United States
Environmental Protection
Agency

Industrial Environmental Research
Laboratory
Research Triangle Park NC 27711

EPA-600/7-79-066
February 1979

EPA

Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil - Fourth Phase

Interagency Energy/Environment R&D Program Report

RESEARCH REPORTING SERIES

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

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

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

EPA REVIEW NOTICE

EPA-600/7-79-066

February 1979

Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil - Fourth Phase

by

A.W. Ramsden and Z. Kowszun

Esso Research Centre
Abingdon, Oxfordshire OX136AE
England

Contract No. 68-02-1479
Program Element No. 1AB013
ROAP 21ADD-BE

EPA Project Officer: Samuel L. Rakes

Industrial Environmental Research Laboratory
Office of Energy, Minerals, and Industry
Research Triangle Park, NC 27711

Prepared for

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

FOREWORD

The Chemically Active Fluidised Bed (CAFB) is a process for gasification and desulphurisation of high sulphur residual fuel oils and coals to produce a low BTU, low sulphur gas for utilisation in conventional combustion equipment. The process presents an alternative means for utilising high sulphur fuels in an environmentally acceptable fashion to meet existing limits for emission of sulphur oxides.

This report presents results of studies in a continuous pilot plant gasifier - regenerator system on the gasification of heavy fuel oil and a heavy vacuum residuum. Results are presented on desulphurisation performance and the effects of process variables on sulphur retention. Preliminary studies on the gasification of coal in a batch reactor and also discussed.

SUMMARY

Phase four of the studies on the CAFB process for gasifying and desulphurising liquid and other potential fuels - including coal - was carried out between June 1975 and December 1976. Objectives originally established for phase four were changed during this period in recognition of the improved means of data handling and evaluation established during phase three (Ref. 3). Essentially, the statistical techniques established enabled each valid hourly data set to be included, thereby avoiding the need to reject data taken under non-lined out operations. The number of data points available for analysis increased considerably as a consequence. Thus, the work planned for alternative fuel feedstocks, particularly coal, could be advanced whilst still meeting the objectives of supporting the design and construction of a large scale demonstration using heavy fuel oil.

Bearing in mind the change in emphasis of the programme mentioned above, the specific objectives of phase four were to evaluate one new limestone and one new fuel, (coal), in the CAFB batch reactor. Secondly, to extend understanding of the CAFB process through analysis of data derived from operation of the continuous CAFB gasifier on conventional fuel, and also on a heavy residuum fuel and coal, and to develop empirical relationships to describe the sulphur removal efficiency and to support processing modelling studies. A further major objective was to support the planned field demonstration programme at San Benito, Texas through evaluation of design, materials, and procedures proposed for this project. Provision was allowed for participation in consultations and discussions with other parties associated with this programme. Finally, ongoing tests were required to establish the fate of trace elements present in the fuel feedstocks used in the CAFB process.

Batch unit studies concluded that a Texas limestone appears to be suitable for the field demonstration unit, and that the gasification and desulphurisation of coal is feasible.

On the continuous CAFB pilot plant, run 10 was the initial test conducted with redesigned and rebuilt equipment incorporating a number of new features, including specific items to be tested in support of the field demonstration.

Essentially, cylindrical gasifier and regenerator reactors were cast in a composite refractory shell, providing

improved insulation, within a steel casing. A two stage gasifier air distributor was fitted with provisions to protect fuel injectors. New high flow, low velocity cyclones were fabricated and fitted external to the reactor vessels, and new fines re-injection equipment was installed. Provisions was made to evaluate a heavy vacuum residuum fuel, and to establish the feasibility of pneumatic injection of coal into the gasifier. For flue gas recycle, equipment was installed to examine tuyere injection, and to identify potential problems of a baghouse filter for flue gas clean up. Numerous detailed changes were made to other sub-systems associated with the pilot unit.

During Run 10, experience with the redesigned gasifier, regenerator, gasifier distributor, product gas cyclones and ducting was more than satisfactory, and the improved insulation greatly reduced the unit skin temperature. However, the cyclone drain/fines reinjection system, the limestone feed equipment, the flue gas recycle bag house filter and to a lesser extent, the bed transfer system proved troublesome, and a prolonged shut down resulted early in the run to enable improvements to be effected.

A number of specific experiments were carried out. The injection of heavy fuel oil through a single injector into the gasifier distributor pit was successfully demonstrated without loss of desulphurisation performance, as was the removal and insertion of fuel injectors without need for plant shut down. The pit refractory walls provided excellent protection for the fuel oil injectors. Heavy vacuum residuum was similarly injected successfully. Pneumatic coal injection was demonstrated and gasification, desulphurisation, and regeneration was not obviously impaired during this short period of operation except when metering problems were encountered with the simple equipment being used.

Flue gas recycle using tuyere injection directly into the gasifier bed proved successful and reduces the degree of recycle gas clean-up required. A baghouse filter system for flue gas clean up was subject to blockage with damp solids which were difficult to drain, and a warm-up stage needs to be considered as part of the operating procedure for such a facility.

Burn out of the product gas ducts from the boiler burner end was successful, but it proved impossible to clear the cyclone entries completely with a reversed gas flow. Air bleeds at the product gas entry point into the ducts leading from the gasifier were completely ineffective in preventing deposit accumulations.

Steam injection was confirmed to hinder the sulphur retention performance of the lime bed in the gasifier, the effect being more pronounced when richer operation of the gasifier was established.

Run 10 was hampered towards the end by excessive boiler water temperature, suggesting either restricted circulation in the primary circuit, or poor performance from the heat exchanger. These were not resolved at this time.

The regression analysis techniques developed during phase three (Ref. 3) for predicting sulphur removal efficiency were extended to the data available from Run 10. The similarity of the equation developed for Run 10 to those previously obtained indicated that the unit configuration, the location and number of fuel injectors, the two stage gasifier distributor and the tuyere injection of flue gas had no significant effect on the sulphur retention performance. A new significant variable, viz bed age, was found as a consequence of prolonged operations without fresh limestone make-up. The equation was successfully applied to TJ 102 Medium vacuum residuum (bitumen) suggesting that fuel characteristics are also of second order importance.

A unified equation covering Runs 8, 9 and 10 was developed as the best available performance predictor. This differed slightly from that presented previously (Ref. 3) as the results from each separate run were initially corrected to the design conditions planned for the demonstration unit. Application to available, selected data from Runs 6 and 7 gave much improved prediction compared to previously.

Retention of a range of trace elements by the lime bed was shown to be a relatively short term effect. Continuous stone make-up is required for optimum retention, though make-up rates lower than 1 molar are adequate provided the necessary bed level is maintained. Prolonged operation without fresh stone addition leads to a deterioration in trace element retention.

This study is generally difficult to carry out due to the difficulties of measuring trace element levels, particularly for the fuel and limestone feeds. Thus, some trace element levels are too low for detection, and the precision of measuring others is poor at ± 3 times the level detected. Balances are thus difficult to carry out.

Vanadium, which is of major interest since its concentration on the limestone may justify recovery, was found to reach a maximum of 0.6 wt% on the gasifier bed.

The investigation of emissions from the CAFB pilot plant, including trace element levels on particulates was carried out during Run 10 by GCA Corporation who were contracted by the EPA to conduct a Level 1 environmental assessment of the process. A separate report (Ref. 8) has been issued which comprehensively covers all aspects of fugitive emissions.

Future work planned will be directed towards further tests on the gasification and desulphurisation of solid fuels, viz coal and lignite. To support this, an efficient and reliable supply and metering system will need to be designed. Included in these studies will be further investigations of proposed design features and operating techniques for field demonstration plant support. Outside these objectives, it will be necessary to make a number of improvements in the current plant configuration to improve performance and data precision. Specifically, the limestone feed system will need improvement, analytical equipment will need updating and the performance of the boiler heat dissipation equipment will have to be resolved and corrected.

CONTENTS

LIST OF FIGURES

LIST OF TABLES

LIST OF TABLES (Continued)

ACKNOWLEDGEMENT

The support of Esso Petroleum Company Ltd. is acknowledged for the provision of a 2930 kW (10 million BTU/hr) Chemically Active Fluid Bed Gasifier facility at the Esso Research Centre, Abingdon, Oxfordshire, England, and for various additions and modifications to this facility which made possible the generation of continuous gasification data for this project.

The authors also thank Dr. G.L. Johnes, Dr. G. Moss, Dr. D. Lyon, Dr. L. Malkin, Mr. M. Alphandary, Mr. A. Brimble, Mr. E. Blissett, Mr. R. Priestnall, Mr. D. Stormes, Mr. A. Jennings, Mr. D. Stone, Mr. H. Towers and Mr. F. Rolf for their invaluable contributions, suggestions and assistance in operating equipment and recording experimental data on which this report is based. Many other Esso Petroleum Company personel are also gratefully acknowledged for assistance in designing, constructing and maintaining equipment, in operation of the pilot plant, and in chemical analysis of test samples.

Finally, the authors acknowledge the valuable discussions and exchange of ideas with the EPA Project Officer, Mr. S.L. Rakes.

INTRODUCTION

General

The Chemically Active Fluid Bed (CAFB) process is a means of reducing 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 re-use. 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 process began at the Esso Research Centre, Abingdon (ERCA) in 1966. In 1969 a six phase programme of work was proposed to take the CAFB process from the laboratory stage through to a demonstration of the process on a 50 to 100 megawatt (electrical) power generation boiler located in the United States. A summary of this six phase programme is shown in Fig. 1. Phase One 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 (Ref. 1) for that contract, dated June 1972.

Work on the second phase of studies was carried out in the period July 1, 1972 through May 1974, and the final report was issued in November 1974 (Ref. 2).

Work on phase three studies carried out under contract 68-02-1359 between November 1973 and June 1975 was covered in report No. EPA-600/2-76-248 issued in September 1976 (Ref. 3).

This report covers work on phase four studies under contract number 68-02-1479 between July 1975 and December 1976.

Gasifier Chemistry

When heavy fuel oil is injected into a bed of fluidised lime under reducing conditions at about 900 °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

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

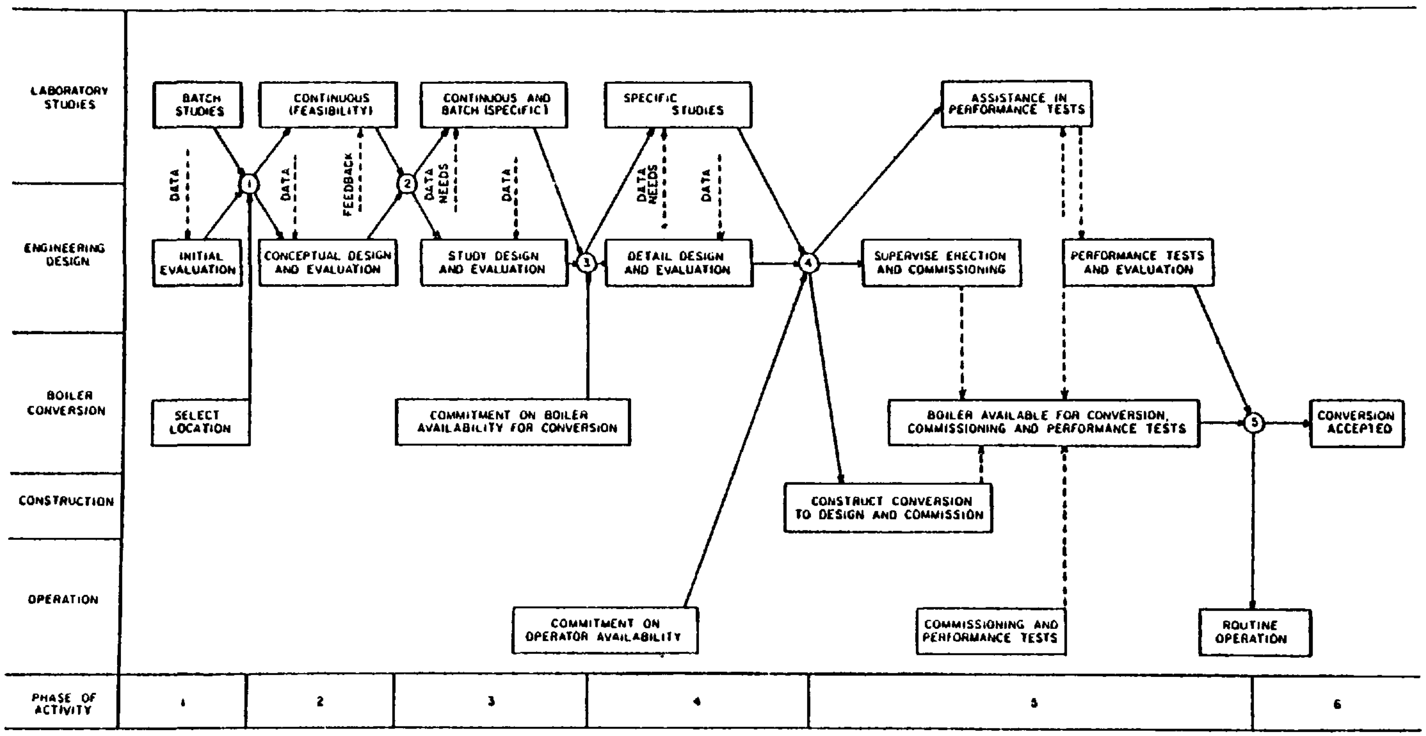


FIG. 1

with H_2S predominating. The sulphur compounds react with CaO to form CaS and water or carbon oxides.

For example:-

$$CaO + H_2S \longrightarrow CaS + H_2$$

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 °C. Other factors however limit gasification temperature to the range of about 850 to 950°C where the equilibrium sulphur removal would be about 99% (see Ref. 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 950°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 bea 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. Previous experimental evidence was that practically all of the fuel vanadium can remain fixed with the lime. This report contains data which indicates the limits within which retention can take place.

Regenerator Chemistry

Calcium sulphide is regenerated to calcium oxide by air oxidation.

$$CaS + 3/2\ O_2 \longrightarrow CaO + SO_2 \qquad H = -485.1\ \text{kJ/mol}$$

A competing reaction also consumes oxygen and forms calcium sulphate.

$$CaS + 2O_2 \longrightarrow CaSO_4 \qquad H = -921.3\ \text{kJ/mol}$$

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

$$CaS + 3CaSO_4 \longrightarrow 4CaO + 4SO_2 \qquad H = 926.8\ \text{kJ/mol}$$

This reaction is strongly endothermic.

The equilibrium constants, (Ref. 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°C.

During the conversion of CaS to CaO and $CaSO_4$ there is evidence for the existence of a transient liquid state (Ref. 4). 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.

Background Influences on Experimental Studies

Details of the results of previous studies are described fully in previous reports (Refs. 1, 2 and 3), but in summary this work had confirmed that good desulphurisation results could be obtained and that the process was stable and controllable. Sufficient information was generated by these studies to enable EPA to request Westinghouse Research Laboratories (WRL), under contract to EPA, to carry out a preliminary technical and cost review and to recommend on future development. A report by WRL (Ref. 5) confirmed that the CAFB process was attractive when compared with the main alternatives for clean power generation viz. flue gas desulphurisation or combustion of heavy fuel oil desulphurised at a refinery.

Acting on behalf of EPA, WRL presented the CAFB process to a number of utilities, with the objective of interesting one of them in participating under EPA support in conversion to CAFB operation of a power generation boiler of about 50 MWe capacity. Early in 1973 New England Electric System (NEES) of Westborough, Mass. agreed in principle to cooperate in such a demonstration to be based on a 50 MWe boiler at Providence, Rhode Island. EPA, NEES and WRL jointly selected Stone and Webster Engineering Corportion (SWEC) as the Architect Engineer for the conversion. Details of the SWEC design and costing have been reported by WRL (Ref. 5), but events were overtaken by the decision of the Federal Energy Authority to schedule NEES boilers for operation on coal only. In view of this, NEES withdrew from the demonstration programme in April 1975.

Fortunately for the progressive continuation of the CAFB development programme, Foster Wheeler Energy Corporation (FWEC) had been pursuing an independent programme as a worldwide licensee of patents held on the CAFB process by Exxon Research and Engineering Company (EREC) and drew this to the attention of EPA in April 1975. Agreement was readily reached between EPA and FWEC to fund engineering studies, and FWEC in turn signed an agreement with Central Power and Light (CP & L) of Corpus Christi, Texas, to convert a 20 MWe boiler in the CP & L plant at San Benito, Texas. A CAFB unit is currently under construction to demonstrate performance at this location.

As set out in Fig. 1, Phase four was planned as a period of specific studies to support detailed design and evaluation of the CAFB process. However, Foster Wheeler Energy Corporation moved rapidly during the period covered by this report and a definitive design and report had been issued for the San Benito demonstration project by April 1976. The design was based essentially on work previously reported (Refs. 1, 2 and 3), and on partial analysis of the data derived during run 10 (Task 3). Consequently, the direction of the work under contract 68-02-1479 was re-directed to prepare the Abingdon CAFB pilot plant for development tests on coal gasification. This work was substituted for that originally planned for further tests on heavy fuel oil for Run 11, (Task 3). Before the elapse of contract 68-02-1479, a further contract with EPA was started in May 1976 to provide for direct support for the San Benito demonstration programme, and this contract 68-02-2159 is still in progress, due for completion in May 1979.

Work Objectives

The initial list of tasks adopted for this contract is set out below:

Task 1 To evaluate three new limestones and a new fuel for CAFB operations in batch tests.

Task 2 To consolidate understanding of the process by analysing data derived from previous tests, and based on design parameters, using mathematical model(s), predict the performance of the CAFB unit.

Task 3 Test proposed design features for the CAFB demonstration plant.

Modify the Esso pilot gasifier to incorporate proposed design features to be tested and if necessary, to alleviate problems encountered during phase three operation. Modificatons relating to proposed design features could include installation of a different distributor, a different product gas burner, and different equipment for cold start-up.

Test proposed design features by examining their effect on start-up, sulphur removal efficiency, and turn down during operation of the gasifier to provide 270 hours of steady state gasification, and 10 data points at lined out conditions.

Cool down unit and inspect.

Task 4 To evaluate one or more of the following:

1. Procedure change and/or modification indicated by results of Task 3.

2. An additional limestone.

3. A heavier fuel, e.g. a vacuum pipestill residuum.

4. Other design features proposed for a demonstration plant.

5. Test programme for a demonstration plant.

6. Control and monitoring instrumentation for the demonstration plant.

Modify Esso pilot gasifier as necessary for evaluation. Operate gasifier to provide 270 hours of steady state gasification, and a maximum of 10 data points at lined out conditions.

Task 5 Provide advice, consultation, and technical expertise to support start-up, shakedown, commissioning and operation of a demonstration unit.

Consult with other EPA contractors involved in design and evaluation of the demonstration plant.

Task 6 Determine effect of the CAFB process on potentially harmful elements other than SO_2 and NO_x.

Analyses of fuel oil, limestone, flue gas, and solids withdrawn from the Esso gasifier, boiler, and stack within the limits of analytical procedures currently available for those elements listed below.

Mercury	Molybdenum	Arsenic	Cobalt
Beryllium	Nickel	Selenium	Boron
Cadmium	Vanadium	Chromium	Lead
Antimony	Tellurium	Manganese	

Examine 6 sets of samples for other potential pollutants such as sulphates and free acids which are potentially emitted by the CAFB process in flue gases or other waste gas streams.

These tasks were modified during the execution of the contract as follows:

Task 1 Only one limestone was provided for evaluation. Since the limestone gave reasonably good performance and was provided from a source within an acceptable distance of the test site in Texas, no further limestones were evaluated, as these would have to be transported a considerable distance to the test site.

The studies on a new fuel for CAFB operations were directed solely towards gasification of a solid fuel.

Task 2 No change.

Task 3 Because of the success of the statistical data analysis techniques developed under the previous contract (Ref. 3), it was found to be possible to utilise data from both steady and transient gasification conditions. Consequently, the number of data points generated was increased from the 10 proposed under this task originally to 192 in actual fact. This made it possible to bring forward into Task 3 most of the studies proposed under Task 4, including gasification of a heavier liquid fuel, and studies on the feasibility of gasifying a solid fuel.

Task 4 Because of the increased amount of data generated under Task 3, and the need to advance the timing of continuous pilot plant operations on solid fuels, this task was redirected into modifying the CAFB pilot plant to be ready for extended operations on solid fuels.

The work carried out on this modified unit formed part of the initial work under contract 68-02-2159, and has already been reported as a topical report under that contract (Ref. 7).

Task 5 Due to change in the demonstration programme timing, support for startup, shakedown and commissioning of a demonstration unit was not carried out.

Task 6 Analysis for trace elements in fuel oil, limestone and solids withdrawn from the gasifier, boiler and stack were carried out.

However, this task was supplemented by the EPA who engaged GCA Corporation, Bedford, Massachusetts to carry out a Level 1 environmental impact study on all the emissions from the CAFB pilot plant. This work has been reported (Ref. 8).

Reporting and Discussion of Results

The discussion is set out in Section 4 by task. Wherever necessary for additional clarity, reference is made to significant events external to this contract which are summarised above, and which influenced the direction or emphasis of the experimental programme.

CONCLUSIONS

Task 1

The Texas limestone tested in the CAFB batch gasifier showed satisfactory performance with regard to grindability, low attrition losses, good sulphur retention and ease of regeneration. It is considered a suitable candidate stone for the San Benito demonstration programme provided it is available in a size range which enables fluidisation under start-up conditions.

Gasification and desulphurisation of Texas lignite and Illinois No. 6 sub-bituminous coal is possible in the CAFB process. A minimum sulphur removal efficiency of 73% was observed, for Illinois No. 6, with 54% of the carbon gasified. Quantitative results for Texas lignite were not obtained due to equipment unreliability. Conditions during these experiments were not optimised and improved performance can be expected with better control of the test conditions. The ash present in the coal and lignite did not present any difficulty, partictularly during regeneration conditions when the higher temperature might lead to fusion. Most of the ash was rejected from the system via the product gas cyclones.

Task 2

The early part of Run 10 covered operations of a new CAFB pilot unit on heavy fuel oil from the same batch as that used during previous tests. A direct comparison is therefore possible between the sulphur removal performance achieved during Run 10 with that observed previously, the major variable being the configuration of the gasifier, regenerator and solids handling equipment.

Statistical evaluation of the Run 10 heavy fuel oil results confirms that the sulphur removal efficiency achieved for the new configuration is similar to that previously observed, and that performance can be predicted using the same operating variables already identified (see below and (Ref. 3)). Thus it appears that the physical configuration of the pilot unit is of at least second order importance in its effect on the sulphur removal efficiency.

A new variable was identified during Run 10, viz bed age. Thus, the time without limestone make-up has a cumulative and significant effect on sulphur retention and therefore must be included with the significant variables already found.

Sulphur removal efficiency can be predicted by a multivariable polynomial equation based on the following operating variables:

Gasifier bed depth
Gasifier bed temperature
Air/fuel ratio
Cyclone drain temperature
Ca/S mole ratio
Added water
Hours without limestone make-up

No correlation was found between variables associated with the gasifier bed characteristics e.g. carbon, sulphur levels, fines concentration, and sulphur retention and these do not appear to have a consistent, independent effect on the process.

The similarity of the regression equations for Runs 8, 9 and 10 independently meant that a generalised equation could be developed. This was done by correcting each equation to standard conditions and then combining them on a weighted basis. The generalised equation was then applied to each run separately and whilst some small loss of precision occurred as a consequence for the individual runs, this is a minor disadvantage compared with the benefit of having a single predictor.

The overall equation was successfully applied to results obtained for TJ 102 Medium Vacuum Bottoms (Bitumen), indicating that fuel characteristics are of secondary importance in the CAFB process. A change of fuel type, e.g. to solid fuels may however, demand separate treatment as additional variables such as feed particle size distribution may be significant.

A considerable improvement in precision is observed compared to previously (Ref. 3) when the updated equation is applied to the selected mean results available from Runs 6 and 7. Nevertheless, further examination of this information is recommended using the hourly data sets. This will further improve the precision of the regression equation, and may lead to the identification of other important variables.

It must be emphasised that regression equations of the type developed here are empirical, and strictly should be applied only within the range of the variables for which they were developed. Also they should not be applied to data produced under significantly different operating regimes, or equipment, unless this can be justified for other reasons. However, physico-chemical reasons can be advanced, and a tenable theory postulated to explain the CAFB process in terms of the major variables identified, and thus the regression equation developed can be applied as a performance predictor as a first approximation in new situations.

Tasks 3 and 4

These can be considered together, since it was possible to bring forward into Task 3 the work planned under Task 4 as a consequence of the success of the new data handling techniques.

Design features incorporated in the CAFB pilot plant specifically to support the field demonstration programme were generally proved to be effective. Thus, the two stage gasifier distributor incorporating a central depression, or pit, did not cause any difficulty with regard to bed fluidisation or deposits. Fall back of the lime bed was evident but this was no worse than experienced previously and essentially is a feature of nozzle design. Improvements need to be made to minimise stone fall back for future operations. The distributor design enabled fuel injectors to be protected from tip burning by being contained in the refractory pit wall whilst still providing good fuel distribution. No nozzle damage occurred as a result. Two entries were provided for evaluation in the distributor viz a hole drilled through the refractory and a "V" channel exposed to the bed. Both alternatives were equally effective and both allowed fuel injectors to be withdrawn and inserted at will.

Fuel injection was carried out at a variety of locations, and combinations of locations during the run. Initially, side injectors were used, but it proved possible eventually to use a single pit injector to deliver all the required fuel into the gasifier without resulting in any loss of desulphurisation performance. It is intended to make this a permanent feature of the pilot unit, thus considerably simplifying the fuel supply system.

Bitumen was successfully injected in the same way and was successfully gasified and desulphurised. Minor difficulties were experienced with the bitumen delivery system due to an inadequately heated filter which resulted in an inability to measure fuel flow. Improvements to this feature are proposed if the need arises to use such fuels in future.

Coal, (Illinois No. 6), was successfully injected pneumatically into the gasifier via a tuyere inserted through the warm-up burner housing. Operations on a mix of coal and bitumen was successfully demonstrated. However, coal delivery was very erratic and it is not possible to provide quantitative information on plant performance and a new coal feed system is required to support further investigations on solid fuels.

Two new means of dealing with carbon deposition in the product gas ducts were considered, both with only limited success. Air jets at the cyclone entry from the gasifier proved to be completely ineffective in preventing deposit accumulations. A procedure was evaluated to burn off the deposits from the boiler burner end rather than the normal burn out from the gasifier end, of the product gas ducts. This was a partial success in that the duct deposits were cleared effectively, but it was not possible to clear the cyclone entries since contacting of oxygen with the carbon deposits in these areas was not very efficient.

Flue gas recycle via a tuyere was successfully tested and it is planned to make this a permanent feature for future operations. Penetration of the flue gas tuyere into the bed must be small to prevent burning. Flue gas clean-up thus becomes a lower priority and it will be adequate to partially clean the gas using a cyclone to remove the larger particles potentially harmful to the gas recycle blowers. Currently, flue gas is taken upstream of the stack cyclones; it is proposed to provide a downstream take off to minimise particle loadings.

As a result, the bag house filter installed for flue gas clean-up is superfluous for the pilot plant. However, as a feature for the field demonstration programme, a number of aspects need improvement. A major requirement is that the bag house filter will require facilities for warm-up prior to use to avoid condensation on the filter fabric. This proved to be a major problem as a solid cake was formed which could not be cleared from the bag, causing excessive pressure drop and greatly reduced flow. Also due to condensation, solids accumulating at the bottom of the filter in the discharge hopper could not be removed. A minor difficulty with the design used was leakage of gas from the filter housing itself, but this could be easily overcome in practice by modifying the housing construction.

Other features not specifically associated with the field demonstration project were evaluated and the following conclusions reached.

The integrity of the gasifier and regenerator construction was excellent with only minor cracking of the new refractory having occurred during Run 10. The redesigned cyclones performed very satisfactorily and there was no excessive solids carry over into the boiler. The cyclone drain and fines returns system were subject to difficulties initially but operational modifications greatly improved their performance. A re-design of the drain lock hoppers is required to eliminate the accumulation of chunks of carbon directly over the discharge line on the perforated retaining plates. The solids transfer system was entirely satisfactory after a solids accumulation caused by condensation in the gasifier to regenerator transfer line was cleared early in the run. However, the rodding ports available for this system need to be enlarged. The limestone feed system was a major problem area, and provision is needed to prevent the excessive damping of the vibrator table experienced in Run 10.

The main gasifier blower was found to have leaks on the casing and correction of the measured air flows were necessary. The blower system was limited in capacity and a new, positive displacement blower will be considered as a replacement.

Regenerator performance was good throughout, but characteristic deposit formations were found around the top of the distributor at the end of the run. These arise when the gasifier warm-up is in progress, when the cold regenerator behaves as a condenser. Agglomerates then form in the relatively quiescent zone below the distributor nozzles when the initial stone flow from the gasifier is established.

The layered insulation within the steel shell of the gasifier and regenerator unit proved to be very effective and the skin temperature was kept to no more than about 60°C, with no hot spots. The reduced heat losses mean that shut downs become of lesser importance since the cooling of the bed is now very slow. This opens up the possibility of running short term tests e.g. daily, using the continuous gasifier.

Persistent troubles were experienced with the gas sampling system because of analyser reliability and leaks in the sampling lines, particularly in the boiler gas sampling train where a water knock out system was difficult to seal.

Task 5

Ongoing discussions were held with all EPA contracters involved in the EPA CAFB programme. In particular, several reviews of the demonstration plant design basis were held with Foster Wheeler Energy Corporation. A formal design review meeting was held at Esso Research Centre, Abingdon during May 1976 (Ref. 9).

Task 6

Advantage was taken of a prolonged period, (68 hours), during Run 10 when no fresh limestone could be added, to investigate retention of trace elements present in the heavy fuel oil feed. Samples taken at the start of this period, after 24 hours, and at the end of the 68 hours were selected for analysis. A comprehensive range of trace elements were checked, but only a few could be estimated with sufficient precision to enable sensible balances to be calculated.

In the short term, virtually all heavy trace metals e.g. vanadium, chromium, nickel and lead are retained. There is some retention of lighter elements such as potassium and sodium.

Longer term, retention efficiency deteriorates and it appears that fines produced in the bed carry trace elements over to accumulate in the stack particulate collection system. This was particularly true of the lighter elements.

The maximum concentration of vanadium observed on the gasifier bed was 0.6 wt%.

The attrition rate of the limestone bed was measured over this period and found to be 0.12 kg/hr./100 kg bed/m^2 of gasifier bed area. At constant bed depth, this represents a feed rate of approximately 0.2 moles of calcium per mole of sulphur, for the fuel oil used in these tests.

RECOMMENDATIONS

1. The fines returns system used during Run 10 will need modification for future test work using coal. Means will have to be incorporated to segregate ash for rejection from the fines collection and return system. Also, the need to re-inject lime and carbon fines from below gasifier bed level at the San Benito demonstration unit must be covered. If possible, a system simpler in operation than that used during Run 10 should be devised.

2. A suitable storage, transfer, metering and delivery system for coal feed to the gasifier is required.

3. The gas analysis instrumentation used for Run 10 proved unreliable with a number of troublesome breakdowns and excessive drift requiring frequent calibration. To improve reliability and to reduce down time, it is recommended that new instrumentation is provided for analysis of gas streams.

4. The flue gas recycle system can be modified to enable recirculation directly via tuyeres into the gasifier bed, thus eliminating the need for bag filters.

   Bag filters, when used for flue gas clean-up, partic-uarly on an intermittent basis, should be designed with provisions for preheating, and insulation to avoid condensation. Collecting hoppers should be designed to enable easy discharge of solids.

5. In view of the improved insulation of the gasifier-regenerator reactors, it is recommended that inter-mittent operation of the pilot unit is investigated as a means of improving the flexibility of operations. Thus, a number of short term tests, spaced out over a period of time could be carried out without need for shift operations.

6. A number of other changes are recommended to improve reliability, and flexibility of operating the pilot unit. These include:

   a) A new gasifier air blower.
   b) Replacement of manometers by differential pressure gauges.
   c) Improved air nozzles to minimise lime fall back.
   d) Fitting a gas oil injection system for the regenerator to improve warm-up.

e) Tapered nozzles for the gasifier and regenerator pressure tappings to prevent blockage by increasing bleed velocities.

f) Increasing the diameter of the solids transfer system rodding ports.

7. A new heat exchanger is required for the boiler system to restore the capacity of the system.

8. Arising from Run 10, the fuel oil feed system can be considerably simplified and can be reduced to a single injector located at the distributor pit.

9. Modifications are required to enable the lime feed metering vibrator to function efficiently over long periods by minimising packing of fines around the vibrator supports.

