.3	58.5
15.1030	76.9	1.47	52.	21.6	1.17	59.2	66.4
15.1130	80.0	1.44	52.	21.2	1.23	58.2	66.2
15.1230	81.3	1.43	52.	21.3	1.03	58.4	66.7
15.1330	81.9	1.42	53.	20.8	1.17	58.2	63.3
15.1430	81.1	1.42	54.	21.5	0.99	60.7	68.2
15.1530	81.7	1.42	55.	21.2	1.01	54.3	59.5
15.1630	81.7	1.42	55.	21.2	0.91	57.0	63.3
15.1730	81.8	1.41	55.	21.2	1.62	60.5	65.3
15.1830	82.3	1.38	55.	21.2	1.45	59.4	62.3
15.1930	83.6	1.40	57.	22.0	1.30	60.2	61.7
15.2030	81.0	1.41	57.	21.2	1.38	60.2	66.9
15.2130	80.0	1.40	55.	21.5	1.26	57.2	63.8
15.2230	80.5	1.37	58.	21.1	1.40	54.0	60.2
15.2330	80.5	1.35	58.?	20.7	1.25	57.0	62.8

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APPENDIX C: TABLE IV.

RUN 6: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
16.0030	80.1	1.35	57.	20.8	1.32	57.2	63.0
16.0130	79.7	1.39	57.	21.4	1.22	59.7	66.6
16.0230	80.1	1.38	55.	21.4	1.34	57.1	62.1
16.0330	78.7	1.38	57.	22.2	1.23	55.6	61.1
16.0430	79.0	1.37	57.	22.0	0.85	55.9	62.9
16.0530	77.9	1.39	57.	21.0	0.93	57.0	63.7
16.0630	77.9	1.37	57.	21.4	1.46	57.2	64.2
16.0730	77.7	1.38	57.	21.4	1.08	56.4	64.5
16.0830	77.8	1.38	55.	21.3	0.87	56.9	64.7
16.0930	76.8	1.38	55.	21.5	0.84	57.7	65.0
16.1030	75.8	1.43	55.	22.0	1.08	53.7	60.7
16.1130	73.8	1.45	55.	22.6	1.17	54.4	61.3
16.1230	73.5	1.45	55.	22.7	1.35	51.4	59.0
16.1330	73.4	1.45	55.	22.8	1.75	49.3	56.9
16.1430	72.3	1.44	53.	22.7	1.50	55.5	63.5
16.1530	74.0	1.42	52.	22.1	1.50	48.1	55.7
16.1630	75.0	1.44	49.	22.1	1.89	52.7	59.4
16.1730	73.5	1.47	45.	22.1	0.86	50.1	60.2
16.1830	73.6	1.45	45.	21.8	1.21	57.0	68.7
16.1930	71.6	1.41	41.	21.6	1.82	65.4	84.6
16.2030	71.6	1.44	41.	21.6	1.71	46.4	58.6
16.2130	72.2	1.36	43.	20.8	2.04	49.8	62.8
16.2230	73.7	1.40	45.	21.3	1.61	55.9	71.4
16.2330	73.5	1.40	45.	21.4	1.36	50.1	64.2
17.0030	71.9	1.41	43.	21.3	1.36	46.4	59.7
17.0130	70.5	1.06	43.	20.4	1.20	48.9	61.8
17.0230	69.5	1.03	40.	20.4	1.02	42.1	55.2
17.0330	70.6	0.97	43.	20.1	1.89	44.7	57.9
17.0430	71.8	0.94	43.	19.6	1.75	51.7	66.2
17.0530	72.6	0.94	43.	20.0	1.61	45.7	58.8
17.0630	72.7	0.91	43.	19.8	1.44	47.1	59.2
17.0730	76.6	0.86	44.	18.7	1.15	50.0	63.4
17.0830	77.2	1.00	43.	21.9	1.90	51.3	65.0
17.0930	77.1	0.94	50.	21.1	1.46	53.9	69.2
17.1030	77.2	1.45	47.	22.2	1.25	53.7	67.4
17.1130	76.5	1.42	47.	22.1	1.43	51.3	64.8

SHUT DOWN AT 17.1130 FOR 40 HOURS

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
19.0430	87.7	1.35	40.	20.6	1.73	49.7	58.6
19.0530	84.1	1.39	40.	20.6	1.02	58.5	70.0
19.0630	80.6	1.39	41.	21.3	1.24	58.6	68.1
19.0730	78.4	1.41	41.	20.8	1.24	56.0	67.8
19.0830	75.7	1.35	38.	20.6	1.09	56.0	66.0
19.0930	73.7	1.30	38.	19.2	1.09	54.7	63.1
19.1030	73.3	1.41	39.	21.0	1.26	56.3	65.9
19.1130	74.1	1.39	39.	20.7	1.21	61.5	67.9
19.1230	75.2	1.37	41.	20.7	1.21	61.3	67.2
19.1330	75.3	1.35	45.	20.4	1.30	66.7	69.9
19.1430	75.2	1.38	40.	21.0	1.29	70.5	75.8
19.1530	77.5	1.37	36.	21.0	1.11	68.6	80.7
19.1630	65.2	1.36	40.	21.6	1.21	57.9	69.2
19.1730	79.8	1.41	40.	21.0	1.24	56.1	62.8
19.1830	82.0	1.35	40.	20.1	1.05	67.1	74.1
19.1930	80.9	1.36	44.	21.3	0.98	65.4	66.4
19.2030	78.1	1.39	43.	21.1	1.40	38.9	19.9
19.2130	81.3	1.35	44.	20.6	1.32	79.0	73.9
19.2230	78.5	1.38	45.	20.6	1.07	69.2	66.6
19.2330	75.0	1.35	44.	20.6	1.26	71.0	72.0
20.0030	78.0	1.36	44.	20.7	1.02	65.1	72.4
20.0130	78.0	1.36	44.	21.1	1.00	71.1	84.1
20.0230	76.7	1.36	44.	20.4	0.86	73.3	79.5
20.0330	78.7	1.34	44.	19.7	0.88	60.0	72.6
20.0430	77.7	1.34	43.	19.5	1.02	71.2	80.7
20.0530	77.4	1.36	43.	20.0	1.16	70.2	83.9
20.0630	74.0	1.34	41.	20.5	0.88	63.6	79.8
20.0730	74.6	1.36	43.	20.9	0.84	63.4	78.0
20.0830	76.4	1.33	43.	20.9	1.66	67.1	74.7
20.0930	77.0	1.31	43.	20.2	1.12	53.3	63.6
20.1030	76.4	1.29	44.	19.7	0.84	60.6	72.4
20.1130	75.8	1.32	44.	20.5	0.84	71.8	81.5
20.1230	67.8	1.26	45.	20.1	1.19	69.9	79.6
20.1330	-	1.26	48.	19.9	1.12	78.1	79.0
20.1430	-	1.27	48.	19.9	0.67	67.7	75.6
20.1530	-	1.24	48.	19.3	0.63	60.5	68.8
20.1630	76.8	1.31	48.	20.4	1.26	55.4	63.6
20.1730	78.7	1.34	48.	20.4	1.57	63.3	89.1
20.1830	80.3	1.38	48.	22.2	1.61	57.4	82.7
20.1930	80.5	1.36	48.	21.6	1.81	62.1	77.1

APPENDIX C: TABLE IV.

RUN 6: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
20.2030	80.3	1.37	55.	21.7	1.27	70.2	63.4
20.2130	80.7	1.39	55.	21.4	1.07	62.2	70.9
20.2230	80.7	1.43	43.	23.2	1.46	68.8	64.1
20.2330	79.8	1.40	48.	22.3	1.43	61.8	63.1
21.0030	79.7	1.42	48.	22.1	1.00	71.5	80.9
21.0130	79.6	1.41	56.	22.2	1.55	63.7	66.8
21.0230	79.1	1.39	53.	21.3	1.69	58.9	74.3
21.0330	78.7	1.42	53.	21.4	1.03	52.8	58.2
21.0430	79.3	1.44	50.	22.3	1.43	68.5	86.1
			STONE CHANGE				
21.0530	79.2	1.45	53.	21.5	1.00	72.5	87.9
21.0630	76.3	1.46	50.	21.5	0.	62.7	78.3
21.0730	73.9	1.44	48.	21.7	0.	56.2	66.0
21.0830	72.1	1.37	45.	19.6	0.	57.8	57.7
21.0930	71.7	1.40	50.	20.9	0.97	58.2	60.3
21.1030	73.6	1.38	50.	21.4	1.07	56.9	66.2
21.1130	74.4	1.36	50.	21.1	1.06	57.9	60.2
21.1230	74.9	1.38	48.	21.5	0.89	55.9	58.6
21.1330	75.8	1.38	48.	21.8	1.37	49.1	49.2
21.1430	75.1	1.38	45.	21.6	1.29	54.5	62.7
21.1530	73.7	1.37	48.	20.7	0.56	51.8	59.1
21.1630	73.5	1.36	48.	20.2	1.10	52.4	59.5
21.1730	74.0	1.35	48.	20.0	1.20	52.8	60.1
21.1830	72.7	1.36	50.	20.3	1.34	51.9	63.1
21.1930	72.1	1.34	50.	20.2	1.28	42.8	50.8
21.2030	72.9	1.37	49.	20.6	1.78	58.3	61.4
21.2130	75.6	1.36	49.	20.5	1.10	57.0	55.4
21.2230	74.2	1.40	46.	21.2	1.30	53.4	54.0
21.2330	73.4	1.40	47.	21.2	1.20	57.5	62.5
22.0030	72.4	1.45	49.	21.3	1.17	51.2	52.6
22.0130	68.5	1.38	48.	20.4	1.14	54.4	51.3
22.0230	69.1	1.41	45.	21.4	1.31	55.2	59.9
22.0330	70.6	1.41	48.	21.2	1.27	53.6	52.5
22.0430	71.9	1.38	47.	20.6	1.22	51.4	53.6
22.0530	76.2	1.39	43.	20.2	1.06	45.4	51.0
22.0630	72.6	1.39	45.	20.4	1.01	47.0	52.8
22.0730	76.2	1.39	45.	20.5	1.11	40.4	44.5
22.0830	72.0	1.40	45.	20.9	1.14	50.1	56.5
22.0930	70.9	1.34	43.	19.1	0.97	52.7	60.1
22.1030	71.4	1.33	45.	19.2	0.90	39.7	43.4

APPENDIX C: TABLE IV.
RUN 6: DESULPHURISATION PERFORMANCE PAGE 12 OF 12

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
22.1130	72.0	1.36	43.	20.0	0.33	54.0	51.5
22.1230	71.2	1.37	42.	20.1	0.	52.1	44.3
22.1330	67.3	1.30	42.	18.9	0.	52.4	52.6
22.1430	66.5	1.32	42.	18.9	0.	54.1	51.2
22.1530	64.8	1.36	43.	19.0	0.73	46.5	51.7
22.1630	64.2	1.42	53.	21.2	1.01	49.2	49.3
22.1730	64.6	1.43	46.	21.4	1.19	36.5	39.7

APPENDIX C: TABLE V.

RUN 6: GAS COMPOSITIONS

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
1.2230	4.5	13.0	12.4	91.	12.00	0.4	0.	17.0	17.5	2.76	2.79
1.2330	3.5	13.5	13.1	192.	3.00	1.7	0.	15.5	16.9	3.27	3.20
2.0030	3.0	13.5	13.5	223.	0.60	2.2	3.9	15.5	16.7	2.28	3.20
2.0130	3.0	13.5	13.5	238.	1.80	0.6	7.0	16.0	16.7	3.27	3.20
2.0230	2.2	14.4	14.1	278.	0.90	3.3	7.4	16.5	16.6	3.54	3.37
2.0330	2.2	14.4	14.1	278.	0.30	2.2	9.4	16.5	16.5	3.45	3.47
2.0430	2.1	14.4	14.2	238.	0.40	2.3	9.4	16.5	16.2	3.45	3.62
2.0530	3.0	13.5	13.5	268.	0.40	3.3	8.6	17.0	16.5	3.10	3.41
2.0630	2.5	14.1	13.9	293.	0.50	4.7	9.0	16.8	16.7	3.10	3.24
2.0730	2.9	14.1	13.6	301.	0.50	4.4	8.6	16.8	16.8	3.10	3.24
2.0830	2.9	14.1	13.6	332.	0.50	4.6	8.6	16.5	17.0	3.82	3.09
2.0930	2.0	14.1	14.3	354.	0.50	5.9	7.4	17.6	16.8	3.10	3.09
2.1030	2.0	14.4	14.3	354.	0.50	3.8	9.0	16.8	17.4	3.01	2.70
2.1130	2.5	13.5	13.9	364.	0.50	3.0	7.4	16.8	16.8	2.93	3.09
2.1230	2.0	14.4	14.3	369.	0.50	5.2	7.2	16.8	16.9	2.93	3.09
2.1330	2.0	14.1	14.3	374.	0.20	6.8	7.4	16.6	16.8	3.01	3.08
2.1430	1.8	14.1	14.4	384.	0.20	8.8	6.2	16.6	16.5	2.93	3.30
2.1530	1.8	14.4	14.4	384.	0.20	7.5	7.0	16.6	16.6	2.93	3.31
2.1630	1.6	14.4	14.5	379.	0.20	6.4	6.8	16.6	16.7	2.93	3.15
2.1730	1.7	14.4	14.5	374.	0.20	7.2	6.6	16.6	16.8	2.93	3.09
2.1830	2.0	14.4	14.3	354.	0.	4.5	7.6	17.0	17.1	2.68	2.98
2.1930	2.1	14.4	14.2	329.	0.	5.0	7.0	17.2	17.4	2.52	2.77
2.2030	2.0	14.4	14.3	319.	0.	3.1	7.6	17.2	17.3	2.60	2.82
2.2130	2.0	14.4	14.3	334.	0.	3.5	7.6	17.2	17.1	2.44	2.93
2.2230	2.0	14.4	14.3	314.	0.	3.9	5.6	17.2	17.1	2.44	2.93
2.2330	2.3	14.1	14.0	293.	0.	3.3	8.0	17.2	17.2	2.44	2.87
3.0030	2.9	13.5	13.6	288.	0.20	2.9	7.4	17.5	17.2	2.36	2.82
3.0130	2.5	14.1	13.9	288.	0.20	2.6	7.4	17.3	17.0	2.44	3.04

3.0230	2.5	13.8	13.9	283.	0.50	4.1	7.0	17.4	17.0	2.44	2.98
3.0330	2.5	14.1	13.9	283.	0.70	3.1	6.4	17.4	17.0	2.44	3.04
3.0430	2.5	13.8	13.9	278.	0.20	2.4	7.2	17.5	17.0	2.36	2.98
3.0530	2.5	13.8	13.9	283.	0.10	1.7	7.4	18.0	17.3	2.20	2.79
3.0630	2.5	14.1	13.9	273.	0.20	1.6	7.8	17.5	17.1	2.28	2.94
3.0730	2.3	14.4	14.0	268.	0.20	3.1	6.8	17.6	17.1	2.28	3.00
3.0830	2.1	14.4	14.2	283.	0.30	2.3	7.0	17.6	17.2	2.28	2.90
3.0930	2.0	14.4	14.3	293.	0.20	2.6	7.2	17.8	17.6	2.28	2.55
3.1030	2.0	14.4	14.3	293.	0.	3.5	6.8	17.8	17.6	2.20	2.55
3.1130	1.9	14.4	14.3	304.	0.	3.3	7.2	17.6	17.6	2.20	2.55
3.1230	1.9	14.4	14.4	319.	0.	3.5	7.6	17.6	17.7	2.12	2.51
3.1330	1.7	14.4	14.5	324.	0.	3.1	7.0	17.6	17.8	2.12	2.35
3.1430	1.7	14.4	14.5	304.	0.	5.2	7.4	17.6	17.8	2.12	2.36
3.1530	1.5	13.8	14.7	283.	0.	3.5	7.6	17.8	18.1	2.28	2.06
3.1630	1.7	13.2	14.5	278.	0.	3.7	8.2	17.8	17.9	2.20	2.13
3.1730	1.5	13.8	14.7	268.	0.	4.1	8.2	17.8	17.9	2.20	2.22
3.1830	1.5	13.8	14.7	283.	0.10	3.9	8.6	17.8	17.9	2.20	2.19
3.1930	1.6	13.5	14.6	268.	0.10	3.0	7.8	17.8	17.9	2.20	2.17
3.2030	1.5	13.8	14.7	253.	0.	3.7	8.6	17.8	17.9	2.12	2.22
3.2130	1.7	14.1	14.5	248.	0.10	4.3	8.6	18.0	18.2	1.96	2.07
3.2230	1.8	13.8	14.4	243.	0.10	5.0	8.8	18.0	18.1	1.96	2.08
3.2330	1.9	13.8	14.4	238.	0.20	2.5	8.4	18.1	18.1	1.96	2.09
4.0030	2.0	13.8	14.3	223.	0.10	3.3	8.6	18.1	18.1	1.96	2.10
4.0130	2.0	13.5	14.3	228.	0.10	8.5	7.8	18.1	18.1	1.96	2.06
4.0230	1.7	13.8	14.5	253.	0.30	4.1	8.0	17.8	15.6	2.12	3.88
4.0330	1.8	13.8	14.5	333.	0.20	1.3	7.8	20.0	20.0	0.34	0.72
4.0430	1.5	13.8	14.7	384.	0.20	3.0	8.2	17.0	17.5	2.76	2.51
4.0530	2.6	13.0	13.9	339.	0.20	3.5	8.8	17.8	17.8	2.20	2.25
4.0630	4.8	11.7	12.2	278.	0.50	2.4	6.8	17.8	18.1	2.12	2.07
4.0730	3.5	12.4	13.2	303.	0.30	3.4	7.6	17.5	17.8	2.28	2.28
4.0830	4.0	12.2	12.8	333.	0.20	3.7	8.2	17.6	17.9	2.20	2.23
4.0930	4.0	12.2	12.8	353.	0.20	3.9	7.8	17.6	17.9	2.12	2.25
4.1030	1.0	14.1	15.1	450.	0.40	3.7	7.0	17.3	17.1	2.44	2.76
4.1130	1.4	14.1	14.8	495.	0.50	3.9	7.4	16.8	17.1	2.36	2.80
4.1230	1.0	14.1	15.0	496.	0.80	3.5	6.4	15.5	15.7	3.01	3.72
4.1330	1.0	14.1	15.0	529.	0.80	4.3	6.4	15.5	15.9	3.10	3.62

APPENDIX C: TABLE V.

RUN 6: GAS COMPOSITIONS

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
4.1430	1.0	14.1	15.0	494.	0.50	4.5	7.0	15.5	16.2	3.01	3.41
4.1530	1.0	13.8	15.1	494.	0.50	3.1	7.4	15.8	17.5	2.28	2.44
4.1630	1.5	13.8	14.7	425.	0.50	2.8	7.0	15.8	17.4	2.28	2.51
4.1730	2.2	13.5	14.2	362.	0.50	3.5	7.4	16.0	17.6	2.28	2.46
4.1830	2.5	13.0	13.9	359.	0.50	3.9	7.4	16.2	17.6	2.12	2.35
4.1930	2.2	13.5	14.1	306.	0.50	3.6	7.8	16.4	17.4	2.12	2.62
4.2030	2.8	13.2	13.7	253.	0.50	3.7	7.2	16.5	17.6	2.36	2.47
4.2130	3.5	12.7	13.2	217.	0.40	2.3	7.4	16.5	17.9	2.28	2.24
4.2230	2.5	13.2	13.9	187.	0.70	2.8	6.8	16.8	17.6	3.63	2.43
4.2330	2.2	13.5	14.1	167.	0.40	2.3	6.6	16.8	17.5	2.12	2.53
5.0030	2.2	13.2	14.1	172.	0.30	4.3	8.2	17.0	17.5	1.96	2.46
5.0130	2.2	13.5	14.1	182.	0.40	2.6	7.8	17.0	17.3	2.04	2.64
5.0230	2.5	13.2	13.9	206.	0.30	3.9	5.4	17.0	17.4	1.96	2.58
5.0330	2.2	13.2	14.2	216.	0.50	2.5	8.2	17.0	17.3	1.88	2.58
5.0430	2.5	13.2	14.0	241.	0.40	2.0	8.6	17.0	17.5	2.20	2.51
5.0530	2.5	13.2	14.0	290.	0.30	2.5	6.8	17.0	17.7	2.36	2.39
5.0630	2.1	13.2	14.3	203.	0.40	3.4	6.2	17.0	17.4	2.52	2.51
5.0730	2.0	13.2	14.4	206.	0.40	3.5	7.2	16.9	17.0	2.44	2.81
5.0830	2.0	13.5	14.3	205.	0.40	3.9	8.2	16.8	17.2	2.20	2.68
5.0930	2.0	13.5	14.3	208.	0.20	4.3	7.8	16.6	17.1	1.96	2.77
5.1030	1.6	13.8	14.6	193.	0.20	5.5	8.2	15.5	17.4	2.36	2.59
5.1130	1.8	13.8	14.4	194.	1.00	5.0	7.8	15.5	17.5	2.28	2.52
5.1230	2.2	13.5	14.1	196.	1.00	2.0	6.2	15.5	17.6	2.28	2.47
5.1330	1.5	14.1	14.7	176.	0.50	2.8	7.6	16.5	17.4	2.28	2.60
5.1430	2.5	13.2	13.9	169.	0.30	6.2	7.4	16.3	17.5	2.20	2.53
5.1530	2.0	13.8	14.3	177.	0.40	3.7	7.4	16.0	17.4	2.28	2.64
5.1630	2.1	13.8	14.2	182.	0.40	4.2	7.0	16.2	17.4	2.28	2.65
5.1730	1.9	14.1	14.3	178.	0.50	4.0	7.0	16.2	17.3	2.28	2.73

5.1830	2.2	13.8	14.1	187.	0.50	4.1	7.4	16.0	17.4	2.28	2.63
5.1930	2.8	13.5	13.7	-	0.40	4.4	7.0	16.0	17.5	2.28	2.62
5.2030	2.1	13.8	14.2	404.	0.40	4.5	6.8	15.9	17.3	2.36	2.68
5.2130	2.0	13.5	14.3	354.	0.50	4.7	8.4	15.9	17.3	2.44	2.61
5.2230	2.5	13.5	13.9	371.	0.60	4.5	7.0	15.9	17.3	2.36	2.67
5.2330	2.0	13.8	14.3	364.	0.70	4.3	7.0	15.9	17.2	2.28	2.78
6.0030	2.0	13.5	14.3	326.	0.60	4.7	7.4	15.8	17.2	2.28	2.67
6.0130	2.2	13.2	14.1	324.	0.60	4.4	7.2	16.0	17.3	2.20	2.61
6.0230	2.2	13.2	14.1	390.	0.50	4.2	7.0	15.9	17.3	2.28	2.57
6.0330	2.2	13.5	14.1	376.	0.50	4.4	7.2	15.9	17.3	2.28	2.65
6.0430	2.3	13.5	14.0	388.	0.60	4.4	7.0	16.0	17.2	2.20	2.72
6.0530	2.2	13.2	14.1	338.	0.50	4.4	7.0	16.0	17.3	2.20	2.57
6.0630	3.2	12.7	13.4	325.	0.60	4.0	7.2	16.1	17.7	2.04	2.39
6.0730	2.5	13.2	13.9	354.	0.50	3.8	7.2	16.2	17.5	2.12	2.49
6.0830	2.2	13.5	14.1	342.	0.60	4.1	7.2	16.8	17.5	2.12	2.51
6.0930	1.9	13.8	14.3	334.	0.60	4.2	7.2	17.6	17.5	2.12	2.56
6.1030	2.1	13.5	14.2	324.	0.50	4.4	7.2	17.6	17.6	2.12	2.41
6.1130	3.0	12.7	13.5	311.	0.30	5.4	7.0	17.9	17.8	1.96	2.24
6.1230	2.0	13.5	14.3	306.	0.50	4.3	7.0	17.4	17.9	2.20	2.23
6.1330	1.8	13.8	14.4	322.	0.60	6.4	6.6	17.6	17.8	2.12	2.28
6.1430	3.0	13.0	13.5	303.	0.40	4.7	6.2	17.8	17.9	2.12	2.20
6.1530	2.0	13.5	14.3	304.	0.30	5.0	6.6	17.7	17.7	2.12	2.33
6.1630	1.9	13.8	14.3	318.	0.30	7.5	5.8	17.7	17.7	2.20	2.37
6.1730	2.0	13.5	14.3	324.	0.30	5.0	7.4	17.7	17.7	2.12	2.33
6.1830	2.0	13.5	14.3	318.	0.20	4.3	7.4	18.1	18.1	2.04	2.06
6.1930	1.9	13.8	14.4	331.	0.80	4.1	7.4	18.1	18.1	2.12	2.09
6.2030	MISSED DATA READING										
6.2130	2.4	13.5	14.0	313.	0.60	4.2	7.0	18.1	18.1	2.04	2.11
6.2230	2.1	13.5	14.2	317.	0.25	4.0	7.0	18.2	18.2	2.20	2.00
6.2330	2.2	13.5	14.1	324.	0.25	5.5	5.0	17.6	17.5	2.36	2.49
7.0030	2.3	13.5	14.0	298.	0.25	5.2	6.8	18.0	17.6	2.20	2.48
7.0130	2.5	13.2	13.9	286.	0.25	3.9	7.6	18.2	18.2	1.96	2.00
7.0230	2.2	13.5	14.1	273.	0.25	5.3	8.6	17.9	17.7	2.20	2.36
7.0330	2.4	13.2	14.0	303.	0.20	3.6	9.0	18.0	17.7	2.12	2.31
7.0430	2.4	13.5	14.0	324.	0.15	4.5	8.8	18.1	18.1	2.04	2.11
7.0530	2.8	13.2	13.7	305.	0.15	2.9	7.8	18.0	18.0	2.04	2.20

APPENDIX C: TABLE V.

RUN 6: GAS COMPOSITIONS

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
7.0630	2.2	13.5	14.1	286.	0.15	4.1	8.0	17.9	17.6	2.12	2.42
7.0730	2.2	13.5	14.1	279.	0.10	4.6	8.4	18.0	17.7	2.04	2.40
7.0830	2.3	13.5	14.0	263.	0.10	3.6	8.6	18.0	17.4	2.12	2.59
7.0930	2.5	13.2	13.9	255.	0.10	4.5	9.0	17.8	17.8	2.12	2.30
7.1030	2.8	13.2	13.7	243.	0.10	4.9	9.0	17.8	17.8	1.96	2.36
7.1130	2.2	13.5	14.1	243.	0.20	4.5	8.8	17.2	17.6	2.12	2.47
7.1230	2.6	13.2	13.9	266.	0.20	4.1	9.0	17.4	17.6	1.96	2.42
7.1330	2.6	13.2	13.8	281.	0.30	5.3	8.0	17.5	17.7	2.04	2.35
7.1430	2.7	13.5	13.8	298.	0.30	4.3	8.6	17.4	17.7	2.04	2.43
7.1530	2.2	13.5	14.2	292.	0.30	4.9	8.2	17.5	17.5	2.04	2.50
7.1630	2.0	13.5	14.3	312.	0.20	5.1	8.2	17.4	17.6	2.04	2.45
7.1730	2.2	13.2	14.1	321.	0.20	5.1	8.2	17.5	17.6	2.04	2.40
7.1830	1.9	13.8	14.4	324.	0.20	5.2	8.0	17.6	17.5	2.12	2.50
7.1930	2.1	13.8	14.2	283.	0.20	4.3	8.2	17.6	17.6	2.12	2.50
7.2030	2.3	13.0	14.1	303.	0.30	4.1	8.2	17.6	17.6	2.12	2.35
7.2130	2.4	13.2	14.0	334.	0.30	4.5	7.8	17.6	17.7	2.04	2.35
7.2230	2.3	13.2	14.1	364.	0.40	4.0	8.0	17.4	17.7	2.12	2.31
7.2330	1.8	13.8	14.5	392.	0.30	6.0	8.2	17.1	17.6	2.44	2.45
8.0030	1.9	13.5	14.4	384.	0.30	5.7	8.6	17.1	17.6	2.44	2.39
8.0130	2.5	13.2	13.9	388.	0.30	4.7	8.6	16.3	17.0	2.84	2.87
8.0230	2.5	13.2	13.9	354.	0.25	6.2	8.2	16.3	16.7	3.01	3.04
8.0330	2.3	13.5	14.1	346.	0.30	4.1	8.6	16.3	16.7	2.93	3.10
8.0430	2.6	13.2	13.8	374.	0.30	4.1	9.0	16.5	17.0	2.84	2.87
8.0530	2.3	13.2	14.1	374.	0.30	4.3	9.2	16.6	17.0	2.84	2.82
8.0630	2.4	13.2	14.0	399.	0.20	4.3	9.0	16.6	17.0	2.84	2.82
8.0730	2.6	13.2	13.8	409.	0.30	3.8	9.0	16.6	17.1	2.84	2.78
8.0830	2.3	13.2	14.1	420.	0.30	3.7	9.2	16.7	17.1	2.76	2.75
8.0930	2.3	13.2	14.1	445.	0.30	4.3	9.0	16.6	17.2	2.84	2.66

8.1030	2.2	13.5	14.1	435.	0.30	5.0	8.8	16.7	17.2	2.93	2.71
8.1130	2.1	13.5	14.2	438.	0.30	4.5	8.6	16.6	17.1	2.93	2.76
8.1230	2.2	13.2	14.1	425.	0.30	4.9	8.6	16.6	17.1	2.93	2.76
8.1330	2.2	13.2	14.1	425.	0.20	5.7	8.4	16.6	17.1	2.93	2.76
8.1430	2.5	13.2	13.9	418.	0.20	5.2	8.2	16.7	17.1	2.84	2.76
8.1530	2.3	13.5	14.1	423.	0.40	5.8	7.8	16.7	17.3	2.93	2.65
8.1630	2.3	13.2	14.1	445.	0.40	5.5	8.0	16.7	17.4	2.84	2.55
8.1730	3.8	12.7	12.9	429.	0.40	4.7	7.8	16.3	16.7	3.36	3.17
8.1830	3.1	13.0	13.5	406.	0.40	6.0	7.4	17.1	17.5	2.76	2.51
8.1930	2.9	13.2	13.6	379.	0.40	5.9	7.8	17.1	17.3	2.76	2.68
8.2030	3.0	13.2	13.5	377.	0.40	6.0	7.8	17.3	17.5	2.60	2.59
8.2130	3.1	13.0	13.5	379.	0.40	6.3	8.0	17.0	17.4	2.84	2.59
8.2230	3.5	13.0	13.1	374.	0.50	5.3	7.8	17.2	17.4	2.76	2.69
8.2330	3.4	12.7	13.2	361.	0.40	5.2	7.8	17.2	17.6	2.76	2.42
9.0030	3.5	12.7	13.2	369.	0.40	4.4	7.4	17.2	17.4	2.93	2.59
9.0130	3.5	12.7	13.1	401.	0.40	4.3	7.2	16.6	17.1	3.27	2.86
9.0230	3.5	12.7	13.1	398.	0.40	5.3	7.4	16.6	16.8	3.27	3.01
9.0330	3.3	13.0	13.3	394.	0.40	5.5	7.2	16.5	17.0	3.36	2.92
9.0430	3.5	12.7	13.1	402.	0.40	5.3	7.4	16.6	17.1	3.36	2.86
9.0530	3.5	12.7	13.1	470.	0.40	3.4	7.4	16.6	16.5	3.27	3.26
9.0630	3.5	12.7	13.2	467.	0.40	3.6	7.2	16.6	17.1	3.45	2.86
9.0730	3.4	12.7	13.2	473.	0.50	3.5	6.8	16.6	17.0	3.45	2.86
9.0830	3.4	13.0	13.2	475.	0.50	3.7	6.6	16.6	17.0	3.54	2.92
9.0930	1.2	14.4	14.9	455.	0.50	4.3	6.4	16.1	16.6	3.82	3.17
9.1030	1.0	14.4	15.0	452.	0.50	4.9	6.6	16.1	16.8	3.82	2.99
9.1130	0.9	14.4	15.1	453.	0.40	4.2	6.6	16.1	16.2	3.82	3.42
9.1230	2.2	13.5	14.1	407.	0.30	4.6	7.4	16.3	16.6	3.54	3.15
9.1330	2.0	13.5	14.2	376.	0.10	6.4	7.8	16.5	16.8	3.45	2.99
9.1430	1.9	13.8	14.3	354.	0.20	5.8	7.6	16.5	16.8	3.45	3.05
9.1530	1.5	13.8	14.6	354.	0.20	6.4	7.6	16.5	16.7	3.45	3.05
9.1630	1.7	13.8	14.5	421.	0.20	5.2	8.2	16.5	16.7	3.54	3.05
9.1730	2.0	13.8	14.3	409.	0.10	5.4	8.2	16.6	17.0	3.27	2.88
9.1830	2.4	13.5	14.0	354.	0.10	5.7	7.4	16.6	17.0	3.27	2.88
9.1930	2.5	13.2	13.9	326.	0.10	5.0	7.4	16.6	17.1	3.18	2.77
9.2030	2.2	13.2	14.2	312.	0.10	4.2	7.8	16.6	16.9	3.27	2.87
9.2130	2.4	13.2	13.9	450.	0.20	3.7	7.2	16.2	16.6	3.73	3.12

APPENDIX C: TABLE V.

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
9.2230	2.4	13.5	13.9	424.	0.20	3.5	6.4	16.2	16.4	3.73	3.34
SHUT DOWN AT 9.2230 FOR 23 HOURS											
10.2130	1.9	13.5	14.4	429.	0.80	0.6	4.8	21.2	21.2	0.02	-0.14
10.2230	1.3	14.1	14.8	395.	0.60	2.4	7.0	17.4	16.9	3.18	2.91
10.2330	1.0	14.1	15.0	375.	0.20	3.5	7.8	17.0	16.6	3.36	3.10
11.0030	1.3	13.8	14.8	367.	0.20	3.3	7.6	17.0	17.0	3.18	2.82
11.0130	1.4	14.1	14.7	351.	0.40	2.6	6.2	17.0	17.0	3.27	2.89
11.0230	1.4	13.8	14.7	324.	0.20	3.1	7.4	17.0	17.0	3.10	2.83
11.0330	1.4	13.8	14.7	294.	0.40	3.1	7.2	17.1	17.2	3.01	2.70
11.0430	1.5	13.8	14.6	293.	0.20	3.5	7.6	17.2	17.2	2.93	2.70
11.0530	2.0	13.5	14.3	273.	0.10	3.4	7.6	17.2	17.3	2.76	2.64
11.0630	1.3	13.8	14.8	263.	0.20	3.0	7.4	17.4	17.1	2.76	2.73
11.0730	1.8	13.8	14.4	247.	0.20	2.2	7.4	17.3	17.2	2.68	2.71
11.0830	1.8	13.8	14.4	227.	0.20	2.8	7.0	17.4	17.3	2.44	2.69
11.0930	1.2	14.1	14.8	229.	0.20	3.1	6.8	17.6	17.4	2.52	2.59
11.1030	1.1	14.1	14.9	244.	0.30	3.5	5.8	17.5	17.2	2.44	2.66
11.1130	1.0	14.1	15.0	273.	0.20	3.1	6.8	17.7	17.5	2.36	2.43
11.1230	1.2	14.1	14.8	269.	0.25	3.6	6.8	17.6	17.3	2.36	2.61
11.1330	1.2	14.1	14.9	253.	0.25	3.9	5.8	17.8	17.8	2.36	2.24
11.1430	1.7	13.8	14.5	233.	0.25	2.6	6.2	17.0	17.1	2.76	2.76
11.1530	2.0	13.5	14.2	211.	0.30	1.9	6.0	16.8	17.1	2.93	2.74
11.1630	1.9	13.5	14.3	213.	0.30	3.1	7.4	16.9	17.0	2.84	2.80
11.1730	1.5	13.5	14.6	218.	0.20	3.0	7.0	16.9	17.0	2.76	2.75
11.1830	1.7	13.8	14.5	233.	0.20	1.5	7.8	16.9	17.1	2.76	2.81
11.1930	1.9	13.5	14.3	260.	0.15	1.6	8.8	16.7	17.1	2.84	2.76
11.2030	2.0	13.2	14.3	256.	0.15	3.1	8.2	16.8	17.2	2.76	2.61
11.2130	2.0	13.5	14.3	286.	0.30	1.9	7.4	16.4	17.0	3.27	2.87
11.2230	2.0	13.8	14.2	279.	0.10	3.5	7.0	15.5	16.5	3.92	3.23

11.2330	2.4	13.5	13.9	243.	0.10	3.3	5.4	16.1	16.6	3.45	3.22
12.0030	2.8	13.0	13.6	223.	0.10	2.2	5.8	16.2	16.5	3.45	3.20
12.0130	2.2	13.1	14.1	216.	0.30	1.6	6.2	16.6	16.8	3.01	2.92
12.0230	2.5	13.2	13.9	216.	0.20	1.6	7.0	16.0	16.7	3.27	3.07
12.0330	2.5	13.2	13.9	263.	0.20	1.7	6.6	16.2	16.9	3.27	2.96
12.0430	2.9	12.7	13.6	278.	0.25	1.4	6.6	16.3	16.9	3.18	2.89
12.0530	2.9	12.7	13.6	260.	0.25	1.6	5.4	16.3	16.9	3.27	2.87
12.0630	2.9	13.0	13.6	228.	0.25	1.6	5.4	16.4	16.9	3.27	2.92
12.0730	2.3	13.5	14.0	203.	0.25	1.7	6.2	16.5	16.6	3.10	3.16
12.0830	2.0	13.8	14.2	207.	0.25	1.8	6.2	16.4	16.6	3.10	3.21
12.0930	2.0	13.8	14.2	203.	0.50	1.3	4.6	16.4	16.8	3.18	3.08
12.1030	2.1	13.5	14.2	220.	0.20	1.5	7.4	16.7	17.3	2.76	2.67
12.1130	2.2	13.5	14.1	228.	0.20	2.1	6.6	16.3	16.8	3.10	3.02
12.1230	2.2	13.8	14.1	234.	0.20	1.7	6.2	16.4	17.0	3.01	2.90
12.1330	2.2	13.5	14.1	241.	0.20	1.9	6.4	16.4	17.0	3.01	2.87
12.1430	2.1	13.5	14.2	221.	0.20	1.9	6.2	16.4	16.9	3.10	2.93
12.1530	2.0	13.5	14.2	215.	0.30	1.6	6.0	16.4	16.9	3.10	2.93
12.1630	2.0	13.5	14.2	220.	0.40	2.0	4.8	15.7	16.4	3.45	3.30
12.1730	1.5	13.8	14.6	243.	0.30	1.6	5.8	15.0	15.7	3.92	3.73
12.1830	2.0	13.2	14.2	261.	0.20	2.5	6.2	15.7	16.2	3.45	3.36
12.1930	1.9	13.8	14.3	276.	0.40	1.6	5.8	15.6	16.3	3.63	3.38
12.2030	2.2	13.5	14.1	271.	0.30	1.6	6.4	15.9	16.7	3.54	3.07
12.2130	2.2	13.2	14.1	263.	0.30	1.7	6.2	15.9	16.4	3.45	3.25
12.2230	2.3	13.2	14.0	243.	0.30	1.7	6.6	15.8	16.3	3.54	3.31
12.2330	2.6	13.0	13.8	236.	0.30	1.6	7.0	15.9	16.2	3.54	3.37
13.0030	3.0	13.0	13.5	233.	0.20	1.4	7.4	15.7	16.5	3.45	3.25
13.0130	2.5	13.2	13.9	279.	0.20	1.9	5.0	15.5	16.4	3.82	3.31
13.0230	2.3	13.5	14.0	284.	0.20	2.6	7.0	15.5	15.9	3.82	3.66
13.0330	3.0	13.0	13.5	256.	0.20	3.9	6.2	15.6	16.2	3.73	3.47
13.0430	3.0	13.0	13.5	243.	0.20	1.5	6.6	15.6	16.1	3.63	3.50
13.0530	2.9	13.0	13.5	250.	0.20	1.5	6.6	15.6	16.1	3.63	3.54
13.0630	2.8	13.2	13.6	256.	0.20	2.1	6.4	15.5	16.0	3.63	3.64
13.0730	3.0	13.2	13.4	237.	0.20	2.1	6.4	15.5	16.3	3.63	3.43

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
13.0830	2.5	13.5	13.8	223.	0.20	1.9	6.2	15.9	16.2	3.54	3.50
13.0930	2.1	13.5	14.1	223.	0.40	1.4	4.6	15.8	16.1	3.63	3.51
13.1030	2.5	13.5	13.9	253.	0.30	1.4	5.6	15.7	16.3	3.63	3.46
13.1130	2.2	13.5	14.1	253.	0.30	1.6	6.0	15.7	16.2	3.63	3.46
13.1230	2.2	13.5	14.1	227.	0.25	1.6	6.4	15.6	16.2	3.63	3.46
13.1330	2.2	13.5	14.1	239.	0.30	1.2	7.0	15.6	16.4	3.45	3.27
13.1430	2.0	13.5	14.2	250.	0.30	1.7	9.0	16.0	16.5	3.27	3.19
13.1530	2.1	13.5	14.2	282.	0.20	2.0	8.2	15.5	16.4	3.45	3.25
13.1630	1.8	13.5	14.4	324.	0.25	1.9	7.6	15.6	16.3	3.54	3.34
13.1730	2.1	13.8	14.2	332.	0.15	1.6	7.6	15.8	16.4	3.45	3.32
13.1830	2.1	13.5	14.2	336.	0.20	1.7	7.0	15.7	16.5	3.54	3.19
13.1930	2.7	13.0	13.7	351.	0.25	1.3	6.6	15.7	16.3	3.45	3.32
13.2030	2.4	13.0	13.9	450.	0.20	1.6	7.0	15.6	16.4	3.54	3.22
13.2130	2.2	13.2	14.1	728.	0.20	1.0	5.0	15.7	16.4	3.45	3.24
13.2230	2.0	13.5	14.2	687.	0.30	1.0	4.2	15.2	16.0	3.82	3.53
13.2330	1.9	13.5	14.3	621.	0.30	0.8	5.2	15.0	15.9	3.92	3.60
14.0030	1.9	13.5	14.3	546.	0.20	0.8	5.4	14.9	15.6	4.02	3.84
14.0130	2.0	13.5	14.2	556.	0.20	0.8	6.2	14.9	15.6	3.92	3.86
14.0230	2.1	13.2	14.1	569.	0.20	0.6	5.8	14.9	15.7	3.92	3.68
14.0330	2.1	13.2	14.2	577.	0.20	0.6	5.4	14.9	15.5	3.92	3.84
14.0430	2.2	13.0	14.1	554.	0.20	1.4	5.8	14.9	15.1	4.02	4.04
14.0530	2.0	13.2	14.2	538.	1.20	0.8	5.4	15.2	15.6	3.82	3.76
14.0630	2.5	13.0	13.9	598.	0.50	0.6	5.8	16.0	16.2	3.45	3.36
14.0730	2.2	13.2	14.1	612.	1.20	1.0	5.4	15.8	16.8	3.45	2.98
14.0830	2.2	13.2	14.1	572.	1.20	0.8	5.4	16.1	16.7	3.27	3.02
14.0930	2.3	13.2	14.0	544.	1.20	1.2	5.4	15.3	16.5	3.63	3.17
14.1030	2.0	13.2	14.2	516.	0.20	0.8	4.8	15.3	15.8	3.82	3.61
14.1130	2.1	13.5	14.2	494.	0.20	1.0	6.2	15.0	15.7	3.92	3.77

14.1230	2.0	13.8	14.2	492.	0.20	1.0	6.2	15.0	15.7	4.02	3.85
14.1330	2.0	13.5	14.2	496.	0.10	0.9	6.4	15.0	15.6	3.82	3.84
14.1430	2.0	13.5	14.2	475.	0.20	1.0	5.8	14.9	15.6	3.82	3.84
14.1530	2.0	13.5	14.2	504.	0.10	1.5	6.4	14.9	15.9	3.92	3.62
14.1630	1.8	13.5	14.4	514.	0.10	1.3	5.8	14.8	15.5	4.02	3.86
14.1730	2.0	13.5	14.2	506.	0.10	1.0	6.4	14.8	15.5	4.02	3.91
14.1830	1.8	13.5	14.4	528.	0.20	1.7	5.8	14.8	15.4	4.02	3.91
14.1930	2.0	13.5	14.2	506.	0.10	1.1	6.0	14.8	15.8	4.02	3.68
14.2030	2.0	13.5	14.2	512.	0.	1.0	6.0	14.8	16.0	3.82	3.58
14.2130	2.1	13.2	14.2	496.	0.50	0.4	3.9	14.8	15.8	3.73	3.62
14.2230	2.1	13.2	14.1	483.	0.20	0.9	5.2	14.5	15.2	4.02	4.07
14.2330	2.1	13.2	14.2	473.	0.20	0.8	5.2	14.5	15.8	3.82	3.63
15.0030	2.1	13.5	14.2	546.	0.50	0.4	4.4	14.8	15.8	3.82	3.70
15.0130	2.0	13.5	14.2	586.	0.40	0.5	5.0	14.5	15.8	3.82	3.69
15.0230	1.9	13.8	14.3	552.	0.80	0.3	5.0	14.6	15.1	4.21	4.23
15.0330	1.2	13.8	14.8	541.	0.60	1.4	4.2	14.7	15.8	4.02	3.64
15.0430	1.4	13.8	14.7	432.	0.50	2.3	5.4	14.7	16.0	3.92	3.50
15.0530	1.3	13.8	14.8	398.	0.10	1.8	6.4	14.7	15.8	3.92	3.62
15.0630	1.4	13.8	14.7	363.	0.20	1.7	6.8	14.7	15.6	3.92	3.77
15.0730	1.2	14.1	14.8	312.	0.30	1.3	6.6	15.1	16.2	3.73	3.42
15.0830	1.0	14.1	15.0	316.	0.30	1.2	6.8	15.1	15.9	3.63	3.57
15.0930	1.0	14.1	15.0	355.	0.	2.7	5.6	15.1	16.0	3.63	3.53
15.1030	2.0	13.5	14.2	346.	0.10	2.2	6.4	15.1	16.5	3.45	3.21
15.1130	2.0	13.5	14.2	299.	0.10	2.2	6.2	15.1	16.0	3.82	3.56
15.1230	2.0	13.5	14.2	281.	0.10	2.2	6.2	15.1	16.0	3.73	3.55
15.1330	2.0	13.5	14.2	270.	0.	2.5	6.8	14.8	16.0	3.73	3.53
15.1430	2.8	13.5	13.6	269.	0.	2.2	7.2	14.8	16.1	3.63	3.62
15.1530	2.2	13.5	14.1	269.	0.	2.2	6.4	14.8	16.0	3.63	3.62
15.1630	2.2	13.5	14.1	269.	0.	1.9	6.8	15.2	16.0	3.92	3.62
15.1730	2.2	13.8	14.1	269.	0.	2.7	7.0	15.2	16.0	4.02	3.70
15.1830	2.0	13.5	14.2	264.	0.	2.7	6.8	15.1	16.1	3.82	3.47
15.1930	1.6	13.8	14.5	250.	0.	3.0	6.8	15.1	16.5	3.73	3.20
15.2030	2.4	13.5	13.9	277.	0.	1.6	7.4	15.1	16.3	3.73	3.43
15.2130	2.8	13.2	13.6	286.	0.	1.8	7.0	15.4	16.4	3.82	3.36
15.2230	2.5	13.2	13.9	283.	0.10	1.6	6.6	15.4	16.1	3.82	3.47
15.2330	2.2	13.5	14.1	288.	0.10	1.6	7.0	15.4	16.4	3.82	3.30

APPENDIX C: TABLE V.

RUN 6: GAS COMPOSITIONS PAGE 6 OF 7

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
16.0030	2.2	13.2	14.1	294.	0.10	1.6	7.0	15.4	16.4	3.82	3.23
16.0130	2.5	13.2	13.9	296.	0.10	1.5	7.4	15.4	16.2	3.82	3.40
16.0230	2.1	13.5	14.2	297.	0.	1.7	7.0	15.4	16.4	3.73	3.26
16.0330	3.0	12.7	13.5	300.	0.	2.2	6.6	15.7	16.6	3.54	3.12
16.0430	2.8	13.2	13.6	299.	0.10	1.7	6.8	15.7	16.3	3.82	3.40
16.0530	2.9	13.2	13.6	314.	0.10	1.7	7.0	15.7	16.0	3.92	3.68
16.0630	2.4	13.2	13.9	324.	0.	1.6	7.0	15.5	16.1	4.02	3.46
16.0730	2.5	13.2	13.9	324.	0.	1.5	7.0	15.5	16.2	3.82	3.46
16.0830	2.4	13.2	13.9	324.	0.10	1.6	7.0	15.5	16.1	3.82	3.46
16.0930	2.5	13.2	13.9	336.	0.10	1.8	7.0	15.5	16.2	3.82	3.46
16.1030	2.4	13.2	13.9	354.	0.10	1.6	6.6	15.4	16.0	3.92	3.54
16.1130	2.8	13.2	13.6	374.	0.	1.9	6.6	15.0	15.9	4.32	3.70
16.1230	2.8	13.2	13.6	379.	0.10	1.3	6.4	15.0	15.9	4.42	3.70
16.1330	3.0	13.2	13.5	377.	0.20	1.0	6.2	15.0	16.0	4.85	3.70
16.1430	2.9	13.0	13.5	394.	0.10	1.6	6.8	15.0	16.1	4.02	3.48
16.1530	2.4	13.5	13.9	380.	0.10	1.2	6.0	15.0	15.7	4.42	3.84
16.1630	2.5	13.5	13.8	364.	0.20	2.2	6.2	15.0	15.5	4.52	3.99
16.1730	2.5	13.5	13.8	385.	0.10	1.3	6.2	15.0	15.5	4.52	3.99
16.1830	2.0	13.5	14.2	395.	0.10	2.0	6.8	15.4	15.4	4.63	3.99
16.1930	2.5	13.2	13.8	415.	0.10	1.4	8.2	15.4	15.6	4.63	3.84
16.2030	2.5	13.5	13.8	415.	0.10	1.0	5.8	16.0	15.4	4.52	4.06
16.2130	2.8	13.2	13.6	400.	0.10	1.3	6.2	16.8	15.5	4.52	3.99
16.2230	2.9	13.2	13.5	374.	0.20	1.3	7.0	17.2	15.6	4.52	3.91
16.2330	2.9	13.2	13.5	377.	0.30	1.0	6.2	17.6	15.6	4.42	3.91
17.0030	2.9	13.0	13.5	400.	0.20	0.7	5.8	18.0	15.6	4.42	3.83
17.0130	2.5	13.2	13.9	430.	0.30	1.7	5.8	18.5	18.5	4.42	1.79
17.0230	2.7	13.2	13.8	439.	0.30	1.0	5.2	18.8	18.8	4.63	1.59
17.0330	2.5	13.5	13.9	430.	0.30	1.4	5.4	19.1	19.1	4.63	1.39

17.0430	2.4	13.5	14.0	414.	0.40	1.6	6.2	19.5	19.5	4.63	1.09
17.0530	2.8	13.2	13.7	394.	0.40	1.1	5.6	19.8	19.8	4.42	0.87
17.0630	2.7	13.2	13.8	394.	0.30	1.7	5.6	20.0	20.0	4.42	0.72
17.0730	2.9	13.2	13.7	333.	0.20	1.9	6.0	20.0	20.0	4.21	0.73
17.0830	3.1	13.2	13.5	323.	0.20	1.7	6.2	20.0	20.0	4.32	0.74
17.0930	3.1	12.7	13.5	323.	0.30	1.3	6.6	20.3	20.3	4.42	0.50
17.1030	2.9	13.2	14.0	336.	0.30	1.7	6.4	14.3	15.7	4.63	3.88
17.1130	2.9	13.2	13.5	334.	0.30	1.4	6.2	14.2	15.8	4.21	3.77

SHUT DOWN AT 17.1130 FOR 40 HOURS

19.0330	4.0	12.4	12.7	112.	0.50	2.3	4.2	15.8	16.0	4.02	3.69
19.0430	4.2	12.2	12.6	162.	0.50	1.1	6.2	15.9	16.1	3.82	3.59
19.0530	3.4	13.0	13.2	218.	0.30	1.3	7.4	15.2	15.7	4.21	3.94
19.0630	3.3	13.0	13.2	268.	0.50	1.3	7.2	15.2	16.0	4.02	3.67
19.0730	3.0	13.0	13.5	304.	0.30	1.4	7.0	15.2	15.5	3.82	3.97
19.0830	2.9	13.0	13.6	344.	0.30	1.4	7.0	15.2	15.9	3.82	3.69
19.0930	3.1	13.0	13.4	369.	0.40	1.8	6.6	15.2	15.6	3.63	3.95
19.1030	2.9	13.1	13.5	379.	0.20	1.4	7.0	15.3	15.8	3.45	3.80
19.1130	2.8	13.2	13.6	369.	0.20	2.0	7.4	15.2	15.7	3.45	3.84
19.1230	2.2	13.8	14.1	365.	0.30	2.6	7.0	15.2	15.8	3.45	3.80
19.1330	2.2	13.8	14.1	365.	0.30	2.3	7.8	15.5	16.0	2.93	3.64
19.1430	2.5	13.1	13.9	361.	0.30	2.5	8.2	15.5	16.2	3.10	3.37
19.1530	2.5	13.2	13.9	326.	0.20	1.1	8.6	16.0	16.2	2.76	3.41
19.1630	2.5	13.2	13.9	506.	0.20	1.6	7.0	15.8	16.2	2.93	3.41
19.1730	2.5	13.2	13.8	293.	0.20	1.6	6.6	14.9	15.7	3.27	3.77
19.1830	2.5	13.2	13.8	260.	0.50	2.5	7.6	14.9	15.6	3.27	3.88
19.1930	2.5	13.2	13.9	276.	0.30	2.3	7.4	15.0	16.1	3.10	3.47
19.2030	2.4	13.2	13.9	322.	0.	10.0	2.2	15.0	15.7	3.10	3.74
19.2130	2.7	13.0	13.7	268.	0.10	4.3	8.2	15.1	15.9	3.10	3.64
19.2230	3.0	13.0	13.5	304.	0.10	3.3	7.6	15.1	15.8	3.10	3.76
19.2330	3.2	12.7	13.3	349.	0.10	3.3	7.8	15.3	16.0	3.01	3.58
20.0030	3.0	12.7	13.5	311.	0.20	2.6	7.4	15.2	15.9	2.93	3.61
20.0130	3.2	12.7	13.3	307.	0.10	1.9	8.6	15.2	15.9	2.93	3.61
20.0230	3.2	12.7	13.3	326.	0.	3.7	8.0	15.0	15.9	3.10	3.61

APPENDIX C: TABLE V.

RUN 6: GAS COMPOSITIONS

	FLUE GAS				REGENERATOR GAS			GASIFIER INLET GAS			
DAY.HOUR	O2	CO2 VOL %		SO2	O2	CO2	SO2	O2 VOL %		CO2 VOL %	
	%	ANAL	CALC	PPM	%	%	%	ANAL	CALC	ANAL	CALC
20.0330	1.2	12.4	14.8	330.	0.15	2.1	7.0	15.1	15.3	3.01	3.61
20.0430	0.4	12.7	15.4	360.	0.15	2.5	8.2	15.1	15.0	3.10	3.68
20.0530	1.0	11.7	15.0	355.	0.10	2.3	8.2	15.1	15.3	3.01	3.35
20.0630	3.6	12.4	13.0	354.	0.15	1.6	7.8	15.2	16.0	3.01	3.55
20.0730	3.2	12.7	13.3	354.	0.10	2.0	7.4	15.2	16.0	3.01	3.55
20.0830	3.0	12.7	13.5	334.	0.20	3.5	7.0	15.2	16.1	2.93	3.46
20.0930	3.0	12.7	13.5	326.	0.20	1.9	6.2	15.2	16.0	2.93	3.53
20.1030	3.0	12.7	13.5	334.	0.20	2.8	6.8	15.2	15.9	2.93	3.61
20.1130	3.0	12.7	13.5	342.	0.20	4.2	7.4	15.3	16.1	2.93	3.46
20.1230	3.0	12.7	13.5	457.	0.20	2.9	7.2	15.2	16.2	2.93	3.38
20.1330	3.0	12.7	13.5	-	0.30	4.7	7.6	15.2	16.2	2.93	3.38
20.1430	3.0	12.7	13.5	-	0.20	3.2	7.4	15.2	16.2	2.93	3.38
20.1530	3.0	12.7	13.5	-	0.30	2.3	7.0	15.2	16.1	2.93	3.45
20.1630	3.1	12.7	13.4	326.	0.30	2.5	6.2	15.0	16.1	3.10	3.46
20.1730	3.0	12.7	13.5	303.	0.30	2.8	7.2	14.8	15.9	3.10	3.61
20.1830	3.1	12.7	13.4	278.	0.40	1.9	6.8	15.6	16.7	2.76	3.06
20.1930	3.0	13.0	13.5	276.	0.50	2.5	7.0	15.5	16.6	2.76	3.19
20.2030	3.1	12.7	13.4	278.	0.70	4.5	6.8	15.5	16.7	2.76	3.06
20.2130	3.0	12.7	13.5	273.	1.40	2.3	6.8	15.5	16.7	2.76	3.06
20.2230	3.0	13.0	13.5	273.	1.10	2.0	7.6	15.8	16.8	2.68	3.01
20.2330	3.2	12.7	13.3	283.	1.00	2.2	7.0	15.9	16.7	2.68	3.06
21.0030	3.4	12.4	13.2	281.	0.40	2.7	8.0	16.3	16.8	2.28	3.00
21.0130	3.4	12.7	13.2	283.	0.10	2.6	7.4	15.9	16.8	2.68	3.06
21.0230	3.5	12.4	13.1	288.	0.20	1.9	7.0	15.7	16.4	2.76	3.26
21.0330	3.5	12.4	13.1	293.	0.10	2.3	6.2	15.5	16.2	2.84	3.40
21.0430	3.6	12.4	13.0	283.	0.10	2.3	8.2	15.6	16.3	2.84	3.34
STONE CHANGE											
21.0530	3.2	12.7	13.3	289.	0.20	3.3	8.0	15.5	16.2	3.01	3.41

21.0630	3.5	12.4	13.1	324.	0.20	2.3	7.4	15.5	16.0	2.84	3.54
21.0730	2.6	13.0	13.8	375.	0.20	3.1	6.2	15.4	16.1	3.01	3.48
21.0830	2.6	13.2	13.8	402.	0.20	3.5	6.2	14.9	15.8	3.10	3.76
21.0930	2.7	13.2	13.7	405.	0.20	3.1	6.4	15.3	16.0	3.01	3.62
21.1030	2.8	13.0	13.6	374.	0.30	2.5	6.4	15.3	16.3	2.93	3.34
21.1130	2.6	13.0	13.8	367.	0.20	3.3	6.2	15.5	16.5	2.93	3.19
21.1230	2.8	13.0	13.6	357.	0.20	2.3	6.4	15.6	16.6	2.84	3.13
21.1330	2.8	13.0	13.6	344.	0.20	2.8	5.4	15.8	16.6	2.84	3.13
21.1430	2.8	13.0	13.6	354.	0.80	2.9	5.8	15.6	16.6	2.76	3.13
21.1530	3.0	13.0	13.5	370.	0.10	3.3	5.6	15.2	16.3	3.10	3.40
21.1630	3.2	12.7	13.3	369.	0.20	3.5	5.6	15.3	15.9	3.01	3.61
21.1730	2.9	13.0	13.5	367.	0.20	4.1	5.4	15.2	15.9	3.10	3.69
21.1830	3.0	13.0	13.5	383.	0.30	3.1	5.6	15.2	15.9	3.10	3.69
21.1930	3.0	12.7	13.5	391.	0.50	2.1	4.8	15.1	15.9	3.10	3.61
21.2030	2.8	13.2	13.6	384.	0.40	3.3	6.2	15.1	15.9	3.10	3.69
21.2130	2.8	13.2	13.6	346.	0.50	3.9	5.8	15.1	15.9	3.10	3.69
21.2230	3.4	12.7	13.2	354.	0.80	2.6	5.8	15.0	16.2	3.01	3.47
21.2330	3.2	12.7	13.3	369.	0.80	3.0	6.2	15.1	16.1	3.01	3.47
22.0030	3.0	12.7	13.5	387.	1.00	3.0	5.4	14.6	15.7	3.27	3.75
22.0130	3.0	12.7	13.5	443.	0.70	3.5	5.6	14.2	15.7	3.36	3.75
22.0230	3.2	12.7	13.3	430.	0.70	3.9	5.6	14.5	15.9	3.36	3.61
22.0330	3.2	12.7	13.3	410.	0.80	3.7	5.4	14.5	15.9	3.27	3.61
22.0430	3.0	12.7	13.5	394.	0.80	3.1	5.4	14.5	15.7	3.27	3.75
22.0530	3.0	12.7	13.5	334.	0.80	2.8	4.8	14.8	15.7	3.10	3.75
22.0630	3.0	12.7	13.5	384.	0.80	2.8	5.0	15.0	15.7	3.10	3.75
22.0730	3.0	12.7	13.5	334.	0.80	3.0	4.2	15.0	15.7	3.10	3.75
22.0830	3.4	12.7	13.2	384.	1.00	2.8	5.4	14.9	16.3	3.10	3.41
22.0930	3.0	12.7	13.5	409.	0.90	3.0	5.6	14.9	15.9	3.18	3.57
22.1030	3.0	12.7	13.5	403.	1.00	2.6	4.2	14.9	15.9	3.10	3.57
22.1130	3.0	12.7	13.5	393.	0.90	2.8	5.6	14.9	16.0	3.10	3.52
22.1230	2.6	12.7	13.8	415.	1.10	3.0	5.4	14.9	15.9	3.10	3.52
22.1330	2.5	13.0	13.9	474.	1.00	3.1	5.4	14.8	15.9	3.18	3.59
22.1430	2.6	12.7	13.8	484.	1.00	4.3	5.0	15.0	16.1	3.01	3.37
22.1530	2.2	13.2	14.1	518.	0.90	2.6	5.0	14.8	16.6	3.18	3.12
22.1630	2.9	13.0	13.6	508.	0.90	3.5	5.0	15.1	16.4	3.01	3.30
22.1730	3.0	12.7	13.5	497.	0.20	2.0	4.2	15.2	16.4	2.93	3.23

APPENDIX C: TABLE VI.

RUN 6: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
1.2230	0.124	0.009	-	0.001	0.115	7.0	1.5	5.5
1.2330	0.261	0.028	(-0.001)	0.003	0.231	10.5	3.0	7.5
2.0030	0.401	0.050	0.064	0.004	0.282	10.5	4.6	5.9
2.0130	0.540	0.074	0.164	0.006	0.297	10.5	6.1	4.4
2.0230	0.684	0.101	0.271	0.007	0.304	10.5	7.6	2.9
2.0330	0.827	0.128	0.412	0.009	0.278	15.1	9.1	6.0
2.0430	0.973	0.152	0.555	0.010	0.256	24.8	10.6	14.2
2.0530	1.118	0.179	0.683	0.013	0.243	32.9	13.3	19.6
2.0630	1.265	0.209	0.820	0.023	0.213	40.7	21.8	18.9
2.0730	1.412	0.240	0.949	0.038	0.185	48.8	32.3	16.5
2.0830	1.558	0.274	1.074	0.042	0.168	57.4	35.2	22.2
2.0930	1.707	0.309	1.184	0.046	0.168	63.6	38.2	25.4
2.1030	1.856	0.344	1.309	0.060	0.143	69.2	47.0	22.2
2.1130	2.005	0.381	1.408	0.090	0.126	75.4	66.1	9.4
2.1230	2.152	0.417	1.505	0.095	0.135	82.4	69.1	13.3
2.1330	2.301	0.454	1.604	0.126	0.117	91.9	87.9	4.0
2.1430	2.451	0.492	1.693	0.158	0.109	98.6	106.7	-8.1
2.1530	2.600	0.530	1.795	0.162	0.113	105.3	109.8	-4.4
2.1630	2.748	0.566	1.893	0.167	0.123	114.2	112.4	1.9
2.1730	2.897	0.602	1.991	0.171	0.132	123.9	114.8	9.1
2.1830	3.046	0.637	2.104	0.176	0.129	134.4	117.2	17.2
2.1930	3.195	0.670	2.206	0.180	0.139	146.0	119.7	26.4
2.2030	3.344	0.702	2.316	0.209	0.117	155.2	137.6	17.6
2.2130	3.491	0.734	2.426	0.216	0.116	164.9	141.4	23.5
2.2230	3.641	0.766	2.507	0.222	0.146	175.7	145.1	30.6
2.2330	3.789	0.795	2.622	0.228	0.144	185.4	148.2	37.1
3.0030	3.938	0.825	2.729	0.233	0.151	196.1	151.3	44.8
3.0130	4.086	0.854	2.836	0.238	0.158	206.4	154.4	52.0

3.0230	4.234	0.883	2.944	0.243	0.164	216.4	157.5	58.9
3.0330	4.382	0.911	3.040	0.267	0.164	226.6	171.9	54.7
3.0430	4.530	0.940	3.148	0.271	0.172	237.9	174.8	63.1
3.0530	4.680	0.968	3.261	0.276	0.175	248.7	177.8	70.9
3.0630	4.828	0.996	3.382	0.280	0.170	258.1	180.7	77.4
3.0730	4.975	1.023	3.485	0.310	0.157	264.9	199.5	65.3
3.0830	5.123	1.051	3.589	0.321	0.162	274.8	206.6	68.2
3.0930	5.271	1.080	3.696	0.335	0.160	282.9	216.0	66.9
3.1030	5.420	1.109	3.803	0.338	0.170	293.1	217.7	75.5
3.1130	5.569	1.139	3.915	0.340	0.175	303.4	219.3	84.1
3.1230	5.720	1.171	4.034	0.342	0.173	310.7	220.9	89.7
3.1330	5.871	1.203	4.140	0.345	0.183	319.3	222.6	96.7
3.1430	6.023	1.233	4.257	0.347	0.186	327.6	224.2	103.4
3.1530	6.173	1.261	4.373	0.372	0.167	335.4	240.8	94.6
3.1630	6.324	1.288	4.500	0.395	0.141	344.1	255.6	88.5
3.1730	6.476	1.315	4.625	0.399	0.137	355.7	258.7	97.0
3.1830	6.628	1.342	4.756	0.404	0.126	370.5	261.8	108.7
3.1930	6.780	1.369	4.870	0.408	0.133	383.9	264.9	119.1
3.2030	6.932	1.394	5.000	0.432	0.107	396.6	281.4	115.2
3.2130	7.083	1.418	5.132	0.437	0.096	408.2	285.0	123.2
3.2230	7.236	1.442	5.265	0.442	0.086	420.9	288.6	132.3
3.2330	7.386	1.466	5.389	0.447	0.084	431.4	292.2	139.2
4.0030	7.539	1.489	5.519	0.450	0.081	444.6	295.0	149.5
4.0130	7.690	1.511	5.641	0.454	0.083	456.2	297.9	158.2
4.0230	7.841	1.536	5.759	0.458	0.087	460.2	300.8	159.4
4.0330	7.992	1.570	5.870	0.462	0.091	460.2	303.6	156.5
4.0430	8.144	1.608	5.989	0.465	0.082	460.2	306.5	153.7
4.0530	8.293	1.643	6.120	0.468	0.062	460.2	308.9	151.3
4.0630	8.438	1.674	6.221	0.471	0.071	460.2	311.1	149.1
4.0730	8.586	1.707	6.336	0.474	0.069	460.2	313.3	146.9
4.0830	8.736	1.744	6.460	0.477	0.055	460.2	315.6	144.6
4.0930	8.887	1.784	6.581	0.480	0.042	460.2	317.9	142.3
4.1030	9.037	1.826	6.687	0.483	0.040	460.2	320.2	140.0
4.1130	9.187	1.875	6.802	0.485	0.026	460.2	321.2	139.0
4.1230	9.338	1.922	6.900	0.486	0.030	460.2	322.1	138.1
4.1330	9.490	1.973	7.000	0.487	0.030	460.2	323.1	137.1

APPENDIX C: TABLE VI.

RUN 6: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
4.1430	9.640	2.020	7.109	0.490	0.022	460.2	324.9	135.3
4.1530	9.791	2.067	7.223	0.513	-0.011	467.2	342.7	124.5
4.1630	9.942	2.108	7.327	0.516	-0.009	475.8	345.4	130.5
4.1730	10.093	2.144	7.437	0.535	-0.025	484.2	360.5	123.7
4.1830	10.243	2.181	7.550	0.537	-0.025	492.5	362.1	130.4
4.1930	10.395	2.213	7.669	0.539	-0.025	499.3	363.5	135.8
4.2030	10.545	2.239	7.778	0.541	-0.013	507.3	364.9	142.4
4.2130	10.696	2.263	7.889	0.543	0.002	517.9	366.3	151.5
4.2230	10.843	2.281	7.992	0.564	0.006	525.9	382.8	143.1
4.2330	10.994	2.298	8.087	0.566	0.043	535.4	384.4	151.0
5.0030	11.151	2.317	8.200	0.568	0.066	546.4	386.4	160.0
5.0130	11.301	2.335	8.305	0.570	0.091	555.6	388.4	167.2
5.0230	11.450	2.356	8.376	0.573	0.146	562.3	390.4	171.9
5.0330	11.600	2.378	8.485	0.576	0.161	571.2	394.0	177.2
5.0430	11.751	2.402	8.606	0.582	0.161	579.8	399.4	180.4
5.0530	11.901	2.432	8.698	0.587	0.184	588.2	404.8	183.4
5.0630	12.051	2.452	8.785	0.592	0.222	596.3	410.2	186.1
5.0730	12.201	2.472	8.881	0.598	0.249	604.6	415.6	189.0
5.0830	12.351	2.493	8.986	0.601	0.270	615.7	419.5	196.1
5.0930	12.499	2.513	9.086	0.605	0.295	624.3	423.0	201.3
5.1030	12.653	2.533	9.197	0.617	0.307	632.9	433.9	199.0
5.1130	12.805	2.552	9.302	0.620	0.330	642.6	437.3	205.3
5.1230	12.955	2.572	9.381	0.626	0.375	653.4	442.4	210.9
5.1330	13.104	2.589	9.493	0.631	0.390	664.2	445.7	218.4
5.1430	13.253	2.606	9.618	0.636	0.393	675.5	449.1	226.4
5.1530	13.404	2.624	9.729	0.640	0.411	685.7	452.4	233.3
5.1630	13.553	2.642	9.834	0.662	0.414	696.8	469.7	227.1
5.1730	13.703	2.660	9.940	0.697	0.406	706.5	496.0	210.4

5.1830	13.852	2.679	10.056	0.700	0.417	714.0	498.4	215.6
5.1930	-	-	-	-	-	721.5	509.8	211.7
5.2030	14.002	2.719	10.160	0.703	0.420	731.2	512.0	219.2
5.2130	14.151	2.755	10.292	0.707	0.397	740.1	515.1	225.0
5.2230	14.299	2.792	10.400	0.711	0.396	747.9	518.4	229.6
5.2330	14.449	2.829	10.507	0.715	0.397	755.8	521.8	233.9
6.0030	14.598	2.861	10.622	0.734	0.381	764.7	536.1	228.5
6.0130	14.747	2.894	10.731	0.739	0.383	771.7	540.5	231.2
6.0230	14.895	2.933	10.836	0.757	0.370	780.0	554.0	226.0
6.0330	15.042	2.970	10.944	0.763	0.365	789.2	559.0	230.1
6.0430	15.194	3.010	11.047	0.769	0.367	804.3	564.0	240.2
6.0530	15.344	3.044	11.154	0.776	0.370	815.3	569.0	246.3
6.0630	15.492	3.078	11.260	0.782	0.372	827.4	573.9	253.5
6.0730	15.642	3.114	11.368	0.788	0.372	841.4	578.8	262.6
6.0830	15.791	3.149	11.474	0.795	0.374	852.8	584.2	268.5
6.0930	15.941	3.182	11.580	0.803	0.377	868.7	590.2	278.5
6.1030	16.093	3.215	11.690	0.823	0.366	876.2	605.3	270.8
6.1130	16.241	3.247	11.797	0.843	0.354	889.7	620.6	269.0
6.1230	16.390	3.278	11.899	0.860	0.354	902.3	633.4	269.0
6.1330	16.539	3.309	12.000	0.880	0.349	912.0	648.6	263.4
6.1430	16.688	3.341	12.093	0.909	0.345	925.2	670.3	255.0
6.1530	16.839	3.371	12.187	0.918	0.363	941.1	676.8	264.4
6.1630	16.991	3.403	12.266	0.926	0.396	952.7	682.0	270.7
6.1730	17.142	3.436	12.368	0.931	0.408	968.6	686.1	282.6
6.1830	17.297	3.468	12.468	0.944	0.416	977.5	695.9	281.6
6.1930	17.450	3.502	12.569	0.951	0.428	992.1	701.1	290.9
6.2030			MISSED DATA READING					
6.2130	17.600	3.534	12.665	0.958	0.444	1003.9	706.3	297.6
6.2230	17.752	3.566	12.758	0.983	0.444	1015.8	724.2	291.5
6.2330	17.903	3.599	12.821	1.008	0.475	1023.3	741.3	282.0
7.0030	18.056	3.630	12.900	1.026	0.499	1038.1	754.2	284.0
7.0130	18.209	3.660	12.985	1.035	0.530	1053.8	760.2	293.5
7.0230	18.360	3.688	13.089	1.049	0.534	1064.5	769.9	294.6
7.0330	18.512	3.719	13.202	1.066	0.525	1078.3	780.8	297.5
7.0430	18.664	3.752	13.313	1.080	0.517	1089.6	790.5	299.1
7.0530	18.815	3.784	13.409	1.096	0.525	1104.4	800.9	303.5

APPENDIX C: TABLE VI.

RUN 6: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
7.0630	18.965	3.813	13.509	1.116	0.526	1115.2	814.6	300.6
7.0730	19.116	3.842	13.615	1.124	0.535	1128.7	820.3	308.4
7.0830	19.267	3.869	13.721	1.131	0.547	1141.1	825.7	315.4
7.0930	19.417	3.895	13.834	1.151	0.536	1150.5	837.8	312.7
7.1030	19.567	3.920	13.950	1.158	0.539	1159.4	842.6	316.8
7.1130	19.718	3.945	14.060	1.164	0.549	1166.1	846.9	319.2
7.1230	19.868	3.972	14.181	1.208	0.507	1169.6	871.4	298.2
7.1330	20.019	4.001	14.289	1.213	0.515	1173.7	874.5	299.1
7.1430	20.170	4.033	14.416	1.219	0.502	1178.0	878.6	299.4
7.1530	20.321	4.062	14.531	1.226	0.501	1183.9	883.5	300.4
7.1630	20.471	4.094	14.648	1.231	0.499	1190.4	887.2	303.1
7.1730	20.621	4.126	14.760	1.235	0.500	1196.8	889.8	307.1
7.1830	20.773	4.158	14.869	1.239	0.507	1200.3	892.0	308.3
7.1930	20.920	4.186	14.982	1.241	0.510	1204.1	894.0	310.1
7.2030	21.069	4.217	15.091	1.246	0.515	1207.3	896.7	310.7
7.2130	21.219	4.251	15.197	1.251	0.519	1211.1	900.0	311.1
7.2230	21.367	4.288	15.306	1.266	0.508	1214.3	908.9	305.4
7.2330	21.526	4.329	15.419	1.271	0.507	1218.4	912.1	306.2
8.0030	21.685	4.369	15.534	1.277	0.505	1221.6	915.3	306.4
8.0130	21.841	4.411	15.648	1.282	0.501	1225.7	918.2	307.4
8.0230	21.992	4.447	15.756	1.286	0.503	1230.0	921.0	309.0
8.0330	22.143	4.482	15.861	1.290	0.508	1234.0	923.5	310.5
8.0430	22.293	4.521	15.976	1.296	0.500	1237.8	926.6	311.2
8.0530	22.444	4.560	16.092	1.302	0.490	1241.0	930.2	310.8
8.0630	22.595	4.601	16.205	1.316	0.474	1244.5	937.6	306.9
8.0730	22.746	4.643	16.318	1.325	0.459	1248.8	942.9	305.9
8.0830	22.896	4.686	16.432	1.331	0.447	1253.4	946.7	306.7
8.0930	23.047	4.731	16.546	1.335	0.434	1258.0	949.1	308.9

8.1030	23.198	4.776	16.665	1.339	0.418	1261.5	951.4	310.1
8.1130	23.349	4.820	16.786	1.343	0.400	1265.0	953.7	311.3
8.1230	23.500	4.863	16.907	1.347	0.383	1268.5	956.1	312.4
8.1330	23.650	4.906	17.028	1.374	0.343	1272.0	970.0	302.0
8.1430	23.800	4.949	17.142	1.377	0.331	1274.7	972.4	302.3
8.1530	23.951	4.992	17.255	1.381	0.322	1279.3	974.5	304.8
8.1630	24.102	5.038	17.368	1.384	0.313	1283.9	976.2	307.6
8.1730	24.245	5.083	17.467	1.386	0.308	1288.7	977.6	311.1
8.1830	24.380	5.122	17.567	1.389	0.303	1294.4	979.0	315.4
8.1930	24.513	5.157	17.671	1.391	0.295	1300.6	980.4	320.2
8.2030	24.646	5.192	17.776	1.394	0.284	1303.8	982.3	321.5
8.2130	24.777	5.228	17.882	1.398	0.270	1306.8	984.8	321.9
8.2230	24.912	5.264	17.982	1.402	0.264	1310.5	987.4	323.2
8.2330	25.045	5.299	18.087	1.436	0.223	1316.2	1005.6	310.6
9.0030	25.178	5.334	18.188	1.440	0.216	1320.2	1008.0	312.2
9.0130	25.312	5.373	18.287	1.443	0.208	1324.8	1010.3	314.5
9.0230	25.445	5.412	18.389	1.447	0.197	1328.0	1012.7	315.4
9.0330	25.579	5.450	18.487	1.451	0.191	1331.3	1015.0	316.3
9.0430	25.713	5.489	18.589	1.454	0.180	1334.0	1017.3	316.7
9.0530	25.847	5.535	18.689	1.458	0.165	1337.5	1019.6	317.9
9.0630	25.981	5.580	18.785	1.462	0.154	1341.2	1021.9	319.3
9.0730	26.116	5.626	18.878	1.466	0.145	1345.6	1024.2	321.3
9.0830	26.250	5.672	18.978	1.469	0.131	1350.1	1026.3	323.8
9.0930	26.385	5.712	19.056	1.472	0.145	1354.7	1028.2	326.5
9.1030	26.513	5.749	19.147	1.475	0.143	1359.0	1030.3	328.7
9.1130	26.642	5.786	19.234	1.478	0.144	1364.4	1032.5	331.9
9.1230	26.771	5.821	19.329	1.482	0.139	1369.8	1034.6	335.2
9.1330	26.902	5.854	19.431	1.485	0.132	1376.0	1036.8	339.2
9.1430	27.035	5.886	19.531	1.488	0.131	1381.7	1038.7	342.9
9.1530	27.167	5.916	19.632	1.490	0.129	1388.4	1040.4	348.0
9.1630	27.298	5.952	19.740	1.494	0.111	1393.8	1043.3	350.5
9.1730	27.436	5.990	19.843	1.500	0.103	1401.1	1047.4	353.6
9.1830	27.574	6.023	19.931	1.503	0.117	1408.9	1049.3	359.5
9.1930	27.698	6.051	20.018	1.506	0.123	1417.5	1051.3	366.2
9.2030	27.827	6.078	20.107	1.533	0.109	1424.2	1070.2	354.0
9.2130	27.957	6.118	20.199	1.534	0.107	1429.6	1071.0	358.6

APPENDIX C: TABLE VI.