10. The successful statistical analysis techniques developed for describing the performance of the CAFB process on liquid fuels should be extended to cover solids fuels.

11. If the limestone provided from Whites Mines, Texas is confirmed for use in the San Benito demonstration unit, it should be substituted for BCR 1359 for future test runs. If an alternative supply source is preferred it should be screened for suitability in the batch unit before final selection for the demonstration plant or the CAFB continuous pilot plant.

DISCUSSION

CONTINUOUS PILOT PLANT STUDIES, RUN 10

PREPARATION OF CAFB CONTINUOUS PILOT PLANT FOR RUN 10

The pilot plant used up to Run 9 had exceeded approximately 5,000 test hours and during that time it had suffered the stresses normally encountered during numerous start and shut-downs together with several unplanned and uncontrolled temperature excursions, pressure surges, and one propane-air explosion. Numerous cracks were evident in the refractory, several redundant penetrations were present, and the integral cyclones for product clean-up were very sensitive to gasifier pressure fluctuations.

It was therefore decided to break out the existing gasifier, regenerator and cyclones and to completely redesign and up-date the pilot unit to give an improved facility for further studies.

Observations During Break Out

The gasifier refractory, apart from deep cracks and some severe surface erosion at about 1m (3 ft) from the base was in good condition. The bottom 1m (3 ft) of the regenerator refractory was very soft up to a depth of 5cm (2 inch), and eroded to as much as 5 cm (2 inches), in places, particularly in the vicinity of the gasifier to regenerator solids transfer outlet. Refractory in the upper portion of the regenerator was in good condition.

Several gasifier refractory cracks were found to continue through to the outside casing, and were stained dark gray to black indicating that product gas, with its associated tars had diffused through to the skin of the pilot unit. Two cracks between the product gas cyclones and the gasifier, and one crack between the right hand product gas cyclone and the regenerator were also continuous and could have caused gas leakage. The crack between the gasifier and regenerator, extending from approximately 15 cm (6 inch) from the base to 75 cm (30 inch) from the base was also continuous and approximately 6.5 mm (0.25 inch) wide in places. The product gas cyclones were badly eroded, (up to 1.5 cm (0.6 inches) opposite the gas entries) and at the bases of the cones.

Calcium silicate insulating slab at the base of the gasifier-regenerator monolith was, in places, completely saturated with tar.

Self bonded silicon carbide pipes, supplied by British Nuclear Fuels, used for product gas cyclone off-takes, lining the lower parts of the cyclone drains and the fines returns injector, and as a sleeve insert around the regenerator distributor were all found to be in excellent condition.

These observations during the break out of the old unit confirmed most of the assumptions based on the initial, external examination. Particularly, the crack between the gasifier and regenerator vessels, and the accumulations of tarry deposits in the insulation were worse than expected.

Features Noted for Redesigned Equipment

The type of refractory used for the gasifier-regenerator monolith was considered suitable for CAFB operations with an expected in-service life of at least 6,000 hours for the gasifier and upper part of the regenerator, and 3,000 hours for the lower portion of the regenerator.

To minimise cracking, the gasifier-regenerator separating wall must have either a gas proof membrane, or some flexibility to allow independent movement of the two components. Corners and severe discontinuities in the refractory should be avoided if possible but it would be desirable to incorporate planes of weakness so that refractory cracks occur in predictable fashion in regions where they are of little consequence.

Provision should be made to pressurise the outer insulating layer to slightly above the gasifier and regenerator operating pressure to prevent leakage of product gas containing tars through cracks in the refractory when these propagate through to the walls of the unit.

The regenerator should be arranged so that, if necessary, it can be broken out and re-cast without affecting the gasifier, as its life expectancy is less than the gasifier.

Improved cyclones need to be less sensitive to pressure fluctuations and to erosion opposite the gas inlet whilst the cyclone drains should be lined with a higher grade refractory.

Self bonded silicon carbide as supplied by British Nuclear Fuels Limited appears to be an eminently suitable material of construction for locations where the environment is particularly hostile but where mechanical stress is not a serious problem. Silicon carbide is brittle and liable to break up when treated to mechanical stock.

Basis of New Design

The experience summarised above proved invaluable as a starting point in arriving at a new design for the CAFB pilot plant. A detailed discussion of the design and construction of the new unit is given in Appendix A in terms of the required range of process operations and within the limitations of the physical facilities and services capacities available. A summary of the major features follows, and includes the changes incorporated as a direct result of experience with the original unit and to accommodate specific items to support the design of the demonstration unit planned at San Benito in Texas.

Improved insulation was incorporated to reduce heat losses and skin temperature to more closely simulate a larger installation, and the skin was pressurised slightly to prevent leakage of product gas and tars into the insulation and the occurrence of local hot spots. The gasifier was designed with a circular section, as was the regenerator and they were cast in separate refractory monoliths. These, and other precautions such as provision of expansion joints, were intended to minimise cracking. Such cracks as were expected to develop were arranged to be in areas of refractory where they would be relatively unimportant, by suitable location of anchor points in the refractory.

External cyclones providing high gas flow with minimum pressure drop and high efficiency were designed with external snail gas entry ducts to increase gas velocity and were confidently expected to be insensitive to normal pressure and flow fluctuations through the system. New product gas ducting was needed to accommodate the revised plant lay out.

Specific items relating to the demonstration unit were planned for in the new design. A two level gasifier air distributor was incorporated, allowing fuel injectors to be inserted into a central pit through a protective refractory layer. Provisions were made to recycle flue gas directly into the gasifier bed via a tuyere to reduce the need for particulate clean-up of the flue gas stream. As an alternative, a bag filter was incorporated to provide operating experience and to evaluate design and performance features. New systems were designed to handle liquid fuels heavier than the normal heavy fuel oil feed used for the pilot unit;

heated storage, lines, metering and pumping facilities were provided, and a temporary system was constructed to demonstrate the feasibility of pneumatic injection of solid fuel into the gasifier.

These major changes meant a larger number of modifications to most of the sub systems of the pilot plant. Particularly, numerous changes were made to the air and nitrogen systems, the gas sampling trains, the pressure monitoring equipment, the heavy fuel oil supply system, the gasifier, regenerator fluidising air and flue gas recycle supply systems and the cyclone drain-fines re-inection system. These in turn demanded changes to the operating procedures.

The equipment lay out and operating procedures for the new pilot plant, as used for run 10, are detailed in Appendix D.

EXPERIENCE WITH NEW PILOT PLANT DURING RUN 10

A summary on a daily basis of the Run 10 log is given in Appendix C; highlights of the operating experience during run 10 are given below.

The integrity of the gasifier, regenerator, cyclones and hot gas ducts was more than satisfactory, but the usual teething problems associated with the functioning, performance, and reliability of new equipment were experienced from time to time. During the run, most of these difficulties were solved, but some experimental equipment did not give satisfactory, reliable performance throughout the test period. As a consequence, the run did not proceed smoothly, particularly during the early periods when most of the difficulties came to light, and moreover, the reasons for some of the automatic plant shut downs could not be traced conclusively since attention was often directed to some aspect of plant performance other than the functioning of the gasifier and regenerator. The progress of the run can conveniently be broken down as follows.

Pre-run refractory curing and warm-up

This proceeded smoothly throughout with only minor unavoidable interruptions due to the need to complete final construction details and to link in various associated auxiliary systems and equipment.

Days 1-7

Gasification of heavy fuel oil was carried out during this time with a service factor on the plant of 77%.

It was obvious however, that there were serious operational difficulties with the cyclone drain, fines returns, solids transfer, limestone feed and gas sampling systems and these caused the eventual shut down on day 7 so that the problems could be examined on a systematic basis.

Days 7-12

Essentially the unit was inoperative throughout this period during which modifications were made to improve those parts of the system proving troublesome as identified above. Also, a new regenerator distributor of a different type was fabricated and fitted.

Days 12-19

Fuel oil gasification was in progress for most of the time with one continuous run of 107.5 hours. Plant control and performance was greatly improved but the limestone feed system still did not function satisfactorily and problems still occurred with the gas sampling systems. The service factor over this period was 82%.

During this period, a variety of experiments were carried out including fuel oil injection through a single injector, steam injection, and an extended period of operation without fresh limestone make-up.

Days 19-21

Bitumen was substituted for the heavy fuel oil and successfully gasified and desulphurised. Flow measurement of the bitumen was not possible due to plugging of inadequately heated filters upstream of the flow meter.

Boiler over heating problems curtailed these experiments.

Days 21-22

Coal gasification was successfully demonstrated though with generally poor plant performance as flow control with a very simple feed system was difficult. Illinois No.6 coal was used at a rate of 50% of the total fuel feed, with the remainder being bitumen.

Throughout the total period, including the prolonged shut down, a total of 520 hours were available for gasification. In fact, fuel oil was gasified for 263.5 hours, bitumen for 34.5 hours and coal + bitumen for 7.5 hours giving an overall service factor of approximately 59% for the run. After elimination of incomplete and obviously erroneous data sets, and taking into account periods when time did not permit data collection, a total of 192 hours data on heavy fuel oil, and 17 hours on bitumen were available for detailed analysis. No data collection was possible for the short period of coal/bitumen gasification.

A summary of the performance and reliability of the equipment during the run is given in Table 1.

During days 4 - 21, personnel from GCA Technology Division were present to conduct an assessment of the environmental impact of the CAFB process.

INSPECTION OF PILOT UNIT, POST RUN 10

The post run inspection revealed that there was no serious damage to the refractory lined reactors and vessels viz gasifier, regenerator, cyclones and product gas ducts, including the gasifier and regenerator distributors and lids. However, considerable deposits of carbon were present in the cooler regions of the gasifier and in the product gas lines as the unit was shut down whilst on gasification so that deposits could be examined.

Some small accumulations of agglomerated material comprising carbon and lime were present in the gasifier pit and approximately 30% of the air nozzle holes were plugged. The plenum contained a quantity of lime which had fallen through the fluidising nozzles during the run.

Deposits were found above the regenerator distributor, and also in the transfer system entry boxes in the gasifier and regenerator.

Accumulation of carbon chunks were present in the cyclone drain system, retained by a perforated plate inserted for this purpose. However, they accumulated immediately

TABLE 1

EQUIPMENT RELIABILITY DURING RUN 10

Component	Details of Malfunction	Number of Incidents
Cyclone Drain/ Fines Re-injection	Valve blockages	6
	Line blockages	24
	Electrical control box faults	2
Solids Transfer	Blocked ducts	12
	Faulty valves	5
Limestone Feed Vibrator	Excessive damping	Continuous
	Electrical supply failure	1
	Blockages	10
Regenerator	Failure to fluidise	5
	Off-take leak	1
Pressure Tappings and Manometers	Blocked tappings	14
	Blown manometers	Continuous
	Leaks	2
Flue Gas Recycle Bag House	Housing leaks	3
	Bags wet, blocked	3
	Drain hopper blocked	2
Gasifier Air Blowers	Leaks	3
Heavy Flue Oil Delivery	Seized pump	1
	Trace heating cold	2
	Faulty secondary heating	3
	Leaks	2
	Blockages	1
Bitumen Delivery	Trace heating inadequate	1
Boiler	Overheating	3
Analytical Equipment	Regenerator sample line blocked	7
	Regenerator sample line leak	1
	Boiler gas sampling train leaks	10
Electrical Equipment	Faulty chart recorders	3

above the fines discharge line, and thus seriously interfered with the flow of fines through this system.

Further details of the unit strip down, together with photographic evidence can be found in Appendix C.

ANALYSIS OF CONTINUOUS RUN 10 RESULTS

Introduction

Previous reports in this series have described how the results from runs 6 to 9 inclusive were analysed (Refs. 1, 2). For run 10, following initial work-up of the results this statistical approach has been continued and each hour's data has been considered separately. The value of this approach was demonstrated during the course of run 10, when the equation derived from runs 8 and 9 was found to give a good prediction of the sulphur removal efficiency observed for the rebuilt CAFB pilot unit. This proved to be a valuable tool for monitoring plant performance on an ongoing basis during the run.

Thus, a stepwise multiple regression analysis technique was used on all the run 10 results to identify the variables of major importance describing the plant performance. The linearity, or otherwise, of the contribution of each variable in turn was investigated and a polynomial expression derived if appropriate. The functions thus derived for each variable then constituted an expression giving the optimum correlation of the measured sulphur removal efficiency for the plant during run 10.

This exercise was conducted for the results available for heavy fuel oil thus enabling a direct comparison to be made between run 10 and the previous runs. Following this comparison, an exercise was conducted to produce a single equation which best described the sulphur removal performance of the CAFB pilot plant.

Data collected for operations on bitumen was not included for the initial analysis, but the heavy fuel oil equation was subsequently applied to the limited results available to establish the relevance of the equation. No such comparison could be made for coal as CAFB fuel on this instance as no readings were taken due to the demands of running the unit with the crude coal feed system.

An outline of the statistical techniques used for analysis follows, after which the detailed data analysis is discussed.

Statistical Analysis Techniques

Stepwise Multiple Regression

Stepwise multiple regression analysis of the data was carried out using a standard programme 'Mul-Correlation' available on the Honeywell MK III Foreground "STATSYST" system. This programme produces linear equations of the form:-

$$y = b_o + b_1X_1 + b_2X_2 + --- b_mX_m + e$$

where y is the dependent variable

X_1, --- X_m are the independent variables

b_o, --- b_m are the regression coefficients to be determined

e is a random error term which gives the difference between the predicted and actual values for the dependent variable

In this stepwise regression routine, the basic premise which distinguishes it from conventional approaches to multiple regression is that intermediate partial regression equations are developed to indicate whether a variable is significant in an early stage of the regression calculation so that it may be entered into the regression at that stage. The final regression equation should, therefore, only contain significant variables. The programme allows the user to specify the level of significance such that the independent variable is entered or removed from the regression equation during the analysis. However, it is possible to override this mechanism and force a variable into the regression equation despite the fact that it may not meet the specified significance level criteria. This is obviously useful when the variable is known to be significant from other sources.

Limitations of Multiple Regression

Although multiple regression analysis, including stepwise, is an extremely powerful tool, it is subject to certain limitations which the user should be aware of in order not to be led astray.

Perhaps the most important point to remember is that it is not possible to infer cause and effect relationships from regression analysis. The fact that a dependent variable is highly correlated with an independent variable in no way suggests that a change in the independent variable causes a change in the dependent variable. Although such relationships may exist they cannot be proved. We can say that the results do not refute the theory, but neither do they prove it.

A second important point is that regression relationships are empirical equations that apply only to the range of data on which they are based. Extrapolation can lead to highly erroneous results. Since regression equations are based on the experimental data, then the equation is no more accurate than the data. If the data is subject to large experimental error, the regression equation will predict inaccurately.

An underlying assumption in regression analysis is that the independent variables are truly independent, and that there are no interactions between variables. In practice great skill in experimental design is needed to achieve this end, and sometimes in complex processes, such as this one, it is only possible to a limited extent.

A final important point to remember is that regression analysis assumes linear relationships between the dependent variables and the independent variables. In order to deal with variables which are obviously non-linear, it is necessary to transform the variable into some algebraic function of itself before carrying out the regression.

Development of Polynomial Equations

In order to develop polynomial expressions for the non-linear variables identified as significant from the linear regression stage, the following steps were taken.

1. Assume as a first approximation that all the regression variables are linear except for one. Correct the % sulphur removal efficiency experimental values using the linear regression coefficients for all the variables except the one for deviations away from the mean values of these variables.

2. Plot the corrected % sulphur removal efficiency values against the selected variable and to improve accuracy, average the data into boxes covering the experimental range of the variable e.g. for bed depth, results would be averaged in the range 90-100 cm, 100-110 cm etc.

3. Repeat for each variable in turn.

4. The optimum polynomial expression for the relationship between the variable and the sulphur removal efficiency may then be found using standard statistical techniques. In this instance, a Polynomial-Fit programme included in the Honeywell MK III "STATSYST" package was used to investigate the linearity, or non linearity of each significant variable.

This programme calculates coefficients and comparative data for fitting polynomials of the form:

$$y = a + bx + cx^2 + dx^3 + \text{-----} \; mx^7$$

where the coefficients a, b ---- m may or may not be zero. The equation has maximum order 7, and the programme allows weighting factors to be used, so that it was very suitable for application to the grouped data approach adopted in this analysis.

Obviously, having carried out this exercise starting with the assumption that the variables were linear, a recycle through the steps above starting with the non-linear equations would be possible with an improvement in the precision of the equation so derived. However, this is costly in both time and money and the resulting benefit is likely to be of doubtful value. It was not included in this analysis.

A useful tool throughout the analysis stage was to plot the residual error viz. % sulphur removal efficiency (measured - calculated) against time throughout the run. This enabled periods to be identified where consistent, and occasionally large discrepancies occurred, and by reference to the run log book and data sheets it was usually possible to pinpoint the causes. Thus additional important variables could be investigated and occassionally errors in input data could be identified.

Computer programmes described previously (Refs. 2, 3) were used for the data grouping and graph plotting exercises.

Run 10 Data Analysis

Editing of Run 10 Raw Data

During the run, a total of 264 hours gasification of heavy fuel oil was accumulated over varying time periods. Generally, data was recorded for each hour's operation for most of the time. However, it was evident at the outset of the analysis that some data sets would have to be rejected for various reasons. Some sets were incomplete so that the % sulphur removal efficiency could not be calculated as one or more important variables were missing. Results taken within an hour or two of start up, particularly after a bed sulphation step, were ignored as the plant performance was not lined out; some periods of gasification were too brief, and occasionally time was not available to record data when plant operational problems demanded priority. Finally, some data was eliminated during the statistical analysis stage when good reasons were identified that plant performance was far removed from normal.

After elimination of the inconsistent and inaccurate data, a total of 192 separate hourly readings were available for statistical evaluation for heavy fuel oil, and 17 hours for bitumen.

Preliminary Data Work-up

The data co-ordination and work up programmes have been described in an earlier report (Ref. 2). These were used with only some minor detailed modifications to the programmes to take into account the different solids sampling and removal pattern during Run 10, and the different configuration of the new unit.

The end result of the data work up is a consolidated data file which contains all the information to produce the output tables and which served as the data source for the starting point of the statistical analysis stage.

Linear Regression Analysis of Run 10 Results

The multiple regression programme "MUL-CORRELATION" produces a table giving the correlation coefficients of each selected variable with the % sulphur removal efficiency and also the intercorrelation of each variable with all the others included in the analysis.

This information is useful in identifying the variables which are truly independent. A negative correlation coefficient indicates that as one variable increases, the other decreases; positive correlation coefficients indicate that the two variables increase or decrease together. The square of the correlation coefficient (x 100%) gives the "% explained" of the variation of one variable by the other. Thus, a correlation coefficient of 1 indicates that the variables are indistinguishable and 0 that they are truly independent. Obviously, low values of the correlation coefficients are desirable to simplify the statistical analysis.

Table 2 lists the correlation matrix for the variables selected for examination for Run 10. Additional variables were included for this analysis compared to Runs 8 and 9. In particular, it was possible to investigate the effect of elapsed time without the addition of fresh limestone i.e. the bed age effect.

An arbitrary value of the correlation coefficient of ± 0.5 has been selected to identify the variables of major interest. This gives a % explained of 25% or greater, indicating that a reasonably high degree of interdependence exists. These are underlined in Table 2.

Correlation of % Sulphur Removal Efficiency with Process Variables

Table 3 compares the correlation of common process variables with % sulphur removal efficiency for Runs 8, 9 and 10.

Good agreement is shown for all three runs with respect to gasifier bed depth, cyclone drain temperature, added water, Ca/S mole ratio, and bed velocity, with the coefficients being comparable in sign i.e. effect, and to some extent in magnitude also. These variables can in general be considered to be good predictors of % sulphur removal efficiency.

Bed temperature, bed carbon and bed fines are also in good agreement in that all have poor correlation with % sulphur removal efficiency. Under these circumstances, the apparent change in effect is not of any significance.

Bed sulphur level intercorrelation shows quite wide fluctuation, with the Run 10 result showing low significance. This may not be entirely unexpected when the results for Runs 8 and 9 are considered.

Strictly, according to the criterion for significance applied, the results for air/fuel are similar. However, directionally Run 10 appears to be different with a much lower correlation coefficient and an unexpected, negative relationship. It was thought that this was due to a lack of variation in air/fuel ratios during Run 10, but as Table 4 shows, this was intermediate between the Run 8 and 9 variation and thus could not be offered as an explanation. Thus, no explanation for the poor correlation of the Run 10 air/fuel ratio can be proposed at this stage.

TABLE 2

CORRELATION MATRIX FOR RUN 10 VARIABLES

VARIABLE	SULPHUR REMOVAL EFFICIENCY %	GASIFIER BED DEPTH (cm)	GASIFIER BED TEMP. (°C)	AIR/ FUEL RATIO	CYCLONE TEMP. (°C)	ADDED WATER (m^3/hr)	Ca/S MOLE RATIO	GASIFIER BED S (wt %)
Sulphur Removal Efficiency (%)	1.00	0.70	-0.12	-0.17	0.58	-0.55	0.51	-0.12
Gasifier Bed Depth (cm)		1.00	-0.20	0.10	0.83	-0.23	0.23	-0.09
Gasifier Bed Temp. (°C)			1.00	-0.06	-0.24	-0.27	-0.40	0.31
Air/Fuel Ratio				1.00	0.18	0.28	-0.12	-0.30
Cyclone Temp. (°C)					1.00	-0.24	0.27	-0.28
Added Water (m^3/hr)						1.00	-0.28	-0.15
Ca/S Mole Ratio							1.00	-0.14
Gasifier Bed S (wt %)								1.00
Gasifier Bed C (wt %)								
Gasifier Bed Sulphate (wt %)								
Gasifier Bed Velocity (m/sec)								
Gasifier Air (m^3/hr)								
Fuel (kg/hr)								
Gasifier Bed Fines, 600-250μ (wt %)								
Time With No Stone Added (hr)								

cont....

TABLE 2 (Continued)

VARIABLE	GASIFIER BED C (wt %)	GASIFER SULFATE (wt %)	GASIFIER BED VELOCITY (m/sec)	GASIFIER AIR (m^3 hr)	FUEL (kg/hr)	GASIFIER BED FINES, 600-250μ (wt %)	TIME WITH NO STONE ADDED (hr)
Sulphur Removal Efficiency (%)	0.30	-0.09	-0.14	0.24	0.40	0.12	-0.72
Gasifier Bed Depth (cm)	0.34	-0.06	0.19	0.41	0.36	0.09	-0.44
Gasifier Bed Temp. (°C)	0.19	-0.15	0.01	0.08	0.13	0.01	-0.03
Air/Fuel Ratio	-0.29	0.08	0.84	0.53	-0.49	-0.11	0.30
Cyclone Temp. (°C)	0.10	-0.06	0.25	0.51	0.35	0.00	-0.42
Added Water (m^3/hr)	-0.19	0.34	0.23	-0.05	-0.34	-0.26	0.57
Ca/S Mole Ratio	0.07	-0.07	-0.24	0.16	0.22	-0.09	-0.43
Gasifier Bed S (wt %)	0.52	-0.09	-0.24	-0.23	0.11	0.21	-0.17
Gasifier Bed C (wt %)	1.00	0.33	-0.23	0.08	0.37	0.09	-0.28
Gasifier Bed Sulphate (wt %)		1.00	0.01	0.11	-0.01	-0.36	0.29
Gasifier Bed Velocity (m/sec)			1.00	0.52	-0.19	-0.07	0.28
Gasifier Air (m^3/hr)				1.00	0.43	-0.32	-0.01
Fuel (kg/hr)					1.00	-0.15	-0.35
Gasifier Bed Fines, 600-250μ (wt %)						1.00	-0.37
Time With No Stone Added (hr)							1.00

TABLE 3

COMPARISON OF CORRELATION COEFFICIENTS WITH % SULPHUR REMOVAL EFFICIENCY FOR RUNS 8, 9 and 10

VARIABLE	RUN 8	RUN 9	RUN 10
Bed Depth (cm)	0.49	0.61	0.70
Bed Temperature (°C)	0.13	-0.19	-0.12
Air/Fuel Ratio	0.35	0.34	-0.17
Cyclone Temperature °C	0.31	0.56	0.58
Added Water (m^3/hr)	-0.87	-0.25	-0.55
Ca/S Mole Ratio	0.27	0.27	0.51
Bed Sulphur (wt %)	0.38	-0.42	-0.12
Bed Carbon (wt %)	0.16	-0.16	0.30
Bed Velocity (m/sec)	-0.21	-0.19	-0.14
Bed Fines 600-250 μ, (wt %)	-0.003	-0.01	0.12

TABLE 4

SUMMARY STATISTICS FOR RUNS 8, 9, 10

VARIABLE	RUN 8 MEAN VALUE	RUN 8 STANDARD DEVIATION	RUN 9 MEAN VALUE	RUN 9 STANDARD DEVIATION	RUN 10 MEAN VALUE	RUN 10 STANDARD DEVIATION
Sulphur Removal Efficiency (%)	77.1	10.8	79.1	6.8	74.9	8.0
Bed Depth (cm)	91.0	11.8	103.1	14.0	110.0	11.6
Bed Temperature (°C)	891	13.2	921	24.5	920	23.2
Bed Velocity (m/sec)	1.62*	0.17*	1.70*	0.23*	1.54	0.27
Air Rate (m^3/hr)	-	-	-	-	320.1	31.2
Fuel Rate (kg/hr)	-	-	-	-	128.6	9.6
Air/Fuel Ratio (% Stoichiometric)	23.9	1.9	26.9	3.4	23.9	2.7
Added Water (m^3/hr)	16.6	13.1	4.8	8.6	4.0	8.9
Ca/S Mole Ratio	1.58	0.7	1.23	0.9	0.82	1.2
Cyclone Drain Temperature (°C)	365	109	356	134	279	113
Bed Carbon (wt %)	0.3	0.5	0.24	0.5	0.34	0.31
Bed Sulphur (wt %)	5.2	1.8	4.5	1.1	3.98	0.98
Bed Sulphate (wt %)					0.04	0.04
Bed Fines (600-250µ) (wt %)	16.4	3.7	18.4	5.9	27.7	2.7
Time Without Fresh Limestone Feed (hr)	-	-	-	-	12.6	20.1

* Corrected values from results provided in (Ref.2)

Table 2 shows that the bed age effect, i.e. the period of time for which no bed make up was added is a very important variable, showing the highest correlation of all with the % sulphur removal efficiency. This effect was quantified for the purpose of statistical evaluation by assigning the natural number sequence (1, 2, 3 ---) for successive hours without limestone make up. The variable does not appear in the analysis for previous runs as stone make up was supplied virtually throughout the run.

Intercorrelation of Other Process Variables

Some significant (i.e. greater than 0.5) correlation coefficients are shown for some of the other dependent variables also. Comments are given below.

Bed depth and Cyclone Drain Temperature

The cyclone drain temperature is a crude measurement of the rate of fines recirculation through the main gas cyclones back into the gasifier bed. It can be expected that the deeper the bed, the greater is the opportunity for fines to enter the gasifier outlet gas ducts and the cyclones, and hence the greater the cyclone drain temperature. This strong correlation is therefore gratifying.

Air/Fuel Ratio and Bed Velocity

This is a natural correlation. As leaner operation is introduced at a fixed fuel flow rate (i.e. the air rate to the gasifier increases) so bed velocity also increases.

Air/Fuel Ratio and Gasifier Air and Fuel Rates

As might be expected, there is a strong correlation between air/fuel ratio and gasifier air, since air rate and bed velocity are correlated (see above). Similarly there is a strong correlation, in the opposite direction, with the fuel rate.

Gasifier Air and Gasifier Bed Velocity

This correlation is significant but not as great as might be expected due to the fact that the gasifier bed velocity is calculated on the basis of the air rate plus flue gas recycle rate. Flue gas recycle was used at various times throughout Run 10.

Added Water and Time Without Stone Addition

This is a purely fortuitous correlation due to the addition of steam towards the end of the no stone addition period. Thus, high water addition rates are naturally associated with high bed age variable values.

Bed Carbon and Bed Sulphur

The correlation matrix indicates that high bed sulphur levels are associated with high bed carbon levels. During the run, there were periods when the unit was operated with high carbon levels on the gasifier bed and as a consequence, the regenerator was overcarboned and out of action. During these periods there would be an accumulation of sulphur on the gasifier bed, leading directionally to be intercorrelation observed.

Summary Statistics for Run 10

Mean values and standard deviations for Run 10 are given in Table 4. These results are for heavy fuel oil, and thus can be compared directly with Runs 8 and 9, both in terms of the mean values and in the standard deviation. The latter gives a measure for how widely conditions were varied during the run, and a high value is desirable, showing that a wide range of operating conditions are covered in order to improve the precision of the statistical analysis of the data. Comments on particular aspects of the summary results are given below.

% Sulphur Removal Efficiency

Overall, Run 10 sulphur removal efficiency was comparable to the previous runs, the slightly lower value for the mean % sulphur removal performance indicating only that the unit was operated during Run 10 under less favourable conditions for sulphur retention than for Runs 8 and 9.

Bed Depth

Generally, the gasifier bed depth was greater during Run 10 than previously, but surprisingly in view of the lack of stone addition for long periods, did not show the expected greater fluctuations.

Bed Velocity and Stoichiometry

Bed velocity results reported for Runs 8 and 9 have been found to be in error for two reasons.

First of all, the recorded air flows to the gasifier are too high due to leakage downstream of the orifice plate and the blower. Secondly, a correction factor had been omitted from the equation used in the work-up programme ZKDAT (Ref. 2) to calculate bed velocity.

Thus, the bed velocity results in Table 4 for Runs 8 and 9 have been corrected for the omission of the correction factor in ZKDAT.

However, correction for the loss of air downstream of the measuring orifice plate is more difficult, since there is no way in retrospect of determining when the leak developed and how it changed with time. As a rough estimate, when the leak was corrected during Run 10, a slight increase in the gasifier bed temperature was observed indicating that the magnitude of the leak was great enough to change the stoichiometry. It was possible to calculate that there was approximately 1% leaning off of the air/fuel ratio and that the gasifier air rate increased by 4.3% (13.5 m^3/hr or 8 cfm) though a significant change in the orifice plate pressure measurement could not be detected at the time. In view of the speculative nature of these deductions, no corrections have been made to the data in Table 4 for leakage, either to bed velocity or stoichiometry.

Nevertheless, the presence of the leak, and the doubts raised as a consequence, are important in explaining differences between runs so far as the stoichiometry - sulphur removal efficiency relationship is concerned. Comments will be made on this point later.

Stone Feed Rate and Time Without Limestone Make-Up

The average fresh limestone make-up rate was relatively lower and more variable for Run 10 due to operational problems with the limestone feed equipment. There was one continuous period of 68 hours during which no fresh limestone was added, and this permitted the bed age effect to be investigated. It was necessary to quantify the time variable, and this was simply done by assigning zero values to hourly data when stone was added, and the natural number sequence 1, 2 --- to successive hours when no fresh limestone was fed. The average value of 12.6 hours thus arises mainly from the 68 hours when the stone feed system was inoperative.

The rate at which stone was added is identified by the Ca/S Mole Ratio.

Cyclone Drain Temperature

A significant difference is seen between Run 10 and previously. The thermocouples used to measure the cyclone drain temperatures during Run 10 were located diffferently in the redesigned unit and resulted in the average temperatures, taken as the mean of the drain temperatures for each of the cyclones, being 80°C lower than before. The variations observed are comparable for the three runs.

In general, it can be concluded that the overall results for Run 10 are very similar to Runs 8 and 9, the only major differences of consequence being the cyclone drain temperature mean, and the operation of the pilot unit without stone make-up for a prolonged period during Run 10.

Linear Regression Equations for Run 10

As for the analysis described for Runs 8 and 9, (Ref. 3), a comprehensive investigation was carried out of the Run 10 results to establish the variables of importance in describing the performance of the pilot plant gasifier sulphur removal performance. As a starting point, the variables identified in Table 4 were selected as likely candidates. Due to intercorrelation effects, it was necessary to evaluate gasifier bed depth and cyclone temperature by combining them into a simple function, using the average coefficients for Runs 8 and 9 to proportion the effects. Thus, a function defined as (0.16 Bed Depth + 0.013 Cyclone Drain Temperature) was derived and found to be of high significance. This was justified on the grounds that the intercorrelation of bed depth with cyclone drain temperature is much higher for Run 10 and therefore the linear regression equation could include either variable, but not both together. An equally good correlation is obtained by combining the two variables, and this step enables a better comparison to be made with previous results.

It was found that the significance of bed temperature was low for the Run 10 data and this variable had to be forced into the regression equation.