RUN 6: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
9.2230	28.088	6.156	20.280	1.535	0.117	1437.7	1071.8	365.9
SHUT DOWN AT 9.2230 FOR 23 HOURS								
10.2130	28.222	6.194	20.338	1.536	0.154	1451.4	1072.6	378.9
10.2230	28.363	6.230	20.433	1.537	0.163	1467.1	1073.4	393.7
10.2330	28.504	6.263	20.542	1.540	0.159	1479.5	1075.6	403.9
11.0030	28.645	6.297	20.645	1.545	0.158	1490.0	1079.1	410.8
11.0130	28.784	6.328	20.727	1.550	0.179	1503.7	1082.7	421.0
11.0230	28.921	6.357	20.827	1.556	0.180	1519.9	1087.1	432.8
11.0330	29.057	6.383	20.929	1.563	0.183	1535.8	1091.8	443.9
11.0430	29.198	6.409	21.025	1.570	0.194	1548.2	1096.5	451.6
11.0530	29.336	6.435	21.125	1.594	0.183	1555.7	1114.0	441.7
11.0630	29.472	6.458	21.221	1.602	0.191	1573.5	1119.8	453.7
11.0730	29.607	6.479	21.315	1.625	0.188	1591.0	1136.0	455.0
11.0830	29.742	6.500	21.404	1.659	0.180	1606.9	1160.5	446.4
11.0930	29.878	6.520	21.491	1.689	0.179	1625.2	1181.9	443.4
11.1030	30.016	6.541	21.566	1.694	0.215	1642.5	1185.5	456.9
11.1130	30.153	6.564	21.656	1.717	0.215	1659.7	1202.6	457.1
11.1230	30.291	6.588	21.748	1.723	0.232	1677.0	1206.7	470.3
11.1330	30.428	6.610	21.820	1.742	0.255	1693.9	1221.0	473.0
11.1430	30.565	6.631	21.884	1.750	0.300	1712.5	1226.5	486.1
11.1530	30.702	6.650	21.942	1.759	0.350	1730.0	1233.1	496.9
11.1630	30.839	6.670	22.014	1.770	0.386	1747.8	1240.8	507.0
11.1730	30.976	6.689	22.091	1.781	0.415	1766.7	1249.6	517.0
11.1830	31.111	6.710	22.167	1.788	0.447	1782.8	1254.4	528.4
11.1930	31.248	6.733	22.269	1.794	0.451	1795.5	1259.2	536.3
11.2030	31.384	6.757	22.375	1.801	0.451	1795.5	1263.9	531.6
11.2130	31.518	6.782	22.474	1.807	0.456	1804.9	1268.6	536.3
11.2230	31.654	6.808	22.554	1.817	0.475	1817.3	1276.5	540.9

11.2330	31.788	6.830	22.613	1.839	0.507	1835.1	1292.6	542.5
12.0030	31.922	6.851	22.674	1.843	0.554	1848.0	1296.2	551.8
12.0130	32.057	6.870	22.742	1.869	0.575	1863.1	1315.2	547.9
12.0230	32.190	6.890	22.819	1.875	0.606	1870.4	1320.1	550.3
12.0330	32.324	6.914	22.898	1.890	0.622	1884.1	1331.6	552.5
12.0430	32.462	6.941	22.969	1.900	0.652	1902.2	1338.7	563.5
12.0530	32.599	6.966	23.027	1.910	0.697	1919.7	1346.6	573.2
12.0630	32.733	6.987	23.084	1.920	0.741	1934.0	1354.4	579.6
12.0730	32.868	7.006	23.153	1.928	0.781	1948.5	1360.2	588.4
12.0830	33.002	7.024	23.221	1.932	0.824	1965.2	1363.8	601.4
12.0930	33.137	7.043	23.271	1.937	0.886	1983.0	1367.4	615.6
12.1030	33.271	7.062	23.353	1.962	0.893	1994.9	1386.1	608.8
12.1130	33.406	7.083	23.421	1.990	0.912	2005.7	1406.6	599.1
12.1230	33.542	7.105	23.484	2.019	0.934	2023.4	1428.5	595.0
12.1330	33.677	7.127	23.553	2.039	0.958	2040.2	1443.3	596.9
12.1430	33.812	7.147	23.619	2.059	0.988	2057.1	1457.7	599.5
12.1530	33.948	7.166	23.687	2.073	1.022	2071.7	1468.0	603.6
12.1630	34.083	7.186	23.739	2.083	1.075	2085.7	1475.4	610.3
12.1730	34.217	7.207	23.804	2.096	1.110	2096.5	1484.8	611.6
12.1830	34.352	7.231	23.874	2.109	1.139	2109.7	1494.3	615.4
12.1930	34.487	7.256	23.938	2.118	1.175	2120.2	1501.5	618.6
12.2030	34.621	7.280	24.008	2.137	1.196	2130.4	1515.3	615.1
12.2130	34.757	7.304	24.071	2.147	1.234	2142.5	1522.5	620.1
12.2230	34.890	7.326	24.147	2.163	1.254	2152.2	1534.2	618.1
12.2330	35.023	7.348	24.224	2.172	1.280	2163.8	1540.8	623.0
13.0030	35.156	7.370	24.305	2.183	1.298	2176.5	1549.1	627.4
13.0130	35.290	7.396	24.361	2.196	1.338	2186.2	1559.1	627.1
13.0230	35.426	7.422	24.443	2.209	1.351	2200.7	1569.2	631.5
13.0330	35.560	7.446	24.513	2.222	1.378	2211.8	1579.4	632.4
13.0430	35.692	7.469	24.588	2.234	1.400	2224.4	1588.3	636.2
13.0530	35.825	7.493	24.662	2.244	1.427	2236.3	1595.8	640.5
13.0630	35.960	7.517	24.730	2.257	1.456	2248.4	1605.5	642.9
13.0730	36.091	7.539	24.793	2.273	1.487	2266.5	1617.5	649.0

APPENDIX C: TABLE VI.

RUN 6: SULPHUR AND STONE CUMULATIVE BALANCE. PAGE 5 OF 7

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
13.0830	36.228	7.560	24.853	2.289	1.525	2279.4	1629.3	650.1
13.0930	36.362	7.580	24.894	2.305	1.583	2294.8	1641.0	653.7
13.1030	36.497	7.603	24.947	2.320	1.625	2306.1	1652.0	654.1
13.1130	36.631	7.626	25.008	2.334	1.663	2319.6	1662.1	657.4
13.1230	36.766	7.647	25.073	2.349	1.697	2333.3	1672.8	660.5
13.1330	36.901	7.669	25.150	2.361	1.721	2346.8	1681.8	665.0
13.1430	37.035	7.692	25.258	2.384	1.702	2356.5	1697.7	658.8
13.1530	37.169	7.717	25.351	2.394	1.707	2363.2	1705.1	658.1
13.1630	37.304	7.746	25.447	2.401	1.710	2373.2	1710.4	662.8
13.1730	37.438	7.776	25.543	2.408	1.711	2382.6	1716.8	665.8
13.1830	37.573	7.807	25.630	2.416	1.720	2393.9	1723.1	670.8
13.1930	37.708	7.840	25.711	2.423	1.734	2405.2	1729.0	676.2
13.2030	37.839	7.880	25.794	2.432	1.733	2415.7	1735.9	679.9
13.2130	37.972	7.945	25.851	2.442	1.734	2425.7	1743.6	682.1
13.2230	38.109	8.008	25.897	2.452	1.752	2437.8	1751.0	686.9
13.2330	38.245	8.064	25.957	2.461	1.763	2448.9	1757.9	690.9
14.0030	38.382	8.114	26.019	2.470	1.779	2458.3	1765.2	693.1
14.0130	38.520	8.165	26.091	2.480	1.784	2468.3	1772.7	695.6
14.0230	38.659	8.218	26.153	2.489	1.799	2477.4	1780.2	697.2
14.0330	38.796	8.271	26.210	2.499	1.816	2483.6	1787.7	695.9
14.0430	38.934	8.323	26.276	2.508	1.827	2492.8	1795.2	697.6
14.0530	39.074	8.373	26.335	2.517	1.848	2503.6	1802.7	700.9
14.0630	39.213	8.430	26.400	2.527	1.856	2513.5	1810.6	702.9
14.0730	39.351	8.487	26.456	2.537	1.871	2525.4	1818.9	706.5
14.0830	39.489	8.540	26.512	2.547	1.890	2538.1	1827.2	710.8
14.0930	39.621	8.589	26.565	2.557	1.910	2545.9	1835.5	710.3
14.1030	39.761	8.637	26.614	2.567	1.944	2555.0	1843.8	711.2
14.1130	39.901	8.683	26.686	2.573	1.958	2563.9	1849.3	714.6

14.1230	40.041	8.729	26.763	2.577	1.972	2571.7	1851.9	719.8
14.1330	40.180	8.775	26.841	2.580	1.984	2580.4	1854.5	725.8
14.1430	40.320	8.820	26.913	2.583	2.005	2588.7	1857.1	731.6
14.1530	40.461	8.867	26.988	2.603	2.002	2596.3	1875.7	720.6
14.1630	40.601	8.915	27.056	2.623	2.007	2603.3	1893.4	709.9
14.1730	40.741	8.962	27.132	2.630	2.018	2613.2	1899.6	713.6
14.1830	40.882	9.011	27.201	2.637	2.033	2625.6	1905.8	719.9
14.1930	41.023	9.059	27.271	2.644	2.049	2635.3	1911.9	723.4
14.2030	41.163	9.107	27.341	2.652	2.064	2644.2	1918.1	726.1
14.2130	41.304	9.153	27.392	2.659	2.101	2654.2	1924.2	729.9
14.2230	41.443	9.198	27.472	2.666	2.107	2663.1	1930.4	732.7
14.2330	41.582	9.243	27.551	2.676	2.113	2669.8	1939.1	730.7
15.0030	41.724	9.295	27.612	2.689	2.129	2679.2	1950.4	728.8
15.0130	41.866	9.350	27.684	2.702	2.130	2690.6	1961.7	728.9
15.0230	42.009	9.402	27.756	2.715	2.137	2700.5	1973.4	727.1
15.0330	42.155	9.453	27.827	2.728	2.148	2710.8	1985.6	725.2
15.0430	42.304	9.494	27.921	2.741	2.148	2720.5	1997.8	722.7
15.0530	42.455	9.533	28.021	2.754	2.148	2728.6	2009.9	718.6
15.0630	42.603	9.568	28.123	2.759	2.153	2736.4	2015.1	721.3
15.0730	42.752	9.598	28.221	2.766	2.167	2746.3	2020.8	725.5
15.0830	42.901	9.628	28.325	2.774	2.174	2756.0	2028.5	727.6
15.0930	43.049	9.661	28.411	2.784	2.194	2764.7	2037.3	727.3
15.1030	43.193	9.694	28.504	2.792	2.203	2774.4	2044.7	729.7
15.1130	43.333	9.722	28.594	2.799	2.219	2784.3	2050.4	733.9
15.1230	43.473	9.748	28.684	2.805	2.236	2792.7	2056.2	736.5
15.1330	43.612	9.773	28.769	2.812	2.258	2802.1	2062.0	740.1
15.1430	43.749	9.799	28.859	2.819	2.272	2809.9	2068.0	741.9
15.1530	43.886	9.824	28.937	2.826	2.299	2818.0	2074.2	743.8
15.1630	44.025	9.850	29.021	2.833	2.321	2825.3	2080.4	744.9
15.1730	44.163	9.875	29.108	2.840	2.340	2838.2	2086.6	751.6
15.1830	44.301	9.899	29.190	2.847	2.364	2849.8	2092.4	757.4
15.1930	44.440	9.922	29.272	2.854	2.392	2860.3	2097.8	762.5
15.2030	44.579	9.948	29.361	2.860	2.409	2871.4	2103.2	768.1
15.2130	44.716	9.976	29.444	2.867	2.429	2881.3	2108.6	772.7
15.2230	44.852	10.002	29.522	2.874	2.453	2892.4	2114.7	777.7
15.2330	44.989	10.029	29.604	2.882	2.474	2902.3	2121.4	780.9

APPENDIX C: TABLE VI.

RUN 6: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
16.0030	45.126	10.056	29.687	2.890	2.493	2912.8	2128.1	784.8
16.0130	45.263	10.084	29.774	2.898	2.507	2922.5	2134.6	787.9
16.0230	45.402	10.112	29.856	2.906	2.529	2933.3	2141.0	792.4
16.0330	45.535	10.140	29.933	2.913	2.548	2942.8	2147.3	795.4
16.0430	45.665	10.167	30.012	2.921	2.565	2949.2	2154.2	795.0
16.0530	45.800	10.197	30.094	2.930	2.579	2956.5	2161.5	795.0
16.0630	45.934	10.227	30.176	2.939	2.592	2967.8	2168.8	799.0
16.0730	46.068	10.257	30.259	2.948	2.605	2976.2	2176.2	799.9
16.0830	46.202	10.286	30.342	2.957	2.616	2982.9	2183.8	799.1
16.0930	46.335	10.317	30.425	2.967	2.626	2989.4	2191.3	798.0
16.1030	46.469	10.350	30.503	2.976	2.641	2997.7	2198.8	799.0
16.1130	46.600	10.384	30.580	2.985	2.651	3006.6	2206.1	800.5
16.1230	46.732	10.419	30.653	3.027	2.633	3016.8	2242.2	774.6
16.1330	46.863	10.454	30.724	3.037	2.649	3030.1	2249.5	780.6
16.1430	46.994	10.490	30.802	3.077	2.625	3041.4	2284.4	757.0
16.1530	47.124	10.524	30.869	3.104	2.628	3052.7	2307.4	745.3
16.1630	47.256	10.557	30.942	3.116	2.641	3067.0	2316.6	750.4
16.1730	47.386	10.591	31.015	3.142	2.638	3073.4	2338.4	735.0
16.1830	47.517	10.626	31.100	3.167	2.624	3082.6	2359.5	723.1
16.1930	47.648	10.663	31.207	3.175	2.603	3096.3	2365.5	730.8
16.2030	47.780	10.701	31.280	3.183	2.617	3109.3	2370.8	738.5
16.2130	47.911	10.737	31.359	3.189	2.626	3124.6	2375.4	749.3
16.2230	48.042	10.771	31.449	3.196	2.626	3136.7	2380.0	756.8
16.2330	48.172	10.806	31.530	3.203	2.633	3147.0	2385.3	761.7
17.0030	48.303	10.843	31.607	3.211	2.642	3157.2	2391.5	765.7
17.0130	48.434	10.882	31.687	3.218	2.647	3166.4	2397.7	768.7
17.0230	48.566	10.922	31.758	3.228	2.658	3174.2	2405.7	768.5
17.0330	48.697	10.960	31.833	3.241	2.663	3188.5	2415.7	772.8

17.0430	48.830	10.998	31.919	3.253	2.660	3201.9	2425.6	776.4
17.0530	48.962	11.034	31.996	3.266	2.666	3214.3	2436.2	778.1
17.0630	49.095	11.070	32.073	3.280	2.672	3225.4	2447.6	777.8
17.0730	49.227	11.101	32.156	3.293	2.678	3234.3	2458.3	776.0
17.0830	49.360	11.131	32.240	3.307	2.682	3248.8	2469.1	779.7
17.0930	49.490	11.161	32.328	3.322	2.679	3259.9	2480.7	779.2
17.1030	49.620	11.191	32.391	3.727	2.312	3269.3	2742.5	526.8
17.1130	49.751	11.222	32.475	3.737	2.318	3280.1	2749.0	531.1

SHUT DOWN AT 17.1130 FOR 40 HOURS

19.0330	49.882	11.233	32.527	3.747	2.376	3290.3	2755.5	534.8
19.0430	50.013	11.249	32.603	3.754	2.407	3303.5	2760.1	543.4
19.0530	50.145	11.270	32.695	3.758	2.422	3311.3	2762.8	548.6
19.0630	50.277	11.295	32.785	3.762	2.434	3320.8	2765.6	555.1
19.0730	50.408	11.323	32.874	3.767	2.444	3330.2	2768.7	561.5
19.0830	50.540	11.356	32.961	3.772	2.452	3338.6	2771.9	566.7
19.0930	50.673	11.390	33.045	3.777	2.461	3346.9	2775.0	571.9
19.1030	50.806	11.426	33.132	3.781	2.467	3356.6	2777.7	578.9
19.1130	50.941	11.461	33.223	3.784	2.472	3366.0	2779.8	586.3
19.1230	51.076	11.494	33.314	3.788	2.480	3375.5	2782.3	593.2
19.1330	51.212	11.528	33.409	3.792	2.483	3385.7	2785.3	600.4
19.1430	51.349	11.562	33.513	3.796	2.478	3395.9	2788.4	607.5
19.1530	51.483	11.592	33.621	3.801	2.469	3404.6	2791.7	612.8
19.1630	51.614	11.638	33.712	3.805	2.459	3413.7	2795.3	618.5
19.1730	51.749	11.665	33.797	3.811	2.477	3423.4	2799.1	624.4
19.1830	51.882	11.689	33.895	3.816	2.483	3431.5	2802.9	628.6
19.1930	52.015	11.714	33.983	3.821	2.497	3439.1	2806.8	632.3
19.2030	52.148	11.743	34.009	3.826	2.570	3449.8	2810.3	639.5
19.2130	52.281	11.768	34.107	3.830	2.576	3460.1	2813.5	646.6
19.2230	52.416	11.797	34.195	3.836	2.587	3468.4	2817.3	651.1
19.2330	52.549	11.831	34.290	3.842	2.586	3478.1	2821.8	656.3
20.0030	52.681	11.860	34.385	3.847	2.589	3485.9	2825.8	660.1
20.0130	52.812	11.889	34.494	3.852	2.577	3493.5	2829.4	664.1
20.0230	52.946	11.920	34.600	3.858	2.568	3500.2	2833.8	666.4

APPENDIX C: TABLE VI.
RUN 6: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
20.0330	53.079	11.948	34.695	3.865	2.571	3506.9	2838.9	668.0
20.0430	53.211	11.978	34.801	3.872	2.560	3514.8	2844.5	670.3
20.0530	53.344	12.008	34.911	3.880	2.545	3523.7	2850.4	673.2
20.0630	53.477	12.042	35.016	3.888	2.531	3530.4	2856.1	674.3
20.0730	53.609	12.076	35.117	3.895	2.521	3536.9	2861.4	675.5
20.0830	53.741	12.107	35.214	3.901	2.518	3549.5	2866.3	683.3
20.0930	53.873	12.137	35.297	3.908	2.532	3558.1	2870.7	687.4
20.1030	54.006	12.169	35.391	3.914	2.533	3564.6	2875.2	689.4
20.1130	54.138	12.201	35.497	3.920	2.521	3571.1	2879.4	691.6
20.1230	54.271	12.243	35.600	3.926	2.502	3580.2	2883.7	696.6
20.1330	-	-	-	-	-	3588.9	2889.3	699.5
20.1430	-	-	-	-	-	3594.0	2895.0	699.0
20.1530	-	-	-	-	-	3598.8	2898.7	700.1
20.1630	54.404	12.274	35.682	3.931	2.517	3608.5	2902.4	706.1
20.1730	54.538	12.303	35.798	3.937	2.501	3620.6	2906.1	714.6
20.1830	54.672	12.329	35.906	3.945	2.492	3633.0	2912.3	720.8
20.1930	54.806	12.355	36.007	3.953	2.490	3647.1	2918.4	728.6
20.2030	54.941	12.382	36.090	3.961	2.508	3657.0	2924.0	733.0
20.2130	55.079	12.409	36.186	3.969	2.516	3665.6	2929.6	736.0
20.2230	55.213	12.434	36.269	3.976	2.534	3677.0	2934.8	742.1
20.2330	55.347	12.462	36.349	3.983	2.553	3688.0	2939.6	748.4
21.0030	55.482	12.489	36.455	3.990	2.548	3695.8	2944.4	751.4
21.0130	55.618	12.517	36.540	4.001	2.559	3707.9	2952.6	755.4
21.0230	55.753	12.545	36.637	4.011	2.560	3721.1	2959.5	761.7
21.0330	55.888	12.574	36.714	4.018	2.582	3729.2	2965.2	764.1
21.0430	56.022	12.602	36.825	4.030	2.566	3740.3	2973.9	766.3
			STONE CHANGE					
21.0530	56.157	12.630	36.937	4.046	2.545	3748.0	2985.8	762.2

21.0630	56.291	12.661	37.036	4.057	2.536	3748.0	2994.2	753.8
21.0730	56.424	12.696	37.121	4.073	2.533	3748.0	3007.8	740.2
21.0830	56.562	12.735	37.197	4.080	2.550	3748.0	3012.8	735.2
21.0930	56.696	12.773	37.274	4.093	2.557	3755.4	3022.9	732.6
21.1030	56.830	12.808	37.358	4.105	2.560	3763.6	3032.2	731.4
21.1130	56.965	12.843	37.434	4.114	2.575	3771.9	3038.3	733.6
21.1230	57.101	12.877	37.508	4.122	2.594	3778.8	3043.9	734.9
21.1330	57.235	12.909	37.569	4.130	2.627	3789.3	3049.1	740.2
21.1430	57.371	12.943	37.648	4.138	2.641	3799.3	3054.3	745.0
21.1530	57.507	12.979	37.723	4.147	2.658	3803.7	3060.4	743.3
21.1630	57.641	13.014	37.797	4.156	2.674	3812.2	3066.5	745.7
21.1730	57.777	13.050	37.872	4.165	2.690	3821.4	3072.5	748.9
21.1830	57.911	13.086	37.950	4.174	2.700	3831.7	3078.4	753.2
21.1930	58.044	13.123	38.012	4.208	2.700	3841.4	3107.9	733.5
21.2030	58.178	13.160	38.086	4.221	2.711	3855.0	3116.7	738.4
21.2130	58.313	13.193	38.154	4.232	2.735	3863.5	3123.4	740.1
21.2230	58.448	13.227	38.220	4.242	2.759	3873.5	3130.2	743.3
21.2330	58.583	13.263	38.298	4.251	2.771	3882.8	3136.1	746.7
22.0030	58.717	13.300	38.364	4.271	2.782	3891.8	3152.8	738.9
22.0130	58.852	13.342	38.427	4.280	2.802	3900.5	3158.6	741.8
22.0230	58.986	13.384	38.502	4.293	2.807	3910.5	3168.5	742.0
22.0330	59.120	13.423	38.566	4.303	2.828	3920.2	3175.2	745.1
22.0430	59.253	13.461	38.630	4.317	2.845	3929.5	3185.7	743.8
22.0530	59.388	13.493	38.693	4.326	2.877	3937.7	3191.5	746.2
22.0630	59.522	13.530	38.758	4.342	2.892	3945.4	3204.5	740.9
22.0730	59.655	13.561	38.813	4.349	2.932	3953.9	3209.2	744.7
22.0830	59.789	13.599	38.885	4.356	2.950	3962.6	3213.7	748.9
22.0930	59.924	13.638	38.961	4.363	2.962	3970.1	3218.7	751.4
22.1030	60.058	13.676	39.015	4.371	2.997	3977.0	3223.7	753.3
22.1130	60.193	13.714	39.082	4.374	3.023	3979.6	3225.8	753.7
22.1230	60.327	13.752	39.140	4.377	3.057	3979.6	3228.0	751.6
22.1330	60.461	13.797	39.208	4.382	3.075	3979.6	3231.2	748.4
22.1430	60.596	13.842	39.273	4.387	3.094	3979.6	3234.4	745.2
22.1530	60.731	13.889	39.339	4.393	3.110	3985.2	3238.1	747.1
22.1630	60.865	13.937	39.401	4.398	3.128	3992.9	3241.8	751.1
22.1730	60.997	13.984	39.448	4.406	3.159	4001.9	3247.1	754.7

APPENDIX C - TABLE VII

ANALYSIS OF SOLIDS REMOVED DURING RUN 6
T O T A L   S U L P H U R   WT. PERCENT

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
2.0730	4.73	3.55	-	-	-	-	-
2.1600	5.44	-	-	-	-	-	-
2.2000	4.93	3.42	6.30	3.70	-	3.83	3.01
3.0130	5.05	5.47	5.41	2.99	-	3.92	3.35
3.1300	4.93	5.11	4.24	3.16	4.11	3.74	3.55
3.2000	4.55	2.91	4.24	1.97	-	3.82	3.78
4.0200	4.17	2.56	3.79	2.26	-	3.80	3.35
4.1400	-	3.40	2.76	1.06	-	4.14	3.46
5.1030	-	3.87	2.40	1.97	4.00	3.87	3.44
5.1800	4.10	2.74	4.19	3.17	3.73	3.66	3.80
6.0000	4.15	2.75	4.06	1.72	1.29	3.87	3.63
6.1900	4.10	3.36	4.52	2.74	3.08	4.04	3.45
7.0800	5.77	3.29	4.06	3.09	4.73	3.20	4.31
7.2000	5.81	4.96	4.69	1.85	5.63	3.57	3.14
8.0000	6.15	4.29	4.02	3.07	5.39	5.39	3.61
8.0800	6.35	4.18	3.57	1.10	5.34	5.33	4.28
8.1600	-	4.05	4.62	2.52	5.30	4.71	4.46
9.0400	5.67	3.90	4.04	2.26	-	4.54	3.81
9.1500	4.71	3.42	4.39	3.22	4.37	3.92	3.96
12.0500	4.16	3.50	4.04	2.67	3.84	3.58	3.88
12.1530	4.42	2.80	4.20	3.06	3.35	3.25	3.22
13.0300	4.51	3.24	4.00	3.93	3.46	4.14	3.89
13.1700	4.84	2.70	4.48	3.80	3.25	3.95	4.07
14.0500	3.27	2.39	3.43	3.15	4.04	4.55	3.22
14.1500	3.32	2.34	3.42	3.97	3.02	4.23	3.49
15.1800	4.35	2.96	-	3.41	3.32	2.67	3.53
16.0100	4.43	-	4.07	3.59	2.99	3.21	3.59
16.1000	3.85	-	4.01	3.35	3.46	3.50	3.78
16.1300	3.35	4.57	4.57	4.19	-	4.48	3.78

APPENDIX C - TABLE VIII

ANALYSIS OF SOLIDS REMOVED DURING RUN 6

SULPHATE SULPHUR WT. PERCENT

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
2.0730	-	-	-	-	-	-	-
2.1600	0.27	-	-	-	-	-	-
2.2000	0.21	1.03	1.29	0.58	-	3.20	0.19
3.0130	0.13	0.89	1.31	0.49	-	3.48	0.16
3.1300	0.14	0.90	1.15	0.33	2.28	2.97	0.11
3.2000	0.24	0.95	1.31	0.31	-	3.08	0.20
4.0200	0.11	0.91	1.25	0.29	-	3.21	0.18
4.1400	-	1.02	1.47	0.13	-	2.42	0.13
5.1030	-	0.96	1.19	0.32	0.72	3.06	0.21
5.1800	0.11	0.90	1.89	0.37	0.84	2.88	0.14
6.0000	0.18	0.83	1.67	0.48	0.81	3.48	0.16
6.1900	0.40	1.07	1.92	0.14	0.77	3.59	0.21
7.0800	0.30	0.64	1.08	0.34	1.63	3.00	0.15
7.2000	0.35	0.91	1.03	0.23	1.26	3.07	0.16
8.0000	0.32	0.44	1.03	0.18	1.38	4.66	0.17
8.0800	0.21	0.62	1.20	0.30	0.94	4.21	0.14
8.1630	-	0.87	1.38	0.38	1.59	3.98	0.22
9.0400	0.41	0.89	1.96	0.28	-	3.65	0.13
9.1500	0.15	1.06	2.09	0.32	1.28	3.30	0.18
12.0500	0.31	1.24	1.37	0.38	1.58	3.00	0.20
12.1530	0.35	1.04	1.29	0.39	1.42	2.87	0.14
13.0200	0.34	1.27	1.29	0.53	0.95	3.88	0.20
13.1700	0.41	1.04	1.45	0.62	0.79	3.63	0.22
14.0500	0.19	1.67	1.68	0.34	3.08	3.77	0.15
14.1500	0.50	1.48	1.38	0.23	1.57	3.59	0.28
15.1800	0.17	1.27	-	0.42	1.40	2.38	0.19
16.0100	0.10	-	1.35	0.42	1.23	3.14	0.16
16.1000	0.18	-	1.33	0.38	1.43	3.23	0.12
16.1300	0.12	1.90	1.90	0.59	-	4.03	-

APPENDIX C - TABLE IX
ANALYSIS OF SOLIDS REMOVED DURING RUN 6
TOTAL CARBON WT. PERCENT

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
2.0730	-	-	-	-	-	-	-
2.1600	0.19	-	-	-	-	-	-
2.2000	0.18	0.	2.06	25.00	-	0.70	4.85
3.0130	0.23	0.04	1.36	28.50	-	4.99	0.99
3.1300	0.11	0.04	0.36	38.90	0.14	13.00	2.85
3.2000	0.25	0.05	2.86	35.00	-	13.00	0.59
4.0200	0.09	0.	1.24	33.10	-	11.00	3.80
4.1400	-	0.02	5.38	45.00	-	29.90	0.78
5.1030	-	0.05	0.63	35.10	0.45	17.70	2.73
5.1800	0.20	0.15	0.93	29.20	0.51	11.30	1.19
6.0000	0.23	0.04	0.94	30.40	0.77	6.37	2.04
6.1900	0.08	0.04	0.52	21.10	0.34	6.04	3.46
7.0800	1.65	0.16	0.61	24.50	0.20	3.14	0.55
7.2000	0.14	0.	2.85	33.70	0.08	8.84	8.37
8.0000	0.32	0.06	3.67	41.13	0.25	6.49	3.23
8.0800	0.37	0.	1.59	31.60	0.32	10.80	1.94
8.1630	-	0.07	2.60	42.70	0.48	17.90	1.42
9.0400	0.33	0.	2.11	46.60	-	14.20	2.17
9.1530	0.68	0.10	5.57	36.20	0.58	10.80	7.75
12.0500	0.13	0.	0.19	17.70	0.08	5.78	2.34
12.1530	0.09	0.08	0.42	21.00	0.14	6.70	3.74
13.0200	0.	0.	0.09	20.10	0.28	3.25	2.16
13.1700	0.08	0.	0.62	16.50	0.16	4.98	0.86
14.0500	0.	0.	0.02	11.20	0.	3.32	0.33
14.1500	0.	0.	0.02	11.50	0.	3.44	0.72
15.1800	0.11	0.	-	7.77	0.04	1.65	0.90
16.0100	0.10	-	0.23	10.10	0.14	2.96	2.91
16.1000	0.11	-	0.41	9.85	0.15	3.55	0.29
16.1300	0.09	-	0.41	16.00	-	2.39	-

APPENDIX C - TABLE X

SOLIDS REMOVED DURING RUN 6, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
0.2107	-	-	-	-	-	4.08	-
0.2352	-	-	0.68	-	-	-	-
1.0118	-	-	-	9.53	-	-	-
1.0240	-	-	0.91	1.36	-	-	-
1.0500	-	-	1.81	6.80	-	8.62	-
1.0655	-	-	1.81	1.36	-	4.54	-
1.0930	-	-	1.59	8.16	-	3.40	-
1.1530	-	-	3.86	0.23	-	20.41	-
1.1800	-	-	1.81	-	22.68	9.53	-
1.1845	-	-	1.70	-	-	1.47	-
1.2000	-	-	1.59	0.23	0.11	8.85	-
2.0440	-	-	3.63	0.68	0.11	7.26	-
2.0835	-	5.90	6.12	0.23	0.11	5.90	-
2.0900	7.94	-	-	-	-	-	-
2.1200	6.12	-	5.67	0.91	-	4.54	-
2.1315	16.78	-	-	-	-	-	-
2.1545	16.33	-	5.90	2.27	-	4.76	-
2.1700	16.33	-	-	-	-	-	-
2.1945	-	-	7.71	0.68	-	2.27	-
2.2235	14.97	-	5.90	0.23	-	5.56	-
3.0200	-	-	8.85	0.23	0.11	2.49	-
3.0450	-	-	-	-	-	-	-
3.0545	11.79	-	-	-	-	-	-
3.0745	-	-	9.75	-	-	-	-
3.0800	-	-	-	2.27	-	7.94	-
3.0930	16.33	-	9.98	2.49	-	5.44	-
3.1615	-	-	8.62	1.13	-	2.72	-
3.1750	15.42	-	-	-	-	-	-
3.1800	13.15	-	-	-	-	-	-
3.2000	-	-	9.53	0.91	-	2.72	-
3.2300	13.61	-	-	-	-	-	-
3.2330	-	-	9.75	0.91	0.11	3.63	-
4.0445	-	-	10.89	1.13	0.11	4.99	-
4.0730	-	-	4.08	0.45	-	2.72	-
4.1030	-	-	3.40	1.36	-	4.08	-
4.1400	-	-	1.81	0.45	0.23	2.27	-
4.1740	15.42	-	6.80	1.81	-	3.63	-
4.1945	12.70	-	-	-	-	-	-
4.2315	-	-	5.44	0.34	-	3.29	-
5.0100	15.42	-	-	-	-	-	-

SOLIDS REMOVED DURING RUN 6, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
5.0300	-	-	4.76	1.13	1.25	2.95	-
5.0745	-	-	9.07	21.09	0.68	4.76	-
5.1200	-	7.71	13.15	0.34	0.57	1.81	-
5.1445	1.81	-	2.04	0.23	1.81	6.80	-
5.1800	14.29	-	10.21	0.45	-	-	-
5.2000	23.59	-	1.81	0.45	1.13	0.68	-
5.2145	10.43	-	-	-	-	-	-
5.2200	-	-	3.97	0.68	0.68	1.59	-
6.0000	-	-	3.86	0.11	1.36	2.27	-
6.0200	-	-	6.24	0.45	-	-	-
6.0230	10.66	-	-	-	-	-	-
6.0400	-	-	7.71	0.45	2.49	3.18	-
6.0445	9.07	-	-	-	-	-	-
6.0600	-	-	6.92	0.57	1.47	1.93	-
6.0800	-	-	6.46	0.45	1.59	2.15	-
6.1000	-	-	7.03	0.34	2.72	2.95	-
6.1200	9.53	-	8.62	0.34	1.81	1.81	-
6.1400	9.75	-	6.35	0.34	1.81	2.72	-
6.1450	12.70	-	-	-	-	-	-
6.1500	7.48	-	-	-	-	-	-
6.1530	10.43	-	-	-	-	-	-
6.1600	7.71	-	10.43	0.45	1.36	1.81	-
6.1630	16.33	-	-	-	-	-	-
6.1800	-	-	5.22	0.23	1.36	2.04	-
6.2000	-	-	4.08	0.45	2.72	3.29	-
6.2200	-	-	4.65	0.68	4.54	2.38	-
7.0000	11.11	-	10.43	0.79	5.22	3.29	-
7.0100	8.28	-	-	-	-	-	-
7.0200	5.22	-	8.05	0.23	3.18	1.59	-
7.0300	5.44	-	-	-	-	-	-
7.0415	4.20	-	6.92	0.68	2.72	2.27	-
7.0500	6.01	-	-	-	-	-	-
7.0600	4.42	-	7.03	0.45	2.83	1.70	-
7.0700	4.20	-	-	-	-	-	-
7.0800	4.88	3.18	7.37	0.68	2.49	1.59	-
7.1000	-	-	7.03	0.45	1.59	2.04	-
7.1130	7.37	-	-	-	-	-	-
7.1200	-	-	5.67	1.02	1.36	1.47	-
7.1400	-	-	3.18	0.45	1.93	1.36	-
7.1430	21.77	-	-	-	-	-	-

SOLIDS REMOVED DURING RUN 6, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
7.1600	-	-	5.44	1.36	1.59	2.72	-
7.1800	-	-	2.72	0.45	1.13	1.36	-
7.2000	-	-	1.36	1.13	1.36	0.79	-
7.2200	-	-	2.61	1.02	2.61	1.47	-
8.0000	2.61	3.18	2.49	0.91	2.72	1.36	-
8.0200	-	-	1.36	0.34	3.18	1.81	-
8.0400	-	-	1.59	0.68	2.27	1.25	-
8.0600	-	-	1.59	0.34	3.29	3.18	-
8.0800	2.95	-	1.47	0.23	7.71	2.27	-
8.1000	-	-	1.36	-	2.83	0.91	-
8.1200	-	-	2.04	0.23	1.81	1.13	-
8.1400	-	-	3.18	0.23	1.47	0.91	-
8.1600	11.79	-	2.61	0.23	1.11	0.91	-
8.2000	-	-	2.72	0.45	1.59	1.70	-
9.0000	-	-	4.54	0.91	-	2.27	-
9.0100	16.33	-	-	-	-	-	-
9.0400	-	-	3.63	0.91	-	2.27	-
9.0800	-	-	1.36	0.68	-	4.99	-
9.1000	-	-	1.13	0.45	14.06	0.79	-
9.1400	-	-	4.54	0.91	3.63	0.91	-
9.1600	-	-	0.45	0.45	2.15	0.68	-
9.1730	-	-	3.97	0.11	2.04	-	-
9.2000	-	-	2.27	0.34	1.47	2.04	-
10.2300	-	-	10.43	1.59	-	12.70	-
11.0145	-	-	7.26	1.13	0.45	2.72	-
11.4150	12.70	-	-	-	-	-	-
11.0600	-	-	16.78	1.13	-	4.08	-
11.0700	13.15	-	-	-	-	-	-
11.0915	9.75	-	20.87	0.68	-	-	-
11.1015	18.14	-	-	-	-	3.18	-
11.1100	15.88	-	-	-	-	-	-
11.1200	-	-	8.28	0.34	-	1.81	-
11.1300	15.88	-	-	-	-	-	-
11.1400	13.61	-	6.80	0.45	-	2.27	-
11.1600	10.21	-	11.79	0.23	-	2.27	-
11.1730	-	-	11.34	0.45	-	2.49	-
11.2000	-	-	4.54	0.23	1.36	8.62	-
11.2200	-	-	6.58	0.91	0.91	1.81	-
11.2300	-	-	9.07	0.68	0.45	1.81	-
12.0100	9.07	-	-	-	-	-	-

SOLIDS REMOVED DURING RUN 6, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
12.0200	-	-	8.62	0.68	0.45	1.93	-
12.0400	15.88	-	5.90	0.91	2.49	4.31	-
12.0600	5.44	-	-	-	-	-	-
12.0700	-	-	18.14	1.36	2.72	3.18	-
12.1000	-	-	7.62	0.45	1.36	2.27	-
12.1200	9.07	-	27.22	0.91	2.27	4.08	-
12.1300	7.26	-	-	-	-	-	-
12.1400	4.54	-	-	-	-	-	-
12.1430	-	-	26.08	0.91	2.04	2.27	-
12.1500	8.16	-	-	-	-	-	-
12.1600	3.18	-	-	-	-	-	-
12.1700	2.72	-	16.33	0.45	0.91	1.81	-
12.1800	3.18	-	-	-	-	-	-
12.1900	2.27	-	-	-	-	-	-
12.2000	2.27	-	17.69	0.68	1.36	3.63	-
12.2200	6.80	-	12.70	0.45	0.45	1.59	-
13.0001	4.99	-	11.34	0.45	0.68	1.81	-
13.0200	-	-	9.07	0.91	5.22	6.35	-
13.0400	-	-	13.61	0.45	4.54	3.18	-
13.0600	-	-	9.98	0.45	3.27	2.36	-
13.0800	-	-	20.87	0.45	2.27	1.81	-
13.1000	-	-	21.21	0.45	1.02	2.27	-
13.1200	-	-	16.78	0.34	2.72	1.81	-
13.1300	-	-	-	-	4.31	2.72	-
13.1400	-	-	9.53	0.45	1.36	0.91	-
13.1600	-	-	2.72	0.45	4.54	2.72	-
13.1645	10.43	-	0.45	-	3.18	-	-
13.1800	2.72	-	1.13	0.45	6.35	1.81	-
13.2000	-	-	4.54	0.45	5.90	1.81	-
13.2200	-	-	9.07	0.23	5.44	1.81	-
14.0000	-	-	8.62	0.45	4.08	1.81	-
14.0200	-	-	9.98	-	4.08	1.81	-
14.0400	-	-	9.98	0.45	4.08	1.81	-
14.0600	-	-	10.21	0.91	3.63	1.36	-
14.0800	-	-	10.21	-	4.54	2.04	-
14.1100	-	-	19.96	2.72	-	3.18	-
14.1430	-	-	4.08	0.91	4.08	2.72	-
14.1800	12.70	-	12.70	0.91	5.90	3.63	-
14.1830	11.79	-	-	-	-	-	-
14.2300	-	-	19.96	0.79	6.12	5.90	-

SOLIDS REMOVED DURING RUN 6, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
15.0200	-	-	14.97	0.34	14.51	6.12	-
15.0530	-	-	2.49	-	35.15	7.26	-
15.0700	-	-	-	-	4.54	-	-
15.0800	-	-	3.52	0.23	4.31	2.49	-
15.1000	-	-	2.27	0.45	7.71	8.62	-
15.1400	-	-	10.89	0.45	7.94	5.22	-
15.1800	-	-	13.61	1.81	6.80	4.08	-
15.2200	-	-	14.29	0.68	4.76	3.18	-
16.0100	-	-	11.34	0.45	5.67	3.86	-
16.0400	-	-	10.43	0.68	6.35	2.83	-
16.0700	-	-	10.43	0.57	9.53	2.72	-
16.1000	-	-	9.98	0.68	8.16	5.44	-
16.1400	29.48	-	12.70	0.91	12.70	4.99	-
16.1600	27.22	-	-	-	-	-	-
16.1700	14.06	-	12.70	0.68	12.25	4.08	-
16.1730	14.06	-	-	-	-	-	-
16.1900	14.51	-	-	-	-	-	-
16.2000	14.51	-	10.89	0.68	5.90	1.81	-
16.2040	15.42	-	-	-	-	-	-
16.2300	-	-	7.26	0.57	4.76	2.27	-
17.0200	-	-	3.18	-	13.15	3.40	-
17.0500	-	-	2.83	0.11	24.49	4.31	-
17.0700	-	-	-	-	18.60	-	-
17.0800	-	-	3.63	0.11	7.94	4.54	-
17.1000	-	-	2.72	0.11	14.74	7.48	-
17.1200	-	-	1.81	0.11	14.29	3.18	-
18.0600	-	-	2.83	-	-	-	-
18.0800	-	-	0.45	-	11.57	8.39	-
18.1000	-	-	-	-	-	1.13	-
18.1800	-	-	-	-	-	174.18	-
18.2145	-	-	-	-	-	65.77	-
19.0030	-	-	1.36	0.45	0.68	1.81	-
19.0400	-	-	13.61	0.45	0.45	5.44	-
19.0600	-	-	0.45	0.23	-	4.54	-
19.0800	-	-	-	-	-	5.90	-
19.1000	-	-	0.23	-	-	6.12	-
19.1200	-	-	0.11	0.45	3.29	3.63	-
19.1400	-	-	-	-	2.95	3.40	-
19.1600	-	-	0.23	0.23	3.40	3.63	-
19.1800	-	-	-	-	4.31	3.63	-

SOLIDS REMOVED DURING RUN 6, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	ELUTR FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
19.2000	-	-	0.23	0.45	4.54	3.63	-
19.2200	-	-	1.36	0.23	1.81	3.63	-
20.0000	-	-	1.25	0.45	4.54	3.52	-
20.0200	-	-	1.70	0.23	3.74	2.15	-
20.0400	-	-	1.59	0.23	5.78	3.52	-
20.0600	-	-	1.59	0.23	7.26	3.86	-
20.0800	-	-	2.27	0.23	6.24	2.72	-
20.1030	-	-	3.63	0.34	4.54	3.63	-
20.1230	-	-	3.40	0.11	3.18	2.49	-
20.1430	-	-	7.48	0.23	2.27	2.27	-
20.1730	-	-	6.80	0.23	2.49	2.49	-
20.1930	-	-	3.63	0.23	6.58	2.83	-
20.2200	-	-	4.54	0.11	6.12	4.42	-
21.0030	-	-	7.26	0.11	3.40	2.15	-
21.0200	-	-	5.90	0.11	4.65	2.49	-
21.0400	-	-	3.18	0.45	6.12	2.49	-
21.0600	-	-	10.43	0.45	11.79	2.72	-
21.0800	-	-	4.76	0.34	4.08	1.36	-
21.1000	8.85	-	6.58	0.23	3.18	1.13	-
21.1130	4.99	-	-	-	-	-	-
21.1200	-	-	8.16	-	2.49	-	-
21.1230	3.86	-	-	-	-	-	-
21.1430	-	-	9.98	0.11	1.13	5.22	-
21.1700	-	-	11.79	-	-	-	-
21.1930	-	-	11.34	1.02	2.61	5.90	-
21.2030	-	-	6.80	-	1.02	-	-
21.2200	24.04	-	-	-	-	-	-
21.2245	-	-	11.79	0.68	1.13	4.54	-
22.0045	-	-	8.16	-	0.91	-	-
22.0245	11.34	-	8.62	0.45	0.91	5.90	-
22.0445	4.08	-	10.89	-	1.81	-	-
22.0645	3.63	-	8.16	0.23	1.13	4.54	-
22.0830	7.71	-	5.44	0.23	0.68	2.27	-
22.1030	-	-	7.14	0.23	1.36	2.04	-
22.1230	-	-	2.72	-	0.45	-	-
22.1430	-	-	4.54	0.11	0.91	-	-
22.1700	-	-	7.82	-	-	4.76	-
22.1830	-	-	9.07	0.	0.68	1.81	-
22.2000	-	-	-	-	-	0.68	-
29.0000	-	33.57	-	-	-	-	-
30.0000	-	-	-	-	17.24	-	-

APPENDIX C - TABLE XI

STONE FEED SAMPLES. **RUN 6**

SIEVE SIZE IN MICRONS

SAMPLE NUMBER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
					WT. PERCENT.					
50471	2.0730	.0	15.6	22.2	31.4	20.8	9.8	.0	.1	.2
50416	3.1300	.0	33.0	22.1	33.3	11.2	.0	.0	.0	.3
50428	3.2000	.0	26.7	19.1	35.7	11.7	5.0	.8	.1	.8
50484	5.1800	.0	21.1	18.3	38.9	13.7	5.7	1.2	.3	.8
50444	5.1030	.0	25.1	19.7	36.3	11.4	5.2	1.1	.4	.9
50503	6.1900	.0	16.4	14.8	31.4	18.1	14.5	3.0	1.1	.6
50417	6.2345	.0	9.2	10.1	28.0	20.6	24.7	4.1	1.5	1.8
50417	7.0845	.0	23.4	18.0	36.0	16.4	4.9	.5	.2	.6
50515	7.2000	.0	16.7	15.8	33.9	18.2	11.6	2.3	.8	.8
50559	7.2359	.0	15.7	16.0	36.6	19.8	9.1	1.4	.4	.8
50514	7.0800	.0	16.1	14.8	32.1	17.9	14.0	3.1	.9	1.1
50417	7.1000	.0	12.6	13.8	36.4	21.8	12.3	1.5	.7	.9
50569	8.0800	.0	16.8	14.3	28.7	17.1	16.9	3.7	1.2	1.2
50417	12.0500	.0	23.0	17.3	34.5	14.6	8.3	1.3	.5	.7
50618	12.1530	.0	19.4	15.9	30.1	18.4	12.2	1.9	.8	1.3
50627	13.0200	.0	25.6	19.8	33.1	12.1	7.5	1.2	.3	.3
50661	13.1700	.0	19.9	17.1	38.6	13.9	8.5	1.2	.3	.5
50670	14.0500	.0	15.5	13.7	30.4	23.6	14.1	1.5	.5	.7

APPENDIX C - TABLE XII
RUN 6

GASIFIER BED SAMPLES.
SIEVE SIZE IN MICRONS

SAMPLE NUMBER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
				WT. PERCENT.						
50473	2.1600	.1	18.9	16.5	33.2	20.3	10.9	.0	.0	.1
50405	2.2000	.0	28.8	13.1	27.6	19.5	10.8	.1	.0	.0
50411	3.0030	.0	20.5	15.4	30.8	20.5	10.3	.0	2.6	.0
50414	3.1300	.0	14.4	14.8	32.9	24.2	13.5	.0	.1	.0
50429	3.2000	.0	13.0	13.7	32.3	25.3	15.5	.2	.0	.0
50438	4.1400	.1	12.0	13.9	32.5	26.6	14.9	.0	.0	.0
50421	4.0200	.7	14.8	15.6	33.3	23.7	11.9	.0	.0	.0
50438	4.1400	.0	12.7	14.7	32.3	28.4	11.7	.1	.0	.1
50445	5.1030	.0	13.7	15.7	34.2	24.8	11.6	.0	.0	.1
50477	5.1800	.0	13.2	16.4	34.1	25.3	11.0	.0	.0	.0
50436	5.2359	.0	15.0	15.9	34.6	24.6	9.9	.0	.0	.0
50605	6.0500	.0	13.9	15.7	31.5	25.8	13.0	.0	.0	.0
50496	6.1900	.0	12.0	15.1	33.5	25.8	13.3	.2	.0	.0
50516	7.2000	.0	10.8	14.3	32.0	27.0	15.8	.0	.0	.0
50510	7.0800	.1	14.1	15.5	33.8	24.4	11.7	.0	.1	.3
50584	7.1530	.0	10.7	13.8	34.0	28.9	12.6	.0	.0	.0
50551	7.2359	.1	13.7	16.3	34.8	24.7	10.3	.0	.0	.0
50661	8.0800	.1	11.6	14.5	33.3	26.1	14.1	.1	.1	.1
50571	8.1630	.0	12.4	15.5	33.3	26.4	12.4	.0	.0	.0
50578	9.0400	.0	9.3	26.4	28.9	28.3	14.9	.0	.0	.1
50584	9.1530	.0	11.0	14.0	33.5	28.7	12.8	.0	.0	.0
50592	9.2330	.0	10.2	13.2	32.2	29.8	14.6	.0	.0	.0
50614	12.1530	.0	7.1	20.5	29.1	28.5	14.4	.1	.0	.2
50623	13.0200	.0	13.4	14.8	31.7	25.1	14.8	.0	.2	.0
50659	13.1700	.0	21.4	7.3	31.3	27.8	12.2	.0	.0	.0
50668	14.0500	.0	12.3	14.6	30.8	26.3	15.9	.1	.0	.0
50633	14.1500	.0	14.4	15.6	30.8	24.1	14.9	.1	.0	.1
50640	15.1800	.0	14.8	16.7	31.5	23.9	12.8	.2	.0	.1
50646	16.0100	.0	13.4	15.1	28.6	26.9	15.1	.8	.0	.0
50678	16.1000	.0	14.4	16.1	32.0	23.4	13.9	.1	.1	.1
50683	16.2300	.0	11.8	14.3	31.1	25.2	17.6	.0	.0	.0
50695	21.1400	.0	12.5	15.1	30.7	24.9	16.7	.0	.0	.1
50700	22.1000	.6	14.9	16.7	23.8	22.6	21.4	.0	.0	.0

APPENDIX C - TABLE XIII
RUN 6
REGENERATOR DRAIN SAMPLES

SIEVE SIZE IN MICRONS

SAMPLE NUMBER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
				WT. PERCENT.						
50472	2.0730	.1	21.5	16.2	31.6	19.7	10.6	.2	.0	.1
50474	2.2000	.1	18.6	14.4	30.5	20.3	14.6	1.1	.3	.0
50410	3.0100	.1	14.7	13.3	28.5	20.4	21.8	1.1	.2	.0
50420	3.1300	.1	14.3	14.0	29.8	23.0	18.1	.5	.1	.0
50431	3.2000	.0	14.0	13.8	29.5	24.5	17.6	.3	.1	.3
50437	4.1400	.1	14.9	14.6	31.7	24.1	14.6	.0	.0	.0
50447	5.1030	.0	22.9	12.6	27.6	22.5	13.8	.3	.1	.1
50480	5.1800	.0	12.5	14.1	31.8	25.5	16.0	.2	.0	.0
50489	5.2359	.1	10.3	12.5	30.1	26.6	19.5	.5	.2	.1
50601	6.0500	.0	9.9	11.8	31.9	25.4	20.4	.6	.0	.0
50499	6.1900	.0	10.9	13.2	30.1	25.8	19.4	.6	.1	.0
50511	7.0800	.0	11.4	13.2	23.0	27.2	24.2	.8	.2	.1
50517	7.2000	.0	11.9	14.2	31.2	25.6	16.7	.3	.1	.0
50554	7.2359	.1	2.4	16.1	38.6	32.0	10.5	.2	.1	.0
50564	8.0800	.0	12.1	14.2	32.2	27.2	14.0	.2	.0	.2
50573	8.1633	.0	.4	1.4	5.4	9.7	17.6	16.9	11.5	37.1
50579	9.0400	.0	10.7	12.9	31.5	27.9	16.6	.2	.0	.1
50593	9.2030	.0	12.2	14.3	31.0	28.5	13.9	.0	.0	.1
50585	9.1530	.1	11.5	14.6	32.9	27.6	13.2	.1	.0	.0
50610	12.1530	.0	13.8	14.9	29.6	25.1	16.3	.3	.1	.0
50619	13.0200	.0	13.9	15.4	29.7	25.2	15.6	.1	.1	.0
50656	13.1700	.0	12.1	13.9	29.2	26.6	17.7	.5	.0	.0
50665	14.0500	.0	11.6	13.3	30.1	24.2	20.3	.4	.1	.1
50634	14.1500	.0	10.9	13.4	32.3	30.6	10.4	2.0	.3	.3
50641	15.1800	.1	12.3	14.3	28.0	23.5	20.8	.8	.1	.0
50694	21.0400	.0	11.3	13.9	28.8	24.0	20.5	1.2	.2	.1
50701	22.1000	.1	16.4	11.1	22.7	22.6	24.9	1.3	.4	.5

APPENDIX C - TABLE XIV

RUN 6

ELUTRIATOR COARSE SAMPLES

SIEVE SIZE IN MICRONS

SAMPLE NUMBER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
		WT. PERCENT.								
50401	2.2000	.1	6.5	5.1	13.3	14.4	46.5	5.7	1.7	6.7
50407	3.0130	.1	3.5	4.2	13.5	17.2	54.2	4.8	.6	1.8
50471	3.1300	.1	6.5	7.1	18.9	20.9	41.5	2.2	.5	2.2
50534	3.2000	.1	6.5	7.0	19.9	24.0	37.5	1.8	.7	2.4
50426	4.0200	.1	7.1	8.6	22.4	23.5	33.7	1.1	.4	3.3
50439	4.1400	.0	4.6	6.2	19.1	24.1	41.0	1.1	.4	3.5
50442	5.1030	.1	5.5	7.2	20.4	24.2	32.2	1.0	.7	8.6
50478	5.1800	.0	8.7	9.6	20.2	20.2	23.1	1.0	.0	17.3
50487	5.2359	.0	6.7	8.7	23.6	26.6	28.4	1.3	.5	4.2
50497	6.1900	.0	2.1	3.3	9.0	13.0	40.8	6.6	3.8	21.5
50513	7.0800	.0	7.3	10.0	23.4	25.2	32.5	1.1	.2	.3
50575	8.1630	.0	3.8	5.5	17.0	24.3	42.1	2.8	.8	3.8
50582	9.0400	.0	2.7	4.4	15.3	23.1	42.3	4.3	1.6	6.1
50588	9.1530	.0	2.7	14.6	13.0	18.1	23.8	4.1	2.1	16.7
50596	9.2030	.0	1.7	2.6	9.4	12.8	25.0	7.8	5.2	35.6
50604	12.0500	.0	4.5	6.0	16.4	22.3	36.0	3.4	1.9	9.4
50613	12.1500	.0	3.3	4.8	14.0	21.3	43.0	4.6	2.2	6.8
50622	13.0200	.0	4.0	5.1	14.5	19.9	44.2	3.5	1.6	7.3
50645	13.1700	.0	3.1	4.5	12.6	18.0	44.8	4.5	1.9	10.5
50663	14.0500	.0	8.9	10.6	24.7	26.2	28.2	.5	.2	.8
50639	15.1800	.0	4.6	6.2	15.7	18.8	33.0	4.1	2.2	15.3
50643	16.0100	.0	5.8	8.0	19.6	23.4	31.7	2.8	1.6	7.1
50679	16.1000	.0	6.3	9.0	20.7	22.8	37.6	1.2	.3	2.3
50696	21.0400	.1	4.0	5.5	14.2	17.0	41.6	6.1	2.0	9.5

C. A. F. B. RUN 6

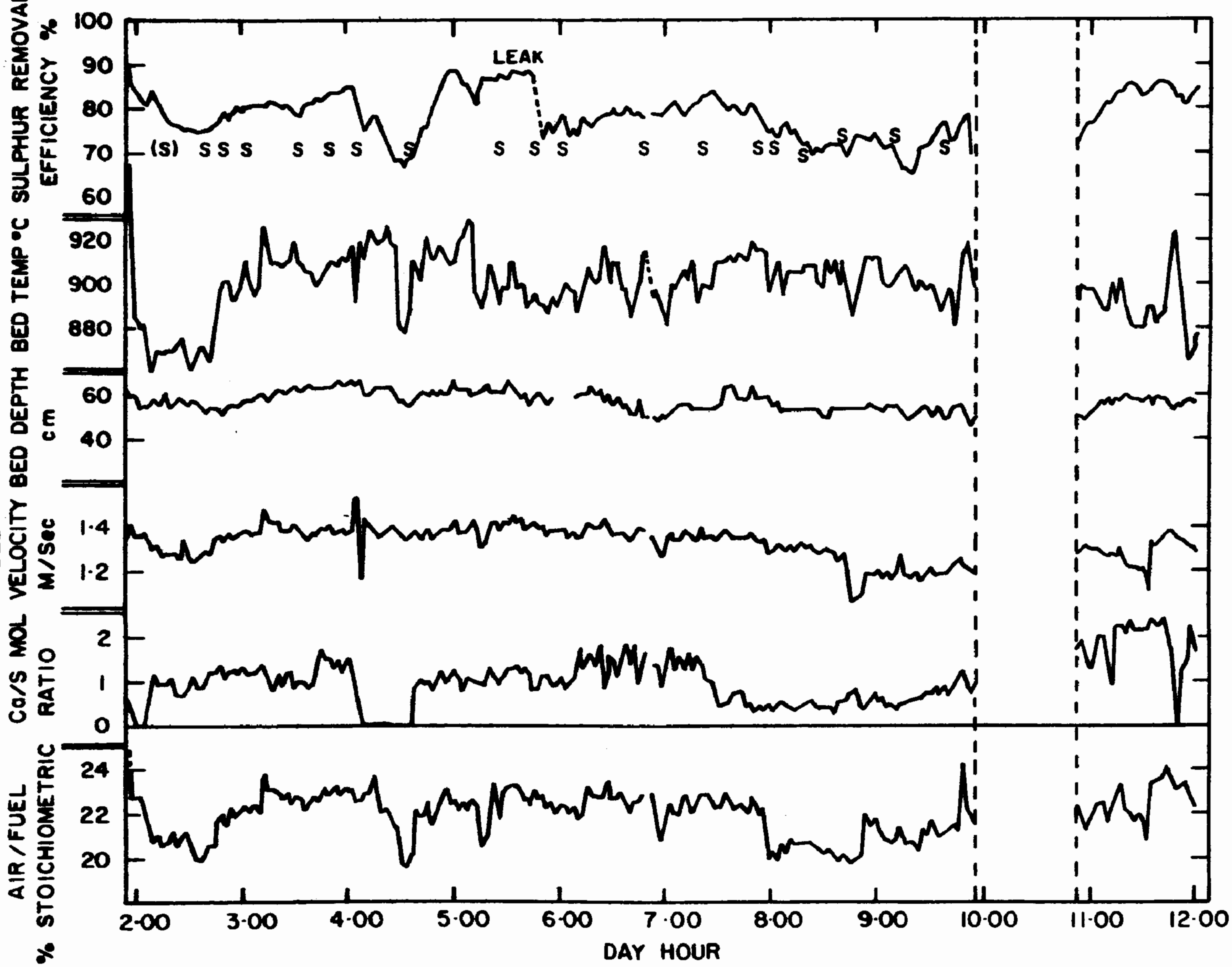


FIG. C15. SHEET. I.

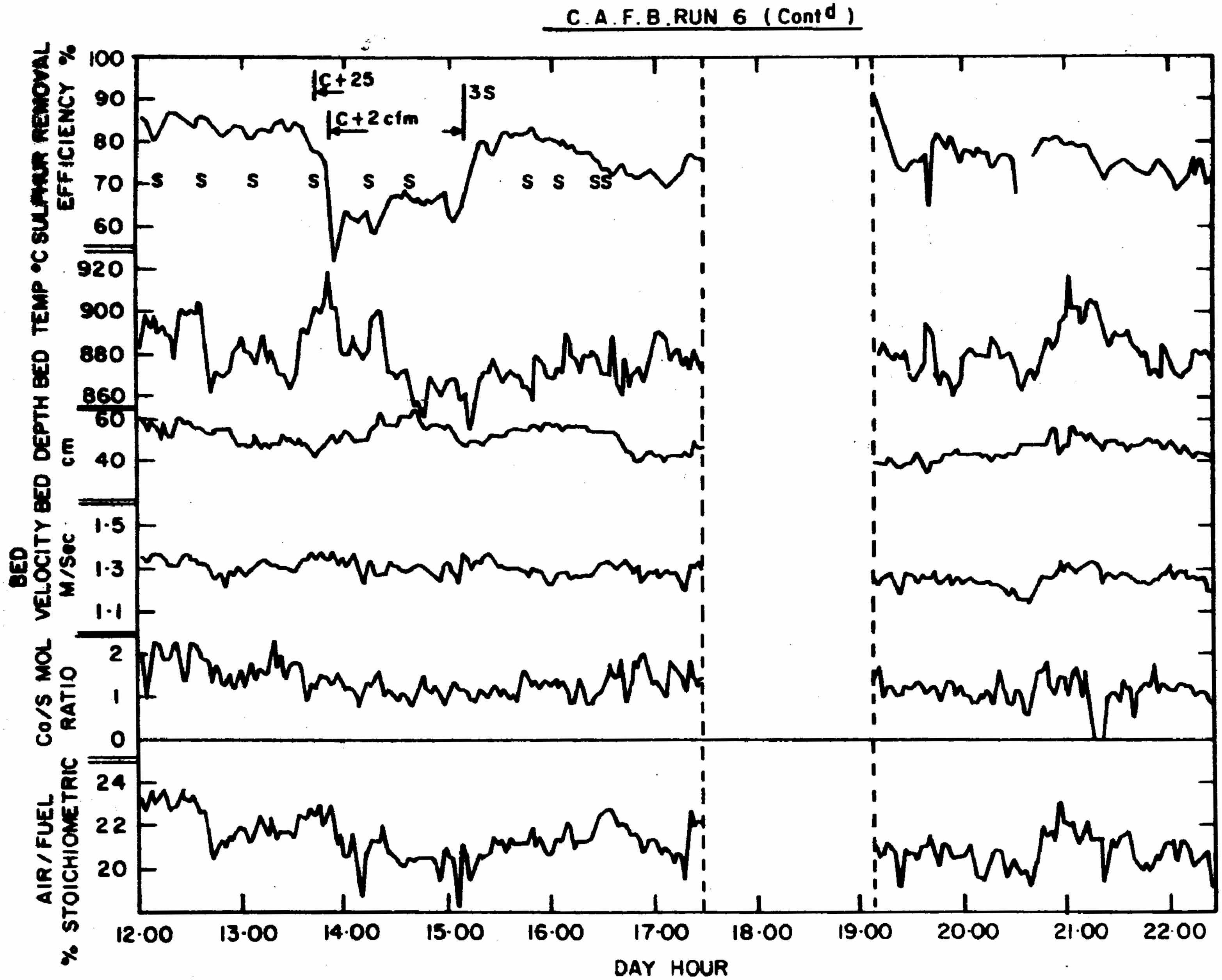


FIG. C15. SHEET 2.

APPENDIX D

RUN 7

Operational Log, Inspection, and Data

APPENDIX D

CAFB RUN 7

OPERATIONAL LOG

10.9.73 to 14.9.73 (Unit warm up)

Warm up commenced at 17.00 and the temperature was increased steadily at 12°C per hour until 500°C and thereafter increased at 20°C per hour bringing in kerosene at 700°C with a final temperature of 850°C. On 13.9.73 bed was added using material retained from the gasifier at the shut down after Run 6 and checks were made on circulation of bed material, fluidisation and fines transfer, the latter needing some adjustment for optimum performance.

On 14.9.73 the various systems were checked out, the fire tube was removed from the boiler rear end and the unit restarted on combusting conditions with kerosene. The first attempt at gasification was unsuccessful with a momentary main flame light up but followed soon afterwards by both main flame and pilot flame failure. Four further trials were also unsuccessful following the same pattern of main and pilot flame failure.

15.9.73 (Day 1 of Gasification)

The main flame was established at 12.15 following adjustments to the metering pumps and gas pilot flame. Conditions began to line out with a stone feed of 13.6 kgs/h (30 lbs/h) of Denbighshire limestone (3200 - 600 micron range), 900°C gasifier temperature and 171 kgs/h (377 lbs/h) fuel flow.

16.9.73 (Day 2)

The shooter for firing small quantities of the gasifier bed into the left hand cyclone inlet was started up so that regular short bursts of coarse material could be ejected into the cyclone entrance with the purpose of minimising material deposition in this critical area. The regenerator lower bed pressure tapping blocked frequently and repeated rodding was required to maintain its operation. Eventually the obstruction became so persistent that the regenerator bed depth measurement was obtained by measuring the pressure drop between the air inlet of the distributor and the gas space above the bed and subtracting the distributor pressure drop from this measurement to obtain the bed depth. Samples of bed material and dust from the various collection points were taken at 16.00.

The regenerator automatic bed transfer controller became irregular but the trouble was found to be caused by a damaged thermocouple cable from the regenerator which was repaired. Some investigations were made into the effect of the fines return to the gasifier and confirmed that the boiler SO_2 level was increased by these injections into the gasifier.

<u>17.9.73 (Day 3)</u>

The unit ran smoothly and trials were made with two methods of boiler SO_2 analysis. The hot gas cyclone sampling stream introduced in Run 6 was retained for this run but comparisons were needed with the system used in earlier runs which drew a much smaller gas sample stream through the boiler door. The pre-run 6 system was set up with the sample gases drawn through a knock out vessel for water removal and a cotton wool filter before passing to the sampling pump and analytical instruments. For convenience of installation the sampling point in the boiler door was 30 cms (11.8 ins) higher than used in previous work and angled downwards by 10°. The results of gas analysis from this point with the pre-run 6 system showed reasonably close agreement with the values obtained with the hot cyclone sampling stream method. The system was then returned to this latter method and a further trial would be made later in the programme.

At 07.00 the stone feed rate was increased to approximately twice stoichmetric so that the bed height could be built up to 63.5 cms (25 ins) and then the feed rate was reduced to about 11.8 kgs/h (26 lbs/h).

<u>18.9.73 (Day 4)</u>

The regenerator temperature controller which operates the bed transfer system was changed to control the regenerator to gasifier transfer with the gasifier to regenerator transfer controlled from the manually set timer. Some blockages were experienced in the regenerator fines return system, the shooter delivery pipe and in the left hand cyclone fines return system but all were cleared without problems.

Samples of bed material and dust from the various collection points were taken at 08.00 and at 12.00 the gasifier temperature was lowered to 880°C by increasing the flue gas rate and at 20.30 conditions were lined out smoothly.

19.9.73 (Day 5)

AT 02.00 a further set of samples was taken at the lower gasifier temperature with approximately the same limestone feed rate, bed depths and fuel flow. The gasifier temperature was then increased to 920°C and at 13.00 a further set of samples was taken with approximately constant limestone feed, bed depth and fuel flows.

The unit ran smoothly without any major problems and as usual maintained a very steady set of conditions. At 16.30 the limestone feed was cut back to about half stoichmetric and the unit left to line out ready for the next data point. At this stage there was no flue gas in use because of the high heat removal rate from the water tubes in the gasifier bed - about 13.8 kwatts (47,000 BTU/h).

20.9.73 (Day 6)

Some investigations were made into the effect of small changes in the air rate to the regenerator. First of all the rate was increased in small steps to find a maximum SO_2 removal rate. Within the short term it was found that the actual SO_2 removal was about constant over about 8.5 m^3/h (5 cfm) air rate change. The investigation was complicated by the change in bed circulation rate inherent in air rate changes which in turn changed pulser frequency and hence the quantity of nitrogen introduced into the regenerator offgas.

At 06.30 further bed material and dust samples were collected and the flue gas reintroduced to lower the gasifier bed temperature back to 880°C and it was observed that the boiler SO_2 level increased from 350 ppm to 430 ppm corresponding fairly exactly with this temperature change. Some problems were encountered with leaks on the chimney stack top washer pump which was then shut off for repair and some blockages were experienced in the fines return from the left hand cyclone. The shooter controller was adjusted to reduce the quantity of material delivered at each operation because of the possible danger of flooding the cyclone return handling system.

At 18.00 a set of bed material and dust samples was taken with the gasifier at 880°C and then the gasifier temperature was increased to 920°C again to determine if the previous effect on SO_2 concentration of a 40°C change in gasifier temperature could be repeated but the level did not return back to 350 ppm.

At 23.00 the water flow in the front bed cooling tube decreased sufficiently to allow the formation of superheated steam at 640°C but the water outlet temperature returned to 75°C when the water flow was increased.

21.9.73 (Day 7)

Preparations were made to feed the narrow cut Denbighshire stone (2180 - 300 micron range) and before the change was made bed material and dust samples were collected at 04.00. At 06.25 the narrow cut Denbighshire stone feed was started and while conditions were lining out further tests were made with the earlier sampling system on the boiler i.e. pre Run 6 system using the identical sampling point in the earlier runs. The initial result of this test showed a nearly zero ppm SO_2 level in the boiler. It was then noticed that the sampling hole which was situated in the centre of the boiler door was obstructed by a large deposit of red hot lime and after this was removed, the gas analysis using the Wöstoff analyser showed 110 ppm SO_2. Draeger tube tests showed 140 ppm upstream of the cotton wool filter in the sample line and 40 ppm downstream of this filter with the Wöstoff SO_2 analyser reading 70 ppm having gradually drifted down from the 110 ppm initial level. Further trials were made on the SO_2 sampling system throughout this day using five methods for analysis.

(a) Pre-run 6 configuration sampling from boiler door centre point.
(b) Pre-run 6 configuration except sample taken from the angled hole 30 cms (11.8 ins) above the boiler door centre point.
(c) Techniques (a) and (b) above but with a through bleed of gas from which a sample was drawn for analysis.
(d) The hot cyclone with continuous bleed as installed in Run 6.
(e) Technique (d) above but with cyclone outlet sealed off.

The results of this work showed that (a) and (b) gave similar results with SO_2 levels below 100 ppm - there being some variation depending upon the dampness of the cotton wool filter in the line. The results of analysis with technique (b) did not agree with the Day 3 trial earlier in this run and could be explained by there being less lime deposits in this sample tube on Day 3 due to the small number of hours run at that time. Trials with technique (c) gave increased SO_2 levels from less than 100 ppm to about 300 ppm.

During the day the sampling system was changed to technique (d) periodically to check the datum case which was averaging 500 ppm. A temperature of 1150°C at the entry of the gas sample stream into the tube through the boiler door hot face was measured by inserting a thermocouple through the sample tube.

At 15.30 technique (e) was tried and an initial SO_2 level of 260 ppm was indicated followed by a gradual drop to 70 ppm over a period of 40 minutes. The analysis was then returned to technique (d).

<u>22.9.73 (Day 8)</u>

Some tests were made upon the effect of increasing the air to the fuel injectors from 8.5 m^3/h (5 cfm) per injector to 17 m^3/h (10 cfm) and the result of this was to marginally increase the boiler SO_2 level. The gasifier space pressure had reached 6.3 kPa (25.5 ins) water gauge at 05.00 and before burning out the carbon,samples of bed material and dust were taken to assess the performance of the narrow cut Denbighshire stone.

At 09.30 the bed sulphation was commenced and proceeded quite normally being completed in about 20 minutes. The carbon burn out procedure was then started and there was some difficulty in achieving a reasonable gas flow rate through the ducts partly caused by a joint on the blower inlet which was found to have blown. There was some erratic behaviour in the gasifier bed thermcouples particularly in the top bed thermcouple which was getting cooled by the cold recycle gas entering the lid. The water flow rate in the bed cooling tubes was cut right back to less than 9 kgs/h (19.8 lbs/h) during this period with the slumped bed and low water outlet temperatures were easily maintained.

It was apparent that the duct thermcouples were all heavily coated with carbon because of their sluggish responses to flow changes and observation through the boiler inspection window showed the bifurcated duct thermcouple encrusted with red hot material. At 14.15 the gasifier temperature had dropped to 680°C and the unit was given a short period of fluidisation with kerosene added to boost the bed temperature. This caused the duct temperatures to rise rather rapidly reaching 1200°C indicating that there was still carbon to burn out. The bed was again slumped and burn out continued until 16.00 when, with duct temperatures dropping the unit was set on combusting conditions without undue duct temperature increases.

Both the cyclone drain legs were emptied to remove any chunks of material which may have broken away during the burn out procedure and then bed circulation was established on combusting conditions. The unit was then shut down to permit the cleaning of the boiler of accumulated lime.

23.9.73 (Day 9)

At 04.00 gasification was restarted and 27.2 kgs/h (59.9 lbs/h) stone feed supplied to build the bed but the response was slow suggesting that the cyclones were not performing very well. The left hand cyclone drain system was slow to build pressure and at every operation there were sharp kicks in the boiler SO_2 level suggesting that the butterfly sealing valve was not shutting off tightly. Adjustments were made to improve the valve seal by resetting the end stop on the actuator. During this period of operation with the 2185 - 300 micron size Denbighshire limestone there was no evidence of any improvement in desulphurisation efficiency and the losses with this fine stone were increased due to the bed cyclone performance. The feed was therefore switched to 3300 - 600 micron range Denbighshire stone at 13.00 hours.

At 18.30 some problems were encountered with the pneumatic controllers on the fines return system for the left hand cyclone and the drainage of this cyclone to the transfer vessel was not running freely. The drain leg was rodded out from above the cyclone and some lumps of material were removed from the transfer vessel following this operation.