In order of importance, the regression analysis selected the following variables to explain the % sulphur removal efficiency observed:

Hours without limestone addition i.e the bed age effect.

Bed depth and cyclone drain temperature (combined variable).

Ca/S mole ratio.

Added water.

Air/fuel ratio.

Bed temperature (strictly not significant but forced).

The coefficients obtained from the regression analysis are worth comparing with previous results - see Table 5.

Generally, the linear regression equations are consistent with regard to the effect of the individual variables. The exception is air/fuel ratio for Run 10 which unexpectedly appears to have an adverse effect of sulphur removal efficiency. An explanation for this result will be given when the results of the non linearity of the variables is discussed. It was found also that a better correlation for Run 10 was obtained with added water, rather than with its square as previously. Comments are made below on the reasons why this should be so.

Reasonably good consistency between the magnitude of the coefficients is seen for bed depth, Ca/S mole ratio, cyclone drain temperature, and to a lesser extent, bed temperature. Similarly, residual errors and % variation are comparable across the three runs.

Analysis of Non Linearity of Process Variables

An outline of the mechanism of deriving polynomial expressions for individual variables to improve the fit of results has been described earlier. Whilst the techniques used are as described, differencies were introduced into the analysis which required some re-working of the results for Runs 8 and 9.

One of the objectives of the statistical analysis is to derive an overall relationship between the % sulphur removal efficiency and the process variables identified as important. Obviously, this is desirable so that a single equation is available for further application. However, the precision of the equation is improved as the amount of data available

TABLE 5

LINEAR REGRESSION EQUATIONS TO PREDICT % SULPHUR REMOVAL EFFICIENCY RUNS 8, 9 and 10

VARIABLE	RUN 8	RUN 9	RUN 10
Bed Depth (cm)	0.15	0.17	0.18
Bed Temperature (°C)	-0.035	-0.073	-0.002
Air/Fuel Ratio (% Stoichiometric)	0.94	0.38	-0.19
Cyclone Drain Temperature (°C)	0.0065	0.017	0.015
Added Water (m^3/hr)	-0.011*	-0.011*	-0.018
Ca/S Mole Ratio	1.96	1.18	1.31
Hours with no limestone addition	-	-	-0.12
Constant	71.8	112.3	58.45
Residual Error	3.94	4.06	4.13
% Explained	86.8	64.5	74.1
Number of Results	408	471	192

* Square of variable used

for analysis increases. So from a number of standpoints it is advantageous to be able to combine the data from Runs 8, 9 and 10, provided that the three runs are statistically similar, i.e. the results are derived for essentially similar experiments in similar equipment.

Little in the way of changes and modifications were carried out between Run 8 and 9 so combining these results is reasonable. However, prior to Run 10, a massive redesign and rebuild of equipment was undertaken and it is therefore fair to question whether these data can be included with the others. This question was looked at very closely and it was decided on the basis of the results of the linear regression analysis that the new unit configuration as used for Run 10 behaved in a manner sufficiently similar to the previous unit to merit combining all the data together to produce one overall equation.

To achieve this, Run 10 results were treated as described, taking a single variable and investigating its relationship with the sulphur removal efficiency after compensating for the values of all the other variables away from their means. Also, the results for each variable were subdivided into groups within the range of the variable in order to improve accuracy.

For Run 8 and 9, the data had already been worked up to give a single equation, corrected to the mean values of the variables for each separate run. In this form, the data was not amenable to combination with Run 10 since the correct weighting factors had to be applied to the individual run means. Neither was it desirable to merge Run 10 data with the combined Run 8 and 9 equation. This method of data evaluation would mean that each time additional data became available the exercise of taking each run results separately before merging would be necessary.

This difficulty was easily overcome by correcting the results for each run separately to a standard set of values ascribed to each variable in the equation. It was convenient to take this as the design parameters used by Foster-Wheeler Energy Corporation as the basis for the San Benito demonstration unit. These are shown in Table 6.

The grouped data for Runs 8, 9 and 10, for which subsequent analyses are carried out are shown in Tables 7 to 9. This is the point at which this analysis departs from that conducted for Runs 8 and 9. There, the next step was to develop an optimum equation by correcting both equations to the means of Run 8 and 9 variables. Here, the next step will involve correcting to the Foster Wheeler design conditions.

TABLE 6

STANDARDISED OPERATING CONDITIONS

(FOSTER WHEELER DESIGN CRITERIA)

VARIABLE	DESIGN VALUE
Bed Depth (cm)	91.4
Bed Temperature (°C)	910
Air Fuel Ratio (% Stoichiometric)	22.5
Water Input (m^3/hr steam)	0
Ca/S Mole Ratio (% Stoichiometric)	1.0
Cyclone Drain Temperature (°C)	50
Time Without Stone Addition (hr)	0

TABLE 7

GROUPED DATA FOR POLYNOMIAL FIT ANALYSIS (UNCORRECTED) FOR RUN 8

BED DEPTH (cm)			BED TEMPERATURE (°C)			AIR/FUEL RATIO (% STOICHIOMETRIC)		
MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS
72.6	56.4	10	78.7	858	2	72.0	16.1	1
76.7	64.6	14	81.2	873.4	68	75.6	21.2	65
77.4	75.7	42	81.5	889	245	78.6	23.5	234
79.0	85.9	80	80.1	907	80	80.9	26.1	100
79.1	94.0	173	97.3	924	13	79.6	28.7	8
81.9	103.4	76						
83.0	112.9	13						

cont/....

TABLE 7 (Continued)

WATER INPUT (m^3/hr STEAM)			Ca/S MOLE RATIO			CYCLONE DRAIN TEMPERATURE (°C)		
MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS
82.8	5.7	141	77.9	0.7	62	79.1	80	17
80.7	14.3	162	80.4	1.4	250	76.6	141	13
75.9	24.9	44	82.7	2.3	77	77.6	228	42
71.6	34.1	31	83.1	3.4	16	81.5	309	68
64.9	41.9	21	82.3	4.6	2	80.7	386	117
62.1	50.1	1	77.3	5.3	1	80.5	453	126
34.4	68.2	3				80.4	522	25
27.7	71.5	5						

TABLE 8

GROUPED DATA FOR POLYNOMIAL FIT ANALYSIS (UNCORRECTED) FOR RUN 9

BED DEPTH (cm)			BED TEMPERATURE (°C)			AIR/FUEL RATIO (% STOICHIOMETRIC)		
MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS
68.0	58	2	77.1	871	11	71.2	17.5	7
71.1	66	8	77.9	893	70	74.6	20.3	14
71.9	75.3	27	77.1	909	175	77.3	23.8	77
78.0	84.9	51	75.8	929	97	77.2	36.4	221
75.5	95.2	74	74.0	948	84	78.8	29.6	113
77.7	104.5	111	71.9	968	28	80.2	31.8	29
80.6	114.5	171	66.6	989	6	77.0	35.1	9

cont/.....

TABLE 8 (Continued)

WATER INPUT (m_3/hr STEAM)			Ca/S MOLE RATIO			CYCLONE DRAIN TEMPERATURE (°C)		
MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS
80.1	0.7	373	75.5	0.5	202	69.6	84	13
75.9	14.4	43	77.7	1.4	197	73.8	28	13
73.2	23.6	48	77.3	2.2	45	74.5	222	74
71.1	35.2	6	79.2	3.4	19	76.0	303	109
			81.1	4.2	5	78.0	375	96
			77.0	5.1	3	78.9	463	74
						79.3	533	52
						80.8	611	25

TABLE 9

GROUPED DATA FOR POLYNOMIAL FIT ANALYSIS (UNCORRECTED) FOR RUN 10

BED DEPTH (cm)			BED TEMPERATURE (°C)			AIR/FUEL RATIO (% STOICHIOMETRIC)		
MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS
71.7	93.6	26	76.1	369	5	73.4	20.0	22
71.9	99.7	42	74.1	891	27	75.8	22.5	91
75.6	106.9	31	75.2	908	60	75.0	25.6	48
75.2	112.0	29	75.1	927	56	73.1	27.7	28
77.0	118.9	21	75.2	951	35	74.3	31.1	2
77.5	124.3	30	72.5	962	9	70.5	34.1	1
79.9	130.0	10						
78.9	134.3	3						

cont/....

TABLE 9 (Continued)

WATER INPUT (m^3/hr STEAM)			Ca/S MOLE RATIO			CYCLONE DRAIN TEMPERATURE (°C)			HOURS WITHOUT STONE MAKE-UP		
MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS	MEAN SRE %	MEAN OF VARIABLE	NUMBER OF DATA POINTS
75.3	2.0	175	73.9	0.2	136	72.5	89	11	76.4	0.6	129
73.7	10.3	8	76.3	1.5	19	74.8	142	26	75.7	10.4	9
71.2	25.6	3	76.8	2.4	25	74.0	217	56	72.9	18.0	7
71.9	37.3	1	81.6	3.4	8	74.1	301	37	72.9	25.0	7
67.1	45.9	4	75.8	4.4	2	76.3	380	35	72.7	32.0	7
66.4	51.4	1	77.5	5.2	1	77.4	446	23	71.9	39.0	7
			81.1	6.6	1	76.3	521	4	70.4	46.0	7
									71.5	53.0	7
									70.0	60.0	7
									67.4	66.0	5

The % sulphur removal efficiencies shown for the grouped data for one variable for one run are corrected for the deviation of the mean value of the variable away from the design condition. This is repeated for other variables for the run, and for appropriate results from the other runs. The method of calculation, and the correction factors are shown in Table 10.

The corrected results were subjected to examination for linearity using the POLYNOMINAL-FIT programme and optimum equations derived for the three runs separately. Some attempts were made to reconcile the equations for the three sets of data for each variable in the sense that polynomials of the same order could be used to describe the data fit. This meant that in one or two instances, a small sacrifice in precision was made to ensure a measure of homogeneity between the result sets. In all instances, the loss of precision was small.

Finally, a weighted mean average equation was produced which best described all the data.

The equations are given in Table 11, and the resulting curves are illustrated in Figs. 2-8. It will be noted that the curves for the three runs show considerable divergence compared with the Run 8 and 9 results previously reported (Ref. 3). This is the consequence of correcting the results to standard conditions away from the common means of the run results.

An explanation is now possible for the apparently anomalous coefficient for the effect of air/fuel ratio seen for Run 10 during the linear regression stage. Reference to Figure 2 shows that the curves of % sulphur removal efficiency vs air/fuel ratio pass through a maximum and that the negative slope of the linear coefficient for Run 10 arises simply because the data in this case lies generally to the right of the maximum. It is nevertheless apparent from the curves that in actual fact the three runs were very similar. Also with regard to this figure, it has already been noted that an air leak was found during Run 10 which may have led to the air/fuel ratios being approximately 1% too lean. It may be expected therefore that in reality, the Run 8 and 9 curves should be displaced somewhat towards the lower air/fuel region and it is interesting to note that this brings the three runs even closer together.

The optimum results for the "Added Water" curves show Runs 9 and 10 with linear relationships. This arises from the comparatively restricted range covered by the variable in comparison with Run 8.

TABLE 10

CORRECTIONS TO % SULPHUR REMOVAL EFFICIENCY VALUES. RUNS 8, 9 and 10

VARIABLE	CORRECTION FACTOR*		
	RUN 8	RUN 9	RUN 10
Bed Depth (cm)	-3.37	-2.0	-3.4
Bed Temperature (°C)	0	+0.8	0
Air/Fuel Ratio (% Stoichiometric)	+0.3	-1.7	+0.3
Added Water Vapour (m^3/hr)	+0.7	+1.1	+0.7
Ca/S Mole Ratio	+0.2	-0.3	+0.2
Cyclone Drain Temperature °C	-3.39	-5.2	-3.4
Hours without Stone Addition	-	-	+1.5

* Correction = Coefficient for variable (design condition - run mean)

TABLE 11

POLYNOMIAL EQUATIONS FOR RUNS 8, 9 and 10

	RUN 8			RUN 9	
	x	x^2	x^3	x	x^2
Bed Depth (cm)	0.15			0.17	
Bed Temperature (°C)	3.25	-18.41×10^{-4}		1.65	-9.32×10^{-4}
Air/Fuel Ratio (% Stoichiometric)	-26.75	1.28	-1.95×10^{-2}	1.84	-2.78×10^{-2}
Added Water (m^3/hr)	7.62×10^{-3}	-1.06×10^{-2}		-0.29	
Ca/S Mole Ratio	5.75	-0.92		2.72	-0.39
Cyclone Drain Temperature (°C)	0.67×10^{-2}			1.72×10^{-2}	
Hours without Stone Make-Up					
Constant		-1200.5		-704.8	
Standard Error		3.85		3.87	
% Explained		87.3		67.4	

cont/....

TABLE 11 (Continued)

	RUN 10		AVERAGE FOR ALL RUNS			
	x	x^2	x	x^2	x^3	x^4
Bed Depth (cm)	0.18		0.16			
Bed Temperature (°C)	0.75	-4.07×10^{-4}	1.48	-8.38×10^{-4}		
Air/Fuel Ratio (% Stoichiometric)	3.52	-7.48×10^{-2}	3.69	-6.86×10^{-2}		
Added Water (m^3/hr)	-0.18		-0.21	-1.33×10^{-2}	4.75×10^{-4}	-5.45×10^{-6}
Ca/S Mole Ratio	2.28		3.79	-0.56		
Cyclone Drain Temperature (°C)	1.4×10^{-2}		1.3×10^{-2}			
Hours without Stone Make-Up	-0.12		-0.12			
Constant	-328.7					
Standard Error	3.90					
% Explained	76.2					

% SRE vs BED DEPTH (cm)

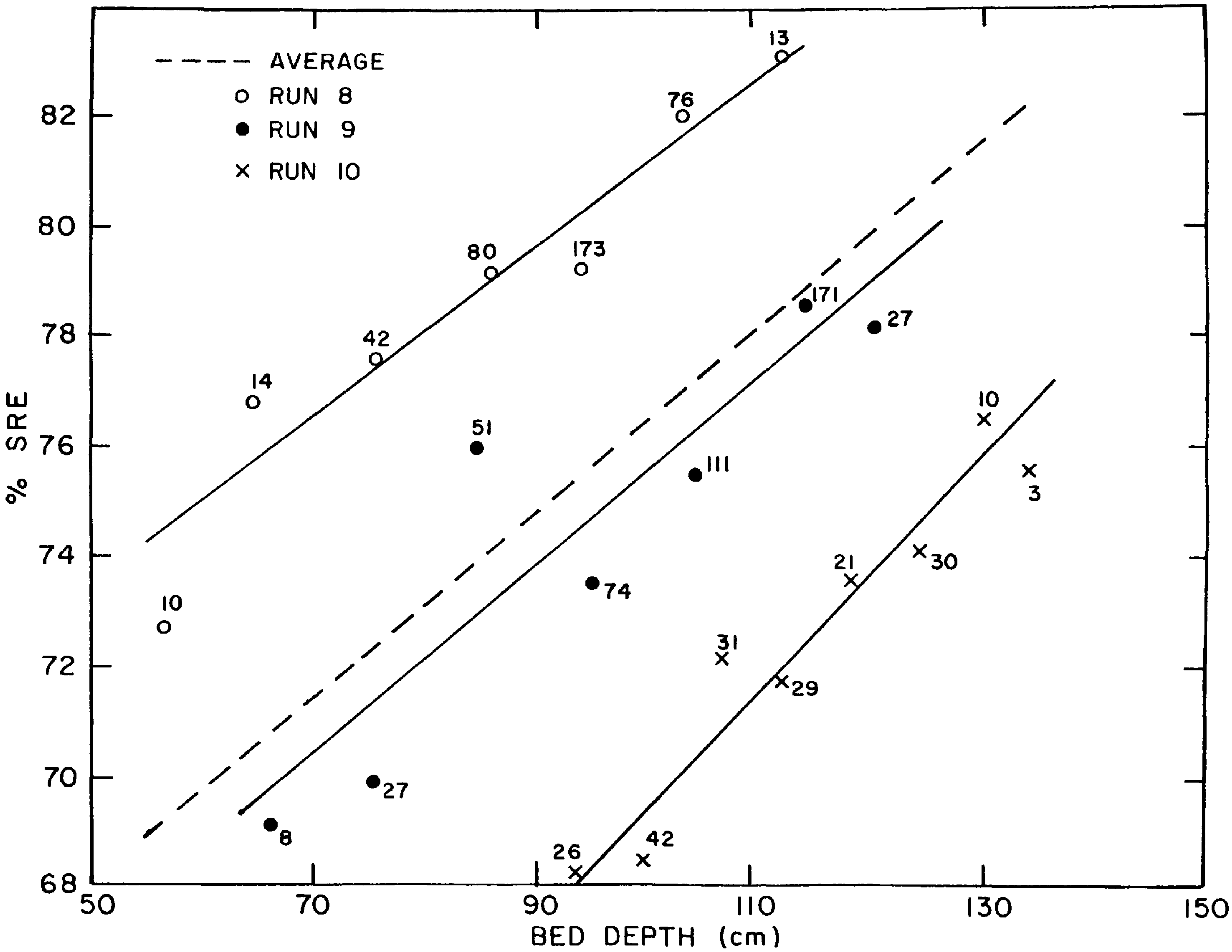


FIG. 2.

% SRE vs GASIFIER BED TEMPERATURE

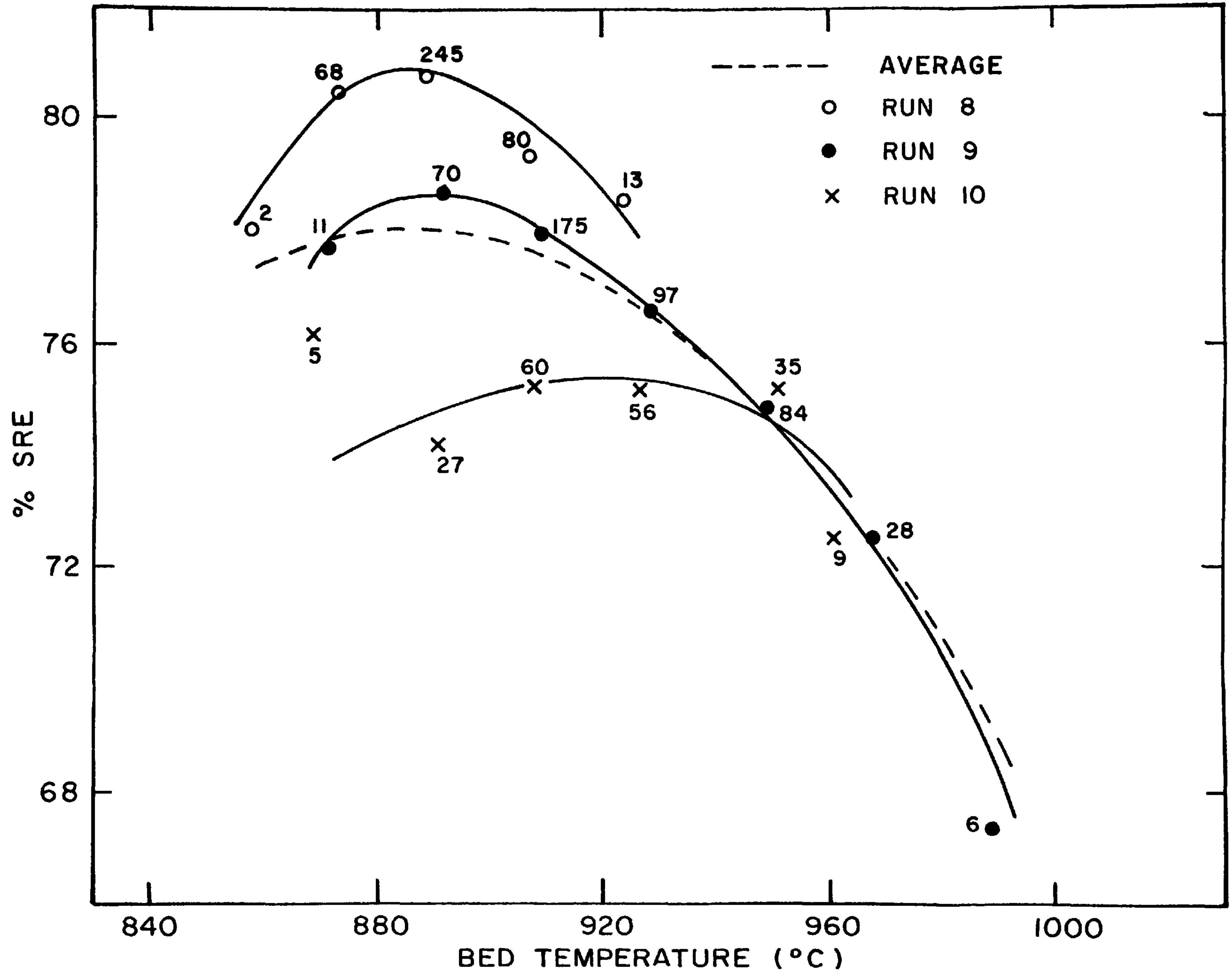


FIG. 3.

% SRE vs AIR/FUEL RATIO

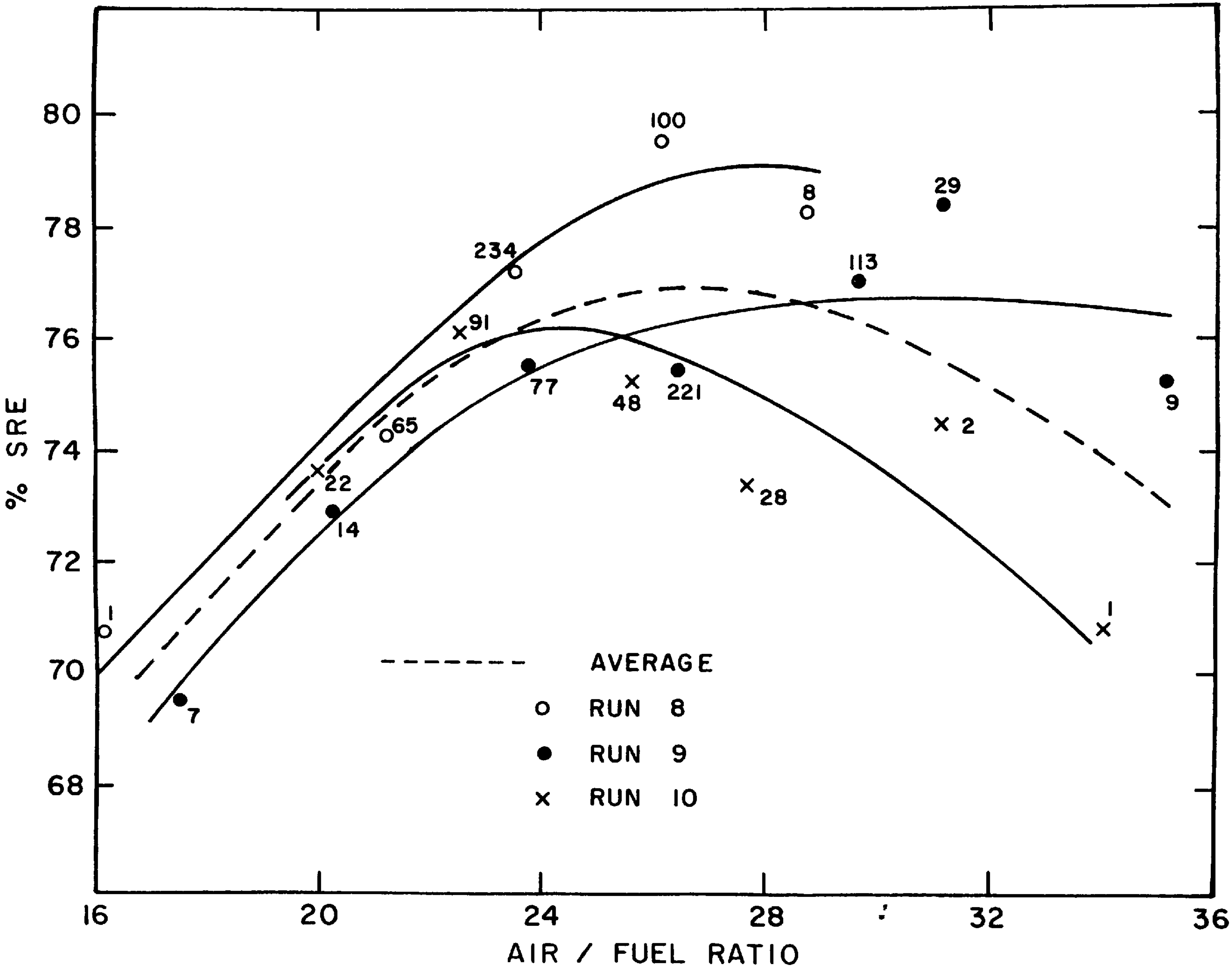


FIG. 4.

% SRE vs ADDED WATER VAPOUR (M^3/HR

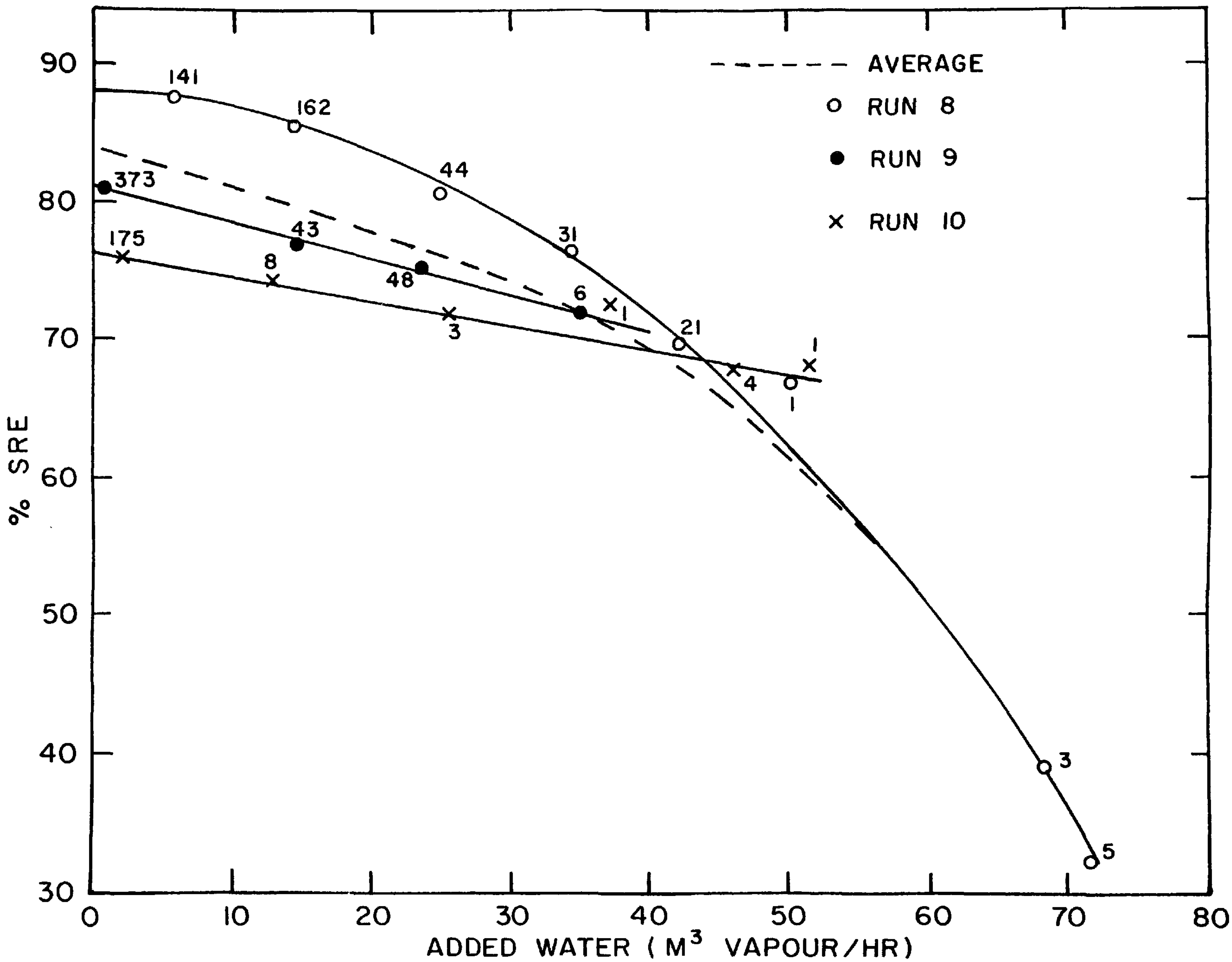


FIG. 5.

% SRE vs Ca/S MOLE RATIO

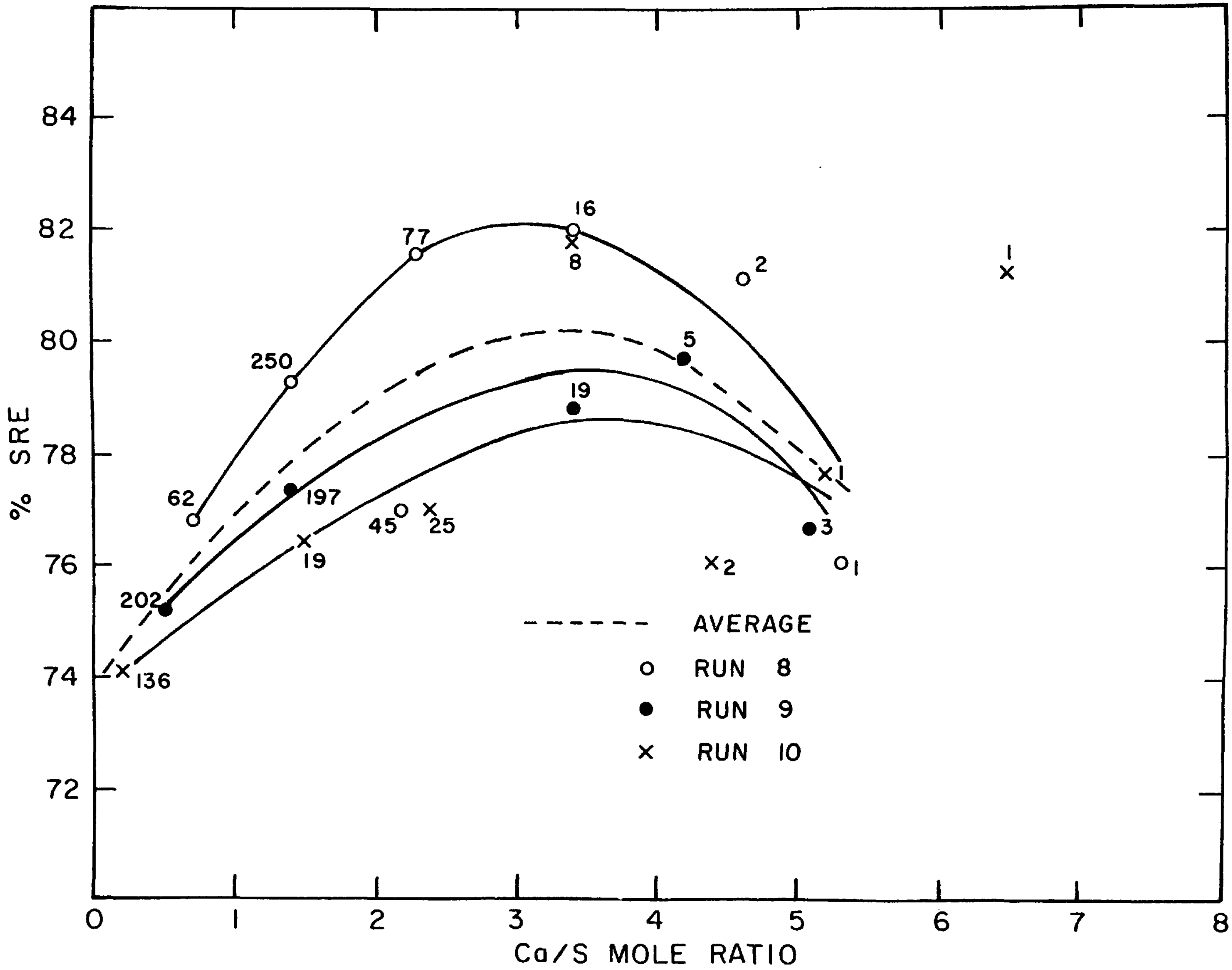


FIG. 6.