24.9.73 (Day 10)

The shooter was reestablished to the left hand cyclone but did not increase the temperatures in the cyclone transfer vessel which would have been expected from past experience. Then followed some problems with the automatic valve controlling the flow of fine material back into the gasifier and during this period whilst the valve controller was under repair the fine material was returned to the gasifier in slugs by manual operation of the return valve. These slugs were sufficiently large to drop the gasifier temperature by up to 20°C and kicks in the boiler SO_2 level up to 100 ppm were associated with the return of each slug.

At 09.00 samples of bed material and dust were taken before raising the gasifier bed depth. Soon after this some problems arose with the fines return transfer pot outlet pipe which blocked periodically indicating that lumps of material were still falling from the cyclone drain leg and passing through the chunk trap installed in the transfer vessel. The bed circulation through the regenerator became a little slow and was improved by rodding through the regenerator to gasifier transfer duct.

25.9.73 (Day 11)

The bed circulation system again showed some erratic behaviour giving some problems in controlling the regenerator temperature but was improved by increasing the pulse rate on the nitrogen transfer controllers.

At 07.00 hours some samples of bed material and dust from the various collection points were taken to determine the effect of the increased gasifier bed depth whilst the unit ran steadily throughout the day. At 17.00 further samples were taken at nearly identical conditions before draining out a quantity of the bed material so that the feedstock could be changed to BCR 1359 (3200 - 600 micron range) with the minimum quantity of Denbighshire stone remaining so reducing the time delay in adequately purging the bed.

26.9.73 (Day 12)

The first eight hours of this period were spent in feeding BCR 1359 limestone into the unit and withdrawing nearly 400 lbs from the regenerator to improve the rate of change of the bed to BCR 1359 specification.

At 12.00 some adjustments were made to the fines return valve to try and slow down the rate of fines return and so prevent large slugs of fines passing through into the boiler. Some trouble was experienced with the fines return system which blocked in the transfer line probably caused by the higher proportion of coarse stone which prevented the material forming into discrete slugs for good transport. The regenerator cyclone which up to this point had been returning its fines into the gasifier via the elutriator was changed to drain into an external vessel for manual emptying.

27.9.73 (Day 13)

The scrubber knock out chamber drain choked and water was drawn into the recycle blower and into the recycle delivery line. The system was drained off and a permanent bleed made from the delivery pipe to prevent water being carried into the gasifier under fault conditions of the scrubber.

At 07.00 a set of bed material and dust samples was taken before studying the effect of 920°C gasifier operation with a high stone feed rate. The bed transfer system became erratic again and high pulsing rates were necessary to keep the regenerator temperature under control.

At 10.00 bed material and dust samples were collected before halving the stone feed from twice stoichmetric. The regenerator cyclone was not collecting much material which suggested that the right hand cyclone was not draining properly into the regenerator bed feed line due to either an obstruction or else an unfavourable pressure balance situation due to a high gasifier bed level and/or high cyclone entry pressure drop. At 12.00 the gasifier bed level was lowered to 53.5 cms (21 ins) to improve the pressure balance across the right hand cyclone drain leg.

At 19.00 a further sample of bed material and dust samples were taken before lowering the gasifier temperature. At 20.00 the fines return system to the gasifier was blocked off to investigate the unit performance without fines return. The collected fines were drained into a bucket and weighed and the two hour period of this operation yielded 18.2 kgs (40 lbs) of fines.

<u>28.9.73 (Day 14)</u>

The period during which the fines were withdrawn from the system did not make any significant difference upon the boiler SO_2 level apart from giving a smoother trace. The fines withdrawn during this test were then replaced into the gasifier with an apparent improvement in the bed circulation because the regenerator temperature dropped with the pulse settings remaining constant.

The shooter to the left hand cyclone was restarted after rodding out and burning the carbon from the exit pipe in the gasifier top space. At 08.00 the shooter stopped due to some control failure which permitted the vessel to overfill and subsequently the vessel would not empty. Some adjustments were made during the day to reduce fines carry over into the boiler by reducing the bleed rate in the cyclone drain legs and the regenerator cyclone was returned to external manual drain.

At 16.00 further problems were encountered with the bed transfer rate which required a fast pulsing rate to control the regenerator temperature. At 20.00 samples of bed material and dust were collected.

29.9.73 (Day 15)

The early part of this day was concerned with investigating the effect of various air flow rates upon the regenerator SO_2 release. The air rate to the regenerator was changed in increments allowing the conditions to steady out between changes. At 11.00 further samples were collected before raising the regenerator temperature. These changes in air flow and temperature in the regenerator were made in an attempt to cut down the stone sulphur loading and with 35.5 m^3/h (20.9 cfm) air flow to the regenerator the unit was given time to settle before taking bed material and dust samples at 17.45.

The regenerator air rate was then increased to 38 m^3/h (22.4 cfm) but the sulphur removal did not improve and there was difficulty in holding the temperature. At 19.45 the regenerator air rate was lowered to about 30.5 m^3/h (17.9 cfm). The elutriator fines return pipe to the gasifier blocked up at 20.45 and it was successfully drilled out.

30.9.73 (Day 16)

Preparations were made to connect up the single fuel injector which was positioned through the centre of the distributor. Prior to switching the fuel to this injector a further set of samples was taken at 15.00.

At 15.15 the centre fuel injector containing six outlet holes was pushed up into the gasifier so that the outlet centre line was at the same height as the side wall fuel injector hole centre i.e. 11.5 cms (4.5 ins) above the distributor nozzle centres. Initially there was a high air pressure through the injector but after some minutes the obstruction partially cleared and the air pressure dropped to 75.8 kPa (11 psi) for 13.6 m^3/h (8 cfm) flow. The centre side wall injector oil supply was diverted to the single injector at 17.20 and after levelling out without any obvious problems, the left hand side wall injector supply was added at 19.40 and at 20.40 the total oil supply was fed through the bottom injector. Air bleeds were left in the three side wall injectors for cooling purposes.

1.10.73 (Day 17)

Some trials were carried out on the effect of the single fuel injector height and there was some uneveness in the gasifier bed temperatures during this period. At 13.00 bed material and dust samples were collected. It was then planned to make quick tests by raising the fuel injector positions until a sharp deterioration in performance was reached but after moving 1.3 cms (.5 ins) upwards the injector jammed and could not be moved. Instead of this injector position test the unit was lined out with a high stone feed rate of 1½ stoichmetric and at 18.00 bed material and dust samples were collected before tests were made upon the effect of raising the water cooled tubes in the bed.

At 19.10 the rear water tube was raised up about 2.5 cms (1 in) with an immediate response in the water outlet temperature which was held at 70°C by increasing the flow rate to 250 kgs/hr (550 lbs/hr). At 20.00 the rear water tube was raised a further 2.5 cms (1 in) and the water outlet temperature approached 100°C with 250 kgs/hr (550 lbs/hr) flow rate. There were fairly marked drops in the gasifier bed temperatures at these changes in the water tube position.

During the previous 24 hours the gasifier space pressure had risen quite sharply and at 21.30 had reached 6.7 kPa (27 ins water gauge) which was near the maximum level recommended for safe loading on the gasifier lid. It was therefore agreed that the unit would be shut down by reducing the fuel flow in steps of approximately 11.4 kgs/hr (25 lbs/hr) keeping the gasifier superficial velocity constant by lowering the air rate to keep constant temperature but increasing the flue gas to maintain constant velocity in the bed. A series of step changes were made, each time adjusting the main burner air rate to keep about 2% oxygen in the boiler flue until at a flow of 82 kgs/hr (180 lbs/hr) of fuel the flame failure alarm came up and shut the unit down. This last period of gasification was the longest achieved with this pilot plant and cool down was made with nitrogen purges at various locations to prevent carbon burning off so that some measurements could be made of the carbon thickness after over 200 hours of continuous gasification.

APPENDIX D

CAFB RUN 7

INSPECTION

Gasifier and Regenerator Refractory

The gasifier walls were generally blackened overall with a thin layer of carbon deposited on the lower sections polished by the action of the bed material. The carbon above the bed close to the lid was up to 6 mm (.23 ins) thick with an irregular surface and in some areas it had bridged across to the underside of the lid and formed a joint to the refractory. The vertical cracks in the gasifier side walls present before this run had not deteriorated and had acquired the usual deposition of carbon along the line of the crack. There was one new crack which ran around the horizontal joint below the top refractory lift i.e. 28 cms (11 ins) below the top face. The concrete on the lid was generally good, again deposited with carbon (fig D.1) on its exposed face about 6 mms (.23 ins) thick. The insulation behind the hot face was cracked and in some areas large pieces had fallen away.

The transfer passages to and from the gasifier were in excellent condition without any cracks. The gas burner quarl did not show any deterioration and unlike previous runs the passage through the quarl was not heavily obstructed although there were some deposition in the lower section.

The regenerator bore was clear of any obstruction and on the wall at the top of the bore, well above the bed there were 1 cm (.4 ins) thick hard deposits of white fine material with a smooth very light purple external surface. A piece of this material was removed from the wall (fig. D.2). The cracks evident before the run had not deteriorated significantly.

The silicon carbide ring inserted as a spacer to lower the distributor was in good condition with some thin deposits on its inner face. The ring was firmly fixed in place by a mixture of fine and coarse material which had penetrated into the annulus between the outside of the ring and the refractory concrete hole into which it had been placed.

Gasifier and Regenerator Penetrations

The thermocouples, fuel injections, pressure tappings and drains were in good order throughout. Some of the bed thermocouples showed a local thinning of approximately .5 mm (.02 ins) on the 17 mm (.67 ins) diameter and in other areas there were deposits of similar magnitude. There was a layer of carbon deposited on the lid gas space thermocouple (fig. D.1) which was typical of the other penetrations in the gas space area. Part of the carbon had fallen away showing the thermocouple sheath beneath. Both the shooter tubes which protruded through the wall in this area were blocked with carbon, one tube had not been in use throughout the run and the other tube had stopped some hours before the shut down. The centre single fuel injector which passed through the distributor was apparently clear although the pressure-flow characteristic during operation suggested some obstruction. There was a small deposit of carbon on the top of this injector but this would not have influenced the fuel distribution or pressure drop characteristic. The regenerator penetrations were generally clean apart from the deposit at the bottom of the pressure tapping (fig. D.3).

Cyclones

The left hand cyclone inlet was heavily obstructed with a predominately carbon deposit around its entry (fig. D.4). The open area of duct remaining represented about 29% of the original area and at a point 4 cms (1.5 ins) from the entry section the remaining area was 32% but after a further 2 cms (.8 ins) the duct opened considerably to about 75% free area. The deposit was very hard and firmly attached to the refractory and analysis showed that it consisted of 80 - 85% carbon with the balance of calcium and sulphur. The carbon around the left cyclone entry was laid down with a corrugated surface finish with the lines parallel to the entry duct.

The origin of the loose piece of carbon (fig. D.4) bridging across the upper section is unknown but it may have fallen from the lid when a sudden pressure rise was observed in the gas space pressure near the termination of the run.

The right hand cyclone entry was heavily obstructed (fig. D.5) and here the obstruction in the duct became greater away from the entry section with a free area of 26% at the entry and 20% at a point of 4 cms (1.6 ins) into the duct. The deposit was again very firmly attached to the refractory and became lighter in colour away from the duct entry. The silicon carbide cyclone outlet tubes were cast into refractory collars which were located on top of the cyclone bodies and sealed with refractory cement. Unlike previous runs it was not possible to lift off these collars and they were gently cut away to show the inside of each cyclone immediately by the entry duct. Here the gas would first strike the centre tube before starting the downward vortex to enter the outlet tube at the bottom. (Fig. D.6) and (fig. D.7) show the left and right hand cyclone silicon carbide tubes were reasonably clean at the face opposite the gas entry but then acquired a heavy irregular deposit around the remaining area. There is some indication that the left hand cyclone was a little less obstructed and the shooter may have helped in this area. The deposits on the cyclone walls were extremely hard and strong and in some areas bridged across from the outer face of the silicon carbide outlet tube to the wall of the cyclone.

The left hand cyclone lower section was clear (fig. D.8) but the right hand cyclone was completely obstructed in the lower section by a build up of fine material with a complete crust over the top about 55 cms above the cyclone drain point. Beneath this crust there were other hard pieces with the finer material filling up the remaining space. (Fig. D.9) shows the drain after some of the fine material had been removed to show up the crust formations attached to the cyclone wall. Some of the crust had fallen away before the photograph was taken. It is most likely that the right hand cyclone was obstructed for a long period because the carbon on the outlet tube (fig. D.10) did not show the vortex profile of the left hand cyclone outlet (fig. D.11) indicating the swirling action of the outgoing gases produced by the correct functioning of a cyclone.

The internal drain leg which branches off the vertical external drain leg of the right hand cyclone to join the gasifier to regenerator transfer line was blocked at one of the bends between the cyclone bottom point and the bed material transfer passage. This obstruction prevented drainage and the cyclone then filled with material until the level rose sufficiently to prevent any further deposition and subsequently all the fine material passed straight out of the cyclone.

The left hand cyclone internal drain leg which branches off the vertical external drain leg to join the regenerator to gasifier transfer duct was sealed off before this run by a 50.1 mm (2 in) diameter stainless steel tube placed in the cyclone drain leg. The tube was removed after the run and there was an area of corrosion (fig. D.12) corresponding to the entry pipe of the internal drain connection. There was more severe thinning in the remaining material immediately around this area. The inside surface of the tube was not corroded indicating the corrosion was caused by gas coming up the drain leg from the regenerator to gasifier transfer line.

Gasifier and regenerator distributor

The gasifier distributor was in good condition with some thin lime deposits on some nozzles, with 5 of the 192 holes completely blocked, these nozzles were grouped near the fines return pipe to the gasifier. There were 53 other randomly placed nozzles which were partially blocked. The refractory which was in good condition had a thin layer of fine bed material in the defluidised zone below the nozzles which protruded higher than earlier distributor designs. The regenerator distributor was in good condition with one or two areas having a thin deposit of fine material. All the holes were quite clear (fig. D.13) apart from the centre drain.

Bed Material

The gasifier was shut down without sulphation and (Fig. D.14) shows the top of the bed after the lid was removed. The carbon debris on the bed fell from the lid which was firmly bonded to the gasifier upper wall by this carbon. The bed material was free from large agglomerates and within the depth of the bed some pieces of carbon were found about 2 cms (0.8 ins) across. The regenerator bed was free flowing and without agglomerates.

Water cooling tubes

Two water cooled tubes 27 mm (1.06 ins) outside diameter type 321 stainless steel were installed through the gasifier distributor each with independent water cooling control (fig. D.15). Initially they were to be placed in the retracted position below the fluidised zone thus minimising their possible heat pick up. Unfortunately due to some accidental displacement, the front tube was slightly exposed to the fluidised region of the bed and the results show that this tube absorbed more heat than the rear tube.

The front tube which was not raised up from this displaced position is shown in (fig. D.16) with the polished areas caused by local air impingement from local distributor nozzles. The tube was slightly distorted after the test and this may be due to a short term very high temperature excursion when the water cooling rate fell low enough, for the generation of steam. It is estimated that 60% of the tube area had carbon-lime deposits less than .25 mm (.01 ins) thick and 2% of the area was covered with deposits less than .90 mm (.035 ins) thick, the remainder being clean.

The rear water tube which is shown in (fig. D.17) was moved up into the fluid bed during the last day of the run with the cooling rate maintained to prevent steam formation. This tube was also slightly distorted from its original shape but like the front tube remained free from leaks. The material deposition was less marked due to the cleaning action of the fluid bed and showed 20% of the area deposited with material less than .25 mm (.035 ins) thick and 5% covered with material less than .90 mm thick with the remainder of the tube clean.

Bifurcated duct

The bifurcated duct between the gasifier cyclone outlet and the burner was generally clear apart from carbon deposits built up around the thermcouple and on the refractory walls. (Fig. D.18) shows the hot gas duct with carbon and lime deposits about 4 mm (.16 ins) at the top wall of the duct, 12 mm (.47 ins) on the bottom and 8 mm (.32 ins) on either side. The difference in the thickness of the deposits arose during operation and burn out when material may have dropped out of the gas stream.

Premix Section

The air premix section, installed between the bifurcated duct and the burner, consists of an inner insulated pipe with an annular gap through which the first stage air is admitted. The inner stainless steel pipe, in contact with the hot gas, was coated with carbon varying in thickness from 7 mm (.28 ins) to 4 mm (.16 ins). The outer steel shroud around the pipe insulation was heavily scaled by local high temperatures at the leading edge close to the mixing zone between the air and hot gas and in some areas the scaled material had broken away (fig. D.19).

Burner

(Fig. D.20) shows the carbon/lime deposits in the burner entry duct deposited around 60% of the duct periphery and leaving 66% of the duct area open. The deposit was hard but not firmly attached to the duct being retained by the geometry of the section. This entry duct is supported by a flange attached at the leading end to the burner body by a stainless steel flange. This flange was cracked along a weld line around approximately 40% of the periphery (fig. D.21) and coincided with the deposited material in the entry duct suggesting local temperature gradients due to shielding from the deposits. The pilot burner was slightly coated with lime but otherwise in good order.

Boiler and Stack

(Fig. D.22) shows the boiler after opening the rear door. All the tube ends were coated with lime deposits and 45% of the tubes were totally obstructed either by the thick layer of material built up in front of the tube plate or in some cases isolated tubes on the right and left hand side near the top of the array were plugged. Generally the plugs were hard and compacted, penetrating up to 10 cms (4 ins) into the tube base. The final pass boiler tubes all contained a thin coating of dust apart from one tube in an identical position on each side which was almost obstructed. The boiler corrugated fire tube had deposits of coarser material laying along the bottom.

The refractory was generally reasonable with light brown flaky deposits on the hot face at the end of the fire tube. The bricks in this area had loosened at their joints and will be rebuilt before the next run.

A total of 213 kgs (470 lbs) of material was removed from the boiler tubes, 121 kgs (267 lbs) removed from the front soot box, i.e. at the exit of the first tube pass and 161 kgs (355 lbs) removed from the area at the end of the flame tube. The stack and cylcone were clear with 3.2 kgs (7 lbs) of material deposited in the collection zone at the bottom of the stack.

Burner probe

The burner test probe was controlled at approximately 600°C by internal air cooling but there were circumstances when the temperature dropped due to the failure of the automatic controller which permitted full flow of cooling air. (Fig. D.23) shows the installed position of the tube with the free end protruding into the main flame path and the root of the tube shielded with the hot gas turning past it to enter the first tube pass.

The side of the tube end facing the burner was lightly covered with a hard tenacious white deposit varying from 0.5 mm (.02 ins) to 1.0 mm (.04 ins) thick and locally pitted by particle impingement. The trailing side of this end had a local brown fine deposit shown in (fig. D.24) with the adjacent area covered with a 0.25 mm (.01 ins) thick soft white deposit which was easily removed. The root end of the tube facing towards the boiler tube entries had one light brown deposit 12.5 mm (5 ins) thick and 12 cms (4.75 ins) long and on the opposite face there was a similar deposit 40 cms (15.7 ins) long and 6 mm (.24 ins) thick. The remainder of the root end was covered by a 0.4 mm (.16 ins) thick deposit which was fairly readily removed.

Fig. D.1 Gasifier Lid

Fig. D.2   Deposit from regenerator top

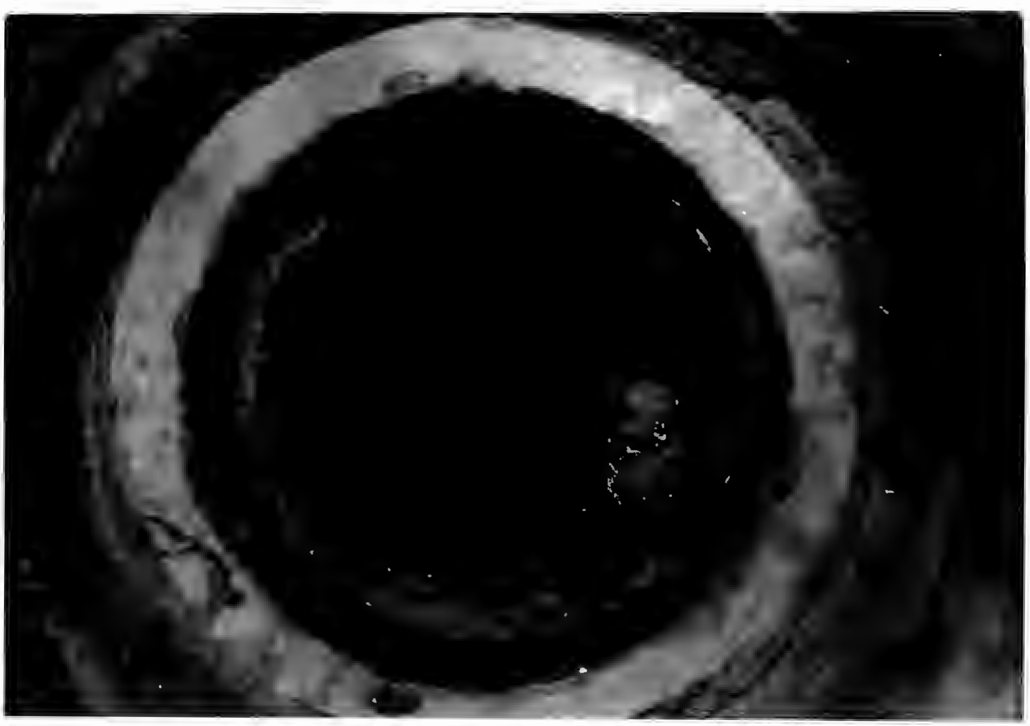


Fig. D.3   Regenerator bottom view

Fig. D.4   L.H. cyclone inlet

Fig. D.5   R.H. cyclone inlet

Fig. D.6 L.H. cyclone outlet tube in situ

Fig. D.7 R.H. cyclone outlet tube in situ

Fig. D.8 L.H. cyclone

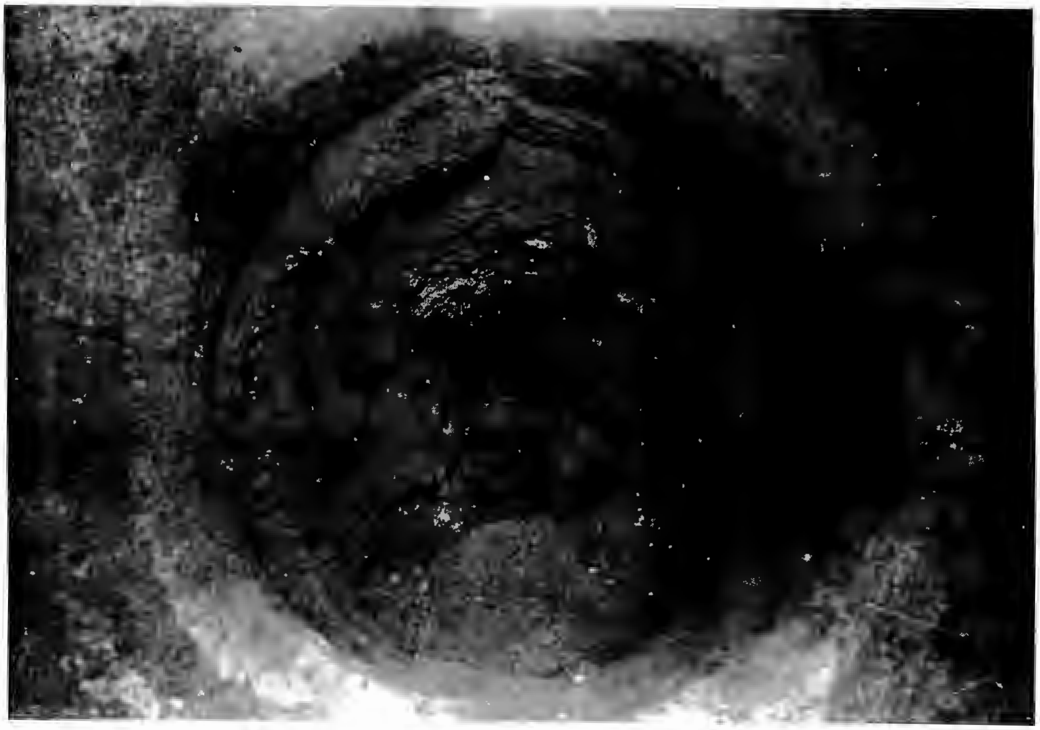


Fig. D.9 R.H. cyclone

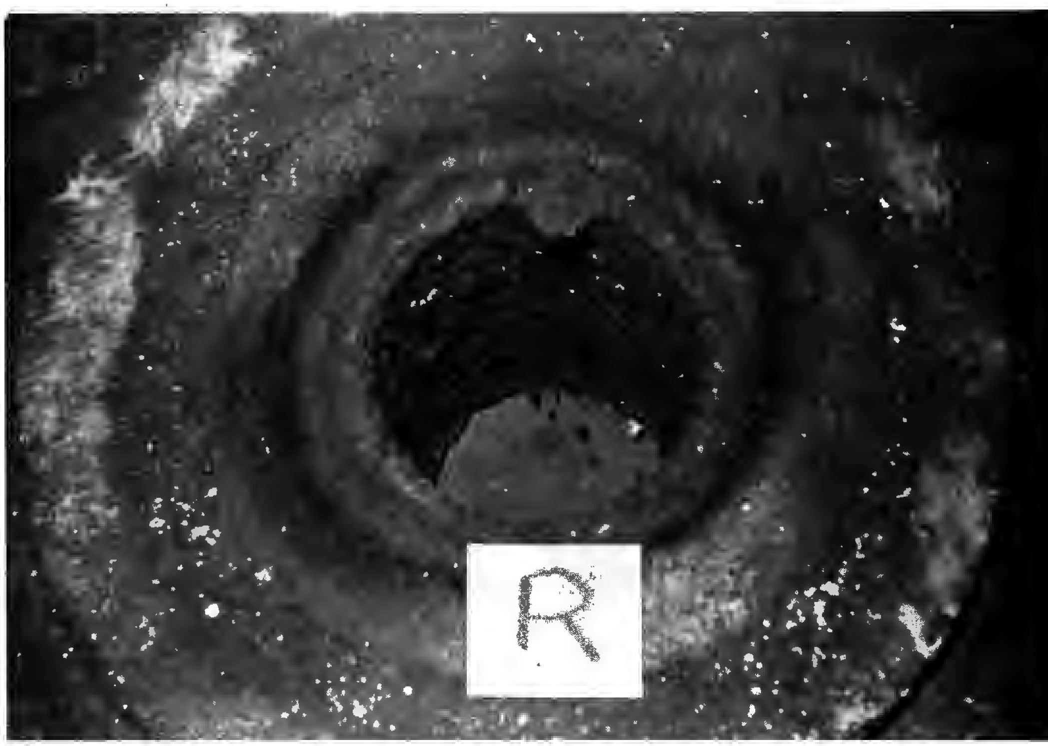


Fig. D.10 R.H. cyclone outlet

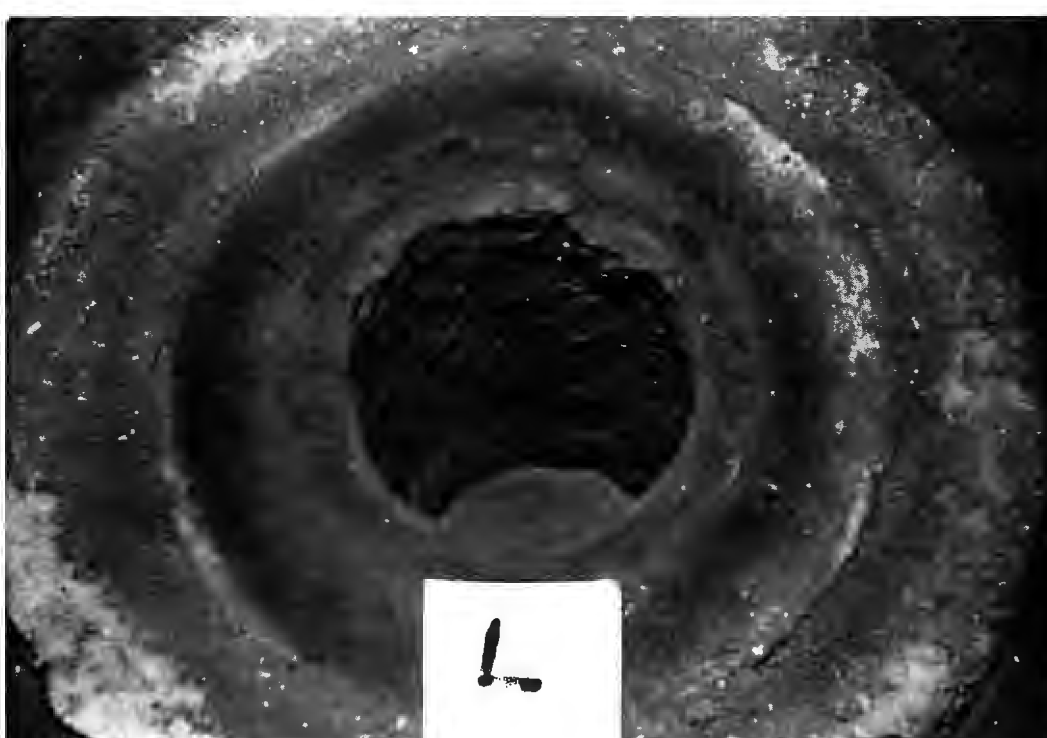


Fig. D.11 L.H. cyclone outlet

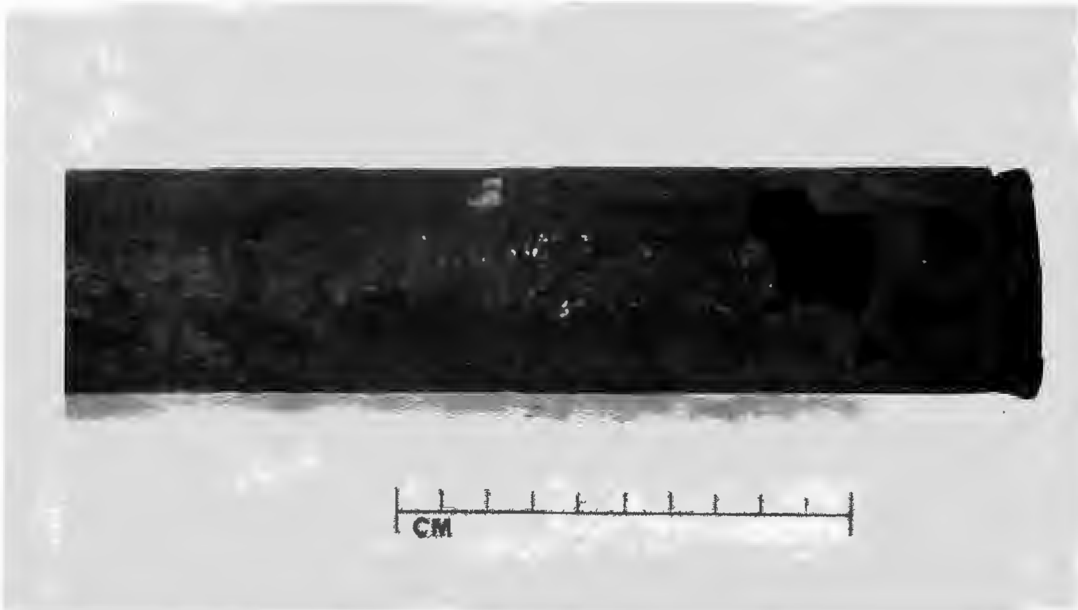


Fig. D.12 L.H. cyclone drain leg seal

Fig. D.13 Regenerator distributor

Fig. D.14 Gasifier Bed

Fig. D.15 Gasifier cooling tubes

Fig. D.16   Front cooling tube

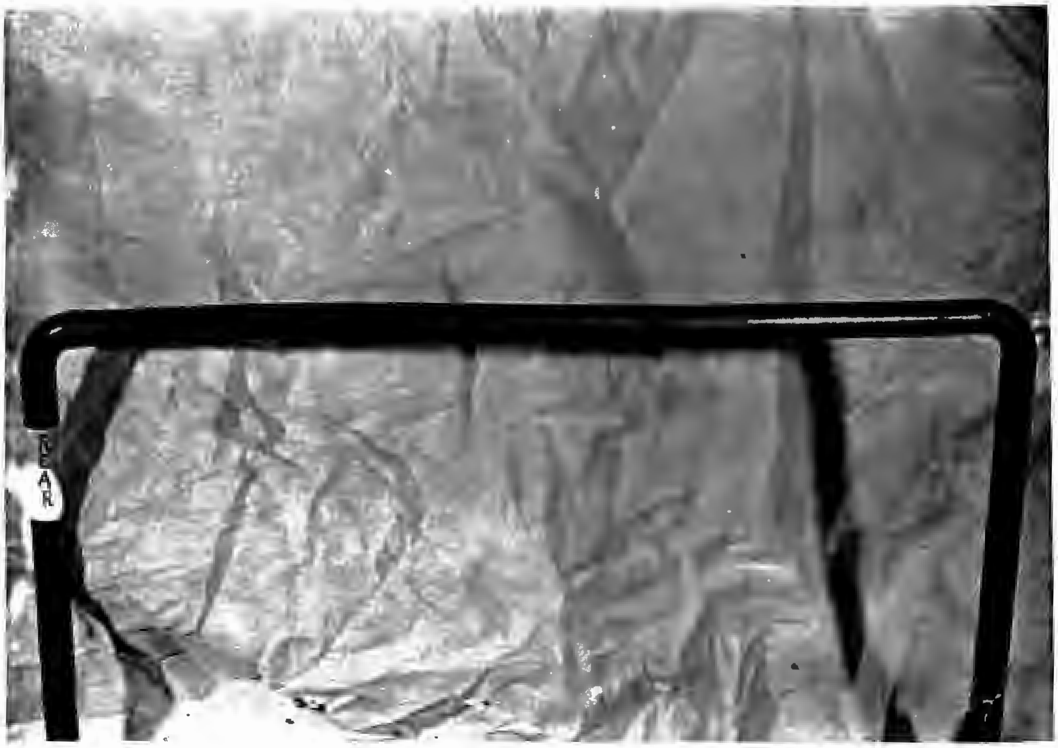


Fig. D.17   Rear cooling tube

Fig. D.18 Bifurcated duct

Fig. D.19 Burner premix section

Fig. D.20 Burner (rear view)

Fig. D.21 Crack in burner flange

Fig. D.22 Boiler back end

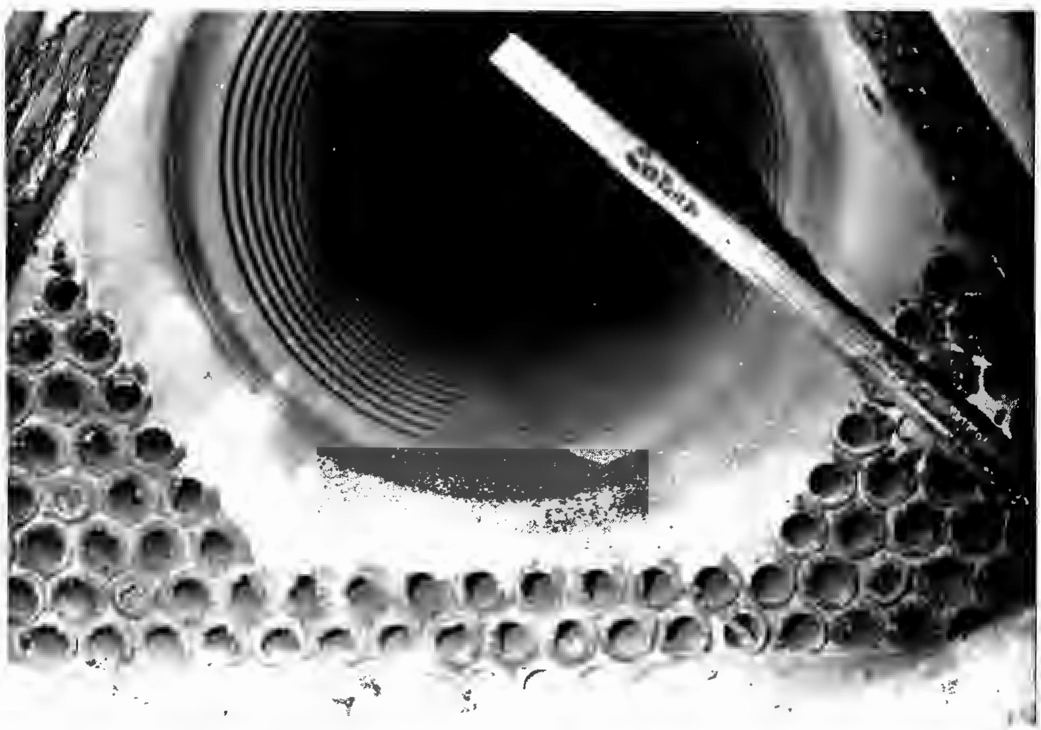


Fig. D.23 Boiler probe installation

Fig. D.24   Boiler probe

APPENDIX D - TABLE I

RUN 7: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
1.1230	940.	974.	62.	177.2	0.
1.1330	905.	1040.	70.	177.2	0.
1.1430	905.	1040.	70.	173.1	0.
1.1530	890.	1045.	70.	174.7	2.3
1.1630	888.	1035.	68.	174.7	15.0
1.1730	898.	1046.	68.	175.6	16.3
1.1830	905.	1032.	68.	175.1	9.5
1.1930	892.	1045.	70.	175.1	10.9
1.2030	876.	1030.	74.	175.1	10.9
1.2130	884.	1060.	75.	176.4	10.4
1.2230	899.	1062.	69.	174.7	10.0
1.2330	907.	1060.	69.	175.6	10.4
2.0030	911.	1065.	69.	175.1	13.2
2.0130	894.	1063.	69.	176.0	14.5
2.0230	898.	1059.	69.	175.6	12.7
2.0330	894.	1069.	69.	175.1	15.4
2.0430	903.	1062.	68.	175.6	17.2
2.0530	901.	1060.	69.	175.6	18.6
2.0630	902.	1055.	68.	174.7	19.1
2.0730	898.	1060.	68.	175.6	19.1
2.0830	899.	1062.	64.	175.1	18.6
2.0930	896.	1066.	69.	175.6	17.2
2.1030	885.	1070.	67.	176.0	20.0
2.1130	895.	1068.	65.	174.3	18.6
2.1230	884.	1068.	64.	175.1	20.9
2.1330	894.	1070.	64.	175.6	18.6
2.1430	900.	1070.	64.	176.0	10.9
2.1530	900.	1070.	65.	176.0	13.2
2.1630	904.	1078.	64.	176.4	14.5
2.1730	905.	1079.	61.	178.9	14.1
2.1830	892.	1082.	61.	173.5	11.8
2.1930	892.	1066.	66.	176.0	13.2
2.2030	896.	1068.	66.	176.0	13.6
2.2130	891.	1048.	65.	176.4	14.5
2.2230	891.	1050.	63.	176.4	14.1
2.2330	884.	1050.	62.	176.4	14.5
3.0030	899.	1055.	69.	176.4	15.0
3.0130	882.	1058.	63.	174.3	15.9
3.0230	889.	1050.	66.	177.6	14.1
3.0330	886.	1050.	63.	176.8	15.0

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
3.0430	881.	1052.	63.	176.4	15.0
3.0530	881.	1050.	69.	176.0	13.6
3.0630	889.	1061.	68.	176.4	21.3
3.0730	885.	1060.	58.	176.8	23.6
3.0830	881.	1060.	66.	176.4	25.9
3.0930	883.	1064.	68.	176.0	24.0
3.1030	889.	1060.	68.	176.4	22.2
3.1130	883.	1061.	69.	176.4	24.5
3.1230	866.	1061.	70.	176.4	32.2
3.1330	883.	1061.	68.	176.0	31.3
3.1430	880.	1062.	67.	178.0	29.0
3.1530	873.	1061.	67.	174.3	30.8
3.1630	880.	1062.	64.	176.0	31.8
3.1730	876.	1060.	64.	176.4	35.8
3.1830	878.	1060.	65.	176.4	27.2
3.1930	880.	1062.	64.	176.4	30.8
3.2030	878.	1063.	62.	176.4	28.6
3.2130	888.	1062.	60.	176.0	28.6
3.2230	899.	1061.	60.	176.4	16.3
3.2330	892.	1062.	60.	177.2	11.8
4.0030	899.	1062.	60.	180.5	10.9
4.0130	899.	1062.	60.	185.0	12.2
4.0230	888.	1070.	60.	185.9	13.2
4.0330	890.	1080.	60.	187.9	11.3
4.0430	890.	1060.	60.	187.1	12.2
4.0530	892.	1072.	60.	187.5	11.3
4.0630	855.	1080.	60.	187.9	10.0
4.0730	895.	1081.	60.	187.5	11.3
4.0830	895.	1082.	60.	187.9	14.1
4.0930	890.	1077.	60.	187.1	11.3
4.1030	897.	1065.	60.	187.9	9.1
4.1130	892.	1066.	60.	186.7	11.3
4.1230	878.	1065.	64.	184.6	11.3
4.1330	881.	1066.	65.	184.2	11.3
4.1430	885.	1060.	68.	184.2	12.2
4.1530	886.	1064.	69.	184.2	10.9
4.1630	886.	1068.	70.	184.2	10.0
4.1730	869.	1068.	66.	184.2	13.2
4.1830	880.	1069.	60.	184.6	14.5
4.1930	882.	1072.	60.	184.2	13.2

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
4.2030	882.	1060.	60.	185.0	11.3
4.2130	889.	1068.	60.	184.2	11.8
4.2230	885.	1068.	60.	185.9	13.2
4.2330	885.	1061.	60.	184.6	12.2
5.0030	887.	1065.	55.	185.0	12.2
5.0130	885.	1072.	54.	185.0	12.2
5.0230	908.	1062.	45.	185.5	14.5
5.0330	919.	1075.	42.	184.6	14.5
5.0430	922.	1064.	42.	185.0	14.1
5.0530	912.	1064.	42.	186.3	15.0
5.0630	917.	1060.	42.	189.6	14.5
5.0730	920.	1064.	42.	190.4	12.2
5.0830	918.	1074.	42.	190.8	15.0
5.0930	918.	1071.	45.	190.8	12.2
5.1030	920.	1069.	46.	190.4	10.4
5.1130	920.	1070.	44.	190.4	12.2
5.1230	920.	1075.	40.	190.4	11.8
5.1330	918.	1069.	42.	190.8	12.7
5.1430	917.	1073.	42.	190.4	12.7
5.1530	916.	1068.	45.	190.0	13.2
5.1630	915.	1070.	45.	190.0	13.6
5.1730	918.	1068.	40.	190.0	11.8
5.1830	925.	1070.	44.	190.4	8.6
5.1930	930.	1069.	30.	190.0	7.7
5.2030	926.	1065.	40.	189.6	6.8
5.2130	930.	1066.	45.	190.4	5.4
5.2230	930.	1062.	45.	190.0	6.8
5.2330	930.	1061.	48.	190.0	6.8
6.0030	930.	1065.	48.	190.0	6.8
6.0130	930.	1065.	48.	190.4	7.3
6.0230	928.	1073.	48.	191.2	8.2
6.0330	929.	1070.	48.	191.6	7.3
6.0430	925.	1070.	48.	191.6	7.7
6.0530	928.	1075.	48.	192.5	8.2
6.0630	929.	1072.	48.	192.0	6.8
6.0730	911.	1074.	60.	191.6	7.7
6.0830	912.	1070.	60.	192.0	10.4
6.0930	912.	1075.	60.	192.0	10.9
6.1030	920.	1071.	60.	191.6	9.1
6.1130	920.	1080.	59.	191.6	6.8

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
6.1230	878.	1078.	69.	192.5	5.0
6.1330	888.	1069.	68.	191.2	6.4
6.1430	878.	1072.	68.	192.0	7.3
6.1530	881.	1070.	68.	191.6	7.7
6.1630	880.	1070.	68.	191.2	7.3
6.1730	874.	1062.	66.	192.5	10.4
6.1830	890.	1070.	68.	190.8	8.6
6.1930	919.	1060.	62.	192.5	6.8
6.2030	923.	1072.	61.	192.0	8.6
6.2130	924.	1070.	60.	192.0	7.3
6.2230	926.	1078.	60.	191.2	8.6
6.2330	921.	1065.	61.	192.0	7.7
7.0030	910.	1072.	61.	192.0	5.9
7.0130	918.	1074.	61.	188.7	9.1
7.0230	912.	1072.	61.	194.9	9.1
7.0330	903.	1072.	60.	192.0	9.1
7.0430	921.	1072.	60.	191.6	6.8
7.0530	920.	1074.	60.	191.6	7.3
7.0630	914.	1078.	60.	191.6	3.2
		STONE CHANGE			
7.0730	908.	1070.	60.	192.5	15.4
7.0830	899.	1069.	59.	192.9	18.1
7.0930	903.	1070.	62.	192.0	13.6
7.1030	903.	1067.	63.	193.3	11.8
7.1130	902.	1065.	63.	183.8	12.7
7.1230	902.	1065.	62.	193.3	13.6
7.1330	898.	1067.	64.	192.0	13.2
7.1430	899.	1070.	65.	192.9	11.8
7.1530	895.	1070.	65.	192.5	14.1
7.1630	890.	1069.	65.	190.0	15.0
7.1730	896.	1069.	62.	193.3	14.5
7.1830	898.	1062.	69.	192.9	10.9
7.1930	905.	1070.	61.	191.6	10.9
7.2030	902.	1066.	61.	193.3	12.7
7.2130	909.	1069.	60.	190.4	10.9
7.2230	907.	1069.	60.	192.0	9.5
7.2330	911.	1069.	60.	191.6	7.7
8.0030	919.	1071.	60.	191.2	9.5
8.0130	910.	1069.	61.	191.6	10.4
8.0230	909.	1069.	61.	192.0	10.9

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
8.0330	908.	1070.	61.	191.6	11.8
8.0430	903.	1067.	61.	191.6	11.8
8.0530	902.	1062.	60.	192.0	13.6
8.0630	888.	1071.	60.	191.2	22.2
8.0730	883.	1069.	60.	191.6	26.3
8.0830	881.	1062.	60.	191.6	28.6
SHUT DOWN AT 8.0830 FOR 21 HOURS					
9.0530	889.	1039.	70.	173.1	26.8
9.0630	882.	1045.	70.	183.8	30.4
9.0730	902.	1052.	70.	178.0	25.4
9.0830	895.	1059.	71.	175.6	25.4
9.0930	882.	1050.	70.	175.1	23.6
9.1030	882.	1056.	68.	175.6	30.8
		STONE CHANGE			
9.1130	872.	1058.	67.	177.6	29.0
9.1230	869.	1060.	65.	181.7	46.7
9.1330	868.	1060.	60.	183.8	54.0
9.1430	880.	1065.	57.	184.2	50.3
9.1530	892.	1066.	53.	183.8	51.3
9.1630	926.	1068.	50.	185.5	39.5
9.1730	950.	1062.	40.	184.6	14.1
9.1830	955.	1068.	50.	194.9	20.4
9.1930	922.	1068.	30.	190.8	19.5
9.2030	922.	1065.	30.	190.8	21.8
9.2130	908.	1064.	32.	190.4	22.7
9.2230	922.	1064.	35.	190.4	20.9
9.2330	922.	1062.	39.	190.8	21.3
10.0030	930.	1062.	40.	190.4	18.6
10.0130	930.	1060.	39.	191.2	20.0
10.0230	919.	1062.	39.	190.4	23.1
10.0330	924.	1062.	38.	190.4	21.8
10.0430	924.	1062.	34.	191.6	22.2
10.0530	922.	1062.	35.	190.8	22.2
10.0630	933.	1062.	35.	191.2	21.8
10.0730	922.	1062.	40.	190.8	23.1
10.0830	918.	1062.	23.	190.4	23.6
10.0930	900.	1063.	25.	191.2	31.3

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
10.1030	905.	1068.	26.	191.6	50.3
10.1130	910.	1070.	26.	189.2	49.0
10.1230	892.	1068.	26.	193.3	49.9
10.1330	888.	1068.	28.	190.8	51.7
10.1430	882.	1070.	29.	190.8	51.3
10.1530	895.	1070.	30.	191.2	48.5
10.1630	889.	1069.	30.	190.8	50.8
10.1730	870.	1070.	30.	191.2	54.9
10.1830	902.	1071.	30.	190.8	29.5
10.1930	919.	1068.	30.	190.8	20.0
10.2030	911.	1067.	28.	191.2	21.3
10.2130	912.	1070.	28.	190.8	20.4
10.2230	912.	1069.	25.	190.8	22.7
10.2330	910.	1073.	26.	191.2	22.2
11.0030	910.	1065.	28.	190.8	19.5
11.0130	912.	1062.	28.	190.4	19.5
11.0230	918.	1067.	28.	190.8	19.5
11.0330	916.	1068.	28.	190.8	21.8
11.0430	918.	1066.	28.	190.4	22.7
11.0530	912.	1075.	28.	191.2	23.6
11.0630	912.	1064.	28.	190.4	22.7
11.0730	914.	1067.	28.	190.8	22.7
11.0830	914.	1068.	26.	190.8	22.7
11.0930	905.	1078.	26.	190.8	23.6
11.1030	904.	1066.	27.	190.4	23.6
11.1130	910.	1065.	27.	190.8	22.2
11.1230	906.	1071.	30.	189.6	23.6
11.1330	905.	1070.	30.	190.8	25.4
11.1430	910.	1070.	30.	190.4	22.2
11.1530	922.	1070.	30.	190.4	18.1
11.1630	912.	1070.	30.	190.0	18.1
11.1730	912.	1070.	30.	190.4	19.1
11.1830	911.	1060.	30.	190.4	23.6
11.1930	915.	1062.	30.	193.7	0.3
11.2030	915.	1065.	30.	189.2	21.8
11.2130	900.	1060.	30.	189.2	21.3
11.2230	895.	1060.	60.	188.7	12.2
11.2330	885.	1053.	71.	187.9	0.9
		STONE CHANGE			
12.0030	874.	1060.	72.	183.0	26.3

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
12.0130	892.	1060.	65.	176.8	30.4
12.0230	900.	1052.	40.	176.0	32.7
12.0330	895.	1054.	35.	175.6	34.9
12.0430	893.	1055.	36.	176.0	39.9
12.0530	894.	1050.	32.	175.6	40.8
12.0630	895.	1050.	34.	176.0	37.2
12.0730	897.	1052.	34.	175.6	37.6
12.0830	889.	1050.	33.	174.3	43.1
12.0930	886.	1050.	32.	173.5	39.5
12.1030	879.	1060.	29.	173.9	36.7
12.1130	889.	1055.	29.	173.5	27.7
12.1230	890.	1059.	32.	173.5	32.2
12.1330	888.	1060.	29.	173.1	33.6
12.1430	883.	1060.	29.	173.5	32.2
12.1530	880.	1060.	28.	173.5	37.6
12.1630	875.	1059.	30.	172.7	38.1
12.1730	870.	1058.	30.	173.1	44.9
12.1830	884.	1059.	30.	173.1	46.3
12.1930	880.	1055.	30.	173.5	42.6
12.2030	879.	1062.	30.	172.7	35.8
12.2130	874.	1058.	32.	177.2	36.7
12.2230	876.	1058.	32.	171.0	45.4
12.2330	885.	1059.	31.	180.9	41.3
13.0030	892.	1060.	42.	173.9	33.6
13.0130	902.	1060.	36.	173.1	34.9
13.0230	909.	1067.	34.	173.5	37.6
13.0330	920.	1072.	36.	174.3	26.8
13.0430	918.	1062.	38.	173.9	29.9
13.0530	919.	1063.	38.	173.9	28.6
13.0630	916.	1060.	38.	173.9	39.0
13.0730	908.	1066.	38.	175.6	30.8
13.0830	919.	1065.	39.	175.6	40.4
13.0930	910.	1052.	38.	173.5	31.8
13.1030	920.	1071.	41.	171.0	32.2
13.1130	915.	1055.	39.	175.6	18.6
13.1230	920.	1061.	32.	173.1	15.0
13.1330	920.	1062.	32.	173.9	10.0
13.1430	918.	1058.	30.	173.1	11.8
13.1530	920.	1058.	29.	173.9	11.8
13.1630	920.	1058.	28.	172.3	13.6

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
13.1730	923.	1051.	28.	169.8	16.8
13.1830	918.	1049.	28.	175.6	18.1
13.1930	923.	1049.	23.	173.9	19.5
13.2030	925.	1058.	23.	174.3	14.1
13.2130	930.	1060.	37.	174.7	14.1
13.2230	928.	1059.	24.	174.7	16.3
13.2330	929.	1059.	29.	174.3	13.6
14.0030	919.	1059.	31.	176.0	12.2
14.0130	920.	1048.	30.	173.9	14.1
14.0230	921.	1059.	30.	174.3	14.1
14.0330	929.	1061.	30.	174.3	12.7
14.0430	883.	1050.	52.	174.7	12.2
14.0530	879.	1059.	60.	173.9	13.2
14.0630	880.	1060.	61.	175.6	14.5
14.0730	882.	1063.	61.	175.1	12.7
14.0830	890.	1063.	65.	174.3	12.7
14.0930	889.	1069.	62.	175.6	12.2
14.1030	881.	1063.	63.	174.3	13.2
14.1130	881.	1060.	62.	175.1	12.7
14.1230	881.	1058.	62.	175.1	11.8
14.1330	880.	1055.	62.	174.7	12.7
14.1430	881.	1050.	62.	174.7	12.7
14.1530	882.	1065.	62.	175.1	12.2
14.1630	881.	1048.	62.	175.1	12.7
14.1730	872.	1048.	60.	174.7	15.4
14.1830	875.	1041.	60.	174.7	20.0
14.1930	870.	1042.	61.	174.7	15.0
14.2030	880.	1048.	60.	175.1	12.7
14.2130	878.	1042.	61.	176.0	14.5
14.2230	875.	1045.	61.	176.0	15.0
14.2330	886.	1048.	61.	173.5	10.9
15.0030	882.	1045.	61.	174.3	11.3
15.0130	880.	1049.	59.	175.6	13.2
15.0230	880.	1041.	60.	175.1	16.8
15.0330	885.	1040.	62.	175.6	15.0
15.0430	882.	1039.	62.	174.3	15.9
15.0530	882.	1036.	58.	174.3	17.2
15.0630	888.	1037.	60.	175.1	15.9
15.0730	888.	1039.	61.	174.3	14.5
15.0830	880.	1038.	59.	173.9	20.4

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
15.0930	879.	1038.	61.	175.1	17.7
15.1030	876.	1038.	62.	174.7	17.2
15.1130	876.	1048.	63.	174.7	19.1
15.1230	878.	1056.	63.	174.7	18.6
15.1330	890.	1058.	63.	174.3	15.4
15.1430	885.	1060.	60.	174.3	18.1
15.1530	884.	1064.	60.	174.7	17.2
15.1630	882.	1064.	60.	174.3	15.9
15.1730	877.	1060.	61.	174.7	17.2
15.1830	885.	1059.	61.	174.7	16.8
15.1930	883.	1064.	62.	174.3	15.0
15.2030	875.	1065.	63.	174.7	19.5
15.2130	898.	1060.	60.	174.7	16.3
15.2230	890.	1060.	60.	173.9	15.9
15.2330	877.	1060.	60.	173.9	18.6
16.0030	880.	1060.	60.	174.3	15.4
16.0130	900.	1080.	60.	172.3	14.5
16.0230	911.	1076.	48.	173.9	16.8
16.0330	932.	1075.	42.	174.3	17.2
16.0430	935.	1078.	40.	173.9	10.9
16.0530	917.	1078.	40.	173.9	13.2
16.0630	910.	1070.	40.	174.7	15.0
16.0730	902.	1072.	40.	174.3	14.1
16.0830	904.	1070.	40.	173.5	13.2
16.0930	898.	1064.	41.	173.9	17.2
16.1030	898.	1063.	42.	174.3	16.3
16.1130	899.	1065.	41.	174.3	15.4
16.1230	896.	1062.	41.	173.9	15.4
16.1330	896.	1061.	41.	174.7	15.4
16.1430	899.	1062.	41.	174.3	12.2
16.1530	898.	1060.	50.	174.3	12.2
16.1630	895.	1060.	48.	175.6	16.3
16.1730	895.	1057.	48.	175.6	17.2
16.1830	899.	1060.	45.	176.4	13.6
16.1930	900.	1066.	46.	176.0	12.2
16.2030	896.	1059.	48.	175.1	15.0
16.2130	895.	1062.	45.	175.1	16.8
16.2230	908.	1060.	45.	175.6	12.2
16.2330	910.	1065.	45.	1755.6	14.1
17.0030	910.	1062.	45.	175.6	13.2

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	FEED RATE KG/HR OIL	STONE
17.0130	902.	1055.	48.	174.3	13.2
17.0230	905.	1055.	48.	174.7	11.3
17.0330	903.	1054.	50.	173.5	9.5
17.0430	962.	1053.	50.	175.6	11.8
17.0530	894.	1055.	50.	174.3	13.2
17.0630	898.	1056.	50.	175.1	12.2
17.0730	890.	1053.	50.	174.7	15.0
17.0830	889.	1050.	50.	174.7	17.2
17.0930	910.	1055.	50.	174.7	13.6
17.1030	899.	1056.	49.	175.1	8.2
17.1130	894.	1055.	49.	175.1	12.7
17.1230	897.	1058.	48.	175.6	13.6
17.1330	898.	1058.	45.	176.0	14.5
17.1430	900.	1057.	45.	173.9	10.4
17.1530	894.	1054.	40.	175.1	19.5
17.1630	897.	1054.	37.	168.1	25.4
17.1730	893.	1057.	35.	166.1	25.9
17.1830	900.	1056.	35.	166.9	26.8
17.1930	902.	1060.	35.	166.1	16.8
17.2030	900.	1059.	36.	166.1	13.6
17.2130	892.	1060.	34.	165.7	16.8

APPENDIX D - TABLE II

RUN 7: GAS FLOW RATES PAGE 1 OF 10

DAY.HOUR	GAS RATES M3/HR					REGEN.
	GASIFIER		PILOT	REGENERATOR		VELOCITY
	AIR	FLUE GAS	PROPANE	AIR	NITROGEN	M/SEC
1.1230	437.	183.	3.7	31.4	4.5	1.54
1.1330	411.	214.	3.7	30.6	4.7	1.59
1.1430	394.	214.	0.	33.2	3.0	1.63
1.1530	403.	224.	0.	32.9	2.6	1.61
1.1630	386.	154.	0.	27.4	3.4	1.39
1.1730	403.	144.	3.7	32.0	3.6	1.62
1.1830	386.	168.	3.7	31.7	3.8	1.59
1.1930	385.	165.	3.7	33.3	3.7	1.68
1.2030	402.	138.	3.7	26.8	2.8	1.33
1.2130	407.	132.	3.7	27.1	2.5	1.36
1.2230	420.	127.	3.7	27.7	2.5	1.39
1.2330	421.	117.	3.7	28.0	2.8	1.42
2.0030	411.	117.	3.7	28.5	2.6	1.43
2.0130	404.	117.	3.7	27.7	2.9	1.40
2.0230	412.	117.	3.7	28.9	2.5	1.44
2.0330	412.	117.	3.7	32.8	2.9	1.65
2.0430	421.	117.	3.7	31.3	2.6	1.55
2.0530	421.	107.	3.7	31.0	2.9	1.55
2.0630	430.	107.	3.7	30.3	2.6	1.50
2.0730	421.	107.	3.7	31.0	2.9	1.55
2.0830	430.	97.	3.7	30.3	2.5	1.51
2.0930	417.	97.	3.7	30.7	2.7	1.53
2.1030	415.	97.	3.7	29.1	2.5	1.46
2.1130	420.	87.	3.8	29.4	2.5	1.47
2.1230	416.	87.	3.8	29.5	2.5	1.48
2.1330	419.	87.	3.7	29.7	2.5	1.49
2.1430	416.	81.	3.7	29.7	2.4	1.48
2.1530	421.	79.	3.7	29.5	2.4	1.47
2.1630	421.	77.	3.7	29.2	2.4	1.47
2.1730	419.	77.	3.6	32.6	2.5	1.63
2.1830	416.	97.	3.6	26.5	2.2	1.34
2.1930	416.	91.	3.6	29.7	4.2	1.56
2.2030	416.	87.	3.7	33.5	2.8	1.67
2.2130	416.	107.	3.6	33.8	4.3	1.73
2.2230	416.	97.	3.6	37.1	4.3	1.88
2.2330	411.	97.	3.6	33.2	4.3	1.70
3.0030	428.	97.	3.6	34.7	4.4	1.78
3.0130	428.	97.	3.6	33.5	4.7	1.74
3.0230	419.	97.	3.6	31.4	2.5	1.54
3.0330	419.	96.	3.6	31.4	2.5	1.54

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
3.0430	419.	87.	3.6	30.5	4.4	1.59
3.0530	410.	97.	3.6	31.1	4.1	1.60
3.0630	428.	97.	3.6	30.8	4.0	1.59
3.0730	438.	96.	3.7	33.5	4.3	1.73
3.0830	438.	97.	3.7	32.6	4.0	1.67
3.0930	435.	97.	3.6	32.6	4.0	1.67
3.1030	437.	97.	3.6	32.6	4.1	1.68
3.1130	437.	97.	3.6	32.4	4.0	1.67
3.1230	436.	97.	3.7	32.4	4.0	1.66
3.1330	437.	87.	3.6	32.4	4.1	1.67
3.1430	436.	87.	3.7	30.8	4.3	1.60
3.1530	436.	77.	3.6	32.0	4.2	1.65
3.1630	437.	77.	3.6	27.9	4.2	1.47
3.1730	436.	77.	3.6	27.3	4.2	1.44
3.1830	436.	67.	3.7	27.0	4.3	1.44
3.1930	437.	67.	3.7	27.1	4.3	1.44
3.2030	436.	67.	3.7	27.2	4.3	1.45
3.2130	454.	47.	3.7	28.6	4.4	1.51
3.2230	436.	47.	3.7	35.6	4.3	1.83
3.2330	436.	47.	3.7	36.8	4.3	1.89
4.0030	437.	47.	3.7	37.1	2.5	1.82
4.0130	419.	67.	3.7	37.0	2.6	1.82
4.0230	437.	47.	3.7	37.5	2.6	1.85
4.0330	428.	47.	3.7	38.6	2.6	1.91
4.0430	437.	47.	3.7	38.0	2.2	1.84
4.0530	437.	47.	3.7	37.4	2.0	1.82
4.0630	428.	47.	3.7	34.1	2.2	1.68
4.0730	428.	47.	3.7	34.3	1.7	1.67
4.0830	428.	47.	3.7	34.4	1.7	1.68
4.0930	428.	47.	3.7	33.1	1.9	1.62
4.1030	428.	47.	3.7	36.0	2.1	1.75
4.1130	428.	47.	3.7	35.2	3.1	1.76
4.1230	419.	47.	3.7	34.7	3.1	1.73
4.1330	419.	57.	3.7	33.7	3.1	1.69
4.1430	420.	57.	3.7	33.7	3.2	1.68
4.1530	420.	58.	3.7	33.2	3.1	1.66
4.1630	432.	58.	3.7	33.3	3.0	1.67
4.1730	423.	67.	3.7	32.8	3.0	1.64
4.1830	431.	67.	3.7	32.1	3.1	1.62
4.1930	431.	67.	3.7	35.6	3.0	1.78

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
4.2030	431.	67.	3.7	38.4	3.1	1.90
4.2130	433.	57.	3.7	37.8	3.6	1.90
4.2230	432.	57.	3.7	37.6	3.7	1.89
4.2330	423.	57.	3.7	37.5	3.6	1.88
5.0030	425.	57.	3.7	37.5	2.9	1.85
5.0130	425.	57.	3.7	37.3	2.6	1.84
5.0230	468.	58.	3.7	36.9	2.7	1.80
5.0330	467.	58.	3.7	35.9	2.9	1.79
5.0430	468.	58.	3.7	35.9	2.9	1.77
5.0530	467.	48.	3.7	35.6	2.8	1.75
5.0630	477.	48.	3.7	35.2	2.9	1.73
5.0730	477.	48.	3.7	36.0	2.7	1.76
5.0830	468.	48.	3.7	36.8	2.3	1.80
5.0930	468.	48.	3.7	38.1	2.4	1.86
5.1030	462.	48.	3.7	38.0	2.2	1.84
5.1130	471.	48.	3.7	38.0	2.4	1.85
5.1230	482.	48.	3.7	38.7	2.4	1.89
5.1330	487.	48.	3.7	36.7	2.5	1.79
5.1430	479.	48.	3.7	36.7	2.2	1.79
5.1530	470.	48.	3.7	35.7	2.2	1.74
5.1630	478.	58.	3.7	31.7	2.2	1.55
5.1730	478.	48.	3.7	31.4	1.9	1.53
5.1830	477.	68.	3.7	31.9	1.9	1.55
5.1930	478.	69.	3.6	31.2	2.0	1.52
5.2030	478.	58.	3.6	31.1	2.0	1.51
5.2130	476.	58.	3.6	31.9	1.9	1.54
5.2230	477.	48.	3.6	31.9	2.0	1.54
5.2330	468.	48.	3.7	32.0	2.0	1.55
6.0030	477.	48.	3.6	31.9	2.0	1.54
6.0130	477.	48.	3.6	30.2	2.0	1.47
6.0230	477.	48.	3.6	28.6	1.9	1.40
6.0330	468.	48.	3.7	26.1	1.7	1.27
6.0430	468.	48.	3.7	26.4	1.7	1.28
6.0530	476.	48.	3.7	23.1	1.7	1.14
6.0630	477.	48.	3.6	23.5	1.7	1.15
6.0730	476.	47.	3.6	23.6	1.7	1.15
6.0830	476.	47.	3.6	23.6	1.6	1.15
6.0930	474.	47.	3.6	23.9	1.7	1.18
6.1030	474.	37.	3.6	27.3	1.6	1.32
6.1130	474.	37.	3.7	27.1	1.9	1.33

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
6.1230	447.	97.	3.7	27.4	1.6	1.33
6.1330	448.	87.	3.7	27.0	1.9	1.32
6.1430	434.	91.	3.7	27.0	1.6	1.31
6.1530	434.	91.	3.7	27.7	1.6	1.34
6.1630	439.	91.	3.7	27.3	1.9	1.33
6.1730	430.	77.	3.7	27.7	1.6	1.33
6.1830	440.	57.	3.7	30.7	1.9	1.49
6.1930	474.	37.	3.7	26.8	1.9	1.30
6.2030	474.	37.	3.7	27.7	1.9	1.35
6.2130	472.	47.	3.6	25.8	1.9	1.26
6.2230	466.	47.	3.6	31.4	1.9	1.53
6.2330	464.	57.	3.7	30.6	1.9	1.47
7.0030	455.	47.	3.7	31.1	2.7	1.54
7.0130	455.	47.	3.7	30.7	2.6	1.51
7.0230	455.	47.	3.7	32.2	2.7	1.59
7.0330	447.	47.	3.7	31.3	2.5	1.54
7.0430	446.	47.	3.7	32.0	2.7	1.58
7.0530	445.	47.	3.7	31.3	2.5	1.54
7.0630	446.	47.	3.7	31.7	2.2	1.55
		STONE CHANGE				
7.0730	446.	47.	3.7	28.5	2.5	1.41
7.0830	446.	47.	3.9	27.4	2.7	1.37
7.0930	447.	47.	3.7	27.4	2.7	1.36
7.1030	447.	47.	3.7	27.3	2.5	1.35
7.1130	456.	37.	3.7	26.8	2.2	1.31
7.1230	456.	37.	3.7	27.4	2.5	1.35
7.1330	456.	37.	3.7	25.9	2.5	1.29
7.1430	454.	47.	3.7	26.2	2.5	1.30
7.1530	454.	47.	3.7	29.0	2.5	1.43
7.1630	457.	47.	3.7	28.7	1.9	1.38
7.1730	456.	47.	3.7	29.1	3.2	1.46
7.1830	456.	47.	3.7	30.4	2.2	1.47
7.1930	456.	47.	3.7	30.3	2.2	1.48
7.2030	456.	47.	3.7	30.8	1.6	1.46
7.2130	446.	47.	3.7	30.3	3.2	1.52
7.2230	445.	47.	3.7	30.6	3.0	1.52
7.2330	470.	47.	3.7	30.0	2.7	1.47
8.0030	472.	47.	3.7	29.3	2.6	1.44
8.0130	448.	47.	3.7	29.6	2.7	1.46
8.0230	447.	47.	3.7	29.6	2.7	1.46

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
8.0330	448.	47.	3.7	29.1	2.6	1.43
8.0430	447.	47.	3.7	29.1	2.7	1.43
8.0530	447.	47.	3.7	28.6	2.6	1.40
8.0630	446.	47.	3.7	28.6	2.7	1.41
8.0730	446.	47.	3.7	28.6	2.7	1.41
8.0830	454.	47.	3.7	28.0	2.3	1.36
SHUT DOWN AT 8.0830 FOR 21 HOURS						
9.0530	421.	127.	3.6	29.5	2.9	1.46
9.0630	437.	108.	3.6	32.6	2.6	1.59
9.0730	439.	108.	3.7	28.0	2.3	1.38
9.0830	423.	106.	3.7	36.7	2.4	1.79
9.0930	410.	96.	3.7	27.6	2.4	1.37
9.1030	386.	105.	3.7	31.1	2.4	1.54
		STONE CHANGE				
9.1130	437.	105.	3.7	30.8	2.7	1.54
9.1230	472.	75.	3.7	28.8	2.1	1.42
9.1330	472.	48.	3.7	28.7	2.6	1.43
9.1430	506.	46.	3.7	28.5	2.6	1.44
9.1530	488.	46.	3.6	28.7	2.7	1.45
9.1630	471.	7.	3.5	28.0	2.9	1.43
9.1730	507.	7.	3.5	28.3	3.2	1.44
9.1830	490.	7.	3.5	28.5	3.4	1.47
9.1930	448.	7.	3.6	29.4	3.4	1.51
9.2030	451.	7.	3.5	29.2	3.1	1.49
9.2130	452.	7.	3.5	29.4	3.0	1.49
9.2230	452.	7.	3.5	29.3	3.0	1.49
9.2330	452.	7.	3.4	29.1	3.0	1.47
10.0030	460.	7.	3.4	29.6	3.1	1.50
10.0130	461.	7.	3.4	29.4	3.1	1.49
10.0230	451.	7.	3.4	29.4	3.0	1.49
10.0330	451.	7.	3.4	29.5	3.0	1.49
10.0430	460.	7.	3.4	29.6	3.2	1.51
10.0530	461.	7.	3.4	29.4	3.0	1.49
10.0630	460.	7.	3.4	29.7	2.9	1.50
10.0730	461.	7.	3.4	29.3	3.1	1.49
10.0830	460.	7.	3.4	29.9	3.1	1.51
10.0930	451.	7.	3.4	29.6	3.0	1.50

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
10.1030	476.	7.	3.4	29.1	3.1	1.48
10.1130	471.	7.	3.4	28.6	3.2	1.47
10.1230	467.	7.	3.4	29.2	3.3	1.50
10.1330	458.	7.	3.4	28.4	3.2	1.46
10.1430	467.	7.	3.4	28.2	3.3	1.45
10.1530	466.	7.	3.4	27.8	2.9	1.42
10.1630	457.	7.	3.4	27.6	3.7	1.45
10.1730	459.	7.	3.4	28.0	3.6	1.46
10.1830	457.	7.	3.4	28.2	3.4	1.46
10.1930	466.	7.	3.4	28.2	3.5	1.46
10.2030	466.	7.	3.4	28.4	3.6	1.48
10.2130	466.	7.	3.4	28.3	3.9	1.49
10.2230	466.	7.	3.4	28.7	3.8	1.50
10.2330	466.	7.	3.4	28.4	4.1	1.50
11.0030	466.	7.	3.4	28.4	4.5	1.51
11.0130	466.	7.	3.4	28.3	4.6	1.51
11.0230	465.	7.	3.5	28.4	4.8	1.3
11.0330	465.	7.	3.5	28.5	4.5	1.52
11.0430	465.	7.	3.5	28.9	5.1	1.56
11.0530	466.	7.	3.5	28.7	4.9	1.56
11.0630	466.	7.	3.5	28.6	5.4	1.56
11.0730	467.	7.	3.5	28.5	4.8	1.53
11.0830	477.	7.	3.4	28.3	5.3	1.55
11.0930	475.	7.	3.4	28.4	5.0	1.55
11.1030	465.	7.	3.4	27.9	6.3	1.57
11.1130	475.	7.	3.4	28.0	4.2	1.48
11.1230	482.	7.	3.4	27.4	7.0	1.59
11.1330	473.	7.	3.4	27.6	7.4	1.61
11.1430	482.	7.	3.4	27.4	7.3	1.60
11.1530	482.	7.	3.4	27.3	7.2	1.59
11.1630	481.	7.	3.4	27.4	7.6	1.61
11.1730	482.	7.	3.4	27.5	7.8	1.62
11.1830	481.	7.	3.4	27.4	3.9	1.43
11.1930	428.	7.	-	-	-	-
11.2030	482.	7.	3.4	29.9	3.0	1.51
11.2130	481.	59.	3.5	29.9	3.1	1.51
11.2230	438.	109.	3.5	30.0	2.8	1.50
11.2330	421.	69.	3.5	31.9	2.9	1.58
		STONE CHANGE				
12.0030	421.	69.	3.5	32.8	2.8	1.63

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	FLUE GAS	PILOT PROPANE	REGENERATOR AIR	NITROGEN	REGEN. VELOCITY M/SEC
12.0130	438.	68.	3.5	30.0	2.8	1.51
12.0230	439.	69.	3.5	29.8	2.8	1.49
12.0330	440.	59.	3.5	29.8	2.8	1.49
12.0430	439.	7.	3.5	29.9	2.0	1.46
12.0530	439.	7.	3.5	28.9	3.3	1.46
12.0630	440.	7.	3.5	28.3	2.5	1.41
12.0730	439.	7.	3.5	28.2	2.4	1.39
12.0830	438.	7.	3.5	28.2	2.4	1.39
12.0930	430.	7.	3.7	28.8	2.3	1.42
12.1030	422.	7.	3.7	30.7	2.5	1.53
12.1130	421.	7.	3.7	30.2	2.2	1.49
12.1230	420.	7.	3.7	30.0	2.7	1.50
12.1330	422.	7.	3.7	29.7	2.2	1.47
12.1430	423.	7.	3.7	29.5	2.4	1.47
12.1530	422.	7.	3.7	31.7	2.4	1.56
12.1630	431.	7.	3.7	34.1	2.7	1.69
12.1730	439.	7.	3.7	31.1	2.5	1.54
12.1830	449.	7.	3.7	29.5	2.9	1.48
12.1930	448.	7.	3.7	25.3	2.6	1.27
12.2030	448.	7.	3.7	28.2	2.7	1.42
12.2130	448.	7.	3.7	28.3	2.6	1.42
12.2230	448.	7.	3.7	28.0	2.5	1.40
12.2330	448.	7.	3.7	27.3	2.7	1.38
13.0030	448.	7.	3.7	27.1	2.9	1.37
13.0130	465.	7.	3.7	27.7	2.9	1.40
13.0230	465.	7.	3.7	27.1	1.6	1.32
13.0330	474.	7.	3.7	28.9	2.9	1.47
13.0430	474.	7.	3.7	28.9	5.6	1.58
13.0530	474.	7.	3.7	28.0	4.2	1.48
13.0630	465.	7.	3.7	29.2	3.3	1.49
13.0730	465.	7.	3.7	27.4	3.4	1.41
13.0830	474.	7.	3.7	28.0	3.1	1.43
13.0930	466.	7.	3.6	29.7	4.0	1.53
13.1030	473.	7.	3.6	27.7	4.5	1.48
13.1130	449.	7.	3.6	29.5	6.1	1.62
13.1230	448.	7.	3.6	27.7	5.4	1.52
13.1330	430.	7.	3.7	28.9	5.0	1.56
13.1430	431.	7.	3.6	29.2	5.3	1.58
13.1530	430.	7.	3.6	29.0	4.5	1.53
13.1630	430.	7.	3.6	28.9	4.3	1.52