% SRE vs CYCLONE DRAIN TEMPERATURE (°C)

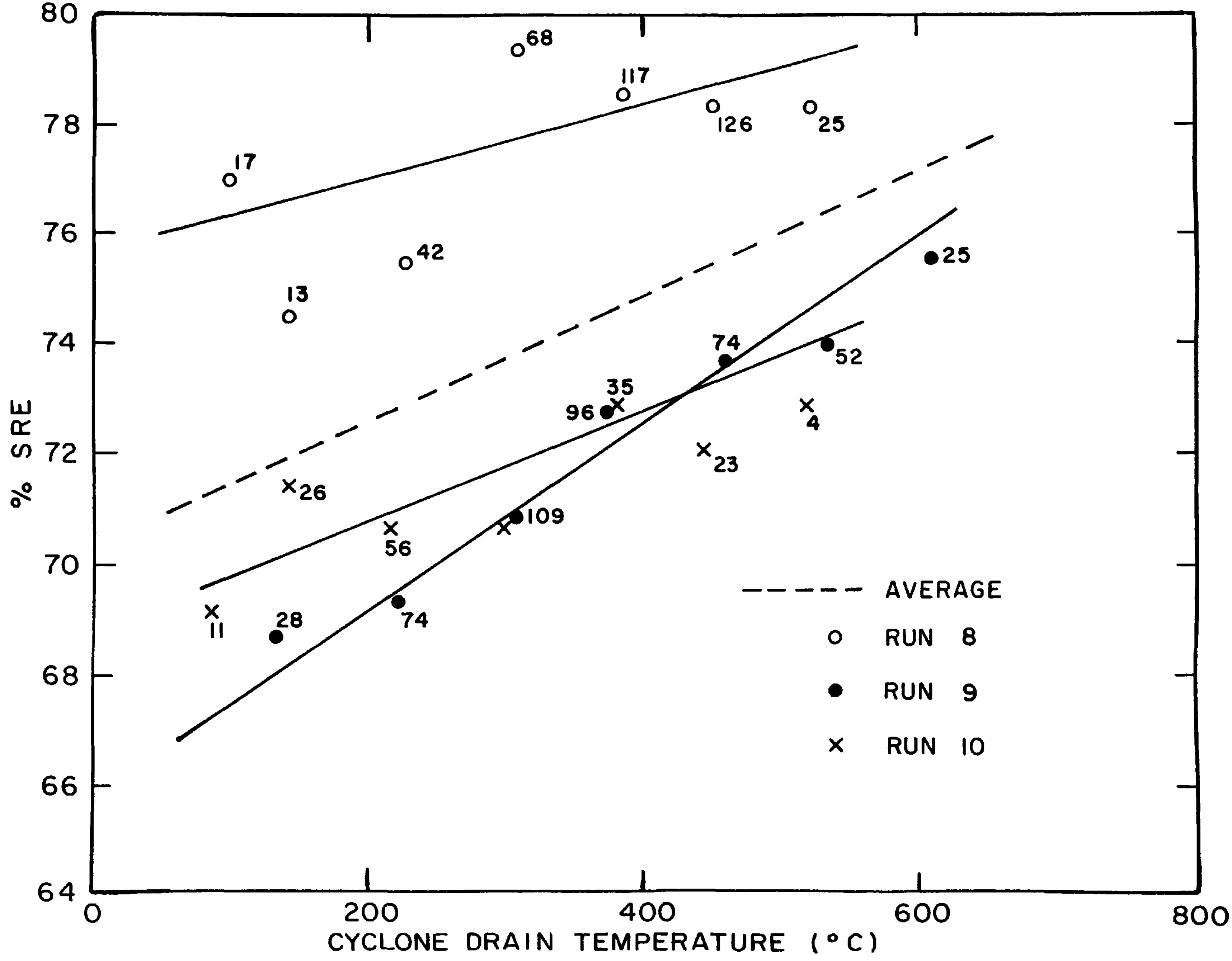


FIG. 7.

% SRE vs HOURS WITHOUT STONE ADDITION

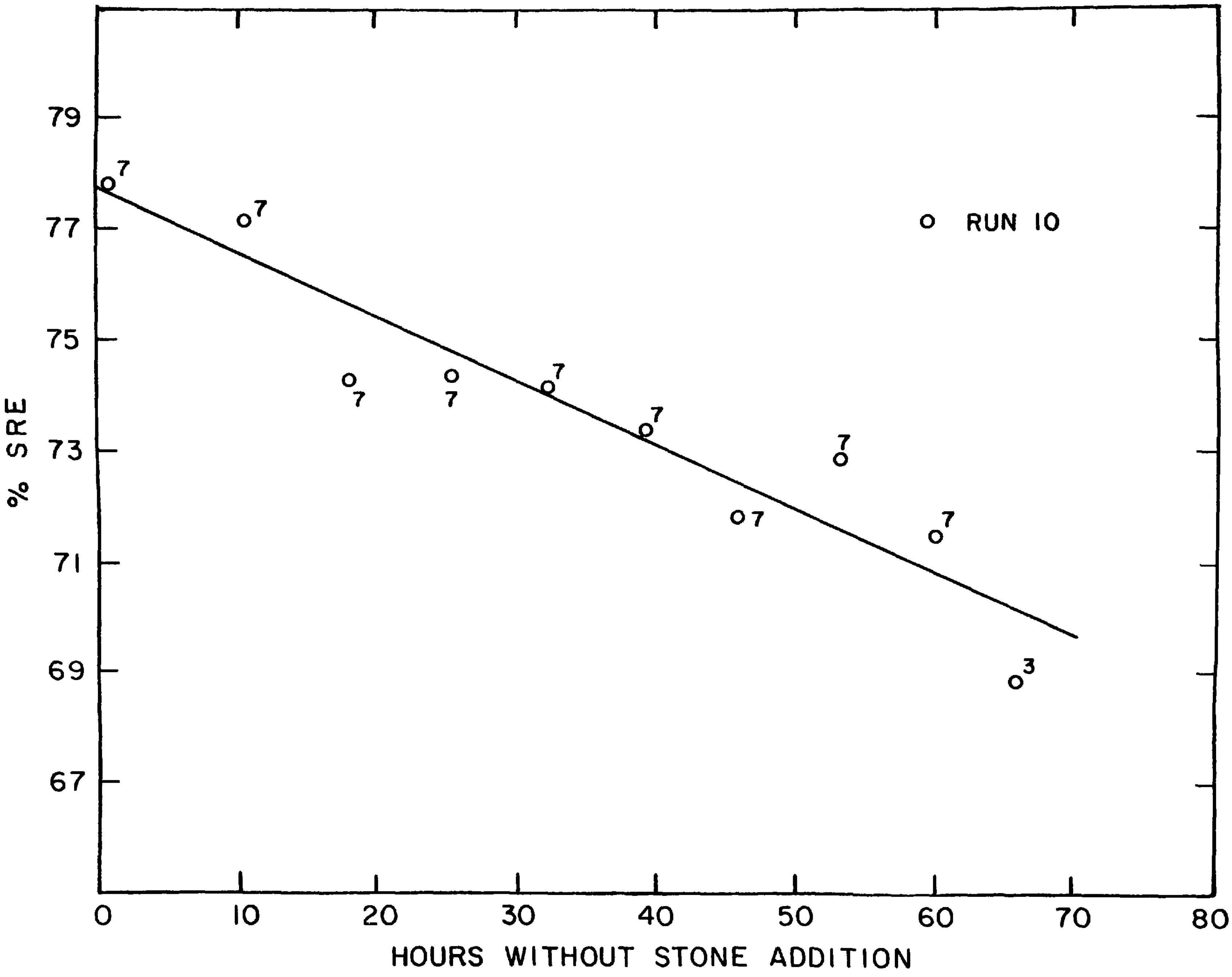


FIG. 8.

The overall equation was applied to the individual run data. The results are compared to the optimum polynomial equations and the initial linear regression equations applied to their respective individual runs in Table 12.

As would be expected there is a loss of precision when applying the average equation to the individual runs compared to when the run optimum equation is used. However, this loss of precision is comparatively minor compared to the advantage of having available a single predictor for future heavy fuel oil runs.

Evaluation of Other Run 10 Variables

After allowing for the contribution of the significant variables, a random error, attributable to experimental error, remains associated with the differences between measured and predicted sulphur removal efficiency. Part of this error may be associated with variables not included in the regression equations, and a number of potential process parameters were investigated for runs 8 and 9.

This evaluation has been extended to include the results from Run 10 and additional variables from the three runs. The analysis was concluded using the residual error data, and again the results were averaged into boxes across the range of the variable.

Results are presented in Figures 8 to 13 and it is readily apparent that there is no systematic trend which would indicate a correlation with the residual error.

It was thus confirmed that bed sulphur, carbon, sulphate and fines (surface area) levels showed no strong effect on sulphur removal performance. A similar conclusion may be drawn for air rate and fuel rate into the gasifier.

Application of the General Equation to Bitumen

Towards the end of Run 10, a period of operation on Bitumen was successfully completed showing that both gasification and regeneration could be carried out continuously with fuels significantly different in chemical composition and physical characteristics from the normal heavy fuel oil feed. Good desulphurisation was generally observed throughout.

TABLE 12

OVERALL EQUATION APPLIED TO INDIVIDUAL RUNS

EQUATION	RUN 8	RUN 9	RUN 10
OVERALL			
Standard Error	4.41	4.14	4.66
% Explained	83.5	63	66
Constant*	-642.7	-644.6	-645.9
INDIVIDUAL			
Standard Error	3.85	3.87	3.90
% Explained	87.3	67.4	76.2
Constant	-1200.5	-704.8	-328.7
LINEAR			
Standard Error	3.94	4.06	4.13
% Explained	86.8	64.5	74.1
Constant	71.8	64.5	58.4

* Note, weighted mean value of constant = -645.0

% SRE (RESIDUAL ERROR) vs BED CARBON (WT%)

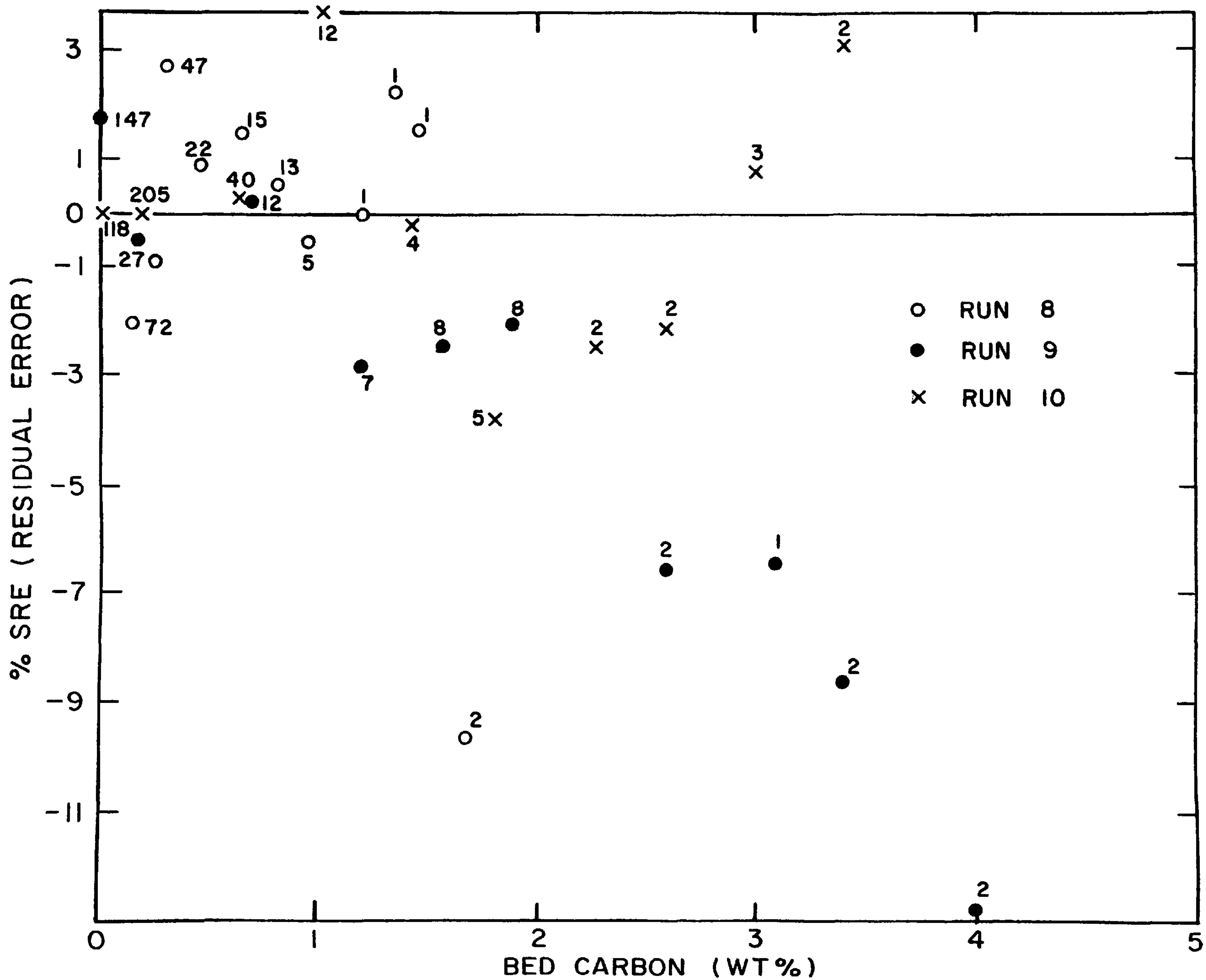


FIG. 9.

% SRE(RESIDUAL ERROR) vs BED SULPHUR (WT %)

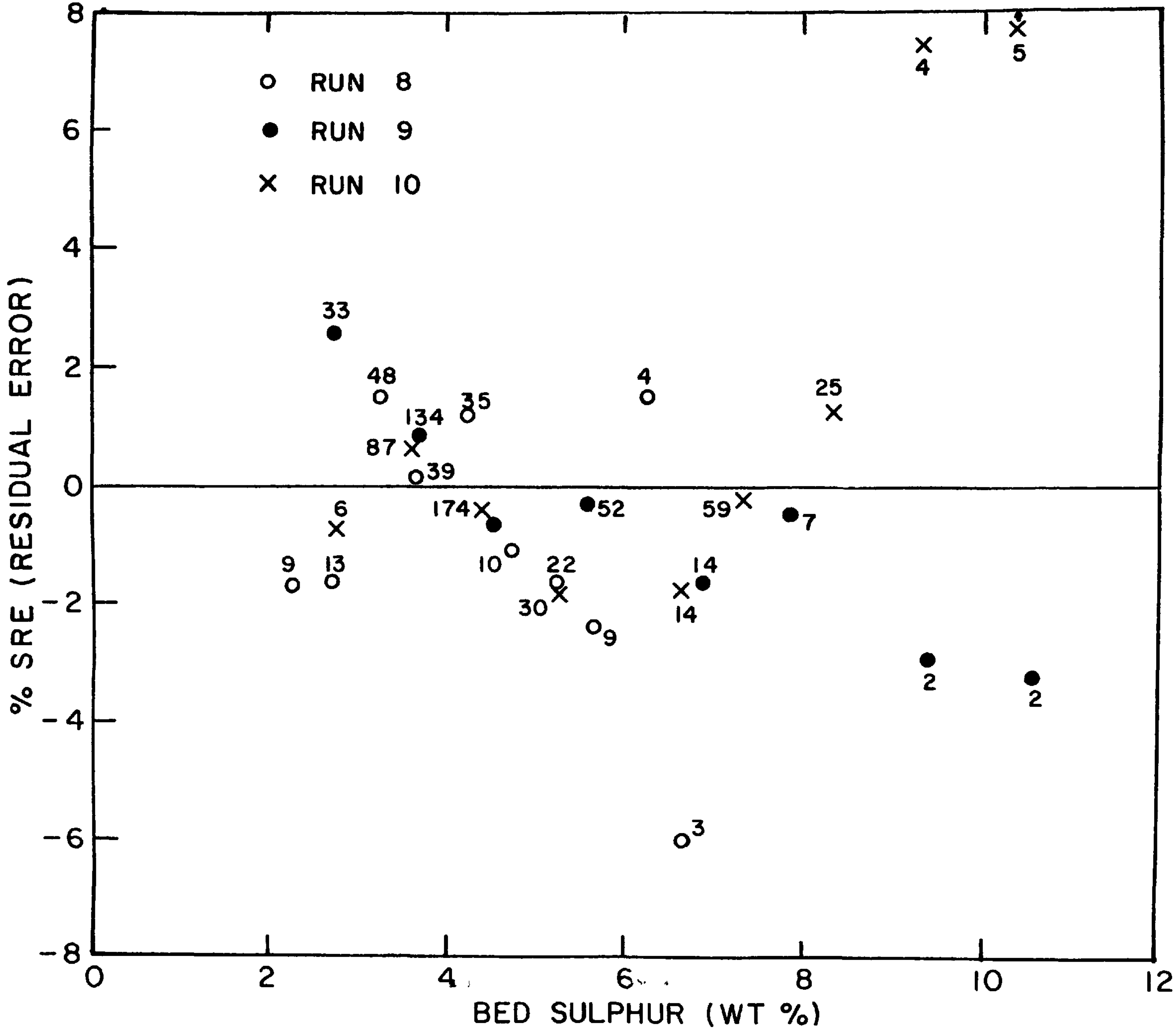


FIG. 10.

% SRE (RESIDUAL ERROR) vs BED SULPHATE (WT %)

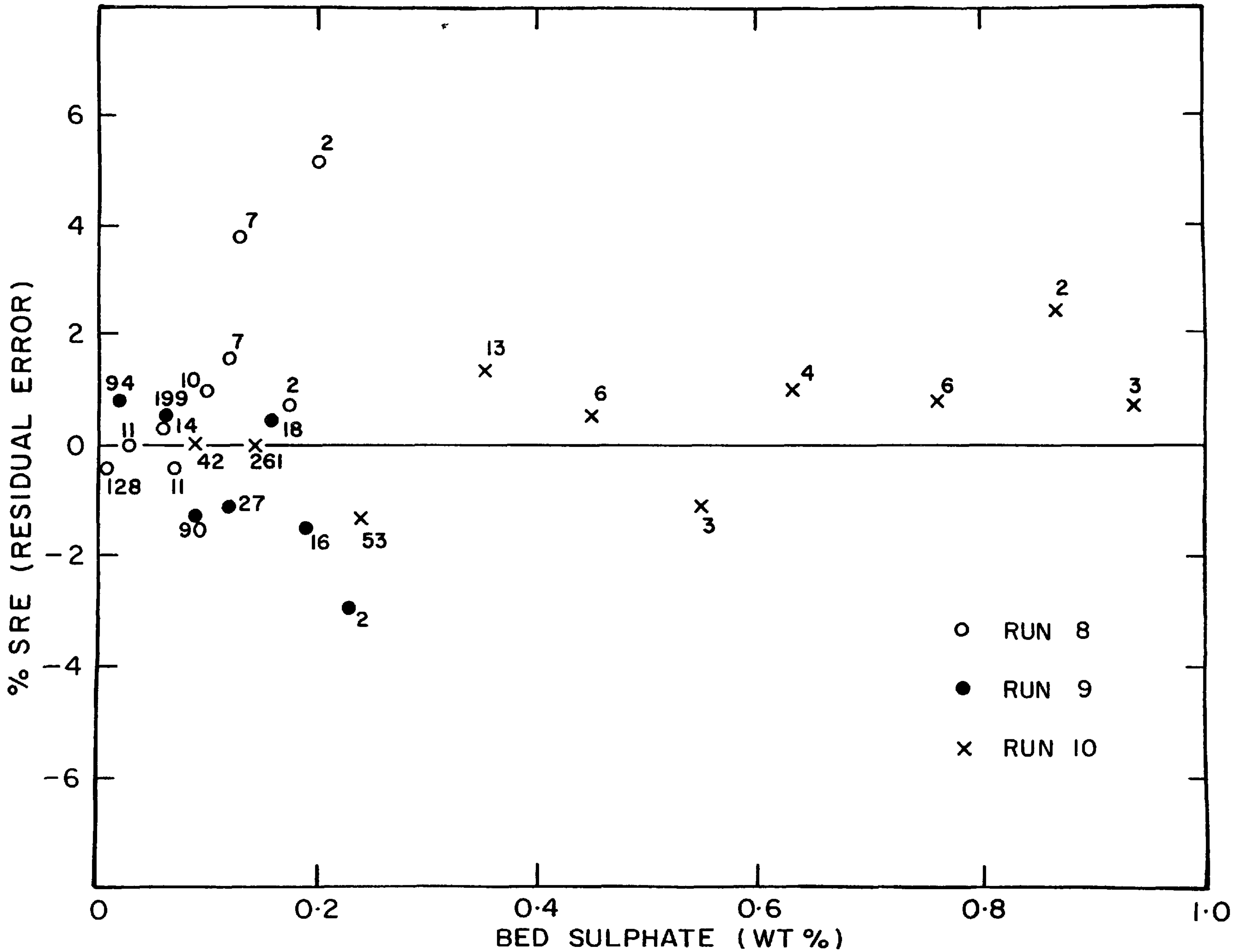


FIG. 11.

% SRE (RESIDUAL ERROR) vs % BED FINES (−600μ)

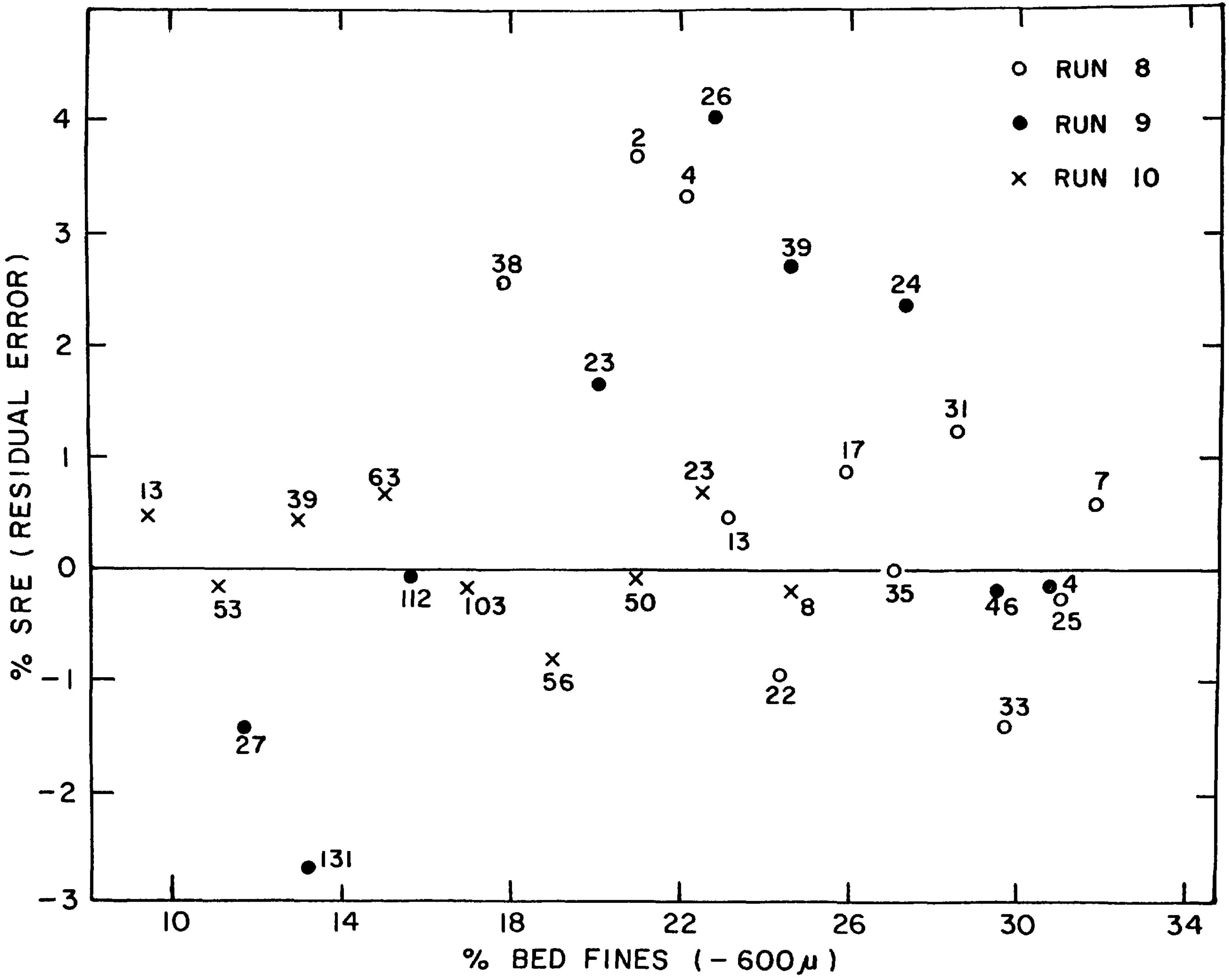


FIG. 12.

% SRE RESIDUAL ERROR) vs GASIFIER AIR RATE (m³/hr)

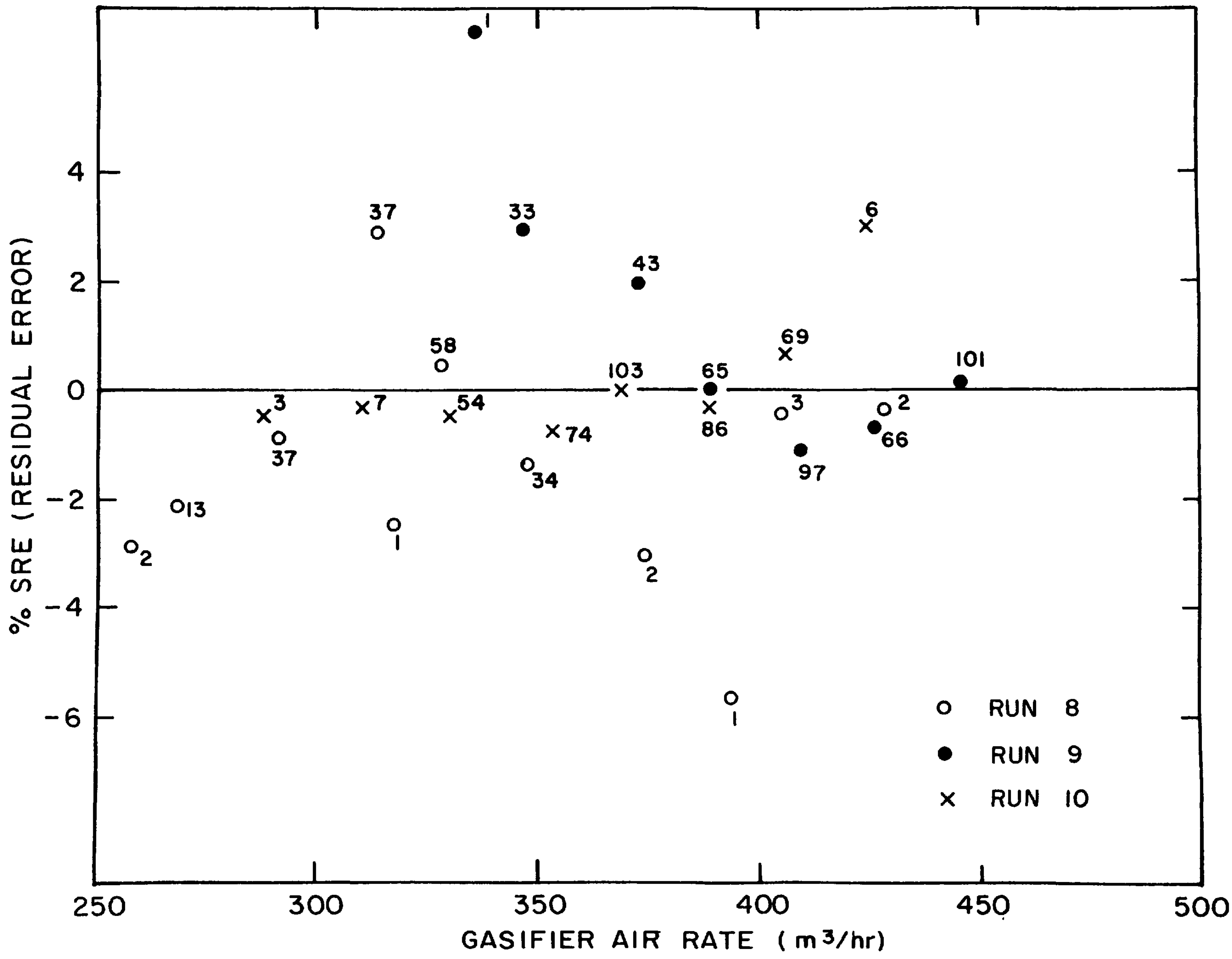


FIG. 13.

% SRE (RESIDUAL ERROR) vs FUEL RATE (Kg/Hr)

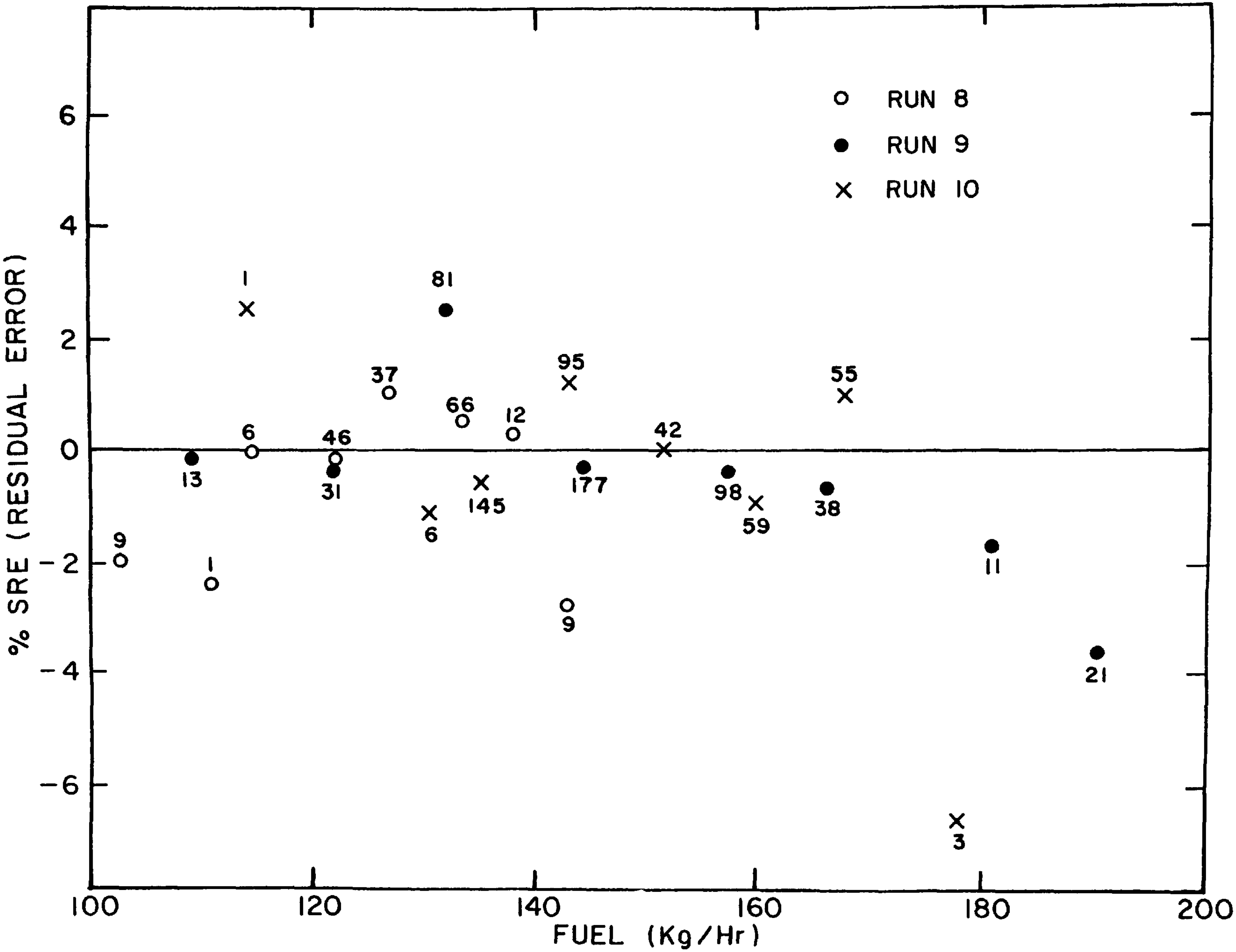


FIG. 14.

A total of 37 hours operation was achieved with complete data available for seventeen hours. However, due to a major blockage in an inadequately heated filter, and the very coarse pump flow indication, it was not possible to obtain a direct measurement of the fuel flow rate throughout this period. Fortunately, this state of affairs was known immediately prior to changing to the bitumen fuel and the conditions already established for heavy fuel oil in the system could be used as a basis for back calculating the vacuum bottoms fuel flow from the stack gas analysis.

The results of the application of the average equation developed for heavy fuel oil to bitumen are summarised in Table 13. Overall, excellent prediction is achieved by the correlation equation, with the average predicted sulphur removal being only 0.6% lower than the measured value. However, a trend can be observed that the equation initially overpredicts performance by approximately 10% and later consistently underpredicts by approximately 3%.

The initial results would not normally be included in the analysis since these represent the first results immediately after restart from a sulphated bed and thus are untypical of normal performance, certainly over the first three hours whilst a reasonable sulphur inventory is accumulated on the lime bed. Thus, it would appear that the equation overpredicts the sulphur removal efficiency in general for this fuel. However, it must be borne in mind that additional uncertanities have been introduced into these results by having to calculate the fuel flow and hence the sulphur input into the system, from the stack gas analysis. Under these circumstances, the results are accepted as indicating that the performance of the pilot plant on oil based fuels other than heavy fuel oil can be predicted reasonably accurately using the fuel oil equation.

Application of the General Equation to Runs 6 and 7 Data

It was intended at the outset that the equation developed would be applied to Runs 6 and 7 to see whether an improvement in the prediction of % sulphur removal efficiency, particularly for Run 6 would occur. Unfortunately, this could not be carried out using the individual hourly data sets as had been intended originally as it was found that the necessary data files had been erased from the computer memory and a very considerable time would be needed to re-insert the original information.

TABLE 13

APPLICATION OF AVERAGE REGRESSION EQUATION TO TJ 102

MEDIUM VACUUM BOTTOMS (BITUMEN)

Time (D.H.)	Measured % SRE	Predicted % SRE	Difference %
20.2330	78.7	67.9	10.8
21.0030	76.7	66.6	10.1
21.0130	71.9	66.0	5.9
21.0230	69.2	65.2	4.0
21.0330	65.7	62.5	3.2
21.0430	62.0	64.4	-2.4
21.0530	59.9	64.3	-4.4
21.0630	59.4	62.4	-3.0
21.0730	59.5	63.1	-3.6
21.0830	59.3	66.7	-7.4
21.0930	63.1	67.4	-4.3
21.1530	58.6	63.1	-4.5
21.1630	62.0	65.6	-3.6
21.1730	63.7	64.0	-0.3
21.1820	64.0	64.6	-0.6
21.1930	60.9	64.7	-3.8
21.2030	58.8	64.8	-6.0
MEAN	64.3	64.9	-0.6

Thus, whilst a complete re-working of the data for the earlier runs using the individual hourly results is not possible at this time, an assessment of some of the data is possible since averaged results over 10 hour periods have been presented previously (Ref. 2). These were examined originally in the light of the regression equation developed for Runs 8 and 9 only, when a reasonably good prediction of the Run 7 results were obtained but the results for Run 6 were badly underpredicted (Ref. 3). It was necessary to take into account an estimate of the cyclone drain temperature variable in order to carry out this analysis and this has been included here also. The cyclone temperature value used as a first approximation is the weighted mean of the average value observed during Runs 8, 9 and 10, viz. 346°C.

The time variable which was established during Run 10 is of no consequence for Runs 6 and 7 since fresh stone was being added continuously throughout the periods for which the grouped data are available.

A correction can be made for the water added via the flue gas recycle during Runs 6 and 7 but this is relatively small, ranging between 0 and approximately -1.1% on the predicted sulphur removal efficiency.

Taking these corrections into account, predicted results for Runs 6 and 7 are shown in Table 14. A reasonably good prediction of the Run 7 results is observed, but by comparison, Run 6 results are underpredicted on average by approximately 6%. This is a very considerable improvement over the previous results (Ref 3) in which the difference values for Runs 6 and 7 were 13.8% and 3.2% respectively.

Results of Experiments to Gasify Solid Fuel

Following the period of gasification on the bitumen the final experiment planned for Run 10 was to examine the feasibility of coal gasification. The basic objectives were to establish that coal could be fed successfully into the gasifier and to establish whether the resulting gas quality would enable a flame to be maintained in the boiler. Illinois No. 6 was used for this experiment.

A very simple coal feed system consisting of a manually replenished lock hopper was used with the coal being injected pneumatically via the flue gas recycle tuyere inserted through the warm-up burner assembly. In the event, the coal feed rate proved to be very erratic and the coal tended to be delivered into the gasifier at extremely high rates on occasions, resulting in the emission of smoke from the exit

TABLE 14

PREDICTION OF AVERAGED % SULPHUR REMOVAL EFFICIENCY FOR RUNS 6 AND 7 USING OVERALL RUN 8, 9 AND 10 REGRESSION EQUATION

Run No.	Time of First Reading (D.H.)	Actual % SRE	Predicted* % SRE	Difference
6	2.0830	75.5	71.3	4.2
	3.0430	80.0	73.5	6.5
	6.2230	80.0	73.5	6.5
	8.0430	71.5	68.8	2.7
	9.0030	71.5	70.2	1.3
	11.0630	84.0	76.0	8.0
	12.2030	82.0	72.7	9.3
	15.1330	82.0	72.8	9.2
	16.1930	71.5	71.1	0.4
	19.1730	78.5	70.1	7.4
MEAN		77.7	72.0	5.7
7	4.0630	77.5	72.9	4.6
	5.0230	80.0	73.3	6.7
	6.1830	67.5	70.7	-3.2
	7.1430	67.5	71.9	-4.4
	9.2130	70.0	72.8	-2.8
	11.0930	78.0	75.2	2.8
	13.0030	81.0	77.9	3.1
	13.1530	80.0	73.5	6.5
	14.1630	77.0	73.2	3.8
MEAN		75.4	73.5	1.9

* Corrected for cyclone temperature and flue gas recycle water variables

stack. A variety of coal particle size ranges were used, viz. below 800 μ, below 1400 μ and 1400 μ up to 3,200 μ (1/8 inch). Similar feed control difficulties were observed for all size ranges.

The coal feed was initiated and simultaneously the bitumen feed rate was cut down to compensate and to maintain approximately similar gasifier stoichiometry to the operation on bitumen alone. The operation was repeated, successively reducing the proportion of bitumen until it represented only approximately 50% of the normal rate. At this point, it was estimated that an average rate of 100 kg/hr of Illinois No. 6 was being supplied to the gasifier.

A flame was maintained throughout in the boiler but there were considerable fluctuations in operating conditions, particularly gasifier temperature, boiler oxygen, and SO_2 levels. No attempt was made to initiate and maintain regenerator performance.

The conclusions drawn from this experiment were that coal gasification could be achieved and that the gas quality would be adequate to maintain combustion in the boiler. However, it was considered essential that a proper coal feed system be developed in order to adequately control the feed rate before it would be possible to operate without any supplementary liquid fuel.

Experiments on Location of Fuel Injection

A number of experiments were planned on injection of heavy fuel oil at different locations in the gasifier bed, to establish whether a single fuel injector would provide the necessary fuel distribution in the bed, and whether the re-designed distributor incorporating a central pit would eliminate tip burning of fuel oil injectors and would allow injectors to be changed whilst maintaining gasification. Two separate access points into the distributor pit were available, one being a hole through the gasifier wall and through the pit wall into the pit and the second having a "V" channel through the distributor pit wall.

Early operation during the run was carried out with fuel oil being injected at two locations on opposite sides of the gasifier, 6.5 cms above the distributor. On day 15 at 03.30 injection via one pit injector was initiated and the side injectors taken out of service completely. The effect on sulphur removal efficiency and other operating parameters is shown in Table 15. Within experimental error, no change

Table 15

EFFECT OF FUEL INJECTOR POSITION

ON SULPHUR REMOVAL EFFICIENCY

Fuel Injection Location	Time D.H.	% SRE	Gasifier Bed Temperature °C
Side, two	14.2330	71.7	921
injectors	15.0030	71.0	924
	15.0130	70.4	925
	15.0230	70.5	932
Distributor Pit,	15.0330	69.8	937
one injector	15.0430	71.3	920
	15.0530	71.9	928
	15.0630	71.3	921

in performance could be considered to have occurred, and the remainder of the run was completed with a single injection point into the distributor pit.

Removal and reinsertion of the pit injectors while gasifying was found to be easy, both via the hole through the refractory wall and distributor, and also through the "V" channel in the distributor.

Experiments on Injection of Steam

Steam was injected into the gasifier on two occasions. Results from the first period were rejected because of difficulties with the steam metering equipment. The second period occurred between D.H. 16.0030 and 16.1030 when approximately 40 m^3/hour of steam was injected in two separate periods. Flue gas recycle rate was reduced to compensate for the steam injection. Average results were tabulated in Table 16.

It is readily apparent that the addition of steam has a deleterious effect on the sulphur removal performance and that recovery is rapid when steam injection is stopped. Also it appears that the effect is more pronounced when running at low air fuel ratios i.e. rich, in the gasifier, and that under these conditions recovery when steam injection is stopped is not as rapid.

Experiments on Burn Back of Ducts and Cyclones

During the normal operation of the gasifier, carbon and condensed tars accumulate in the ducts and cyclones leading from the gasifier through the cyclones into the boiler. As a consequence, the gasifier top space pressure increases and removal of the deposits becomes necessary.

The normal procedure is to burn out the ducts from the entry in the gasifier through into the boiler and procedures are available for conducting this operation. Such a burn out was carried out successfully during run 10 on days 4-5.

Further tests were conducted during Run 10 to establish whether it would be possible to clear the ducts by a burn back procedure, initiating the combustion at the boiler burner end and burning the deposits out through the cyclones and through into the gasifier. This test was carried out on day 19.

Table 16

EFFECT OF STEAM INJECTION ON PERFORMANCE

Period (D.H.)	Average steam input (m^3/hr)	A/F Ratio % Stoichiometric	Average % SRE
15.1930 to 15.2330	9.4	26.7	64.8
16.0030 to 16.0430	37.7	25.5	58.5
16.0530 to 16.0630	0	23.2	70.4
16.0730 to 16.1030	40.2	22.1	57.3
16.1130 to 16.1430	0	22.0	63.4

The initial part of the burn out through the ducts as far as the cyclones was completed successfully. However, the gas flow through the cyclones in the reverse direction was unable to attack the deposits in the cyclone itself and consequently this approach had to be considered to be unsuccessful. The test was terminated and the normal burn out procedure instituted and the ducts and cyclones cleared satisfactorily.

Attempts were made also to prevent or limit the deposition of carbon and tars in the ducts by providing air jets adjacent to the walls at the entry to one of the main gasifier cyclones. It was expected that a small flow of air along the wall would oxidise any deposits which tended to form. In the event, these jets proved to be completely unsuccessful and did not prevent the accumulation of deposits, and the tests were terminated during the early part of the run before day 14.

Experiments on Tuyere Injection of Flue Gas

Provision is available to use flue gas recycle through the gasifier plenum in order to control bed temperature. The flue gas recycle stream has to be cleaned in order to remove small quantities of particulates, and also provision has to be made for continuous recirculation of flue gas bypassing the gasifier plenum in order to prevent condensation in the flue gas recycle system. These precautions are necessary to avoid blocking the air nozzles in the gasifier distributor with damp lime particulates.

In would be an obvious advantage if flue gas could be injected directly into the gasifier bed as this would eliminate the need for clean-up. During Run 10, flue gas was injected directly into the gasifier bed via a tuyere inserted through the warm-up burner assembly. The experiments were carried out during days 20 and 21 whilst operating on bitumen. Whilst a direct comparison of tuyere versus plenum injection of flue gas was not possible due to the re-routing of pipework necessary to change the injection location, bed temperature control was excellent using the tuyere system and no adverse efects could be observed, for example, poor temperature distribution in the gasifier bed or increased fines carry over into the cyclones. It was concluded that the system performed equally efficiently as when plenum injection of flue gas was used but that the tuyere system offered the advantage of being less critical to particle and moisture loading of the flue gas.

Cooling Rate of Gasifier After Shut-down

With the improved insulation of the new unit, a check was made to establish the cooling rate of the gasifier bed at the end of Run 10.

Initially the gasifier bed was at 880°C with a bed depth of 91 cm (36 inch). It can be seen from Fig. 15 that it takes more than 24 hours for the bed to cool to 620°C which is the minimum temperature at which it can be reheated using fuel oil, assuming that it is in a sulphated state. Should the shut-down be made with a sulphided bed which retains its carbon coating on the lime, a restart from a lower temperature would be possible. This has not been investigated and therefore it is not recommended that the gasifier bed should be allowed to drop below 620°C if a reliable restart is desired.

However, the time available before a reheat is necessary is sufficient for safe overnight shut-down of the gasifier and thus operation on a daily basis, becomes possible. This offers advantages for commissioning equipment and for running short experiments using the continuous gasifier without requiring the setting up of shift operations. Continuous operation will however still be essential for long term testing under fully lined out conditions.

MATERIAL BALANCES

Results have been reported previously on the fate of trace elements present in the fuel oil feed to the CAFB gasifier during periods when fresh limestone was added continuously to the gasifier (Refs. 2, 3). The stable operation of the pilot unit during Run 10 between days 13 and 16 presented an opportunity to investigate bed attrition and trace element retention whilst no fresh limestone was being added, and enabled the effect of bed age to be taken into account. It is of obvious interest whether trace elements continue to be captured by the aged bed material or whether release into the atmosphere increases progressively with time. Capture of trace elements minimises pollution, reduces potential corrosion problems in the boiler due mainly to sodium and vanadium, and the efficient capture of certain elements present in high concentration, such as vanadium, could provide an economically attractive route to recovery of valuable material.

COOLING CURVE FOR GASIFIER BED AFTER SHUTDOWN

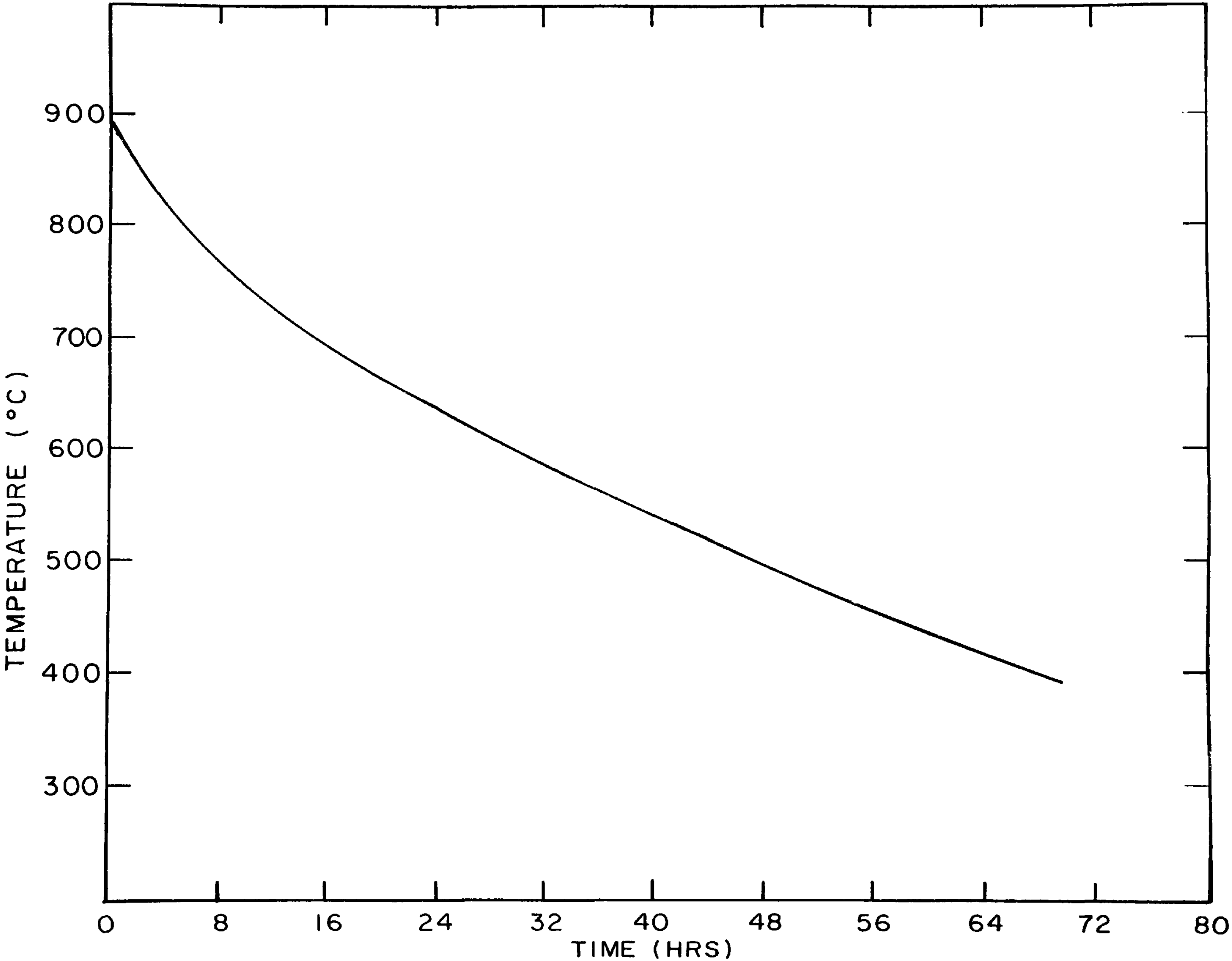


FIG. 15.

Samples of spent stone were removed from the pilot unit at six hourly intervals throughout the period when no fresh limestone could be added. Due to the difficulty and expense of analysing for trace elements, three sets of samples were selected for evaluation. These were at the beginning of the period, after 24 hours operation, and at the end after 68 hours without stone feed. Trace element levels on those samples were established using atomic absorbtion, in addition to the usual analysis for total sulphur, carbon, sulphate sulphur and acid insolubles. Preliminary runs were carried out to establish those elements for which the level present could be measured with reasonable precision.

Calculation of mass balances for individual trace elements present in heavy fuel oil is affected by two major factors. Firstly, there are the inaccuracies present due to the difficulty of measuring the trace element level in the fuel, the fresh limestone feed if present, and the stone samples extracted from the system. Secondly, the stone balance itself may not close, in which case the individual elemental balances cannot be established regardless of the precision of the analytical data.

Thus, before attempting trace element balances, checks were made to establish how well the stone and sulphur balances approached closure.

Limestone and Sulphur Balances

Reference to Table C-6, Appendix C gives the stone and sulphur balances on an hourly basis. These have been established from records of material added to and removed from the system, and in the case of sulphur, including analysis of gas streams from the boiler and regenerator and the changes in sulphur levels observed on the lime samples removed. They do not include the changes in the gross quantity of stone and sulphur present in the gasifier and regenerator arising from changes in bed depth. Thus, all that is needed to complete the stone and sulphur balances is to supplement the information provided in Table C-6, Appendix C, with the changes in the gasifier and regenerator stone and sulphur inventories. This information is given in Table 17 for the stone balance, and Table 18 for the sulphur balance. These are within the anticipated precision of the measurements of the stone and sulphur levels and sampling errors associated with stone removals and thus an evaluation of trace element balances will not be significantly affected by errors associated with the stone balance.

TABLE 17

STONE BALANCE FOR PERIOD D.H. 13.2330-16.1800

TIME (D.H.)		STONE* IN-OUT (kg)	GASIFIER + REGENERATOR BED INVENTORY (kg)
Start	13.2330	407.1	479
	14.2330	349.8	414.5
END	16.1800	249.2	325.5
GAIN (kg)			
	13.2330-14.2330	-57.3	-64.5
	13.2320-16.1800	-157.9	-153.5
NET GAIN (kg)			
	13.2330-14.2330	7.2	
	13.2330-16.1800	-4.4	

* From Appendix C, Table C-6

TABLE 18

SULPHUR BALANCE FOR PERIOD D.H. 13.2330-16.1800

TIME (D.H.)		SULPHUR* IN-OUT, kg	GASIFIER + REGENERATOR SULPHUR, kg: GASIFIER BED (kg)	GASIFIER SULPHUR (wt%)	REGENERATOR BED (kg)	REGENERATOR SULPHUR (wt%)	TOTAL BED SULPHUR (kg)
START	13.2330	2.602	417.3	2.88	61.6	2.03	13.27
	14.2330	2.818	358.3	3.49	56.4	3.17	14.29
END	16.1800	2.832	283.0	4.26	42.8	3.97	13.75
GAIN (kg)							
13.2330-14.2330		-0.216					1.02
13.2330-16.1800		-0.23					0.48
NET GAIN (kg)							
13.2330-14.2330		0.804 (1%)					
13.2330-16.1800		0.250 (0.1%)					

TOTAL SULPHUR ADDED (kg) — D.H. 13.2330-14.2330 = 3158.8 x 0.256 = 80.86
(= Fuel (kg) x wt % sulphur) — D.H. 13.2330-16.1800 = 8438.4 x 0.256 = 216.02

* From Appendix C Table C-6

Before proceeding with the evaluation of the trace element balances, it is useful to consider the rate at which stone attrition proceeds in the absence of fresh stone make up.

Bed Attrition

The stone balance information cannot be used directly to establish the attrition rate since it includes material deliberately removed from the gasifier and regenerator beds as samples. Also, scrutiny of the stone removal records, Appendix C Table C-7 shows that a relatively large quantity of stone was removed from the gasifier plenum during this period which again should not be included in the calculation. Thus, the "total stone" data in Table 19 includes the weight of bed material lost due to sampling, having estimated a normal stone sample removal from the gasifier at D.H. 16.1700.

The overall attrition rate throughout the period was 0.45 kg/hr/100 kg bed material and occurred almost entirely from the gasifier bed.

It is possible to calculate the attrition rate for the 24 hour period immediately prior to the shut down of the stone feed system. During this period, fresh limestone was being added at an average rate of 2.1 molar and the attrition rate was found to be 1.40 kg/hr./100 kg bed inventory. The bed velocity during this period was 1.52 m/sec. which is similar to that for the period when no stone was added. The increase in solids escaping from the bed as the result of adding fresh material is thus very pronounced.

Taking periods of 24 hours from the start of the period (D.H. 13.2330) the rate of attrition can be seen to be related to the bed velocity, the relationship being linear over the range and of the form

$$\text{Bed Loss/hr/100 kg} = 0.659 \text{ Bed Velocity (m/sec)} - 0.689$$

with a correlation coefficient of 0.97.

It can be seen also from Table 19 that there is a decrease in the quantity of fines present in the gasifier bed though this does not account for the total loss of bed material observed. The explanation lies in the continuous generation of fines in the bed. Larger particles continuously break down to produce fines which leave the bed and

TABLE 19

LIME ATTRITION RATE

Time (Day Hour)	Duration (hr)	Bed Depth (m)	Total (1) Limestone (kg)	Loss (kg)	Rate (2) of Loss (kg/hr)	Average Loss (2) (kg/hr/100 kg Bed)	Average Bed Velocity (m/sec.)	Gasifier Bed fines wt% (<600μ)
Start (13.2330)	0	1.30	479	-	-	-	-	30.5
13.2330-14.2330	24	1.13	422	57.0	2.4	0.53	1.77	27.5
14.2330-15.2330	24	0.97	368	54.0	2.3	0.57	1.96	25.1
15.7330-16.1830	20	0.91	351.0	17.0	0.9	0.23	1.43	24.6
Overall	68	-	-	128.0	1.88	0.45(3)	1.73	-

(1) Gasifier + Regenerator bed + samples removed

(2) Based on mean bed limestone for beginning and end of period

(3) 0.12 kg/hr/100 kg bed/m^2

the fines levels measured in no way provide an estimate of the amount which has been removed in the interval between sampling. The gradual decrease in fines level is indicative only of the increasing resistance of the stone particles remaining in the bed to abrasion and attrition.

Trace Element Balances

The levels measured by atomic absorbtion techniques for a number of trace elements in the fuel oil feed, and stone samples taken during the period when no fresh limestone was added are shown in Table 20. As stated above, full sets of samples were analysed after 24 hours, and 68 hours operation.

In the case of heavy fuel oil, some elements such as chromium and cadmium were present at levels too low to be measured accurately. For the purpose of the balance calculations, the maximum level of detection has been taken. For a particular element, the % recovery has been calculated as:

$$\frac{\text{(Gasifier + Regenerator) Final Inventory (kg)} + \text{Sample Inventory (kg)}}{\text{(Gasifier + Regenerator) Initial Inventory (kg)} + \text{Weight supplied via Fuel (kg)}} \times 100\%$$

Thus, when the weight of the element in question provided by the fuel is over estimated by taking the maximum detection level, e.g. for chromium, the % recovery is under-estimated.

Table 21 shows the recovery % for the trace elements which could be measured with reasonable precision.

It is readily apparent that most elements are recovered to a lesser extent as the bed age increases. Exceptions are silicon and aluminium. In both cases, the level observed in the fuel oil is very low compared to the level present on the limestone itself, and under these circumstances the calculations are open to greater error because of the difficulty of measuring small differences in level on the stone samples, when the background level is comparatively high.

From the results, it is evident that both sodium and potassium, being relatively volatile, are not retained to any great extent on the recovered stone samples and the unrecovered material is assumed to leave the system via the fine material present in the stack gases. This is supported

TABLE 20

ELEMENTAL ANALYSIS OF FUEL OIL AND LIMESTONE SAMPLES

SAMPLE SOURCE	LIME-STONE (CALCINED)	HEAVY FUEL OIL	D.H. 13.2330					
			GAS. BED	REGEN. BED	GAS. CYC.	BOILER BACK	STACK CYC.	REGEN. CYC.
SODIUM ppm	50	47	57	166	57	112	890	321
IRON ppm	486	27	1472	1115	1094	943	1550	1212
NICKEL ppm	77	42	376	233	352	290	430	226
VANADIUM %	67*	338*	0.38	0.23	0.37	0.22	0.33	0.13
POTASSIUM ppm	42	3	23	47	9	47	319	80
MAGNESIUM % (1)	0.87	6*	0.89	0.92	0.84	0.84	0.72	0.69
CALCIUM % (2)	96.4	21*	90.2	91.8	91.8	85.3	79.1	81.8
SILICON % (3)	1.84	<3*	0.9	0.8	0.8	0.8	0.6	0.8
ALUMINIUM % (4)	0.50	2.5*	0.6	0.8	0.8	0.6	0.6	0.6
MANGANESE % (5)	0.017	<2	0.013	0.013	0.016	0.012	0.017	0.018
LEAD ppm	75	1	68	69	63	57	90	83
CHROMIUM ppm	7	<2	47	33	152	30	25	7
CADMIUM ppm	13	<1	11	11	10	10	10	12

(1) As MgO; (2) As CaO; (3) As SiO_2; (4) As Al_2O_3; (5) As Mn_2O_3

* ppm

cont/....

TABLE 20 (Continued)

SAMPLE SOURCE	D.H. 14.2330 GAS. BED	REGEN. BED	GAS. CYC.	BOILER BACK	STACK CYC.	REGEN. CYC.
SODIUM ppm	39	48	55	111	1600	210
IRON ppm	1203	1180	1045	1010	1521	1436
NICKEL ppm	490	424	378	374	716	492
VANADIUM %	0.63	0.43	0.40	0.29	0.60	0.47
POTASSIUM ppm	7	7	3	10	125	12
MAGNESIUM % (1)	0.89	0.89	0.84	0.29	0.58	0.72
CALCIUM % (2)	88.3	89.4	85.0	85.7	63.5	79.7
SILICON % (3)	0.9	0.8	1.3	1.5	1.1	1.2
ALUMINIUM % (4)	0.6	0.7	0.8	0.9	0.4	0.5
MANGANESE % (5)	0.014	0.013	0.017	0.012	0.015	0.017
LEAD ppm	63	63	70	64	102	79
CHROMIUM ppm	42	46	39	33	29	31
CADMIUM ppm	11	11	10	10	8	9

(1) As MgO; (2) As CaO; (3) As SiO_2; (4) As Al_2O_3; (5) As Mn_2O_3

cont/....

TABLE 20 (Continued)

SAMPLE SOURCE	D.H. 16.1800					
	GAS. BED	REGEN. BED	GAS. CYC.	BOILER BACK	STACK CYC.	REGEN. CYC.
SODIUM ppm	32	104	52	137	1460	1050
IRON ppm	1403	1159	1644	1221	1175	1605
NICKEL ppm	583	399	548	556	579	870
VANADIUM %	0.64	0.48	0.63	0.46	0.55	0.67
POTASSIUM ppm	<1	7	2	15	50	59
MAGNESIUM % (1)	0.86	0.77	0.86	0.77	0.34	0.71
CALCIUM % (2)	91.5	89.9	86.0	80.8	35.3	77.3
SILICON % (3)	1.3	1.7	1.5	1.6	1.1	1.3
ALUMINIUM % (4)	0.6	1.2	0.9	0.6	0.4	0.4
MANGANESE % (5)	0.013	0.011	0.016	0.016	0.012	0.023
LEAD ppm	68	62	62	63	57	87
CHROMIUM ppm	40	35	195	34	24	34
CADMIUM ppm	10	10	10	10	5	9

(1) As MgO; (2) As CaO; (3) As SiO_2; (4) As Al_2O_3; (5) As Mn_2O_3

TABLE 21

TRACE ELEMENT BALANCES

	% Recovered	
	24 hours	68 hours
Sodium	39.0	22.3
Potassium	66.7	37.4
Iron	79.2	78.0
Nickel	78.2	53.0
Vanadium	100.7	64.2
Magnesium (as MgO)	101.2	88.6
Silicon (as SiO_2)	105.3 (min)	156.6 (min)
Aluminium (as Al_2O_3)	104.5	123.8
Manganese (as Mn_2O_3)	102.3 (min)	89.3 (min)
Lead	103.0	97.3
Chromium	73.0 (min)	61.4 (min)
Cadmium	59.9 (min)	37.8 (min)

by the observation that these elements tend to be concentrated on the fine particles which are recovered at the boiler cyclone since it can be inferred that the still finer material which will not be captured will contain a higher concentration of these elements. Confirmation is available from the measurements conducted by GCA on stack emissions during the run (8) when up to 0.43% wt of sodium was found on the stack solids emitted. Similar trends are observed for potassium.

Enrichment factors have been calculated for the trace elements, as the ratio of the concentration of the element detected on the stone sample to the level observed on the original limestone feed (calcined). These are shown in Table 22.

It is interesting to note that for sodium, there is a trend for the concentration on the finer material (e.g., stack cyclone, regenerator cyclone) to increase throughout the period. The reason for this is undoubtedly the continued generation of fresh active surface for capture by fines generation due to abrasion processes occurring in the fluidised beds of the gasifier and regenerator. The gradual diminution in the enrichment factor for the gasifier bed is unexpected and is possible only if the active sites generated by particle breakdown become occupied with more stable species when competition occurs for these sites.

Similar results apply to potassium also, except that the capture by fines also decreases suggesting that other trace elements are preferentially captured.

One of these elements may in fact be vanadium which is the major trace element present in the heavy fuel oil. Over the 24 hour period, all the vanadium was accounted for even without fresh stone addition. Even under these conditions, the concentration reached only 0.63% and thereafter no further increase was observed on the bed material, though the concentration on the fines continued to increase. Vanadium is also detected on the fines leaving the system via the stack, and it is suspected that this increased only after at least 24 hours from the start of the period when no fresh stone was added. It appears from these results that concentrations on the bed material are unlikely to exceed approximately 0.6-0.7 wt % and the commercial exploitation of spent gasifier stone for vanadium recovery may be economically unattractive as a consequence.

TABLE 22

TRACE ELEMENT ENRICHMENT FACTORS

SAMPLE SOURCE	D.H. 13.2330 GAS. BED	REGEN. BED	GAS. CYC.	BOILER BACK	STACK CYC.	REGEN. CYC.
SODIUM	1.1	2.3	1.1	2.2	19.8	6.4
IRON	3.0	2.3	2.3	1.9	3.2	2.5
NICKEL	4.9	3.0	4.6	3.8	5.6	2.9
VANADIUM	56.7	34.3	55.2	32.8	49.3	19.4
POTASSIUM	0.55	1.1	0.21	1.1	7.6	1.9
MAGNESIUM	1.0	1.1	1.0	1.0	0.8	0.8
ALUMINIUM	1.2	1.6	1.6	1.2	1.2	1.2
MANGANESE	0.8	0.8	0.9	0.7	1.0	1.1
LEAD	0.9	0.9	0.8	0.8	1.2	1.1
CHROMIUM	6.7	4.7	21.7	4.3	3.6	1.0
CADMIUM	0.8	0.8	0.8	0.8	0.8	1.0

cont/....