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
13.1730	435.	7.	3.6	28.6	4.0	1.48
13.1830	435.	27.	3.6	28.5	3.6	1.46
13.1930	436.	28.	3.6	28.2	2.4	1.39
13.2030	435.	28.	3.6	28.4	4.8	1.52
13.2130	435.	27.	3.6	28.4	4.3	1.50
13.2230	435.	28.	3.6	28.4	4.3	1.49
13.2330	435.	27.	3.6	28.9	4.4	1.52
14.0030	444.	48.	3.5	27.4	3.6	1.42
14.0130	444.	48.	3.5	26.8	4.5	1.42
14.0230	435.	48.	3.5	27.1	3.7	1.41
14.0330	435.	48.	3.5	27.1	3.6	1.40
14.0430	418.	47.	3.7	30.1	4.0	1.54
14.0530	418.	77.	3.7	25.9	3.7	1.35
14.0630	418.	47.	3.7	28.6	3.7	1.47
14.0730	426.	47.	3.7	28.0	3.8	1.45
14.0830	426.	47.	3.7	28.0	3.7	1.45
14.0930	426.	47.	3.6	28.0	4.3	1.48
14.1030	416.	47.	3.7	28.5	4.4	1.50
14.1130	417.	47.	3.7	28.2	4.8	1.50
14.1230	408.	47.	3.7	28.2	4.4	1.48
14.1330	408.	47.	3.7	27.9	4.6	1.47
14.1430	416.	47.	3.7	28.2	4.2	1.46
14.1530	408.	47.	3.7	28.2	4.2	1.48
14.1630	408.	47.	3.7	28.0	3.6	1.43
14.1730	408.	47.	3.7	28.2	4.8	1.49
14.1830	408.	47.	3.7	28.2	4.7	1.48
14.1930	407.	47.	3.7	28.0	5.6	1.51
14.2030	408.	47.	3.7	28.3	6.2	1.55
14.2130	416.	47.	3.7	28.2	5.9	1.53
14.2230	416.	47.	3.7	28.2	6.0	1.54
14.2330	416.	47.	3.7	28.0	5.8	1.52
15.0030	416.	47.	3.7	28.9	8.3	1.67
15.0130	416.	47.	3.7	29.8	8.3	1.72
15.0230	416.	47.	3.7	30.1	6.6	1.65
15.0330	416.	47.	3.7	30.8	3.7	1.54
15.0430	425.	47.	3.7	28.9	4.0	1.47
15.0530	425.	47.	3.7	27.7	3.4	1.39
15.0630	416.	47.	3.7	28.0	3.0	1.39
15.0730	416.	47.	3.7	27.7	3.4	1.39
15.0830	416.	47.	3.7	27.7	3.0	1.38

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
15.0930	416.	57.	3.5	27.6	3.0	1.37
15.1030	424.	57.	3.5	27.6	3.2	1.37
15.1130	416.	57.	3.5	29.1	3.5	1.47
15.1230	424.	67.	3.5	34.9	5.2	1.82
15.1330	423.	67.	3.5	36.1	6.0	1.91
15.1430	423.	67.	3.5	36.1	6.6	1.94
15.1530	414.	67.	3.5	37.5	6.9	2.02
15.1630	406.	67.	3.5	38.8	9.2	2.19
15.1730	406.	57.	3.5	38.7	5.8	2.02
15.1830	406.	67.	3.5	41.4	6.2	2.16
15.1930	407.	67.	3.5	43.6	6.1	2.26
15.2030	406.	67.	3.5	38.4	6.8	2.06
15.2130	407.	67.	3.4	37.2	3.2	1.83
15.2230	398.	67.	3.4	37.0	5.0	1.90
15.2330	407.	67.	3.4	37.2	5.1	1.92
16.0030	407.	67.	3.4	34.1	5.0	1.77
16.0130	441.	47.	3.4	34.6	8.1	1.97
16.0230	442.	47.	3.4	31.3	6.6	1.75
16.0330	442.	47.	3.4	30.4	6.0	1.67
16.0430	442.	47.	3.5	30.0	7.7	1.73
16.0530	476.	47.	3.5	30.3	7.5	1.74
16.0630	442.	47.	3.5	30.9	7.0	1.74
16.0730	442.	47.	3.5	30.9	6.7	1.73
16.0830	441.	47.	3.4	30.8	5.9	1.68
16.0930	408.	47.	3.4	30.7	5.5	1.65
16.1030	407.	47.	3.4	30.5	4.9	1.61
16.1130	406.	58.	3.4	30.4	4.8	1.61
16.1230	398.	58.	3.4	30.4	4.5	1.59
16.1330	398.	47.	3.4	30.1	4.4	1.57
16.1430	398.	58.	3.4	29.8	4.6	1.57
16.1530	407.	47.	3.4	29.3	4.4	1.53
16.1630	387.	47.	3.4	29.3	4.1	1.51
16.1730	410.	47.	3.4	29.4	3.7	1.50
16.1830	411.	47.	3.4	28.5	4.2	1.48
16.1930	409.	47.	3.4	29.7	3.8	1.52
16.2030	392.	57.	3.4	29.9	3.9	1.53
16.2130	392.	57.	3.4	29.9	4.2	1.55
16.2230	392.	57.	3.5	29.7	4.2	1.54
16.2330	392.	57.	3.5	29.7	4.4	1.55
17.0030	389.	57.	3.5	29.5	5.3	1.58

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
17.0130	384.	57.	3.5	29.4	3.2	1.47
17.0230	384.	57.	3.5	29.4	4.1	1.51
17.0330	385.	57.	3.4	29.8	4.2	1.52
17.0430	384.	57.	3.4	29.8	4.0	1.52
17.0530	385.	57.	3.5	28.4	3.6	1.44
17.0630	385.	57.	3.5	28.3	3.5	1.43
17.0730	384.	57.	3.5	34.3	3.4	1.69
17.0830	385.	57.	3.5	28.2	3.0	1.39
17.0930	372.	57.	3.4	28.0	3.1	1.40
17.1030	381.	57.	3.4	28.1	3.8	1.43
17.1130	380.	57.	3.4	28.0	3.5	1.42
17.1230	380.	47.	3.4	27.7	3.6	1.41
17.1330	380.	57.	3.4	29.0	3.9	1.48
17.1430	380.	57.	3.4	30.8	4.0	1.56
17.1530	379.	7.	3.4	30.6	3.8	1.54
17.1630	372.	7.	3.4	30.7	3.4	1.53
17.1730	371.	7.	3.4	30.4	4.5	1.56
17.1830	380.	7.	3.4	30.4	3.9	1.54
17.1930	372.	7.	3.4	30.1	4.7	1.56
17.2030	371.	7.	3.4	29.9	4.5	1.54
17.2130	371.	7.	3.4	29.3	4.1	1.49

APPENDIX D - TABLE III

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
1.1230	3.9	4.2	4.9	1.05	6.0
1.1330	4.1	4.2	4.7	1.00	7.0
1.1430	4.0	4.2	5.0	1.05	5.2
1.1530	3.9	4.2	4.7	1.05	6.3
1.1630	3.6	3.2	4.9	1.05	6.0
1.1730	3.6	3.5	5.0	1.05	6.5
1.1830	3.9	3.5	5.0	1.05	6.2
1.1930	3.7	3.5	5.2	1.05	6.2
1.2030	3.7	3.4	5.1	1.05	5.5
1.2130	3.7	3.2	5.0	1.05	5.0
1.2230	3.7	3.4	5.0	1.00	4.7
1.2330	3.7	3.5	4.7	1.10	4.2
2.0030	4.1	3.4	4.5	1.10	4.5
2.0130	4.0	3.2	4.4	1.10	5.0
2.0230	4.1	3.1	4.4	1.10	4.1
2.0330	4.0	3.2	4.4	1.00	4.2
2.0430	4.1	3.2	4.5	1.10	4.2
2.0530	4.1	3.2	4.5	1.00	4.5
2.0630	4.1	3.2	4.2	1.00	4.2
2.0730	4.0	3.2	4.5	1.05	5.0
2.0830	4.1	3.1	4.5	1.05	5.0
2.0930	4.1	3.1	4.7	1.05	5.2
2.1030	3.9	3.1	4.9	1.00	5.5
2.1130	4.0	3.1	4.7	1.05	5.4
2.1230	4.0	3.0	4.6	1.00	5.2
2.1330	3.9	3.1	4.6	1.00	6.2
2.1430	4.0	3.1	4.5	1.05	5.5
2.1530	4.0	3.1	4.5	1.05	5.5
2.1630	4.0	3.1	4.5	1.05	5.5
2.1730	3.9	3.1	4.5	1.05	5.5
2.1830	4.0	3.2	4.5	1.00	5.5
2.1930	4.1	3.1	4.6	1.05	5.5
2.2030	4.2	3.1	4.7	1.05	5.5
2.2130	4.1	3.2	4.7	1.10	5.5
2.2230	4.2	3.2	5.0	1.10	5.5
2.2330	4.1	3.2	5.0	1.00	5.5
3.0030	4.2	3.4	5.0	1.00	7.5
3.0130	4.1	3.2	5.0	1.00	7.3
3.0230	4.1	3.2	5.1	1.00	7.5
3.0330	4.0	3.2	5.2	1.00	7.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
3.0430	4.0	3.2	5.2	1.05	7.5
3.0530	4.1	3.4	5.2	1.00	7.5
3.0630	4.2	3.5	5.2	1.00	7.7
3.0730	4.1	3.6	5.2	1.05	7.6
3.0830	4.2	3.5	5.5	1.05	7.5
3.0930	4.2	3.5	5.4	1.05	7.6
3.1030	4.2	3.5	5.4	1.00	7.5
3.1130	4.1	3.5	5.4	1.05	7.6
3.1230	4.1	3.5	5.6	1.00	7.6
3.1330	4.1	3.4	5.7	1.00	7.7
3.1430	4.1	3.4	5.7	1.00	7.7
3.1530	4.1	3.4	5.7	1.00	8.0
3.1630	4.1	3.2	5.8	1.00	8.1
3.1730	4.1	3.4	5.8	1.00	8.1
3.1830	4.0	3.2	6.0	1.00	8.0
3.1930	4.0	3.2	6.1	1.00	8.0
3.2030	4.0	3.2	6.2	1.00	8.1
3.2130	4.0	3.2	6.1	1.00	8.5
3.2230	3.9	3.2	6.1	1.00	8.0
3.2330	4.0	3.1	5.7	1.00	7.7
4.0030	4.0	3.0	6.2	1.00	7.7
4.0130	4.1	3.2	6.1	1.05	8.0
4.0230	4.0	3.1	6.1	1.00	9.0
4.0330	4.0	3.1	6.2	1.05	8.5
4.0430	4.0	3.1	6.2	1.05	8.7
4.0530	4.1	3.1	6.2	1.05	8.2
4.0630	4.1	3.0	6.2	1.05	8.2
4.0730	4.1	3.1	6.2	1.05	8.0
4.0830	4.1	3.1	6.2	1.05	8.5
4.0930	4.1	3.0	6.0	1.05	8.5
4.1030	4.0	3.0	6.0	1.05	8.5
4.1130	3.9	2.9	6.1	1.00	8.6
4.1230	4.0	3.0	6.1	1.00	8.6
4.1330	4.1	3.1	6.1	1.00	9.0
4.1430	4.2	3.1	6.1	1.00	9.0
4.1530	4.2	3.1	6.1	1.00	8.7
4.1630	4.2	3.1	6.1	1.00	8.7
4.1730	4.5	3.2	$ 6.1	1.00	8.7
4.1830	4.4	3.2	6.2	1.05	9.0
4.1930	4.4	3.1	6.2	1.05	8.7

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
4.2030	4.2	3.1	6.2	1.00	8.7
4.2130	4.4	3.1	6.2	1.05	8.7
4.2230	4.4	3.1	6.3	1.05	8.7
4.2330	4.1	3.1	6.3	1.10	8.7
5.0030	4.2	3.1	6.3	1.10	8.7
5.0130	4.3	3.0	6.3	1.10	8.7
5.0230	4.6	3.4	6.5	1.10	9.2
5.0330	4.6	3.4	6.5	1.05	8.7
5.0430	4.6	3.4	6.5	1.05	8.7
5.0530	4.7	3.4	6.5	1.05	8.7
5.0630	4.7	3.4	6.5	1.05	8.5
5.0730	4.7	3.4	6.5	1.05	8.5
5.0830	4.7	3.4	6.5	1.05	8.5
5.0930	4.7	3.4	6.5	1.05	8.7
5.1030	4.7	3.4	6.5	1.05	8.7
5.1130	4.7	3.4	6.5	1.05	8.7
5.1230	4.7	3.4	6.2	1.05	8.7
5.1330	4.9	3.1	6.2	1.05	9.5
5.1430	4.7	3.1	6.2	1.05	9.0
5.1530	4.6	3.1	6.2	1.05	9.2
5.1630	4.7	3.0	6.3	1.05	9.0
5.1730	4.6	3.0	6.2	1.05	9.0
5.1830	4.6	3.0	6.2	1.05	9.0
5.1930	4.7	3.0	6.2	1.05	9.0
5.2030	4.9	3.1	6.2	1.05	8.7
5.2130	4.9	3.2	6.0	1.05	8.7
5.2230	4.9	3.2	6.0	1.05	8.7
5.2330	5.0	3.2	6.0	1.05	8.7
6.0030	5.0	3.2	6.0	1.05	8.7
6.0130	5.0	3.2	5.8	1.10	7.5
6.0230	4.8	3.2	5.8	1.10	8.0
6.0330	4.9	3.2	5.7	1.10	8.2
6.0430	4.9	3.2	5.8	1.10	8.2
6.0530	5.0	3.2	5.8	1.10	8.5
6.0630	5.1	3.2	5.8	1.10	8.5
6.0730	5.2	3.5	5.5	1.05	8.5
6.0830	5.2	3.4	5.5	1.10	8.5
6.0930	5.2	3.5	5.6	1.10	8.5
6.1030	5.1	3.5	5.7	1.10	8.5
6.1130	5.1	3.4	5.7	1.10	8.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
6.1230	5.0	3.7	5.7	1.10	8.5
6.1330	5.1	3.7	5.7	1.10	8.5
6.1430	5.1	3.5	5.6	1.05	8.5
6.1530	5.1	3.6	5.5	1.10	8.5
6.1630	5.1	3.6	5.5	1.10	8.5
6.1730	5.1	3.6	5.6	1.10	8.5
6.1830	5.0	3.6	5.6	1.10	8.5
6.1930	5.4	3.7	5.6	1.10	7.5
6.2030	5.5	3.5	5.6	1.00	8.0
6.2130	5.5	3.6	5.5	1.00	8.2
6.2230	5.6	3.7	5.5	1.00	8.0
6.2330	5.6	3.7	5.4	1.00	8.0
7.0030	5.6	3.6	5.2	1.00	8.2
7.0130	5.6	3.5	5.2	1.00	8.5
7.0230	5.6	3.6	5.4	1.00	8.5
7.0330	5.6	3.6	5.4	1.00	8.7
7.0430	5.7	3.6	5.1	1.00	8.2
7.0530	5.6	3.6	5.1	1.00	8.5
7.0630	5.7	3.6	5.1	1.00	8.7
		STONE CHANGE			
7.0730	5.7	3.5	5.2	1.00	8.7
7.0830	5.7	3.6	5.2	1.00	8.7
7.0930	6.0	3.6	5.2	1.10	8.2
7.1030	5.7	3.5	5.2	1.10	8.2
7.1130	5.7	3.5	5.2	1.10	8.2
7.1230	5.8	3.5	5.2	1.00	8.5
7.1330	5.8	3.6	5.2	1.00	8.5
7.1430	5.8	3.7	5.2	1.00	8.5
7.1530	6.0	3.7	5.2	1.03	8.2
7.1630	5.8	3.7	5.5	1.00	8.2
7.1730	5.6	3.7	5.5	1.00	8.2
7.1830	5.8	3.6	5.2	1.00	8.2
7.1930	5.8	3.7	5.2	1.00	8.2
7.2030	6.0	3.7	5.5	1.00	8.2
7.2130	6.0	3.6	5.4	1.05	8.2
7.2230	6.0	3.7	5.5	1.00	8.5
7.2330	6.3	3.7	5.4	1.00	8.5
8.0030	6.3	3.7	5.2	1.05	8.5
8.0130	6.1	3.7	5.1	1.05	8.5
8.0230	6.1	3.6	5.1	1.00	8.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
8.0330	6.3	3.7	5.2	1.00	8.0
8.0430	6.3	3.7	5.2	1.00	8.0
8.0530	6.3	3.7	5.0	1.00	8.0
8.0630	6.3	3.7	5.0	1.00	8.5
8.0730	6.4	3.6	5.1	1.00	8.5
8.0830	6.5	3.7	5.2	1.00	8.5
SHUT DOWN AT 8.0830 FOR 21 HOURS					
9.0530	3.7	4.6	4.2	1.00	7.5
9.0630	3.8	4.6	4.5	1.00	7.5
9.0730	3.7	4.5	4.5	1.00	7.7
9.0830	3.6	4.2	4.5	1.00	7.7
9.0930	3.5	3.9	4.6	1.00	7.5
9.1030	3.6	4.1	4.7	1.00	7.2
STONE CHANGE					
9.1130	3.6	4.3	4.7	1.00	7.5
9.1230	3.6	3.9	4.9	1.00	7.5
9.1330	3.6	3.6	5.2	1.00	7.5
9.1430	3.7	3.7	5.2	1.00	7.7
9.1530	3.9	3.9	5.2	1.00	8.0
9.1630	3.9	3.4	5.2	1.00	7.5
9.1730	4.0	3.9	5.0	1.00	7.5
9.1830	4.0	3.5	5.0	1.00	7.5
9.1930	3.7	3.0	4.7	1.00	7.2
9.2030	3.7	3.0	5.0	1.00	7.2
9.2130	3.7	2.9	4.9	1.00	7.3
9.2230	3.7	2.9	5.0	1.00	7.3
9.2330	3.7	3.0	4.9	1.00	7.2
10.0030	3.9	3.0	4.7	1.00	7.2
10.0130	3.9	3.0	4.7	1.00	7.2
10.0230	3.9	3.0	4.7	1.00	7.3
10.0330	3.9	3.0	4.7	1.00	7.3
10.0430	3.9	2.9	4.7	1.00	7.2
10.0530	3.9	2.9	4.7	1.00	7.2
10.0630	3.9	3.0	5.0	1.00	7.2
10.0730	3.8	3.0	4.7	1.00	7.2
10.0830	3.9	3.0	4.6	1.00	7.2
10.0930	3.9	3.0	4.7	1.00	7.2

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
10.1030	4.0	2.9	5.0	1.00	7.5
10.1130	4.0	2.2	5.0	0.95	8.0
10.1230	3.9	2.2	5.2	0.95	8.2
10.1330	3.9	2.2	5.4	1.00	8.2
10.1430	3.9	2.2	5.5	1.00	8.5
10.1530	3.9	2.0	5.5	1.00	8.5
10.1630	3.9	2.0	5.5	1.00	8.5
10.1730	3.9	2.1	5.6	1.00	8.5
10.1830	3.9	2.2	5.5	1.00	8.2
10.1930	4.0	2.2	5.5	1.00	8.2
10.2030	4.0	2.2	5.5	1.00	8.2
10.2130	4.0	2.2	5.5	1.00	8.2
10.2230	4.0	2.2	5.4	1.00	8.1
10.2330	4.0	2.2	5.4	1.00	8.2
11.0030	4.0	3.2	5.5	1.00	8.2
11.0130	4.0	3.2	5.5	1.00	8.0
11.0230	4.0	3.2	5.5	1.00	8.0
11.0330	4.0	3.2	5.5	1.00	7.7
11.0430	4.2	3.2	5.5	1.00	8.0
11.0530	4.1	3.4	5.5	1.00	8.0
11.0630	4.1	3.4	5.6	1.00	8.1
11.0730	4.1	3.4	5.5	1.00	8.2
11.0830	4.1	3.2	5.5	1.00	8.2
11.0930	4.0	3.2	5.6	1.00	8.0
11.1030	4.0	3.2	5.6	1.00	8.7
11.1130	4.0	3.2	5.6	1.00	9.0
11.1230	4.1	3.4	5.7	1.00	8.5
11.1330	4.1	3.4	6.0	1.00	8.5
11.1430	4.1	3.4	5.8	1.00	8.5
11.1530	4.1	3.5	5.7	1.00	8.5
11.1630	4.1	3.5	5.7	1.00	8.7
11.1730	4.1	3.4	5.7	1.00	8.7
11.1830	4.1	3.4	5.7	1.00	9.0
11.1930	4.2	3.4	5.7	1.00	8.7
11.2030	4.2	3.4	5.7	1.00	8.7
11.2130	4.2	3.7	5.7	1.00	8.7
11.2230	4.1	3.7	5.0	1.00	7.0
11.2330	4.0	3.9	4.5	0.95	7.0
		STONE CHANGE			
12.0030	4.0	3.4	4.7	1.00	7.0

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
12.0130	3.7	3.0	4.9	0.95	7.0
12.0230	3.7	3.0	4.5	0.95	6.7
12.0330	3.7	3.0	4.4	0.95	6.7
12.0430	3.7	2.9	4.4	0.95	6.5
12.0530	3.7	3.0	4.4	0.95	6.5
12.0630	3.7	3.0	4.4	0.95	6.7
12.0730	3.9	3.0	4.2	0.95	6.7
12.0830	3.7	2.9	4.4	0.95	6.7
12.0930	3.7	2.9	4.6	0.95	7.0
12.1030	3.6	2.7	4.7	0.95	7.2
12.1130	3.7	2.7	5.0	0.95	7.5
12.1230	3.7	2.7	5.1	0.95	7.5
12.1330	3.7	2.7	5.2	1.00	7.0
12.1430	3.7	2.7	5.5	1.00	7.0
12.1530	3.7	2.7	5.6	1.00	7.0
12.1630	3.7	2.9	5.6	1.00	7.2
12.1730	3.9	2.9	5.8	1.00	8.5
12.1830	3.9	3.0	5.8	0.95	8.7
12.1930	3.9	3.1	6.0	0.95	8.5
12.2030	3.9	3.0	6.1	0.95	8.2
12.2130	3.9	3.0	6.0	0.90	8.2
12.2230	3.7	3.0	5.8	0.90	8.2
12.2330	3.9	3.0	5.8	0.90	8.2
13.0030	4.0	3.0	5.8	0.95	8.2
13.0130	4.1	3.2	5.7	0.90	8.2
13.0230	4.2	3.4	5.6	0.90	8.2
13.0330	4.1	3.2	5.5	0.95	8.2
13.0430	4.1	3.4	5.6	1.00	7.2
13.0530	4.1	3.2	5.6	1.00	7.5
13.0630	4.1	3.2	5.6	1.00	7.5
13.0730	4.1	3.4	5.6	1.00	7.5
13.0830	4.2	3.4	5.6	0.90	7.5
13.0930	4.2	3.4	5.6	0.90	8.0
13.1030	4.2	3.2	5.5	0.90	8.5
13.1130	4.0	3.0	5.5	0.90	7.7
13.1230	4.0	3.0	5.2	0.95	7.5
13.1330	3.9	2.7	5.2	0.95	7.0
13.1430	3.7	2.7	5.2	0.95	7.0
13.1530	3.9	2.9	5.2	0.95	7.0
13.1630	3.9	2.9	5.2	0.95	7.0

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
13.1730	3.9	2.9	5.4	0.95	7.2
13.1830	4.0	2.9	5.4	0.95	7.5
13.1930	4.0	2.9	5.2	0.95	7.5
13.2030	4.0	2.9	5.2	0.95	7.5
13.2130	4.1	2.9	5.2	0.95	7.2
13.2230	4.2	2.9	5.1	1.00	7.2
13.2330	4.2	2.9	5.2	0.95	7.2
14.0030	4.1	3.0	5.1	1.00	7.2
14.0130	4.1	3.0	5.5	1.05	7.7
14.0230	4.2	3.0	5.5	1.00	7.7
14.0330	4.2	3.0	5.5	1.00	7.7
14.0430	4.6	3.6	5.5	1.00	8.0
14.0530	4.5	3.7	5.4	1.05	8.0
14.0630	4.6	3.6	5.4	1.00	8.0
14.0730	4.6	3.4	5.4	1.05	8.0
14.0830	4.6	3.4	5.4	1.00	8.0
14.0930	4.6	3.4	5.4	0.95	8.0
14.1030	4.6	3.5	5.2	1.00	8.0
14.1130	4.6	3.5	5.4	0.95	8.0
14.1230	4.6	3.5	5.2	1.00	8.0
14.1330	4.7	3.5	5.2	0.95	8.0
14.1430	4.7	3.5	5.2	0.95	8.0
14.1530	4.7	3.5	5.2	0.95	7.7
14.1630	4.7	3.5	5.2	1.00	7.5
14.1730	4.7	3.5	5.4	0.95	7.0
14.1830	4.6	3.4	5.5	1.00	7.2
14.1930	4.7	3.4	5.4	1.00	7.5
14.2030	4.9	3.5	5.2	1.00	7.2
14.2130	4.7	3.5	5.2	1.00	7.0
14.2230	4.7	3.4	5.2	1.00	7.0
14.2330	4.7	3.5	5.2	1.00	7.0
15.0030	4.7	3.5	5.2	1.00	7.2
15.0130	4.7	3.6	5.2	1.00	7.0
15.0230	4.7	3.5	5.2	1.00	7.0
15.0330	4.7	3.5	5.2	1.00	7.0
15.0430	4.7	3.4	5.2	1.00	7.0
15.0530	4.7	3.4	5.2	0.95	7.0
15.0630	4.7	3.4	5.2	1.00	7.2
15.0730	4.7	3.4	5.2	1.00	7.2
15.0830	4.7	3.2	5.2	1.00	7.5

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
15.0930	4.7	3.2	5.5	1.00	7.5
15.1030	4.7	3.4	5.5	0.95	7.5
15.1130	4.7	3.4	5.5	0.95	7.5
15.1230	4.7	3.2	5.5	1.00	7.5
15.1330	4.8	3.2	5.5	0.95	7.5
15.1430	4.8	3.2	5.4	0.95	7.5
15.1530	4.8	3.2	5.4	0.95	7.5
15.1630	4.7	3.1	5.4	0.95	7.5
15.1730	4.7	3.1	5.4	0.95	7.3
15.1830	4.7	3.2	5.4	0.95	7.5
15.1930	4.8	3.1	5.2	0.95	7.5
15.2030	5.0	3.2	5.2	0.95	7.5
15.2130	4.9	3.2	5.2	0.95	7.0
15.2230	5.1	3.2	5.0	0.95	7.0
15.2330	5.0	3.2	5.1	0.95	7.0
16.0030	5.0	3.2	5.2	0.95	7.0
16.0130	5.0	3.2	5.2	0.95	8.0
16.0230	4.9	2.9	5.2	0.95	8.5
16.0330	5.1	3.1	5.2	0.95	8.5
16.0430	5.1	3.1	5.2	0.95	8.7
16.0530	4.7	2.7	5.2	0.95	8.0
16.0630	4.7	2.7	5.2	0.95	8.2
16.0730	4.6	2.6	5.2	0.95	8.2
16.0830	4.7	2.5	5.4	1.00	8.2
16.0930	4.7	2.6	5.4	0.95	8.2
16.1030	4.7	2.6	5.4	0.95	8.2
16.1130	4.7	2.5	5.5	1.00	8.2
16.1230	4.7	2.5	5.5	1.00	7.7
16.1330	4.8	2.5	5.5	0.95	8.2
16.1430	4.8	2.5	5.6	0.95	8.5
16.1530	5.0	2.6	5.6	1.00	8.2
16.1630	4.9	2.4	5.6	1.00	8.2
16.1730	5.1	2.4	5.6	0.95	8.2
16.1830	5.1	2.5	5.6	0.95	8.2
16.1930	5.1	2.5	5.6	1.00	8.2
16.2030	5.1	2.4	5.6	0.95	8.2
16.2130	5.1	2.4	5.6	0.95	8.2
16.2230	5.1	2.4	5.6	0.95	8.2
16.2330	5.2	2.4	5.4	0.95	8.2
17.0030	5.1	2.4	5.5	0.95	8.2

RUN 7: PRESSURES

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	GASIFIER P. KILOPASCALS DISTRIB. D.P.	GASIFIER P. KILOPASCALS BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
17.0130	5.5	2.5	5.5	0.95	8.5
17.0230	5.5	2.5	5.5	0.95	8.7
17.0330	5.6	2.5	5.4	1.00	8.7
17.0430	5.6	2.5	5.4	1.00	8.7
17.0530	5.7	2.5	5.4	0.95	8.2
17.0630	5.8	2.5	5.2	0.95	8.2
17.0730	5.7	2.5	5.2	1.00	8.2
17.0830	5.8	2.5	5.4	0.95	8.2
17.0930	5.9	2.5	5.4	1.00	8.2
17.1030	5.6	2.4	5.4	1.05	8.2
17.1130	5.7	2.4	5.4	1.00	5.0
17.1230	5.8	2.4	5.4	1.00	8.2
17.1330	5.8	2.2	5.4	1.00	8.2
17.1430	6.0	2.2	5.4	1.00	8.2
17.1530	6.0	2.2	5.4	1.00	8.5
17.1630	5.9	2.1	5.5	1.00	8.2
17.1730	6.0	2.1	5.5	0.95	8.2
17.1830	6.3	2.2	5.4	1.00	8.2
17.1930	6.4	2.1	5.4	1.00	8.0
17.2030	6.6	2.2	5.4	1.00	8.2
17.2130	6.7	2.1	5.2	1.00	8.5

APPENDIX D: TABLE IV.

RUN 7: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
1.1230	22.2	1.61	47.	24.6	0.	25.2	24.7
1.1330	20.2	1.63	48.	23.4	0.	4.7	3.2
1.1430	57.6	1.59	48.	22.5	0.	15.2	14.5
1.1530	31.2	1.63	45.	23.4	0.16	25.1	24.6
1.1630	51.2	1.37	47.	21.1	1.08	33.8	28.8
1.1730	47.6	1.39	48.	22.0	1.18	39.9	40.6
1.1830	47.1	1.43	48.	20.9	0.69	39.9	38.8
1.1930	49.9	1.41	50.	20.8	0.79	43.6	46.4
1.2030	51.2	1.37	49.	21.6	0.79	51.3	44.0
1.2130	51.4	1.37	48.	21.7	0.75	43.9	39.8
1.2230	53.0	1.38	50.	22.6	0.72	69.5	63.1
1.2330	51.5	1.36	43.	22.5	0.75	54.5	47.8
2.0030	48.2	1.34	41.	22.2	0.95	69.8	64.7
2.0130	53.6	1.30	40.	21.7	1.04	58.4	52.4
2.0230	59.0	1.33	40.	22.2	0.92	59.9	54.2
2.0330	58.0	1.33	44.	22.3	1.12	49.5	52.7
2.0430	61.4	1.35	41.	22.7	1.24	51.4	53.5
2.0530	62.4	1.32	45.	22.6	1.34	65.9	66.4
2.0630	65.8	1.34	43.	23.2	1.38	77.8	72.1
2.0730	66.8	1.32	43.	22.6	1.37	48.2	49.8
2.P830	68.3	1.30	43.	23.2	1.34	45.5	46.4
2.0930	69.9	1.28	45.	22.4	1.24	63.4	61.7
2.1030	73.3	1.26	49.	22.3	1.44	65.4	60.3
2.1130	73.6	1.24	45.	22.7	1.35	66.8	66.6
2.1230	74.8	1.22	46.	22.4	1.51	75.2	69.9
2.1330	77.5	1.24	46.	22.5	1.34	74.1	73.3
2.1430	76.5	1.22	43.	22.2	0.78	69.5	68.5
2.1530	73.4	1.23	43.	22.5	0.95	64.5	63.9
2.1630	69.4	1.22	43.	22.4	1.04	79.2	79.2
2.1730	66.9	1.22	43.	22.1	1.00	69.4	73.7
2.1830	66.9	1.25	45.	22.6	0.86	62.5	56.5
2.1930	65.5	1.24	44.	22.3	0.95	69.6	70.6
2.2030	66.2	1.24	45.	22.3	0.98	59.3	64.2
2.2130	68.6	1.29	43.	22.3	1.04	67.2	66.9
2.2230	70.7	1.25	46.	22.2	1.01	63.0	72.0
2.2330	72.7	1.23	50.	22.0	1.04	70.5	74.3
3.0030	73.1	1.30	50.	22.8	1.07	66.5	72.5
3.0130	70.2	1.27	50.	23.3	1.15	89.3	83.9
3.0230	72.7	1.27	52.	22.2	1.00	64.0	65.3
3.0330	74.0	1.26	52.	22.2	1.07	68.6	69.9

APPENDIX D: TABLE IV.

RUN 7: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
3.0430	75.3	1.22	50.	22.3	1.07	75.4	73.7
3.0530	75.7	1.25	53.	21.9	0.98	72.9	74.0
3.0630	76.3	1.29	53.	22.8	1.53	66.7	68.3
3.0730	75.5	1.29	50.	23.3	1.69	66.2	73.9
3.0830	74.1	1.30	53.	23.4	1.86	67.6	75.7
3.0930	74.1	1.30	52.	23.3	1.73	64.9	73.3
3.1030	75.5	1.31	54.	23.3	1.60	66.8	73.9
3.1130	75.3	1.31	52.	23.3	1.76	66.0	67.6
3.1230	75.8	1.29	57.	23.3	2.31	71.2	77.9
3.1330	77.4	1.28	58.	23.4	2.25	68.3	72.1
3.1430	78.8	1.27	58.	23.1	2.07	82.7	80.3
3.1530	78.6	1.23	58.	23.5	2.24	75.3	85.1
3.1630	80.0	1.25	59.	23.4	2.29	78.1	75.3
3.1730	79.4	1.22	59.	23.2	2.57	77.7	73.9
3.1830	80.5	1.23	60.	23.2	1.95	86.3	80.3
3.1930	81.1	1.23	62.	23.3	2.21	79.3	70.5
3.2030	81.4	1.20	63.	23.2	2.05	74.5	68.3
3.2130	81.8	1.24	62.	24.2	2.06	65.1	61.4
3.2230	81.8	1.20	62.	23.2	1.17	75.8	90.9
3.2330	82.2	1.19	58.	23.1	0.84	77.6	87.1
4.0030	82.4	1.18	63.	22.6	0.76	77.2	86.4
4.0130	81.6	1.18	59.	21.3	0.84	71.5	77.2
4.0230	80.1	1.18	62.	22.0	0.90	83.4	85.9
4.0330	79.5	1.14	60.	21.3	0.76	98.3	89.7
4.0430	77.9	1.19	60.	21.9	0.83	89.0	82.2
4.0530	75.5	1.17	60.	21.8	0.77	72.9	70.0
4.0630	78.2	1.11	60.	21.3	0.67	69.7	69.2
4.0730	80.6	1.15	60.	21.4	0.77	75.2	73.2
4.0830	80.2	1.15	60.	21.3	0.95	75.4	75.2
4.0930	79.1	1.14	58.	21.4	0.77	82.5	76.3
4.1030	78.7	1.15	58.	21.3	0.61	75.0	75.4
4.1130	77.3	1.15	62.	21.4	0.77	96.0	77.5
4.1230	75.6	1.11	62.	21.2	0.78	79.3	73.0
4.1330	75.8	1.22	62.	21.3	0.78	80.7	75.3
4.1430	76.5	1.15	62.	21.3	0.84	78.4	71.4
4.1530	76.6	1.15	62.	21.3	0.75	78.9	77.8
4.1630	77.7	1.30	62.	22.0	0.69	77.8	81.3
4.1730	76.3	1.16	62.	21.5	0.90	93.5	74.9
4.1830	76.9	1.21	60.	22.0	1.00	68.6	69.8
4.1930	78.2	1.22	60.	22.0	0.90	70.4	77.1

APPENDIX D: TABLE IV.
RUN 7: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
4.2030	79.2	1.20	63.	21.9	0.78	63.0	78.9
4.2130	80.9	1.19	60.	22.0	0.81	81.6	80.7
4.2230	81.6	1.18	61.	21.8	0.90	81.3	75.4
4.2330	81.7	1.16	58.	21.5	0.84	104.4	82.8
5.0030	84.5	1.15	58.	21.5	0.84	83.1	78.4
5.0130	81.9	1.14	58.	21.5	0.84	73.3	72.0
5.0230	79.6	1.18	60.	23.5	0.99	95.6	84.9
5.0330	79.5	1.19	62.	23.6	1.00	69.5	79.2
5.0430	79.8	1.19	62.	23.6	0.96	64.0	75.8
5.0530	80.9	1.18	62.	23.4	1.02	75.2	78.9
5.0630	81.0	1.19	62.	23.4	0.97	71.8	78.2
5.0730	80.3	1.21	62.	23.4	0.81	76.4	83.7
5.0830	79.6	1.19	62.	22.9	0.99	80.9	79.4
5.0930	79.6	1.19	62.	22.9	0.81	77.5	79.6
5.1030	79.6	1.16	62.	22.6	0.69	75.7	82.9
5.1130	79.6	1.18	62.	23.0	0.81	70.2	80.1
5.1230	79.3	1.21	60.	23.6	0.78	74.8	84.3
5.1330	79.3	1.22	60.	23.8	0.84	75.4	80.6
5.1430	78.9	1.20	60.	23.4	0.84	77.1	80.6
5.1530	79.0	1.16	60.	23.0	0.88	72.6	81.5
5.1630	78.8	1.18	61.	23.4	0.91	85.2	78.5
5.1730	79.1	1.17	60.	23.4	0.79	110.8	80.8
5.1830	79.4	1.17	60.	23.3	0.57	82.4	76.0
5.1930	79.0	1.17	60.	23.4	0.51	98.3	86.2
5.2030	78.3	1.20	60.	23.5	0.45	81.0	72.6
5.2130	78.3	1.20	58.	23.3	0.36	90.5	80.7
5.2230	78.3	1.21	58.	23.4	0.45	110.5	101.6
5.2330	74.5	1.18	58.	23.0	0.45	66.1	68.4
6.0030	74.9	1.20	58.	23.4	0.45	102.3	98.1
6.0130	74.2	1.20	54.	23.3	0.48	83.1	67.5
6.0230	73.5	1.20	54.	23.2	0.54	77.0	71.7
6.0330	74.1	1.18	53.	22.7	0.48	80.7	67.5
6.0430	74.1	1.18	54.	22.7	0.51	83.9	68.5
6.0530	74.4	1.20	54.	23.0	0.54	84.3	61.6
6.0630	74.8	1.20	54.	23.1	0.45	79.0	59.3
6.0730	73.7	1.24	53.	23.2	0.51	83.9	64.1
6.0830	73.7	1.30	50.	23.2	0.69	86.2	68.2
6.0930	74.4	1.26	51.	23.0	0.72	103.4	72.1
6.1030	74.8	1.38	53.	23.2	0.60	90.6	79.0
6.1130	75.8	1.23	53.	23.0	0.45	89.7	79.4

APPENDIX D: TABLE IV.

RUN 7: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
6.1230	73.1	1.32	53.	21.8	0.33	87.3	75.4
6.1330	71.4	1.30	53.	22.0	0.42	91.8	74.9
6.1430	71.0	1.26	54.	21.3	0.48	99.5	73.3
6.1530	71.0	1.27	50.	21.2	0.51	101.2	77.2
6.1630	69.3	1.28	50.	21.5	0.48	94.2	80.7
6.1730	68.3	1.21	51.	20.9	0.69	88.3	76.4
6.1830	58.6	1.19	51.	22.0	0.57	104.0	84.9
6.1930	68.3	1.33	51.	23.1	0.45	88.7	71.9
6.2030	69.9	1.31	57.	23.1	0.57	84.8	66.2
6.2130	69.2	1.31	55.	23.0	0.48	88.4	68.0
6.2230	68.5	1.32	55.	22.8	0.57	75.3	79.6
6.2330	67.6	1.34	54.	22.6	0.51	71.9	65.5
7.0030	66.3	1.28	53.	22.2	0.39	76.4	72.2
7.0130	67.2	1.31	53.	22.6	0.61	72.8	71.5
7.0230	66.8	1.28	54.	21.9	0.59	89.0	75.6
7.0330	66.7	1.28	54.	21.9	0.60	72.5	68.3
7.0430	66.4	1.30	52.	21.9	0.45	80.7	71.7
7.0530	64.4	1.28	52.	21.8	0.48	69.7	71.1
7.0630	63.9	1.26	52.	21.7	0.21	71.8	61.9
STONE CHANGE							
7.0730	63.1	1.23	53.	21.7	1.01	68.4	61.9
7.0830	63.9	1.22	53.	21.6	1.19	88.1	64.4
7.0930	65.4	1.24	48.	21.8	0.90	82.0	65.3
7.1030	66.6	1.23	48.	21.6	0.77	95.8	69.2
7.1130	66.5	1.25	48.	23.2	0.88	81.7	68.7
7.1230	66.6	1.25	53.	22.0	0.89	82.6	67.7
7.1330	66.8	1.17	53.	22.1	0.87	96.8	69.6
7.1430	66.8	1.32	53.	22.0	0.77	102.0	64.3
7.1530	66.3	1.32	51.	22.1	0.93	97.5	66.8
7.1630	66.2	1.33	55.	22.6	1.00	92.2	65.0
7.1730	66.3	1.32	55.	22.1	0.95	93.7	64.3
7.1830	66.6	1.36	53.	22.2	0.71	87.9	71.2
7.1930	67.6	1.32	53.	22.3	0.72	71.3	62.5
7.2030	68.4	1.32	55.	22.2	0.83	88.3	67.2
7.2130	69.8	1.28	52.	22.0	0.72	85.9	70.5
7.2230	68.5	1.26	55.	21.8	0.63	84.4	69.9
7.2330	68.4	1.34	54.	23.1	0.51	79.4	65.7
8.0030	68.7	1.35	50.	23.2	0.63	74.1	64.9
8.0130	67.7	1.29	49.	22.0	0.69	77.6	66.3
8.0230	67.2	1.29	52.	21.9	0.72	77.6	66.1

APPENDIX D: TABLE IV.

RUN 7: DESULPHURISATION PERFORMANCE PAGE 5 OF 10

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
8.0330	68.0	1.30	53.	22.0	0.78	74.9	67.6
8.0430	68.4	1.29	53.	21.9	0.78	73.4	65.9
8.0530	68.1	1.29	50.	21.9	0.90	80.5	61.2
8.0630	67.3	1.27	50.	21.9	1.47	78.0	67.4
8.0730	67.0	1.25	52.	21.9	1.74	77.9	64.3
8.0830	69.2	1.24	53.	22.2	1.89	79.7	63.9
SHUT DOWN AT 8.0830 FOR 21 HOURS							
9.0530	52.5	1.38	43.	23.2	1.96	2.9	2.8
9.0630	53.2	1.35	45.	22.5	2.09	46.8	52.0
9.0730	54.0	1.38	45.	23.4	1.81	53.1	53.0
9.0830	58.0	1.33	45.	22.8	1.83	53.3	69.6
9.0930	58.3	1.25	46.	22.2	1.71	53.6	50.9
9.1030	61.4	1.22	48.	21.0	2.23	46.4	50.3
STONE CHANGE							
9.1130	63.7	1.32	48.	23.3	2.07	54.2	56.0
9.1230	64.7	1.32	49.	24.4	3.26	54.8	52.2
9.1330	68.2	1.32	53.	24.1	3.72	64.6	56.8
9.1430	71.3	1.30	53.	25.7	3.46	66.1	61.5
9.1530	71.7	1.26	53.	24.8	3.53	69.5	55.0
9.1630	71.4	1.14	53.	23.5	2.70	79.9	74.3
9.1730	72.0	1.25	50.	25.5	0.96	73.7	76.3
9.1830	70.4	1.21	50.	23.3	1.33	66.8	61.3
9.1930	71.3	1.08	48.	21.8	1.29	70.9	63.6
9.2030	72.2	1.09	50.	21.9	1.45	77.7	63.7
9.2130	71.8	1.08	49.	22.0	1.51	73.8	63.4
9.2230	71.3	1.10	50.	22.0	1.39	74.5	66.7
9.2330	70.4	1.09	49.	22.0	1.42	78.1	66.4
10.0030	71.5	1.12	48.	22.4	1.24	70.4	66.5
10.0130	69.5	1.12	48.	22.4	1.32	76.6	67.0
10.0230	69.3	1.09	48.	22.0	1.54	71.8	66.3
10.0330	69.3	1.09	48.	22.0	1.45	76.4	70.4
10.0430	69.3	1.11	48.	22.3	1.47	73.1	65.2
10.0530	69.3	1.11	48.	22.4	1.48	71.3	64.4
10.0630	69.7	1.12	50.	22.3	1.44	73.2	66.6
10.0730	70.7	1.11	48.	22.4	1.54	72.0	64.5
10.0830	70.8	1.11	46.	22.4	1.57	70.4	64.0
10.0930	70.1	1.07	48.	21.9	2.07	63.0	58.5

APPENDIX D: TABLE IV.

RUN 7: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
10.1030	70.8	1.13	50.	23.0	3.33	74.7	65.9
10.1130	74.8	1.12	53.	23.1	3.28	76.3	66.3
10.1230	76.9	1.10	56.	22.4	3.27	77.5	66.4
10.1330	75.6	1.07	54.	22.3	3.43	74.9	65.0
10.1430	75.2	1.09	55.	22.7	3.40	62.2	53.7
10.1530	78.3	1.10	55.	22.6	3.21	74.6	63.1
10.1630	78.9	1.08	55.	22.2	3.37	75.5	61.5
10.1730	80.0	1.06	57.	22.3	3.64	89.0	65.2
10.1830	81.0	1.09	55.	22.2	1.96	91.0	77.5
10.1930	82.4	1.12	55.	22.7	1.32	90.7	77.6
10.2030	82.0	1.11	55.	22.6	1.41	83.2	77.0
10.2130	81.2	1.12	55.	22.7	1.35	89.3	78.5
10.2230	80.9	1.12	54.	22.7	1.51	84.8	76.9
10.2330	79.6	1.11	54.	22.6	1.47	85.9	76.9
11.0030	78.6	1.11	55.	22.7	1.29	83.4	74.0
11.0130	77.6	1.12	55.	22.7	1.30	88.3	80.4
11.0230	78.2	1.12	55.	22.6	1.29	72.5	64.6
11.0330	77.5	1.12	55.	22.6	1.45	67.3	60.2
11.0430	77.5	1.12	55.	22.7	1.51	70.4	64.3
11.0530	78.6	1.11	55.	22.6	1.56	85.2	79.5
11.0630	79.2	1.11	57.	22.7	1.51	79.4	72.6
11.0730	79.1	1.12	55.	22.7	1.51	76.9	70.5
11.0830	78.8	1.14	55.	23.2	1.51	80.3	73.6
11.0930	79.5	1.13	57.	23.1	1.57	80.2	74.8
11.1030	78.8	1.10	57.	22.6	1.57	88.2	78.3
11.1130	78.3	1.13	57.	23.1	1.48	77.2	69.9
11.1230	79.3	1.15	58.	23.6	1.58	93.9	79.8
11.1330	79.0	1.12	60.	23.0	1.69	93.7	82.5
11.1430	79.0	1.15	59.	23.5	1.48	89.1	77.6
11.1530	78.6	1.16	58.	23.5	1.21	93.5	78.0
11.1630	79.0	1.15	58.	23.5	1.21	98.4	87.5
11.1730	79.6	1.15	58.	23.5	1.27	102.7	92.5
11.1830	79.3	1.15	58.	23.5	1.57	75.1	67.8
11.1930	79.3	1.15	58.	23.5	1.48	86.6	84.2
11.2030	79.6	1.15	58.	23.6	1.46	90.3	88.0
11.2130	78.8	1.20	58.	23.7	1.43	76.2	75.2
11.2230	76.9	1.34	50.	22.0	0.82	71.6	70.4
11.2330	72.9	1.20	48.	21.0	0.06	69.6	73.1
STONE CHANGE							
12.0030	73.4	1.23	48.	21.7	1.87	68.4	72.0

APPENDIX D: TABLE IV.

RUN 7: DESULPHURISATION PERFORMANCE PAGE 7 OF 10

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
12.0130	77.5	1.09	52.	23.1	2.24	68.8	70.3
12.0230	79.7	1.04	48.	23.1	2.42	16.2	14.7
12.0330	78.7	1.04	46.	23.2	2.59	65.5	69.6
12.0430	80.0	1.04	46.	23.2	2.95	74.7	76.2
12.0530	81.3	1.04	46.	23.2	3.03	78.2	79.3
12.0630	81.9	1.04	46.	23.2	2.75	77.4	74.3
12.0730	81.3	1.04	45.	23.2	2.79	78.3	75.6
12.0830	83.2	1.03	46.	23.3	3.22	70.3	69.4
12.0930	82.8	1.01	49.	23.0	2.96	77.9	72.5
12.1030	82.5	0.99	50.	22.5	2.75	63.4	70.9
12.1130	82.8	0.99	53.	22.5	2.08	79.7	80.1
12.1230	83.5	0.99	54.	22.4	2.42	87.3	81.9
12.1330	83.5	0.99	53.	22.6	2.52	82.2	75.7
12.1430	85.9	0.99	55.	22.6	2.42	77.0	74.6
12.1530	84.6	0.99	57.	22.6	2.82	77.6	79.5
12.1630	83.9	1.00	57.	23.2	2.87	74.8	81.6
12.1730	83.8	1.02	59.	23.5	3.38	63.4	68.3
12.1830	82.6	1.05	62.	24.0	3.48	72.3	74.4
12.1930	82.1	1.05	64.	24.0	3.20	72.6	64.1
12.2030	82.4	1.04	65.	24.1	2.70	73.7	71.7
12.2130	83.6	1.04	67.	23.5	2.70	74.5	73.5
12.2230	83.4	1.04	66.	24.3	3.45	79.7	75.7
12.2330	83.4	1.05	66.	23.0	2.97	73.6	69.5
13.0030	83.3	1.05	62.	23.9	2.51	74.0	72.6
13.0130	81.5	1.10	64.	24.9	2.63	74.3	74.4
13.0230	78.6	1.11	63.	24.9	2.82	73.0	73.7
13.0330	79.5	1.14	58.	25.2	2.00	76.1	82.7
13.0430	81.3	1.14	57.	25.3	2.24	63.1	65.5
13.0530	82.0	1.14	57.	25.3	2.14	74.5	76.6
13.0630	80.7	1.12	57.	24.8	2.92	70.3	74.2
13.0730	80.5	1.11	57.	24.6	2.29	69.9	69.9
13.0830	81.9	1.14	63.	25.0	2.99	59.0	58.7
13.0930	80.0	1.11	63.	24.9	2.38	66.5	68.1
13.1030	80.9	1.14	62.	25.7	2.45	62.9	63.6
13.1130	82.2	1.08	62.	23.7	1.38	72.1	72.6
13.1230	82.5	1.08	56.	24.0	1.13	77.5	78.9
13.1330	81.3	1.04	56.	23.0	0.75	73.5	75.9
13.1430	81.0	1.04	56.	23.1	0.89	74.3	75.9
13.1530	80.4	1.04	56.	23.0	0.88	70.1	72.2
13.1630	80.4	1.04	56.	23.2	1.03	67.7	70.3

APPENDIX D: TABLE IV.
RUN 7: DESULPHURISATION PERFORMANCE PAGE 8 OF 10

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
13.1730	80.9	1.05	57.	23.8	1.29	66.7	69.5
13.1830	80.7	1.06	57.	23.0	1.34	70.3	66.8
13.1930	79.7	1.06	56.	23.3	1.46	64.9	63.2
13.2030	79.1	1.06	56.	23.2	1.05	75.4	74.5
13.2130	79.1	1.07	56.	23.1	1.05	74.1	76.2
13.2230	78.8	1.06	52.	23.1	1.22	74.1	76.1
13.2330	79.1	1.08	56.	23.2	1.02	74.2	77.9
14.0030	79.1	1.11	52.	23.4	0.91	72.7	71.5
14.0130	78.4	1.11	53.	23.7	1.05	70.6	67.7
14.0230	77.5	1.09	55.	23.2	1.05	68.4	65.9
14.0330	79.1	1.08	55.	23.2	0.95	70.6	71.1
14.0430	78.7	1.00	55.	22.2	0.91	70.7	78.9
14.0530	76.1	1.18	52.	22.4	0.98	71.4	68.6
14.0630	76.5	1.10	54.	22.2	1.08	74.0	74.8
14.0730	77.0	1.12	52.	22.8	0.94	76.8	74.4
14.0830	77.7	1.13	54.	22.9	0.95	72.3	70.3
14.0930	76.7	1.13	57.	22.7	0.91	74.0	71.5
14.1030	76.2	1.10	53.	22.3	0.98	74.0	69.7
14.1130	76.2	1.10	57.	22.2	0.94	79.1	74.1
14.1230	75.9	1.08	53.	21.8	0.88	73.0	72.1
14.1330	76.1	1.08	56.	21.8	0.95	75.4	72.5
14.1430	76.3	1.10	56.	22.2	0.95	72.9	66.6
14.1530	76.4	1.08	56.	21.8	0.91	74.6	70.2
14.1630	76.1	1.08	53.	21.7	0.94	71.5	63.1
14.1730	77.7	1.07	57.	21.8	1.15	70.9	64.0
14.1830	80.3	1.07	55.	21.8	1.49	75.4	62.5
14.1930	77.9	1.06	54.	21.8	1.11	78.9	72.0
14.2030	76.6	1.07	53.	21.8	0.94	78.9	70.1
14.2130	73.7	1.09	53.	22.1	1.07	79.4	68.9
14.2230	75.7	1.09	53.	22.1	1.11	77.1	68.7
14.2330	75.2	1.10	53.	22.5	0.82	84.5	74.0
15.0030	74.0	1.10	53.	22.4	0.85	87.2	73.0
15.0130	74.9	1.09	53.	22.1	0.98	89.4	83.1
15.0230	74.7	1.09	53.	22.1	1.25	73.6	69.6
15.0330	74.4	1.19	53.	22.1	1.11	64.3	63.3
15.0430	74.3	1.22	53.	22.9	1.19	68.3	61.7
15.0530	75.5	1.20	56.	22.8	1.29	68.0	59.7
15.0630	76.6	1.18	53.	22.2	1.18	67.0	59.4
15.0730	77.2	1.19	53.	22.4	1.08	68.1	60.0
15.0830	77.7	1.17	53.	22.4	1.53	70.9	61.3

APPENDIX D: TABLE IV.
RUN 7: DESULPHURISATION PERFORMANCE

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
15.0930	77.1	1.12	55.	22.2	1.31	67.1	60.0
15.1030	77.4	1.21	58.	22.7	1.28	71.8	61.1
15.1130	79.1	1.19	58.	22.3	1.42	78.8	66.0
15.1230	80.4	1.21	55.	22.7	1.39	82.5	88.3
15.1330	78.6	1.20	58.	22.7	1.15	78.1	92.3
15.1430	79.6	1.18	57.	22.7	1.35	78.5	89.0
15.1530	79.7	1.16	57.	22.2	1.28	79.1	87.5
15.1630	79.9	1.13	57.	21.8	1.19	77.8	89.5
15.1730	78.2	1.15	57.	21.7	1.28	73.8	84.3
15.1830	77.7	1.13	57.	21.7	1.25	70.7	82.1
15.1930	76.9	1.16	56.	21.8	1.12	65.2	84.8
15.2030	76.8	1.12	56.	21.7	1.45	66.1	70.1
15.2130	77.2	1.14	56.	21.8	1.22	57.0	67.1
15.2230	76.5	1.11	53.	21.5	1.19	66.8	74.4
15.2330	78.5	1.12	54.	21.8	1.39	63.5	72.0
16.0030	78.5	1.12	56.	21.8	1.15	65.7	68.8
16.0130	77.7	1.21	56.	24.0	1.10	76.6	92.3
16.0230	79.9	1.10	56.	23.6	1.26	79.6	83.9
16.0330	78.5	1.10	56.	23.6	1.29	78.2	81.9
16.0430	76.3	1.10	56.	23.6	0.81	85.6	93.1
16.0530	77.0	1.18	56.	25.5	0.98	87.9	85.3
16.0630	77.1	1.09	56.	23.5	1.11	87.8	80.5
16.0730	78.4	1.08	56.	23.6	1.05	95.2	81.1
16.0830	78.6	1.07	54.	23.7	0.99	96.0	79.8
16.0930	78.8	0.99	57.	21.8	1.29	94.3	76.0
16.1030	78.5	0.99	57.	21.7	1.22	80.0	72.2
16.1130	78.6	0.98	55.	21.7	1.15	92.2	75.5
16.1230	79.5	0.96	55.	21.3	1.15	87.8	70.5
16.1330	78.8	0.95	58.	21.2	1.15	89.1	73.5
16.1430	78.4	0.96	60.	21.2	0.91	90.5	75.7
16.1530	78.4	1.00	57.	21.7	0.91	89.1	73.9
16.1630	78.1	0.95	57.	20.5	1.21	86.2	72.1
16.1730	78.1	1.01	60.	21.8	1.28	87.9	69.8
16.1830	78.8	1.00	60.	21.7	1.00	92.7	73.1
16.1930	79.1	0.99	57.	21.6	0.91	89.1	70.8
16.2030	77.6	0.96	60.	20.8	1.11	85.2	73.0
16.2130	77.0	0.96	60.	20.8	1.25	79.5	74.6
16.2230	74.0	0.97	60.	20.8	0.91	73.5	73.3
16.2330	72.3	0.97	57.	20.8	1.04	73.4	75.4
17.0030	72.0	0.97	58.	20.7	0.98	74.0	77.1

APPENDIX D: TABLE IV.

RUN 7: DESULPHURISATION PERFORMANCE PAGE 10 OF 10

DAY.HOUR	SULPHUR REMOVAL %	GAS VEL. M/S	G-BED DEPTH CENTIM	AIR/ FUEL % ST.	CAO/S RATIO MOL.	% CAS TO CAO	REGEN. S OUT % OF FED
17.0130	71.6	1.02	58.	20.7	0.98	67.7	67.9
17.0230	70.9	1.02	58.	20.6	0.84	68.2	69.9
17.0330	74.6	1.01	54.	20.6	0.71	82.8	81.4
17.0430	72.3	1.07	54.	20.5	0.87	69.4	66.8
17.0530	73.0	1.01	57.	20.7	0.98	71.8	64.1
17.0630	73.6	1.02	56.	20.6	0.91	76.1	63.9
17.0730	74.4	1.01	53.	20.6	1.11	78.8	80.4
17.0830	74.7	1.01	57.	20.6	1.28	75.2	59.2
17.0930	75.4	0.92	54.	19.8	1.01	84.7	67.4
17.1030	74.7	0.94	52.	20.3	0.61	90.3	69.9
17.1130	73.6	0.93	54.	20.2	0.94	93.2	67.4
17.1230	73.6	0.91	54.	20.1	1.01	83.6	67.4
17.1330	73.7	0.91	54.	20.1	1.07	91.0	71.5
17.1430	73.6	0.91	54.	20.3	0.78	85.3	76.0
17.1530	73.1	0.88	54.	20.1	1.45	84.9	72.3
17.1630	75.7	0.86	55.	20.5	1.97	75.5	69.0
17.1730	76.5	0.86	58.	20.8	2.03	82.7	74.7
17.1830	76.3	0.88	54.	21.1	2.09	83.1	79.1
17.1930	76.4	0.86	54.	20.8	1.32	80.3	78.6
17.2030	76.4	0.86	54.	20.7	1.07	81.6	74.1
17.2130	76.4	0.85	53.	20.8	1.32	72.4	68.5

APPENDIX D: TABLE V.

RUN 7: GAS COMPOSITIONS

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	FLUE GAS SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
1.1230	4.6	11.9	12.3	1008.	1.10	0.5	2.7	16.9	16.5	3.27	3.24
1.1330	5.0	11.2	12.0	1008.	6.00	0.2	0.3	15.5	16.4	3.92	3.24
1.1430	3.5	12.4	13.2	606.	3.00	0.4	1.5	15.0	15.8	4.02	3.72
1.1530	5.0	11.4	12.1	908.	2.00	0.8	2.7	15.0	16.1	4.02	3.48
1.1630	2.0	13.2	14.3	757.	1.50	1.0	3.5	16.0	16.3	3.10	3.26
1.1730	2.5	13.0	13.9	757.	1.20	1.0	4.2	16.5	16.8	2.84	2.93
1.1830	1.5	13.8	14.6	808.	1.60	1.2	4.0	16.1	15.9	3.01	3.62
1.1930	1.4	13.5	14.7	768.	1.00	1.3	4.6	16.1	16.1	2.93	3.40
1.2030	1.2	13.8	14.9	753.	1.00	1.6	5.4	15.9	17.0	3.10	2.75
1.2130	1.4	13.8	14.7	743.	1.00	0.6	5.0	16.3	17.3	2.76	2.59
1.2230	1.0	14.1	15.0	733.	1.00	2.1	7.4	16.3	17.1	2.68	2.73
1.2330	1.2	14.1	14.8	747.	1.00	2.1	5.6	16.5	17.4	2.60	2.56
2.0030	2.0	13.5	14.3	767.	1.00	2.2	7.4	16.8	17.5	2.44	2.50
2.0130	2.0	13.8	14.3	687.	1.00	1.7	6.2	16.8	17.4	2.44	2.60
2.0230	2.0	13.8	14.3	606.	1.00	2.3	6.2	16.8	17.5	2.44	2.55
2.0330	2.0	13.8	14.3	621.	1.00	1.4	5.4	16.8	17.5	2.36	2.55
2.0430	2.0	13.8	14.3	571.	1.00	1.1	5.8	17.0	17.5	2.28	2.54
2.0530	2.0	13.8	14.3	556.	1.00	2.0	7.0	17.0	17.8	2.12	2.33
2.0630	2.0	13.5	14.3	505.	1.00	3.3	7.6	17.0	17.8	2.12	2.27
2.0730	2.0	13.5	14.3	490.	1.00	1.0	5.4	17.2	17.8	2.12	2.31
2.0830	2.1	13.5	14.2	465.	1.00	0.8	5.2	17.3	17.9	2.12	2.19
2.0930	2.0	13.5	14.3	445.	1.00	2.3	6.6	17.3	18.0	2.12	2.13
2.1030	2.0	13.8	14.3	394.	1.00	2.4	6.8	17.2	17.9	2.12	2.22
2.1130	2.0	13.8	14.3	389.	0.90	1.7	7.4	17.2	18.2	2.12	2.06
2.1230	1.9	13.8	14.3	374.	0.90	3.0	7.6	17.4	18.1	2.12	2.09
2.1330	1.9	13.5	14.4	334.	1.00	2.1	8.0	17.5	18.1	2.04	2.04
2.1430	1.9	13.5	14.4	349.	0.90	2.0	7.6	17.6	18.3	1.96	1.93
2.1530	1.9	13.8	14.4	394.	1.00	1.6	7.2	17.7	18.4	1.96	1.89

2.1630	1.9	13.8	14.4	455.	0.90	1.9	8.8	17.7	18.4	1.88	1.87
2.1730	2.0	13.8	14.3	490.	1.00	1.9	7.6	17.7	18.3	1.81	1.93
2.1830	2.0	13.8	14.3	490.	1.30	1.2	7.0	17.1	17.8	2.28	2.35
2.1930	2.0	13.8	14.3	510.	1.00	1.3	7.4	17.2	18.1	2.20	2.13
2.2030	2.0	13.8	14.3	500.	1.00	1.9	6.4	17.2	18.2	2.20	2.05
2.2130	2.0	13.8	14.3	465.	1.00	3.1	6.2	16.9	17.6	2.44	2.46
2.2230	2.0	13.8	14.3	434.	0.20	3.1	6.2	17.1	17.8	2.44	2.31
2.2330	2.0	13.8	14.3	404.	0.20	3.0	7.0	17.0	17.7	2.44	2.36
3.0030	1.7	13.8	14.5	404.	0.	3.1	6.6	17.2	18.0	2.28	2.11
3.0130	3.0	13.2	13.5	419.	0.	5.2	7.4	17.2	18.1	2.28	2.16
3.0230	1.5	14.1	14.7	414.	0.	2.6	7.0	17.2	17.9	2.28	2.27
3.0330	1.3	14.1	14.8	399.	0.	2.9	7.4	17.2	17.7	2.60	2.34
3.0430	1.3	14.1	14.8	379.	0.	3.1	7.4	17.2	18.0	2.12	2.14
3.0530	1.5	14.1	14.7	369.	0.	2.9	7.4	17.0	17.9	2.28	2.23
3.0630	1.8	13.8	14.4	354.	0.	2.4	7.0	17.1	18.0	2.28	2.13
3.0730	1.6	13.8	14.6	369.	0.	2.3	7.0	17.2	17.8	2.20	2.30
3.0830	1.9	13.8	14.3	384.	0.	2.0	7.4	17.2	18.0	2.12	2.14
3.0930	1.9	13.8	14.4	384.	0.	1.8	7.2	17.4	18.1	2.12	2.09
3.1030	1.9	13.8	14.4	364.	0.	2.2	7.2	17.4	18.1	2.12	2.09
3.1130	2.0	13.5	14.3	364.	2.00	1.5	6.6	17.4	18.2	2.12	2.03
3.1230	1.9	13.8	14.3	359.	0.	2.4	7.6	17.3	18.2	2.20	2.04
3.1330	2.0	13.5	14.3	334.	0.	2.8	7.0	17.9	18.4	1.81	1.88
3.1430	2.0	13.5	14.3	313.	0.	3.5	8.0	18.0	18.3	1.65	1.90
3.1530	2.1	13.5	14.2	313.	0.	2.1	8.2	18.0	18.6	1.81	1.71
3.1630	2.1	13.8	14.2	293.	0.	2.3	8.2	18.4	18.4	1.49	1.90
3.1730	2.0	13.8	14.3	303.	0.	2.2	8.2	18.6	18.6	1.49	1.74
3.1830	2.0	13.5	14.3	288.	0.	2.6	8.8	18.6	18.6	1.33	1.71
3.1930	2.1	13.5	14.2	278.	0.	3.0	7.8	18.6	18.6	1.33	1.72
3.2030	2.1	13.5	14.2	273.	0.	2.5	7.6	18.8	18.8	1.24	1.57
3.2130	2.4	13.5	14.0	263.	0.	2.5	6.6	19.0	19.0	1.16	1.45
3.2230	2.4	13.5	14.0	263.	0.	2.6	8.0	19.0	19.0	1.16	1.45
3.2330	2.1	13.5	14.2	263.	0.	3.9	7.4	19.0	19.0	1.16	1.43
4.0030	1.8	13.8	14.5	263.	0.	4.0	7.8	19.2	19.2	0.99	1.29
4.0130	2.0	13.8	14.3	273.	0.	3.9	7.2	18.0	18.6	1.96	1.73
4.0230	1.8	13.8	14.5	298.	0.	5.0	7.8	19.0	19.0	1.33	1.44
4.0330	1.8	14.1	14.5	308.	0.	6.8	7.8	19.2	19.2	1.16	1.32

RUN 7: GAS COMPOSITIONS

DAY.HOUR	FLUE GAS O2 %	FLUE GAS CO2 VOL % ANAL	FLUE GAS CO2 VOL % CALC	SO2 PPM	REGENERATOR GAS O2 %	REGENERATOR GAS CO2 %	REGENERATOR GAS SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	GASIFIER INLET GAS O2 VOL % CALC	GASIFIER INLET GAS CO2 VOL % ANAL	GASIFIER INLET GAS CO2 VOL % CALC
4.0430	2.0	13.8	14.3	328.	0.	6.4	7.4	19.0	19.3	1.16	1.22
4.0530	2.0	13.8	14.3	364.	0.	5.4	6.6	19.2	19.3	1.08	1.22
4.0630	2.0	13.8	14.3	323.	0.	3.7	7.2	19.2	19.3	1.08	1.24
4.0730	2.0	13.8	14.3	288.	0.	4.3	7.6	19.2	19.3	1.08	1.24
4.0830	2.0	13.8	14.3	293.	0.	4.0	7.8	19.2	19.3	1.08	1.24
4.0930	1.8	13.8	14.5	313.	0.	4.7	8.0	19.2	19.3	1.08	1.24
4.1030	1.6	13.8	14.6	323.	0.	4.4	7.4	19.2	19.2	1.08	1.24
4.1130	1.7	13.8	14.5	344.	0.	6.9	7.2	19.2	19.3	1.08	1.24
4.1230	1.7	13.8	14.5	369.	0.	5.2	7.0	18.0	19.3	1.08	1.22
4.1330	1.5	13.8	14.7	369.	0.	4.7	7.4	18.2	19.0	1.81	1.43
4.1430	1.7	13.8	14.5	354.	0.	5.0	7.0	18.0	19.1	1.81	1.38
4.1530	1.6	13.5	14.6	354.	0.	3.9	7.8	18.0	19.1	1.81	1.34
4.1630	1.8	13.5	14.5	334.	0.	3.1	8.2	18.2	19.2	1.65	1.27
4.1730	1.9	13.5	14.4	354.	0.	6.4	7.4	17.8	18.8	2.04	1.57
4.1830	1.9	13.5	14.4	344.	0.	2.6	7.4	18.4	18.7	1.57	1.63
4.1930	2.0	13.5	14.3	323.	0.	3.1	7.4	18.4	18.7	1.49	1.63
4.2030	2.0	13.5	14.3	308.	0.	2.0	7.2	18.6	18.7	1.41	1.63
4.2130	2.0	13.8	14.3	283.	0.	5.4	7.0	18.8	19.0	1.24	1.44
4.2230	2.0	13.8	14.3	273.	0.	5.9	6.6	18.8	19.0	1.16	1.45
4.2330	2.0	13.8	14.3	273.	0.	7.8	7.0	18.8	19.0	1.16	1.48
5.0030	1.8	13.8	14.4	233.	0.	5.9	7.0	18.8	18.9	1.08	1.54
5.0130	2.0	13.8	14.3	268.	0.	5.2	6.6	19.0	18.9	1.08	1.55
5.0230	2.0	13.8	14.3	303.	0.	6.7	7.6	20.0	19.0	0.34	1.49
5.0330	2.0	13.5	14.3	303.	0.	2.7	7.6	20.0	18.9	0.34	1.48
5.0430	2.4	13.2	14.0	293.	0.	1.9	7.4	20.0	19.0	0.34	1.45
5.0530	2.0	13.5	14.3	283.	0.	3.8	7.6	20.0	19.3	0.34	1.24
5.0630	1.9	13.5	14.4	283.	0.	2.8	7.8	20.2	19.3	0.34	1.22
5.0730	2.0	13.8	14.3	293.	0.	3.1	8.2	20.0	19.3	0.23	1.24

5.0830	2.0	13.8	14.3	303.	0.	5.0	7.6	20.0	19.3	0.23	1.27
5.0930	2.0	13.5	14.3	303.	0.	4.7	7.4	20.0	19.3	0.13	1.22
5.1030	2.0	13.5	14.3	303.	0.	3.9	7.8	20.2	19.3	0.13	1.24
5.1130	2.0	13.5	14.3	303.	0.	3.1	7.6	20.2	19.3	0.13	1.23
5.1230	2.0	13.5	14.3	308.	0.	3.7	7.8	20.2	19.3	0.23	1.23
5.1330	2.0	13.5	14.3	308.	0.	3.7	7.8	20.2	19.3	0.23	1.24
5.1430	2.0	13.5	14.3	313.	0.	4.1	7.8	20.2	19.2	0.23	1.27
5.1530	1.9	13.5	14.4	313.	0.	2.6	8.2	20.4	19.2	0.23	1.27
5.1630	1.8	13.8	14.5	318.	0.	4.1	8.6	20.4	18.9	0.13	1.51
5.1730	1.9	13.8	14.4	313.	0.	7.2	8.6	20.6	19.2	0.02	1.30
5.1830	1.6	13.5	14.6	313.	0.	4.1	8.4	20.6	18.5	0.02	1.72
5.1930	2.0	13.5	14.3	313.	0.	5.0	9.4	20.8	18.5	0.02	1.81
5.2030	2.0	13.5	14.3	323.	0.	4.1	8.2	20.2	18.9	0.23	1.52
5.2130	2.0	13.5	14.3	323.	0.	4.7	8.8	20.2	18.9	0.23	1.49
5.2230	2.0	13.5	14.3	323.	0.	5.1	10.7	20.2	19.2	0.23	1.25
5.2330	2.0	13.5	14.3	379.	0.	1.9	7.8	20.2	19.2	0.23	1.26
6.0030	2.0	13.2	14.3	374.	0.	4.3	10.5	20.2	19.3	0.23	1.21
6.0130	2.0	13.5	14.3	384.	0.	5.0	7.8	20.2	19.3	0.23	1.23
6.0230	2.0	13.2	14.3	394.	0.	2.2	9.0	20.2	19.3	0.23	1.21
6.0330	1.8	13.5	14.5	389.	0.	2.5	9.2	20.2	19.2	0.23	1.25
6.0430	1.8	13.2	14.5	389.	0.	3.1	9.2	20.2	19.2	0.23	1.23
6.0530	1.8	13.2	14.5	384.	0.	2.8	9.4	20.2	19.2	0.23	1.21
6.0630	1.8	13.2	14.5	379.	0.	2.5	9.0	20.2	19.2	0.23	1.21
6.0730	1.8	13.2	14.5	394.	0.	2.5	9.6	19.6	19.4	0.63	1.11
6.0830	1.8	13.5	14.5	394.	0.	2.1	10.3	19.0	19.4	1.41	1.13
6.0930	1.8	13.2	14.5	384.	0.	4.5	10.3	18.0	19.4	1.33	1.11
6.1030	1.5	13.2	14.7	384.	0.	3.0	10.3	18.2	19.7	1.16	0.89
6.1130	1.8	13.0	14.5	364.	0.	2.6	10.3	19.8	19.7	0.81	0.87
6.1230	1.3	13.5	14.8	415.	0.	3.0	9.9	17.0	18.0	2.60	2.03
6.1330	1.8	13.0	14.5	429.	0.	3.7	9.6	17.5	18.3	2.28	1.80
6.1430	1.9	13.0	14.4	435.	0.	5.2	9.4	17.2	18.2	2.44	1.92
6.1530	1.0	13.5	15.1	455.	0.	5.2	9.6	17.3	18.0	2.36	2.00
6.1630	1.2	13.2	14.9	475.	0.	3.3	10.3	17.5	18.1	2.28	1.94
6.1730	1.0	13.2	15.1	495.	0.	3.1	9.9	17.5	18.3	2.12	1.76
6.1830	6.1	13.2	11.2	483.	0.	5.7	9.4	17.5	19.5	2.12	1.29
6.1930	2.1	12.4	14.3	469.	0.	3.5	9.4	18.8	19.8	1.24	0.81

DAY.HOUR	FLUE GAS				REGENERATOR GAS			GASIFIER INLET GAS			
	O2	CO2 VOL %		SO2	O2	CO2	SO2	O2 VOL %		CO2 VOL %	
	%	ANAL	CALC	PPM	%	%	%	ANAL	CALC	ANAL	CALC
6.2030	1.5	13.2	14.7	459.	0.	4.3	8.4	18.9	19.7	1.08	0.87
6.2130	1.5	13.0	14.7	470.	0.	3.7	9.2	18.9	19.4	1.24	1.07
6.2230	1.3	13.2	14.9	485.	0.	1.6	9.2	18.5	19.3	1.57	1.11
6.2330	1.3	13.0	14.9	500.	0.	3.1	7.8	18.3	19.0	1.49	1.29
7.0030	1.5	13.0	14.7	515.	0.	3.0	8.2	18.5	19.3	1.57	1.10
7.0130	1.5	13.0	14.7	500.	0.20	2.2	8.2	18.3	19.3	1.57	1.10
7.0230	1.6	12.7	14.6	505.	0.	5.0	8.2	18.5	19.4	1.49	1.08
7.0330	1.8	12.7	14.5	500.	0.	3.0	7.8	18.2	19.4	1.65	1.09
7.0430	1.8	12.7	14.5	505.	0.	4.3	7.8	18.2	19.4	1.57	1.09
7.0530	1.8	12.7	14.5	535.	0.	1.7	8.2	18.4	19.4	1.49	1.09
7.0630	1.2	13.0	15.0	560.	0.	4.3	7.0	18.5	19.3	1.49	1.11
			STONE CHANGE								
7.0730	1.4	12.7	14.8	565.	0.	2.0	7.8	18.7	19.3	1.49	1.09
7.0830	1.5	13.0	14.7	550.	0.	4.9	8.0	18.7	19.3	1.49	1.12
7.0930	1.4	13.0	14.8	530.	0.	3.7	8.2	18.6	19.3	1.49	1.09
7.1030	1.5	13.0	14.7	510.	0.	5.2	8.6	18.8	19.4	1.33	1.08
7.1130	1.3	13.0	14.9	515.	0.	3.3	8.6	18.8	19.7	1.33	0.85
7.1230	1.3	13.0	14.9	515.	0.	3.3	8.6	18.8	19.7	1.33	0.86
7.1330	1.2	13.0	14.9	515.	0.	4.7	9.0	17.0	19.7	1.81	0.84
7.1430	1.2	13.2	14.9	515.	0.	6.3	8.2	18.1	19.4	1.81	1.07
7.1530	1.5	13.0	14.7	515.	0.	6.4	7.8	18.1	19.4	1.65	1.04
7.1630	1.5	13.1	14.7	515.	0.	6.2	7.8	18.1	19.4	1.81	1.05
7.1730	1.5	13.1	14.7	515.	0.	6.2	7.4	18.1	19.4	1.65	1.08
7.1830	1.5	12.7	14.7	510.	0.	5.0	8.2	18.1	19.5	1.65	0.96
7.1930	1.5	12.7	14.7	495.	0.	3.5	7.4	18.1	19.4	1.65	1.06
7.2030	1.8	13.0	14.5	475.	0.	5.9	7.8	18.1	19.4	1.65	1.08
7.2130	1.8	13.0	14.5	454.	0.	4.7	7.8	18.2	19.4	1.65	1.11
7.2230	1.8	12.7	14.5	475.	0.	4.6	7.8	18.4	19.4	1.65	1.09