TABLE 22 (Continued)

SAMPLE SOURCE	D.H. 14.2320					
	GAS. BED	REGEN. BED	GAS. CYC.	BOILER BACK	STACK CYC.	REGEN. CYC.
SODIUM	0.8	1.0	1.1	2.2	32.0	4.2
IRON	2.5	2.4	2.2	2.1	3.1	3.0
NICKEL	6.4	5.5	4.9	4.9	9.3	6.4
VANADIUM	94.0	64.2	59.7	43.3	89.6	70.1
POTASSIUM	0.17	0.17	0.07	0.24	3.0	0.29
MAGNESIUM	1.0	1.1	1.0	1.0	0.7	0.8
ALUMINIUM	1.2	1.4	1.6	1.8	0.8	1.0
MANGANESE	0.8	0.8	1.0	0.7	0.9	1.0
LEAD	0.8	0.8	0.9	0.8	1.4	1.1
CHROMIUM	6.0	6.6	5.6	4.7	4.1	4.4
CADMIUM	0.8	0.8	0.8	0.8	0.8	0.6

cont/....

TABLE 22 (Continued)

SAMPLE SOURCE	D.H. 16.1800					
	GAS. BED	REGEN. BED	GAS. CYC.	BOILER BACK	STACK CYC.	REGEN. CYC.
SODIUM	0.6	2.1	1.0	2.7	29.2	21.0
IRON	2.9	2.4	3.4	2.5	2.4	3.3
NICKEL	7.6	5.2	7.1	7.2	7.5	11.3
VANADIUM	95.5	71.6	94.0	68.7	82.1	100.0
POTASSIUM	<0.2	0.17	0.05	0.36	1.2	1.4
MAGNESIUM	1.0	0.9	1.0	0.9	0.4	0.8
ALUMINIUM	1.2	2.4	1.8	1.2	0.8	0.8
MANGANESE	0.8	0.6	0.9	0.9	0.7	1.3
LEAD	0.9	0.8	0.8	0.8	0.8	1.2
CHROMIUM	5.7	7.0	28.0	4.9	3.4	4.9
CADMIUM	0.7	0.8	0.8	0.8	0.4	0.6

Good retention of iron is observed throughout, the stable enrichment factors being due mainly to the relatively small amount of the element present in the fuel oil compared to the limestone (calcined).

Nickel also shows good retention though it shows a decrease with time, and tends to be progressively captured by the bed fines.

Magnesium, aluminium, silicon and manganese all show good retention in the short term, but these results are difficult to interpret due to the large discrepancy in the fuel oil and calcined limestone levels.

Finally, both chromium and cadmium balances are open to question because of the difficulty of measuring the levels present in the fuel oil feed. As stated above, the enrichment factors are under-estimated for these elements and the retention may well be considerably higher than reported here. The results for chromium are interesting nevertheless in that it seems that the major concentrations appear in the larger particles of bed material rather than the fines.

Arising from these observations, further work may be justified to examine the effect of particle size on trace element retention in greater detail. However, of paramount importance in any further attempts to better quantify the capture of the trace elements included here, and to extend the investigation to other elements, is the necessity of having more sensitive and precise analytical techniques for both limestone and fuel oil. A further factor influencing the balance outcome is the sampling techniques employed. Checks may be needed to establish a suitable sampling procedure to ensure that the small sample taken for analysis is truly representative of the material being analysed.

BATCH UNIT TESTS

Introduction

Under the terms of contract 68-02-1479, Phase 4 required evaluation of three new limestones and one new fuel in the ERCA Batch Reactor. In fact, only one limestone was examined, viz. from Whites Mines, Texas, as a candidate limestone for the proposed CAFB demonstration project at San Benito in Texas. This was selected by Foster Wheeler Energy Corporation as a likely limestone for the programme on the grounds of the proximity of the source to the test site and availability of supply. Tests were carried out to examine the sulphur absorbtion, regeneration and attrition characteristics of this limestone.

Tests were conducted also on coal as a fuel for the CAFB gasifier. Initial development work on equipment was necessary to develop a suitable feeding system for solid fuel, following which gasification tests were carried out using Illinois No. 6 sub-bituminous coal and Texas lignite. The main objectives of this programme were to establish that solid fuels could be injected into the gasifier at suitable controllable rates, that the solid fuel could be gasified and desulphurised and to quantify as far as possible the potential problems associated with coal and lignite, such as carbon fines losses and ash accumulation in the gasifier bed. The primary objectives of this aspect of the contract were achieved, but the programme was curtailed by the need to divert effort to preparation and operation of the continuous CAFB gasifier.

EVALUATION OF TEXAS LIMESTONE

Equipment and procedures

The batch CAFB reactor, and the procedure employed for evaluating limestones under a variety of operating conditions have been described previously (Ref. 2). During the current evaluation, tests were conducted to measure attrition losses during calcination, combustion, gasification and regeneration conditions as well as the sulphur retention and regeneration performance.

Fuels and limestone

The composition of the test limestone provided is given in Table 23 and is compared with the standard Grove limestone BCR 1359 selected as the standard for this test work at Abingdon.

The Texas limestone as received was of uniform size within the range 6400μ to 3200μ (1/4 inch to 1/8 inch). Prior to the test work, it was ground to below 3200μ (1/8 inch) and sieved to remove fines below 600μ. No difficulties were experienced during the grinding operations and qualitatively the Texas limestone appeared to be similar to the usual Grove limestone in this respect.

Amuay heavy fuel oil with a sulphur level of 2.4 wt% was used for the gasification tests.

Results and Discussions

The standard conditions employed for the start up of the batch unit and the calcination, combustion and regeneration test stages were found to be entirely satisfactory for the Texas limestone, and its fluidisation characteristics when calcined appeared to be no different from the limestone examined previously (Ref. 2).

Fines losses were measured for propane/kerosine calcination, kerosine combustion and fuel oil gasification modes. Results compared to previously evaluated limestones (Ref. 2) are summarised in Table 24. It can be readily seen that the attrition losses under a variety of conditions compare favourably with BCR 1359 and are superior to the other limestones evaluated.

A 5-hour gasification test was conducted to evaluate the sulphur retaining performance of the test stone. The conditions of operation are summarised in Table 25 and the sulphur removal efficiency - time relationship shown in Fig. 16. This graph is typical of the performance of limestones in the test and as a first approximation it appears that the Texas limestone would be a suitable candidate stone for the CAFB process pending further, more detailed evaluation should this material be confirmed as the supply for the San Benito demonstration project.

Regeneration presented no difficulty following the 5-hour gasification test and the sulphur burden on the stone was reduced from 4.86% to 0.66% sulphur in approximately 10 minutes subsequent to the combustion of the carbon laid down on the limestone during the gasification stages.

TABLE 23

INSPECTION OF TEXAS LIMESTONE EX WHITES MINES

Composition	Texas Limestone	BCR 1359
CaO (wt %)	57.1	54.1
MgO (wt %)	0.48	0.6
SiO_2 (wt %)	3.03	0.75
Fe_2O_3 (wt %)	0.42	0.09
Al_2O_3 (wt %)	1.00	0.31
CO_2 (wt %)	38.3	44.0
Total Sulphur (wt %)	0.21	0.12
Vanadium ppm	<25	50
Sodium ppm	102	<20

Particle Size Distribution

Sieve Size u	Wt % passing through sieve	
3200	100	100
2800	69.6	99.6
1400	32.9	87.5
1180	22.8	78.0
850	8.9	56.2
600	0.6	32.6
250	0.6	6.6
150	0.6	2.1
106	0.2	1.4

TABLE 24

FINES LOSSES FOR TEXAS LIMESTONE

		Test Conditions			
	Calcination Loss (% of Calcined) Stone	Kerosine Combustion		Fuel Oil Gasification	
		Temperature (°C)	Losses (g/min)	Temperature (°C)	Losses (g/min)
Texas Limestone	6	900	1.6	896	1.5
BCR 1359	6	870	2.1	870	3.6
BCR 1691	18	870	22.6	870	7.8
		1050	5.7		
Denbighshire	16	870	4.3	870	9.7
		1050	3.3		
Pfizer Calcite	18	870	6.8	-	-

TABLE 25

TEXAS LIMESTONE GASIFICATION TEST CONDITIONS : BATCH TESTS

Duration	5 hours
Fuel	Amuay Heavy Fuel Oil, 2.4% Sulphur
Gasifier Temperature °C	896
Air Rate l/min	520
Fuel Rate gm/min	210.9
Air/Fuel Ratio (% stoichiometric)	22.6
Bed Velocity (m/sec)	1.5
Bed Depth (cm)	63.5

SULPHUR REMOVAL EFFICIENCY vs TIME

TEXAS LIMESTONE

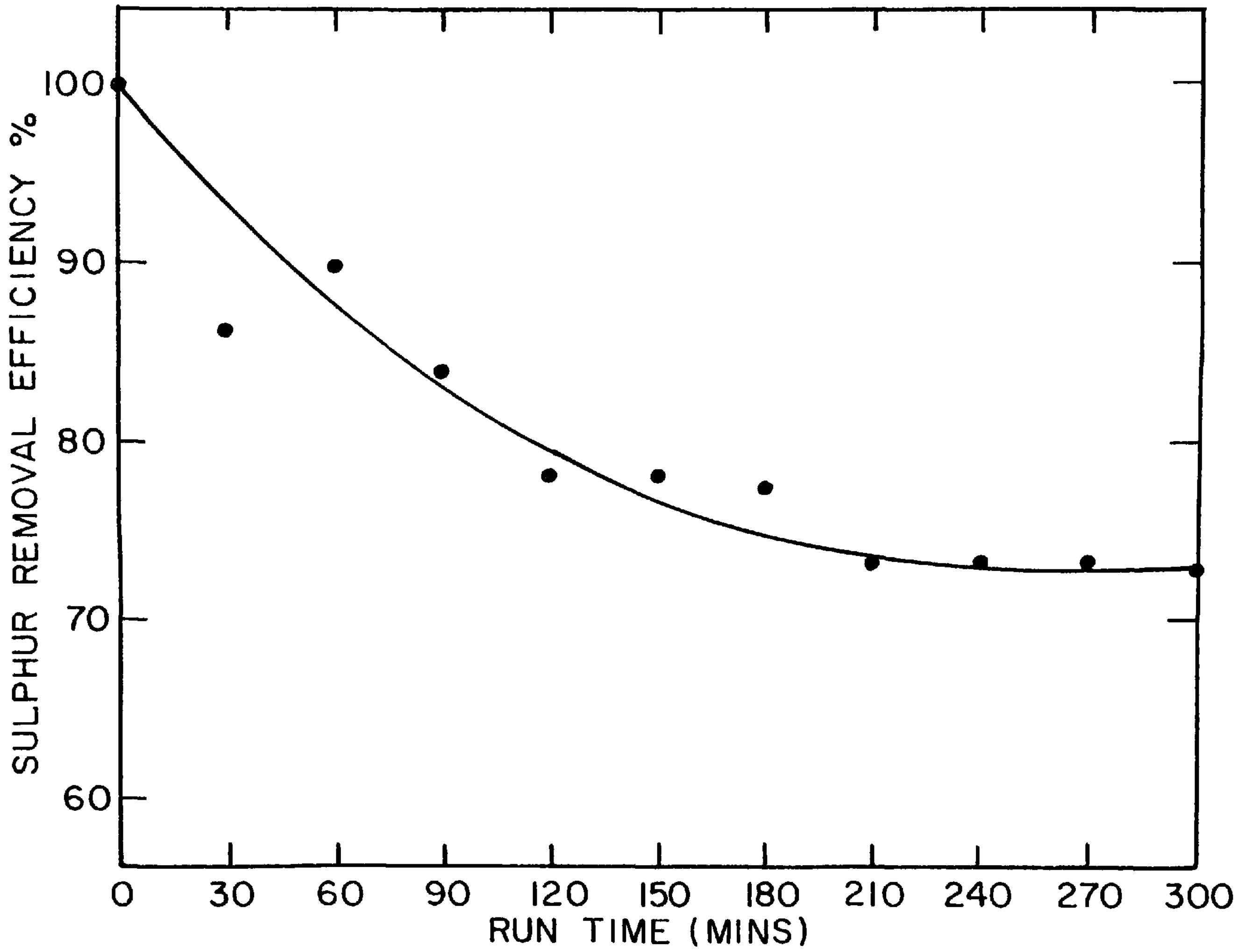


FIG. 16.

Conclusions

No problems can be foreseen for the Texas limestone tested as bed material for the CAFB process provided it is available in the appropriate size range to permit easy fluidisation. Its grindability is satisfactory and it is not subject to any unusual decrepitation or attrition losses under a variety of typical operating conditions. The sulphur retention properties are satisfactory and the sulphided stone can be readily regenerated. No further work is planned until it is confirmed that the Whites Mines, Texas will be the source for limestone for the San Benito demonstration project, when a more careful comparison with the standard BCR 1359 stone would be justified.

COAL GASIFICATION STUDIES

Summary

Preliminary work using solid fuels in the CAFB batch gasifier had identified a number of difficulties which had to be resolved before further work could be attempted. The major items were erratic coal feed rates, accompanied by blockages in the feed system, and the need to improve the fines collection and disposal system to minimise carbon losses from the process.

After preliminary work to resolve these problems, a test programme was initiated to evaluate Illinois No. 6 coal and Texas lignite as feedstocks for the CAFB process. The major objectives were to demonstrate injection, gasification and regeneration and to provide quantitative information on carbon utilisation and sulphur retention.

The qualitative targets were successfully achieved - both Illinois No. 6 and lignite could be injected and gasified. However, it proved more difficult to obtain quantitative information and only one run was completed successfully to the point where reasonable carbon and sulphur balances could be obtained. The main difficulties occurred with the coal feed system, which, although improved, was not sufficiently reliable to enable consistent feed rates of coal to be maintained over the minimum desired test duration of 5 hours.

The results from the successful run on Illinois No. 6 show that approximately 55% of the carbon was gasified with 73% desulphurisation of the feed. Optimisation of the performance of the process was hampered by the difficulties already mentioned.

The experience gained during these tests shows that improvements should be made to the coal feed system for future tests. These will not be achieved easily as the coal feed rates required are relatively low and any fluctuations thus become more important. A second problem, more amenable to solution, is the need to improve the gas sampling and monitoring system for the combusted product gas.

Equipment Modifications

A number of changes were made to the batch gasifier equipment in preparation for the test work described below.

1. A new coal feed system was designed and constructed. Tests were carried out to optimise the design and to calibrate the metering equipment, and whilst excellent performance was obtained during these stages, difficulties were encountered later, when the equipment was used to feed coal into the gasifier.

   A diagram of the revised coal feed system is shown as part of the configuration of the batch gasifier in Fig. 17.

   Coal stored in the weighed pressurised hopper was fed through a specially designed cone and a side entry to a variable speed metering screw. A number of nitrogen bleeds were found to be essential to ensure the free flow of coal. Downstream of the screw, the coal was injected pneumatically into the gasifier.

2. An improved fines handling system for product gas clean up was installed.

   This consisted of a two stage cyclone arrangement. The first stage cyclone was expected to collect carbon fines released from the gasifier and to deliver them into a pneumatic re-injection system. This cyclone was manufactured from refractory to minimise heat loss as it was anticipated that carbon utilisation in the gasifier would be maximised in this case.

LINE DIAGRAM OF BATCH COAL GASIFIER

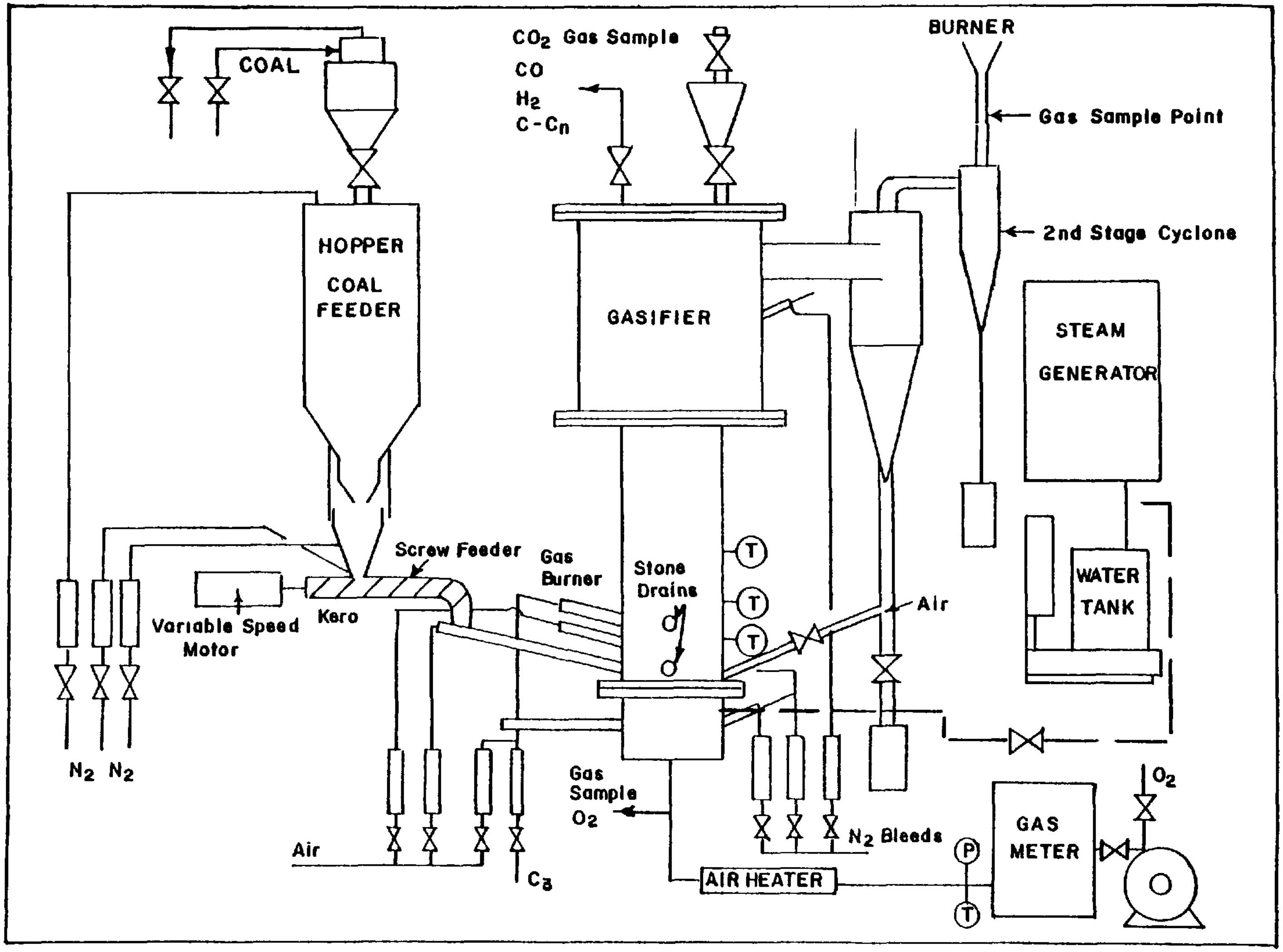


FIG.17.

The second stage cyclone was fabricated from stainless steel and was intended to collect finer ash and carbon particulates not trapped by the first cyclone and to discharge them from the system.

3. A new distributor was made and fluidising nozzles of a design similar to that used for the continuous CAFB gasifier (see Appendix A) were fitted in order to limit the fall back of bed material into the fluidising air plenum.

4. A number of minor changes were made to improve the reliability of the measurements taken for batch operations. A major leak in the gasifier air heater system was cured. An improved product gas burner was fitted and it was expected that better sampling of the flared product gas would be achieved. Facilities were installed to sample the product gas prior to combustion in order to carry out gas chromatographic analyses.

Fuels and Limestones

The fuels examined in this series of gasification tests were Illinois No. 6 sub-bituminous coal, and Texas lignite. Typical inspection details are given in Table 26.

The lignite contains considerable proportion of ash and moisture and is consequently of low calorific value.

These feedstocks were available in a variety of size distributions and ranges. Typically, the size range employed was -1400μ which was the maximum upper size limit on the capability of the coal feed system. A typical size analysis for Illinois No. 6 feed is given in Table 27.

The limestone used was the standard Grove stone, BCR 1359 in the size range 600μ to 3200μ. Typical inspections are given in Table 23.

Operating Procedures

The preparation and warm-up of the batch gasifier have been described previously as have the changeover to gasification and the procedures for bed regeneration (Ref. 2).

For the coal operations, no changes were necessary to the initial stages of the procedures up to the point where

TABLE 26

TYPICAL COMPOSITION OF ILLINOIS NO. 6 AND TEXAS LIGNITE

	Illinois No. 6	Texas Lignite
Carbon (wt %)	65.3	38.1
Hydrogen (wt %)	4.5	3.0
Sulphur (wt %)	2.8	0.5
Nitrogen (wt %)	1.2	0.7
Ash (wt %)	9.0	19.5
Moisture (wt %)	8.2	27.3
Oxygen (wt %) (by difference)	9.0	10.9
Calorific value, (kJ.kg) (Dry, ash free basis)	7,270	5,820

TABLE 27

ILLINOIS NO. 6 PARTICLE SIZE DISTRIBUTION

Particle Size Range (μ)	wt %
1400-1180	2.2
1180-850	8.5
850-600	13.7
600-250	35.9
250-150	12.6
150-106	4.1
106-52	17.4
52-0	5.6

the bed was calcined, on kerosine combustion, and ready for the gasification stage.

At this point the coal system was primed and lined through to the gasifier with the necessary pressurisation, fluidisation, and injection gas streams connected and operational. The kerosine pump was then stopped and the coal feed system started and adjusted to maintain bed temperature at 900°C. When stable, the coal rate was increased to purge the product gas ducts and then quickly increased again to achieve gasification conditions.

Normal procedures were followed for regeneration.

Results and Discussion

A number of test runs were initiated on both Illinois No. 6 and Texas Lignite feeds. Normally, no difficulties were encountered in gasifying either feedstock and in general no major differences in principle from the more usual oil gasification tests were observed. However, the reliability of the coal feed system was unsatisfactory when operating against the fluctuating back pressures encountered from the fluidised lime bed. These problems were difficult to identify and overcome because of their apparently random nature and frequency. It was considered however, that the coal flow properties, as influenced by particle size and moisture content, were important variables since most of the stoppages involved coal packing in different parts of the feed system.

Thus, most of the runs attempted had to be aborted during their early stages before any meaningful quantitative data could be collected to calculate material balances.

However, it could be seen that coal gasification was easily achieved and that the gas produced was of comparable quality to that obtained for fuel oil feed - see Table 28. Differences in the concentrations of hydrocarbon species and moisture in the gases are due to the compositional differences of the oil and coal feedstocks.

One run only was successfully completed to the point when sufficient data were collected to enable balance calculations to be made. This run lasted approximately 2 hours and was curtailed because of loss of coal feed.

TABLE 28

PRODUCT GAS ANALYSIS : ILLINOIS NO. 6

Composition (Vol %)	Illinois No. 6	Heavy Fuel Oil
H_2	12.0	11.0
N_2 + Ar	60.1	56.4
CO	10.7	13.3
CO_2	10.4	8.0
CH_4	3.0	6.4
C_2H_4	0.6	4.2
C_2H_6	NIL	0.6
H_2O	3.3	NIL

The summary of results and balance calculations for this run is given in Table 29 and the detailed calculations shown in Appendix B.

In general, the balances are reasonable bearing in mind the accuracy of analysing the product gas and collection and analysis of the solids.

The sulphur removal efficiency is calculated as 73% based on the sulphur retained on the solids. However, based on the SO2 level in the product gas, a figure of 90% sulphur removal is obtained.

Carbon gasification is calculated at 53.5% with a further 6.7% being burnt as fines inthe burner. 15.0% of the carbon was captured in the product gas cyclones, together with ash, and rejected from the system. A further 7.6% was released during regeneration, giving an overall carbon recovery of 82.8%. The apparent loss of 17.2% is due to inaccuracies in measurements.

Conclusions

These experiments have demonstrated that gasification and desulphurisation of lignite and Illinois No. 6 sub-bituminous coal is possible in the CAFB process. A carbon utilisation of 61% was recorded for Illinois No. 6 with 54% of the carbon fed being gasified. Sulphur removal efficiency measured by sulphur retained and sulphur in the flared gas was 73% and 90% respectively.

Ash present in the coal and lignite did not appear to present any practical problems, particularly during the high temperature regeneration phase when ash fusion might have occurred. A large proposition of the ash was rejected from the system via the product gas cyclones.

Further development work is required to improve the reliability of the coal feed system and the gas analysis systems.

TABLE 29

SUMMARY OF MATERIALS BALANCES

Illnois No.6

Material	% Recovery
Lime	114.2
Ash	96.7
Carbon	82.8
Hydrogen	98.4
Oxygen	85.9
Sulphur	82.4

REFERENCES

1. Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulphur Oxide Emissions. Final Report, Contract No. CPA 70-46, June 1972.

2. J.W.T. Craig _et. al._ CAFB Process for Removal of Sulphur During Gasification of Heavy Fuel Oil, Second Phase. Esso Research Centre, Abingdon. Report No. EPA-650/2-74-109, November 1974.

3. J.W.T. Craig _et. al._ CAFB Process for Removal of Sulphur During Gasification of Heavy Fuel Oil, Third Phase. Esso Reseach Centre, Abingdon. Report No. EPA-600/2-76-248, September 1976.

4. G.P. Curran, C.E. Fink 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.

5. D.L. Keairns _et. al._ Fluidisied Bed Combustion Process Evaluation. Westinghouse Research Laboratories, Pittsburgh. Report No. EPA-650/2-75-027-b, 1975.

6. R.W. Cox _et. al._ J. Inst. Fuel, 498.

7. D. Lyon. First Trials of CAFB Pilot Plant on Coal. Esso Research Centre, Abingdon. Report No. EPA-600/7-77-027, March 1977.

8. A.S. Werner _et. al._ Preliminary Environmental Assessment of the CAFB. GCA Corporation, Bedford, Massachusetts. Report No. EPA-600/7-76-017, October 1976.

9. G.L. Johnes to S.L. Rakes, minutes and action points agreed at CAFB design review meeting, Esso Research Centre, Abingdon, May 10th-13th 1976.

APPENDIX A

DESIGN AND CONSTRUCTION OF NEW CONTINUOUS PILOT UNIT

DESIGN BASIS

1. Process Requirements

Desirable process requirements for the new plant, based on previous experience and expected range of operation, were specified as follows:

	Range	Design
Bed depth, m	0.76-1.52	0.91
Bed velocity, m/sec	0.91-1.83	1.37
Bed temperature, °C	850 - 950	900
Bed stoichiometry, %	15 - 30	20

At design rate the gasifier must also be capable of gasifying bitumen. The regenerator must be able to cope with the sulphur absorbed over the whole operating range.

2. Limitations

The gasifier and regenerator must both fit inside the existing pit 3.35m x 3.66m x 2.01m deep (11ft x 12ft x 6.6ft deep) with personnel access all round; the product gas output must be within the rating of the existing boiler; layout of the plant must take into consideration the existence of plant ancilliaries (blower house etc.). Any removable plant items must be within the lifting capabilities of the existing crane (508 kg, or 1120 lb) while crane height will limit the positioning of the highest item of the plant.

3. Desirable features

Plant must be insulated to simulate a much larger unit; preferred surface temperature of metal work to be <60°C (140°F). Cyclones should be insensitive to gasifier pressure and flow fluctuations, should be accessible for modifications, and cyclone drainage should be easily accessible for modifications to the fines return system. The regenerator should be either in a separate vessel or it should be so arranged that it can be broken out, if necessary, and re-cast without disturbing the gasifier.

Tops of the gasifier and regenerator must be free to expand independently; gasifier top must be fixed to line up with cyclone inlets. Since refractory will crack, planes of weakness must be provided so that cracks do not follow undesirable directions.

DESIGN OF NEW GASIFIER/REGENERATOR UNIT

1. Gasifier

The existing boiler rating is 2.93×10^6W (10^7 B.Th.U/hr). Allowing a 10% safety margin, the maximum fuel rate should be $<2.64 \times 10^6$W ($<9 \times 10^6$ B.Th.U/hr). Characteristics of the design fuels are shown in Table A1.

Maximum fuel oil rate = 223.6 kg/hr (492.6 lb/hr).

Maximum gasifier air at 20% stoichiometric = 8.05 sm^3/min (284.2 scfm).

Bed bottom conditions; temperature = 900°C (1652°F)
assumed pressure = 17.5kPa (70" WG)

Actual air flow = 29.5 a m^3/min (1042 acfm)

Bed area for 1.83 m/sec. (6 ft/sec) = $0.882m^2$ (2.89 ft^2)

Bed diameter at base = 58.4 cm (23")

Allowance for increase in gas volume ($H_2 + CO + C_nH_m$) = 28%

Bed diameter after gas expansion = 66.0 cm (26")

Above the bed divergence to facilitate mould withdrawal to 71.1 cm (28").

2. Regenerator

Sulphur loading at maximum fuel oil rate = 5.57 kg/hr (12.27 lb/hr)

Sulphur loading at design bitumen rate = 5.47 kg/hr (12.06 lb/hr)

Therefore a regenerator sized to cope with the maximum fuel oil rate should also cope with bitumen.

At 80% (assumed) sulphur removal efficiency in the gasifier the balancing SO_2 removal from the regenerator is 0.052 sm^3/min (1.84 scfm). Assuming typical values, from previous runs, of 4% v/vCO_2 and 7.0% v/vSO_2 in the regenerator product gas the required air rate is 0.85 m^3/min (30 scfm) and, at 1050°C (1922°F), 20 kPa (80" WG) and 1.83 m/sec (6 ft/sec) bed velocity the required regenerator bottom diameter can be calculated as 20 cm (7.87"). Since the aspect ratio of the regenerator would be undesirably large at high bed depths it was decided to make the regenerator a truncated cone in shape, 19.05 cm, (7.5") at the distributor and 22.86 cm (9") at the 107 cm (42") level.

3. Plant Geometry

The internal diameters of the two reactor cavities and the intention to use the same system for transferring solids between the gasifier and the regenerator as in the old unit, fixed the centre lines of the reactor cavities at approximately 68.6 cm (27") apart and the thickness of the hot face refractory wall (sufficient to contain the solids transfer system) at 15.2 cm (6"), with both cavities cast inside one steel vessel. Heat transfer calculations showed that 10.2 cm (4") of pearlite based insulating refractory backed by 5.1 cm (2") of fibrous insulation would meet the requirement of maximum steelwork temperature of 60°C (140°F). This resulted in a plant section as shown in Fig. A1. The independence of gasifier and regenerator expansion was achieved by a vertical membrane of Kaowool blanket while the gasifier top was suspended from the wall anchors with an expansion joint below. The 3.2 mm (1/8") thick, externally braced metal casing was joined by a flange along the gasifier/regenerator separating membrane and was supported in such a way that the flange could be split and the regenerator removed, if necessary, without affecting the integrity of the gasifier.

4. Product Gas Cyclones

In order to minimise the effect of gasifier pressure fluctuations on cyclone performance it was decided to use twin ter Linden type cyclones with over-square external snail gas inlets extending for 180° at a loading of approximately 230 $m^3/min/m^2$ (750 $scfm/ft^2$) of cyclone cross-sectional area and 15 m/sec (50 ft/sec) inlet and outlet velocity.

At design conditions of 1.37 m/sec (4.5 ft/sec):-

Air flow	= 6.02 sm^3/min (212.5 scfm)
Product gas flow (assumed 28% increase)	= 7.70 sm^3/min (272.0 scfm)
Product gas flow (2.5 k Pa, 900°C)	= 32.3 am^3/min (1140.7 acfm)
Product gas flow per cyclone	= 16.2 am^3/min (570.5 acfm)

Therefore:

Cyclone cross-sectional area	= 0.071 m^2 (0.76 ft^2)
Cyclone diameter	= 30.5 cm (12")
Cyclone inlet area	= 0.017 m^2 (0.19 ft^2)
Cyclone inlet, dimensions	= 11.4 cm x 15.2 cm (4.5" x 6")

Although the desired cyclone outlet diameter was calculated as 15 cm (5.9"), the self-bonded silicon carbide tubes salvaged from the old unit were re-used; these were 14 cm I.D. (5.5" I.D.). The estimated pressure drop of these cyclones was:

at minimum flow conditions	1 kPa (4" W.G.)
at design flow conditions	2 kPa (8" W.G.)
at maximum flow conditions	3.5 kPa (14" W.G.)

The mechanical details of the cyclones are shown in Fig. A2.