7.2330	2.0	12.7	14.4	469.	0.	4.3	7.6	18.2	19.4	1.65	1.09
8.0030	2.0	12.7	14.3	464.	0.	3.1	7.8	18.2	19.4	1.65	1.09
8.0130	1.8	12.7	14.5	485.	0.	3.7	7.8	18.2	19.4	1.65	1.08
8.0230	1.9	12.7	14.4	490.	0.	3.7	7.8	18.2	19.4	1.65	1.08
8.0330	1.8	12.7	14.5	480.	0.	2.6	8.2	18.1	19.4	1.65	1.08
8.0430	1.6	12.7	14.6	480.	0.	2.6	8.0	18.1	19.4	1.65	1.08
8.0530	1.8	12.4	14.5	480.	0.	4.7	7.4	18.1	19.4	1.65	1.07
8.0630	1.8	12.4	14.5	490.	0.	3.1	8.2	18.1	19.4	1.65	1.07
8.0730	1.8	12.7	14.5	495.	0.	3.7	7.8	18.2	19.4	1.65	1.09
8.0830	1.7	12.7	14.5	464.	0.	3.9	8.0	18.5	19.4	1.57	1.07

SHUT DOWN AT 8.0830 FOR 21 HOURS

9.0530	3.0	13.0	13.5	666.	0.	0.0	0.3	16.5	17.5	3.36	2.49
9.0630	2.5	13.0	13.9	676.	0.	0.5	5.8	17.0	18.0	2.93	2.09
9.0730	3.0	13.0	13.5	646.	0.	0.6	6.6	17.1	18.1	2.76	2.09
9.0830	2.5	13.0	13.9	606.	0.	1.0	6.6	17.2	18.0	2.76	2.10
9.0930	2.8	13.0	13.7	591.	0.	1.6	6.2	17.6	18.2	2.52	2.00
9.1030	2.8	12.7	13.7	545.	0.	1.0	5.6	17.2	17.7	2.76	2.32

STONE CHANGE

9.1130	2.4	13.0	14.0	525.	0.	1.7	6.2	17.4	17.9	2.52	2.13
9.1230	2.0	13.0	14.3	520.	0.	1.7	6.4	18.7	18.8	1.73	1.53
9.1330	2.0	13.0	14.3	469.	0.	3.0	6.8	18.4	19.4	1.33	1.07
9.1430	2.0	12.7	14.3	424.	0.	2.2	7.4	19.6	19.5	1.16	0.97
9.1530	2.2	12.4	14.1	414.	0.	4.5	6.4	19.6	19.5	1.08	1.01
9.1630	2.0	12.2	14.3	424.	0.	2.4	8.8	21.0	20.7	0.02	0.18
9.1730	1.8	12.2	14.5	419.	0.	0.7	9.0	21.0	20.7	0.02	0.17
9.1830	1.6	12.2	14.7	449.	0.	1.6	7.6	21.0	20.7	0.02	0.17
9.1930	1.4	12.2	14.8	439.	0.	2.8	7.4	21.0	20.7	0.02	0.21
9.2030	2.0	11.7	14.4	414.	0.	4.1	7.4	21.0	20.7	0.02	0.20
9.2130	2.0	11.4	14.4	419.	0.	3.5	7.4	21.0	20.7	0.02	0.20
9.2230	2.1	11.0	14.3	424.	0.	3.0	7.8	21.0	20.7	0.02	0.19
9.2330	2.1	11.0	14.3	439.	0.	3.6	7.8	21.0	20.7	0.02	0.18
10.0030	2.0	10.7	14.4	424.	0.	2.2	7.8	21.0	20.7	0.02	0.18
10.0130	2.0	10.7	14.4	454.	0.	3.3	7.8	21.0	20.7	0.02	0.18

	FLUE GAS				REGENERATOR GAS			GASIFIER INLET GAS			
DAY.HOUR	O2 %	CO2 VOL % ANAL	CO2 VOL % CALC	SO2 PPM	O2 %	CO2 %	SO2 %	O2 VOL % ANAL	O2 VOL % CALC	CO2 VOL % ANAL	CO2 VOL % CALC
10.0230	2.1	10.7	14.3	454.	0.	2.5	7.8	21.0	20.7	0.02	0.18
10.0330	2.1	10.3	14.3	454.	0.	2.7	8.2	21.0	20.7	0.02	0.17
10.0430	2.1	10.3	14.3	454.	0.	3.0	7.6	21.0	20.7	0.02	0.17
10.0530	2.1	10.0	14.3	454.	0.	2.7	7.6	21.0	20.7	0.02	0.17
10.0630	2.1	9.8	14.3	449.	0.	2.8	7.8	21.0	20.7	0.02	0.16
10.0730	2.1	9.8	14.3	434.	0.	2.8	7.6	21.0	20.7	0.02	0.16
10.0830	2.0	9.8	14.4	434.	0.	2.9	7.4	21.0	20.7	0.02	0.17
10.0930	2.0	9.8	14.3	444.	0.	2.1	7.0	21.0	20.7	0.02	0.16
10.1030	2.0	9.8	14.3	434.	0.	3.0	7.8	21.0	20.7	0.02	0.15
10.1130	2.0	9.8	14.3	374.	0.	3.1	7.8	21.0	20.7	0.02	0.16
10.1230	2.0	9.8	14.3	343.	0.	3.3	7.8	21.0	20.7	0.02	0.16
10.1330	1.9	9.8	14.4	364.	0.	2.9	7.8	21.0	20.7	0.02	0.16
10.1430	2.0	9.8	14.3	369.	0.	2.4	6.6	21.0	20.7	0.02	0.16
10.1530	2.0	9.6	14.3	323.	0.	3.0	7.8	21.0	20.7	0.02	0.15
10.1630	3.2	9.0	13.4	293.	0.	3.3	7.4	21.0	20.7	0.02	0.15
10.1730	2.0	9.4	14.3	298.	0.	5.1	7.6	21.0	20.7	0.02	0.15
10.1830	2.0	9.4	14.4	283.	0.	3.6	9.0	21.0	20.7	0.02	0.15
10.1930	2.0	9.2	14.4	263.	0.	3.5	9.0	21.0	20.7	0.02	0.15
10.2030	2.0	9.4	14.4	268.	0.	2.3	9.0	21.0	20.7	0.02	0.15
10.2130	2.1	9.2	14.3	278.	0.	3.1	9.0	21.0	20.7	0.02	0.15
10.2230	2.1	9.2	14.3	283.	0.	2.8	8.8	21.0	20.7	0.02	0.15
10.2330	2.0	9.2	14.4	303.	0.	2.8	8.8	21.0	20.7	0.02	0.15
11.0030	2.0	9.2	14.4	318.	0.	2.8	8.4	21.0	20.7	0.02	0.15
11.0130	2.0	9.2	14.4	333.	0.	2.6	9.0	21.0	20.7	0.02	0.15
11.0230	2.1	9.2	14.3	323.	0.	2.3	7.4	21.0	20.7	0.02	0.15
11.0330	2.1	9.0	14.3	333.	0.	2.2	7.0	21.0	20.7	0.02	0.14
11.0430	2.1	9.0	14.3	333.	0.	2.2	7.2	21.0	20.7	0.02	0.14
11.0530	2.0	9.0	14.4	318.	0.	2.3	8.8	21.0	20.7	0.02	0.14

11.0630	2.1	9.0	14.3	308.	0.	2.3	8.0	21.0	20.7	0.02	0.14
11.0730	2.2	8.8	14.2	308.	0.	2.2	8.0	21.0	20.7	0.02	0.14
11.0830	2.1	9.0	14.3	313.	0.	2.2	8.2	21.0	20.7	0.02	0.14
11.0930	2.1	9.0	14.3	303.	0.	2.0	8.4	21.0	20.7	0.02	0.14
11.1030	2.1	9.0	14.3	313.	0.	2.6	8.4	21.0	20.7	0.02	0.14
11.1130	2.0	9.0	14.4	323.	0.	2.2	8.2	21.0	20.7	0.02	0.14
11.1230	2.0	9.0	14.4	308.	0.	3.1	8.4	21.0	20.7	0.02	0.14
11.1330	2.0	9.0	14.4	313.	0.	2.6	8.6	21.0	20.7	0.02	0.14
11.1430	2.0	9.0	14.4	313.	0.	2.5	8.2	21.0	20.7	0.02	0.14
11.1530	2.0	9.0	14.4	318.	0.	3.1	8.2	21.0	20.7	0.02	0.14
11.1630	2.0	9.0	14.4	313.	0.	2.6	9.0	21.0	20.7	0.02	0.14
11.1730	2.0	9.0	14.4	303.	0.	2.6	9.4	21.0	20.7	0.02	0.14
11.1830	2.0	9.0	14.4	308.	0.	1.9	8.2	21.0	20.7	0.02	0.14
11.1930	2.0	9.0	14.4	308.	0.	2.2	9.9	21.0	20.7	0.02	0.14
11.2030	2.0	9.0	14.4	303.	0.	2.8	9.9	21.0	20.7	0.02	0.14
11.2130	2.1	9.0	14.3	313.	0.	2.0	8.6	20.0	18.9	0.81	0.99
11.2230	2.0	9.0	14.3	344.	0.	1.9	8.2	18.0	17.6	0.99	1.60
11.2330	2.0	8.8	14.4	404.	0.	1.9	8.0	17.6	18.9	2.60	0.96
STONE CHANGE											
12.0030	2.3	8.8	14.1	394.	0.	2.4	7.6	18.6	19.0	1.88	0.95
12.0130	2.3	8.8	14.1	333.	0.	2.0	7.8	20.4	18.9	0.63	1.00
12.0230	2.2	8.8	14.2	303.	0.	1.6	1.8	21.0	18.5	0.02	1.18
12.0330	2.2	8.8	14.2	318.	0.	1.3	7.8	21.0	18.8	0.13	1.04
12.0430	2.0	8.8	14.3	303.	0.	2.3	8.6	21.0	20.7	0.13	0.15
12.0530	2.0	9.0	14.3	283.	0.	1.9	8.8	21.0	20.7	0.02	0.15
12.0630	2.0	8.8	14.3	273.	0.	2.5	8.6	21.0	20.7	0.02	0.15
12.0730	2.0	8.8	14.3	283.	0.	2.4	8.8	21.0	20.7	0.02	0.15
12.0830	2.1	8.8	14.2	253.	0.	1.8	8.2	21.0	20.7	0.02	0.15
12.0930	2.1	8.8	14.2	258.	0.	3.3	8.2	21.0	20.7	0.02	0.15
12.1030	2.0	13.0	14.3	263.	0.	1.0	7.8	21.0	20.7	0.02	0.23
12.1130	2.0	13.0	14.3	258.	0.	3.1	8.6	21.0	20.7	0.02	0.23
12.1230	2.0	13.0	14.3	248.	0.	4.1	8.6	21.0	20.7	0.02	0.23
12.1330	2.0	13.0	14.3	248.	0.	4.1	8.2	21.0	20.7	0.02	0.23
12.1430	2.0	13.0	14.3	212.	0.	3.1	8.2	21.0	20.7	0.02	0.23
12.1530	2.0	13.0	14.3	232.	0.	3.3	8.2	21.0	20.7	0.02	0.23
12.1630	2.0	13.0	14.3	243.	0.	3.4	7.8	21.0	20.7	0.02	0.22

DAY.HOUR	FLUE GAS O_2 %	FLUE GAS CO_2 VOL % ANAL	FLUE GAS CO_2 VOL % CALC	FLUE GAS SO_2 PPM	REGENERATOR GAS O_2 %	REGENERATOR GAS CO_2 %	REGENERATOR GAS SO_2 %	GASIFIER INLET GAS O_2 VOL % ANAL	GASIFIER INLET GAS O_2 VOL % CALC	GASIFIER INLET GAS CO_2 VOL % ANAL	GASIFIER INLET GAS CO_2 VOL % CALC
12.1730	2.2	13.2	14.2	243.	0.	1.7	7.4	21.0	20.7	0.02	0.22
12.1830	2.0	13.0	14.3	263.	0.	2.0	8.2	21.0	20.7	0.02	0.21
12.1930	2.2	13.0	14.2	268.	0.	2.0	8.2	21.0	20.7	0.02	0.21
12.2030	2.1	13.0	14.2	263.	0.	2.3	8.2	21.0	20.7	0.02	0.21
12.2130	1.6	13.5	14.6	253.	0.	1.9	8.6	21.0	20.7	0.02	0.22
12.2230	1.5	13.6	14.7	258.	0.	2.9	8.6	21.0	20.7	0.02	0.22
12.2330	1.8	13.5	14.5	253.	0.	1.6	8.6	21.0	20.7	0.02	0.22
13.0030	1.8	13.5	14.5	253.	0.	1.6	8.6	21.0	20.7	0.02	0.22
13.0130	2.0	13.2	14.3	278.	0.	1.6	8.6	21.0	20.7	0.02	0.21
13.0230	2.2	13.2	14.2	318.	0.	1.1	9.2	21.0	20.7	0.02	0.21
13.0330	2.2	13.2	14.2	303.	0.	1.1	9.2	21.0	20.7	0.02	0.20
13.0430	1.8	13.2	14.5	283.	0.	1.1	6.8	21.0	20.7	0.02	0.20
13.0530	1.8	13.2	14.5	273.	0.	1.3	8.4	21.0	20.7	0.02	0.20
13.0630	1.8	13.2	14.5	293.	0.	1.3	8.2	21.0	20.7	0.02	0.21
13.0730	1.6	13.5	14.6	298.	0.	1.1	8.2	21.0	20.7	0.02	0.21
13.0830	1.6	13.5	14.6	278.	0.	1.0	7.0	21.0	20.7	0.02	0.21
13.0930	1.8	13.5	14.5	303.	0.	1.9	7.2	21.0	20.7	0.02	0.21
13.1030	1.6	13.2	14.6	293.	0.	1.1	7.0	21.0	20.7	0.02	0.20
13.1130	1.2	13.2	15.0	278.	0.	2.1	7.2	21.0	20.7	0.02	0.21
13.1230	1.1	13.5	15.0	273.	0.	1.6	8.2	21.0	20.7	0.02	0.22
13.1330	1.0	13.5	15.1	293.	0.	1.8	7.8	21.0	20.7	0.02	0.23
13.1430	1.0	13.5	15.1	298.	0.	2.2	7.6	21.0	20.7	0.02	0.23
13.1530	0.9	13.5	15.2	308.	0.	1.7	7.6	21.0	20.7	0.02	0.23
13.1630	0.9	13.5	15.2	308.	0.	1.6	7.4	21.0	20.7	0.02	0.23
13.1730	0.8	13.8	15.3	303.	0.	1.6	7.4	21.0	20.7	0.02	0.24
13.1830	1.4	13.5	14.8	298.	0.	2.5	7.4	20.8	19.8	0.02	0.84
13.1930	1.4	13.8	14.8	313.	0.	2.0	7.4	20.8	19.8	0.02	0.87
13.2030	1.6	13.5	14.6	318.	0.	2.2	7.8	20.8	19.8	0.02	0.85

13.2130	1.3	13.8	14.9	323.	0.	1.6	8.2	20.8	19.8	0.02	0.83
13.2230	1.3	13.8	14.9	328.	0.	1.6	8.2	20.8	19.8	0.02	0.87
13.2330	1.0	13.8	15.1	328.	0.	1.6	8.2	20.5	19.8	0.02	0.85
14.0030	1.0	13.8	15.1	328.	0.	1.6	8.2	20.3	19.0	0.44	1.38
14.0130	1.0	13.9	15.1	339.	0.	1.6	7.6	20.3	19.0	0.44	1.40
14.0230	1.0	14.4	15.1	354.	0.	1.6	7.6	20.3	19.0	0.44	1.47
14.0330	1.0	14.4	15.1	328.	0.	1.1	8.2	20.5	19.0	0.53	1.47
14.0430	1.0	14.4	15.1	334.	0.	1.1	8.2	20.5	19.1	0.53	1.39
14.0530	1.0	14.7	15.1	374.	0.	1.1	8.2	17.0	18.2	1.96	2.07
14.0630	1.0	14.7	15.1	369.	0.	1.9	8.2	17.5	19.2	1.96	1.33
14.0730	1.9	14.1	14.4	344.	0.	2.3	8.2	18.0	19.3	1.96	1.24
14.0830	1.9	14.1	14.4	333.	0.	2.2	7.8	18.0	19.4	1.81	1.19
14.0930	1.9	14.1	14.4	349.	0.	2.2	7.8	18.0	19.3	1.81	1.23
14.1030	1.8	14.1	14.5	359.	0.	2.8	7.4	17.5	19.3	2.12	1.24
14.1130	1.8	14.1	14.5	359.	0.	2.8	7.8	17.6	19.3	2.12	1.26
14.1230	1.7	14.1	14.6	364.	0.	1.9	7.8	17.7	19.2	2.04	1.28
14.1330	1.3	14.1	14.9	369.	0.	2.2	7.8	17.8	19.2	1.96	1.28
14.1430	1.4	14.1	14.8	364.	0.	3.0	7.2	17.8	19.3	1.96	1.26
14.1530	1.3	13.8	14.8	364.	0.	2.6	7.6	17.8	19.2	1.81	1.26
14.1630	1.3	13.8	14.8	369.	0.	3.3	7.0	17.8	19.2	1.88	1.26
14.1730	1.4	13.8	14.8	344.	0.	3.0	6.8	17.8	19.2	1.96	1.28
14.1830	1.4	13.8	14.8	303.	0.	4.1	6.6	17.8	19.2	1.96	1.28
14.1930	1.5	13.2	14.7	339.	0.	3.0	7.4	17.8	19.2	1.96	1.22
14.2030	2.0	13.2	14.3	349.	0.	3.3	7.0	17.8	19.2	1.88	1.23
14.2130	1.9	13.2	14.4	394.	0.	3.5	7.0	17.8	19.3	1.96	1.19
14.2230	1.9	13.2	14.4	364.	0.	3.1	7.0	17.5	19.3	1.81	1.19
14.2330	2.0	13.0	14.3	369.	0.	3.7	7.4	17.9	19.3	1.81	1.17
15.0030	2.2	13.0	14.2	384.	0.	4.2	6.6	17.9	19.3	1.73	1.17
15.0130	1.5	13.2	14.7	384.	0.	3.5	7.4	17.9	19.2	1.81	1.21
15.0230	1.2	13.5	14.9	394.	0.	3.1	6.6	18.0	19.2	1.73	1.23
15.0330	1.1	13.2	15.0	399.	0.	2.8	6.6	18.1	19.2	1.65	1.18
15.0430	1.5	13.0	14.7	394.	0.	3.3	6.6	18.1	19.3	1.65	1.14
15.0530	1.3	13.0	14.8	379.	0.	3.1	6.8	18.2	19.2	1.49	1.18
15.0630	1.2	13.1	14.9	364.	0.	3.1	6.8	18.2	19.2	1.49	1.19
15.0730	1.5	13.1	14.7	349.	0.	3.1	6.8	18.2	19.2	1.41	1.18
15.0830	1.4	13.2	14.8	344.	0.	3.5	7.0	18.2	19.2	1.33	1.22

RUN 7: GAS COMPOSITIONS

	FLUE GAS				REGENERATOR GAS			GASIFIER INLET GAS			
DAY.HOUR	O2	CO2 VOL %		SO2	O2	CO2	SO2	O2 VOL %		CO2 VOL %	
	%	ANAL	CALC	PPM	%	%	%	ANAL	CALC	ANAL	CALC
15.0930	1.4	13.2	14.8	354.	0.	2.8	7.0	18.0	18.9	1.65	1.42
15.1030	1.4	13.0	14.8	349.	0.	3.6	7.0	18.1	19.0	1.57	1.35
15.1130	1.4	13.0	14.8	324.	0.	4.6	7.0	18.2	18.9	1.49	1.36
15.1230	1.4	13.2	14.7	303.	0.	3.9	7.6	18.2	18.7	1.49	1.58
15.1330	1.2	13.2	14.9	334.	0.	3.0	7.6	18.4	18.6	1.41	1.58
15.1430	1.2	13.0	14.9	319.	0.	3.4	7.2	18.4	18.6	1.49	1.59
15.1530	1.4	13.0	14.7	313.	0.	4.1	6.8	18.4	18.5	1.33	1.62
15.1630	1.2	13.0	14.9	313.	0.30	3.6	6.4	18.4	18.5	1.33	1.65
15.1730	1.0	13.0	15.1	344.	0.20	3.9	6.6	18.2	18.8	1.49	1.42
15.1830	1.2	13.0	14.9	349.	0.30	4.3	6.0	18.0	18.5	1.65	1.64
15.1930	1.3	13.0	14.8	359.	0.30	3.5	6.0	18.2	18.5	1.65	1.62
15.2030	1.4	13.0	14.7	359.	0.30	4.2	5.4	17.6	18.6	1.96	1.60
15.2130	1.3	13.0	14.8	354.	0.30	2.6	6.0	18.0	18.5	1.81	1.65
15.2230	1.9	12.7	14.4	354.	0.30	3.5	6.2	18.0	18.5	1.88	1.65
15.2330	1.0	13.2	15.0	339.	0.30	3.2	6.0	18.0	18.5	1.81	1.69
16.0030	1.0	13.2	15.1	339.	0.20	3.2	6.2	18.0	18.5	1.81	1.69
16.0130	2.0	13.2	14.3	333.	0.20	2.0	7.4	18.9	19.4	1.41	1.14
16.0230	1.9	13.2	14.4	303.	0.20	2.5	7.6	20.2	19.2	0.23	1.24
16.0330	1.9	13.0	14.4	323.	0.20	2.2	7.8	20.4	19.2	0.13	1.25
16.0430	2.0	13.0	14.3	353.	0.20	1.7	8.4	20.4	19.2	0.13	1.26
16.0530	2.0	13.2	14.3	343.	0.20	3.2	7.6	20.4	19.3	0.13	1.20
16.0630	2.0	12.7	14.3	343.	0.20	4.0	7.2	20.4	19.2	0.02	1.24
16.0730	2.0	12.7	14.3	323.	0.20	5.0	7.2	20.4	19.2	0.02	1.24
16.0830	2.2	12.7	14.2	318.	0.10	5.4	7.2	20.5	19.2	0.13	1.24
16.0930	2.3	12.7	14.1	313.	0.10	5.7	7.0	20.5	19.0	0.13	1.33
16.1030	2.3	12.7	14.1	318.	0.20	4.1	7.0	20.5	19.0	0.13	1.33
16.1130	2.2	12.7	14.2	318.	0.30	5.3	7.2	20.5	18.7	0.13	1.58
16.1230	2.0	13.0	14.3	308.	0.20	5.5	6.8	20.5	18.6	0.13	1.65

16.1330	2.0	13.0	14.3	318.	0.20	5.2	7.2	20.6	19.0	0.13	1.39
16.1430	2.0	13.0	14.3	323.	0.10	5.1	7.4	20.6	18.6	0.02	1.65
16.1530	2.0	13.0	14.3	323.	0.10	5.0	7.4	20.2	19.1	0.53	1.30
16.1630	2.0	13.0	14.3	328.	0.10	4.7	7.4	20.2	18.9	0.53	1.40
16.1730	2.0	13.0	14.3	328.	0.10	5.4	7.2	20.2	19.0	0.44	1.35
16.1830	2.0	12.7	14.3	318.	0.10	5.2	7.6	20.4	19.0	0.34	1.34
16.1930	2.0	12.7	14.3	313.	0.10	5.5	7.2	20.4	19.0	0.34	1.33
16.2030	2.1	12.7	14.2	333.	0.10	4.7	7.4	20.2	18.6	0.53	1.63
16.2130	2.1	12.7	14.2	343.	0.10	3.5	7.6	20.2	18.5	0.53	1.65
16.2230	2.0	12.7	14.3	389.	0.10	2.5	7.6	20.2	18.5	0.53	1.65
16.2330	2.0	12.7	14.3	414.	0.10	2.0	7.8	20.2	18.5	0.53	1.65
17.0030	2.0	12.7	14.3	419.	0.10	1.7	7.8	20.2	18.5	0.53	1.65
17.0130	2.0	13.0	14.3	424.	0.10	2.2	7.4	19.0	18.6	1.16	1.66
17.0230	2.0	13.0	14.3	434.	0.10	1.8	7.4	19.0	18.6	1.16	1.66
17.0330	0.	9.4	15.8	420.	0.10	3.1	8.2	19.0	18.3	0.99	1.19
17.0430	2.0	13.2	14.3	414.	0.10	2.8	7.0	19.0	18.6	0.99	1.68
17.0530	2.0	13.2	14.3	404.	0.10	3.3	7.0	19.0	18.6	0.99	1.68
17.0630	2.0	13.2	14.3	394.	0.10	4.1	7.0	19.0	18.6	0.99	1.68
17.0730	2.0	13.2	14.3	384.	0.10	4.3	7.4	19.0	18.6	0.99	1.68
17.0830	2.0	13.2	14.3	379.	0.10	4.9	6.6	19.0	18.6	0.99	1.68
17.0930	2.0	13.0	14.3	369.	0.10	5.0	7.4	20.2	18.5	0.53	1.68
17.1030	2.0	13.0	14.3	379.	0.10	5.3	7.4	20.1	18.6	0.63	1.65
17.1130	1.6	13.2	14.6	404.	0.10	6.0	7.2	20.1	18.5	0.63	1.69
17.1230	1.8	13.2	14.5	399.	0.10	4.5	7.4	20.5	18.9	0.44	1.43
17.1330	1.8	13.0	14.5	399.	0.10	5.4	7.4	20.5	18.5	0.44	1.69
17.1430	1.6	13.0	14.6	404.	0.10	4.7	7.4	20.5	18.5	0.23	1.69
17.1530	2.0	13.0	14.3	404.	0.10	5.1	7.2	21.0	20.6	0.02	0.25
17.1630	2.4	13.0	14.0	358.	0.10	4.5	6.8	21.0	20.6	0.02	0.26
17.1730	2.0	13.2	14.3	353.	0.20	4.6	7.0	21.0	20.6	0.02	0.26
17.1830	1.6	13.2	14.6	364.	0.20	4.1	7.6	21.0	20.6	0.02	0.26
17.1930	2.0	13.0	14.3	353.	0.10	3.7	7.4	21.0	20.6	0.02	0.26
17.2030	2.0	13.0	14.3	353.	0.10	4.5	7.0	21.0	20.6	0.02	0.26
17.2130	2.0	13.0	14.3	353.	0.10	3.6	6.8	21.1	20.6	0.02	0.26

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
1.1230	0.141	0.110	0.035	0.005	-0.009	0.	2.0	-2.0
1.1330	0.282	0.222	0.039	0.010	0.011	0.	4.0	-4.0
1.1430	0.418	0.280	0.059	0.015	0.065	0.	6.0	-6.0
1.1530	0.557	0.376	0.093	0.020	0.069	1.3	7.9	-6.7
1.1630	0.695	0.443	0.132	0.025	0.095	9.8	9.9	-0.2
1.1730	0.833	0.515	0.188	0.029	0.101	19.0	12.0	7.0
1.1830	0.971	0.588	0.242	0.034	0.108	24.4	14.0	10.4
1.1930	1.109	0.657	0.306	0.036	0.110	30.6	15.3	15.3
1.2030	1.247	0.725	0.366	0.038	0.118	36.7	15.9	20.8
1.2130	1.385	0.792	0.421	0.039	0.133	42.6	16.5	26.2
1.2230	1.522	0.856	0.508	0.040	0.118	48.3	17.1	31.2
1.2330	1.660	0.923	0.573	0.071	0.093	54.2	34.1	20.1
2.0030	1.798	0.994	0.662	0.072	0.069	61.6	34.7	26.9
2.0130	1.936	1.058	0.734	0.073	0.070	69.8	35.3	34.5
2.0230	2.073	1.115	0.809	0.102	0.048	77.0	53.4	23.6
2.0330	2.210	1.172	0.881	0.105	0.052	85.7	54.8	31.0
2.0430	2.348	1.225	0.954	0.107	0.061	95.5	56.2	39.3
2.0530	2.485	1.277	1.045	0.136	0.027	106.0	76.1	29.9
2.0630	2.622	1.324	1.142	0.167	-0.011	116.8	98.0	18.8
2.0730	2.759	1.369	1.209	0.176	0.005	127.6	103.6	24.0
2.0830	2.896	1.412	1.272	0.205	0.007	138.1	126.0	12.2
2.0930	3.033	1.454	1.356	0.210	0.013	147.9	129.3	18.6
2.1030	3.170	1.490	1.438	0.214	0.027	159.2	132.7	26.5
2.1130	3.306	1.526	1.529	0.219	0.032	169.7	136.1	33.6
2.1230	3.443	1.561	1.624	0.222	0.035	181.5	138.9	42.6
2.1330	3.579	1.591	1.724	0.250	0.014	192.0	167.6	24.4
2.1430	3.717	1.624	1.818	0.253	0.022	198.2	170.0	28.2
2.1530	3.854	1.660	1.906	0.255	0.033	205.6	172.3	33.4

2.1630	3.991	1.702	2.014	0.258	0.017	213.9	174.6	39.3
2.1730	4.131	1.748	2.117	0.272	-0.006	221.8	191.2	30.6
2.1830	4.267	1.793	2.193	0.274	0.006	228.5	193.4	35.1
2.1930	4.404	1.841	2.290	0.276	-0.003	235.9	195.6	40.3
2.2030	4.541	1.887	2.378	0.279	-0.003	243.6	197.5	46.1
2.2130	4.679	1.930	2.470	0.280	-0.002	251.9	199.2	52.6
2.2230	4.817	1.971	2.569	0.282	-0.005	259.8	200.9	58.9
2.2330	4.955	2.008	2.671	0.284	-0.009	268.0	202.5	65.5
3.0030	5.092	2.045	2.771	0.286	-0.010	276.5	204.2	72.3
3.0130	5.228	2.086	2.885	0.288	-0.030	285.5	205.9	79.6
3.0230	5.367	2.124	2.975	0.290	-0.022	293.4	207.5	85.9
3.0330	5.505	2.160	3.071	0.291	-0.018	301.9	209.2	92.7
3.0430	5.643	2.194	3.173	0.293	-0.017	310.4	210.9	99.5
3.0530	5.780	2.227	3.274	0.295	-0.016	318.1	212.5	105.6
3.0630	5.917	2.260	3.368	0.297	-0.007	330.2	214.2	116.0
3.0730	6.055	2.294	3.469	0.299	-0.006	343.5	215.9	127.7
3.0830	6.193	2.329	3.573	0.305	-0.014	358.1	222.0	136.2
3.0930	6.330	2.365	3.674	0.312	-0.020	371.7	229.5	142.2
3.1030	6.468	2.398	3.775	0.319	-0.025	384.3	236.6	147.8
3.1130	6.605	2.432	3.868	0.325	-0.020	398.2	242.9	155.3
3.1230	6.743	2.466	3.975	0.331	-0.029	416.4	248.9	167.5
3.1330	6.880	2.497	4.074	0.337	-0.028	434.1	254.8	179.3
3.1430	7.019	2.526	4.185	0.343	-0.036	450.6	260.8	189.8
3.1530	7.155	2.555	4.300	0.350	-0.050	468.0	266.8	201.3
3.1630	7.292	2.583	4.403	0.357	-0.052	486.0	274.2	211.7
3.1730	7.429	2.611	4.505	0.366	-0.052	506.3	282.2	224.0
3.1830	7.566	2.638	4.615	0.374	-0.060	521.7	290.3	231.4
3.1930	7.704	2.664	4.711	0.383	-0.055	539.1	299.0	240.1
3.2030	7.841	2.689	4.805	0.393	-0.046	555.3	308.6	246.7
3.2130	7.978	2.714	4.889	0.403	-0.028	571.5	318.2	253.3
3.2230	8.115	2.739	5.013	0.412	-0.049	580.7	326.6	254.2
3.2330	8.253	2.764	5.133	0.419	-0.063	587.4	333.7	253.8
4.0030	8.394	2.788	5.255	0.427	-0.076	593.6	340.8	252.8
4.0130	8.538	2.815	5.366	0.432	-0.075	600.5	346.1	254.4
4.0230	8.682	2.844	5.490	0.438	-0.089	607.9	351.5	256.4
4.0330	8.829	2.874	5.621	0.447	-0.113	614.4	360.5	253.9

APPENDIX D: TABLE VI.

RUN 7: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
4.0430	8.974	2.906	5.740	0.452	-0.124	621.3	365.5	255.8
4.0530	9.120	2.941	5.842	0.460	-0.124	627.7	372.6	255.1
4.0630	9.267	2.973	5.943	0.465	-0.115	633.4	377.6	255.7
4.0730	9.412	3.002	6.050	0.469	-0.108	639.8	381.5	258.3
4.0830	9.559	3.031	6.160	0.472	-0.104	647.7	384.0	263.8
4.0930	9.704	3.061	6.271	0.474	-0.102	654.2	386.4	267.8
4.1030	9.850	3.092	6.381	0.477	-0.100	659.3	388.8	270.5
4.1130	9.996	3.125	6.494	0.480	-0.103	665.7	391.2	274.5
4.1230	10.139	3.160	6.599	0.484	-0.104	672.1	395.7	276.4
4.1330	10.283	3.195	6.707	0.489	-0.108	678.5	400.2	278.3
4.1430	10.426	3.229	6.809	0.494	-0.105	685.5	404.7	280.8
4.1530	10.570	3.262	6.921	0.499	-0.112	691.6	409.2	282.4
4.1630	10.713	3.294	7.037	0.503	-0.122	697.3	413.8	283.5
4.1730	10.857	3.328	7.145	0.508	-0.124	704.7	418.3	286.5
4.1830	11.001	3.361	7.245	0.512	-0.118	712.9	421.9	291.1
4.1930	11.144	3.393	7.356	0.516	-0.120	720.4	425.2	295.2
4.2030	11.288	3.423	7.470	0.519	-0.123	726.8	428.6	298.3
4.2130	11.432	3.450	7.585	0.523	-0.126	733.5	431.9	301.6
4.2230	11.576	3.477	7.694	0.526	-0.121	740.9	435.2	305.7
4.2330	11.720	3.503	7.813	0.532	-0.129	747.9	440.4	307.4
5.0030	11.864	3.525	7.926	0.537	-0.124	754.8	444.6	310.2
5.0130	12.008	3.551	8.030	0.541	-0.114	761.7	448.8	313.0
5.0230	12.152	3.581	8.152	0.546	-0.127	769.9	452.9	317.0
5.0330	12.296	3.610	8.266	0.550	-0.130	778.2	457.0	321.2
5.0430	12.440	3.639	8.375	0.554	-0.129	786.1	461.1	325.0
5.0530	12.585	3.667	8.489	0.558	-0.130	794.6	465.2	329.4
5.0630	12.732	3.695	8.605	0.563	-0.131	802.8	470.2	332.6
5.0730	12.880	3.724	8.728	0.568	-0.141	809.7	475.3	334.5

5.0830	13.028	3.754	8.846	0.573	-0.145	818.2	480.2	338.0
5.0930	13.177	3.785	8.964	0.591	-0.163	825.1	501.4	323.8
5.1030	13.325	3.815	9.087	0.596	-0.173	831.0	506.3	324.8
5.1130	13.473	3.845	9.205	0.600	-0.178	838.0	511.2	326.8
5.1230	13.621	3.876	9.330	0.605	-0.190	844.7	515.8	328.9
5.1330	13.769	3.907	9.450	0.609	-0.196	851.8	520.3	331.6
5.1430	13.917	3.938	9.569	0.613	-0.203	859.0	524.8	334.3
5.1530	14.065	3.969	9.689	0.618	-0.211	866.5	529.2	337.2
5.1630	14.212	4.000	9.805	0.622	-0.215	874.2	533.9	340.3
5.1730	14.360	4.031	9.924	0.627	-0.222	880.9	538.6	342.2
5.1830	14.508	4.061	10.037	0.631	-0.222	885.7	543.4	342.3
5.1930	14.656	4.092	10.164	0.636	-0.237	890.1	548.2	341.9
5.2030	14.803	4.125	10.271	0.641	-0.233	893.9	553.1	340.8
5.2130	14.951	4.157	10.390	0.646	-0.242	897.0	558.2	338.9
5.2230	15.099	4.189	10.540	0.651	-0.281	900.9	563.2	337.7
5.2330	15.246	4.226	10.641	0.656	-0.277	904.7	568.2	336.5
6.0030	15.394	4.263	10.786	0.659	-0.315	908.6	572.1	336.4
6.0130	15.542	4.302	10.886	0.663	-0.309	912.7	576.0	336.6
6.0230	15.691	4.341	10.993	0.667	-0.310	917.3	579.9	337.4
6.0330	15.840	4.380	11.093	0.671	-0.304	921.4	583.9	337.6
6.0430	15.989	4.418	11.195	0.682	-0.307	925.8	593.8	332.0
6.0530	16.138	4.456	11.287	0.688	-0.294	930.4	599.7	330.7
6.0630	16.288	4.494	11.376	0.694	-0.276	934.3	605.7	328.6
6.0730	16.437	4.533	11.471	0.698	-0.266	938.6	609.4	329.2
6.0830	16.586	4.573	11.573	0.703	-0.262	944.5	613.1	331.4
6.0930	16.736	4.611	11.681	0.707	-0.263	950.7	616.8	333.9
6.1030	16.885	4.648	11.799	0.712	-0.274	955.8	620.5	335.3
6.1130	17.034	4.685	11.917	0.722	-0.289	959.7	628.2	331.5
6.1230	17.184	4.725	12.031	0.726	-0.297	962.5	631.9	330.6
6.1330	17.334	4.768	12.142	0.730	-0.307	966.1	635.5	330.6
6.1430	17.483	4.811	12.252	0.734	-0.314	970.2	638.9	331.2
6.1530	17.633	4.855	12.367	0.744	-0.333	974.6	646.4	328.2
6.1630	17.782	4.900	12.488	0.747	-0.353	978.7	649.6	329.0
6.1730	17.932	4.948	12.602	0.751	-0.369	984.6	652.6	332.0
6.1830	18.081	5.009	12.729	0.754	-0.411	989.4	655.6	333.8
6.1930	18.231	5.057	12.836	0.764	-0.426	993.3	662.5	330.8

APPENDIX D: TABLE VI.

RUN 7: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
6.2030	18.381	5.102	12.936	0.767	-0.424	998.2	665.5	332.7
6.2130	18.530	5.148	13.037	0.772	-0.427	1002.3	669.7	332.6
6.2230	18.679	5.195	13.156	0.778	-0.449	1007.2	674.2	333.0
6.2330	18.829	5.244	13.254	0.789	-0.457	1011.5	682.7	328.9
7.0030	18.979	5.294	13.362	0.794	-0.472	1014.9	687.2	327.7
7.0130	19.127	5.343	13.468	0.800	-0.483	1020.0	691.7	328.3
7.0230	19.279	5.393	13.583	0.810	-0.507	1025.1	699.5	325.7
7.0330	19.429	5.443	13.685	0.814	-0.513	1030.3	702.6	327.7
7.0430	19.578	5.493	13.792	0.817	-0.524	1034.1	705.7	328.5
7.0530	19.728	5.546	13.899	0.821	-0.538	1038.2	708.8	329.5
7.0630	19.877	5.600	13.991	0.824	-0.538	1040.0	711.8	328.2
			STONE CHANGE					
7.0730	20.027	5.656	14.084	0.827	-0.540	1048.8	714.4	334.3
7.0830	20.178	5.710	14.181	0.835	-0.549	1059.0	720.3	338.7
7.0930	20.328	5.762	14.279	0.838	-0.551	1066.7	722.8	343.9
7.1030	20.478	5.812	14.383	0.841	-0.558	1073.4	725.2	348.2
7.1130	20.622	5.860	14.482	0.844	-0.564	1080.6	727.7	352.9
7.1230	20.772	5.911	14.583	0.847	-0.568	1088.3	730.1	358.1
7.1330	20.922	5.960	14.688	0.849	-0.575	1095.7	732.6	363.1
7.1430	21.073	6.010	14.784	0.852	-0.574	1102.4	735.1	367.3
7.1530	21.223	6.061	14.885	0.861	-0.583	1110.4	741.0	369.4
7.1630	21.371	6.111	14.981	0.863	-0.584	1118.8	743.5	375.4
7.1730	21.522	6.162	15.078	0.868	-0.586	1127.1	747.7	379.3
7.1830	21.673	6.212	15.185	0.875	-0.600	1133.2	753.8	379.4
7.1930	21.822	6.261	15.279	0.888	-0.605	1139.4	763.3	376.0
7.2030	21.973	6.308	15.380	0.894	-0.610	1146.6	769.4	377.2
7.2130	22.122	6.353	15.485	0.904	-0.620	1152.7	778.5	374.3
7.2230	22.272	6.401	15.590	0.919	-0.638	1158.1	793.5	364.7

7.2330	22.421	6.448	15.688	0.931	-0.645	1162.5	804.4	358.1
8.0030	22.571	6.494	15.785	0.938	-0.647	1167.9	811.1	356.7
8.0130	22.720	6.543	15.884	0.946	-0.652	1173.8	817.9	355.9
8.0230	22.870	6.592	15.983	0.954	-0.658	1179.9	824.7	355.2
8.0330	23.020	6.640	16.084	0.961	-0.665	1186.6	831.5	355.1
8.0430	23.169	6.687	16.182	0.971	-0.671	1193.3	839.9	353.4
8.0530	23.319	6.735	16.274	0.981	-0.671	1201.0	848.1	352.9
8.0630	23.468	6.784	16.375	0.994	-0.684	1213.6	857.9	355.7
8.0730	23.618	6.833	16.471	1.001	-0.686	1228.5	864.2	364.2
8.0830	23.768	6.879	16.566	1.010	-0.688	1244.6	873.8	370.8
	SHUT DOWN AT 8.0830 FOR 21 HOURS							
9.0530	23.903	6.944	16.570	1.023	-0.633	1259.8	886.7	373.1
9.0630	24.047	7.011	16.645	1.035	-0.644	1277.0	899.5	377.5
9.0730	24.187	7.075	16.719	1.048	-0.655	1291.4	913.2	378.2
9.0830	24.324	7.133	16.814	1.063	-0.686	1305.7	929.5	376.2
9.0930	24.461	7.190	16.884	1.067	-0.680	1319.1	932.8	386.3
9.1030	24.598	7.243	16.953	1.070	-0.668	1336.6	936.0	400.5
			STONE CHANGE					
9.1130	24.737	7.293	17.031	1.074	-0.661	1353.0	939.3	413.7
9.1230	24.879	7.343	17.105	1.078	-0.647	1379.4	942.5	436.9
9.1330	25.022	7.389	17.186	1.086	-0.638	1410.0	950.7	459.3
9.1430	25.166	7.430	17.274	1.094	-0.632	1438.5	958.9	479.6
9.1530	25.309	7.471	17.353	1.102	-0.616	1467.5	967.4	500.1
9.1630	25.453	7.512	17.460	1.110	-0.629	1489.8	976.1	513.7
9.1730	25.596	7.552	17.569	1.117	-0.642	1497.8	983.0	514.7
9.1830	25.748	7.597	17.662	1.122	-0.634	1509.3	988.2	521.1
9.1930	25.896	7.639	17.756	1.132	-0.631	1520.4	995.9	524.4
9.2030	26.044	7.681	17.850	1.141	-0.628	1532.7	1004.0	528.7
9.2130	26.192	7.722	17.944	1.150	-0.624	1545.5	1012.5	533.0
9.2230	26.340	7.765	18.043	1.159	-0.626	1557.3	1021.0	536.4
9.2330	26.488	7.809	18.141	1.169	-0.630	1569.4	1029.9	539.5
10.0030	26.636	7.851	18.239	1.179	-0.633	1579.9	1039.3	540.7
10.0130	26.784	7.896	18.339	1.189	-0.640	1591.2	1048.7	542.6

APPENDIX D: TABLE VI.

RUN 7: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
10.0230	26.932	7.941	18.437	1.200	-0.646	1604.3	1058.0	546.3
10.0330	27.080	7.987	18.541	1.210	-0.658	1616.6	1068.0	548.6
10.0430	27.229	8.032	18.638	1.221	-0.662	1629.2	1078.2	551.0
10.0530	27.377	8.078	18.733	1.232	-0.665	1641.8	1088.4	553.4
10.0630	27.526	8.123	18.832	1.240	-0.669	1654.1	1096.2	557.9
10.0730	27.674	8.166	18.928	1.248	-0.668	1667.2	1103.2	564.0
10.0830	27.822	8.209	19.022	1.256	-0.665	1680.6	1110.7	569.8
10.0930	27.970	8.254	19.109	1.264	-0.656	1698.3	1118.7	579.6
10.1030	28.119	8.297	19.207	1.286	-0.671	1726.8	1143.0	583.8
10.1130	28.266	8.334	19.304	1.296	-0.669	1754.5	1152.3	602.2
10.1230	28.416	8.369	19.404	1.307	-0.664	1782.8	1161.7	621.1
10.1330	28.564	8.405	19.500	1.324	-0.665	1812.0	1177.6	634.4
10.1430	28.713	8.442	19.580	1.339	-0.649	1841.0	1192.0	649.0
10.1530	28.861	8.474	19.674	1.372	-0.659	1868.5	1213.6	654.9
10.1630	29.009	8.505	19.765	1.384	-0.644	1897.3	1223.2	674.0
10.1730	29.158	8.535	19.861	1.397	-0.636	1928.3	1235.0	693.3
10.1830	29.306	8.563	19.976	1.413	-0.647	1945.0	1249.1	695.9
10.1930	29.454	8.589	20.091	1.428	-0.654	1956.3	1261.4	694.9
10.2030	29.603	8.616	20.205	1.439	-0.658	1968.4	1271.3	697.1
10.2130	29.751	8.644	20.322	1.450	-0.664	1979.9	1279.6	700.3
10.2230	29.899	8.672	20.436	1.460	-0.669	1992.8	1287.9	704.8
10.2330	30.048	8.702	20.550	1.484	-0.689	2005.3	1308.6	696.7
11.0030	30.196	8.734	20.659	1.495	-0.692	2016.4	1317.0	699.4
11.0130	30.344	8.767	20.778	1.505	-0.707	2027.4	1325.3	702.1
11.0230	30.492	8.800	20.874	1.515	-0.697	2038.5	1333.4	705.0
11.0330	30.640	8.833	20.963	1.525	-0.681	2050.8	1341.3	709.4
11.0430	30.788	8.866	21.058	1.536	-0.672	2063.6	1349.3	714.4
11.0530	30.936	8.898	21.176	1.546	-0.683	2077.0	1356.9	720.1

11.0630	31.084	8.929	21.283	1.573	-0.701	2089.8	1373.7	716.1
11.0730	31.232	8.960	21.388	1.583	-0.697	2102.6	1381.1	721.6
11.0830	31.381	8.991	21.497	1.592	-0.699	2115.5	1388.1	727.3
11.0930	31.529	9.021	21.607	1.600	-0.700	2128.8	1394.9	733.9
11.1030	31.677	9.053	21.723	1.613	-0.712	2142.2	1406.3	735.9
11.1130	31.825	9.085	21.827	1.622	-0.709	2154.8	1413.5	741.3
11.1230	31.972	9.115	21.944	1.632	-0.719	2168.1	1420.6	747.5
11.1330	32.120	9.147	22.066	1.641	-0.733	2182.5	1427.8	754.7
11.1430	32.268	9.178	22.181	1.650	-0.741	2195.1	1435.0	760.1
11.1530	32.416	9.209	22.296	1.660	-0.749	2205.3	1442.2	763.2
11.1630	32.564	9.240	22.425	1.674	-0.776	2215.6	1453.8	761.8
11.1730	32.712	9.270	22.562	1.683	-0.804	2226.4	1460.9	765.5
11.1830	32.859	9.301	22.662	1.693	-0.796	2239.7	1468.1	771.6
11.1930	33.007	9.332	22.786	1.704	-0.815	2252.3	1476.6	775.7
11.2030	33.154	9.361	22.915	1.716	-0.839	2264.6	1485.6	779.0
11.2130	33.301	9.393	23.026	1.728	-0.845	2276.7	1494.6	782.1
11.2230	33.448	9.427	23.129	1.813	-0.921	2283.6	1568.1	715.6
11.2330	33.595	9.466	23.236	1.835	-0.943	2284.2	1592.1	692.1
			STONE CHANGE					
12.0030	33.739	9.505	23.340	1.860	-0.965	2299.8	1619.6	680.2
12.0130	33.879	9.536	23.439	1.880	-0.976	2317.8	1642.7	675.1
12.0230	34.018	9.565	23.459	1.901	-0.906	2337.2	1665.8	671.4
12.0330	34.157	9.594	23.556	1.921	-0.914	2358.0	1689.9	668.1
12.0430	34.297	9.622	23.662	1.941	-0.928	2381.7	1713.5	668.2
12.0530	34.437	9.648	23.773	1.961	-0.946	2405.9	1738.0	668.0
12.0630	34.576	9.674	23.876	1.967	-0.941	2428.0	1742.8	685.2
12.0730	34.716	9.700	23.982	1.973	-0.939	2450.4	1747.9	702.5
12.0830	34.854	9.723	24.078	1.979	-0.926	2476.0	1753.3	722.7
12.0930	34.992	9.747	24.178	1.986	-0.918	2499.4	1759.4	740.0
12.1030	35.130	9.771	24.275	1.993	-0.909	2521.3	1765.5	755.8
12.1130	35.267	9.794	24.385	1.999	-0.912	2537.7	1771.6	766.1
12.1230	35.404	9.817	24.497	2.007	-0.917	2556.8	1778.3	778.5
12.1330	35.541	9.839	24.601	2.016	-0.915	2576.8	1786.4	790.4
12.1430	35.679	9.859	24.703	2.025	-0.908	2595.9	1794.5	801.4
12.1530	35.816	9.880	24.813	2.033	-0.910	2618.3	1802.6	815.7
12.1630	35.953	9.902	24.925	2.046	-0.920	2640.9	1814.4	826.5

APPENDIX D: TABLE VI.

RUN 7: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
12.1730	36.091	9.924	25.019	2.059	-0.911	2667.6	1826.2	841.4
12.1830	36.229	9.948	25.121	2.072	-0.912	2695.0	1838.0	857.0
12.1930	36.367	9.973	25.210	2.087	-0.903	2720.4	1852.2	868.1
12.2030	36.504	9.997	25.308	2.105	-0.905	2741.7	1868.9	872.7
12.2130	36.645	10.020	25.411	2.123	-0.909	2763.5	1885.6	877.8
12.2230	36.781	10.043	25.514	2.138	-0.913	2790.4	1900.2	890.3
12.2330	36.925	10.067	25.613	2.145	-0.900	2814.9	1908.1	906.9
13.0030	37.063	10.090	25.713	2.153	-0.893	2834.9	1916.0	918.8
13.0130	37.200	10.115	25.814	2.160	-0.890	2855.6	1924.0	931.7
13.0230	37.337	10.145	25.916	2.168	-0.890	2878.0	1931.9	946.1
13.0330	37.475	10.173	26.029	2.175	-0.902	2893.9	1939.7	954.2
13.0430	37.612	10.198	26.119	2.183	-0.887	2911.7	1947.5	964.1
13.0530	37.750	10.223	26.224	2.190	-0.887	2928.6	1955.3	973.3
13.0630	37.888	10.250	26.326	2.198	-0.886	2951.8	1963.1	988.8
13.0730	38.027	10.277	26.422	2.206	-0.878	2970.1	1970.8	999.4
13.0830	38.166	10.302	26.504	2.213	-0.853	2994.1	1978.4	1015.7
13.0930	38.303	10.329	26.597	2.238	-0.862	3013.0	2000.7	1012.3
13.1030	38.439	10.355	26.683	2.280	-0.880	3032.1	2037.4	994.7
13.1130	38.576	10.380	26.783	2.321	-0.907	3043.1	2074.2	969.0
13.1230	38.712	10.403	26.890	2.341	-0.923	3052.0	2096.4	955.7
13.1330	38.848	10.429	26.993	2.353	-0.927	3058.0	2107.9	950.0
13.1430	38.983	10.455	27.096	2.365	-0.932	3065.0	2119.5	945.4
13.1530	39.119	10.481	27.194	2.374	-0.930	3072.0	2128.1	943.8
13.1630	39.254	10.508	27.289	2.383	-0.925	3080.1	2136.7	943.3
13.1730	39.387	10.533	27.381	2.391	-0.918	3090.0	2145.3	944.7
13.1830	39.525	10.560	27.473	2.405	-0.912	3100.8	2159.6	941.2
13.1930	39.662	10.587	27.559	2.419	-0.904	3112.4	2174.6	937.8
13.2030	39.798	10.616	27.661	2.428	-0.906	3120.7	2183.5	937.2

13.2130	39.935	10.645	27.765	2.433	-0.907	3129.1	2188.3	940.8
13.2230	40.072	10.674	27.869	2.438	-0.909	3138.8	2193.0	945.8
13.2330	40.209	10.702	27.975	2.443	-0.912	3146.9	2197.8	949.1
14.0030	40.346	10.731	28.074	2.449	-0.908	3154.2	2203.5	950.7
14.0130	40.483	10.760	28.166	2.457	-0.900	3162.5	2211.0	951.5
14.0230	40.619	10.791	28.256	2.465	-0.892	3170.9	2218.5	952.3
14.0330	40.756	10.820	28.353	2.473	-0.890	3178.4	2226.2	952.2
14.0430	40.892	10.849	28.461	2.481	-0.898	3185.7	2233.8	951.9
14.0530	41.029	10.882	28.554	2.489	-0.896	3193.5	2241.5	952.0
14.0630	41.167	10.914	28.657	2.497	-0.902	3202.1	2249.4	952.7
14.0730	41.304	10.945	28.759	2.506	-0.906	3209.7	2257.3	952.4
14.0830	41.441	10.976	28.855	2.514	-0.905	3217.2	2265.2	952.0
14.0930	41.578	11.008	28.954	2.522	-0.906	3224.5	2272.9	951.6
14.1030	41.715	11.041	29.049	2.530	-0.905	3232.3	2280.2	952.1
14.1130	41.852	11.073	29.151	2.538	-0.910	3239.8	2287.4	952.4
14.1230	41.990	11.106	29.249	2.546	-0.912	3246.8	2294.7	952.2
14.1330	42.127	11.139	29.349	2.554	-0.916	3254.4	2302.2	952.1
14.1430	42.264	11.171	29.440	2.563	-0.910	3261.9	2309.7	952.2
14.1530	42.401	11.204	29.536	2.571	-0.910	3269.2	2317.0	952.3
14.1630	42.538	11.237	29.623	2.579	-0.900	3276.7	2324.2	952.5
14.1730	42.675	11.267	29.711	2.585	-0.887	3285.9	2329.6	956.3
14.1830	42.813	11.294	29.797	2.589	-0.867	3297.8	2333.0	964.8
14.1930	42.950	11.325	29.895	2.592	-0.862	3306.7	2336.0	970.6
14.2030	43.087	11.357	29.992	2.596	-0.857	3314.2	2339.1	975.1
14.2130	43.226	11.393	30.087	2.599	-0.854	3322.8	2342.2	980.7
14.2230	43.364	11.427	30.182	2.603	-0.848	3331.7	2345.2	986.5
14.2330	43.500	11.460	30.282	2.625	-0.867	3338.2	2361.8	976.4
15.0030	43.636	11.496	30.382	2.642	-0.884	3344.9	2374.9	970.0
15.0130	43.774	11.530	30.496	2.653	-0.906	3352.7	2382.4	970.4
15.0230	43.912	11.565	30.592	2.664	-0.909	3362.7	2389.8	972.9
15.0330	44.050	11.601	30.679	2.675	-0.905	3371.6	2397.3	974.3
15.0430	44.187	11.636	30.764	2.687	-0.900	3381.0	2404.8	976.2
15.0530	44.324	11.670	30.846	2.700	-0.891	3391.3	2413.2	978.0
15.0630	44.462	11.702	30.928	2.714	-0.882	3400.7	2421.9	978.8
15.0730	44.599	11.733	31.010	2.729	-0.873	3409.3	2430.6	978.7
15.0830	44.736	11.764	31.094	2.744	-0.865	3421.4	2439.4	982.0

APPENDIX D: TABLE VI.

RUN 7: SULPHUR AND STONE CUMULATIVE BALANCE.

	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
DAY.HOUR	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
15.0930	44.874	11.795	31.176	2.760	-0.857	3431.9	2448.4	983.5
15.1030	45.012	11.826	31.260	2.776	-0.851	3442.2	2457.5	984.7
15.1130	45.149	11.855	31.350	2.795	-0.851	3453.5	2467.5	986.0
15.1230	45.287	11.882	31.471	2.814	-0.881	3464.5	2478.0	986.6
15.1330	45.424	11.912	31.598	2.833	-0.919	3473.7	2488.4	985.3
15.1430	45.561	11.940	31.719	2.851	-0.949	3484.5	2498.0	986.5
15.1530	45.698	11.967	31.839	2.866	-0.975	3494.7	2506.7	988.0
15.1630	45.835	11.995	31.962	2.881	-1.003	3504.1	2515.4	988.7
15.1730	45.973	12.025	32.077	2.897	-1.026	3514.4	2524.1	990.3
15.1830	46.110	12.056	32.190	2.913	-1.049	3524.4	2533.8	990.6
15.1930	46.248	12.087	32.306	2.932	-1.078	3533.2	2544.5	988.8
15.2030	46.385	12.119	32.402	2.950	-1.087	3544.8	2555.1	989.7
15.2130	46.523	12.151	32.494	2.969	-1.091	3554.5	2565.7	988.8
15.2230	46.659	12.183	32.595	2.987	-1.106	3564.0	2576.3	987.7
15.2330	46.796	12.212	32.694	3.000	-1.109	3575.0	2583.7	991.3
16.0030	46.933	12.242	32.788	3.011	-1.107	3584.2	2590.5	993.6
16.0130	47.069	12.272	32.912	3.023	-1.139	3592.8	2597.4	995.4
16.0230	47.205	12.299	33.027	3.032	-1.153	3602.8	2603.1	999.7
16.0330	47.342	12.329	33.139	3.041	-1.167	3613.0	2608.7	1004.3
16.0430	47.478	12.361	33.265	3.052	-1.200	3619.5	2615.3	1004.2
16.0530	47.614	12.392	33.381	3.065	-1.224	3627.3	2622.8	1004.5
16.0630	47.751	12.424	33.491	3.076	-1.240	3636.2	2630.0	1006.2
16.0730	47.888	12.453	33.601	3.088	-1.254	3644.5	2636.8	1007.7
16.0830	48.024	12.482	33.709	3.099	-1.267	3652.3	2643.7	1008.6
16.0930	48.160	12.511	33.813	3.110	-1.274	3662.6	2650.6	1012.0
16.1030	48.297	12.541	33.911	3.121	-1.276	3672.3	2657.5	1014.8
16.1130	48.433	12.570	34.014	3.130	-1.281	3681.4	2663.3	1018.1
16.1230	48.570	12.598	34.110	3.139	-1.277	3690.6	2668.9	1021.7

16.1330	48.707	12.627	34.210	3.148	-1.278	3699.8	2674.4	1025.4
16.1430	48.843	12.656	34.313	3.157	-1.283	3707.0	2680.0	1027.0
16.1530	48.980	12.686	34.413	3.167	-1.286	3714.3	2686.1	1028.2
16.1630	49.117	12.716	34.512	3.176	-1.287	3724.0	2692.3	1031.7
16.1730	49.255	12.746	34.608	3.186	-1.285	3734.2	2698.4	1035.8
16.1830	49.393	12.775	34.709	3.197	-1.288	3742.3	2705.3	1037.0
16.1930	49.531	12.804	34.806	3.208	-1.287	3749.6	2712.2	1037.4
16.2030	49.668	12.835	34.906	3.219	-1.291	3758.5	2719.0	1039.4
16.2130	49.806	12.867	35.008	3.233	-1.302	3768.5	2728.1	1040.4
16.2230	49.943	12.902	35.108	3.248	-1.315	3775.7	2737.1	1038.6
16.2330	50.081	12.940	35.212	3.259	-1.331	3784.1	2744.4	1039.7
17.0030	50.218	12.979	35.318	3.271	-1.349	3791.9	2751.6	1040.3
17.0130	50.355	13.018	35.410	3.284	-1.357	3799.7	2759.9	1039.8
17.0230	50.492	13.058	35.506	3.300	-1.371	3806.5	2769.2	1037.2
17.0330	50.628	13.092	35.616	3.312	-1.392	3812.1	2776.7	1035.4
17.0430	50.766	13.130	35.708	3.321	-1.394	3819.1	2782.4	1036.7
17.0530	50.902	13.167	35.795	3.331	-1.391	3826.9	2788.2	1038.8
17.0630	51.040	13.203	35.883	3.340	-1.387	3834.2	2794.0	1040.2
17.0730	51.177	13.239	35.993	3.355	-1.409	3843.1	2802.7	1040.4
17.0830	51.314	13.273	36.074	3.369	-1.402	3853.3	2811.4	1041.9
17.0930	51.451	13.307	36.166	3.393	-1.414	3861.4	2828.0	1033.5
17.1030	51.588	13.342	36.261	3.401	-1.416	3866.3	2833.5	1032.7
17.1130	51.725	13.378	36.35	3.410	-1.416	3873.8	2839.1	1034.7
17.1230	51.863	13.414	36.446	3.417	-1.414	3881.9	2843.5	1038.4
17.1330	52.001	13.450	36.544	3.423	-1.417	3890.5	2847.4	1043.1
17.1430	52.137	13.486	36.647	3.454	-1.451	3896.7	2869.7	1027.0
17.1530	52.274	13.523	36.747	3.460	-1.456	3908.3	2873.7	1034.6
17.1630	52.407	13.555	36.838	3.498	-1.485	3923.4	2898.0	1025.4
17.1730	52.538	13.586	36.936	3.501	-1.486	3938.7	2900.2	1038.5
17.1830	52.669	13.617	37.040	3.505	-1.493	3954.6	2902.4	1052.2
17.1930	52.800	13.648	37.143	3.508	-1.499	3964.6	2904.7	1060.0
17.2030	52.930	13.679	37.239	3.511	-1.499	3972.7	2906.9	1065.8
17.2130	53.060	13.710	37.328	3.515	-1.492	3982.7	2909.1	1073.6

APPENDIX D - TABLE VII

ANALYSIS OF SOLIDS REMOVED DURING RUN 7
TOTAL SULPHUR WT. PERCENT

DAY.HOUR	GASIFIER	REGEN.	REGEN. CYCLONE	ELUTR. FINES	BOILER BACK	BOILER FLUE	ELUTR. COARSE	STACK K.O.
2.1600	3.24	2.48	5.83	4.79	3.81	2.83	5.11	-
4.0800	3.44	2.17	-	4.40	3.57	2.64	4.01	2.90
5.0200	3.45	2.22	4.98	3.19	3.78	2.98	4.43	-
5.1300	-	2.62	4.58	3.57	3.45	2.42	3.67	-
6.0630	5.41	2.84	4.34	3.21	3.92	2.52	4.18	-
6.1800	5.62	3.46	-	2.96	4.91	3.32	5.43	3.35
7.0400	5.58	4.12	4.01	4.34	4.27	2.65	4.59	2.86
8.0500	5.97	4.61	4.19	4.28	4.61	2.79	5.60	3.22
10.0900	4.33	3.64	6.45	4.43	4.20	2.45	5.30	-
11.0700	4.30	2.41	5.49	2.01	4.71	2.88	5.53	-
11.1900	4.22	2.60	4.82	4.18	4.90	2.79	5.30	4.36
13.0100	3.35	1.89	4.12	3.68	3.37	2.18	4.88	2.66
13.1000	3.22	2.07	4.08	4.00	3.42	2.40	5.19	3.28
13.1900	4.02	3.03	4.86	4.58	2.66	3.41	5.07	3.96
14.2000	5.35	3.82	5.70	3.12	3.18	3.89	5.99	5.29
15.1100	6.39	5.13	5.25	3.61	5.79	4.87	6.29	-
15.1800	5.63	3.78	5.27	3.26	5.50	4.06	5.88	-
16.1500	4.97	3.76	4.71	2.99	4.85	3.81	5.30	-
17.1300	6.04	4.32	5.10	3.63	-	3.71	5.87	-
17.1800	5.50	3.94	4.93	4.32	5.28	2.88	6.02	-

APPENDIX D - TABLE VIII

ANALYSIS OF SOLIDS REMOVED DURING RUN 7
S U L P H A T E S U L P H U R WT. PERCENT

DAY.HOUR	GASIFIER	REGEN.	REGEN. CYCLONE	ELUTR. FINES	BOILER BACK	BOILER FLUE	ELUTR. COARSE	STACK K.O.
2.1600	0.16	1.00	2.06	0.73	2.45	2.07	0.19	-
4.0800	0.14	0.78	-	0.77	1.50	2.06	0.13	1.73
5.0200	0.14	0.63	1.31	0.63	1.36	1.99	0.12	-
5.1300	0.22	0.81	1.33	0.79	1.27	1.94	0.23	-
6.0630	0.15	0.51	1.00	0.48	1.08	2.06	0.25	-
6.1800	0.29	0.77	-	0.46	1.32	2.22	0.26	1.90
7.0400	0.18	0.76	1.51	0.64	1.21	2.02	0.23	1.62
8.0500	0.24	0.77	1.86	0.64	1.54	1.94	0.16	1.66
10.0900	0.23	0.80	4.34	1.14	1.44	1.78	0.30	-
11.0700	0.17	0.93	2.87	0.66	1.08	1.59	0.31	-
11.1900	0.13	0.80	2.30	0.57	0.99	1.80	0.29	1.42
13.0100	0.21	0.95	1.95	0.39	0.72	1.60	0.17	1.00
13.1000	0.11	0.76	2.08	0.28	3.21	1.56	0.11	1.13
13.1900	0.12	0.74	1.92	0.44	2.36	1.75	0.16	1.84
14.2000	0.21	0.78	2.60	0.29	2.82	2.55	0.18	2.25
15.1100	0.15	-	2.56	0.32	1.46	1.59	0.20	-
15.1900	0.14	0.87	2.04	0.21	1.62	2.46	0.17	-
16.1500	0.19	0.62	1.53	0.32	2.84	2.21	0.21	-
17.1300	0.14	0.54	1.62	0.26	-	2.17	0.14	-
17.1800	0.22	0.73	1.54	0.21	0.92	1.57	0.16	-

APPENDIX D - TABLE IX

ANALYSIS OF SOLIDS REMOVED DURING RUN 7
T O T A L   C A R B O N   WT. PERCENT

DAY.HOUR	GASIFIER	REGEN.	REGEN. CYCLONE	ELUTR. FINES	BOILER BACK	BOILER FLUE	ELUTR. COARSE	STACK K.O.
2.1600	0.05	0.	2.92	17.20	0.18	4.12	0.40	-
4.0800	0.55	0.	-	9.13	0.03	3.49	1.65	0.78
5.0200	0.51	0.04	0.71	11.00	0.46	2.67	2.67	-
5.1300	0.39	0.	2.08	11.30	0.92	4.99	1.41	-
6.0630	0.25	0.	1.53	11.70	0.30	4.17	1.48	-
6.1800	0.17	0.	-	8.04	0.40	2.99	2.00	1.04
7.0400	1.54	0.	0.48	15.00	0.31	5.50	0.20	1.54
8.0500	0.15	0.	0.51	8.37	0.44	2.25	0.33	0.70
10.0900	0.08	0.06	0.80	17.07	0.41	2.15	1.52	-
11.0700	0.06	0.	0.74	22.90	0.22	5.42	6.08	-
11.1900	0.29	0.	0.70	23.40	0.23	6.60	8.49	1.78
13.0100	0.05	0.05	0.24	20.90	0.11	2.00	1.90	0.40
13.1000	0.06	0.	0.44	16.18	0.	5.00	4.20	1.02
13.1900	0.14	0.	1.10	21.90	0.	12.10	4.64	3.46
14.2000	0.28	0.	3.76	22.50	0.	3.25	2.56	0.75
15.1100	0.24	0.	1.32	22.80	0.56	1.75	3.22	-
15.1800	0.41	0.02	0.30	23.20	0.46	3.12	3.72	-
16.1500	0.26	0.03	0.32	28.60	0.38	8.62	5.88	-
17.1300	0.49	0.	1.52	24.40	-	6.69	3.79	-
17.1800	0.30	0.16	1.40	14.00	0.58	3.97	3.94	-

APPENDIX D - TABLE X
SOLIDS REMOVED DURING RUN 7, KG. (RAW DATA)

DAY.HOUR	GASIFIER	REGEN.	REGEN. CYCLONE	ELUTR. FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
1.1900	-	-	-	2.75	-	35.00	-
2.0135	-	41.00	-	-	-	7.60	-
2.0450	-	41.00	2.25	1.50	-	10.30	-
2.0750	-	38.00	-	-	-	-	-
2.0800	-	39.00	5.25	2.50	-	36.50	-
2.1040	-	42.50	-	-	-	-	-
2.1200	-	-	-	2.75	-	29.50	-
2.1600	-	61.00	2.00	1.50	-	20.00	-
2.2000	-	32.80	-	1.50	-	18.50	-
3.0745	-	-	5.50	2.50	-	40.00	-
3.1000	-	-	-	-	-	40.00	-
3.1120	-	-	-	1.00	5.00	20.00	-
3.1545	-	-	-	1.20	13.00	48.00	-
3.1900	-	-	-	1.50	19.00	41.00	-
3.2200	-	-	-	1.00	22.00	45.00	-
4.0030	-	-	-	-	15.00	26.00	-
4.0330	-	-	-	1.50	12.00	25.00	-
4.0530	8.00	-	-	-	-	-	-
4.0700	-	-	7.00	-	-	30.00	-
4.0800	-	-	-	1.00	13.00	-	5.00
4.1130	-	-	-	2.00	7.00	14.00	-
4.1745	-	-	-	3.00	22.00	42.00	-
4.2230	-	-	-	-	-	26.00	-
4.2330	-	-	-	2.00	12.00	-	-
5.0200	-	-	1.00	0.50	7.00	24.00	3.00
5.0530	-	-	-	-	-	27.00	-
5.0730	-	-	-	2.50	9.00	20.00	-
5.1150	-	-	-	2.00	14.00	35.00	-
5.1130	-	37.00	-	-	-	-	-
5.1600	-	-	-	2.00	12.00	31.00	-
5.2000	-	-	-	3.00	13.00	30.00	-
5.2330	-	-	1.00	2.50	11.00	29.00	-
6.0330	-	-	-	-	11.00	24.00	-
6.0630	4.50	4.50	1.00	3.50	12.00	29.00	-
6.1100	-	-	0.50	2.50	-	-	-
6.1300	4.50	4.50	-	-	29.00	24.00	-
6.1600	-	-	-	-	-	22.00	-
6.1700	4.50	4.50	-	-	-	-	-
6.2045	-	-	0.50	5.50	4.00	29.00	-
6.2120	4.50	4.50	-	-	-	-	-

SOLIDS REMOVED DURING RUN 7, KG. (RAW DATA)

DAY.HOUR	GASIFIER	REGEN.	REGEN. CYCLONE	ELUTR. FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
7.0140	4.50	4.50	-	-	-	-	-
7.0200	-	-	0.25	4.25	17.25	36.00	-
7.0400	4.50	4.50	-	1.00	-	-	-
7.0620	-	-	-	-	-	19.50	-
7.0650	-	-	-	1.50	12.00	-	-
7.1000	4.00	4.00	-	-	-	-	-
7.1700	4.00	4.00	0.50	7.75	12.75	42.00	-
7.2110	4.00	4.00	-	-	-	40.00	-
7.2300	-	-	-	4.50	24.75	57.00	-
8.0400	-	-	0.75	4.00	22.00	54.50	-
8.0500	-	-	-	-	10.75	12.75	-
8.0800	4.00	4.00	-	1.00	11.50	32.50	-
8.2300	9.00	9.00	-	-	-	40.00	8.00
9.0715	-	-	0.50	5.00	0.50	54.00	0.
9.0830	-	-	-	0.50	1.00	47.00	-
9.1230	-	-	-	-	-	20.00	-
9.1500	-	-	-	-	-	42.00	-
9.1700	-	-	-	3.50	-	36.00	-
9.1830	-	-	-	-	22.00	-	-
9.2000	-	-	-	3.50	12.00	27.00	-
9.2300	-	-	-	2.00	20.00	38.00	-
10.0245	-	-	-	3.00	32.00	48.00	-
10.0545	-	-	-	1.00	24.00	47.00	-
10.0800	-	-	-	1.00	16.00	20.00	-
10.0945	-	-	-	-	-	22.00	-
10.1030	-	-	-	-	15.00	48.00	-
10.1230	-	-	-	-	-	20.00	-
10.1415	-	-	-	-	44.00	45.00	-
10.1700	-	-	-	-	-	27.00	-
10.1730	-	-	-	-	41.00	-	20.00
10.1900	-	-	-	-	-	40.00	-
10.2000	-	-	-	-	33.00	-	-
10.2030	-	-	-	3.00	-	17.00	-
10.2130	-	-	3.50	-	-	-	-
10.2300	-	-	-	1.00	30.50	22.75	-
11.0200	5.00	5.00	-	2.50	29.00	28.00	-
11.0500	-	-	-	1.00	31.00	24.00	-
11.0800	-	-	-	2.50	28.00	22.00	23.00
11.1000	-	-	-	1.50	13.00	18.00	-
11.1245	5.00	5.00	-	1.00	21.00	25.00	-

SOLIDS REMOVED DURING RUN 7, KG. (RAW DATA)

DAY.HOUR	GASIFIER	REGEN.	REGEN. CYCLONE	ELUTR. FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
11.1845	5.00	5.00	-	3.00	50.00	50.00	-
11.2200	-	-	-	3.00	36.00	31.00	-
12.0045	100.00	50.00	-	-	-	-	-
12.0200	-	42.00	-	-	-	-	-
12.0245	-	50.00	-	2.00	32.00	28.00	-
12.0330	-	40.00	-	-	-	-	-
12.0405	-	40.00	-	-	-	-	-
12.0500	-	40.00	-	-	-	-	-
12.0540	-	45.00	-	1.00	14.00	11.00	-
12.0645	-	45.00	-	-	-	-	-
12.0740	-	47.00	-	-	-	-	-
12.0810	-	-	-	2.00	14.00	14.00	-
12.1210	-	-	-	1.00	28.00	28.00	-
12.1530	-	-	-	0.50	37.00	25.00	-
12.1900	-	-	6.00	0.50	64.00	31.00	-
12.2215	-	-	-	-	93.00	30.00	-
13.0900	-	-	8.00	-	101.00	90.00	-
13.1130	-	-	-	-	212.00	8.00	-
13.1430	-	24.00	1.00	-	67.00	16.00	-
13.1800	-	-	0.50	1.50	55.00	-	-
13.1900	-	-	-	-	42.00	22.00	-
13.2015	-	-	-	-	25.00	-	-
14.0010	-	-	-	-	27.50	-	-
14.0200	-	-	-	-	-	30.00	-
14.0315	-	-	2.00	-	-	-	-
14.0500	-	-	-	-	65.00	-	-
14.0915	-	-	-	-	60.00	34.00	-
14.1245	-	-	-	-	46.00	-	-
14.1400	-	-	-	4.00	-	19.50	-
14.1700	-	-	-	-	60.00	-	-
14.1800	-	-	-	-	-	12.50	-
14.2300	-	-	-	1.00	32.50	9.00	-
14.2345	-	-	-	-	50.00	-	-
15.0445	-	-	-	-	65.00	26.00	-
15.0820	-	-	2.00	3.00	-	-	-
15.1045	-	-	-	1.00	84.00	38.00	-
15.1400	-	-	-	1.00	63.00	14.00	-
15.1800	-	-	7.00	1.00	60.00	19.00	-
15.2015	-	-	-	1.00	44.00	11.00	-
15.2240	-	-	-	-	50.00	-	-

SOLIDS REMOVED DURING RUN 7, KG. (RAW DATA)

DAY.HOUR	GASIFIER	REGEN.	REGEN. CYCLONE	ELUTR. FINES	BOILER BACK	BOILER FLUE	ELUTR COARSE
15.2330	-	-	-	-	-	12.00	-
16.0130	-	-	-	-	33.00	-	-
16.0400	-	-	-	-	22.00	18.00	-
16.0600	-	-	-	-	26.00	-	-
16.0700	-	-	7.00	4.00	-	13.00	-
16.0800	-	-	-	-	23.00	-	-
16.1045	-	-	-	5.00	32.00	12.00	-
16.1415	-	-	-	1.50	32.00	11.00	-
16.1730	-	-	-	1.50	33.00	12.00	-
16.2030	-	-	11.00	1.00	34.00	13.00	-
16.2230	-	-	-	-	32.00	-	-
17.0100	-	-	-	-	29.00	23.00	-
17.0300	-	-	-	-	38.00	-	-
17.0500	-	-	-	-	-	10.00	-
17.0630	-	-	-	-	36.00	-	-
17.0830	-	-	7.00	4.00	34.00	10.00	-
17.1145	-	25.00	3.00	2.50	21.00	18.00	-
17.1615	-	42.00	-	1.00	28.00	13.00	-
17.1820	-	31.00	-	-	-	-	17.20
17.2200	-	-	-	-	15.00	-	-
20.1200					335.66	-	-
25.1200	289.51	15.42	-	-	-	-	-