5. Closures

The gasifier lid was a cylindrical block of refractory, insulated and attached by anchors to a braced metal plate. Details of the gasifier lid dimensions are shown in Fig. A3. The estimated weight was:

	kg	lb
Hot face refractory	210.0	463
Insulating refractory	5.0	11
Fibrous insulation	5.4	12
Structural steelwork	45.8	101
Fittings	43.6	96
Total	309.8	683

Therefore Lift-off pressure = 7.68 kPa (30.7" W.G.)

To allow the lid to act as a safety valve the flue gas connection to the lid was made from flexible metal pipe which would allow free movement over about 38 cm (15") in the vertical direction.

The gasifier distributor was intended to consist of a 7.5 cm (3") thick refractory slab mounted on a steel disc and pierced for 16 air nozzles, 4 in an inner ring and 12 in an outer ring. The plenum chamber was to be constructed from 3.2 mm (0.125") steel with a sloping base to facilitate solids removal. Since this arrangement was not used, no further details are provided. Design information for the modified distributor are given below.

The regenerator lid was a cylindrical refractory plug connected via a refractory lined duct of 6.6 cm I.D. (2.6") off-take pipe and a bare, 51.0 x 44.5 mm (2" x 1.75") internal, branch pipe to a 10.2 cm (4.0") diameter cyclone, geometrically similar to the product gas cyclone. The off-take pipe was connected via spring loaded bolts to the regenerator top steel plate with the spring loading allowing 20 mm (0.8") vertical movement to allow for refractory expansion. This lid was not designed to act as a relief valve. Details are shown in Fig. A4.

The regenerator distributor consisted of a 5 cm (2") thick refractory slab pierced to take 3 air nozzles which communicated with a 5 cm high 25 cm dia. (2" high x 10" dia.) air distributor plenum; details in Fig. A5.

6. Fines re-injection

The available height was insufficient to fit cyclone dipleg seals, therefore the product gas cyclones were drained into lock hoppers from which fines were re-injected into the gasifier via 21 mm (0.81") injectors. Angle of repose seals with pulsed seal breakers were used to control the rate of re-injection. Perforated steel plate "chunk traps" were provided to remove large agglomerates which might form and be dislodged during burn-out. Details of the lock hoppers are shown in Fig. A6.

CONSTRUCTION

Most of the plant steelwork was constructed in the ERCA workshops. The refractory was subcontracted to A.P. Green Ltd., with their recommended refractories described in Table A2. The plant was piped into existing service supplies (air, steam, flue gas recycle, analysers, etc.) and connected to the existing G.W.B. boiler. Positions of various penetrations are listed in Table A3.

BITUMEN INJECTION SYSTEM

The bitumen injection system consisted of two separate parts: storage and circulating ring main and the injection system.

Storage of bitumen was provided in an insulated trailer normally used for road surfacing operation. The trailer was provided with two gas oil (No. 2 heating oil) burners, which needed to be fired intermittently to maintain the bitumen in a fluid state, and a hydraulically operated pump for circulating and mixing the contents. The outlet from the circulating pumps was connected to a circulating ring main which consisted of a concentric double pipe with 5.5 bar (80 psi) steam in the annular space and insulated outside. The installation also included a stand-by diesel driven circulating pump, fire shut off valves and temperature sensors.

Hot bitumen was taken from the ring main and pumped via a Plenty 200 metering pump and a flow meter to electrically traced lines which could be connected to any one of six oil injectors (four side injectors and two pit injectors).

The Plenty 200 pump had modified bearings and seals suitable for high temperature operation and both the pump and the meter were electrically heat traced and insulated. Preliminary trials showed that the flow meter would not meter bitumen and it therefore had to be by-passed and flow estimated from the pump calibration.

MODIFICATIONS IN SUPPORT OF DEMONSTRATION PLANT

Resulting from discussions with Foster Wheeler, it was decided to incorporate and test out some possible design features of the proposed demonstration plant. These were:

Modified air distributor with enclosed oil injectors.

Tuyere injection of flue gas.

Bag filter for recycled flue gas.

Feasibility of coal injection

1. Modified air distributor

A 12.7 cm (5") deep by 29.2cm (11.5") square pit was incorporated in the centre of the distributor so that the fuel injectors could penetrate through the refractory with only the injector tip exposed to the fluid bed. As can be seen from Fig. A7, provision was made for two withdrawable fuel injectors: one through a hole in the refractory and the other through a channel which would fill with slumped lime and shield the injector. It was also arranged so that either fuel oil or bitumen could be injected through either of the two injectors. In addition, a withdrawable vertical central injector, which passed through a gland seal in the air plenum, was installed and piped in to take either fuel oil or kerosene. The gasifier was now equipped with seven injectors piped in so that any combination of one or more could be selected to inject fuel oil or kerosene and up to six (i.e. excluding the central vertical injector) could be selected to inject bitumen. Details of the fuel injector seal arrangements are shown in Fig. A8.

2. Tuyere injection of flue gas

This modification allowed the possibility of routing the moderating flue gas either mixed with air through the plenum or separately, through the tuyere, directly into the fluid bed. Successful tuyere injection of flue gas would allow its use in an unfiltered state with possible cost savings. The tuyere consisted of a 5cm (2") I.D. EN 312 S.S. pipe inserted up to 38cm (15") into the fluid bed at an angle of 45° downward through a sealing gland which replaced the start-up burner.

3. Bag filter for recycled flue gas

Since this seemed to be an excellent opportunity for checking out the performance of a bag filter in flue gas cleaning service, the existing flue gas venturi scrubber was replaced with a rectangular filter box containing four 20cm dia. by 1.52m long (8" dia. x 5' long) Nomex 40 filter bags. Flue gas entered a plenum near the top of the box, flowed radially outward through the filter bags and exited near the base. Cleaning was by back flow of nitrogen and manual withdrawal of solids through a lock hopper.

4. Feasibility of Coal Injection

Equipment to test qualitatively the feasibility of injecting coal into the gasifier was prepared for temporary installation towards the end of the test run. It consisted of a pressurised hopper which discharged via a 5 cm (2") pipe, valve and a pulsed angle of repose flow controller (identical to that used with the fines return lock hoppers) to a 32 mm (1.25") I.D. injector which was to be inserted through the centre of the flue gas recycle tuyere. A bleed of flue gas through the tuyere, which would act as a shroud, and a stream of air through the injector would reproduce approximately the conditions found suitable for coal injection on the batch units. The intention was to charge the coal hopper manually with Illinois No.6 coal sieved into various cuts in the range 0-3.2 mm (0-0.125") and inject coal intermittently, while backing-off liquid fuel, to check whether the injector would coke-up or remain free. Provision was made also to use the limestone injection system for injection of coal by providing a manual filling connection and inerting with nitrogen.

NEW PILOT PLANT GEOMETRY

(Approx. Scale, 1:39)

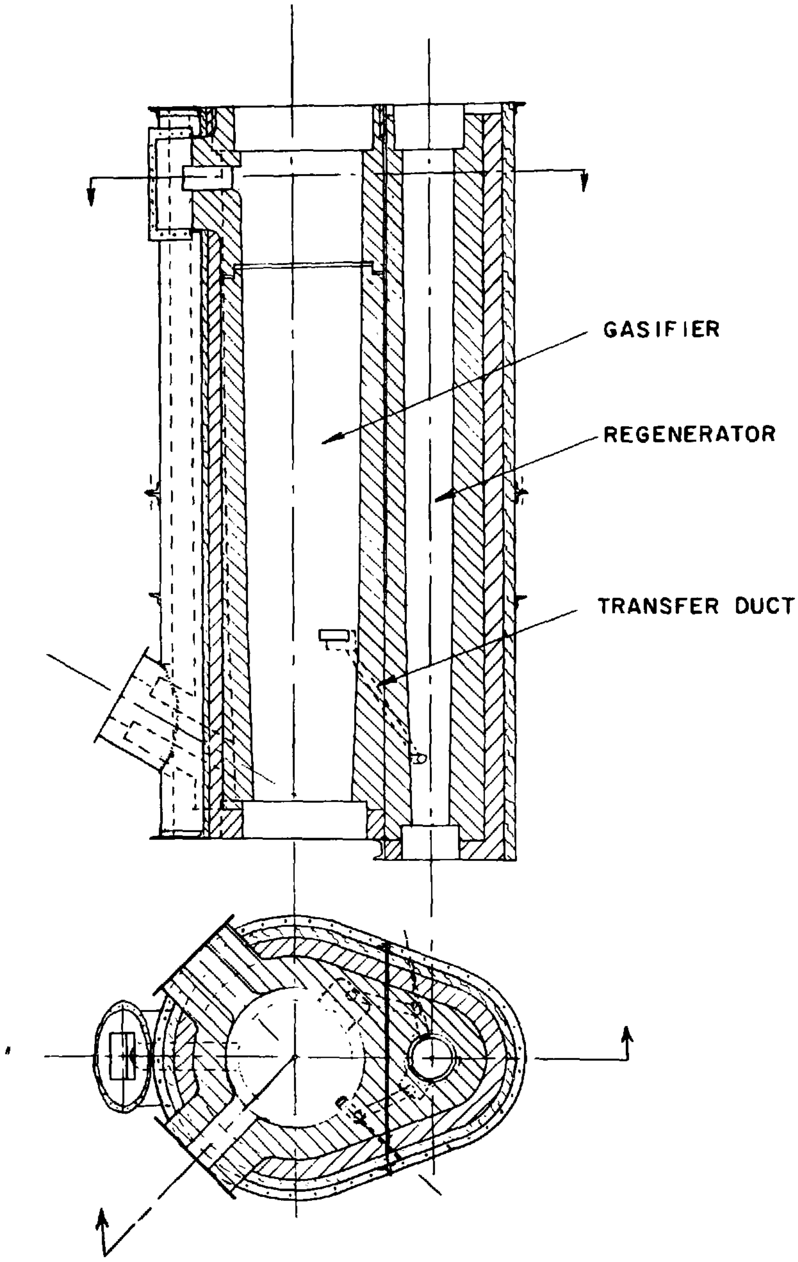


FIG. A1

PRODUCT GAS CYCLONE

(Approx. Scale, 1:19)

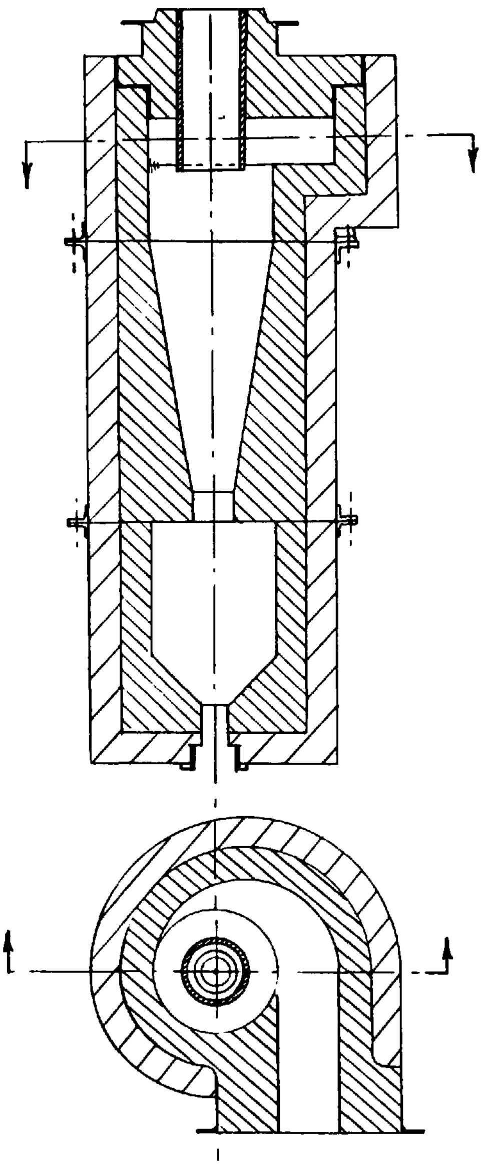


FIG. A2

GASIFIER LID

(Approx. Scale, 1:13)

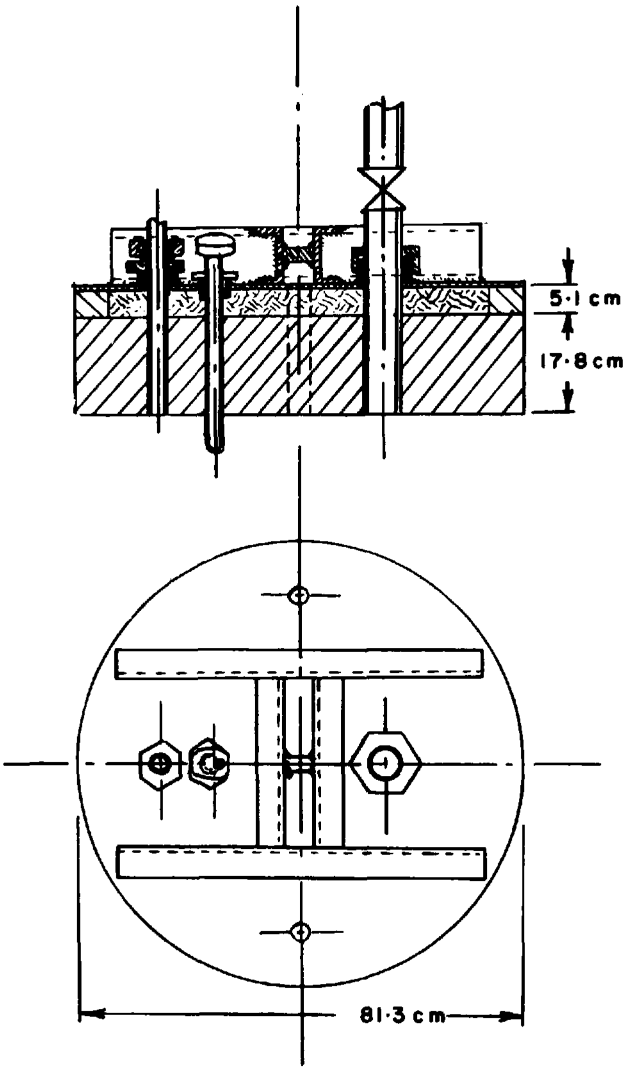


FIG. A3

REGENERATOR TOP

(Approx. Scale, 1:7)

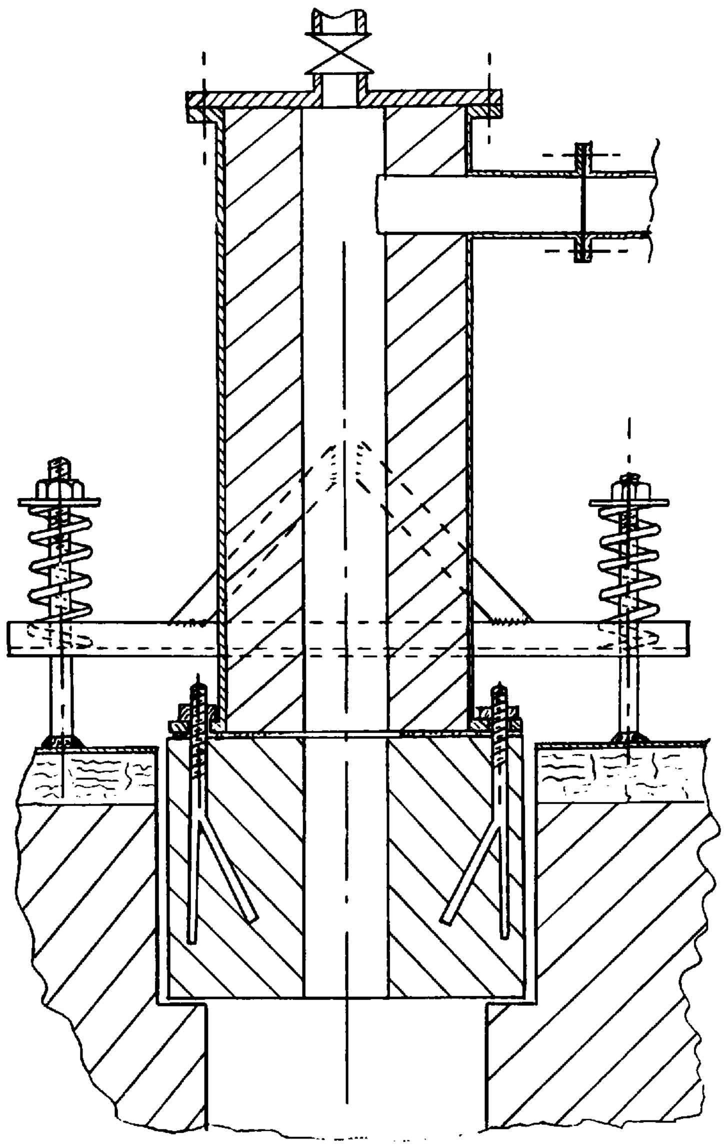


FIG. A4.

REGENERATOR DISTRIBUTOR

(Approx. Scale, 1:3.5)

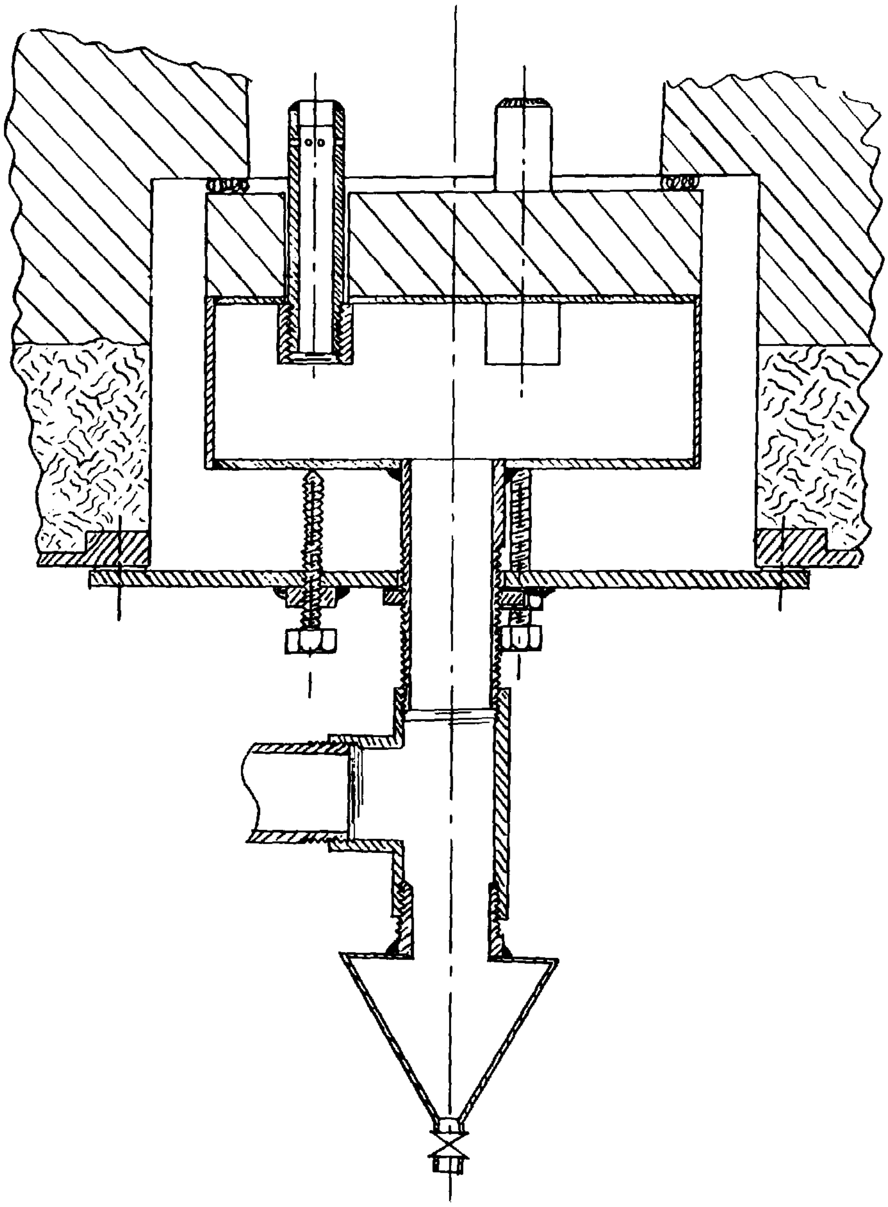


FIG. A5

FINES RETURN LOCK HOPPER

(Not to Scale)

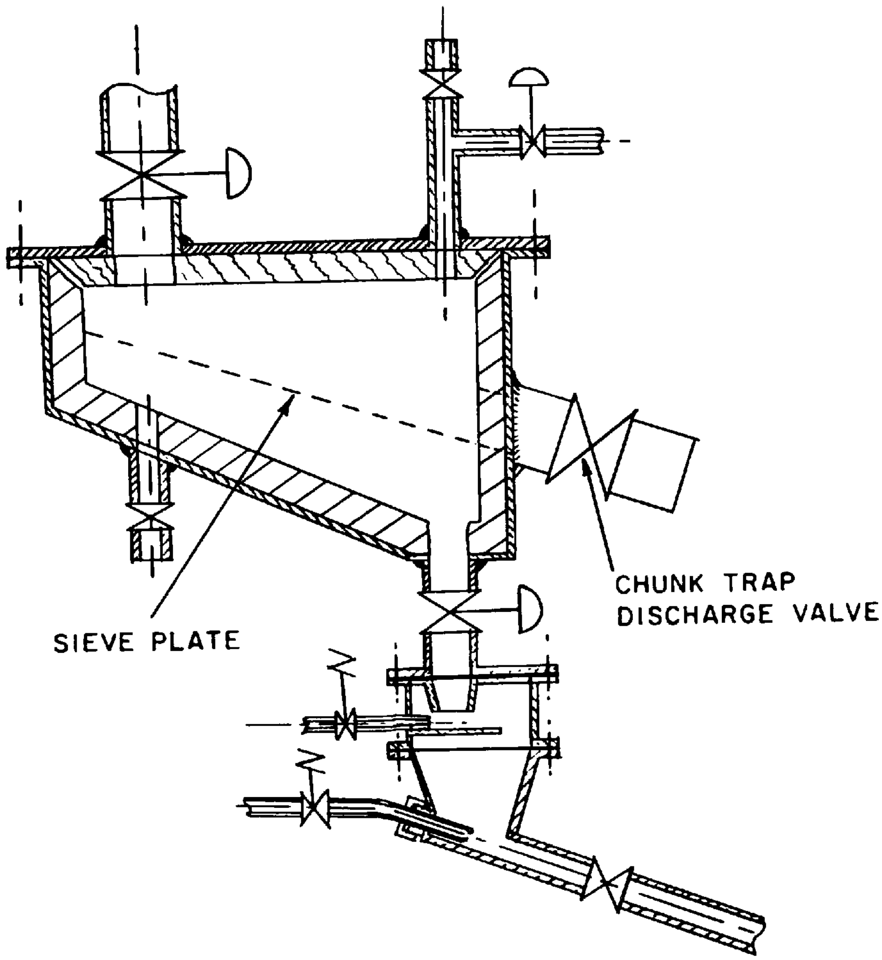


FIG. A6

MODIFIED GASIFIER DISTRIBUTOR

(Approx. Scale, 1:12)

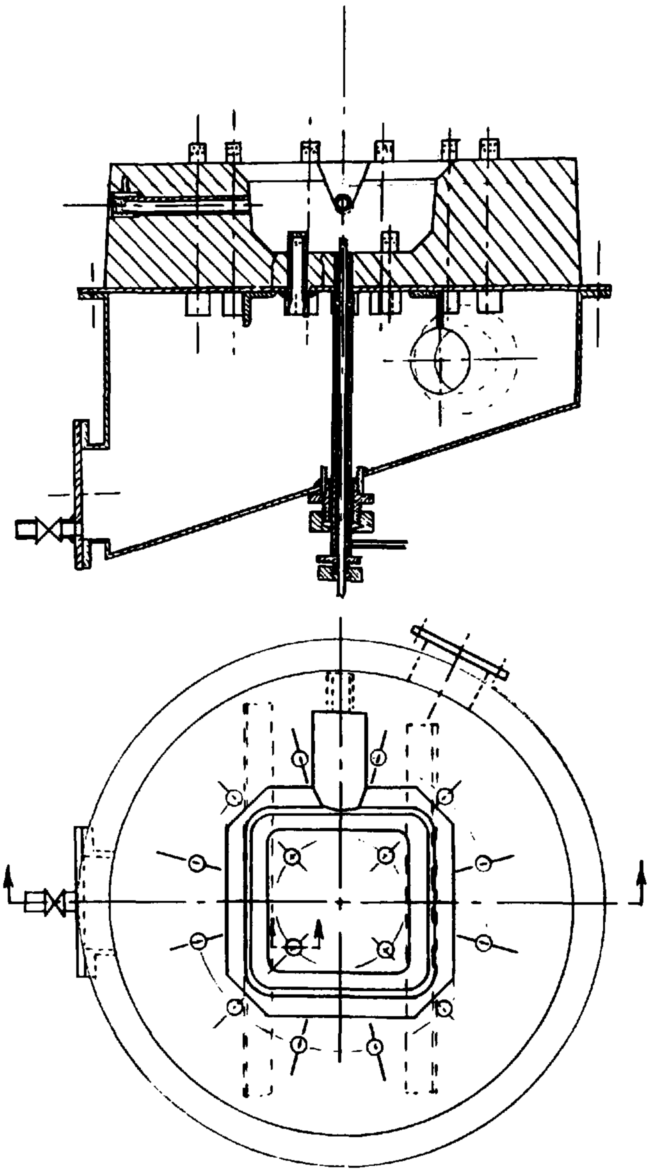


FIG. A7

FUEL INJECTOR SEAL ARRANGEMENT

(Not to Scale)

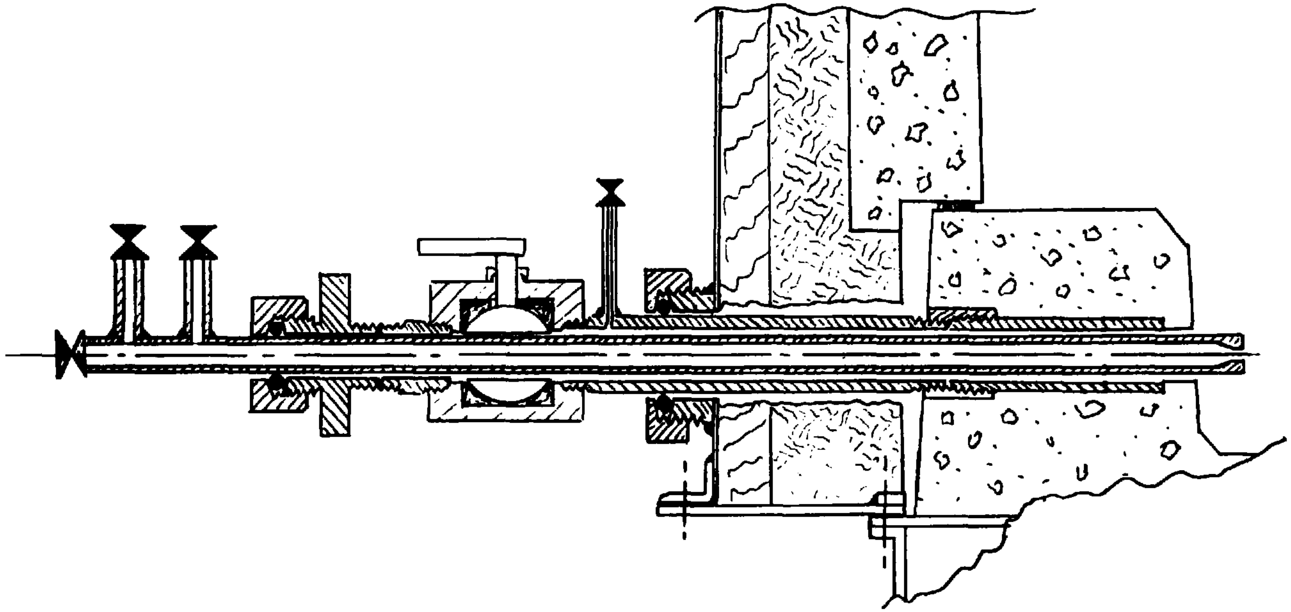


FIG. A8

APPENDIX A

TABLE A1

DESIGN FUELS

	Fuel Oil	Bitumen
C, wt %	85.31	85.83
H, wt %	11.28	10.46
S, wt %	2.49	3.22
Cv, MJ/kg (B.Th.U/lb)	42.5 (18270)	41.98 (18050)
Stoichiometric air*, m^3/kg (scf/lb)	10.81 (173.1)	10.66 (170.8)

* Note: air taken at 1.5% moisture

APPENDIX A

TABLE A2

REFRACTORIES USED IN CONSTRUCTION

Refractory	Type	Maximum Service Temperature, °C (°F)	Cured density $kg.m^{-3}$ (lb/ft^3)	Use
Greencast 94(1)	Tabular alumina	1870 (3400)	2600 (162)	Cyclone liners
High Temperature Castable(1)	70% alumina low iron, low lime	1650 (3000)	2250 (140)	Gasifier regenerator, duct lining, cyclone tops, lock hopper lining, lids, distributors.
LW-22(1)	Pearlite based lightweight castable refractory	1200 (2200)	802 (50)	Insulating refractory for gasifier, regenerator & cyclones.
Eagle-Pitcher(1)	Mineral fibre insulating block	1040 (1900)	270 (17)	Internal insulation of gasifier, regenerator, gas ducts & gasifier lid.
Refel(2)	Self-bonded silicon carbide	1400+(2550+)	3100 (194)	Cyclone off-take pipes.

Notes
(1) Supplied by A.P. Green Refractories Ltd., Dock Road South, Bromborough, Merseyside, U.K.
(2) Supplied by British Nuclear Fuels Ltd., Risley, Warrington, U.K.

APPENDIX A

TABLE A3

POSITIONS OF GASIFIER PENETRATIONS

	Item	Height from distributor, cm (inches)
1.	Warm-up burner, bottom edge	8.9 (3.5)
	Warm-up burner, top edge	30.5 (12.0)
2.	Sample drain, lower	19.1 (7.5)
	Sample drain, upper	63.5 (25.0)
3.	Side oil injector, lower (one each side)	16.5 (6.5)
	Side oil injector, upper (one each side)	33.0 (13.0)
4.	Fines return injection (one each side)	19.1 (7.5)
5.	Thermocouple, bottom (protrudes 5.1cm, 2")	15.2 (6.0)
	Thermocouple, middle (protrudes 5.1cm, 2")	50.8 (20.0)
	Thermocouple, top (protrudes 8.9cm, 3.5")	85.1 (33.5)
6.	Pressure tapping, lower (protrudes 1.9cm, 0.75")	8.9 (3.5)
	Pressure tapping, upper (protrudes 2.2cm, 0.88")	35.6 (14.0)
7.	Lime injection, centre of 6cm (2.4") aperture	48.9 (19.3)
8.	Transfer duct, "post box", lower edge	77.5 (30.5)
	Transfer duct, "post box", upper edge	84.5 (33.3)
	Transfer duct, return from regen., centre of 6.7cm (2.6") aperture	22.5 (8.9)
9.	Underside of gasifier lid	348.0(137.0)

Note:

Warm-up burner and lime injection penetrations are angled at approximately 45°.

All other penetrations are angled at 30° to horizontal

APPENDIX B

BATCH UNIT STUDIES

MATERIALS BALANCE : ILLINOIS No. 6 COAL RUN

Balance calculations for the batch unit operation on Illinois No. 6 coal are detailed below, established over the total gasification period. Due to operational difficulties, the gas composition during regeneration is not available and assumptions with regard to sulphur and carbon balances must be made. These are identified in the calculations.

Balance Calculations

1. General information

(a) Run duration : 107 minutes of gasification.
(b) Temperature conditions are shown in Fig. B1.

2. Input materials

(a) Weight of BCR 1359 limestone added = 16.0 kg.

For the sample of limestone used, the total loss of material as CO_2 and moisture during the calcining stage amounted to 41% giving an initial lime bed weight of 9.44 kg.

The composition of the lime is detailed below in Table 1.

TABLE 1

LIME COMPOSITION
(weight %)

Ca as CaO	97.30
Mg as MgO	0.92
Al as Al_2O_3	0.59
Si as SiO_2	0.89
Fe as Fe_2O_3	0.20

It has been assumed that 2% of the calcined lime bed would behave as ash derived from the coal feed, appearing as hydrochloric acid insolubles during analysis.

Thus, there is assumed to be an initial "ash" burden on the lime bed of 0.189 kg.

(b) Coal fed during the run amounted to 29.994 kg.

The composition and weights of individual components are given in Table 2.