APPENDIX D - TABLE XI

RUN 7 - STONE FEED
SIEVE SIZE IN MICRONS

SAMPLE NUMPER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
				WT. PERCENT.						
50926	1.0830	8.9	24.8	5.3	9.3	8.7	29.8	7.3	2.6	3.2
50926	1.1800	8.2	56.1	10.3	18.4	5.5	.7	.1	.1	.7
50926	1.2145	.9	57.8	11.6	21.5	6.1	1.3	.3	.2	.4
50926	2.0030	.8	56.4	12.5	22.7	5.8	.8	.2	.0	.9
50926	3.0900	3.7	63.4	10.1	18.2	4.0	.0	.0	.0	.5
50926	3.1200	1.9	63.2	10.0	18.0	5.8	.7	.1	.0	.2
50926	3.1500	1.6	57.5	10.5	21.6	7.5	.8	.1	.0	.3
50926	3.1600	1.1	64.8	9.8	17.4	5.2	.8	.1	.1	.6
50939	4.0600	.6	60.9	11.5	22.1	4.1	.5	.0	.0	.3
50938	4.0130	1.2	64.6	9.9	18.3	5.3	.4	.0	.0	.2
50926	4.1200	1.7	62.9	9.2	18.4	6.8	.7	.0	.0	.3
50948	4.2330	1.3	60.5	12.1	20.3	4.7	.7	.2	.0	.3
51007	5.0303	.8	57.7	9.7	23.0	7.8	.5	.1	.1	.3
51008	5.0730	.7	54.6	17.7	21.0	4.8	.6	.1	.1	.3
51017	6.0300	1.7	47.3	9.4	17.1	8.1	7.3	1.5	.1	7.5
50926	6.1315	.6	56.8	13.4	21.8	6.5	.5	.1	.0	.3
50926	6.1715	1.0	70.1	3.7	18.1	5.7	.8	.0	.0	.6
50926	6.1715	1.2	61.8	11.7	18.8	5.3	1.0	.3	.1	.0
51035	6.1800	1.1	72.9	9.0	15.9	.3	.4	.0	.1	.2
51044	7.0400	.9	45.8	15.6	20.8	8.3	4.3	.5	.0	3.6
51026	7.0510	.0	21.9	11.3	21.5	19.1	25.7	.0	.0	.4
51026	8.0230	.0	19.0	12.6	24.4	19.7	23.9	.0	.0	.4
51026	8.0620	.0	22.4	11.4	22.2	17.9	25.8	.0	.0	.2
51026	9.0200	.0	20.6	11.2	22.3	19.0	26.5	.0	.0	.5
51055	9.1030	.1	24.4	15.9	29.9	19.2	7.6	1.2	.1	1.7
51057	10.0900	.0	48.2	11.2	21.8	10.1	5.8	1.5	.4	1.0
50164	11.0700	.1	33.2	12.4	23.1	17.9	13.2	.0	.1	.1
51026	11.1600	.6	47.3	13.5	27.9	8.9	1.0	.2	.0	.6
50169	11.0700	.9	57.8	12.2	20.7	6.5	1.2	.2	.1	.4
50927	11.1600	.4	44.5	14.9	29.3	9.4	1.1	.2	.0	.2
51082	11.1900	1.9	66.6	10.3	16.0	4.7	.3	.0	.1	.1
51090	12.0100	.0	9.2	9.2	22.3	20.5	34.8	2.3	1.6	.0
50927	12.1145	.0	12.6	12.1	29.4	17.0	20.4	5.0	.0	3.4
51100	13.1900	.0	8.1	9.8	29.5	24.7	25.3	.8	.2	1.5
51161	15.1800	.0	15.0	11.1	31.9	17.4	23.1	.8	.2	.6
51153	15.1100	.0	12.5	9.6	19.9	19.6	35.1	1.4	.2	1.6
50927	15.1500	.0	17.0	12.8	23.8	18.6	26.3	.9	.4	.3
51166	16.1500	.0	9.5	9.6	21.1	22.4	35.5	1.4	.1	.5
51127	17.1300	.1	10.7	9.5	21.7	19.9	35.1	2.2	.2	.6

APPENDIX D - TABLE XII

RUN 7 - GASIFIER BED

SIEVE SIZE IN MICRONS

SAMPLE NUMBER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
					WT. PERCENT.					
51001	5.0200	.5	37.0	14.1	20.3	16.1	12.0	.0	.0	.0
51009	5.1300	.0	30.1	16.9	21.1	17.2	14.7	.0	.0	.0
51027	6.1800	.0	28.9	10.7	18.5	20.8	20.9	.0	.0	.1
51036	7.0400	.0	23.4	10.2	19.2	21.8	25.2	.0	.0	.2
51061	10.0900	.2	27.2	11.7	20.0	18.5	22.2	.0	.0	.2
51133	13.1900	.0	22.4	15.8	27.6	19.7	14.5	.0	.0	.0
51146	15.1100	.0	16.2	13.8	27.9	23.8	18.4	.0	.0	.0
51156	15.1800	.0	16.1	13.8	28.6	23.4	18.0	.0	.0	.0
51146	15.1100	.0	16.1	14.3	28.4	23.4	17.8	.0	.0	.0
51165	16.1500	.0	15.0	13.9	28.6	23.6	18.8	.1	.0	.0
51184	17.1800	.0	12.5	11.7	27.7	24.4	23.6	.0	.0	.2
51170	17.1300	.0	13.2	13.2	22.9	26.1	24.7	.0	.0	.0

APPENDIX D - TABLE XIII

RUN 7 - REGENERATOR BED

SIEVE SIZE IN MICRONS

SAMPLE NUMBER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
					WT. PERCENT.					
50941	4.0800	.0	36.6	12.8	21.2	15.8	13.6	.0	.0	.0
51060	10.0900	.3	26.6	10.7	19.7	19.1	23.3	.0	.0	.3
50170	11.0700	.0	34.2	12.8	22.0	17.7	13.0	.3	.0	.0
51147	15.0000	.0	17.1	13.6	28.2	23.4	17.6	.1	.0	.1
51157	15.1800	.0	15.7	13.6	28.5	23.0	19.1	.0	.0	.1
51164	16.1500	.0	13.9	12.8	28.8	23.9	20.4	.0	.0	.1
51171	17.1300	.0	12.9	12.8	27.6	24.4	22.1	.0	.1	.1
51185	17.1800	.1	13.8	13.1	27.0	23.5	22.4	.0	.0	.1

APPENDIX D - TABLE XIV
RUN 7 - BOILER BACK END

SIEVE SIZE IN MICRONS

SAMPLE NUMBER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
		WT. PERCENT.								
51058	10.0900	.0	5.1	2.7	4.4	9.3	62.8	.2	.2	15.2
51067	11.0700	.0	.2	1.3	11.4	26.3	57.0	.9	.0	2.8
51094	13.1000	.0	21.9	12.9	21.7	14.1	22.1	2.6	1.3	3.3
51163	16.1500	.0	4.5	5.0	15.2	19.0	45.9	5.5	2.7	2.1
51173	17.1300	.0	3.0	3.8	11.7	21.1	53.3	4.1	1.3	1.7
51182	17.1800	.1	1.4	1.8	8.6	19.3	67.7	.2	.4	.4

APPENDIX D - TABLE XV
RUN 7 - ELUTRIATOR COARSE
SIEVE SIZE IN MICRONS

SAMPLE NUMBER	DAY-TIME	3200 2800	2800 1400	1400 1180	1180 850	850 600	600 250	250 150	150 100	100
					WT. PERCENT.					
50947	4.0800	.2	12.8	6.0	12.1	14.5	43.8	3.1	2.9	4.8
50950	5.0200	.2	6.4	3.0	2.7	12.4	36.9	8.8	.2	29.4
51015	5.1300	.1	7.9	3.9	7.9	7.1	42.6	3.0	1.0	26.6
51024	6.0630	.0	3.3	1.3	3.3	5.7	2.3	1.3	18.0	64.7
51038	7.0400	.0	3.8	1.7	3.5	6.7	66.5	.0	.8	17.1
51062	10.0000	.1	16.9	6.6	11.2	11.7	42.4	.6	.1	10.3
51068	11.0700	.0	7.9	3.7	9.7	13.8	40.9	.3	2.9	20.9
51126	13.1900	.0	2.1	2.3	6.7	10.2	59.3	.1	.0	19.4
51083	13.0100	1.4	2.2	1.7	5.2	7.5	75.2	.0	3.2	3.5
51148	15.1100	.1	3.6	4.1	11.0	14.2	50.2	.0	1.3	15.5
51160	15.1800	.1	4.0	4.6	11.9	14.6	54.0	.1	5.5	5.1
51168	16.1500	.0	2.0	2.8	7.6	11.4	54.6	.0	2.3	19.3
51178	17.1800	.0	.9	1.3	3.9	6.5	37.9	.2	14.3	35.0
51062	10.0900	.0	14.3	5.5	10.3	10.3	33.7	8.5	5.1	12.4

APPENDIX D - TABLE XVI

CAFB Stack Cyclone Fines (Sample 1)

Dry sieving with 200 mesh sieve:

+74 microns = 18.70 gm = 19.06%
-74 microns = 79.41 gm = 80.94%

Total Sample= 98.11 gm =100.00%

BAHCO Analysis: - 74 micron fraction

Throttle No.	Upper Size Microns	Weight gm	Per Cent	Cumulative % less than stated size
18	1.2	0.25	0.25	0.25
17	1.9	1.65	1.68	1.94
16	4.2	9.15	9.33	11.26
14	7.0	11.80	12.03	23.29
12	10.0	11.60	11.82	35.11
8	18.0	15.20	15.49	50.61
4	30.0	5.10	5.20	55.80
-	74.0	24.66	25.14	80.94
			80.94%	
		+74 microns	19.06%	
		Total =	100.00%	

APPENDIX D - TABLE XVI

CAFB Stack Cyclone Fines (Sample 2)

Dry sieving with 200 mesh sieve:

+74 microns	=	21.84 gm	=	21.67%
-74 microns	=	78.93 gm	=	78.33%
Total Sample	=	100.77 gm	=	100.00%

BAHCO Analysis: -74 micron fraction

Throttle No.	Upper Size Microns	Weight gm	Per Cent	Cumulative % less than stated size
18	1.2	0.24	0.24	0.24
17	1.9	1.74	1.73	1.97
16	4.2	9.64	9.57	11.53
14	7.0	12.01	11.92	23.45
12	10.0	12.28	12.19	35.64
8	18.0	16.05	15.93	51.56
4	30.0	5.32	5.28	56.84
-	74.0	21.65	21.48	78.33
			78.33%	
		+74 Microns	21.67%	
		Total =	100.00%	

C. A. F. B. RUN 7

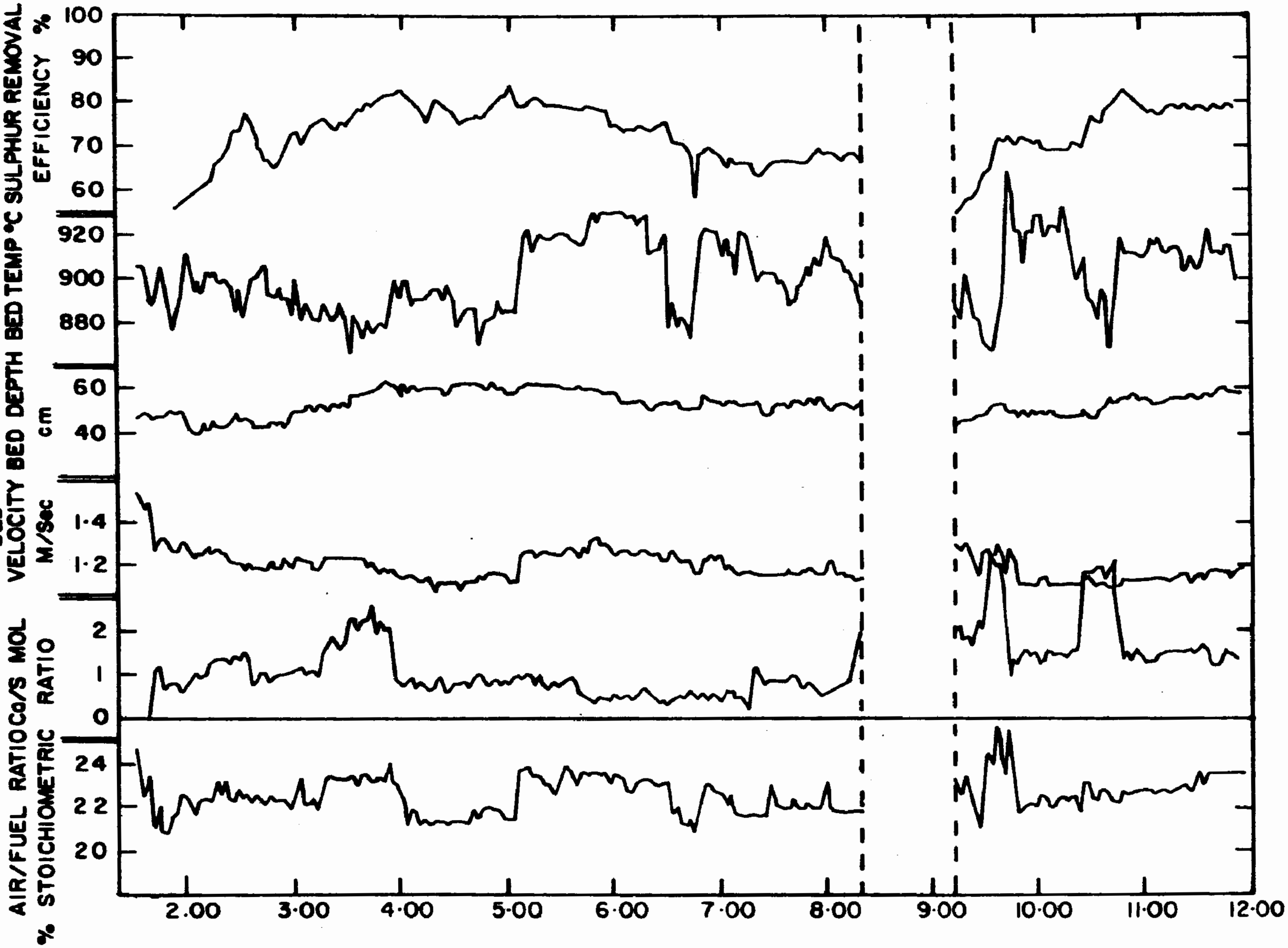


FIG. D25 SHEET. I.

C.A.F.B. RUN 7 (Contd)

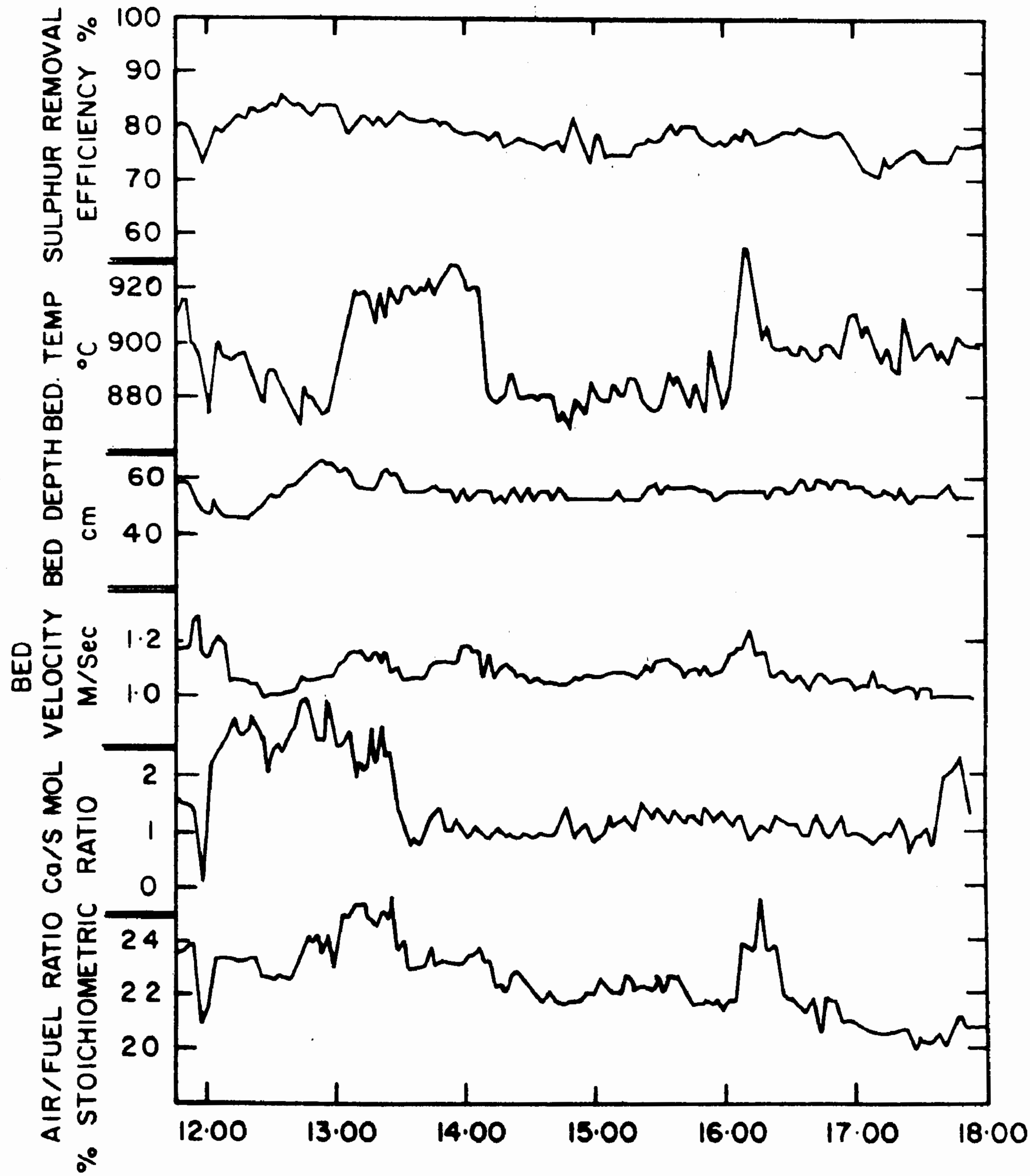


FIG. D25 SHEET 2

C.A.F.B. CYCLONE FINES

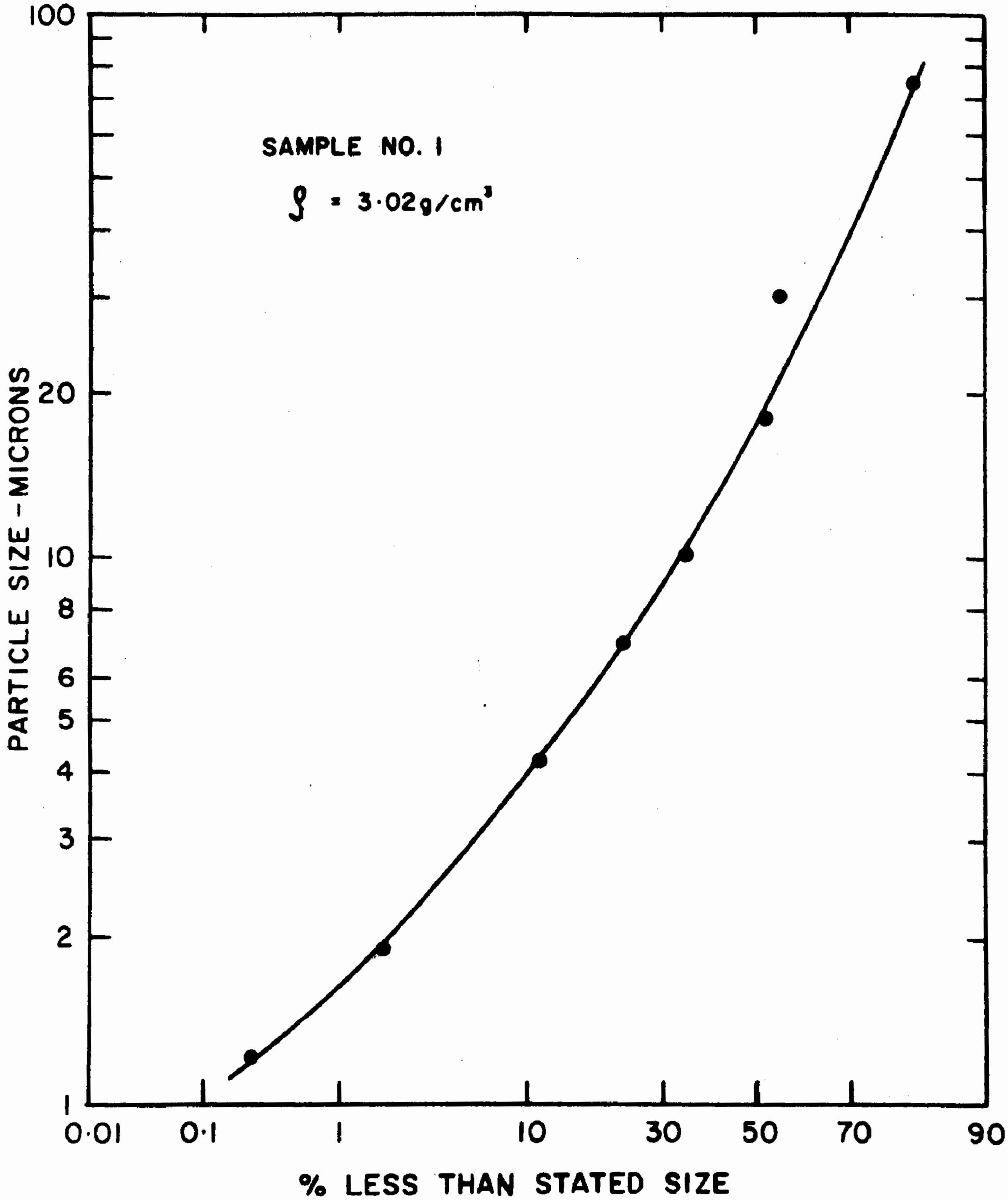


FIG. D26

C. A. F. B. CYCLONE FINES

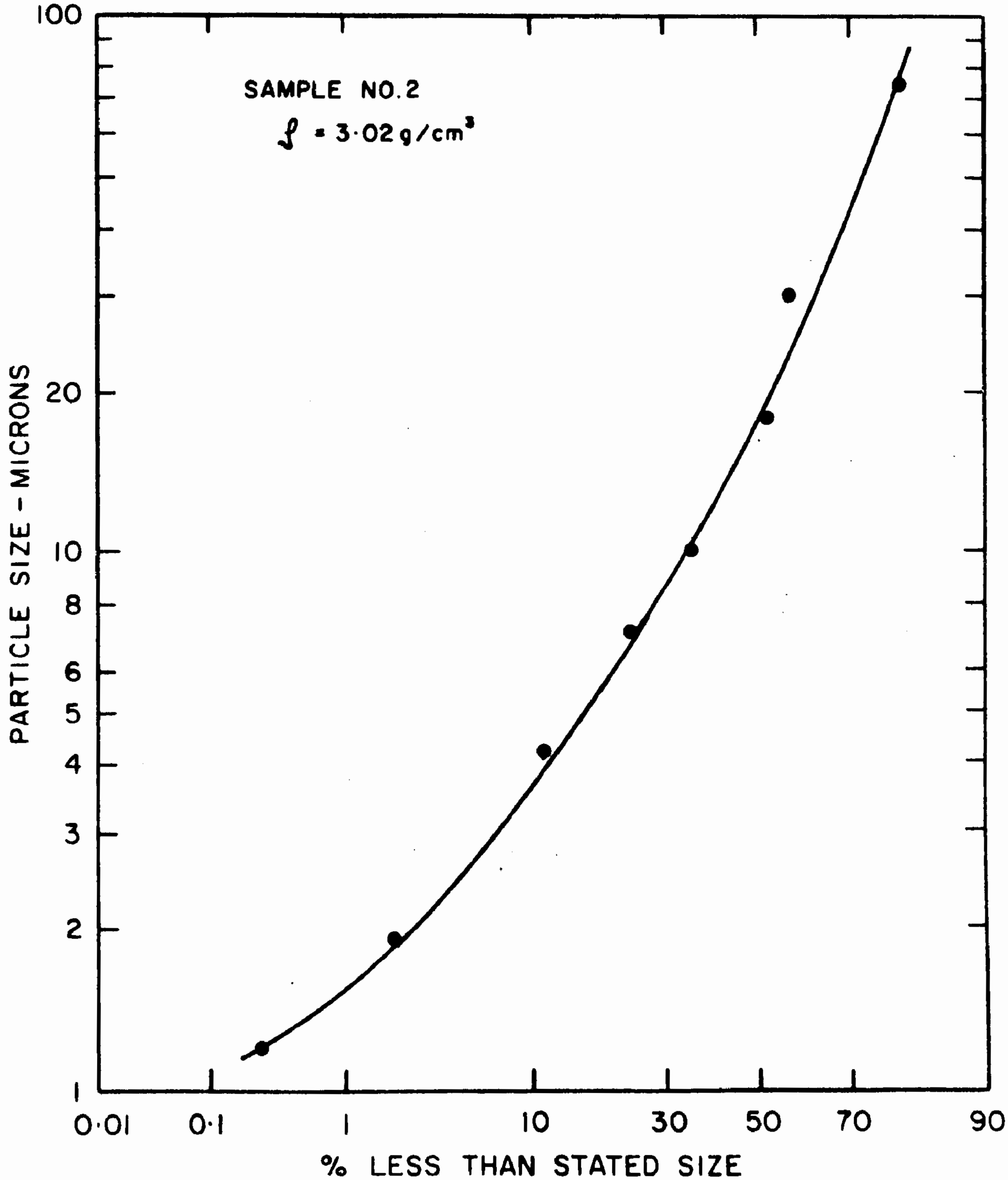


FIG. D27.

APPENDIX E

CAFB PILOT PLANT OPERATING PROCEDURES

A. PREPARATION FOR RUNNING

1. Operation of Services

(a) Oil Ring Main

- Check tank contents.
- Check outflow temperature is about 140/150°F.
- Start compressor/oil pump on boiler by switching on at panel. Do not run for more than 5 seconds at first few runs in order to set thick oil moving. Continue until temperature gauge on outflow heater reaches 200/210°F and then pump may be left on.
- The boiler heater circuit is now on automatic operation and is ready for use.

(b) Kerosene

- Stored in 500 gal. tank - check that there is enough for anticipated usage. First check that isolating valve is turned off on pump supply feed pipe. Then open valve at barrel.

(c) Nitrogen

- Stored in liquid N_2 tank check that there is sufficient for anticipated usage.
- Check all valves are turned off at the bleed locations on gasifier before opening valve on manifold.

(d) Propane

- Stored in 2 banks of cylinders outside building.

- Commence runs on the bank with least available so that replacement bottles may be ordered in reasonable time. Warn Purchasing at least 7 days prior to anticipated use and thereafter replacements can be obtained on 48 hours notice. Only turn on outside valve immediately prior to use for safety reasons.

(e) Boiler and Cooling System

(1) Cooling System

- Open drain valve on cooling pump delivery which bleeds water to waste.

- Go up on roof and check valve is open on water feed to cooler - listen to check that water is flowing. If not - check all valves back to tower.

- Turn off valve on drain to conserve water until heating takes place.

- Warn Services of an anticipated soft water usage - maximum of 500 gph. Please give as much notice as possible.

- Turn on secondary side i.e. cooler pump - hold button in for at least 10 seconds to prevent a shutdown alarm.

- Turn on boiler circulating pump. No button to hold on this circuit.

- Turn on supply to cooler fan located on end wall (main stores end) n.b. fan is thermostatically controlled and will not cut in until pond water reaches 100°F.

- Check temperature setting of automatic mixing valve is set to 180°F and turn on electrical supply to valve. (Controller set on wall by pressurisation unit).

(2) Pressurisation System

- Check water level in storage tank is at rubber band marker and make up is turned on.

- Check nitrogen bottle pressure is above 500 psi.

- Turn on pressurisation system and check that pressure reaches 50 psi approximately. On start up bell will ring, cancel bell, then turn off N2 at cylinder.

(3) Boiler

- Check that main flue butterfly valve is at marked position. (Do not take any notice of 'open' and 'shut' marked positions as this was incorrectly fabricated).

- Check that water pressure in boiler is approximately 50 psi.

- Check that any sampling lines are correctly installed in flue.

2. Preparation for Running Gasifier

(a) General Check

- Carry out general inspection to ensure all main supply lines are complete and that no sampling or other instrumentation holes have been left open.

- Check condition of all drive belts.

- Check oil level in metering pumps, drain all air filters for moisture.

- Check all sampling pumps work.

- Check all analytical equipment and calibrate if necessary.

(b) Control Unit

- Set alarm switch to "all alarms will show" position.

- Set alarm bell to "mute".

- Set automatic shut down switch to "inoperative" position.

- Turn on main power supply.

- "Burner or Boiler water" and "regenerator low bed" warning lights should come on. The former alarm is caused by pilot flame being out.

- Turn on regenerator blower and check that it will shut down if "automatic shut down" is selected. Reset alarm panel. Replace switch to "shutdown inoperative".

- Check that original two alarm lights show.

- Check and fill manometers if required.

- Start bleeds on injectors using direct N_2 bypass to avoid starting Compton compressors.

- Set pilot burner air supply to 6 cfm, turn on propane at external manifold, at plug cock on the wall, at valve over pit. Start main burner blower to pressurise plenum. Turn on pilot operating switch. The pilot should light and lock on. The pilot should stay alight and the warning light on the control panel go out.

- Set up recorders to 20 psi air supply. Switch in charts to check movements and recording pens are working. Turn off charts until required. Switch is in back of panel - plug and socket on each recorder.

B. WARM UP

The gasifier is heated over a 3 day period with a rate of temperature rise of about 12°C per hour.

It is important that heat be passed through the regnerator to make heat up of the refractory as uniform as possible. To secure a gas flow through the regenerator, the regenerator outlet valve must be open wide and the main boiler outlet tube damper closed. Be sure there is a nitrogen bleed through all injectors and pressure taps before starting.

The heat up sequence is as follows:

1. Start second stage air blower and line up 200 CFM air flow to gasifier. Remote valves set as follows:-

1st Stage outlet	Closed
2nd Stage inlet	Open
Flue Gas Recycle	Closed

2. Turn on regenerator blower and set flow to about 3 CFM through regenerator.

3. Open air flow to fuel injector lines. Set to 5 CFM on each injector.

4. Start air flow to boiler main burners - Set to 900 CFM.

5. Turn on propane supply up to gasifier start up burner. Set ganged propane valve to "start" position.

6. Line up air flow to start up burner.

7. Turn on cooling water spray to gas burner casing.

8. Turn on electrical power to start up burner control and start burner.

9. Adjust propane air/fuel ratio to obtain about 3 times stoichiometric air to fuel.

10. Monitor temperatures and adjust main and regenerator dampers to obtain uniform temperature rise in gasifier and regenerator.

11. Adjust gas and air rate to burner to follow desired temperature schedule on TC 7 of the L & N Recorder (gas space below Lid).

12. Do not let air rate to distributor of gasifier fall below 100 CFM. When necessary, increase total air supply to gasifier. Remember that air to distributor is difference between total air and air to burner.

13. When gasifier temperature reaches 700°C begin kerosene firing. Do not start kero firing without technical man on unit with operator.

14. Kerosene should be cut in very slowly at first, one pump at a time, to avoid upset to temperatures. Once kero feed is started, gradually raise kero rate to follow temperature schedule to 850°C. Air rate to plenum may have to be increased to provide sufficient air for kero combustion. There should be at least 30 CFM air through the plenum for each gallon/hr of kerosene feed.

15. When gasifier temperature reaches 850°C begin reducing gas and air to burner while increasing kero and air to gasifier to maintain temperature. Gas rate should be reduced to about 100 CFM, burner air to about 60 CFM, and total air about 350 CFM.

16. Fill limestone hopper while heat up is in progress.

17. Start air flow at 5 CFM to limestone feed system.

18. Start limestone feed vibrator at low setting (about 1) and begin adding stone to unit. Do not start limestone addition without technical man present.

19. Observe temperatures and fluidisation as soon as limestone addition starts to be sure that stone is being heated and fluidised throughout. If necessary adjust air and firing rates to obtain uniform heat up. Avoid letting stone temperature fall below 800°C.

20. Gradually add limestone to reach specified bed level on manometer. Do not exceed specified vibrator setting while adding bed, or fill line will block.

21. Start Compton compressors, raise regenerator air flow to 15 CFM, and begin bed circulation between gasifier and regenerator as soon as bed depth reaches 10 in. w.g.

22. When bed level reaches 12 in. w.g. the start up gas burner can be turned off. Start a 4 CFM air flow through burner cooling inlet to keep burner head cool and block main air line to burner.

C. INITIATING GASIFICATION

When the gasifier is at operating temperature with a desired bed level established and circulating, gasification can be started. The procedure used is to switch from kerosene combustion to fuel oil combustion, increase oil rate to the stoichiometric ratio to eliminate oxygen from the system, and then increase oil rate to that required for gasification.

1. Starting point is with gasifier at about 850°C with at least 12 in. w.g. bed depth, and good solids circulation in both directions between gasifier and regenerator. Temperature is maintained by kerosene combustion in the bed with excess air.

2. Increase air flow to main burner to 1100 CFM total of which 150 CFM is passed through the premix nozzle.

3. Check that pilot flame is strong and stable. Do not proceed until it is.

4. Check cooling system to be sure it is operating satisfactorily.

5. Line up flue gas recycle and adjust total flow to gasifier plenum to 260 CFM; 30 CFM flue gas recycle and 230 CFM fresh air.

6. Stop oil pumps and switch pump suction lines to fuel oil supply from kerosene supply.

7. Set centre pump to deliver 78 lb/hr fuel oil (stoichiometric) and start pump.

8. Set pumps 1 and 3 to deliver 130 lb/hr fuel oil each, and when gasifier bed temperature reaches 900°C, switch on both pumps. Normally pumps 1 and 3 will be required about 15 seconds after starting pump 2 at stoichiometric oil rate.

9. Main boiler flame should ignite within about 15 seconds of starting pumps 1 and 3.

10. Raise pump 2 setting to deliver 130 lb/hr fuel oil.

11. Adjust flue gas recycle rate to control temperature at desired level.

12. Check that cooling tower fan starts when secondary water temperature reaches the set point.

13. Begin limestone feed addition.

14. Bring all conditions to running specifications.

D. SOLIDS HANDLING

1. Bed Replacement

It is necessary to add fresh bed to the gasifier to maintain efficiency of operation. If required, material may be removed from the regenerator through a valve which can be operated on an automatic cycling basis from a variable controller. The fresh bed material is supplied to the gasifier through a feed pipe which passes through the side wall and enters the bed about 61 cms (24 ins) above the distributor. The stone is fed from a storage hopper onto a vibrator and then dropped into the feed pipe to be transported in a dilute phase with air. The feed hopper and vibrator are contained within a pressure vessel which is purged with nitrogen and maintained just above gasifier pressure to prevent the back flow of gasifier products which could lay down tar in the feed system. The feed hopper is continuously monitored with a remote reading load cell so that the bed feed rate can be accurately controlled at all times. The hopper can be refilled from an upper lock hopper without stopping the feed to the gasifier thus maintaining constant conditions in the unit. The upper lock hopper is filled by dilute phase material transfer from a small storage hopper on the ground into which the bags of stone are emptied.

2. Bed Removal

Hot bed material may be removed from the regenerator by automatic operation of the plug valve on the regenerator drain, setting the valve opening time to achieve the desired material removal rate. Frequent short periods of drainage are preferable to long periods followed by long periods without material movement because blockages are more likely to occur in the drain pipe. Gasifier bed material can be removed by manual operation of either the upper or lower drain valves.

3. Draining Regenerator Cyclone

The regenerator cyclone fines could be drained externally in Runs 5 and 6. Provision was made in Run 7 to feed the fines back into the gasifier cyclone fines into the gasifier bed with manual external draining available if required. The frequency of draining the material is determined by the running conditions but in all circumstances the material is removed by a lock hopper system to prevent the release of SO_2.

Provision was made to include a pneumatic rapper on the regenerator cyclone which comprised of a small metal hammer which struck the cyclone opposite the gas entry port, at regular intervals to prevent the adherence of fine dust.

4. Gasifier Bed Sampling

Samples of bed material may be taken from the upper or lower drain points taking care to flush the line of residual material and purging with nitrogen before collecting the sample.

E. CARBON BURN-OUT PROCEDURE

The carbon burn out procedure is initiated when the pressure in the gasifier above the bed reaches approximately 60 cms w.g.

1. Preparations

(a) Preparations should be started as soon as there are signs of the pressure build up increasing its rate of change. Additional help should be called out if required.

(b) Connect recycle and nitrogen supply pipes to the lid and check that the valves are closed.

(c) Install gas sampling lines from the plenum and lid recycle pipework to O_2 and CO_2 meters.

(d) Shut off any air bleeds to the lid.

2. Bed Sulphation

(a) Shut off bed feed, nitrogen to hopper purge and shut hopper outlet valve.

(b) Reduce gasifier temperature to below 850°C by increasing flue gas recycle rate.

(c) Switch bed circulation to manual and increase the rate to lower the regenerator temperature below 1000°C if possible.

(d) Check that the automatic control on the cooler water is functioning and the set point is at or just above the boiler water return temperature.

(e) Disconnect sample lines to boiler SO_2 analysis and clip off so that other analysers cannot suck air.

(f) Shut outlet valve on boiler sampling cyclone to prevent backflow of air into the boiler gas sampling system.

(g) Disconnect both orifice plate gauges on the 1st stage blower inlet to prevent liquid being blown out of the instruments.

(h) Open the by-pass on the flue gas recycle orifice plate.

(j) Shut off air to the gasifier start up burner purge.

(k) Switch fuel injectors to nitrogen and shut off air into the bed feed.

(l) Shut off fuel to the gasifier by stopping the pump and closing the isolating valves on the feed lines.

(m) Shut off air to the main burner when the main flame has gone out by closing the valves - do not shut off the blower.

(n) Watch the pressurisation unit pressure and be prepared to maintain pressure above the low level alarm by nitrogen addition if required.

(p) Close the scrubber gas return valve to provide maximum recycle supply and open recycle valve fully securing plugs across the 1st stage air inlet orifice plate. Watch the gasifier distributor pressure drop to assess the flow rate through the bed and if possible maintain a higher pressure in the gasifier than the regenerator.

(q) Turn on the nitrogen supply to the lid and reduce water supply to the scrubber to give a pressure drop of approximately 20". The water may need a further reduction if there is an inadequate flue gas recycle supply.

(r) Control the bed temperature to 925 - 950°C by regulating the flow of air to the main burner which should be set initially to provide about 10% O_2 concentration in the gas stream.

(s) When sulphation is near completion the bed temperature will start to fall and the duct temperatures rise due to the breakthrough of oxygen. At this point shut down the gasifier blowers, close the gate valve on the plenum supply pipe, disconnect the plenum sampling line and reduce the regenerator air flow to 3 cfm.

(t) Shut off air to the main burner by closing all the valves.

(u) Shut off the nitrogen with the fines return system.

3. <u>Carbon Burn-out</u>

(a) Restart the gasifier blowers and open the valve in the lid recycle line, adjusting the flow to about 100 cfm.

(b) Adjust oxygen concentration to about 7% by bleeding in air using the main burner air supply system.

(c) Gradually increase the rate of recycle to 150 cfm but maintain duct temperatures below 1000°C by controlling the oxygen concentration to prevent damage to the unit.

(d) Burn out is complete when the duct temperatures start to fall, the operation may take some hours depending upon the amount of carbon to be removed. The gasifier bed temperature will gradually fall during this period and should not go below 600°C at the bed centre otherwise there will be difficulty in restarting with kerosene combustion. It is possible to give the bed a short temperature boost by fluidising and feeding kerosene but duct temperatures will rise sharply and great care must be taken.

(e) When burn-out is complete close the valve in the line to the lid and shut off the blowers.

(f) Restart on kerosene combustion with a fluidised gasifier bed.

(g) Prepare to inspect the boiler rear end and commence the check out for resumption of gasification.

(h) Just prior to gasification drain out the chunk traps in the cyclone drain pots and inspect the cyclones through the top access holes to ensure that the drain lines are not obstructed.

APPENDIX F

CAFB PILOT PLANT ALARM SYSTEMS

1. ALARM ACTIONS

The installation is protected by a number of safety detection systems and there are four main actions that may be triggered off by an alarm depending upon which alarm is energised.

(a) Alarm Action A

- Fire Valves Close on oil line flow and return at entry point into the building.
- Oil circulation pump stops.
- Gasifier control panel is alarmed - the consequences of this alarm may be controlled and are described below.
- Interior and exterior bells ring inside and outside laboratory.
- Red light comes up on auxiliary panel located on laboratory wall close to main door to 3A.

(b) Alarm Action B

- Gasifier control panel is alarmed - consequences of this alarm may be controlled and are described below.

(c) Alarm Action C

- Alarm light comes up on auxiliary panel.
- Interior and exterior bells ring inside and outside laboratory.

(d) Alarm Action D

- Alarm light only comes up on main control panel.

(e) Choice of main control panel action

There are various actions that may be preselected to follow an alarm signal to the main control panel.

- An internal bell can ring or may be made inoperative by a switch on the panel.
- Automatic shut down of all blowers, pumps, compressors except burner air supply which is manually controlled at all times. Alternatively this procedure can be made inoperative by a switch on the panel.
- All alarm sensors may show as lights on the panel when alarmed.
- The first alarm sensor only will show as a light and any other alarms which are energised as a result of the automatic shut down will not show.
- The low hopper alarm has been arranged to show a warning light only and cannot be arranged to sound a bell or cause shut down.

Generally it is proposed to run the unit with the alarm bell in circuit and automatic plant shut down selected - this is necessary because failure of the gas pilot flame could result in unburnt gas forming in the boiler and ultimately an explosion.

(f) Method of alarm display on main panel

The particular alarm might well be dependent upon the source of the alarm. Table F-1 lists the alarms and how they will be shown.

2. ALARM SYSTEMS

The installation is best considered as four main systems:

- The boiler and its cooling system
- The gasifier
- The experimental burner on the boiler
- General alarms

(a) <u>The boiler and cooling system</u>

The water in the boiler circuit is pressurised to 50 psi and pumped through a heat exchanger. The secondary side of the heat exchanger is pumped to an evaporative cooler located on the roof of the building. The following alarms are installed in this system.

(1) Failure of the cooler circulating pump will operate a differential pressure switch across the pump feed and delivery lines -

<u>Result - Action (A) above</u>

(2) High water temperature in the cooler water feed line - set to operate at 190°F.

<u>Result Action A above</u>

(3) High water temperature in the boiler - set to operate at 245°F.

<u>Result Action B above</u>

(4) Low pressure in the pressurisation unit - set to operate at 43 psi.

<u>Result - Action B above</u>

(b) <u>Gasifier</u>

The gasifier has many alarm circuits whose action has been described earlier. The following parameters are monitored:-

(1) High temperature in the gasifier bed - set to 950°C. Shown and alarmed from Guardian indicator on main panel.

(2) High temperature in regenerator bed - set to 1100°C. Shown and alarmed from Leeds and Northrup recorder.

(3) Gasifier distributor blocked - shown on switch and set to 15".

(4) Regenerator distributor blocked - shown on switch and set to 15".

(5) Regenerator bed low level - shown on switch and set to 10".

(6) Regenerator bed high level - shown on switch and set to 24".

(7) Down stream blockage indicated by pressure in gasifier gas space - shown on switch and set to 10".

All these alarms <u>Result - Action B</u>

(8) Hopper low level will cause a warning light only on panel.

<u>Result - Action D</u>

(c) <u>Experimental burner</u>

The experimental burner will cause an alarm signal if there is a failure of:-

- pilot flame
- main flame
- low gas pressure set to 4" w.g.
- low air plenum pressure - set to 2" w.g.
- low pilot air pressure - set to .3 psi

Any one of these alarms will shown as alarm J on main panel.

<u>Result - Action B</u>

(d) <u>General Alarms</u>

(1) <u>Fire Detector</u>

A fire detector is situated over the boiler and will alarm at 150°F.

<u>Result - Action A</u>

(2) <u>Sump Level</u>

In the event of the build up of liquid in the pit to about 1" deep over the floor of the pit an alarm will sound.

<u>Result - Action C</u>

3. RESTART PROCEDURE

In the event of an alarm showing it is important to determine the cause of the alarm and if it is safe to restart.

(a) Restart after Action A

(1) Gasifier

The gasifier restarting procedure is described in the operating section, Appendix E.

(2) Auxiliary Panel

- If the audible alarm has not been muted it may be done by pushing the mute button on the auxiliary panel.
- The panel may be reset as soon as the alarm signal has ceased and the red light will go out.
- The two fire valves must be reset by lifting the lever and releasing the rewind mechanism by raising the rubber covered lever on the side of the panel and tightly winding up the string in a clockwise direction until the levers are set up as high as possible. The rubber covered levers can now be switched down. The oil circulating pump is restarted by depressing the green button on the side of the starter located near the pump.

(b) Restart after Action B

Gasifier

The cause of the alarm must be established before restart according to the instructions listed in Appendix E.

(c) Restart after Action C

Assuming that the automatic shut down was operative for main flame failure, the plant should be put under static conditions by shutting off the plenum air valve and turning nitrogen to all air injection points using minimum flow to retain bed temperature. Check pilot flame and flame scanners for correct operation and when all is functioning restart as described in Appendix E.

(d) Action D and E

These alarms will not cause a shutdown and are intended to draw the attention of the controller to some change in unit conditions which could develop into a problem if ignored.

(e) Action G

This alarm will have caused a shut down and the unit should be put under static condition with nitrogen bleeds when appropriate. Investigation should be made into the cause of low pressure, i.e. pump failure, valve leaking, or boiler water circuit leak. When the cause has been established and the problem resolved the unit may be either sulphated and burnt out or restarted as described in Appendix E.

Table F-1

Summary of Pilot Plant Alarm System

	Source of Alarm	Auxiliary Panel	Main Control Panel
1.	Failure of water circulating pump.	Red Light	Red light titled "Auxiliary Panel"
2.	High water temperature on cooler feed line	Red Light and bells	Red light titled "Auxiliary Panel"
3.	High water temperature in the boiler	None	Red light titled "Boiler water or Burner"
4.	Low Pressure in pressurisation unit	None	Red light titled "Auxiliary Panel"
	n.b. Red light shown on pressurisation panel, and its own bell rings.		
5.	Gasifier high temperature	None	Gasifier high temperature warning.
6.	Regenerator high temperature	None	Regenerator high temperature warning.
7.	Gasifier distributor-low pressure	None	Gasifier distributor low pressure warning.
8.	Regenerator distributor-high pressure	None	Regenerator-distributor high pressure warning.
9.	Regenerator low bed level	None	Regenerator low bed level warning.
10.	Regenerator high bed level	None	Regenerator high bed level warning.
11.	Gasifier high pressure	None	Gasifier high pressure warning.

Table F-1 (Continued)

Source of Alarm	Auxiliary Panel	Main Control Panel
12. Experimental Burner Failure of:		
Pilot flame	None	Pilot flame failure
Main flame	"	Main flame failure
13. Fire Detector	Red light and bells	Red light titled Auxiliary panel.
14. Sump Level	Red light	Nothing.
15. Emergency Stop Buttons in building	Red light and bells	Red light titled Auxiliary panel.
16. Emergency Stop Button on panel	None	None

APPENDIX G

CAFB CYCLONE EXTERNAL DRAIN SYSTEM

CAFB Cyclone Fines Return System

A major modification to the CAFB pilot plant made between runs 4 and 5 was installation of a system for externally draining the main gasifier cyclone legs and returning the coarse fraction of recovered solids to the gasifier. The system is shown in figure G-1.

Beneath each cyclone drain line is a conical receiver which can be isolated from its cyclone by a butterfly valve A. A line leads from the bottom of each conical vessel to a common fines receiver mounted above the gasifier. A pulser controlling valve D injects bursts of N_2 into this transfer line at a "N_2 knife" location just beneath conical vessel. Another supply of nitrogen enters each conical vessel through valve C.

Each transfer line is isolated from the fines receiver by a ball valve E. The fines receiver drains into the side of an elutriator vessel fluidised by nitrogen and fitted with a slug breaker near its top. Gas from the top of the elutriator, together with gas from the top of the fines receiver goes to a filter vessel in which fines are retained. The coarse solids fraction from the bottom of the elutriator drops through a water cooled heat exchanger to a bottom outlet from which it falls into a pick-up line through which it is air injected back into the gasifier. A valve in the vertical line just above the air pickup is controlled by a differential pressure switch to prevent flow reversal from the gasifier toward the elutriator.

The operating sequence of valves A, C, D, and E is regulated from a control panel in the control room. There is a separate control panel and set of valves for each of the two gasifier cyclones.

The controllers operate pneumatically with fluidic elements performing all logic and time delay functions. The sequence of operations of the controller is as follows:

1. Valve A, is open, valves C, D and E are closed. The vessel is filling with solids from the cyclone. A small bleed of N_2 into the body of valve A purges the cyclone drain leg of gasifier product gas.

2. When a timer reaches its end point, a signal closes valve A. After a short delay to allow A to shut completely, valve C opens to admit N_2 to the vessel to aerate and pressurise the solids in it. When the pressure reached a suitable level, say 5 psi, valve D begins opening and closing in rapid sequence to inject pulses of N_2 into the transfer line. Simultaneously, valve E opens to deliver material to the fines receiver. Material forced from the vessel is broken into short slugs by the action of the N_2 pulses. These slugs travel with reduced pressure drop and with less gas consumption than would be required for conventional pneumatic transport.

When the vessel is empty, the pressure within falls rapidly because there is no longer a resistance imposed by the solids. The pressure sensor detects this fall and closes valves C, D and E. Valve A then opens and a new filling period begins. We expect that fill times will be of the order of 15 minutes and emptying times about 1 minute or less. An interlock between the two parallel systems prevents both operating at the same time.

Solids drop from the fines receiver into the elutriator vessel where they are contacted with nitrogen at about 3 ft/sec. Fine solids are carried overhead by this gas while coarser material falls into the downcomer which also is fluidised by N_2. This downcomer is water cooled to reduce solids temperature enough to prevent rapid reaction with air. Fluidising N_2 for the downcomer is injected about one foot above the bottom outlet. Nitrogen from this injector passes both upward to fluidise the downcomer and downward to seal against backflow of air. A restriction at the bottom of the downcomer limits the flow rate of solids to prevent choking the air pickup. If the height of the fluid bed in the downcomer falls too low to provide an adequate seal against air backflow, the pressure detector senses the reduction and closes the bottom valve. The valve opens again when the pressure difference is restored.

A detailed description of the operation of the fluidic controller is given after Fig. G-1.

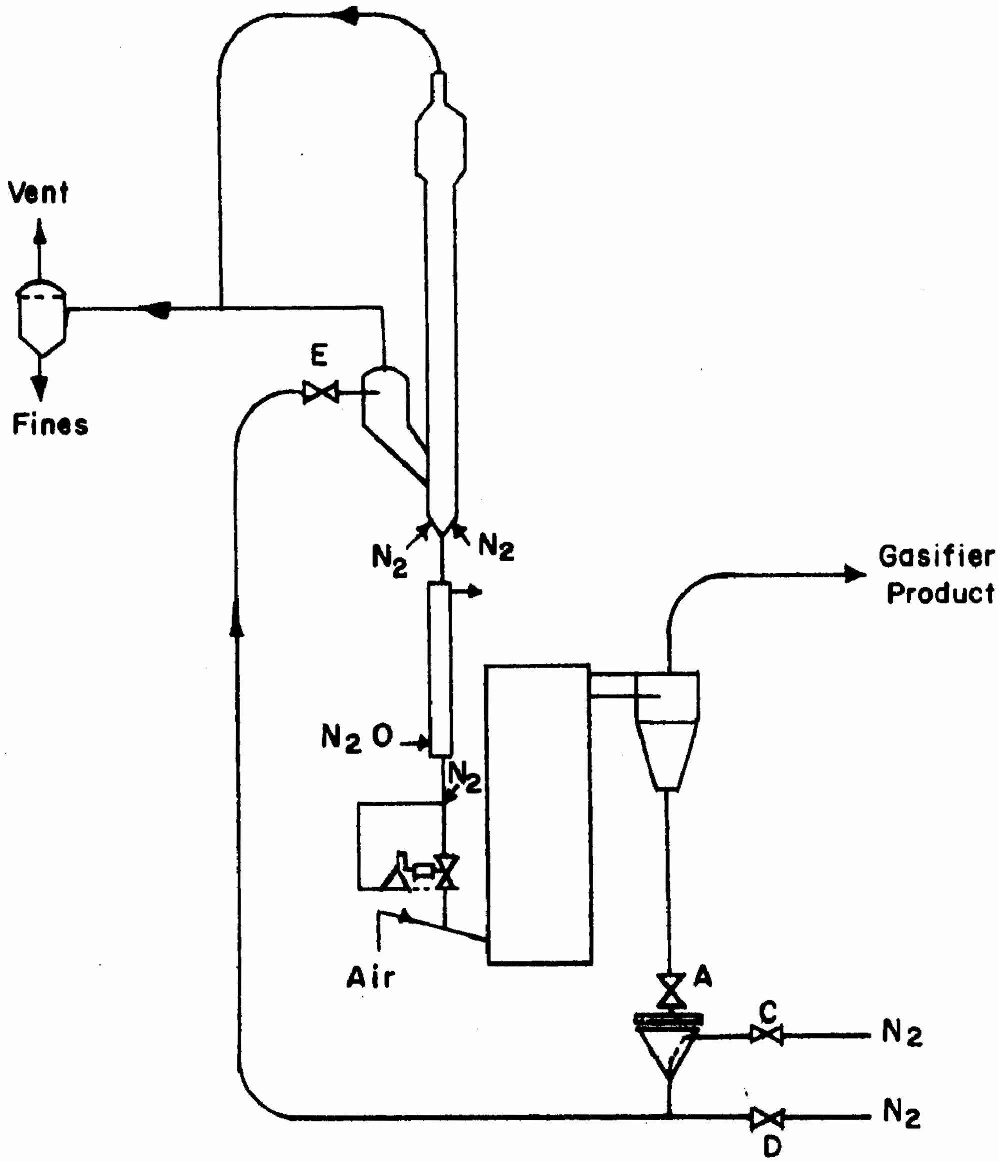


FIG. G-1 CYCLONE DRAIN SYSTEM

Description of Pneumatic Circuit

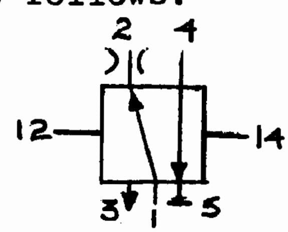
There are two parallel cyclone drain systems which empty to a common elutriator. Each drain and transfer system is controlled by its own pneumatic control panel. The two systems are interlocked to prevent simultaneous transfer by both. The controllers are based on "Fluid Log" pneumatic elements made by Lang Pneumatic Ltd., Telford, Shropshire. A diagram of the pneumatic circuit appears in Figure G-2 Table I compares the numbering system used in the circuit diagram with numbers on the physical panel.

Most of the logic elements in the circuit are 5 port valves with either pilot inlets at each end or a spring at one end and pilot pressure at the other.

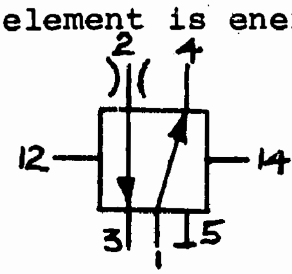
The symbols used in this memo, for these elements are as follows:

- If an element is energised left, it means that the left pressure is higher than right. In this case, the element lines are as follows:

  L Pressure Higher



- If an element is energised Right, the opposite applies:

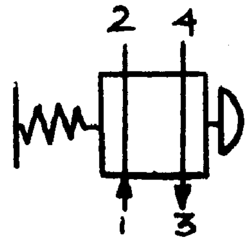


  R Pressure Higher

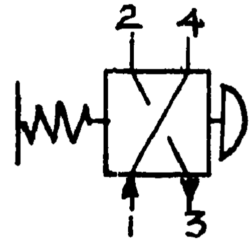
  The numbers refer to the manner in which the element parts are numbered.

The symbol used at point 2 on the above example is a restrictive orifice. Point 3 is open to atmosphere with no restriction. Point 5 is plugged.

Element No.3 in the diagram is a four port valve, air operated spring loaded. With no air signal, it takes the configuration;



When air energized, it reverses to:



(P > 5 psig)

Elements 6 and 9 are two port valves, spring loaded. They are normally closed, that is, no flow passes through them unless pressure is applied to their pilot sides.

The oscillator consists of four elements and a regulator which cause pulses of air to be applied to the pilot of element 9 when a signal is applied to the oscillator. Details of this circuit are not illustrated here.

The figure illustrates the configuration of the valves when operating air at 80 psig has been applied to all points marked "—•", but the switch is in the "off" position.

At this point there is no pressure on either pilot side of element 12. The position shown was maintained from the previous operation by a "detent" feature which prevents movement of the element when pressure is off. The element is normally in this position except for the brief period during which tank pressure is building before the outlet valve has opened.

The sequence of operations is listed in Table II and described briefly below.

Pressure is supplied to the switch from element 12 through element 7, but is blocked to the pilot of 10 and inlet valve 3 by the switch. Thus the inlet is closed while the switch is off.

When switch is turned on, pressure is applied to 3 to open the inlet valve and to reverse 10 to pressurised the R-1 delay accumulator and start R-2 accumulator exhausting through an orifice. The exhausting of R-2 supplies the main time delay for solids to drain from the cyclone to the transfer tank. This time is expected to be of the order of 10-15 minutes.

Because of the pressure applied through R-1 to the pilot of 2, 2 reverses. At this point there is still no pressure on either side of 12, but it remains held in position by a "detent" spring. The pilot of 4 is pressurised by the air pressure to valve A, and it reverses. Element 5 remains held by "detent" with no pressure on either side.

When the pressure in R-2 accumulator drops below that supplied by "Delay-2" regulator, element 1 reverses. This starts R-1 exhausting and pressurises the pilot of 14 which reverses. Pressure is removed from 3 pilot which reverses to close the inlet valve.

Pressure is released from 4 pilot which reverses in readiness to supply pressure to the left pilot of 5.

When R-1 pressure falls below that of Delay 1 regulator, 2 reverses. This delay was provided to give inlet valve about 15 seconds to close fully. Pressure now is supplied through 2 to the left pilot of 12 which reverses. Pressure from 12 then goes through 4 to reverse 5. Pressure from 12 also reverses 13 to put "x" pressure on the pilot of 7. The reversal of 5 applies pressure from 11 to the pilot of 6 which opens to admit nitrogen to the transfer tank aerator. Element 11 supplies this signal to operate its aerator so long as the second system is not operating the other tank aerator. If the other system is operating, the signal to open valve 6 is delayed until transfer from the other system is completed. Likewise, a signal from this system prevents the second from starting if the first is aerating. The purpose of this interlock is to prevent simultaneous high nitrogen demand from both transfer units.

While the transfer tank is building pressure, the outlet valve remains closed until tank pressure exceeds "x" pressure say 5 psig. When this happens, element 7 reverses to apply a signal to the oscillator, to element 8, and to the right side of 12. Element 7 removes the signal from right of 5 and from 10. The oscillator starts pulsing N_2 through the knife via 9. Element 8 reverses to open the outlet valve. Solids should now start transferring from the transfer tank to the elutriator. The element 12 reverses to change the signal to 13 which reverses to apply "w" pressure to the right pilot of 7.

When transfer of solids is completed, tank pressure falls quickly. When it drops below pressure "w", element 7 reverses to restore conditions to the starting point. Signal is removed from the oscillator, from 8, and from the right side of 12. The oscillator stops pulsing the knife through 9 and element 8 reverses to close the outlet valve. Pressure is applied through 7 to 5 which reverses to stop the N_2 flow to the aerator, and signal is applied to 10 and to 3 which reverses to open the inlet valve. Reversal of 10 starts the long time delay exhausting of R-2 and repressurises R-1 to reverse element 2. The transfer tank is now again receiving solids from the cyclone.

Table G-1

Composition of Element Numbering System

Number of Fig.1	Number on Panel	Catalogue No.
1	17	PLV 52/16F/B
2	18	PLV 52/16F/B
3	-(on front)	
4	9	PLV 52/15H/B
5	11	PLV 52/15D/B
6	-(in pit)	
7	13	PLV 52/16F/B
8	15	PLV 52/15H/B
9	-(in pit)	
10	16	PLV 52/15H/B
11	int	PLV 52/15H/B
12	7	PLV 52/16D/B
13	12	PLV 52/16F/B
14	behind panel	PLV 52/15H/B

Table G-2

Sequence of Controller Operations

	Supply On			Turn on Switch			R.2 < Delay 2 P 8 min			R-1 P < Delay 1 P 10 sec			Tank R > X			Powder Empties Tank P < W	Repeats Cycle
Switch	OFF	(1)	*	ON			ON			ON	.		ON			ON	
Valve 1	L			L	(1)	*	R	(4)	*	L			L			L	
2	R	(3)	*	L			L	(1)	*	R			R	(3)	*	L	
3	L	(2)	*	R	(3)	*	L			L			L	(2)	*	R	
4	R	(3)	*	L short delay	(4)	*	R - short delay			R			R	(3)	*	L short delay	
5	R			R			R	(3)	*	L			L	(2)	*	R	
6	R			R			R	(4)	*	L			L	(3)	*	R	
7	R			R			R			R	(1)	*	L	(1)	*	R	
8	R			R			R			R	(2)	*	L	(2)	*	R	
9	R			R			R			R	(2)	*	Pulsing	(2)	*	R	
10	L	(2)	*	R			R	(3)	*	L			L	(2)	*	R	
11	R			R			R			R			R			R	
12	R			R			R	(2)	*	L	(2)	*	R			R	
13	L			L			L	(3)	*	R	(3)	*	L			L	
14	L			L	(2)	*	R	(5)	*	L			L			L	
R-1 P O	80			80			Exhausting >			⟶ O			O			80	
R-2 P	80			Exhausting >			< Delay 2			⟶ 80			80			Exhausting > Delay 2 P	
W	1 PSI			1						1			1			1	
X	5 PSI			5						5			5			5	
Inlet	Closed		*	Open		*	Closed			Closed			Closed		*	Open	
Outlet	Closed			Closed			Closed			Closed		*	Open		*	Closed	
Aerator	Off			Off			Off		*	On			On		*	Off	
Knife	Off			Off			Off			Off		*	On		*	Off	
Tank P	O			O			O			Rising < X			Steady			O < W	

* indicates a change

(1) number in parentheses indicates order of event

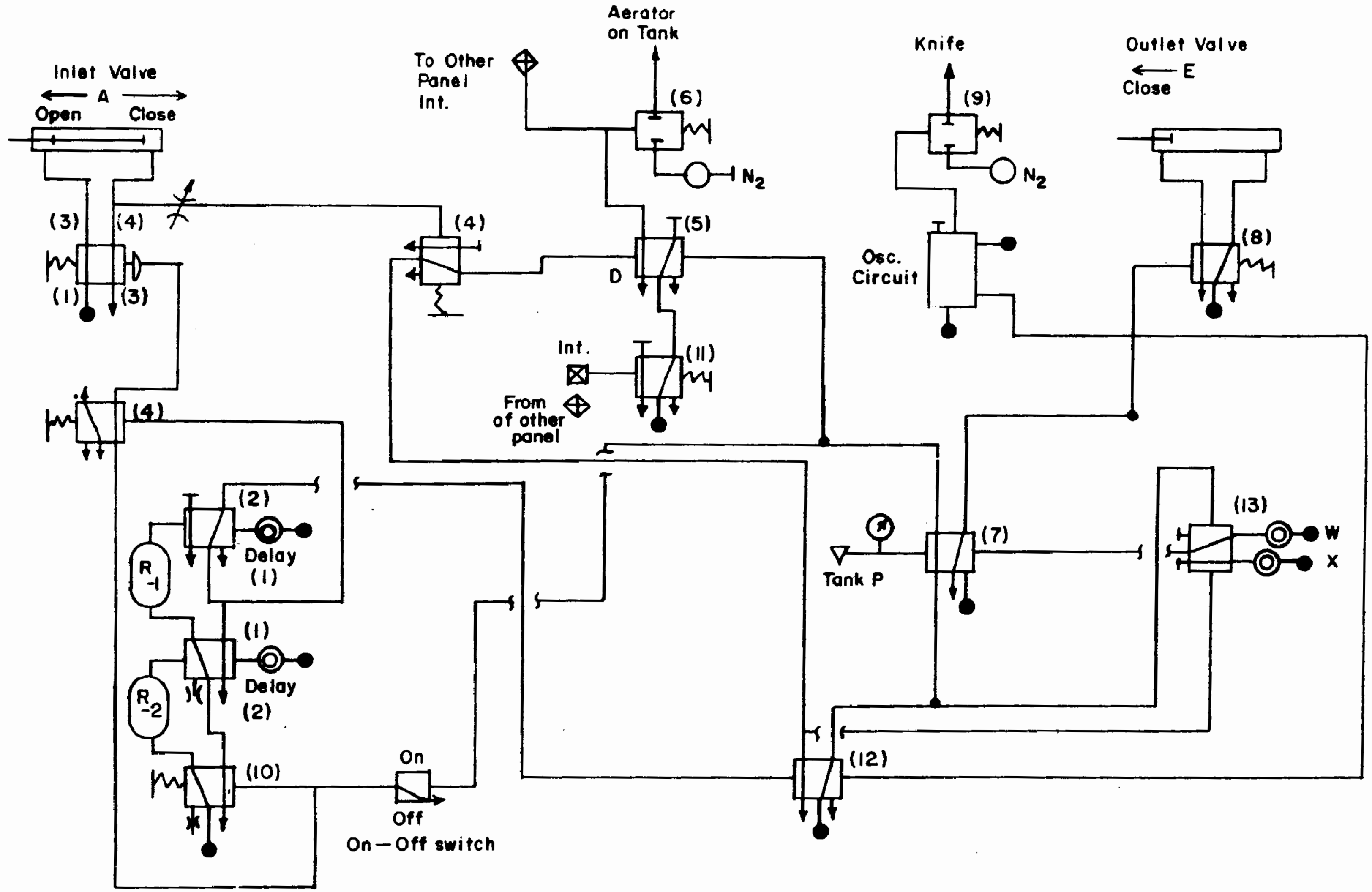


FIG. G-2 CIRCUIT OF CYCLONE FINES PUMP CONTROLLER

APPENDIX H

GAS ANALYSIS

BATCH UNITS

The gas analysis equipment used in the batch units is the same as used in Phase I, and is listed in Table 1.

Appendix H - Table 1

Batch Unit Gas Analysis Equipment

Analyser	Type	Manufacturer	Model	Response	Range
SO_2	Infra-red	Maihak	Unor 6	Continuous	0-1,000 ppm
SO_2	Infra-red	Maihak	Unor 6	"	0-20% by vol.
SO_2	Conducti-metric	Wostoff	-	"	0-1000 ppm
CO_2	Infra-red	Maihak	Unor 6	"	0-20% by vol.
CO	Infra-red	Maihak	Unor 6	"	0-20% by vol.
O_2	Para-magnetic	Servomex	OA 137	"	0-25% by vol.

During fully combusting conditions and also regeneration of sulphide, the gas from the reactor is pulled through a water condenser and filter and monitored directly for CO_2, CO, O_2, and SO_2. The appropriate information on combustion efficiency, sulphur removal efficiency or sulphur release is then deduced. During gasification, a portion of the product gas is burned in a sample flame located just above the reactor and the combustion products analysed for SO_2, O_2, CO and CO_2. The desulphurising efficiency of the gasifier is calculated from this analysis of the fully combusted gas. Because of the wide range of sulphur compounds present in the gasifier product itself, this is the only practical method for measuring gasifier sulphur removal efficiency. During this operation no in-line filter is used. Efficient cyclone operation and counterflow sampling are relied upon to separate particulates from the gas stream. This system has been shown to provide a representative sample of combusted gas and consequently, a reliable measure of SRE. It has been tested by passing sample gases containing the appropriate amounts of SO_2 through it and sulphur balances have been obtained.

CONTINUOUS PILOT PLANT

The gas analysers used during continuous pilot plant operation are the same as used in Phase I and are listed in Table II along with their applications.

Appendix H - Table II

CAFB Pilot Plant Gas Analysers

Gas Stream	Component	Analyser	Operating Principle	Range
Air-Flue Gas	O_2	Servomex OP 250	Paramagnetic	0-25% by vol
Mix to Gasifier Plenum	CO_2	Maihak Unor 6	Infra-Red	0-10% by vol
Boiler Flue Gas	O_2	Servomex OA 137	Paramagnetic	0-5% by vol
Sampled at fire tube outlet	CO_2	Maihak Unor 6	Infra-Red	0-20% by vol
	CO	Maihak Unor 6	Infra-Red	0-20% by vol
	SO_2	Maihak Unor 6	Infra-Red	0-1000 ppm
	SO_2	Wostoff	Electrical conductivity of H_2O_2 - SO_2 reactor products in solution	0-1000 ppm
Regenerator	O_2	Servomex OA 137	Paramagnetic	0-2.5% by vol
	CO_2	Maihak Unor 6	Infra-Red	0-10% by vol
	SO_2	Maihak Unor 6	Infra-Red	0-20% by vol
	CO_2) SO_2)	Pye Series 104	Gas Chromatography	0-100% by vol*

* Variable Range.

With the exception of the gas chromatograph, all the analysers give a continuous response.

Figure H-1 is a flow diagram for the pilot plant gas analysing equipment as used in Runs 4 and 5. The boiler gas sample was drawn through a hole in the refractory brickwork in the centre of the boiler rear door. This point was selected to provide a sample stream immediately after combustion in the main fire tube and before the gases passed through the water tubes where lime not trapped by the gasifier cyclones might be expected to deposit and might absorb some of the SO_2 in the gas stream. The regenerator gas sample was drawn from a point immediately after the regenerator cyclone where it was expected to be relatively free of solids.

During many of the test periods in Run 5, a poor sulphur balance was obtained. It was recognised that one possible reason for this could have been loss of SO_2 in the boiler sampling system. Lime was indeed observed to build up quite rapidly in the sampling line during Run 5 due both to poor gasifier cyclone efficiency and to the relatively high attrition rate of BCR 1691 limestone which was used for part of the run. To counteract this, therefore, the sampling system was modified before Run 6. The new arrangement incorporated a much larger gas stream through the sampling point in the boiler door which then passed through a cyclone before venting to atmosphere. The small sampling stream was taken from the cyclone outlet and passed through a filter, a water separator and then to the gas analysers as before. This design ensured that most of the dust was removed from the sample stream and did not obstruct the filter and possibly absorb SO_2.

The Run 6 gas analysis showed a generally higher level of SO_2 in the boiler than Run 5 and hence lower sulphur removal efficiency. The sulphur balance during Run 6 was closed within experimental error.

During Run 7, the boiler sampling system was investigated further. The results of the tests are recorded in the log of Run 7 in Appendix D. It has been concluded that a low sampling flow rate through the boiler brickwork in the presence of hot lime will not give accurate results because of absorption of SO_2 on the lime. A high volume sample stream achieves two improvements in that the residence time in the hot brickwork is considerably reduced and the increased velocity minimises the deposition of material in the duct which still needed regular cleaning during operation to eliminate errors due to reabsorption. Sulphur balance was also closed in Run 7.

It would appear, therefore, that the problems in accurately measuring SO_2 in the combusted gas from the continuous pilot plant which were not evident in operating the batch reactors were due primarily to the relatively low efficiency of the gasifier cyclones. The consequent higher solids loading in the boiler gas and the positioning and operation of the sampling point then combined to produce unreliable SO_2 measurements unless the sampling system was cleaned very frequently.

CAFB PILOT PLANT

FLOW DIAGRAM FOR GAS ANALYSING EQUIPMENT(RUNS 4 AND 5)

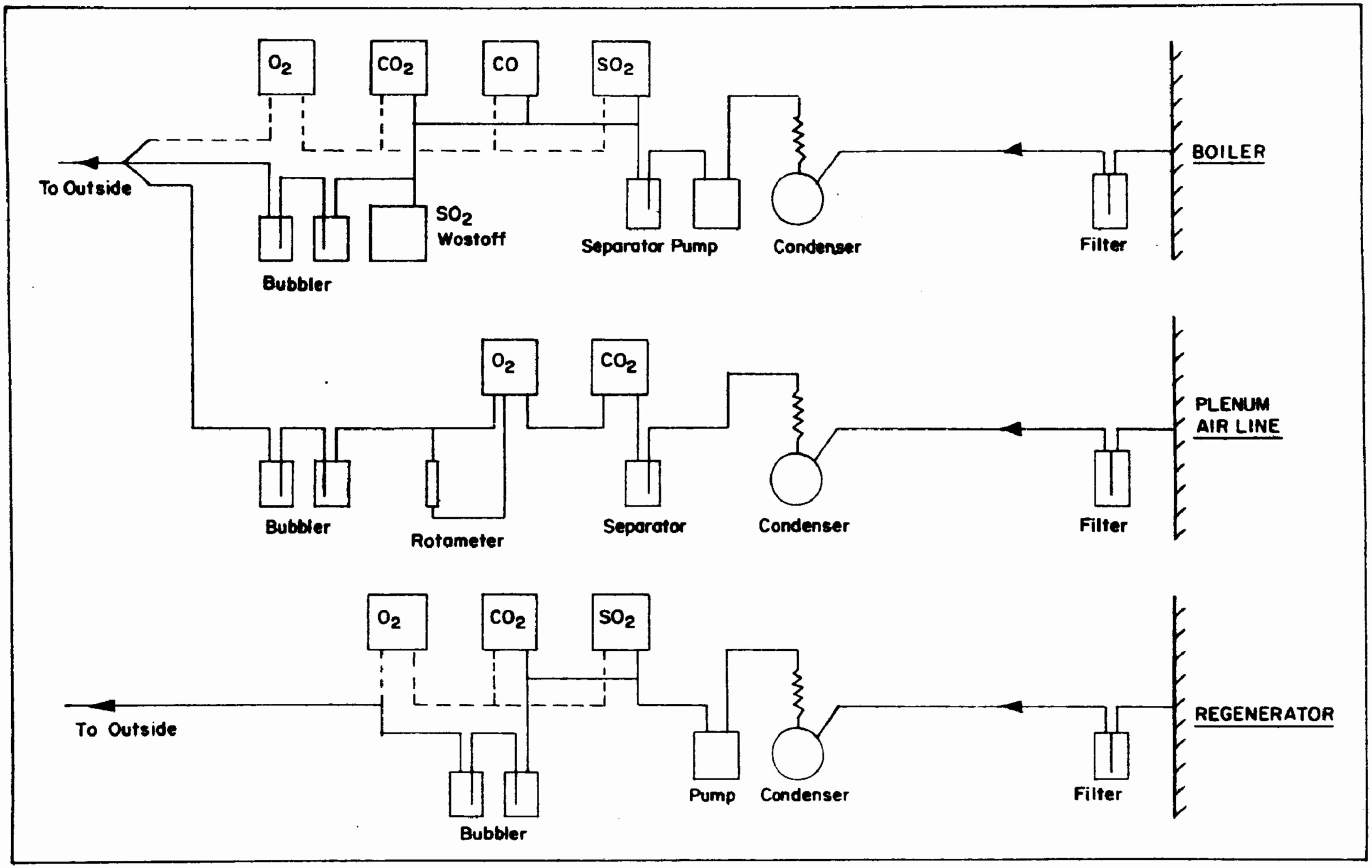


Figure H-1

APPENDIX I

GASIFIER HEAT BALANCE

A heat balance around the regenerator has been used to calculate the stone circulation rate. This balance and the consequent computer programme has been reported in the Final Report of Phase I of this work (Reference 1). This programme was used to calculate the stone circulation rate from data for Run 5, which values are reproduced in Appendix B of this report. An improved programme, which allows for the internal recycle of solids from the right hand cyclone, allows for solids drained from the regenerator and lost as fines, uses improved enthalpy data and an improved heat loss calculation, is in the process of being developed as a part of the mathematical model studies in Phase III.

The gasifier heat balance programme presented in the Interim Report of Phase II of this work is reproduced here together with outputs calculated from Run 5 data. This programme is being extensively modified and expanded as a part of the mathematical model studies of Phase III of this work, and will be converted to metric (S.I.) units.

The heat released in the gasifier under partial oxidation conditions depends on the air fuel ratio and on the relative extents of the various competing reactions within the bed.

As a first step in development of a mathematical model to predict gasifier operating conditions, a series of heat and material balance equations have been programmed to compute the heat release rate in the gasifier from measured flow rates and temperatures. A second set of equations is used to estimate the heat release rate from heats of reaction and from estimates of the relative amounts of hydrogen and carbon oxidised and the CO/CO_2 ratio produced.

List of Variable Names

Variable	Meaning	Units
F	Fuel Rate	lb/Hr
A	Air rate	SCFM
D	Gasifier bed depth	inches
S	Average bed particle size	microns
STNRT	Fresh stone feed rate	lb/Hr
TC	Gasifier bed temp.	Deg.C
TR	Regenerator bed temp.	Deg.C

Variable	Meaning	Units
V	Superficial gas velocity	ft/sec
CPCT	Carbon in fuel oil	wt.%
HPCT	Hydrogen in fuel oil	wt.%
SPCT	Sulphur in fuel oil	wt.%
CAOPCT	CaO in stone	wt.%
O2INFG	Oxygen in flue gas	Vol.%
STNCIR	Lime circulation rate	lb/Hr
G	Flue gas inlet rate	SCFM
CO2FG	Carbon dioxide in flue gas	Vol.%
TAIN	Temperature of inlet plenum	Deg.F
O2FED	Oxygen feed to gasifier	lb Mole/hr
RATIO	CO/CO_2 ratio produced in gasifier bed combustion	Dimensionless
FRCO2	Fraction of CO_2 in CO and CO_2 produced in gasifier bed combustion	Dimensionless
TG	Gasifier temperature	Deg.F
HCCO2	Heat of Combustion for C to CO_2	Btu/lb mole C
HCCO	Heat of combustion for C to CO	Btu/lb mole C
HH2H2O	Heat of combustion for H_2 to H_2O	Btu/lb mole H_2
STO2	Oxygen needed for stoichio-metric combustion of fuel oil	lb Mole O_2/100 lb oil
R	Oxygen fed as percent of stoichiometric	percent
TRF	Regenerator temperature	Deg.F
BTUCIR	Heat transferred from regenerator to gasifier by circulating lime	Btu/Hr
BTUOIL	Heat required to raise temperature of oil from inlet to gasifier temperature	Btu/Hr
BTUCRK	Heat required to crack oil to H_2 + C	Btu/Hr
BHA	Heat required to raise inlet air from plenum temperature to 1600 Deg. F.	Btu/Hr-CFM
BTUAIR	Heat required to raise inlet air from plenum temperature to gasifier temperature	Btu/Hr
C	Ratio of CaO to S fed	Dimensionless

Variable	Meaning	Units
BTUCAL	Heat required to calcine limestone	Btu/Hr
BTUCO2	Heat required to raise temperature of CO_2 from limestone to gasifier temperature	Btu/Hr
BTUSTN	Heat required to raise nonvolatile part of lime-stone to gasifier temp.	Btu/Hr
BTULOS	Heat lost from gasifier bed through walls	Btu/Hr
BTUFXD	Net heat input to gasifier except for reaction and flue gas contributions	Btu/Hr
AA	Constant in equation for percent of carbon oxidised	Dimensionless
COXDP	Portion of fuel carbon oxidised	Percent
COXDM	Amount of fuel carbon oxidised	lb Mole/100 lb fuel
BTUCCO2	Heat released by oxidising carbon to CO_2	Btu/100 lb fuel
BTUCCO	Heat released by oxidising carbon to CO	Btu/100 lb fuel
BTUH2O	Heat released by oxidising hydrogen to H_2O	Btu/100 lb fuel
BTUFG	Heat required to raise flue gas from plenum to gasifier temperature	Btu/Hr
BTUBRN	Heat released by combustion in gasifier	Btu/Hr
HTREL	BTUBRN x 10^{-6}	(Btu/Hr) x 10^{-6}
BTULB	Heat released per lb oil by combustion in gasifier	Btu/lb oil
HOXD	Heat released per mole oxygen by combustion in gasifier	Btu/lb mole
HOXDM	Amount of fuel hydrogen oxidised in gasifier	lb Mole/100 lb fuel
HOXDP	Portion of fuel carbon oxidised	Percent
	HOXDM and HOXDP are based on heat balance calculation	
CO2M	Amount of fuel carbon oxidised to CO_2	lb Mole/100 lb fuel

Variable	Meaning	Units
COM	Amount of fuel carbon oxidised to CO	lb Mole/100 lb fuel
O2H2O	Amount of oxygen converted to H_2O	lb Mole/100 lb fuel
HOXDM2	Amount of fuel hydrogen oxidised	lb Mole/100 lb fuel
HOXDP2	Portion of fuel hydrogen oxidised	Percent
	HOXDM2 and HOXDP2 are based on material balance calculation	
HDFL	Heat release per mole oxygen by combustion in gasifier, calculated by stoichiometry and reaction heats	Btu/lb mole

Thermal Equations

The thermal equations are linearised forms which adjust for departure from tabulated values at 1600 Deg.F.

```
HCCO2  =  169790 +  .434 * (TG-1600)
HCCO   =  48525  + .7375 * (TG-1600)
HH2H2O =  106941 + 1.327 * (TG-1600)
```

The air heat equations are based on published enthalpies at 100°F and 1600°F and on specific heats (C_p) at these levels.

$$BHA = \left\{\frac{60}{380}\,\frac{\text{min}}{\text{hr}}\,\frac{\text{lb Mole}}{\text{cu.ft.}}\right\} x\ (H_{1600} - (H_{100} + C_{p100}(TAIN-100)))$$

BHA = Enthalpy change to heat air from inlet temp. to 1600°F.
The term $\frac{60}{380}$ converts from CFM to lb Mole/Hr

BHA = .1579* (15057 -(3825 + 6.98*(TAIN-100.)))

$$BTUAIR = A *(BHA + \left\{\frac{60}{380} *C_{p1600}\right\} * (TG-1600))$$

BTUAIR = A *(BHA + .0439 *(TG-1600)

The enthalpy change to heat the flue gas is based on a flue gas of assumed composition:

$2.44\%\ O_2$; $75\%\ N_2$; $10.1\%\ H_2O$; $12.5\%\ CO_2$

Weighted average values of enthalpies and specific heats were combined to obtain the final equation.

$$BTUFG = G * \left\{\frac{60}{380}\right\} * (H_{1600} - H_{300} + C_{p1600} * (TG-1600) + C_{p300}\ (300-TAIN))$$

$$BTUFG = G * .1579 * (10719 + 8.902*TG - 14343 - 7.423*TAIN + 2227)$$

$$BTUFG = G * (1.4056 * TG - 1.172*TAIN - 204.79)$$

Heat of limestone calcination is taken at 1410 Btu/lb of CaO

$$BTUCAL = STNRT * CAOPCT * .01 \times 1410$$

$$BTUCAL = STNRT * CAOPCT * 14.10$$

Heat to raise temperature of CO_2 liberated in calcination.

$$BTUCO2 = STNRT * CAOPCT * \frac{.01}{56} * (H_{1600} + C_{p1600}\ (TG-1600))$$

$$BTUCO2 = STNRT * CAOPCT * \frac{.01}{56} * (17935 + 13.36\ (TG-1600)$$

$$BTUCO2 = STNRT * CAOPCT * (3.203 + .00239 * (TG-1600))$$

Heat to raise temperature of nonvolatile portion of stone.

$$BTUSTN = (STNRT) * (\text{Fraction Nonvolatile}) * (H_{1600} + C_{p1600} * (TG-1600))$$

$$BTUSTN = STNRT * (1 - \frac{44}{56}\ \frac{CAOPCT}{100} * (362.9 + .284 * (TG-1600))$$

$$BTUSTN = STNRT * (1 - .00786 * CAOPCT) * (362.9 + .284 * (TG-1600))$$

Heat supplied to oil.

Heat uptake by the oil is made up of sensible heat, latent heat, and heat of cracking.

Sensible heat and latent heat are included in a linear equation based on the oil enthalpy at 1200°F and the specific heat of the oil between 1200°F and gasifier temperature.

BTUOIL = F * (780. + 0.84 * (TG-1200))

The heat of cracking is assumed to be 600 Btu/lb

BTUCRK = F * 600

The heat loss from the gasifier bed is estimated from an analysis which indicated that the effective product of thermal conductivity and bed area is equal to 0.242 x D. The heat loss equation is therefore:

BTULOS = 0.242 * D * (TG-70)

Material Balance Equations

Oxygen fed.

$$O2FED = 0.21 * \frac{60}{380} * A + \frac{60}{380} * G * \frac{O2INFG}{100}$$

O2FED = .0332 * A + .00158 * G * O2INFG

CO/CO_2 Ratio

The ratio of CO to CO_2 produced by partial combustion of oil in the fluid bed is a complex function of themodynamics, reaction kinetics, and contacting in the bed.

Analytical results indicate that the values obtained are much lower than equilibrium for the temperatures encountered.

It is believed that considerable CO_2 forms by reaction of carbon on lime in the highly oxidising region near the air inlet nozzles and that insufficient contacting time is available for this CO_2 to reach equilibrium with CO and carbon in the upper portion of the bed.

The equation used here is an empiricial equation relating CO/CO_2 ratio to temperature, based on gas analysis data obtained in the batch CAFB reactors during the Phase I study and reported in Appendix D of reference (1). The equation is:

RATIO = 2.91 E-4 * EXP(9.76 E-3 * TC)

Fraction of Carbon Oxidised

In the current analysis it is assumed that components of the flue gas, recycled for temperature control, do not react in the gasifier. The validity of this assumption remains to be verified. With this assumption, fixing the CO/CO_2 ratio and the fraction of feed carbon oxidised also fixes the fraction of feed hydrogen oxidised since the oxygen fed must appear as CO, CO_2 or H_2O.

The fraction of carbon oxidised is based on an empirical equation first derived from batch unit studies but modified to improve match with the heat results of the continuous pilot plant runs. This equation is:

$$\text{COXDP} = \text{EXP}(-\text{AA}) * (\text{R} \uparrow .942) * (\text{TC} \uparrow 1.336)$$

A value of 8.4 for (AA) has given the best match between measured and calculated heat release rates.

The moles of carbon oxidised per 100 lb of fuel is the product of the fuel carbon content and fraction oxidised.

$$\text{COXDM} = \frac{1}{12} * \text{CPCT} * \frac{\text{COXDP}}{100}$$

$$\text{COXDM} = 8.333\ \text{E-4} * \text{CPCT} * \text{COXDP}$$

Calculated Heat Release of Fuel Oil

The calculated heat release per 100 pounds of fuel is given, for each element, by the moles of that element oxidised multiplied by the heat of combustion of that element.

$$\text{BTUCCO2} = \text{HCCO2} * \text{COXDM} * \text{FRCO2}$$

$$\text{BTUCCO} = \text{HCCO} * \text{COXDM} * (1 - \text{FRCO2})$$

$$\text{BTUH2O} = \text{HH2H2O} * \text{HOXDM}$$

The total heat release is the sum of that for the elements.

"Measured" Heat Release

The "measured" heat release is the difference between the sensible heat inputs to the gasifier and the heat required to raise the products to gasifier temperature.

BTUBRN = BTUFG + BTULOS + BTUCAL + BTUOIL + BTUCRK + BTUAIR + BTUSTN + BTUCO2 - BTUCIR

Hydrogen Oxidised

The amount of hydrogen oxidised can be estimated from the heat balance or from the oxygen balance.

Derivation of the heat balance equation for HOXDM is as follows:

$$\text{Heat release per mole } O_2 = \frac{\text{Heat of Combustion of elements}}{O_2 \text{ needed to burn elements}}$$

$$\text{HOXD} = \frac{\text{BTUCCO2 + BTUCCO + BTUH2O}}{(.5 * (\text{HOXDM + COXDM} * (1 + \text{FRCO2})))}$$

Also from the measured heat release in the gasifier,

HOXD = BTUBRN/O2FED

Substituting, BTUH2O = HOXDM * HH2H2O

gives,

$$\text{HOXD} = \frac{\text{BTUCCO2 + BTUCCO + HH2H2O} * \text{HOXDM}}{(.5 * (\text{HOXDM + COXDM} * (1+\text{FRCO2})))}$$

Rearranging and solving for HOXDM gives:

HOXDM = (BTUCCO2 + BTUCCO - (.5*COXDM * (1. + FRCO2))*HOXD)/(.5*HOXD-HH2H2O)

Material balance hydrogen oxidised

Since all oxygen fed to the gasifier is assumed to make CO, CO_2, or H_2O, fixing the amounts of CO and CO_2 by empirical expressions also fixes the quantity of H_2O.

```
CO2M   = COXDM * FRCO2
COM    = COXDM - CO2M
O2H2O  = O2FED * 100/F - CO2M - COM/2
HOXDM2 = O2H2O * 2
```

Calculated Heat Release per Mole O_2

A purely calculated heat release per mole oxygen can be computed for comparison with the "measured" value. For this purpose a value of BTUH2O is calculated using the material balance value of hydrogen oxidised.

```
BTUH2O = HOXDM2 * HH2H2O
```

then,

```
HDEL = .01 * F * (BTUCCO2 + BTUCCO + BTUH2O)/O2FED
```

These heat and material balance equations have been used in several computer programs to analyse experimental data and to predict new operating conditions. Table I-1 lists the Fortran statements of programme JHTHOC which calculates the gasifier heat release from experimental operating conditions and compares the value with the predicted value. Results of calculations on CAFB data appear in Table I-II.

APPENDIX I Table 1

FORTRAN LISTING OF CAFB HEAT BALANCE PROGRAMME

JHTHOC

```
100C CALCULATE HEAT OF COMBUSTION FROM CAFB CONDITIONS
110C  PROGRAM NAME JHTHOC
120  PRINT 600
130  DIMENSION P(25)
140 $FILE JHTFI1
150 50 CONTINUE
160 READ (1) (P(I), I=1,17)
170 IF (ENDFILE 1) 10, 800
180 10 CONTINUE
190C DEFINE VARIABLE VALUES
200  100 F = P(1)    'FUEL RATE, LB/HR
210  A = P(2)
220 D = P(3)
230  S = P(4)        'AVERAGE PARTICLE SIZE, MICRONS
240  STNRT  = P(5)
250  TC = P(6)       'GASIFIER TEMP DEG C
260  TR = P(7)       'REGENERATOR TEMP DEG C
270  V = P(8)        'GAS VELOCITY (INITIAL ASSUMPTION)
280  CPCT = P(9)     'CARBON IN FUEL, WT. PCT
290  HPCT = P(10)    'HYDROGEN IN FUEL, WT. PCT
300  SPCT = P(11)    'SULFUR IN FUEL, WT. PCT
310  CAOPCT = P(12) 'CAO IN STONE, WT. PCT
320  O2INFG = P(13) 'O2 IN FLUE GAS, VOL PCT
330  STNCIR = P(14) 'LIME CIRCULATION RATE, LB/HR
340 G = P(15)
350  CO2FG = P(16)
360  TAIN = P(17)*1.8+32.       'AIR INLET TEMP, DEG F.
370  TG  = 1.8*TC + 32.
380  TRF = 1.8*TR+32.
390  BTUCIR = STNCIR*(TRF-TG)*.284  'HEAT SUPPLIED BY CIRCULATING LIME
400  BTUOIL = F*(780. + .84*(TG-1200.))
410  BTUCRK = F*600.                         ' HEAT TO CRACK OIL
420  BHA = .1579*(15057.-(3825.+6.98*(TAIN-100.)))
430  BTUAIR = A*(BHA+.0439*(TG-1600.))
440 C = STNRT*CAOPCT/(1.75*F*SPCT)
450 BTUCAL = STNRT*CAOPCT*14.1
460 BTUCO2 = STNRT*CAOPCT*(3.203+.00239*(TG-1600.))
470  BTUSTN = STNRT*(1.-.00786*CAOPCT)*(362.9+.284*(TG-1600.))
480 BTULOS = .242*D*(TG-70.)
490  BTUFXD = BTUCIR-BTULOS-BTUCAL-BTUOIL-BTUCRK-BTUAIR-BTUSTN-BTUCO2
500  BTUFG = (1.4056*TG-1.172*TAIN-204.79)*G
510  BTUBRN = BTUFG-BTUFXD
520 HTREL = BTUBRN/1.E+06
530  BTULB = BTUBRN/F
540  O2FED = .0332*A+.00158*G*O2INFG   'MOLES O2 TO GASIFIER/HR
550  HOXD = BTUBRN/O2FED
560  150 RATIO = 2.91E-4*EXP(9.76E-3*TC)
570  FRCO2 = 1./(RATIO + 1.)            'FRACTION CO2 IN PRODUCT GAS
580 STO2 = CPCT/12. +HPCT/4. + SPCT/32.   'STOIC O2, MOL/100 LB
590 R = 1.E4*O2FED/(F*STO2)
```

APPENDIX I Table I

FORTRAN LISTING OF CAFB HEAT BALANCE PROGRAMME

JHTHOC CONTINUED

```
600 AA = 8.4
610 200 COXDP = EXP(-AA)*(R↑.942)*(TC↑1.336)
620  COXDM = 8.333E-4*CPCT*COXDP  'MOLE C OXIDIZED/100 LB FUEL
630  HCCO2 = 169790. + .434*(TG - 1600.)
640  HCCO = 48525 + .7375*(TG-1600.)  ' REACT HEAT, C TO CO
650  HH2H20 = 106941. + 1.327*(TG-1600.)  '  REACT HEAT,  H2 TO H20
660  BTUCCO2 = HCCO2*COXDM*FRCO2    'BTU/100 LB FUEL FOR C TO CO2
670 BTUCCO = HCCO*COXDM*(1.-FRCO2)
680 HOXDM=(BTUCCO2+BTUCCO-(.5*COXDM*(1.+FRCO2))*HOXD)/(.5*HOXD-HH2H20)
690 HOXDP = 2.E2*HOXDM/HPCT
700 CO2M = COXDM*FRCO2
710 COM = COXDM-CO2M
720 O2H20 = O2FED*100./F -CO2M-COM/2.
730 HOXDM2 = O2H20*2.
740 HOXDP2 = HOXDM2/(HPCT*5.E-3)
750 BTUH20 = HOXDM2*HH2H20
760 HDEL = .01*F*(BTUCCO2+BTUCCO+BTUH20)/O2FED
765 PRINT ,HOXDP,HOXDP2
780 610 FORMAT (4X,F5.1,4X,F6.2,4X,F7.0,4X,F8.0,4X,F7.0,4X,F5.1)
790 GO TO 50
800 600 FORMAT (/,6X,1HR,7X,5HHTREL,4X,6HBTU/LB,6X,6HBTU/02)
810 800 STOP
820  END
```

APPENDIX I TABLE II

Gasifier Thermal Performance

Computed by Programme JHTHOC

Run 3

Run Time Day-Hour	Percent Stoic. Air	Heat Release in Gasifier: Total Btu x 10^6	Per lb Oil Btu	Btu/lb Mole O_2 Measured	Btu/lb Mole O_2 Model	Percent H_2 Oxidised: Mass Balance	Percent H_2 Oxidised: Heat Balance
1,2130	20.4	1.16	3071.	149479.	153831.	14.5	9.5
10,0030	20.5	1.14	3241.	157451.	154603.	14.4	18.0
3,0030	19.0	1.12	2974.	155696.	154216.	13.2	14.8
3,1330	18.3	1.11	2925.	158537.	152165.	12.7	20.1
4,0430	21.9	1.16	3080.	139622.	153441.	15.8	1.4
4,0530	22.0	1.16	3074.	138590.	149615.	16.2	4.8
7,1130	19.4	1.20	3134.	160568.	153010.	13.6	23.3
8,2300	20.8	1.17	3314.	158708.	153759.	14.8	21.3
8,1130	19.7	1.15	3102.	156458.	153065.	13.9	17.9
10,1330	21.5	1.04	3258.	150388.	154904.	15.4	9.9
2,1230	20.8	1.19	3136.	150195.	154269.	14.7	10.0
5,0630	20.6	1.15	3170.	153128.	152576.	14.7	15.3
6,0430	17.7	1.02	2845.	160022.	153325.	12.1	19.8
7,0330	20.3	1.22	3180.	155426.	152149.	14.5	18.5
1,0930	19.9	1.14	3001.	150019.	154376.	14.0	9.1
1,1030	19.8	1.16	3046.	152706.	154112.	13.9	12.3
2,0030	20.5	1.19	3135.	152254.	154476.	14.5	11.8
2,0830	20.6	1.20	3148.	151735.	154502.	14.6	11.3
2,1130	20.7	1.18	3097.	148491.	154774.	14.7	7.6
3,0230	20.3	1.15	3014.	147617.	154447.	14.3	6.9
3,0630	19.5	1.14	3009.	153449.	154308.	13.6	12.6
3,0730	19.5	1.15	3024.	153891.	154315.	13.7	13.1
3,0830	18.6	1.12	2941.	157529.	154138.	12.8	16.7
3,1130	19.1	1.14	2993.	155765.	154238.	13.3	15.0
7,0830	20.6	1.24	3204.	154295.	154505.	14.6	14.3
7,0930	19.3	1.19	3060.	158475.	154277.	13.5	18.6
7,2130	17.6	1.18	3044.	172273.	153689.	12.0	39.6
7,2230	17.6	1.17	3026.	171259.	154332.	11.9	36.5
7,2330	17.7	1.19	3088.	173057.	154237.	12.1	40.9
8,1230	19.7	1.15	3100.	156746.	154082.	13.8	17.0
8,2330	19.5	1.15	3217.	163932.	154310.	13.6	26.8
9,0730	19.9	1.16	3251.	162581.	154630.	13.9	24.8
9,0830	20.7	1.17	3299.	158678.	154764.	14.6	19.7
9,1130	20.1	1.14	3204.	158843.	154660.	14.1	19.4
9,1830	20.3	1.12	3171.	155178.	154195.	14.3	15.5
9,2330	19.3	1.12	3167.	162757.	154408.	13.5	24.6
10,0230	20.3	1.15	3241.	158765.	154319.	14.3	20.0
10,1430	21.4	1.04	3261.	151476.	154883.	15.3	11.1
10,1530	21.4	1.05	3276.	152170.	154375.	15.3	12.5
4,1330	20.4	1.18	3134.	152815.	149320.	14.7	18.8
4,1530	20.0	1.18	3117.	155241.	148983.	14.4	21.8
8,0630	18.5	1.22	3161.	169463.	155029.	12.7	34.0
7,1530	19.2	1.21	3134.	162067.	152075.	13.5	26.5
2,0030 ♦	23.3	1.19	3232.	137605.	154295.	17.0	-.9
2,0430 ♦	23.4	1.23	3339.	141778.	154685.	17.1	2.2

♦ Run 4 data

APPENDIX I TABLE II

Gasifier Thermal Performance

Computed by Programme JHTHOC

Run 5

Run Time Day-Hour	Percent Stoic. Air	Heat Release in Gasifier: Total Btu x 10^6	Heat Release in Gasifier: Per lb Oil Btu	Heat Release in Gasifier: Btu/lb Mole O_2 Measured	Heat Release in Gasifier: Btu/lb Mole O_2 Model	Percent H_2 Oxidised: Mass Balance	Percent H_2 Oxidised: Heat Balance
2,2130	19.8	1.26	3173.	159492.	152692.	14.0	22.6
3,0530	20.4	1.23	3102.	151492.	152409.	14.5	13.4
3,1530	20.6	1.26	3198.	154496.	152704.	14.7	16.8
3,2130	20.3	1.29	3195.	156095.	151379.	14.6	20.4
5,1030	21.3	1.31	3277.	152803.	156889.	15.1	10.0
6,0730	20.4	1.24	3237.	157959.	156234.	14.3	16.4
12,1530	20.6	1.23	3105.	149565.	154505.	14.6	9.0
12,1930	20.6	1.20	3026.	145809.	154375.	14.6	5.4
13,0330	20.7	1.26	3216.	154398.	154770.	14.6	14.1
12,0830	20.6	1.26	3165.	152395.	153613.	14.7	13.2
17,1230	21.3	1.25	3129.	145675.	155509.	15.2	4.3
17,1730	21.9	1.28	3187.	144341.	155854.	15.7	2.8
21,0930	19.7	1.39	2968.	149845.	150096.	14.1	13.7
21,1830	19.0	1.40	2977.	155948.	151771.	13.3	18.1
22,0730	19.6	1.27	3066.	155877.	153294.	13.7	16.8
22,1830	19.8	1.26	3079.	154165.	153987.	13.9	14.1
24,0730	19.2	1.29	3146.	162969.	155273.	13.3	23.5
25,0530	19.2	1.27	3131.	162098.	153230.	13.4	25.0
25,1530	19.1	1.25	3085.	160530.	153597.	13.3	22.0
26,0530	18.9	1.25	3048.	160349.	152916.	13.2	22.4
26,1830	18.7	1.25	3046.	161566.	153916.	13.0	22.6
13,0630	20.8	1.26	3226.	153912.	153772.	14.8	15.0
26,1130	18.1	1.24	3024.	166220.	154561.	12.4	28.0

APPENDIX J

ANALYSIS AND ESTIMATION OF ERRORS

A thorough analysis of the probable errors associated with data used in the calculation of sulphur removal efficiency, overall sulphur balance and regeneration selectivity as well as errors associated with a carbon, hydrogen and oxygen balance around the gasifier and regenerator unit has been carried out as a preparatory step to using our data for the development of a mathematical model in Phase III.

An illustration of the approach used in the error analysis is given here, based on data from the early part of Run 6 (at day time 6.1900) at which time two simultaneous gasifier product gas samples were taken. Any equations used in data conversion are in the listing of programme ZKDAT, in Appendix K.

1. Calculation of Regenerator Errors

The streams entering and leaving the regenerator, and subject to error are:

Input
Regenerator air (total flow meter and stop watch)
Nitrogen to solids transfer (total flow meter and stop watch, some slip).

Output
Analysis of regenerator product gas for SO_2, CO_2 and O_2 (analysers).
Solids drained from the regenerator (weighed).

Interpolated values
Solids drained from regenerator cyclone (weighed).
Analysis of solids drained from regenerator cyclone, regenerator and gasifier (chemical analysis of samples for S, SO_4 sulphur and C).

Unmeasured values
Regenerator product gas flow rate.
Stone transfer from gasifier to the regenerator.
Stone transfer from regenerator to gasifier.
Internal stone recycle from right hand cyclone.

Errors

- Regenerator air $\pm$ 1% (estimated)
- Nitrogen to solids: although most of the gas flows in the direction of the solids, there is some flow the other way introducing an error (estimated from likely maximum and minimum flow resistance) of $\pm$ 20%.

- Regenerator product gas

SO_2: scale reads 10 to 100, corresponding to 0-20% SO_2. The scale is read to the nearest 1 (mean error 0.5) division, calibration contributes another 0.5 of a division, swings of the needle (based on 12 successive readings at 15 sec intervals) gave a standard deviation of 0.5 i.e. a 95% confidence limit of 1 division.

$$\text{Total reading error} = (0.5^2 + 0.5^2 + 1^2)^{\frac{1}{2}} = 1.2 \text{ divisions.}$$

Manufacturer's tolerance on calibration gas $\pm$ 5% of the SO_2 concentration, supplied in 5% and 10% concentrations.

At 5% the reading is 33, reading error =

$$\frac{(1.2\ \ )}{(33-10)} \times 100 = 5.2\%$$

$$\text{Total \% error} = (5.2^2 + 5^2)^{\frac{1}{2}} = 7.2\%$$

7.2% of 5% concentration = 0.36%

Therefore at a meter reading of 33 the SO_2 = 5 $\pm$ 0.36%. With 10% calibration gas the meter reading is 56.

$$\text{Reading error} = \frac{(1.2\ \ )}{(56-10)} \times 100 = 2.6\%$$

$$\text{Total \% error} = (2.6^2 + 5^2)^{\frac{1}{2}} = 5.6\%$$

5.6% of 10% concentration = 0.56%

Therefore at a meter reading of 56 the SO_2 = 10 $\pm$ 0.56%

Assuming linear variation of error with SO_2 concentration, we have

<u>Absolute error of regenerator SO_2 concentration $= 0.04 \times \%SO_2 + 0.16$</u>

<u>CO_2</u>: Similar analysis to above yielded scale readings $\pm$ 0.5, oscillations $\pm$ 2.

$$\text{Total reading error} = (0.5^2 + 0.5^2 + 2^2) = 2.1 \text{ divisions.}$$

5% CO_2 gave a reading of 66, reading error = 3.8%

$$\text{Total error} = (3.8^2 + 5^2)^{\frac{1}{2}} = 6.3\%$$

Absolute error at 5% CO_2 = 0.32%

10% CO_2 gave a reading of 96, reading error = 2.4%

$$\text{Total error} = 2.4^2 + 5^2)^{\frac{1}{2}} = 5.5\%$$

Absolute error = 0.55%

<u>Absolute error of regenerator CO_2 concentration $= 0.046 \times \%CO_2 + 0.09$</u>

<u>O_2</u>: needle can be read to $\pm$ 0.1, calibration gas errors are negligible (pure nitrogen and air and no oscillations observed therefore manufacturer's tolerance of $\pm$ 0.2% absolute error accepted.

<u>H_2O</u>: concentrations observed by gas chromatographic measurements corresponded to a carbon composition of approximately $CH_{0.5}$. H_2O was therefore calculated as 0.25 x measured CO_2 and the error of this estimate is a combination of the error associated with the CO_2 measurement combined with the error in the assumed composition for the reaction

$$CH_n + O_2 \rightarrow CO_2 + \frac{n}{2} H_2O$$

a tolerance of $\pm$ 0.2 on n = 0.5 corresponds to a percentage error of 40% which is unlikely to be optimistic and also completely swamps any contribution from the CO_2 measurement.
Therefore H_2O concentration = $0.25 \times CO_2 \times (1 \pm 0.4)$

$\underline{N_2}$: obtained by difference:

$$N_2 = 100 - O_2 - SO_2 - 1.25\ CO_2$$

Absolute error of N_2 = (sum of squared absolute errors of O_2, SO_2 and $H_2O)^{\frac{1}{2}}$.

Regenerator gas flow was calculated by nitrogen balance:

$$0.79 \times \text{Air} + \text{Reg.}N_2 = \text{Regen. gas flow} \times \frac{N_2 \text{ concentration}}{100}$$

The errors are combined in the usual way by summing sqaured absolute errors for added or substracted quantities and by summing squared percentage or fractional error for multiplication or division.

The above estimates are illustrated by the following example which illustrates the measurement errors and calculation of regeneration selectivity.

2. <u>Example of Regenerator Measurement Errors - Regenerator Selectivity</u>

Run 6, 6.19 Day time, averages of Day time 6.1830 and 6.1930 used.

Regenerator air = 33.75 m^3/h, 1% error = 0.34 m^3/h.
" " N_2 = 1.1 " , 20% error = 0.22 m^3/h.

Therefore: O_2 in = .21 x 33.75 = 7.088 m^3/h,
error = .21 x .34 = 0.071 m^3/h.

N_2 in = .79 x 33.75 + 1.1 = 27.76 m^3/h

$$\text{error} = ((.79 \times .34)^2 + .22^2)^{\frac{1}{2}} = 0.35\ m^3/h$$

Regenerator SO_2 = 7.4 ± (0.04 x 7.4 + .16) = 7.4 ± 0.46%

Regenerator CO_2 = 4.2 ± (0.046 x 4.2 + 0.09) = 4.2 ± 0.28%

Regenerator O_2 = 0.5 ± 0.2%

Regenerator H_2O = 0.25 x 4.2 (1 ± 0.4) = 1.05 ± 0.42%

Regenerator nitrogen = 100 - 7.4 - 4.2 - 0.5 - 1.05 = 86.85%

$$\text{error} = (0.46^2 + 0.28^2 + 0.2^2 + 0.42^2)^{\frac{1}{2}} = \pm 0.71\%$$

Therefore: N_2 = 86.85 ± 0.71%

Let regenerator product gas flow = Q m^3/h, then:

$$27.76 = Q\,\frac{86.85}{100}\ ,\ Q = 31.96$$

$$\text{fractional error} = \left\{\left(\frac{0.35}{27.76}\right)^2 + \left(\frac{0.71}{86.85}\right)^2\right\}^{\frac{1}{2}} = 0.015$$

absolute error = 0.015 x 31.96 = 0.48

therefore: $Q = 31.96 \pm 0.48\ m^3/h$

We can now calculate the selectivity of regeneration for:

$$CaS + 1.5\ O_2 \rightarrow CaO + SO_2$$

$$\text{mols CaS oxidised to } SO_2 = Q \times \frac{SO_2}{100} \times \frac{1}{22.711}$$

$$= \frac{31.96 \times 7.4}{2271.1} = 0.1041$$

$$\text{fractional error} = \left\{\left(\frac{.48}{31.96}\right)^2 + \left(\frac{.46}{7.4}\right)^2\right\}^{\frac{1}{2}} = 0.064$$

absolute error = 0.064 x 0.1041 = 0.0067

therefore: CaS oxidised to SO_2 = 0.1041 $\pm$ 0.0067 kmols/h

Oxygen not used for above and for oxidation of $CH_{0.5}$ is used to make sulphate:

$$CaS + 2\ O_2 \rightarrow CaSO_4$$

$$\text{mols } O_2 \text{ accounted for} = Q\left\{1.5 \times \frac{SO_2}{100} + \frac{O_2}{100} + \frac{CO_2}{100} + 0.5 \times \frac{H_2O}{100}\right\} \times \frac{1}{22.711}$$

$$= \frac{31.96}{2271.1}\,(1.5 \times 7.4 + 0.5 + 4.2 + 0.5 \times 1.05)$$

$$= \frac{31.96}{2271.1} \times 16.325 = 0.2397$$

$$\text{Error of concentration term} = \left((1.5 \times .46)^2 + .2^2 + .28^2 + .42^2\right)^{\frac{1}{2}} = 0.88$$

$$\text{fractional error} = \left\{\left(\frac{.48}{31.96}\right)^2 + \left(\frac{.88}{16.33}\right)^2\right\}^{\frac{1}{2}} = 0.0559$$

absolute error = 0.0559 x 0.2397 = 0.013

therefore: O_2 accounted for = 0.2397 $\pm$ 0.013 kmols/hr;

" : O_2 used to make sulphate = $\frac{7.088}{22.711}$ - 0.2397

$= 0.0724$ kmols/hr

$$\text{error} = \left\{\left(\frac{0.07}{22.711}\right)^2 + 0.013^2\right\} = 0.0134 \text{ kmols/hr}$$

therefore: Mols $CaSO_4$ made = 0.0362 $\pm$ 0.0067 kmols/hr (since 2 mols O_2 are required to make 1 mol of $CaSO_4$)

$$\text{Selectivity} = \frac{\text{mols CaS to CaO x 100}}{\text{mols CaS to CaO + mols CaS to } CaSO_4}$$

$$= \frac{0.1041 \times 100}{0.1041 + 0.0362} = 74.2$$

$$\text{fractional error} = \left\{\left(\frac{0.0067}{0.1041}\right)^2 + \frac{.0067^2 + .0067^2}{.1041 + .0362}\right\}^{\frac{1}{2}}$$

$$= 0.0691$$

absolute error = 0.0691 x 73.5 = 5.1

therefore: Selectivity = 74.2 $\pm$ 5.1%

Similar calculations for different conditions, which yielded selectivities in the range 50 - 100%, indicate that the error remains fairly constant at about 7% of the actual value. The estimation and associated errors of the stone transfer will be considered under balances, later on in this report.

3. Calculation of Gasifier Errors

Gasifier air: the main air flow is measured by a calibrated inlet orifice in cfm. The air added with stone feed is measured by a rotometer; the injector and burner purge air are recorded as a sum of 4 rotometer readings. All the rotometer readings are corrected for pressure and temperature and can be taken as correct to $\pm$ 7%.

The orifice pressure drop is read to the nearest 5 cfm with the following cumulative errors:

reading round-off = 1.25 cfm
fluctuations = 2 cfm
orifice tolerance $\pm$ 1% = 2.5 cfm (on a reading of 250)
calibration $\pm$ 1% = 2.5 cfm (on a reading of 250)

total = 4.25 cfm = 6.9 m^3/hr.

Air with stone feed, 7% on 8.6 m^3/hr = 0.6 m^3/hr
Injector and burner purge, 4 readings of about 6.8 m^3/hr
error = $(4 \times (.07 \times 6.8)^2)^{\frac{1}{2}}$ = 1.0 m^3/hr.
Total error on gasifier air = $(6.9^2 + 0.6^2 + 1.0)^{\frac{1}{2}}$
= $\pm$ 7.0 m^3/hr.

Recycle gas: This is measured by an orifice plate which is read to the nearest 5 cfm. The readings fluctuate mildly when higher than about 45 cfm, appreciably at 40 cfm and wildly at 25 cfm. In addition the recycle gas is approximately saturated with water vapour. Since the actual degree of saturation is not measured, the water vapour content is calculated by assuming 95% saturation at the exit temperature from the gas recycle venturi scrubber. The error of temperature measurement has been estimated at $\pm$ 2°C; assuming a temperature of 70°C and the water vapour equation:

H_2O = exp (0.4329 Tr + 0.3232) % by volume
which gives 31.19% v/v at 72°C and
26.23 at 68°C or an error of $\pm$ 2.48%
on a value of 28.6% v/v

Combining this value with $\pm$ 5% likely error in the assumption of 95% saturation we have:

$$\% \text{ error of } H_2O \text{ content} = 100 \left(\left(\frac{2.48}{28.6}\right)^2 + \left(\frac{5}{95}\right)^2 \right)^{\frac{1}{2}}$$

= 10.1%
say $\pm$ 10%.

Because of the wide variation in the fluctuation error, the cumulative reading errors are estimated at 3 levels:

All errors in cfm.

Level (cfm)	25	40	90
reading round-off	1.25	1.25	1.25
fluctuations	10	5	2
orifice tolerance (1%)	0.25	0.4	0.9
orifice calibration (1%)	0.25	0.4	0.9
Total	10.1	5.2	2.7

The cfm readings are corrected for pressure, temperature and for the effect of water vapour content on the specific gravity. The error of the pressure reading is $\pm$ 2 on a reading of 12 inches water gauge and that of temperature reading is $\pm$ 2°C, as mentioned above.
The calculations are (see ZKDAT in Appendix K):

$$Z = \text{cfm} \left\{ \frac{34.247 \times (12 + 407)}{(70 + 273)(30.4 - .124 \times \%H_2O)} \right\}^{\frac{1}{2}} \quad \ldots\ldots(1)$$

wet flue gas flow = 1.628825 (4.556 + .97121Z) ......(2)

where term in brackets represents the orifice correction and 1.628825 is the conversion factor from SCfm to m^3/hr and incorporates temperature correction from 60°F to 0°C and pressure corrections from 1 at to 1 bar.

viz: scfm at 60°F, 1 at to cfm at 0°C, 1 at,
conversion = $\frac{492}{520}$

cfm at 0°C, 1 at to m^3/h at 0°C, 1 at,
conversion = 1.699011

m^3/h at 0°C, 1 at to m^3/h at 0°C, 1 bar,
conversion = 1.013250

$$\text{therefore } \frac{492 \times 1.699011 \times 1.013250}{520}$$

$$= 1.628825$$

dry flue gas rate = wet flue gas rate (1 - .01 x H_2O)..(3)

Errors: Terms in brackets, eq. (1):

Top line (12 has error of $\pm$ 2) = 14349 $\pm$ 68

Bottom line, value = 9210.8

$$\text{Fractional error} = \left\{\left(\frac{2}{(70 + 273)}\right)^2 + \left(\frac{.124 \times .1 \times 28.6}{(30.4 - .124 \times 28.6)}\right)^2\right\}^{\frac{1}{2}}$$

$$= 0.0144$$

therefore: bottom line = 9210.8 $\pm$ 133

fractional error of the term multiplying cfm in eq.(1) is

$$\tfrac{1}{2}\left\{\left(\frac{68}{(14349)}\right)^2 + \left(\frac{133}{(9210.8)}\right)^2\right\}^{\frac{1}{2}} = 0.0076$$

on a value of 1.2481

$$\text{the fractional error on Z} = \left(.0076^2 + \left\{\frac{\text{error in cfm}}{\text{cfm}}\right)^2\right\}^{\frac{1}{2}}$$

Equation (2) does not introduce any error.

Equation (3) introduces an error through the water term, the fractional value of which at 70°C is:

$$\frac{0.01 \times 28.6 \times 0.1}{1 - 0.01 \times 28.6} = 0.04$$

We can now calculate the total errors associated with the recycle gas measurement and calculations at the three levels considered above:

	Recycle gas error calculation		
level (cfm)	25	40	90
total reading error (cfm)	10.9	2.5	2.5
multiplying term (T,P,H_2O correction)	←	1.2481	→
multiplying term fract. error	←	0.0076	→
Z (eq.(1))	31.20	49.93	112.33
error of Z	0.436	0.130	0.031
Wet gas flow m^3/h (eq.(2))	56.77	66.41	185.12
Water vapour content	←	28.6% v/v	→
Dry flue gas rate m^3/h (eq.(3))	40.5	61.7	132.2
Fractional error of dry flue gas	0.438	0.136	0.0506
Absolute error of dry flue gas, m^3/h	17.7	8.4	6.7

Plenum Gas: The mixed fluidising air and recycle gas are sampled from the plenum chamber and are analysed for oxygen and carbon dioxide.

A comparison between the analysed and calculated oxygen and carbon dioxide gives a check on our measurements, indeed, the analysis of the likely errors of these measurements led us initially to realise that the flue gas recycle is approximately saturated with water. The plenum gas is cooled and dried.

O_2: in the plenum is determined by a Servomex paramagnetic analyser which gives a very steady reading. Since the reading is roughly half-way between the calibration points the likely error is probably higher than 0.2% assumed for the lower readings (regenerator and flue gas).

Assumed error $\pm$ 0.3% (absolute).

CO_2: in the plenum is measured by an identical instrument to that used for the regenerator CO_2; the error can therefore be assumed the same, i.e.

$0.046 \times \% \ CO_2 + 0.09$ (absolute).

Flue Gas: it is sampled via a hot cyclone and filter and then a two-stage water condenser. It is analysed for:

O_2: by a Servomex, error $\pm$ 0.2% absolute

CO_2: scale reading 10 - 100, reading error 2.1 divisions (see regenerator SO_2 and CO_2).
Calibration gas tolerance $\pm$ 5%
5% CO_2 gave a reading of 42, reading error 6.6%.
total error $= (6.6^2 + 5^2)^{\frac{1}{2}} = 8.3\%$
absolute error at 5% = 0.41%
10% CO_2 gave a reading of 67, reading error = 3.7%.
total error $= (3.7^2 + 5^2)^{\frac{1}{2}} = 6.2\%$
absolute error at 10% = 0.62%

Absolute error of flue gas $CO_2 = 0.042 \times \% \ CO_2 + 0.20$

SO_2: measured by a Wostoff conductimetric analyser calibrated in ppm with a reading error of $\pm$ 5% (estimated) and calibration gas error of $\pm$ 5% giving a total error of $\pm$ 5.8%. The SO_2 content is corrected for solubility in condensed water.

4. Example of Gasifier Measurement Errors - Plenum Gas

Using data from Run 6, 6.1900: Fluidising air = 432.8 ± 6.9 m^3/h

O_2 = 90.9 ± 1.4 m^3/h
N_2 = 341.9 ± 5.5 m^3/h

Dry flue gas = 106.5 ± 7.3 m^3/h; water vapour = 42.7 ± 4.5 m^3/h.

Analysis: $O_2 = 1.95 \pm 0.2$
$CO_2 = 13.65 \pm 0.77$
$N_2 = 84.4 \pm 0.8$ (by difference)

Rates: $O_2 = 2.1 \pm 0.3$ m^3/h
$CO_2 = 14.5 \pm 1.3$ m^3/h
$N_2 = 89.9 \pm 6.2$ m^3/h

Dry plenum gas = 539.3 ± 10.0 m^3/h

Measured $O_2 = 18.1 \pm 0.3\%$, calculated $O_2 = 17.24 \pm 0.42\%$
Measured $CO_2 = 2.05 \pm 0.18\%$, calculated = $2.69 \pm 0.25\%$

The discrepancy between the calculated and measured concentrations is marginally outside the likely error range and it is probably caused by solution of CO_2 in the wash water. The discrepancy increases at lower recycle rates, at which the error of recycle gas measurement also increases. For these reasons, at plenum gas O_2 concentrations in excess of 17.5%, the dry flue gas recycle rate was calculated from the oxygen contents.

5. Balances

Apart from the streams already described, data on flue gas rate, oil analysis, solids analyses and product gas analyses were needed.

Repeat oil analyses are given in Appendix J Table I. Since the same oil supply was used in both Run 6 and 7 and a number of samples were taken during each run, sufficient data was obtained to calculate precision.

A number of repeat analyses were carried out on BCR 1359 limestone. These are shown in Appendix J Table II.

The gasifier product gas samples were taken at 6.1900 during Run 6. The analyses, correction for air contamination and errors (as estimated by our Analytical Dept.) are listed in Appendix J Table III. The nitrogen, as analysed by gas chromatography, contains nitrogen only; the corrected nitrogen also contains inerts.

The solids were analysed by wet chemistry with the precision of $\pm$ 0.15 on total sulphur and $\pm$ 0.1 on sulphate sulphur. The precision of carbon analysis has not been established and has been assumed, arbitrarily, to be $\pm$ 20% of the value for C content of less than 1% and $\pm$ 10% for C contents of 1% or more. The departures of the interpolated values from actual are, of course, unknown; the interpolated sulphur content have been assumed to have an error of 0.25%. The interpolated removed solids are assumed to be $\pm$ 20% correct for a spot reading; cumulative totals are, of course, much more exact.

Gasifier product gas flow was calculated from nitrogen balance:

Total Air = 468.5 $\pm$ 7.0 m^3/h, at 79%, N_2 = 370.1 $\pm$ 5.53 m^3/h

Dry flue gas = 106.5 $\pm$ 7.3 m^3/h, at 84.4 $\pm$ 0.8% N_2 = 89.9 $\pm$ 6.2 m^3/h

Total N_2 in = 460.0 $\pm$ 8.3 m^3/h.

Let dry product gas flow = Q m^3/h at 56.35 $\pm$ 1.0%

therefore: Q = 816 + 21 m^3/h = 35.93 + 0.92 kmols/h

6. Gasifier Carbon Balance (kmols/h) Run 6, 6.1900

Input:

oil 195.35 $\pm$ 0.39 kg/h at 85.3 $\pm$ .36% = 13.88 $\pm$ 0.06 kmols/h

Recycle CO_2 = 14.5 $\pm$ 1.3 m^3/h = 0.64 $\pm$ 0.06 kmols/h

Limestone = 19.75 $\pm$ 1 kg/h at 40.6 $\pm$ 1.3 wt% CO_2 = 0.66 $\pm$ 0.04 kmols/h

Total input = 15.18 + 0.09 kmols/h

Output:

Product gas = 35.93 $\pm$ 0.92 kmols/hr, composition as in Table III.

CO	4.78 ± 0.16
CO_2	2.86 ± 0.095
CH_4	2.31 ± 0.078
C_2H_6	0.43 ± 0.065
C_2H_4	3.02 ± 0.13
C_3H_6	0.07 ± 0.010
C_4's	0.20 ± 0.029
Total	13.67 ± 0.25

Regenerator product gas flow (see p. 526) = 31.96 ± 0.48 m^3/h

Regenerator CO_2 = 4.2 ± 0.28 %

therefore: carbon out of the regenerator = 0.06 ± 0.004 kmols/hr.

Carbon out in solids

	Interpolated solids rate, kg/hr.	Interpolated carbon wt%	Carbon out kmols/hr.
Regenerator cyclone	3.38 ± 0.68	0.53 ± 0.11	0.0015 ± 0.0004
Elutriator fines	0.38 ± 0.08	21.4 ± 2.1	0.0068 ± 0.0016
Boiler back	2.44 ± 0.49	0.34 ± 0.07	0.0007 ± .0002
Boiler flue	2.03 ± 0.41	5.88 ± 0.59	0.0099 ± 0.0022
		Total	0.019 ± 0.003

therefore: Total output = 13.75 + 0.25 kmol/h

Carbon recovery = 90.6 ± 1.7%

Since tar and possibly light hydrocarbons >C_4 may have been present in the gas but were not analysed, it is probable that they account for the missing 9.4 ± 1.7% carbon.

7. Gasifier Hydrogen Balance

Input:

oil 195.35 $\pm$ 0.39 kg/h at 11.28 $\pm$ 0.18
= 21.86 $\pm$ 0.35 katoms/h
steam 42.7 $\pm$ 45 m^3/h = 3.76 $\pm$ 0.40 katoms/h

Totat input = 25.62 $\pm$ 0.53 katoms/h

Output:

product gas 35.93 $\pm$ 0.92 kmols/h, composition as in Table III.

H_2	=	7.87 $\pm$ 0.27 katoms/h
CH_4	=	9.24 $\pm$ 0.31
C_2H_6	=	1.29 $\pm$ 0.20
C_2H_4	=	6.04 $\pm$ 0.27
C_3H_6	=	0.13 $\pm$ 0.02
C_4	=	0.40 $\pm$ 0.06
Output total	=	24.97 $\pm$ 0.53 katoms/hr

Hydrogen recovery = 97.5 $\pm$ 2.9%

Water vapour was not measured in the product gas and also the tar (possibly 1.43 $\pm$ 0.27 kmols/h of carbon and therefore 1.0 - 1.5 katoms/h of hydrogen) are not accounted for.

8. Gasifier oxygen Balance

Input:

Air O_2	468.5 $\pm$ 7.0 m^3/h	=	8.66 $\pm$ 0.13 katoms/h.
Recycle O_2	106.5 $\pm$ 7.3 m^3/h		
at 1.95 $\pm$ 0.2%		=	0.18 $\pm$ 0.02 katoms/h.
recycle CO_2	14.5 $\pm$ 1.3 m^3/h	=	1.28 $\pm$ 0.11 "
recycle H_2O	42.7 $\pm$ 4.5 m^3/h	=	1.88 $\pm$ 0.20 "
Transfered as SO_4^{-} from regenerator (see p. 527)		=	0.14 $\pm$ 0.03 "
	Total Input	=	12.14 $\pm$ 0.27 "

Output:
Product gas 35.93 ± 0.92 kmols/hr, composition as in Table III.

$$\begin{aligned} CO &= 4.78 \pm 0.16 \\ CO_2 &= 5.72 \pm 0.19 \end{aligned}$$

Total output = $\underline{10.50 \pm 0.25}$

Oxygen recovery = $\underline{86.5 \pm 2.8\%}$

This departure from balance has so far eluded us and it will be investigated during Phase III of this work.

9. Gasifier Stoichiometry

Carbon from oil = 13.88 ± 0.06 kmols/h requiring 13.88 ± 0.06 kmols O_2/h
Hydrogen from oil = 21.86 ± 0.35 katoms/h requiring 5.465 ± 0.088 kmols O_2/h
Sulphur from oil = 0.1517 ± 0.0009 katoms/h requiring 0.1517 ± 0.0009 kmols O_2 h

Total oxygen required = 19.50 ± 0.11 kmols O_2/h

Actual oxygen supplied as oxygen gas = 8.84 ± 0.13 katoms/h

therefore: Stoichiometry of gasification = 22.67 ± 0.36%

10. Overall Sulphur Balance

The dry flue gas rate for the calculation of sulphur balance was calculated from the nitrogen balance in the following way:

Overall oil reactions are

$$C + O_2 \rightarrow CO_2$$
$$H_2 + \tfrac{1}{2}O_2 \rightarrow H_2O$$

N_2 associated with each mole of O_2 = 3.762 mols.

195.35 kg oil contains 13.88 ± 0.06 kmol C which reacts with 13.88 ± 0.06 kmols of O_2. Associated with the O_2 are 52.22 ± 0.22 kmol N_2. Also, the oil contains 21.86 ± 0.35 katoms H_2 which react with 5.465 ± 0.088 kmol O_2 associated with which is 20.56 ± 0.33 kmol N_2.

$3.5 \pm 0.25\ m^3$ of pilot propane required 0.46 ± 0.03 mols of O_2 making $0.46 \pm .03$ kmols of CO_2 and 0.31 ± 0.05 kmols of H_2O. Associated with the O_2 required for combustion are 2.90 ± 0.21 kmols of N_2.

We can now sum the dry gases:

CO_2 from oil combustion	13.88 ± 0.06 kmol/h
N_2 from oil carbon combustion	52.22 ± 0.22 kmol/h
N_2 " " H_2 "	20.56 ± 0.33 kmol/h
CO_2 " propane "	0.46 ± 0.03 kmol/h
N_2 " " "	2.90 ± 0.21 kmol/h
CO_2 from limestone	0.66 ± 0.04 kmol/h
CO_2 from recycle gas	0.64 ± 0.06 kmol/h
Total	91.32 ± 0.46 kmol/h

Let X = mols excess air, then:

$$\%O_2 = \frac{21\ X}{91.32 \pm 0.46 + X}$$

but $\%O_2 = 1.95 \pm 0.2\%$ measured.

therefore $X = 9.35 \pm 0.96$ kmol/h

therefore total mols of Dry flue gas = 100.67 ± 1.06 kmols/hr.

Also required for the balance are the contributions from collected solids:

	Interpolated solids rate, kg/hr	Interpolated total S, wt%	Sout kmols/h
Regenerator cyclone	3.38 ± 0.68	4.49 ± 0.25	$0.0047 \pm .001$
Elutriator fines	0.38 ± 0.08	2.75 ± 0.25	$0.0003 \pm .00007$
Boiler back	2.44 ± 0.49	3.15 ± 0.25	$0.0024 \pm .00052$
Boiler flue	2.03 ± 0.41	3.99 ± 0.25	$0.0025 \pm .00053$
		Total	0.0099 ± 0.0013

Sulphur input:

Oil: 195.35 ± 0.39 kg/h at 2.49 ± 0.014 %S
= 0.1517 ± 0.0009 kmol/h

Limestone: 19.75 ± 1 kg/h at 0.25 ± 0.06 %S
= 0.0015 ± 0.0004 "

Recycle: 106.5 ± 7 m^3/h at 163 ppm S
= 0.0005 ± 0.0001 "

Total Input = 0.1537 ± 0.0010 "

Sulphur output:

In flue gas: 100.67 ± 1.06 kmols/h at 325 ± 16 vol.ppm
= 0.0327 ± 0.0017 kmol/h

In regenerator gas: 31.96 ± 0.48 m^3/h at 7.4 ± 0.46 wt.%
= 0.1041 ± 0.0067 kmol/h

In solids (see above) = 0.0099 ± 0.0013 kmol/h

Total output = 0.1467 ± 0.0070 kmol/h

Sulphur recovery = **95.4 ± 4.6%**

This calculation also gives the sulphur content of flue gas and total sulphur input. Therefore we can calculate the overall sulphur removal efficiency:

$$\text{S.R.E.} = \frac{\text{total sulphur in} - \text{total sulphur in flue gas}}{\text{total sulphur in}} \times 100$$

therefore: S.R.E. = **78.7 ± 1.4%**

11. Summary:

Approximate errors of the derived quantities as a percentage of the value:

Sulphur removal efficiency = ± 2%
Regenerator selectivity = ± 7%
Sulphur balance (overall) = ± 5%
Gasifier carbon balance = ± 2% (balance not closed)
Gasifier hydrogen balance = ± 3%
Gasifier oxygen balance = ± 3% (balance not closed)
Gasifier stoichiometry = ± 2%

APPENDIX J - Table I

Analysis of Fuel Oil Used in Runs 6 and 7

	Run 6			Run 7								
Sample No.	50494	50495	50591	51072	51073	50928	51025	51135	51136	Mean	standard error	95% confidence limits
C wt.%	86.23	85.95	84.93	85.19	85.30	85.12	85.25	84.88	84.97	85.31	.156	.36
H wt.%	10.92	10.95	11.45	11.21	11.33	11.22	11.64	11.42	11.34	11.28	.077	.18
S wt.%	2.50	2.45	2.49	2.50	2.5	2.50	2.50	2.50	2.47	2.490	.006	.014
N wt.%	.35	.37	.44	.35	.31	.38	.37	.30	.31	0.353	.015	.034
V ppm	310	300	310	300	305	320	310	305	300	306.7	2.2	5.1
Na ppm	39	38	59	33	46	23	40	32	37	38.6	3.3	7.7
Fe ppm	5	3	3	3	2	2	2	2	2	2.67	.33	.77
Ni ppm	54	52	38	37	37	44	35	36	35	40.9	2.5	5.7
Sp gravity	.9580	.9590	.9571	-	-	-	-	-	-	.9580	.00055	.0024
Conradson carbon	11.12	11.16	10.54	-	-	10.81	10.58	-	-	10.84	0.29	0.81
Asphaltenes	5.1	5.6	5.43	-	-	5.67	5.47	-	-	5.45	0.22	0.61
								Total		99.472		±0.40

APPENDIX J - Table II

Analyses of Limestone BCR 1359

Sample No.	50444	50428	50475	50416	50489	50515	50618	50661***	50669	Mean	standard error	95% confidence limits
CaO wt.%	56.7	57.1	55.6, 57.4	57.0	55.8	57.4	57.0	56.6	56.6	56.72	0.19	0.44
MgO wt.%	0.58	0.58	0.57, 0.55	0.58	0.54	0.55	0.55	0.55	0.6	0.565	0.0061	0.014
SiO_2 wt.%	0.8	0.8	0.8, 1.1	0.8	0.8	1.1	0.8	0.8	0.8	0.86	0.045	0.10
Al_2O_3 wt.%	0.2	0.2	0.2, 0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.0095*	0.02
CO_2 wt.%	-	-	38.6	-	-	41.0	38.4	38.6	38.6	40.6	0.48	1.3
Fe_2O_3 ppm	-	-	660	-	-	490	491	536	543	544	31	86
Na_2O ppm	-	-	655**	-	-	160	126	113	102	125	12.6	40
S ppm	-	-	-	-	-	0.24	-	0.28	0.24	0.25	0.013	0.06
V ppm	-	-	<50	-	-	79	<50	<50	<50	<50	-	-
Ni ppm	-	-	11	-	-	6	16	56	57	29	11	31

Notes:
- * estimated from other data
- ** outlier, left out of error calculations
- *** this sample was analysed for free CaO (as opposed to total Ca): found to be ∿ 1%.

APPENDIX J - Table III

Gasifier Product Gas analyses, Run 6, 6.1900

Sample No.	50505 (1)			50505 (2)				
	Determined %	Corrected %	Error	Determined %	Corrected %	Error	Mean	Error
O_2 + Ar	1.3	-	-	1.0	-	-	-	-
N_2	57.0	-	-	54.9	-	-	-	-
N_2 + inerts	-	56.87	1.42	-	55.84	1.40	56.35	1.0
CO	13.0	13.36	0.40	12.9	13.27	0.40	13.31	0.29
CO_2	7.8	8.01	0.24	7.7	7.92	0.24	7.96	0.17
H_2	10.4	10.68	0.32	10.9	11.21	0.34	10.95	0.24
CH_4	6.21	6.38	0.19	6.3	6.48	0.19	6.43	0.14
C_2H_6	0.45	0.46	0.09	0.71	0.73	0.15	0.60	0.09
C_2H_4	3.95	4.06	0.20	4.22	4.34	0.22	4.20	0.15
C_3H_8	-	-	-	-	-	-	-	-
C_3H_6	0.05	0.05	-	0.06	0.06	-	0.06	(0.01)
C_4	0.13	0.13	0.03	0.15	0.15	0.03	0.14	0.02

APPENDIX K

COMPUTER PROGRAMMES FOR ANALYSIS OF THE CONTINUOUS RUN DATA

GENERAL

Time sharing computing facilities, rented on the Honeywell Mk.III Foreground system, were used for data analysis. All the programmes were written in FORTRAN IV. Data was entered on paper tape into listable files (to allow easy correction). After checking, correcting and interpolating the data is written into a set of consistent binary files which are used as the data base for calculating the sulphur removal efficiency, balances, etc. The chart in Table K-1 indicates how the various programmes and files are used.

GENERATION OF DATA BASE

The hourly run data, consisting of selected temperatures, pressures, flows and continuous analyser readings are read into files RXDATA, where X = run number. The actual data points used are described in Table K-2.

Files RXDATA are checked manually and translated by programme ZKABCON into binary files BRXDATA. During the conversion any instrument correction factors, which are pertinent to run X, are applied so that files BRXDATA are consistent regardless of any run-to-run instrumentation changes.

The binary files BRXDATA are scanned by programme ZKSD which identifies intervals of more than one hour and records these as well as the day-time of start-up and final shut-down on a listable file SDX. This file is used to ensure day-time coincidence of interpolated data with that of hourly run data.

Weights of solids removed are entered into files STX in the following order:

A(1) = Day-time
A(2) = solids drained from the gasifier
A(3) = solids drained from the regenerator
A(4) = solids from the regenerator cyclone

A(5) = elutriator fines
A(6) = solids removed from the back of the boiler
A(7) = combined solids from the stack K.O. pot and cyclone
A(8) = elutriator coarse solids

Missing values at any one day time are filled in with -1 for easy identification. This data is processed by programme ZKSTINT in the following manner: weight units are converted to kilograms, A(2), A(3) and A(8) are used directly but the day-time is corrected for the delay between draining, weighing and recording; A(3) -A(7) are interpolated linearly to the equivalent hourly collection rates. Any solids removed during shut-down are ignored and any solids removed during missing data periods are allowed for (N.B. "missing data" periods are periods of normal gasification during which one or more essential readings were missed). File SDX is used to ensure compatibility of day-time with run data files and the converted data on solids removals is written on to a binary data file STBINX. A listing of programme ZKSTINT is provided in this Appendix.

Analytical data on the removed solids is written, in the same order as the solids removal, into files YX, where Y = analytical identification (Y=S for total sulphur, Y=O for sulphate sulphur, Y=C for total carbon, Y=V for vanadium, etc.) and X = run number. These files are interpolated linearly to equivalent hourly values using programme ZKANINT ensuring compatibility of day-time by reference to file SDX. The interpolated analyses are written into files YBX where Y and X have the same meaning as above. A listing of ZKANINT is provided in this Appendix.

Chemical analysis of the oil and limestones as well as the day-times at which limestones are changed are entered into files ZKPROPX and checked manually. Programme ZKANFIL combines the analytical data from ZKPROPX, SBX, OBX and CBX and the solids removed data STBINX into a single analytical data file BRXAN. The two sets of files: hourly run data BRXDATA and the interpolated hourly analytical and solids removal BRXAN as well as the derived data files JIMX (described in the next section) constitute our data base.

DATA ANALYSIS

The data base files are worked up by programme ZKDAT which converts all the gas flow meter readings into cubic meters per hour at 0°C, the oil and limestone readings into kilograms per hour and analyser readings into percentage concentrations. Flue gas recycle is corrected for 95% saturation with water vapour. Measured SO_2 concentration in the flue gas is corrected for solubility in water knocked out as the sample is cooled. Sulphur removal efficiency is calculated as:

$$\% \text{ Efficiency} = 100 \left(\frac{\text{total sulphur in} - \text{S in flue gas} \times \text{flue gas rate}}{\text{total sulphur in}} \right)$$

where: total sulphur includes the sulphur from oil and limestone and sulphur in the flue gas recycle at half the concentration of SO_2 in the flue gas. The figure of half the concentration is based on some rough measurements before and after the gas scrubber, measuring the SO_2 with a Drager tube. Flue gas rate was calculated from nitrogen balance as follows:

Atoms C from oil and propane -C out of regen -C in fines = C_o

Atoms H from oil and propane -H out of regen -H in fines = H_o

Mols oxygen to combust the above = $C_o + \frac{1}{4}H_o$

Mols nitrogen associated with oxygen = $\frac{79}{21}(C_o + \frac{1}{4} H_o)$

Note: the contribution from N_2 associated with O_2 for sulphur combustion is negligible.

CO_2 from flue gas recycle and limestone = CO_2

with X mols excess air, the total boiler dry mols are:

$$BDM = C_o + CO_2 + \frac{79}{21}(C_o + \frac{1}{4} H_o) + X$$

Measured oxygen in the flue gas = O_f

$$\text{therefore } O_f = \frac{0.21X}{BDM} \times 100 \%$$

elimination of X and rearrangement gives

$$BDM = \left(\frac{21 - O_f}{21}\right)\left(\frac{100}{21} C_o + CO_2 + \frac{79}{84} H_o\right) \text{ mols/hr}$$

It has been assumed that the carbon deposits have an average composition $CH_{0.5}$ and therefore 0.25 moles of H_2O are formed in the regenerator for every mole of CO_2 measured. Gas chromatographic measurement of water content in the regenerator off gas have confirmed the presence of 0.3 - 0.4% H_2O.

> NOTE: This value of BDM can be used to calculate the O_2 and N_2 required for combustion of sulphur and recalculated. However, such a correction was found to be negligible.

Selectivity of oxidation of CaS to SO_2 is calculated as follows:

Mols CaS to CaO = mols SO_2 out of the regenerator

This represents the reaction

$$CaS + 1\tfrac{1}{2}O_2 \rightarrow CaO + SO_2$$

The balance of oxygen is assumed to react to form $CaSO_4$ by the reaction

$$CaS + 2O_2 \rightarrow CaSO_4$$

Therefore:

Mols $CaSO_4$ = ½(mols O_2 in -1.5 x mols O_2 out of regenerator as O_2, SO_2, CO_2 and ½ H_2O)

$$\%\ \text{selectivity} = 100\ \frac{(\text{Mols CaS to } SO_2)}{(\text{Mols CaS to } SO_2 + \text{mols CaS to } CaSO_4)}$$

The total mols of regenerator off gas are calculated from nitrogen balance.

Other calculations should be self explanatory from the programme ZKDAT, a listing of which is attached. The values calculated by ZKDAT are written into binary files JIMX, where X = run number.

The actual output tables are produced by a series of programmes ZKWRT1, ZKWRT2 etc. which read files JIMX and produce the output tables in the correct format.

CUMULATIVE SULPHUR BALANCE

All the data for an overall sulphur balance are contained in files JIMX. The following data are abstracted.

<u>Sulphur input</u> = moles of sulphur in the fuel oil
+ moles of sulphur in the limestone
+ moles of sulphur in recycle gas (taken at half the concentration in the flue gas).

Note: the contributions from the limestone and recycle gas are usually negligible.

<u>Sulphur output</u> = SO_2 in flue gas x boiler dry moles
+ SO_2 in regenerator off gas
+ total sulphur in solids removed (calculated as interpolated total sulphur x interpolated solids).

The error in the interpolated values may be considerable but, since the contribution of sulphur in the solids to the total balance amounts to only a few percent, the absolute error of this contribution is also small.

The accumulated hourly sulphur inputs and outputs and the accumulated difference are printed out in Appendix tables BV1,CV1 & DV1. In both runs 6 and 7 the cumulative balance runs close to 100% throughout the run.

CUMULATIVE SOLIDS BALANCE

Data for these balances was taken from files JIMX. Solids balance was calculated in terms of kilograms equivalent burned stone (e.b.s.)

For limestone:

$$\text{e.b.s.} = \text{wt of limestone} \times (1.0 - 0.01 \times \%\ CO_2 \text{ in stone})$$

For removed solids:

$$\text{e.b.s.} = \text{wt of solids} \times (1.-.01 \times C - .00501 \times S - .01996 \times SO_4)$$

where C = total carbon wt.%
S = total sulphur, wt.%
SO_4 = total sulphur present as sulphate, wt.%

input e.b.s. = limestone e.b.s.

output e.b.s. = sum of interpolated e.b.s. removed from gasifier, regenerator, regenerator cyclone, elutriator fines, back of the boiler, flue K.O. pot and cyclone and elutriator coarse material when it was not returned to the process.

The cumulative balances printed in Appendix tables, B.VI to D.VI do not include any material removed during shut downs.

SNAPSHOTS

A programme ZKSNAP picks out any specified period of time and calculates average values for that period, using data in files JIMX.

Concentration averages are biased:

$$\text{mean concentration} = \frac{\Sigma \text{concentration} \times \text{rate}}{\Sigma \text{rate}}$$

Other values are calculated as arithmetic averages.

In addition to the mean values, the programme calculates root mean square deviations which serve as a measure of stability of the data during that period. For values which are averaged directly the root mean square deviation is:

$$\text{S.D.} = \sqrt{\frac{\Sigma X_i^2 - (\Sigma X_i)^2/N}{N}}$$

where X_i = i th value of a variable measured
N = total number of values averaged.

For more complex values the r.m.s. deviations are combined by the usual rules of combining standard deviations.

A listing of ZKSNAP is attached.

ZKSTINT 05/22/74

```
C THIS PROG. READS DATA FROM STX AND INTERPOLATES IT TO HOURLY
C VALUES OF SOLIDS WITHDRAWALS.  THE INPUT DATA IS EXPECTED IN THE
C FORM:  D.H, GASIFIER, REGENERATOR, REGEN. CYCLONE, ELUTR. FINES
C BOILER BACK, BOILER FLUE (CYCLONE+K.O.), ELUTR. COARSE,  ALL
C QUANTITIES IN LB.  PROG. CONVERTS THE WEIGHTS TO KILO.
C DHE=END OF RUN PERIOD,  DHB=END OF SHUT-DOWN PERIOD, BOTH D.H
C SD=TOTAL ACCUMULATED LOST HOURS FROM D.H=0.0000
C DH0=START ZERO=START D.H-.01,IN D.H, DHF=FINAL HOUR, D.H
 FILELIST "ST6","STBIN6","SD6"
 DIMENSION A(8,400),B(8,500),D(8),DHE(20),DHB(20),SD(21)
 REWIND 1
 REWIND 2; ENDFILE 2
 REWIND 3
 DO 145 J=1,8; DO 144 K=1,500;144 B(J,K)=0.;145 CONTINUE
 K=1
 READ(3,161)DH0,DHF,NUM,(DHE(J),DHB(J),J=1,NUM)
 DHF=FNDH(DHF)
 160 READ(1,161,END=210)D; 161 FORMAT(V)
 IF(D(1)-DH0+.0001)160,180,180
 180 A(1,K)=FNDH(D(1))
 DO 201 J=2,8; 201 A(J,K)=D(J)
 K=K+1; IF(A(1,K-1)-DHF)160,160,210
 210 KMAX=K-1
 SD(1)=FNDH(DH0)
 DO 241 J=1,NUM
 DHE(J)=FNDH(DHE(J)); DHB(J)=FNDH(DHB(J));SD(J+1)=SD(J)+DHB(J)
& -DHE(J)-1.
 241 CONTINUE
 250 DO 460 J=2,8
 IC=1;TK=0.; S=0.; TB=0.; ACUM=0.
 DO 452 K=1,KMAX
 RT=A(1,K)
 300 IF(RT-DHE(IC))340,340,310
 310 IF(RT-DHB(IC))320,330,330
 320 IF(A(J,K))450,321,321
 321 IF(J-3)450,450,322; 322 IF(J-7)323,323,450
 323 ACUM=ACUM+A(J,K); GO TO 450
 330 IC=IC+1; IF(SD(IC)-SD(IC+1)-3.1)331,331,332
 331 TB=TB-(SD(IC)-SD(IC-1))
 IF(A(J,K))450,333,333;333 A(J,K)=A(J,K)+ACUM;ACUM=0.;GO TO 341
 332 IF(A(J,K))450,334,334;334 ACUM=0.; GO TO 341
 340 IF(A(J,K))450,341,341
 341 RT=RT-SD(IC)
 A(J,K)=A(J,K)*.45359
 IF(J-3)370,370,360; 360 IF(J-7)380,380,370
 370 KN=INT(RT-1.5)
 IF(KN)450,450,371
 371 IF (A(J,K))450,373,373
 373 B(J,KN)=A(J,K); GO TO 450
 380 IF(RT-(TK+1.))390,400,400
```

```
390 S=S+A(J,K); TB=RT; GO TO 450
400 EM=A(J,K)/(RT-TB); DEL=(TK+1.-TB)*EM
KN=INT(TK+1.1); B(J,KN)=S+DEL; S=0.; A(J,K)=A(J,K)-DEL
420 TK=TK+1.
IF(RT-(TK+1.))390,390,440
440 KN=INT(TK+1.1); B(J,KN)=EM; A(J,K)=A(J,K)-EM; GO TO 420
450 CONTINUE
452 CONTINUE
460 CONTINUE
KMAX=INT(DHF-SD(IC+1)+.1); IC=1
DO 580 J=1,KMAX
RT=J; RT=RT+SD(IC)
IF(RT-DHE(IC)+.01)520,510,510
510 IC=IC+1
520 X=INT(RT/24.); XX=RT-24.*X; XXX=INT(XX)
XX=INT((XX-XXX)*6.+.1)
B(1,J)=X+XXX/100.+XX/1000.
DO 550 I=1,8; 550 D(I)=B(I,J)
WRITE(2)D

580 CONTINUE
STOP; END
FUNCTION FNDH(V)
X=INT(V); XX=INT(100.*(V-X)+.001)
FNDH=24.*X+XX+(V-X-.01*XX)*1000./6.
RETURN; END
```

ZKANINT   05/24/74

```
100C INTERPOLATION OF ANALYTICAL SOLIDS DATA FROM SX,OX OR CX
110 FILELIST "S6","SB6","SD6"
120 DIMENSION A(8,400),B(8,500),D(8),C(8),DHE(20),DHB(20),SD(21)
130 REWIND 1
140 REWIND 2; ENDFILE 2
150 REWIND 3
160 READ(1,161)V1
170 DO 145 J=1,8; DO 144 K=1,500; 144 B(J,K)=0.; 145 C(J)=V1
180 K=1
190 READ(3,161)DH0,DHF,NUM,(DHE(J),DHB(J),J=1,NUM)
200 DHF=FNDH(DHF)
210 160 READ(1,161,END=210)D; 161 FORMAT(V)
220 IF(D(1)-DH0+.0001)160,180,180
230 180 A(1,K)=FNDH(D(1))
240 DO 201 J=2,8; 201 A(J,K)=D(J)
250 K=K+1; IF(A(1,K-1)-DHF)160,160,210
260 210 KMAX=K-1
270 SD(1)=FNDH(DH0)
280 DO 241 J=1,NUM
290 DHE(J)=FNDH(DHE(J)); DHB(J)=FNDH(DHB(J));SD(J+1)=SD(J)+DHB(J)
300& -DHE(J)-1.
310 241 CONTINUE
320 250 DO 460 J=2,8
330 IC=1; KT=1;TB=0.
340 DO 452 K=1,KMAX
350 RT=A(1,K)
360 IF(A(J,K))450,300,300
370 300 IF(RT-DHE(IC))340,340,310
380 310 IF(RT-DHB(IC))450,330,330
390 330 IC=IC+1; IF(SD(IC)-SD(IC+1)-3.1)331,331,332
400 331 TB=TB-(SD(IC)-SD(IC-1))
410 332 CONTINUE
420 340 CONTINUE
430 341 RT=RT-SD(IC)
440 IF(RT-TB-.01)401,401,400
450 401 CONST=A(J,K); EM=0.; GO TO 431
460 400 EM=(A(J,K)-C(J))/(RT-TB); CONST=A(J,K)-EM*RT
470 431 CONTINUE
480 KTMAX=INT(RT)
490 DO 440 KN=KT,KTMAX; T=KN;440 B(J,KN)=EM*T+CONST
500 C(J)=A(J,K); TB=RT; KT=KTMAX+1
510 450 CONTINUE
520 452 CONTINUE
530 460 CONTINUE
540 KMAX=INT(DHF-SD(IC+1)+.1); IC=1
550 DO 580 J=1,KMAX
560 RT=J; RT=RT+SD(IC)
570 IF(RT-DHE(IC)+.01)520,510,510
580 510 IC=IC+1
590 520 X=INT(RT/24.); XX=RT-24.*X; XXX=INT(XX)
```

ZKANINT   05/24/74

```
XX=INT((XX-XXX)*6.+.1)
B(I,J)=X+XXX/100.+XX/1000.
DO 550 I=1,8
IF(B(I,J))551,550,550; 551 B(I,J)=0.; 550 C(I)=B(I,J)
WRITE(2)C

580 CONTINUE
STOP; END
FUNCTION FNDH(V)
X=INT(V); XX=INT(100.*(V-X)+.001)
FNDH=24.*X+XX+(V-X-.01*XX)*1000./6.
RETURN; END
```

ZKDAT 06/10/74

```
100C PROG FOR WORKUP OF FINES AND ANALYTICAL DATA FROM FILE BRXAN
110C AND RUN DATA FROM FILE BRXDATA, OUTPUT TO JIMX, X=RUN NO.
120 FILELIST "BR7AN","BR7DATA","JIM7"
130 DIMENSION D(35),ZA(29),CAO(6),CO2(6),S(6),DHF(6),A(95)
140 REWIND 1$ REWIND 2$ REWIND 3$ENDFILE 3
150 READ(1)ZA
160 COIL=ZA(1)/1201.
170 HOIL=ZA(2)/201.6
180 SOIL=ZA(3)/3206.
190 SG60=ZA(4)
200 TOIL=200.
210 NLIMES=INT(ZA(5)+.01)
220 160 FORMAT(V)
230 DO 190 I=1,NLIMES
240 CAO(I)=ZA(4*I+2)$ CO2(I)=ZA(4*I+3)
250 S(I)=ZA(4*I+4)$ DHF(I)=ZA(4*I+5)
260 190 CONTINUE
270 DO151 J=1,29$ 151 A(J)=ZA(J)
280 DO 154 N=1,2$ READ(1)ZA$ DO 153 J=1,29$ 153 A(J+29*N)=ZA(J)
290 154 CONTINUE
300 DO 155 J=88,95$ 155 A(J)=0.
310 WRITE(3)A
320 IHR=0
330 IDT=1
340 ILIME=1
350 240 READ(2,END=2200)D
360 READ(1)ZA
370 IF(INT(10000.*(D(1)+.00001))-INT(10000.*(ZA(1)+.00001)))156,157,156
380 156 PRINT 147,D(1),ZA(1)$ 147 FORMAT(1H ,"INCOMPATIBLE D.H ",2F10.4)
390 157 EBST=0.$ SS=0.$ CT=0.
400 DO 158 J=1,7$ EBS=100.-ZA(22+J)-.50094*ZA(8+J)-1.9963*ZA(15+J)
410 EBST=EBST+.01*EBS*ZA(1+J)$ SS=SS+.01*ZA(8+J)*ZA(1+J)
420 CT=CT+.01*ZA(22+J)*ZA(1+J)
430 158 CONTINUE
440 SRT=.01*ZA(4)*ZA(12)
450 DO 159 J=1,29$ 159 A(66+J)=ZA(J)
460 241 A(52)=0.
470 IF (D(1)-DHF(ILIME))250,250,242
480 242 ILIME=ILIME+1
490 A(52)=1.
500 250 A(1)=D(1)
510 F=1.782E-4*CAO(ILIME)/SOIL
520 DO 270 I=2,5
530 270 A(I)=.24884*D(I)
540 A(51)=D(6)/10.
550 A(6)=25.4*D(4)/D(6)
560 DO 300 I=7,10
570 300 A(I)=D(I)
580 IF(A(10)-10.)424,424,422
590 422 A(10)=10.
```

ZKDAT   06/10/74

```
 424 CONTINUE
C     COMPUTE GAS COMPOSITIONS
C           BOILER CO2
 Z=(D(11)/10.)-1.
 A(11)=.14181778E-1+2.321528*Z-.51096758*Z↑2+.13242664*Z↑3
&-.13988354E-1*Z↑4+6.0512861E-4*Z↑5
C           REGENERATOR CO2
  Z = (D(13)/10.)-1.
 A(13)=1.1608245E-3+1.1574973*Z-.27330138*Z↑2+.075628236*Z↑3
&-.00837005*Z↑4+.36410274E-3*Z↑5
C           REGENERATOR S02
 A(14)=0.10395*(D(14)-10.)↑1.148+1.426*(1.-EXP(-.18*(D(14)-10.)))
 IF(A(14))428,428,429;  428 A(14)=0.0;  429 CONTINUE
 A(15)=D(15)
C           PLENUM GAS CO2
 Z=(D(16)/10.)-1.
 A(16)=.17024395E-1+1.1080589*Z-.16916171*Z↑2+.36508513E-1*Z↑3
&-2.844985E-3*Z↑4+9.7723649E-5*Z↑5
 A(17)=D(17)
C CONVERSION OF MEASURED FLOWS TO M3/H
C GASIFIER AIR
 A(19)=1.628825*(2.42+1.06091*D(19))
C FLUE GAS RECYCLE, P,T AND MOISTURE CORRECTION
 H2OFR=EXP(.04329*D(9)+.3232)
 IF(A(17)-18.)615,615,612
 612 DFGR=A(19)*(21.-A(17))/(A(17)-A(10))
 A(18)=DFGR/(1.-.01*H2OFR)
 GO TO 641
 615 CONTINUE
 Z=D(18)*SQRT(34.247*(D(28)+407.)/((D(9)+273.)*(30.4-.124*H2OFR)))
 A(18)=1.628825*(4.556+.97121*Z)
 DFGR=A(18)*(1.-.01*H2OFR)
 641 CONTINUE
 H2OR=.01*A(18)*H2OFR
C AIR TO INJECTOR AND BURNER
 A(20)=1.628825*D(20)
C AIR TO STONE FEED
 A(21)=D(21)*SQRT(D(31)+14.7)*1.5001E-2
C AIR TO FINES RETURN
 D(34)=12.969932+4.133439*D(34)+.066331712*(D(34)↑2.)
&-1.3322994E-3*(D(34)↑3.)+1.5393257E-5*(D(34)↑4.)
 A(34) = 1.628825*D(34)*SQRT(D(35)+407.)*1.7505E-3
C REGEN AIR
 A(25)=1.628825*D(25)*(407.+D(23))/24420.
C N2 TO STONE
 A(22)=.4248*D(22)*SQRT(D(31)+14.7)
C REGEN N2
 A(29)=1.628825*D(29)*(407.+D(30))/24420.
C PILOT PROPANE
 A(27)=.3427*D(27)*SQRT(D(26)+14.7)
```

```
1100 A(24)=4.536924*D(24)*(.9627+.06*SG60-.00036*TOIL)*SG60
1110C REGEN GAS FLOW BY N2 BALANCE
1120 QREG=(79.*A(25)+100.*A(29))/(100.-A(14)-A(13)*1.25-A(15))
1130 A(32)=.453592*D(32)
1140C BOILER GAS FLOW
1150 QPROP=A(27)/22.1209
1160 BC=A(24)*COIL+3.*QPROP-A(13)*QREG/2212.09
1170 CSTG=A(32)*CO2(ILIME)/4401.
1180 CFGR=DFGR*A(11)/2271.1
1190 CT=CT/12.01; CHT=CT/4.
1200 BH2O=A(24)*HOIL+4.*QPROP
1210 BN=3.7619*(BC-CT)+1.881*(BH2O-CHT)+A(22)/22.711
1220 BDM=BC-CT+CSTG+CFGR+BN
1230C CORRECTION FOR EXCESS AIR
1240 Z=A(10)/(21.-A(10))*BDM
1250 BO=.21*Z
1260 BN=BN+.79*Z
1270 BDM=BDM+Z
1280C CORRECTION OF BOILER SO2 FOR SOLUTION IN CONDENSED WATER
1290 BSO2=1.02362E-6*BDM*D(12)
1300 IF(BSO2)1039,1039,990
1310 1039 A(12)=0.; BSO2=0.;  GO TO 1040
1320 990 BSO2C=EXP((ALOG(7.6E-4*D(12))-1.246)/1.282)
1330 BSO2C=BSO2C*(BH2O-CHT+H2OR/22.711-.02362*BDM)*2.8143E-4
1340 BSO2=BSO2+BSO2C
1350 A(12)=BSO2/(BDM*1.02362E-6)
1360C SULPHUR REMOVAL
1370 1040 SIN=A(24)*SOIL+A(32)*S(ILIME)/3206.+DFGR*A(12)/4.542E7
1380 SS=SS/32.06
1390 SREM=100.*(SIN-BSO2)/SIN
1400C S OUT IN REGENERATOR
1410 SRT=SRT/32.06
1420 SREG=QREG*A(14)/2271.1
1430 SR=100.*SREG/SIN
1440C SELECTIVITY
1450 CASO4=.5*(.21*A(25)-.01*QREG*(A(15)+1.5*A(14)+1.25*A(13)))/22.711
1460 CAOR=QREG*A(14)/2271.1
1470 CAOS=100.*CAOR/(CAOR+CASO4)
1480C  STOICHIOMETRY
1490 O2AC=(A(19)+A(20)+A(34))*.0092466+DFGR*A(10)/2271.1
1500 O2ST=A(24)*(COIL+.5*HOIL+SOIL)
1510 ST=O2AC/O2ST*100.
1520C  PLENUM CONC. OF O2 AND CO2
1530 PO2=(A(19)*21.+DFGR*A(10))/(A(19)+DFGR)
1540 PCO2=DFGR*A(11)/(A(19)+DFGR)
1550C BED VELOCITY
1560 BV=8.5386E-4*(A(18)+A(19)+A(20)+A(34)+H2OR)*(A(8)+273.)/(D(5)+407.)
1570C  REGEN VELOCITY
1580 RV=.01453*(A(25)+A(29))*(A(7)+273.)/(D(5)+407.)
1590C CAO/S STONE FEED RATIO, MOLAR
```

```
CASR=F*A(32)/A(24)
C CONVERSION TO ARRAY AND OUTPUT TO FILE
A(23)=CASR
A(26)=BV
A(28)=PO2
A(30)=PCO2
A(31)=ST
A(35)=SR
A(36)=SREM
A(37)=SIN
A(38)=SREG
A(39)=BSO2
A(40)=BDM
A(41)=BO
A(42)=BN
A(43)=BC
A(44)=CSTG
A(45)=RV
A(46)=O2ST
A(47)=O2AC
A(48)=CAOR
A(49)=CASO4
A(58)=CAOS
A(50)=A(19)+A(20)+A(34)
A(53)=BH2O
A(54)=QREG
A(55)=H2OR
A(56)=DFGR
A(57)=D(12)
A(59)=CFGR; A(60)=EBST; A(61)=SS; A(62)=CT; A(63)=CHT
A(64)=SRT; A(65)=A(32)*(1.-.01*CO2(ILIME))
IF(IHR)2080,2080,2060
2060 IDT=(A(1)-XDT+.00001)*100.
GO TO 2090
2080 IHR=1
XDT=A(1)
2090 IF(IDT-77)2100,2160,2110
2100 IF(IDT-1)2160,2170,2120
2110 IDT=IDT-76
2120 ISHUT=0
2130 IF(IDT-76)2150,2140,2140
2140 ISHUT=ISHUT+24; IDT=IDT-100
GO TO 2130
2150 IDT=IDT+ISHUT; GO TO 2170
2160 IDT=1
2170 A(33)=XDT
XDT=A(1); A(66)=IDT; A(66)=A(66)+.00001
WRITE(3)A
GO TO 240
2200 STOP ; END
```

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```
100 FILELIST "JIM5"
110 DIMENSION A(95),CAO(6),DHF(6),X(24),P(8),Y(8),SUM(40),SQ(40)
120 DIMENSION Z(42),Z1(28)
130 DIMENSION DR(6),DC(6)
140 REWIND 1
150 DR(1)=2.1530; DC(1)=3.0030; DR(2)=3.0530; DC(2)=3.1230
160 DR(3)=11.2130; DC(3)=12.1030; DR(4)=12.1430; DC(4)=13.0330
170 DR(5)=16.1030; DC(5)=16.1830; DR(6)=17.1030; DC(6)=17.1930
180 READ(1)A
190 NLIMES=INT(A(5)+.01)
200 DO 150 J=1,NLIMES; CAO(J)=A(4*J+2);150 DHF(J)=A(4*J+5)
210 DO 1155 JDH=1,6
220 DH1=DR(JDH); DH2=DC(JDH)
230 ILIME =1;160 IF(DH1-DHF(ILIME)+.00001)171,171,170
240 170 ILIME=ILIME+1; GO TO 160; 171 CONTINUE
250 DO 180 J=1,40; SUM(J)=0.; 180 SQ(J)=0.; NUM=0
260 190 READ(1,END=1900)A
270 IF(INT(10000.*DH1+.1)-INT(10000.*A(1)+.1))210,210,190
280 210 IF(INT(10000.*DH2+.1)-INT(10000.*A(1)+.1))500,220,220
290 220 IF(A(36)-99.99)230,230,190
300 230 NUM=NUM+1
310 X(1)=A(8); X(2)=A(7); X(3)=A(9)
320C X1=GAS'R T,X2=REG T, X3=G.REC. T
330 X(4)=A(24); X(5)=A(32); X(6)=A(19)+A(20)+A(34); X(7)=A(56)
340C X4=OIL RATE, X5=STONE, X6=TOT. AIR,X7=DRY F.G.
350 X(8)=A(55); X(9)=A(25); X(10)=A(27);X(11)=A(6); X(12)=A(51)
360C X8=REC.ST.,X9=REG AIR,X10=PILOT C3,X11=BED DEPTH,X12=BED SG.
370 X(13)=A(2)/.098064; X(14)=A(37); X(15)=A(26); X(16)=A(45)
380C X13=REG.DP.,X14=SIN MOLS,X15=BED VEL.,X16=REG. VEL.
390 IF(X(11)-53.34)360,370,370
400 360 X(17)=X(11)*(4177.4+.095238*X(11)*(69.215+.0317*X(11)))
410& *X(12)/1000.; GO TO 380
420 370 X(17)=X(12)*(242.035+4.9064*(X(11)-53.34))
430 380 CONTINUE
440 X(18)=A(39)
450C X17=BED WT. KG., X18=SOUT MOLS
460 X(19)=A(75); X(20)=A(82); X(21)=A(89)
470C X19=GS, X20=GSO4, X21=GC
480 X(22)=A(76); X(23)=A(83); X(24)=A(90)
490C X22=RS, X23=RSO4, X24=RC
500 Y(1)=A(15); Y(2)=A(13); Y(3)=A(14); P(1)=A(54); P(2)=P(1);P(3)=P(1)
510C Y1=RO2, Y2=RCO2, Y3=RSO2, P1=REG. GAS FLOW
520 Y(4)=A(17); Y(5)=A(16); P(4)=A(19)+A(56);P(5)=P(4)
530C Y4=PLENUM O2, Y5=PLENUM CO2, P4=PLENUM GAS FLOW (DRY)
540 Y(6)=A(10); Y(7)=A(11); Y(8)=A(12); P(6)=A(40); P(7)=P(6); P(8)=P(6)
550C Y6=BO2, Y7=BCO2, Y8=BSO2, P6=BOILER DRY MOLES
560 DO 450 J=1,24; SUM(J)=SUM(J)+X(J); 450 SQ(J)=SQ(J)+X(J)*X(J)
570 DO 460 J=1,8; X(J)=P(J)*Y(J); SUM(24+J)=SUM(24+J)+X(J)
580 SQ(24+J)=SQ(24+J)+X(J)*X(J); SUM(32+J)=SUM(32+J)+P(J)
590 460 SQ(32+J)=SQ(32+J)+P(J)*P(J)
```

```
GO TO 190
C BEGINNING OF AVERAGING
500 XNUM=NUM
DO520 J=1,40; SQ(J)=SQ(J)-SUM(J)*SUM(J)/XNUM
IF(SQ(J))521,521,523
521 SQ(J)=0.; GO TO 524
523 SQ(J)=SQRT(SQ(J)/(XNUM-1.)); 524 CONTINUE
520 SUM(J)=SUM(J)/XNUM
DO 550 J=25,32; X(J-24)=SUM(J)/SUM(J+8)
IF(SUM(J))542,542,544; 542 SQ(J)=0.; SUM(J)=1.
544 IF(SUM(J+8))546,546,548
546 SQ(J+8)=0.; SUM(J+8)=1.
548 CONTINUE
SQ(J)=X(J-24)*SQRT((SQ(J)/SUM(J))**2+(SQ(J+8)/SUM(J+8))**2)
550 SUM(J)=X(J-24)
C CALC. OF SRE
X(1)=SUM(18)/SUM(14)
X(8)=(1.-X(1))*SQRT((SQ(18)/SUM(18))**2+(SQ(14)/SUM(14))**2)*100.
X(7)=(1.-X(1))*100.
X(5)=SUM(5)*CAO(ILIME)/5607.; X(6)=SQ(5)*CAO(ILIME)/5607.
X(5)=X(5)/SUM(14)
X(6)=X(5)*SQRT((SQ(14)/SUM(14))**2+(X(6)/X(5))**2)
C CALC. OF SELECTIVITY
X(1)=SQ(25)*SQ(25)+1.5*SQ(27)*SQ(27)+1.125*SQ(26)*SQ(26)
X(2)=SUM(25)+SUM(26)*1.125+SUM(27)*1.5
X(3)=.01*((SQ(33)/SUM(33))**2+X(1)/(X(2)*X(2)))*SUM(33)*X(2)
X(1)=.5*SQRT(.21*SQ(9)*SQ(9)+X(3))/22.711
X(2)=.5*(.21*SUM(9)-.01*SUM(33)*X(2))/22.711
C CALC. OF CA/S RATIO
X(4)=SUM(33)*SUM(27)/2271.1
X(3)=X(4)*SQRT((SQ(33)/SUM(33))**2+(SQ(27)/SUM(27))**2)/2271.1
SEL=100.*X(4)/(X(4)+X(2))
SELSQ=SEL*SQRT((X(3)/X(4))**2+(X(1)*X(1)+X(3)*X(3))/(X(4)+X(2))**2)
Z(1)=DH1; Z(2)=DH2; NUM=INT(100.*(DH2-DH1)+.1)+1
IF(NUM-76)781,781,760; 760 NUM=NUM-76; N=0
770 IF(NUM-100)780,771,771; 771 NUM=NUM-100; N=N+24; GO TO 770
780 NUM=NUM+N; 781 Z(3)=NUM; Z(4)=XNUM
DO 790 J=1,13; Z(2*J+3)=SUM(J); 790 Z(2*J+4)=SQ(J)
Z(31)=X(5); Z(32)=X(6); Z(33)=SUM(15); Z(34)=SQ(15)
Z(35)=SUM(16); Z(36)=SQ(16); Z(37)=X(7); Z(38)=X(8)
Z(39)=SEL; Z(40)=SELSQ; Z(41)=SUM(17); Z(42)=SQ(17)
PRINT 1000,Z
DO 840 J=1,8; Z1(2*J-1)=SUM(24+J); 840 Z1(2*J)=SQ(24+J)
DO 850 J=1,6; Z1(15+2*J)=SUM(18+J); 850 Z1(16+2*J)=SQ(18+J)
PRINT 1200,Z1
1000 FORMAT(1H ,//////22X15HRUN 5  SNAPSHOT/
&14X4HFROM,F8.4,3H TO,F8.4,F6.0,10H HOURS RUN/19XF5.0,
&19H USEFUL DATA POINTS//3X8HVARIABLE12X5HUNITS7X4HMEAN7X4HS.D.
&/17H GASIFIER   TEMP.6X5HDEG.C4XF8.1,F10.1/16H REGENERATOR -"-
&8X3H-"-5XF8.1,F10.1/8H RECYCLE5X3H-"-8X3H-"-5XF8.1,F10.1/9H OIL FED
```

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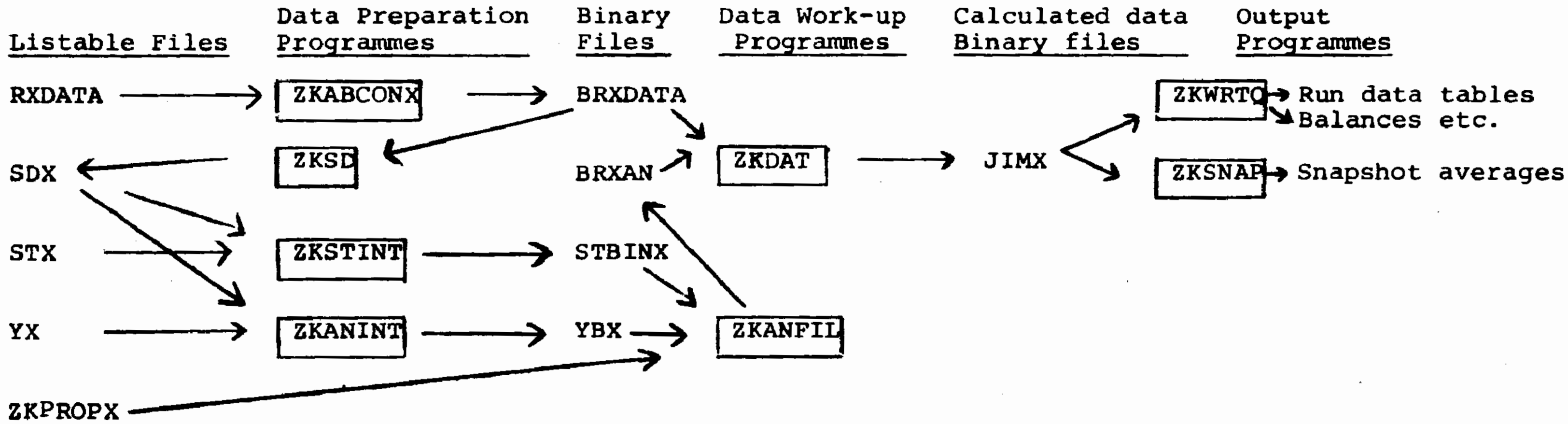
```
&14X5HKG/HR4XF8.1,F10.1/6H STONE5H FEED13X3H-"-5XF8.1,F10.1/
&10H TOTAL AIR13X6HNM3/HR3XF8.1,F10.1/13H DRY FLUE GAS11X3H-"-5X
&F8.1,F10.1/16H STEAM WITH F.G.8X3H-"-5XF8.1,F10.1/10H REGEN.AIR13X
&3H-"-5XF8.1,F10.1/14H PILOT PROPANE10X3H-"-5XF8.1,F10.1/
&10H BED DEPTH13X2HCM7XF8.1,F10.1/9H BED S.G.11X13HDIMENSIONLESS
&F9.3,F10.3/16H REGEN BED DEPTH7X2HCM7XF8.1,F10.1/15H CA/S MOL. RAT.
&5X13HDIMENSIONLESS,F8.2,F10.2/15H GAS'R BED VEL.8X5HM/SEC5XF8.2,
&F10.2/14H REGEN -"- -"-10X3H-"-6XF8.2,F10.2/7H S.R.E.16X7HPERCENT
&2XF8.1,F10.1/12H SELECTIVITY12X3H-"-5XF8.1,F10.1/11H BED WEIGHT13X
&2HKG6XF8.1,F10.1/)
1200 FORMAT(1H ,/20X22HANALYSES WT. OR VOL. %/23X2HO214X3HCO211X
&3HSO2/3X5HGASES12X4HMEAN3X4HS.D.4X4HMEAN3X4HS.D.4X4HMEAN3X4HS.D./
&12H REGENERATOR4XF8.2,F7.2,F8.1,F6.1,F8.1,F6.1/13H PLENUM (DRY)
&F10.1,F7.1,F9.1,F6.1,6X1H-5X1H-/15H FLUE GAS (DRY),F8.1,F7.1,F9.1,
&F6.1,F7.0,5H(PPM),F4.0//3X6HSOLIDS13X6HTOT. S8X5HSO4 S10X6HTOT. C/
&9H GASIFIER3XF11.2,F7.2,F10.2,F5.2,F9.2,F7.2/12H REGENERATOR,F11.2,
&F7.2,F10.2,F5.2,F9.2,F7.2//////)
1155 CONTINUE
1900 STOP; END
```

Table K-1

Data Flow : Analysis of Continuous Run Data



Note: X = 5, 6 or 7
Y = S, O or C
Q = 1, 2 or 3

Table K-2

DESCRIPTION OF DATA POINTS ENTERED IN RXDATA

D(1) = Day-time in the form X.YYZZ where X = day from the start of the run, YY = hour (24 hr clock), ZZ = minutes. e.g. 8.37 in the morning on 5th day = 5.0837
D(2) = pressure drop across the regenerator bed inches w.g.
D(3) = " " " " gasifier distributor, inches w.g.
D(4) = " " " " " bed, inches w.g.
D(5) = gasifier gas space pressure, inches w.g.
D(6) = pressure drop across 10" of gasifier bed, inches w.g.
D(7) = regenerator temperature, °C
D(8) = gasifier temperature, °C
D(9) = flue gas recycle temperature, °C
D(10) = flue gas O_2 content (Servomex), % by volume
D(11) = " " CO_2 (Maihak), 10-100 scale reading
D(12) = " " SO_2 (Wostoff), ppm by volume
D(13) = regenerator CO_2 (Maihak), 10-100 scale reading
D(14) = " SO_2 ("), 10-100 " "
D(15) = " O_2 (Servomex), % by volume
D(16) = gasifier plenum CO_2 (Maihak), 10-100 scale
D(17) = " " O_2 (Servomex), % by volume
D(18) = flue gas recycle rate, orifice plate reading, c.f.m.
D(19) = air rate to gasifier, " " " "
D(20) = injector + burner air, sum of several rotameter readings, c.f.m.
D(21) = air rate with stone feed, rotameter reading, liters/min
D(22) = nitrogen rate with stone feed, rotameter reading, c.f.m.
D(23) = regenerator air meter pressure, inches w.g.
D(24) = oil rate, positive displacement meter read at hourly intervals (using stop-watch), I.gal/hr
D(25) = regenerator air rate, positive displacement flow meter read at hourly intervals (using stop-watch), c.f.h.
D(26) = pilot propane pressure, p.s.i.
D(27) = pilot propane rate, rotameter reading, c.f.m.
D(28) = flue gas recycle pressure, inches w.g.
D(29) = stone transfer nitrogen to regenerator, positive displacement meter hourly reading (using stop-watch) c.f.h.
D(30) = stone transfer nitrogen to regenerator pressure, inches w.g.
D(31) = limestone feed hopper pressure p.s.i.
D(32) = limestone feed rate, from load cell chart, lb/hr
D(33) = spare
D(34) = air to fines return, rotameter reading, 0-30 scale
D(35) = " " " " pressure, inches w.g.

APPENDIX L

GASIFIER PRODUCT COMPOSITION

Samples of gasifier product were collected in metal containers and analysed for composition several times during Run 5. Since the coke, tar, and heavy hydrocarbon portion of the product is not measured by this procedure, an attempt has been made to estimate the quantity and composition by material balance on the gasifier itself.

The dry gas compositions measured were listed in Table 11 of the text, and a summary of the material balance results and estimated heavy product compositions was listed in Table 12.

The calculations made involve several assumptions. Examples of these calculations and assumptions are presented in this appendix.

Data from the operating period 26.0330 are used in the examples.

- Air flow to Gasifier, SCFM

231	To plenum and fuel injectors
2.8	with stone feed
2	to cyclone inlet
235.8 SCFM	

$$O_2 \text{ Feed} = 235.8 \times .03316 \ \frac{\text{lb Mole}/O_2/\text{Hr}}{\text{SCFM Air}}$$

$$= 7.819 \text{ lb Mole/Hr } O_2$$

- Flue Gas to Gasifier

Oxygen and CO_2 are measured on dried samples of flue gas in continuous analysers. The rate of flue gas flow to the gasifier is measured by orifice meter on wet gas. A correction is therefore needed for the composition change due to water.

$$\text{Dry Gas Rate} = (\text{Wet Gas Rate}) \left\{\frac{100/Y}{100/Y + X/Z}\right\}$$

Where Y = Vol.% CO_2 in Dry Gas
X = Vol.% H_2O in Wet Gas
Z = Vol.% CO_2 in Wet Gas

The ratio X/Z is assumed to be .818 based on the composition of the fuel oil. This assumption implies that composition of the flue gas does not change during passage through the flue gas scrubber.

For the example period, Y = 14.1

Wet Gas Rate = 86 SCFM

$$\text{Dry Gas Rate} = 86 \left\{\frac{100}{14.9}\right\} \Big/ \left\{\frac{100}{14.1} + .818\right\} = 77.1 \text{ SCFM}$$

$$\text{Dry Gas Rate} = 77.1 \times \frac{60}{380} = 12.17 \text{ lb Mole/hr}$$

Inputs with Flue Gas

Mole/Hr CO_2 = Mole/Hr Flue Gas x Mole Fraction CO_2

CO_2 = 12.17 x .141 = 1.717 Mole/hr
O_2 = 12.17 x .022 = .268
N_2 = 12.17 x .837 = 10.19

Mole/Hr H_2O = Wet Gas Rate - Dry Gas Rate

$$H_2O = (86 - 77.1) \times \frac{60}{380} = 1.404$$

Inputs from Solids

Oxygen from regenerator sulphate

$$= (\text{Lime Circulation, lb/Hr}) (\text{Fraction S as } SO_4) * \frac{2}{32}$$

$$O_2 \text{ from Sulphate} = 833 \times \frac{.0137}{16} = .713 \text{ Mole/Hr}$$

Oxygen from fresh stone Carbonate

= (stone rate, lb/Hr) (Wt. Fraction CO_2 in Stone)/44

= 32 x .434/44 = 0.316 Mole O_2/Hr.

Oxygen from $CaO + H_2S \rightarrow CaS + H_2O$

= (Mole/Hr S to gasifier) (S removal efficiency)/2

= (409.1 x .0248/32) (.906)/2 = .144 Mole O_2/Hr

Total O_2 Inputs to Gasifier

Air		7.819
Flue Gas	O_2	.268
	CO_2	1.717
	H_2O	.702
Sulphate		.713
Stone CO_2		.316
CaO		.144
		11.679 lb Mole O_2/Hr

Nitrogen Inputs

Bed Air + Cyclone inlet flow + Stone feed Air

= O_2 Rate x .79/21

= 7.819 x .79/.21 = 29.414 Mol/Hr N_2

Bed Feed N_2 Bleed	=	.351	Mole/Hr
Injector N_2	=	.16	Mole/Hr
Flue Gas	=	10.19	Mole/Hr
Pressure Tapping Bleeds	=	.03	
Total N_2 Inputs		40.145	Mole/Hr

Product Gas Rate = N_2 Rate/fraction N_2

= 40.145/.634 = 63.321 Mole/Hr

Oxygen Outputs in Product Gas C oxides

= (Fraction CO_2 + FrCO/2) (Product Gas Rate)

= (.1088 + .0936/2) (63.321) = 9.85 Mole O_2/Hr

O_2 Out as Sulphate

= (lime Circulation) Fr S as SO_4 = /32) x 2

= 825 x .0029/16 = .149 Mole O_2/Hr

O_2 Out as H_2O

The oxygen leaving the gasifier as water is assumed to be the difference between the O_2 inputs and other known O_2 outputs.

O_2 in Water = 11.679 - 9.85 - .149 = 1.68 Mole/Hr

H_2 Inputs to Gasifier

Hydrogen in with fuel = Fuel Rate x wt. fraction H_2/2

= 409.1 x 0.114/2 = 23.319 Mole H_2/Hr

Hydrogen in with flue gas = 1.404

Total H_2 In = 24.723 Mole/Hr

Hydrogen Output in Product H_2 + Hydrocarbons

Σ (Mole/Hr Product) (Mole fraction x H_2 Multiple)

As H_2	= 63.321 x .0642	= 4.065
As CH_4	= 63.321 x .0609 x 2	= 7.712
As C_2H_4	= 63.321 x .0381 x 2	= 9.825
	Total	16.602

Hydrogen Out as H_2O

Moles H_2 in Water = 2 x Moles O_2 in water

H_2 out in Water = 2 x 1.68 = 3.356

Hydrogen Missing

= Known Inputs - Known Outputs

= 24.723 - 16.602 - 3.356 = 4.765 Mole H_2/Hr

Missing hydrogen is assumed to be in tars, coke and liquid portions of products not measured by gas chromatograph.

Carbon Inputs

With Fuel = Fuel Rate x fraction C/12
= 409.1 x .856/12 = 29.182 Mole C/Hr

With flue gas = CO_2 input = 1.717 Mole C/Hr

With limestone = Stone Rate x Mole fraction CO_2

= 32 x .434/44 = .316 Mole C/Hr

Total Carbon in = 31.215 Mole/Hr

Carbon Outputs

Carbon Out in Product Gas

= Σ Product gas x Mole. fraction x C multiple

As CO_2	= 63.321 x .1088	= 6.889
As CO	= 63.321 x .0936	= 5.927
As CH_4	= 63.321 x .0609	= 3.856
As C_2H_4	= 63.321 x .0381 x 2	= 4.825

Dry gas total C = 21.497 Mole/Hr

Carbon Out Regenerator

= (Mole/Hr Regenerator Gas) x (Mole fraction CO_2)

= N_2 to Regenerator = 2.902 Mole/Hr
N_2 in Regenerator Gas = .951 Mole fraction
O_2 in Regenerator Gas = .017 Mole fraction

CO_2 out = 2.902 x $\frac{.017}{.951}$ = .052 Mol./Hr

Carbon Missing

= Known Inputs - Known Outputs

= 31.215 - 21.497 - .052 = 9.666 Mole C/Hr

Like Hydrogen, the missing carbon is assumed to be in the form of coke, tar, and other heavy components of the product gas not measured by the gas chromatograph.

CO/CO_2 Made in Gasifier

The CO_2 made in the gasifier is taken as CO_2 out less CO_2 from lime stone and flue gas recycle

CO_2 made = 6.889 - .316 - 1.717 = 4.856

CO made = 5.927

CO/CO_2 = 5.927/4.856 = 1.221

H/C Ratio of Heavy Products

The H/C ratio of missing products is taken as 2 x H_2 missing/C missing

H/C = 2 x 4.765/9.666 = .99

Results of these component material balance calculations for four gas samples are summarised here in Table L-1 and in Table 12 of the text.

Table L-1

Summary of Gasifier Component Material Balances

Time	22.1030	22.1730	26.0330	26.1730
O_2 In, Mole/Hr	11.797	11.715	11.679	11.240
O_2 Out, "	10.084	9.753	9.999	8.931
O_2 to H_2O (Diff)	1.713	1.962	1.678	2.309
H_2 to H_2O	3.426	3.924	3.356	4.618
H_2 inputs	24.901	24.852	24.723	24.677
H_2 in products	20.472	19.792	16.602	12.301
H_2 missing	1.003	1.136	4.765	7.758
C inputs	31.441	31.356	31.215	31.255
C in Products	24.439	22.544	21.549	20.367
C missing	7.002	8.812	9.666	10.888
H/C in missing	.286	.258	.986	1.43
CO/CO_2 Made	1.102	.97	1.22	1.36
H missing, % of feed	4.3	4.8	20.4	33.3
H oxidised % of feed	14.6	16.8	14.4	19.9
C missing, % of feed	23.9	30.0	33.1	37.4
C oxidised, % of feed	35.8	34.0	37.0	31.9

APPENDIX M

BATCH UNIT PROCEDURES

GENERAL

Several test procedures were employed to measure sulphur absorption and dust producing characteristics of the stones. These included fresh bed tests in which a new batch of calcined lime was used for each test, cyclic gasification tests in which a single batch of lime was cycled between gasification and regeneration conditions, and kerosene and fuel oil combustion tests in which the appropriate fuel was burned in the fluid bed with excess air to control temperature.

START UP AND CALCINATION

The bed is calcined by combustion of propane below the distributor and by direct injection of kerosene into the bed. First of all, the bed space temperature is raised to 950 deg. C. by gas combustion below the distributor. Then, 4000g of limestone is added. This is heated to 750 deg. C. using gas before switching to direct kerosene injection. Due to the strongly endothermic nature of the calcination reaction, the temperature remains in the region of 800 deg. C. until CO_2 evolution ceases. When the temperature rises to 950 deg. C, indicating that calcination of the original charge is complete, 2500g limestone is added and calcined. Further batches of this size are added and the procedure repeated until the target bed depth is reached.

COMBUSTION TEST

For a test with kerosene combustion, the injection of kerosene is continued with the bed past the point of complete calcination. For fuel oil combustion, the fuel supply is simply switched to heavy fuel oil from its heated supply drum. The fuel rate is set to give a bed temperature slightly in excess of the target, and fine adjustment is carried out with the aid of a cooling coil. Relevant data on gas analysis and unit behaviour are recorded and appropriate bed and cyclone samples taken. Combustion is continued for the required length of time.

GASIFICATION TEST

To achieve gasification, the fuel supply is changed from kerosene to fuel oil when calcination is complete, and oil rate is increased to obtain an air/fuel ratio of about 25% of stoichiometric. The sample flame burner and external flare are lit and the gas analysers connected. Relevant data are recorded and bed samples and cyclone samples taken at prescribed intervals. Gasification is continued for the requisite length of time.

REGENERATION

Regeneration of the sulphided stone from a gasification cycle is performed by stopping the fuel supply and continuing the flow of air. Oxidation of carbon and CaS in the bed raises temperature to the regeneration level. Relevant temperature and analytical data are collected, and a bed sample is taken when regeneration is complete.

SHUT DOWN

At the end of a run, the bed temperature is allowed to drop to 700 deg. C. before the bed is removed through the drain point just above the distributor. Draining the unit is much easier when the bed is hot since the solids flow better under these conditions. When all the bed has been removed, all ancillary equipment is shut off.

CYCLE TESTS

Cyclic tests are the nearest simulation to continuous gasifier operation that can be obtained in batch units. The same charge of lime is subjected to repeated cycles of sulphur absorption and and regeneration. After each regeneration a portion of lime is removed and replaced by an equivalent amount of fresh limestone. The limestone calcines to lime during the early part of the next gasification cycle. Without replacement, the activity of the lime bed gradually declines. With replacement, the activity falls initially, but in a few cycles lines out at an equilibrium level which is influenced by the rate of replacement.

Enough cycles are performed at each set of operating conditions to establish the lined-out sulphur removal efficiency for those conditions.

Cyclic tests use the same calcination and start-up procedures as the fresh bed tests. However, after the initial start, a series of gasification and regeneration cycles follow each other. Sampling and gas analysis procedures also are the same as used in fresh bed tests.

When a regeneration cycle is complete, the fluidisation cools the bed very rapidly. Therefore the following procedure was adopted.

(1) When regeneration was complete as indicated by end of SO_2 emission and fall of bed temperature, fluidising air was stopped.
(2) Replacement limestone was added.
(3) Fluidising air was resumed and when temperature reached the desired point, oil feed was resumed.

For the purpose of stone comparison, the target conditions in each test were the same.

APPENDIX N

Batch Data

TABLE N-1. CYCLE TEST CONDITIONS AND SRE'S (TEST 1-E)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range, μ	Fuel Rate, g/min	Total Air, l/min	Superficial Gas Velocity, m/sec	% Stoich. Air	Bed Depth centimetres Start	Bed Depth centimetres End	Bed Specific Gravity	Average Bed Temp. (Absorption), °C	Max Bed Temp (Regen), °C	Duration of Absorption, min	Make-up Weight (Limestone), g/cycle	Cyclone Dust, g	SRE, %
7/72	1	Amuay(2.3%S)	BCR 1691	600-3175	199	656	1.71	30.2	64.8	52.6	0.67	830	920	40	-	500	100
8/72	2	"	"	"	190	704	1.89	34.0	45.5	43.4	0.75	850	980	"	700	810	100
9/72	3	"	"	"	210	676	1.80	29.5	49.5	47.5	0.75	845	1000	"	"	530	100
10/72	4	"	"	"	200	633	1.68	29.0	52.1	43.4	0.73	845	990	"	"	300	91
11/72	5	"	"	"	204	629	1.68	28.3	49.0	41.6	0.77	855	1000	"	"	285	86
12/72	6	"	"	"	201	638	1.71	29.1	-	44.2	0.79	850	1000	"	"	205	71
13/72	7	"	"	"	216	634	1.68	26.9	48.3	-	0.80	850	1000	"	"	260	72
14/72	8	"	"	"	236	640	1.71	24.9	55.9	48.0	0.76	845	1010	"	"	240	79
15/72	9	"	"	"	206	661	1.74	29.4	45.7	42.7	0.83	830	955	"	"	305	78
16/72	10	"	"	"	186	668	1.77	32.9	50.8	46.5	0.77	850	1000	"	"	190	81
17/72	11	"	"	"	212	664	1.74	28.7	47.2	45.7	0.83	835	1005	"	"	225	57
18/72	12	"	"	"	233	669	1.77	26.3	50.3	45.7	0.83	840	1015	"	"	200	57
19/72	13	"	"	"	224	659	1.74	27.0	54.8	48.8	0.83	840	1015	"	"	245	72
20/72	14	"	"	"	233	647	1.71	25.5	59.9	50.3	0.81	840	1010	"	"	242	72
21/72	15	"	"	"	227	641	1.68	25.9	53.3	50.3	0.83	835	1010	"	"	250	72
22/72	16	"	"	"	224	630	1.68	25.8	62.5	47.0	0.85	840	1010	"	"	260	74
23/72	17	"	"	"	213	656	1.74	28.3	47.2	43.7	0.82	840	990	"	"	300	77
24/72	18	"	"	"	216	638	1.68	27.1	48.8	47.2	0.81	840	1010	"	"	170	73
25/72	19	"	"	"	227	634	1.68	25.6	54.4	44.6	0.83	840	1010	"	"	160	74
26/72	20	"	"	"	233	622	1.62	24.5	54.6	46.5	0.83	830	1010	"	"	182	73
27/72	21	"	"	"	226	633	1.68	25.7	55.6	48.0	0.84	840	1010	"	"	207	72
28/72	22	"	"	"	238	690	1.83	26.6	53.3	49.3	0.84	845	1010	"	"	290	72
29/72	23	"	"	"	244	679	1.80	25.5	54.9	50.8	0.83	845	1020	"	"	260	76
30/72	24	"	"	"	247	677	1.80	25.1	56.4	47.0	0.88	840	1015	"	"	295	71

Table N-2. CYCLE TEST CONDITIONS AND SULPHUR REMOVAL EFFICIENCY (TEST 2-E)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range, u	Fuel Rate, g/min	Total Air, l/min	Superficial Gas Velocity, m/sec	% Stoich. Air	Bed Depth centimetres Start	Bed Depth centimetres End	Bed Specific Gravity	Average Bed Temp. (Absorption), °C	Max Bed Temp (Regen), °C	Duration of Absorption, min	Make-up Weight (Limestone), g/cycle	Cyclone Dust, g	SRE, %
38/72	1	Amuay (2.3%S)	Denbighshire	600-3175	184	700	1.86	34.9	48.0	39.1	0.83	880	980	40	-	577	100
39/72	2	"	"	"	203	643	1.71	29.1	47.2	40.9	0.84	850	1010	"	550	394	100
40/72	3	"	"	"	214	671	1.83	28.8	49.3	37.8	0.86	880	1010	"	"	296	90
41/72	4	"	"	"	221	633	1.74	26.3	45.7	40.6	0.85	890	1010	"	"	231	87
42/72	5	"	"	"	241	608	1.62	23.2	38.9	40.9	0.90	850	1015	"	"	177	77
43/72	6	"	"	"	241	645	1.74	24.6	41.4	40.6	0.94	860	1010	"	"	195	85
44/72	7	"	"	"	248	614	1.68	22.7	41.9	36.8	0.97	870	1020	"	"	160	76
45/72	8	"	"	"	238	664	1.80	25.6	41.4	37.3	0.88	870	970	"	"	180	75
46/72	9	"	"	"	243	633	1.71	23.9	44.5	35.8	0.92	870	1010	"	"	183	76
47/72	10	"	"	"	253	645	1.74	23.4	42.9	37.3	0.94	870	1020	"	"	150	70
48/72	11	"	"	"	254	629	1.68	22.7	42.2	38.3	0.95	850	1030	"	"	151	74
49/72	12	"	"	"	247	637	1.77	23.7	41.7	40.1	0.91	870	1030	"	"	190	77
50/72	13	"	"	"	258	633	1.71	22.5	43.2	40.4	0.92	850	1035	"	"	120	76
51/72	14	"	"	"	244	658	1.77	24.7	42.9	42.2	0.92	860	1025	"	"	158	78
52/72	15	"	"	"	213	661	1.80	28.5	55.4	49.0	0.74	875	930	"	"	160	74
53/72	16	"	"	"	238	643	1.71	24.8	56.1	55.4	0.72	845	980	"	"	160	79
54/72	17	"	"	"	238	616	1.65	23.8	59.2	52.8	0.73	850	1015	"	"	162	76
55/72	18	"	"	"	241	617	1.65	23.5	60.2	45.7	0.79	855	1020	"	"	160	77
56/72	19	"	"	"	227	605	1.65	24.5	42.7	42.7	0.92	870	1020	"	"	170	77
57/72	20	"	"	"	238	601	1.62	23.2	49.3	52.8	0.83	865	1030	"	"	155	75
58/72		"	"	"	233	601	1.62	23.7	-	-	-	865	1030	"	"	150	76

Table N-3. CYCLE TEST CONDITIONS AND SULPHUR REMOVAL EFFICIENCY (TEST 3-C)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range, μ	Fuel Rate, g/min	Total Air, l/min	Superficial Gas Velocity, m/sec	% Stoich. Air	Bed Depth centimetres Start	Bed Depth centimetres End	Bed Specific Gravity	Average Bed Temp. (Absorption) °C	Max Bed Temp (Regen), °C	Duration of Absorption, min	Make-up Weight Limestone, g/cycle	Cyclone Dust, g	SRE %
62/72	1	Amnay (2.3%S)	BCR 1359	600-3175	199	688	1.89	31.7	48.0	36.9	0.85	880	935	40	-	279	94
63/72	2	"	"	"	193	697	1.89	33.1	43.4	37.3	0.90	860	990	"	550	495	85
64/72	3	"	"	"	213	700	1.92	30.2	41.7	37.6	0.92	880	1020	"	"	361	82
65/72	4	"	"	"	227	691	1.89	27.9	40.6	36.3	0.95	880	1030	"	"	345	82
66/72	5	"	"	"	230	684	1.83	26.9	39.7	37.8	0.95	845	1030	"	"	215	75
67/72	6	"	"	"	233	682	1.86	26.9	39.9	34.3	0.98	880	1030	"	"	137	76
68/72	7	"	"	"	236	664	1.80	25.8	41.4	35.6	0.96	875	1035	"	"	138	71
69/72	8	"	"	"	227	664	1.80	26.8	39.1	36.8	1.00	875	1025	"	"	122	80
70/72	9	"	"	"	222	652	1.80	26.9	37.6	36.8	1.04	880	1025	"	"	130	80
71/72	10	"	"	"	210	659	1.80	28.8	36.3	35.6	1.08	880	-	"	"	125	77
72/72	11	"	"	"	219	660	1.77	27.8	44.5	41.7	0.92	845	990	"	"	76	69
73/72	12	"	"	"	184	650	1.74	32.4	45.7	42.7	0.92	860	995	"	"	133	76
74/72	13	"	"	"	179	637	1.74	32.7	47.0	43.4	0.92	870	1010	"	"	103	78
75/72	14	"	"	"	182	639	1.74	32.2	46.5	42.7	0.96	870	1010	"	"	86	77
76/72	15	"	"	"	196	650	1.71	30.4	43.7	40.6	1.00	840	1020	"	"	135	76
77/72	16	"	"	"	201	631	1.71	28.8	47.2	41.9	0.99	870	1020	"	"	140	73
78/72	17	"	"	"	196	635	1.74	29.7	41.9	42.7	1.00	875	1015	"	"	105	72
79/72	18	"	"	"	193	609	1.65	29.0	47.0	43.2	1.00	870	1020.	"	"	92	71
80/72	19	"	"	"	193	645	1.74	30.7	51.3	41.1	1.04	965	1020	"	"	85	75
81/72	20	"	"	"	179	663	1.80	34.0	-	-	-	870	990	"	"	125	75

Table N-4 CYCLIC TEST CONDITIONS AND S.R.E.'S (TEST 3-D)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range, μ	Fuel Rate, g/min	Total Air, l/min	Superficial Gas Velocity, m/sec	% Stoich. Air	Bed Depth centimetres Start	End	Bed Specific Gravity	Average Bed Temp. (Absorption), °C	Max Bed Temp (Regen), °C	Duration of Absorption, min	Make-up Weight (Limestone), g/cycle	Cyclone Dust, g	SRE, %
41/73	1	Amuay(2.3%S)	BCR 1359	600-3175	172	685	1.81	34.6	48	50	0.77	860	960	40	-	139	92
42/73	2	"	"	"	170	690	1.81	35.3	51	52	0.75	855	1020	"	750	115	92
43/73	3	"	"	"	172	695	1.84	34.7	55	50	0.78	855	1070	"	"	128	89
44/73	4	"	"	"	176	683	1.81	33.6	54	50	0.82	860	1060	"	"	143	83
45/73	5	"	"	"	179	685	1.81	33.4	54	49	0.83	860	1040	"	"	148	89
46/73	6	"	"	"	199	673	1.78	28.7	43	42	0.78	865	1040	"	"	151	84
47/73	7	"	"	"	211	674	1.78	28.1	49	45	0.83	860	1090	"	"	169	80
48/73	8	"	"	"	210	676	1.73	28.1	52	46	0.82	860	1030	"	"	172	80
49/73	9	"	"	"	207	708	1.88	29.6	49	47	0.75	850	985	"	"	157	82
50/73	10	"	"	"	204	692	1.84	29.6	50	45	0.78	855	1010	"	"	185	84
51/73	11	"	"	"	204	700	1.84	29.7	47	46	0.82	865	1020	"	"	234	84
52/73	12	"	"	"	216	690	1.84	27.9	46	46	0.83	865	1030	"	"	418	82
53/73	13	"	"	"	213	700	1.88	28.5	50	47	0.78	870	1030	"	"	497	85
54/73	14	"	"	"	225	700	1.88	27.0	44	44	0.83	875	1030	"	"	572	83
55/73	15	"	"	"	247	785	2.09	27.5	40	39	0.82	860	1025	"	"	130	75
56/73	16	"	"	"	254	680	1.78	23.2	45	41	0.78	855	1030	"	"	146	78
57/73	17	"	"	"	252	680	1.78	22.3	46	45	0.81	860	1040	"	"	180	70
58/73	18	"	"	"	243	690	1.84	24.7	48	45	0.83	860	1035	"	"	203	68
59/73	19	"	"	"	244	688	1.83	24.4	44	42	0.84	855	1035	"	"	232	79
60/73	20	"	"	"	281	685	1.82	21.0	44	-	0.80	860	1030	"	"	225	-[a]

[a] FUEL PUMP FAILED AFTER 25 MINS.

Table N-5. CYCLIC TEST CONDITIONS AND S.R.E.'S (TEST 3-E)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range, µ	Fuel Rate, g/min	Total Air, l/min	Superficial Gas Velocity, m/sec	% Stoich. Air	Bed Depth centimetres Start	Bed Depth centimetres End	Bed Specific Gravity	Average Bed Temp. (Absorption), °C	Max Bed Temp (Regen), °C	Duration of Absorption, min	Make-up Weight (Limestone), g/cycle	Cyclone Dust, g	SRE, %
61/73	1	Amuay (2.3%S)	BCR 1359	600-3175	-	-	-	-	-	-	-	-	-	-	-	-	-[a]
62/73	2	"	"	"	196	687	1.78	32.6	48	46	.78	830	960	40	950	242	96
63/73	3	"	"	"	164	679	1.81	38.3	50	48	.75	865	990	"	"	134	93
64/73	4	"	"	"	170	665	1.81	36.3	50	49	.78	870	1010	"	"	207	89
65/73	5	"	"	"	170	655	1.74	35.4	52	52	.78	865	1020	"	"	190	86
66/73	6	"	"	"	244	698	1.52	22.7	51	52	.83	860	1030	"	"	185	79
67/73	7	"	"	"	184	662	1.76	33.2	55	55	.78	860	1015	"	"	210	80
68/73	8	"	"	"	173	700	1.87	37.2	53	54	.83	870	1010	"	"	260	87
69/73	9	"	"	"	225	720	1.95	29.7	44	43	.77	880	1040	"	"	239	82
70/73	10	"	"	"	230	670	1.78	27.0	48	46	.79	860	1040	"	"	258	83
71/73	11	"	"	"	199	660	1.73	30.7	47	43	.77	855	1010	"	"	159	90
72/73	12	"	"	"	201	730	1.95	33.6	41	41	.87	870	1015	"	"	200	85
73/73	13	"	"	"	183	675	1.80	32.4	49	46	.81	865	1020	"	"	217	86
74/73	14	"	"	"	204	663	1.76	29.9	46	46	.83	860	1020	"	"	205	93
75/73	15	"	"	"	190	655	1.75	32.0	47	47	.83	870	1025	"	"	188	90
76/73	16	"	"	"	193	645	1.72	31.0	54	47	.83	865	1020	"	"	190	90
77/73	17	"	"	"	196	682	1.81	32.3	44	44	.83	860	1000	"	"	119	88
78/73	18	"	"	"	201	660	1.74	30.4	46	46	.83	850	1015	"	"	173	88
79/73	19	"	"	"	197	643	1.70	30.2	48	47	.92	853	1015	"	"	181	90
80/73	20	"	"	"	204	675	1.79	30.7	45	44	.90	860	1020	"	"	227	83
81/73	21	"	"	"	193	693	1.86	34.3	47	47	.89	870	1020	"	"	296	89

[a] FUEL PUMP FAILED AFTER 25 MINS.

Table N-6 Cycle Test Conditions and SREs (Test 4-C)

Run No.	Cycle No.	Fuel	Lime-Stone	Limestone Particle Size Range (μ)	Fuel Rate (g/min)	Total Air (l/min)	Superficial Gas Velocity (m/sec)	% Stoich. Air	Bed Depth (centimetres) Start	End	Bed Specific Gravity	Average Bed Temp (Absorption) (°C)	Max. Bed Temp (Regen.) (°C)	Duration of Absorption (min.)	Make-up Weight (limestone) (Kg/cycle)	Cyclone Dust (g)	SRE %
6/73	1	Amuay (2.3%S)	Pfizer	600-3175	210	792	2.24	34.6	48.0	47.2	0.71	880	975	40	-	677	95
7/73	2	"	Calcite	"	231	672	1.84	26.8	46.0	38.9	0.75	845	1015	"	550	296	92
8/73	3	"	"	"	204	669	1.84	30.1	43.9	36.6	0.83	845	1025	"	"	623	90
9/73	4	"	"	"	207	687	1.91	30.5	40.6	36.6	0.77	855	1035	"	"	1335	91
10/73	5	"	"	"	203	683	2.08	30.9	36.6	35.1	0.68	920	1040	"	"	1580	87
11/73	6	"	"	"	191	639	1.81	30.6	-	-	-	880	1040	"	"	731	84
12/73	7	"	"	"	210	680	1.84	29.7	38.9	32.3	0.71	875	1040	"	"	703	88
13/73	8	"	"	"	269	743	2.10	25.3	37.1	31.5	0.68	840	1055	"	"	662	81
14/73	9	"	"	"	259	662	1.81	23.6	51.6	46.5	0.76	835	1015	"	11000	2129	79
15/73	10	"	"	"	-	-	-	-	-	-	-	-	1010	"	550	-	80
16/73	11	"	"	"	255	698	1.90	25.1	53.8	38.1	0.83	835	1000	"	2000	1343	88
17/73	12	"	"	"	181	679	1.84	34.3	42.7	43.4	0.79	880	1020	"	750	976	90
18/73	13	"	"	"	191	683	1.94	33.0	44.2	38.6	0.79	880	1025	"	"	1185	91
19/73	14	"	"	"	193	696	1.97	33.1	38.1	35.8	0.71	880	1035	"	"	1533	88
20/73	15	"	"	"	216	692	1.96	29.5	30.5	29.0	0.79	875	1040	"	"	1616	88
21/73	16	"	"	"	213	689	1.95	29.7	33.8	30.5	-	885	1040	"	3000	1933	87
22/73	17	"	"	"	199	697	1.98	32.2	39.6	21.2	-	840	1040	"	"	2139	88
23/73	18	"	"	"	207	654	1.79	29.0	48.3	43.4	0.82	850	955	"	9000	2963	90
24/73	19	"	"	"	180	648	1.79	33.0	42.2	42.7	0.83	855	995	"	1000	491	95
25/73	20	"	"	"	181	639	1.77	32.3	48.3	45.5	0.79	860	1015	"	750	455	94
26/73	21	"	"	"	179	631	1.76	32.4	49.0	46.5	0.79	865	1030	"	"	409	97
27/73	22	"	"	"	179	655	1.83	33.7	45.2	44.2	0.83	885	1030	"	"	385	97
28/73	23	"	"	"	180	660	1.88	33.6	47.2	41.7	0.92	890	1030	"	"	547	91
29/73	24	"	"	"	183	669	1.91	33.6	46.2	44.2	0.83	885	1030	"	"	531	90
30/73	25	"	"	"	190	674	1.91	32.5	47.0	46.0	0.81	880	1035	"	"	547	89
31/73	26	"	"	"	179	647	1.83	33.1	44.2	41.1	0.83	870	1010	"	2000	563	89
32/73	27	"	"	"	179	621	1.74	32.4	40.1	37.3	0.92	865	1020	"	750	234	90
33/73	28	"	"	"	179	629	1.76	32.8	42.4	42.2	0.88	880	1015	"	"	466	90
34/73	29	"	"	"	179	680	1.93	35.0	42.4	43.4	0.88	885	1025	"	"	542	89
35/73	30	"	"	"	181	659	1.88	33.3	42.9	41.7	0.92	885	1025	"	"	506	89
36/73	31	"	"	"	201	648	1.84	29.6	49.3	41.7	0.92	880	1030	"	"	468	88
37/73	32	"	"	"	199	665	1.88	30.7	49.3	41.7	0.92	880	1030	"	"	519	88
38/73	33	"	"	"	196	641	1.82	30.0	48.5	39.1	0.96	880	1025	"	"	-	88

Table N-7 CYCLIC TEST CONDITIONS AND S.R.E.'S (TEST 5-D)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range, µ	Fuel Rate, g/min	Total Air, l/min	Superficial Gas Velocity, m/sec	% Stoich. Air	Bed Depth centimetres Start	Bed Depth centimetres End	Bed Specific Gravity	Average Bed Temp. (Absorption), °C	Max Bed Temp (Regen), °C	Duration of Absorption, min	Make-up Weight (Limestone), g/cycle	Cyclone Dust, g	SRE, %
82/73	1	AMUAY VAC RESID BOTTOM	BCR 1359	600-3175	198	700	1.95	31.3	43	41	0.83	860	1050	35	-	119	98
83/73	2	"	"	"	194	700	1.96	31.8	41	41	0.83	865	1020	"	550	140	93
84/73	3	"	"	"	227	700	1.98	27.2	41	41	0.83	875	1060	"	"	149	83
85/73	4	"	"	"	240	670	1.90	24.7	39	39	0.87	880	1070	"	"	177	80
86/73	5	"	"	"	220	683	1.91	27.5	47	44	0.81	855	1030	"	"	181	87
87/73	6	"	"	"	220	709	2.00	28.5	40	40	0.86	870	1045	"	"	120	77
88/73	7	"	"	"	240	683	1.93	25.2	34	35	0.93	870	1080	"	"	-	74
89/73	8	"	"	"	-	670	-	-	50	48	0.79	875	-	"	"	107	-
90/73	9	"	"	"	230	671	1.90	25.8	46	47	0.80	865	1040	"	"	104	78
91/73	10	"	"	"	250	660	1.88	23.7	50	47	0.83	875	1030	"	"	133	80
92/73	11	"	"	"	233	648	1.85	24.6	44	46	0.87	875	1050	"	"	141	56
93/73	12	"	"	"	272	698	1.99	22.7	46	47	0.83	875	1040	"	"	183	58
94/73	13	"	"	"	194	680	1.89	30.9	46	45	0.83	850	1030	"	"	79	80
95/73	14	"	"	"	185	704	1.98	33.7	46	46	0.83	860	1020	"	"	94	80
96/73	15	"	"	"	201	656	1.85	28.9	46	47	0.83	865	1040	"	"	79	70
97/73	16	"	"	"	207	608	1.72	26.0	44	46	0.87	863	1040	"	"	69	69
98/73	17	"	"	"	207	574	1.60	29.5	44	40	0.92	850	1065	"	"	70	60
99/73	18	"	"	"	217	661	1.88	26.9	41	42	0.87	875	1060	"	"	155	69
100/73	19	"	"	"	269	676	1.88	22.2	41	41	0.86	850	1070	"	"	146	62

Table N-8 CYCLIC TEST CONDITIONS FOR HIGH SULPHUR PITCH (TEST 6-C)

Run No.	Cycle No.	Stone	Stone Size Range Microns	Fuel Rate g/min	Total Air l/min	Superficial Gas Vel. ft/sec	% Stoic Air	Bed Pressure Drop (inches H_2O) Start	End	Average Bed Temp (Absorption) °C	Max Bed Temp (Regen) °C	Duration of Absorption Mins	Make-up Weight (Limestone) g/cycle	SRE %
130/73	1	BCR 1359	600-3175	208	490	1.34	23.3	42	43	870	1030	35	-	74.5
131/73	2	"	"	201	548	1.49	27.0	43	43	870	1040	"	1030	70.9
132/73	3	"	"	175	575	1.55	32.0	43	42	900	1035	"	1300	84.5
133/73	4	"	"	156	524	1.43	32.9	46	46	880	1040	"	"	79.4
134/73	5	"	"	136	513	1.40	36.9	46	47	900	1020	"	"	82.5
135/73	6	"	"	156	517	1.40	32.5	48	47	890	1040	"	"	76.8
136/73	7	"	"	156	528	1.43	33.1	47	47	910	1030	"	"	76.8
137/73	8	"	"	157	510	1.40	31.8	48	50	900	1030	"	"	81.0
138/73	9	"	"	182	520	1.43	27.9	43	41	890	1050	"	"	75.0
139/73	10	"	"	169	521	1.43	30.2	43	43	880	1030	"	"	80.0
140/73	11	"	"	182	485	1.34	26.0	44	44	900	1050	"	"	69.5
141/73	12	"	"	188	491	1.34	25.5	46	48	880	1050	"	"	71.0
142/73	13	"	"	156	553	1.52	34.7	46	46	895	1050	"	"	82.0
143/73	14	"	"	149	530	1.46	34.8	46	46	900	1010	"	"	84.0
144/73	15	"	"	157	520	1.43	32.3	48	48	900	1040	"	"	79.0
145/73	16	"	"	170	510	1.40	29.4	47	47	910	1030	"	"	76.0
146/73	17	"	"	162	525	1.43	31.6	48	48	915	1030	"	"	82.0
147/73	18	"	"	149	533	1.46	35.0	47	46	900	1030	"	"	84.0
148/73	19	"	"	135	533	1.46	38.6	46	46	900	1030	"	"	83.0
149/73	20	"	"	158	539	1.46	33.4	46	46	910	1025	"	"	82.5

Table N-9 KEROSENE COMBUSTION (TESTS 1-A, 2-A, 3-A,4-A

Stone:	BCR 1691				Denbighshire				BCR 1359				Pfizer Calcite			
Time, hr	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C
0	.59	0.75	-	875	.53	0.75	-	870	.58	0.75	-	850	0.51	0.75	-	870
½	.47	0.79	170	900	.49	0.79	280	870	.55	0.79	90	842	0.50	0.79	303	865
1	.40	0.79	920	880	.50	0.75	135	865	.52	0.83	70	860	0.48	0.79	210	820
1½	.39	0.75	1005	855	.49	0.75	125	870	.53	0.79	60	860	0.48	0.73	268	865
2	.37	0.77	390	890	.47	0.79	80	870	.51	0.83	50	865	0.43	0.79	221	890
2½	.32	0.77	420	880	.46	0.77	85	885	.51	0.83	87	865	0.42	0.78	185	859
3	.31	0.77	290	893	.45	0.79	35	890	.49	0.83	55	872	0.42	0.79	90	853
3½	.30	0.77	280	900	.46	0.79	40	880	.49	0.88	40	870	0.43	0.75	96	845
4	-	-	-	-	.44	0.82	30	860	.49	0.86	30	880	0.40	0.79	143	880
4½	-	-	-	-	-	-	-	-	.49	0.83	30	880	0.40	0.79	90	855
5	-	-	-	-	-	-	-	-	.51	0.82	27	860	0.39	0.79	91	859
5½	-	-	-	-	-	-	-	-	.52	0.80	23	860				
6	-	-	-	-	-	-	-	-	.50	0.82	35	860				

Table N-10 KEROSENE COMBUSTION (TESTS 1-B, 2-B)

Stone:	BCR 1691				Denbighshire			
Time, hr	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C
0	.55	0.73	0	1020	.50	0.83	-	1010
½	.52	0.73	317	1080	.45	0.90	270	1015
1	.55	0.71	55	1065	.44	0.91	90	1080
1½	5.1	0.75	235	1078	.39	1.00	27	1050
2	.49	0.75	380	1080	-	-	-	-
2½	.46	0.75	130	1070	.42	0.91	47	1080
3	.47	0.76	80	1058	.39	0.98	18	1080
3½	.46	0.76	75	1054	.39	1.00	20	1060
4	.47	0.75	75	1078	.39	1.00	14	1050
4½	.45	0.77	60	1090	.38	1.01	12	1050
5	.45	0.80	50	1075	.39	1.00	11	1050
5½	.44	0.82	50	1065	.39	1.00	10	1060
6	.43	0.80	30	1065	.39	1.00	6	1080

Table N-11. FUEL OIL COMBUSTION (TESTS 1-C, 2-C)

Stone:	BCR 1691					Denbighshire					Pfizer Calcite				
Time, hr	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Fuel Rate, g/min	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Fuel Rate g/min	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Fuel Rate g/min
0	.62	0.67	-	870	-	.58	0.75	-	850	-	0.51	0.75	-	895	-
½	.58	0.67	180	810	34	.54	0.75	160	850	27	0.50	0.71	318	836	41.5
1	.52	0.71	155	870	47	.50	0.79	260	850	23	0.47	0.75	173	858	45.4
1½	.45	0.79	113	875	44	.48	0.79	310	868	27	0.47	0.75	166	876	49.0
2	.46	0.75	170	870	49	.46	0.83	100	868	23	0.43	0.79	126	875	41.5
2½	.45	0.75	70	870	49	.45	0.85	75	872	23	0.44	0.78	92	848	41.5
3	.45	0.75	70	880	47	.43	0.86	60	878	25	0.42	0.78	93	846	41.5
3½	.42	0.79	57	860	48	.45	0.83	85	878	25	0.42	0.78	58	-	37.8
4	.41	0.79	23	850	48	.43	0.85	68	965	25	0.42	0.79	65	-	39.7
4½	.42	0.79	35	840	40	.44	0.83	62	890	25	0.42	0.79	47	-	39.7
5	.42	0.80	35	840	38	.40	0.91	45	850	23	0.42	0.79	58	-	39.7
5½	.40	0.83	45	850	38	.40	0.91	67	860	23	0.42	0.79	44	-	-
6	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

Table N-12. FUEL OIL GASIFICATION (TESTS 1-D, 2-D, 3-B)

Stone:	BCR 1691					Denbighshire					BCR 1359				
Time, hr	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Fuel Rate, g/min	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Fuel Rate g/min	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Fuel Rate g/min
0	0.55	0.79	-	868	-	0.58	0.75	-	900	-	0.56	0.75	-	920	-
½	0.56	0.75	40	860	170	0.49	0.82	490	850	178	0.52	0.79	180	835	173
1	0.49	0.83	395	860	178	0.43	0.87	290	860	204	0.52	0.79	127	860	180
1½	0.50	0.82	135	850	170	0.41	0.92	150	850	178	0.49	0.83	90	853	192
2	0.52	0.82	95	860	197	0.41	0.92	60	848	208	0.48	0.88	90	868	204
2½	0.50	0.79	10	855	200	0.41	0.92	130	862	206	0.49	0.83	60	860	204
3	0.48	0.82	100	855	204	0.41	0.92	40	862	211	0.42	0.83	67	868	214
3½	0.48	0.83	70	855	211	0.90	0.92	75	850	221	0.46	0.88	62	870	221
4	0.49	0.83	70	850	239	0.48	0.92	33	840	242	0.47	0.88	60	860	221
4½	0.49	0.83	86	862	221	0.41	0.92	60	850	238	0.46	0.89	60	860	242
5	0.46	0.87	75	870	231	0.40	0.96	50	860	224	0.50	0.92	50	860	234
5½	0.46	0.87	70	870	185	0.39	0.96	60	865	221	0.47	0.96	52	860	234
6	0.45	0.87	55	867	216	0.40	0.96	35	870	212	0.47	0.96	-	860	234

Table N-13. KEROSENE COMBUSTION ON CYCLED BED (TESTS 1-F, 2-F)

Stone:	BCR 1691				Denbighshire			
Time, hr	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C	Bed Depth, Metres	Bed Specific Gravity	Cyclone Dust, g	Bed Temp., °C
0	0.58	0.75	-	900	0.47	0.96	-	850
½	0.51	0.83	240	880	0.46	1.0	18	830
1	0.49	0.83	175	815	0.45	1.0	10	843
1½	0.48	0.83	170	830	0.45	1.0	12	853
2	0.46	0.83	280	800	0.45	1.02	8	860
2½	0.45	0.83	100	800	0.45	1.0	12	870
3	0.45	0.83	40	850	0.46	1.0	10	872
3½	0.44	0.82	230	820	0.46	1.0	10	862
4	0.44	0.82	100	770	0.44	1.03	8	865
4½	0.42	0.83	40	850	0.43	1.04	5	868
5	0.41	0.83	40	850	0.44	1.04	8	875
5½	0.40	0.83	100	850	0.43	1.04	7	880
6	0.40	0.79	140	870	0.42	1.04	7	880

Table N-14. **GASIFICATION OF HEAVY RESIDUAL FUEL OILS FRESH BED TEST RESULTS (TESTS 5-A, 5-B, 5-D)**

Run:	Run No 101/73 Amuay Vacuum Bottoms					Run No 102/73 Amuay Residual Fuel Oil					Run No 104/73 Amuay Vacuum Bottoms with 24.5% O_2				
Time, Hrs	Totals % wt. Bed	Carbon % wt. Bed	Fuel/Air Ratio % Stoich	Bed Temp., °C	Cyclone Dust, g	Total Sulphur %wt Bed	Carbon % wt. Bed	Fuel/Air Ratio % Stoich	Bed Temp., °C	Cyclone Dust, g	Total Sulphur %wt Bed	Carbon % wt Bed	Fuel/Air Ratio % Stoich	Bed Temp., °C	Cyclone Dust, g
0	-	-	23	820	-	-	-	27	880	-	-	-	26.7	900	-
½	-	-	"	-	-	-	-	"	-	138	-	-	"	-	213
1	2.25	2.85	"	860	150	2.30	1.00	"	860	96	3.36	0.97	"	900	119
1½	-	-	"	-	100	-	-	"	-	112	-	-	"	-	223
2	4.76	11.42	"	850	103	4.31	0.30	"	880	114	5.73	7.18	"	920	-
2½	-	-	"	-	-	-	-	"	-	124	-	-	"	-	296
3	5.83	15.7	"	875	257	6.53	1.58	"	860	90	7.83	12.4	"	940	263
3½	-	-	"	-	-	-	-	"	-	136	7.89	11.3	"	-	220
4	7.62	20.7	"	860	284	8.14	2.27	"	860	68	-	-	"	-	-

Table N-15. GASIFICATION OF HEAVY RESIDUAL FUEL OILS FRESH BED TEST RESULTS (TESTS 5-C, 6-A, 6-B)

Run:	Run No 107/73 Amuay Vacuum Bottoms with 25% O_2					Run No 108/73 High Sulphur Pitch					Run No 109/73 High Sulphur Pitch				
Time, Hrs	Total Sulphur %wt Bed	Carbon % wt. Bed	Fuel/Air Ratio % Stoich	Bed Temp., °C	Cyclone Dust, g	Total Sulphur %wt Bed	Carbon % wt. Bed	Fuel/Air Ratio % Stoich	Bed Temp., °C	Cyclone Dust, g	Total Sulphur %wt Bed	Carbon % wt. Bed	Fuel/Air Ratio % stoich	Bed Temp., °C	Cyclone Dust, g
0	-	-	28.1	940	-	-	-	24.5	860	-	-	-	31.6	-	-
½	-	-	"	-	526	-	-	"	-	88	-	-	"	890	120
1	1.94	0.23	"	940	241	3.59	12.39	"	890	70	2.91	8.31	"	860	150
1½	-	-	"	-	300	-	-	"	-	130	4.73	11.76	"	880	75
2	3.76	0.24	"	950	265	4.96	17.82	"	905	140	5.82	13.02	"	900	175
2½	-	-	"	-	311	6.29	-	"	910	215					
3	5.82	1.62	"	950	288	6.90	-	"	910	115					
3½	-	-	"	-	281										
4	8.02	3.27	"	940	272										

APPENDIX O

OIL AND LIMESTONE

OIL

The oil used in the present work have been heavy fuel oils from Venezuelan crudes obtained from Amuay refinery of Creole Petroleum Co. One supply, obtained in drums direct from Amuay has been used in batch unit tests. A second supply from bulk storage in the U.K. has been used in pilot plant work. An additional supply of very heavy residue has been obtained for batch unit studies which have not yet begun. This oil, also from Amuay, is vacuum pipe still bottoms produced when ordinary atmospheric pipe still bottoms is further distilled to give a vacuum gas oil which can be hydrofined to give a reduced sulphur fuel oil. Chemical and physical inspection results of these three oils are listed in Table O-1.

LIMESTONE

Pilot plant runs have employed U.S. limestone BCR 1691 and a U.K. limestone from Denbighshire. Batch unit work has employed these two stones as well as U.S. stone BCR 1359. Two additional U.S. stones have been recieved. They are Tymochtee Dolomite and a limetstone selected by New England Electric System: Pfizer Calcite. This was selected because of its proximity to the proposed demonstration unit. Chemical inspections of all the stones are listed in Table O-2.

Additional analyses on fuel oil and limestone used in Runs 6 and 7 is given in Appendix J, Tables I and II.

Table O-1

Properties of
CAFB Test Fuel Oils

Property		Amuay Batch Unit Tests	Amuay Pilot Plant Tests			Vacuum Bottoms	High Sulphur Pitch
			1/12/71	14/3/73	26/3/73		
Specific Gravity		0.957	0.955	0.960	0.960	1.015	1.101
Kinematic Viscosity							
CS at 140 deg. F		201	221	-	254	-	-
210 deg. F		41.4	39.3	40.3	45.6	3180	-
280 deg. F		-	-	-	-	321	-
350 deg. F		-	-	-	-	71	-
400 deg. G		-	-	-	-	240	-
Carbon	% by wt.	85.9	85.6	85.8	85.3	85.7	84.5
Hydrogen	"	11.3	11.4	11.6	11.3	10.3	8.7
Sulphur	"	2.35	2.48	2.42	2.43	2.95	5.4
Nitrogen	"	0.35	0.26	0.09	0.35	0.63	0.64
Conradson Carbon	"	11.6	10.9	11.1	10.8	17.4	33.0
Asphaltenes	"	7.1	4.8	5.3	6.0	6.9	15.0
Vanadium	ppm	366	345	300	315	530	155
Nickel	"	43	40	65	41	6	65
Sodium	"	36	35	37	38	9	17
Iron	"	3	4	-	3	108	15

Table O-2

Chemical Properties of Test Limestones

Limestone Composition

Stone:		BCR 1691	BCR 1359	Denbighshire	Tymochtee Dolomite	Pfizer Calcite
Component						
CaO	% by wt.	45.6	54.1	55.2	31.1	55.0
MgO	"	3.35	0.60	0.30	20.9	0.6
SiO_2	"	13.65	0.75	0.68	3.1	0.8
Fe_2O_3	"	0.35	0.09	0.10	0.4	0.09
Al_2O_3	"	2.80	0.31	0.25	1.13	0.3
CO_2	"	35.7	44.0	43.4	43.6	43.2
S (Total)	"	0.44	0.12	< 0.05	0.13	0.03
Vanadium	ppm	41	50	26	< 25	< 20
Sodium	"	219	< 20	59	175	250
Nickel	"	40	30	21	10	50

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-650/2-74-109	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil--Second Phase		5. REPORT DATE November 1974
		6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) J. W. T. Craig, G. L. Johnes, Z. Kowszun, G. Moss, J. H. Taylor, and D. E. Tisdall		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Esso Research Centre, Abingdon, Berkshire, England		10. PROGRAM ELEMENT NO. 1AB013; ROAP 21ADD-BE
		11. CONTRACT/GRANT NO. 68-02-0300
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development, NERC-RTP, Control Systems Laboratory, Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Second Phase July72-May74
		14. SPONSORING AGENCY CODE

15. SUPPLEMENTARY NOTES

16. ABSTRACT The report describes the second phase of studies on the CAFB process for desulfurizing gasification of heavy fuel oil in a bed of hot lime. The first continuous pilot plant test with U.S. limestone BCR 1691 experienced local stone sintering and severe production of sticky dust during startup. Batch tests confirmed that BCR 1691 produced more dust than the purer Denbighshire or U.S. BCR 1359 stones. With BCR 1691, 10 times more dust was produced during kerosene combustion at 870C than during gasification/regeneration. The continuous pilot plant was modified to improve operability under dusty conditions: 332 gasification hours were spent in a second run with Denbighshire and BCR 1691 stones in six operating periods, the longest being 109 hours. Sulfur removal efficiency was comparable for the two stones, ranging from 60 to 95%. Regenerator performance was less satisfactory than in earlier tests. A poor sulfur material balance indicates need for improved analytical procedures. Total CAFB development through a large demonstration test will probably take about 6-7 years and require $3,320,000 in engineering effort.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution	Calcium Oxides	Air Pollution Control	13B
Regeneration (Engineering)	Limestone	Stationary Sources	08G
Sulfur		CAFB Process	07B
Desulfurization		Chemically Active Fluidized Bed	07A, 07D
Gasification		Fluidized Lime Bed	13H
Fuel Oil		Heavy Fuel Oil	21D

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