TABLE 2

COAL FEED COMPOSITION AND WEIGHTS FED

	Composition (wt %)	Weight Fed (kg)
Carbon	65.3	19.586
Hydrogen	4.5	1.350
Sulphur	2.8	0.840
Nitrogen	1.2	0.360
Ash	9.0	2.699
Moisture	8.2	2.460
Oxygen (by difference)	9.0	2.699

(c) Air input to the gasifier during the gasification period amounted to $58.514 m^3$, comprising $46.226 m^3$ N_2 and $12.288 m^3$ O_2.

Thus, the total N_2 input from air and coal

$$= 46.226 + \frac{0.360 \times 22.4}{28} = 46.514 m^3 \text{ or } 58.143 \text{ kg}$$

and total O_2 input

$$= 12.288 + \frac{2.669 \times 22.4}{32} = 14.156 m^3 \text{ or } 20.223 \text{ kg}$$

3. Materials removed during gasification and regeneration

Tables 3 and 4 give the weights of materials removed as solids samples during the run and at the end of regeneration, and their compositions.

4. Product Gas

Averaged values for the composition and resulting weights of elements for the product gas from the gasifier are given in Table 5. The volume of product gas is established on the basis of the input N_2:

Input N_2 = 46.415 m^3, representing 59.7% of the product gas volume.

Therefore, product gas volume = $\frac{46.514 \times 100}{59.7}$ = 77.913 m^3

TABLE 5

PRODUCT GAS COMPOSITION

	Volume %	Volume m^3	Weight kg	Weight of Element (kg) N_2	H_2	C	O_2
N_2	59.7	46.514	58.143	58.143	-	-	-
H_2	11.9	9.272	0.828	-	0.828	-	-
CO	10.6	8.259	10.324	-	-	4.425	5.899
CO_2	10.3	8.025	15.763	-	-	4.299	11.464
CH_4	3.0	2.337	1.669	-	0.417	1.252	-
C_2H_4	0.6	0.467	0.584	-	0.083	0.501	-
H_2O	3.9	3.039	2.442	-	-	-	-
	100	77.913	89.753	58.143	1.328	10.477	17.363

It is assumed also that there is no net loss or gain or moisture through the gasifier, and that the SO_2 level in the product gas, typically at ppm levels is insignificant.

TABLE 3

SOLIDS SAMPLES, WEIGHTS AND COMPOSITION

ORIGIN	TIME (min)	SAMPLE WEIGHT (kg)	COMPOSITION (wt %)					
			TOTAL SULPHUR	SULPHATE SULPHUR	CARBON	ACID INSOLS.	LIME	ASH
1st Cyclone	30	0.190	3.01	<0.01	28.8	43.0	54.0	14.2
2nd Cyclone	60	4.763	1.65	<0.01	43.0	91.0	7.4	48.0
Gasifier	107	0.100	2.96	<0.01	12.1	16.0	81.0	3.9
Gasifier	After Regeneration	9.800	1.04	0.83	0.1	2.1	96.9	2.0
1st Cyclone	After Regeneration	0.170	3.22	<0.01	55.3	84.0	12.8	28.7
2nd Cyclone	After Regeneration	1.474	3.57	<0.01	48.3	62.0	34.4	13.7

TABLE 4

WEIGHTS REMOVED, gm

ORIGIN	TIME (min)	SAMPLE WEIGHT (kg)	WEIGHT (gm) TOTAL SULPHUR	SULPHATE SULPHUR	CARBON	ACID INSOLS.	LIME	ASH
1st Cyclone	30	0.190	6	-	55	82	103	27
2nd Cyclone	60	4.763	79	-	2048	4332	353	2286
Gasifier	107	0.100	3	-	12	16	81	4
Gasifier	After Regeneration	9.800	102	81	10	206	9496	196
1st Cyclone	After Regeneration	0.170	6	-	94	143	143	49
2nd Cyclone	After Regeneration	1.474	53	-	712	914	914	202
	TOTALS		249	81	2931	5695	10563	2764

5. Combusted Gas Composition

The typical composition of the flared product gas is given below in Table 6.

TABLE 6

FLARED GAS COMPOSITION

	Vol. %
O_2	3.7
CO_2	14.6
SO_2	0.042

These data will be used later in calculating total sulphur and carbon balances.

6. Materials Balances

(a) Lime Weight lime fed $= \frac{98 \times 9.44}{100} = 9.251$ kg

Weight lime recovered = 10.562 kg

$$\% = \frac{10.562 - 9.251}{9.251} \times 100\% = 14.2\ \%$$

(b) Ash Weight ash fed $= 2.669 + \frac{2 \times 9.44}{100} = 2.858$ kg

Weight ash recovered = 2.764 kg

$$\% = \frac{2.764 - 2.858}{2.858} \times 100\% = -3.3\%$$

It is noted that most of the ash recovery was from the 2nd stage cyclone where 2.488 kg (87%) of the ash fed appeared.

(c) Hydrogen Weight hydrogen fed = 1.350 kg

Weight hydrogen recovered = 1.328 kg

$$\% = \frac{1.328 - 1.350}{1.350} \times 100\% = -1.6\%$$

(d) Carbon

Weight carbon fed = 19.586 kg

This has to be accounted for by the carbon recovered from samples and bed material, the carbon gasified, and the fines present in the product gas combusted in the gas burner.

Carbon recovered from solid samples = 2.931 kg
Carbon present in product gas = 10.477 kg

Carbon burn off during regeneration cannot be estimated from gas composition sa these are not available. However, an estimate may be made based on bed weights and compositions.

Let x = weight of gasifier bed at start of regeneration, and assume that the solids collected in the cyclones accumulated during the regeneration phase.

$$\frac{81}{100}(x + 0.1) = \frac{96.9}{100} \times 9.496 + \frac{12.8 \times 0.17}{100} + \frac{34.4 \times 1.474}{100}$$

$$x = 12.277 \text{ kg}$$

Thus, carbon content of gasifier at end of gasification, burnt of during regeneration

$$= \frac{12.1 \times 12.277}{100} = 1.486 \text{ kg}$$

It remains to calculate the carbon fines which have been combusted in the product gas burner.

The O_2 require for stoichiometric combustion of the product gas

= weight for hydrogen combustion + weight for carbon combustion - weight of oxygen in product gas

$$= 1.328 \times \frac{16}{2} + 10.477 \times \frac{32}{12} - 17.363 \text{ kg}$$

$$= 21.190 \text{ kg}$$

Equivalent weight of N_2 = $\frac{21.190 \times 79 \times 28}{21 \times 32}$ = 69.750 kg

Thus, the composition of the combusted product gas under stoichiometric conditions is:

$$N_2 = 58.143 + 69.750 = 127.893 \text{ kg}$$

$$CO_2 = 10.477 + 10.477 \times \frac{32}{12} = 38.415 \text{ kg}$$

$$H_2O = 1.328 + 1.328 \times \frac{16}{2} = 3.099 \text{ kg}$$

$$(\text{Check } O_2 = 21.195 + 17.363 - 10.477 \times \frac{32}{12} - 1.328 \times \frac{16}{2} = \text{Nil})$$

This ignors the insignificant amount of oxygen required to convert any sulphur species to SO_2.

The additional amount of carbon burnt as fines may now be calculated.

TABLE 7

FLARED GAS COMPOSITION

	Weight (kg)	Volume (m^3)	Volume (%)
N_2	127.893	102.314	83.92
CO_2	38.415	19.559	16.08

A measured 14.6% CO_2 was obtained at 3.7% excess oxygen giving a level under stoichiometric conditions of

$$14.6 \times \frac{21}{21-3.7} = 17.72 \text{ vol \%}$$

and a corresponding N_2 level of 82.28 vol %. The actual N_2 volume is 102.315 m^3 and thus the CO_2 volume

$$= \frac{17.72}{82.28} \times 102.314 \text{ m}^3 = 22.034 \text{ m}^3.$$

The additional CO_2 = 22.034 - 19.559 = 2.435 m^3 representing 1.304 kg of carbon as fines in the product gas.

Thus, total carbon recovered

$$= 2.931 + 10.477 + 1.486 + 1.304$$

$$= 16.198 \text{ kg}$$

$$= \frac{16.198 - 19.586}{19.586} \times 100\% = -17.2\%$$

$$\text{Carbon gasified} = \frac{10.477}{19.586} \times 100\% = 53.5\%$$

(e) Sulphur

Sulphur fed = 0.840 kg

Sulphur recovered on solid samples = 0.249 kg

Sulphur lost during regeneration

$$= \frac{2.96}{100} \times 12.277 = 0.363 \text{ kg}$$

Sulphur recovered in combusted gas

$$= \frac{32 \times 10.477 \times 0.042}{12 \times 14.6} = 0.080 \text{ kg}$$

Total sulphur recovery = 0.692 kg

$$= \frac{0.692 - 0.840}{0.840} \times 100\% = -17.6\%$$

Sulphur removal efficiency (based on combusted gas analysis)

$$= \frac{0.080}{0.840} \times 100\% = 90\%$$

Sulphur removal efficiency (based on sulphur retained on bed material) = 73%

(f) Oxygen

Weight oxygen fed = 20.223%

Weight oxygen recovered = 17.363%

$$\% = \frac{17.363 - 20.223}{20.223} \times 100\% = -14.1\%$$

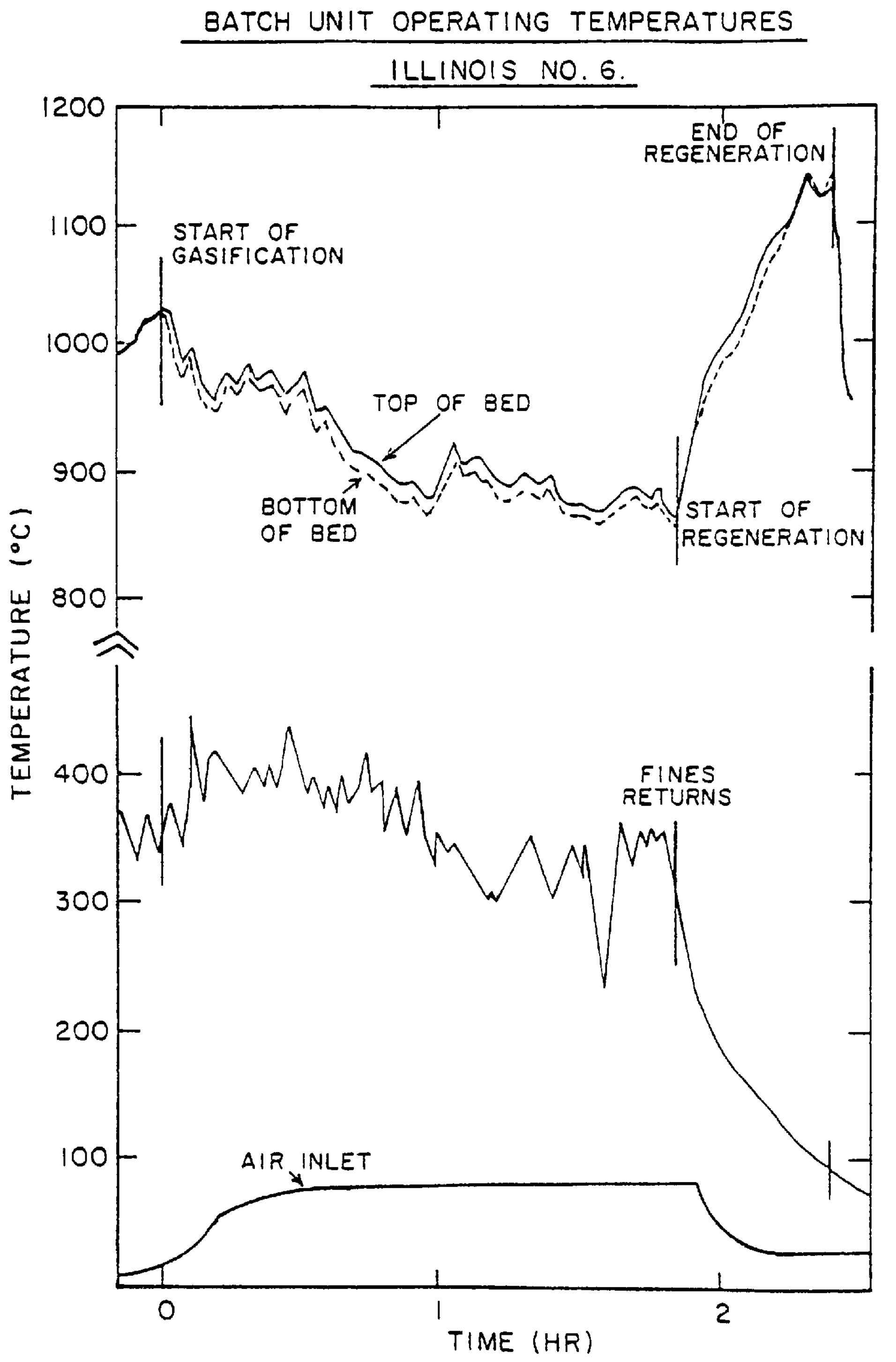

BATCH UNIT OPERATING TEMPERATURES
ILLINOIS NO. 6.
1200
1100
1000
900
800
400
300
200
100
TEMPERATURE (°C)
START OF GASIFICATION
END OF REGENERATION
TOP OF BED
BOTTOM OF BED
START OF REGENERATION
FINES RETURNS
AIR INLET
0
1
2
TIME (HR)

FIG. B.1.

APPENDIX C

OPERATIONAL LOG, TEST DATA, INSPECTION, EQUIPMENT PERFORMANCE

OPERATIONAL LOG

10.11.75 to 20.11.75 : Refractory curing and warm-up

The warm-up burner was lit at 11.15 on 10.11.75 to commence the curing and warm-up sequence for the new refractory according to the schedule provided by the refractory suppliers, A.P. Green Services Ltd.

Temperature Range	Rate (°C/hr)	Duration (hr)
0°C to 120°C	15	8
Hold at 120°C		9.75
120°C to 260°C	15	9.33
Hold at 260°C		9.75
260°C to 540°C	15	18.67
Hold at 540°C		9.75
540°C to 815°C	15	18.33
Hold at 815°C		9.75
	TOTAL	93.33

All gasifier and regenerator thermocouples were withdrawn 2.5 cm (1 inch) into their respective refractory walls during this period in order to monitor the refractory temperatures more accurately.

The above schedule was followed as closely as possible for both the gasifier and regenerator sections of the unit by modulating propane and air rates to the warm-up burner situated in the gasifier. Further control was possible, particularly to cure the regenerator refractory, by varying hot gas flows within the unit using the gasifier-regenerator pressure balance blower system, the gasifier, regenerator, boiler and flue gas recycle blowers, and the various dampers controlling gas flows around the system.

During the initial stages of the warm-up sequence, copious amounts of water were released from the new refractory, and internal wooden and wax moulds were burnt out.

On 15.11.75, the gasifier and regenerator refractory wall temperatures reached 710°C and 570°C respectively, the maxima possible using the propane burner. Further temperature boost was achieved by initiating kerosine injection into the gasifier distributor pit at 15.35 on 15.11.75, and the remainder of the recommended curing schedule completed for the gasifier. A further curing period was then necessary for the regenerator as its temperature lagged the gasifier temperature throughout.

Propane was gradually backed out and replaced with kerosine during this period, and limestone addition was started at 18.00 on 16.11.75. The kerosine was replaced with Heavy Fuel Oil at 15.30 on 11.19.75 with the bed depth in the gasifier at 115 cm (45 ins), and the temperature at 940°C. The gasifier and regenerator thermocouples were inserted so that they protruded 5 cm (2 inch) into their respective limestone beds on 17.11.75 at 15.00 hours.

Curing of other refractory lined channels and vessels, such as the hot gas ducts and cyclones was assumed to be complete when the regenerator curing sequence ended. Since these would not be subject to such extreme thermal and mechanical stress as the main reaction vessels, it was not considered essential to adhere as closely to the recommended curing schedule.

Throughout the period, construction, commissioning of instrumentation, calibration and other preparatory work was being carried out around the pilot unit and its associated subsystems. Shut downs were frequently necessary to accommodate this work safely, and there were also numerous unscheduled shut downs arising from a variety of minor operational and equipment difficulties.

Two major problems were identified during the latter stages of the warm-up. The product gas cyclone drains and fines returns system did not function according to expectations, and secondly, there were signs of erratic delivery from the limestone feed equipment. Some attention was given to both problems, but no satisfactory solutions could be found before gasification was commenced. Both systems subsequently gave persistent trouble virtually throughout the entire running period.

21.11.75 : Day 1 (of gasification)

Gasification was initiated at 23.30 on 20.11.75; day 1 of gasification was taken as 21.11.75. During the early part of the day, conditions were stabilised with rich operation and low bed velocity in the gasifier. Bed depth was initially 99 cm but was increased to approximately 120 cm by 07.30. The stone feed rate was erratic, and eventually stopped completely at 08.30 when the gasifier temperature increased to 940°C as a result.

A high stone feed rate was re-established at 13.30 and conditions were lined out with lean A/F ratio, low bed velocity and a deep bed with the gasifier temperature just above 900°C. Sulphur removal efficiency under these conditions ranged between 80% and 85%.

The regenerator came on stream at 09.30 with a SO_2 level of 4.8% in the off gas.

The fines returns system did not function consistently throughout the day with the right hand cyclone apparently not collecting much solids, and the left hand cyclone not draining efficiently. Gas leaks were plugged around the cyclone off-takes with asbestos rope and refractory.

A leak in the boiler gas sampling line was repaired at 15.45.

Leaks were found also in the manometer connecting lines to the gasifier and regenerator. These were traced and eliminated.

<u>22.11.75</u> : <u>Day 2</u>

Initially, gasifier conditions were lined out with a high stone feed rate, deep bed and rich operation at 22% stoichiometric air. During the day, running conditions were established at low stone addition rate, less than 1 molar, and the gasifier temperature allowed to rise. Towards the latter part of the day, fuel and air rates were increased to establish a higher velocity. Bed depth dropped steadily during the day from 126 cm to 110 cm due to poor stone feed control and unsatisfactory performance from the fines collection and returns system.

The low stone feed rate was caused by excessive damping of the vibrator metering table which only performed even moderately well immediately after it had been thoroughly cleaned.

The fines returns left hand cyclone drain was found to have a valve actuator incorrectly connected to the pneumatic air supply. When this was rectified at 06.15, an immediate response in gasifier bed depth and temperature was seen indicating that fines were draining and being re-injected.

A steam supply and metering system was fabricated and lined through to the gasifier plenum and lid.

Problems with leaks in the boiler gas sampling system persisted and were traced to the joints on a glass moisture knock out-vessel located in a cold box. Attempts were made to eliminate these. Difficulties with the regenerator off gas sampling system were also encountered due to leaks and line plugging.

No readings were taken between 10.30 and 16.30 when work was in hand to improve the stone feed and solids handling systems.

<u>23.11.75</u> : <u>Day 3</u>

With conditions lined out following the changes made during Day 2, the steam feed lines were purged through and the condensate drained in preparation for steam injection into the gasifier. After increasing fuel flow into the gasifier, steam injection was started at maximum rate at 04.45, and continued until 07.40. Results were inconclusive and no significant change in the SO_2 level in the boiler flue gas was observed. Doubts subsequently raised over the functioning of the steam flow meter led to the results taken during this period to be scrapped. Indeed, no further readings of value were taken for the remainder of the day as a number of difficulties occurred more or less together.

The stone feed system was out of action during virtually the whole day with its typical behaviour of giving stone delivery only for a short period immediately after cleaning the hopper and vibrator.

The fines returns system, particularly the right hand cyclone was still not collecting and discharging fines efficiently.

Serious leaks were found in the flue gas recycle baghouse filter housing causing air ingress and the gasifier bed temperature to rise when flue gas was being injected. This was overcome by trimming the flue gas recycle flow until it just leaked out of the filter, but under these conditions, flow was restricted to 85 m^3/hr (50 cfm).

The boiler off gas CO_2 meter was unserviceable and it was replaced by the flue gas recycle CO_2 meter at 04.30. No sensible reading could be obtained and it was suspected that leaks of air into the system were still present.

The regenerator off gas sample line was plugged. The blockage was traced to a stainless steel valve in which a screw-in fitting partially blocked an internal channel thus causing the valve to be susceptible to plugging when passing unfiltered regenerator gas. The valve was modified to eliminate the obstruction.

The regenerator suddenly defluidised for about two hours. It eventually cleared itself, but left the regenerator bottom thermocouple reading approximately 30°C lower than the middle and uppper thermocouples and an accumulation of solid deposits above the regenerator distributor was suspected. This supposition was supported by loss of the regenerator lower pressure signal, the pressure tapping being located adjacent to the lower thermocouple.

24.11.75 : Day 4

Work continued until 04.30 on the problems identified during Day 3. The gasifier Fuel Oil input was then reduced by about 10% to reduce the carbon burden on the stone in order to improve regenerator performance. Flue gas recycle was started at the same time to prevent the bed temperature rising too far. The gasifier bed temperature had dropped to 860°C when a main flame failure occurred at 05.40 without obvious cause.

Attempts to restart were unsuccessful and a burn out over a sulphided bed was carried out to clear the severely restricted cyclone entry ports, particularly on the right hand side. This was completed successfully and the gasifier bed reheated in preparation for further gasification.

During the rebuild of the unit, air bleed jets had been provided at the cyclone inlets to investigate whether continuous carbon burn off could be effected. It was evident that these were unsuccessful in preventing carbon build up - no evidence of carbon removal around the jet entries could be observed through the cyclone entry viewing ports.

It was found that the Fuel Oil circulation through the secondary heating system had been cut off so that cold fuel was being delivered into the gasifier. This was the likely cause of the flame failure and the subsequent difficulties of relighting; the fuel flow would be reduced and its dispersal within the gasifier would be impaired.

Personnel from GCA technology arrived on site and prepared equipment for an assessment of the emissions released from the pilot unit.

25.11.75 : Day 5

Gasification was recommenced, and a boiler flame established at 04.30 at the third attempt.

Low velocity conditions were established in the gasifier with lean operation and a low stone feed rate. Sulphur removal was low initially at 68% but improved to 80% during the day.

Though no flue gas was being recycled to the unit, the recycle system through the baghouse filter was stopped at 06.40 to investigate the appearance of moisture on the outside of the housing, gas leakage, and loss of pressure drop across the filter bags.

The gasifier cyclone drains were working intermittently during this time, resulting in transference of fines into the boiler. The sulphur removal efficiency was low as a consequence, and results confirmed the beneficial effect of fines recirculation on sulphur removal performance. At the same time, the stone feed system was giving its characteristic erratic performance.

At 21.00, a leak was found on the second stage gasifier blower casing, located downstream of the metering orifice plate. It was repaired but no observable difference was noticed on the pressure drop across the orifice plate. Nevertheless, there was a small increase of about 10°C in the gasifier bed temperature, equivalent to approximately 1% increase in the gasifier stoichiometry, or a 13.5 m^3/hr (8 cfm) air flow increase.

Thus, previous readings were taken to indicate a slight over estimate of stoichiometry, and conditions in the gasifier were in fact slightly richer than had been calculated.

26.11.75 : Day 6

At 05.40, the fuel oil was decreased to run leaner at the same bed velocity, and flue gas recycle was started into the gasifier plenum to control the temperature.

Blockages were cleared in both cyclone drain legs during the day and it was becoming apparent that a thorough overhaul of this system was needed. Further problems later in the day at 19.30, including malfunctioning of the control sequencer boxes made a prolonged stoppage inevitable, and the unit was shut down at 20.15. Prior to the shut down, air and flue gas rates to the gasifier were increased to run lean at higher velocity. A sharp increase in the Wosthoff reading on the boiler flue gas from 380 to 470 ppm was observed, but no explanation found. This continued to rise to 600 ppm just prior to the shut down, and may have been due to the failure of the fines collection system.

At 19.35, the gasifier temperature rose by 70°C due to a major air leak into the baghouse filter.

A Hartmann and Braun SO_2 analyser was connected through its own cooler unit to sample the boiler off gas, and it was zeroed and spanned.

27.7.75 : Day 7

The cyclones were again cleared, the sequencer boxes repaired and after reassembly, gasification was restarted at 03.00. A flame out occurred at 04.30. The fuel flow control valve through the secondary heating system was found closed again, so that the fuel delivery was cold. Correct flow was re-instated and after checking both fuel and air flows, a further restart was made at 07.15. A flame out with immediate restart recurred at 08.15 and the flame was subsequently maintained until 10.15.

Flows were again checked and confirmed, repairs were made to regenerator and gasifier sampling lines and a new temperature recorder installed to replace the main gasifier/ regenerator instrument which had become faulty.

Gasification recommenced at 15.00. It proved to be impossible to get the regenerator functioning properly and poor fluidisation and transfer were diagnosed. This resulted eventually in a complete draining of the regenerator at 17.00, and a thorough clearing of the solids transfer ducts.

A minor problem at 17.30 occurred when the nitrogen lance become stuck in the gasifier to regenerator duct, and the gasifier bed had to be slumped in order to retrieve it.

A further period of gasification ensued from 18.45 until 21.00 during which adjustments were made to boiler air in an attempt to improve an unstable flame. No response was observed from the regenerator.

It was decided to shut down at this stage in order to further investigate the persistent problems which had been troublesome from more or less the start of the run. Particular attention was to be given to the fines collection and return systems, stone feed equipment, solids transfer system and it was planned to remove the regenerator distributor for examination.

28.11.75 : Day 8

The gasifier bed was sulphated and subsequently kept hot using kerosine combustion. A burn-out was initiated to clear all ducts and cyclones through to the boiler. Temperature control in the gasifier during burn-out was poor due to insufficient fluidising air. Increasing the air rate stabilised the temperature.

54.5 kg (120 lb) of gasifier bed were removed to drop the bed level below the entry hole to the gasifier to regenerator transfer duct. The regenerator was drained and the distributor removed. Some hard lime deposits were found adhering to the regenerator walls above the distributor, and on the distributor itself. A new regenerator distributor was cast from refractory with three fluidising nozzles to replace the single top hat nozzle used until this time. It was expected that this would provide better fluidisation of the regenerator bed.

A tuyere was inserted through the low level warm up propane burner to enable flue gas to be injected directly into the gasifier bed above the distributor.

The stone feed hopper and vibrator were cleared out. The problems associated with this equipment were traced to accumulations of limestone dust underneath and around the feet of the vibrator table damping the vibrating action. Bleeds were installed to give purge air to blow the dust away from these critical areas.

29.11.75 : Day 9

The solids transfer ducts were cleared by rodding, and an obstruction just below the pocket in the gasifier scraped away. It was decided to install viewing ports into the entry points in both gasifier and regenerator transfer duct pockets and after locating the points for drilling on the casing, jigs were made to support a water cooled, diamond tipped drilling rig in the correct attitude. The fibrous and insulating refractory were drilled out by hand up to the hard, hot face refractory.

The drilling rig was positioned on the gasifier side and a 2.5 cm (1 inch) diameter hole drilled through into the gasifier post box. Entry was effected at 05.45. A similar hole was drilled into the regenerator post box by 07.15. Sight glasses and valves were fitted at both locations.

The transfer ducts were rodded through again and the lance was visible through both sight glasses indicating that the ducts were free.

The cyclone drain systems were overhauled and procedural changes made regarding their maintainence in order to minimise the hold up of draining solids.

30.11.75 : Day 10

The new regenerator distributor was fitted and the air supply, bleeds and pressure tapping fittings reconnected.

The gasifier bed was built up under combusting conditions and stone circulation through the regenerator started. It was not possible to fluidise the regenerator bed until a massive air leak on one of the valves on the air line was repaired. The cold bed was drained from the regenerator and normal fluidisation and stone circulation restored.

Fines recirculation through the gasifier cyclones were improved but occasional blockages were still occurring.

1.12.75 : Day 11

Stone circulation between gasifier and regenerator was still sluggish, though obviously much improved. Further steps were taken to free the ducts but no obvious obstruction could be detected. An investigation was made of the effect of transfer gas flow and pressure and it was found that a large improvement was possible when running at a higher transfer gas pressure. This gave an immediate improvement in the regenerator temperature, showing efficient bed circulation and the regenerator bed stabilised at 140°C below the gasifier bed under combustion conditions.

All analytical systems were checked, zeroed and spanned.

By 18.00, the gasifier and boiler systems were being checked out in preparation for gasification. Heavy Fuel Oil was lined through to the left and right hand side (upper) injectors and the delivery lines purged. A check on flow rate was made and found to be marginally low in comparison with the original pump calibrations.

12.2.75 : Day 12

Gasification was commenced at 05.10 following a strip down of the cyclone drain lock hoppers. The internal chunk traps were found partially blocked with large lumps of carbon resulting from the two burn outs which had been carried out. Points were noted where design improvements could be made for future runs.

The regenerator gas sample line was cleared through after a blockage was detected at 06.00.

The stone feed system was started at 09.25 but the stone was found to be damp and would not flow so that the system had to be emptied and refilled. It was restarted at 11.15.

Lean operations with a deep bed and low velocity were established in the gasifier by about 09.30. Gasifier temperature was at 900°C and a high stone feed rate (approx. 2 molar) was started when the stone feed system was ready. Air flow to the gasifier was increased to maintain bed temperature for the high stone rate. No flue gas was being added. Sulphur removal efficiency was in excess of 80% throughout.

The regenerator came on stream nicely at 07.00 and it was obvious that the problems hitherto had been due to operating the transfer system at too low a gas pressure.

A flame out occurred at 14.15 and several unsuccessful attempts to restart were made until a relight was achieved at 17.30 with adjusted gasifier air and fuel flows.

A systematic investigation of the boiler flame stability dependance on boiler primary, secondary, and main air settings was started and the air flows adjusted to give the optimum flame stability.

A short time cycle was in operation for the cyclone drains and fines returns and the system seemed to be behaving better than at any time so far. Similarly, there was some improvement in the stone feed system after introducing purge air streams around the vibrator table supports.

By the end of the day, it was very obvious that much better control was available as the result of the changes made during the shut down. The unit operation was stable and all subsystems performance was much improved.

3.12.75 : Day 13

A series of different operating conditions were established in the gasifier during the day. Initially, conditions prevailing during the latter part of Day 12 were continued until 04.30 when the bed velocity was increased to run in a leaner mode. At 09.30, the stone feed system was shut down and the bed velocity reduced to restore richer operation. The bed depth dropped and the gasifier temperature increased during this time.

At 13.30, further changes were made when a high stone feed rate was started. There was a consequent drop in bed temperature, and bed depth increased throughout the remainder of the day.

Sulphur removal efficiency remained in excess of 80% throughout.

Minor difficulties occurred in the fines returns system but by 20.00 hrs these appeared to have been overcome and the system stabilised nicely.

The solids transfer system was trouble free throughout the day, the only attention required being to adjust the circulation rate to maintain a high SO_2 level in the regenerator off gas between 07.30 and 15.30.

At the end of the day, the flue gas recycle filter was being purged in preparation for injecting flue gas via the tuyere.

4.12.75 : Day 14

The stone feed was stopped at 00.30 when the usual control problems reappeared. Bed depth at this time was 130 cm. The gasifier temperature increased sharply and though some adjustments were made to the A/F ratio, the gasifier operated under lean conditions until 04.30 when flue gas recycle was started via the tuyere and the fuel rate reduced in order to maintain the gasifier bed temperature. Bed velocity at this time was relatively high due to the flue gas injection.

At 12.30, the final change of the day was made when both air and fuel input were increased to further raise bed velocity at fixed stoichiometry. The objective was to establish the maximum gasifier output, and the limitation, e.g. solids elutriation, boiler capacity, on further increase in output. In the event, the restriction was found to be the capacity of the gasifier air system and with the existing bed depth (123 cm) the maximum air rate possible was only 343 m^3/hr. Fuel rate was 128 kg/hr and the air/fuel ratio 27.4% stoichiometric. Lean operation at high bed velocity resulted.

These conditions were maintained throughout the remainder of the day, and preparations were made to commence injection of fuel oil into the distributor pit.

5.12.75 : Day 15

Day 14 running conditions were continued until 05.30. For the remainder of the day, lean operation at a high bed velocity applied. No stone was added throughout the day and a relatively low bed depth developed as a result. This was accompanied by a gradual deterioration in sulphur removal efficiency from 70% to 60% approximately.

At 03.10, the right hand plenum fuel oil injector was inserted through a shallow trough in the distributor refractory. No difficulty was experienced. Fuel Oil injection was commenced into the pit and an equivalent amount backed off the side injectors. No change in performance was observed and this process was continued in rapid stepwise fashion until all the Fuel Oil was injected in the distributor pit. Experiments were conducted varying the air used to carry the Fuel Oil into the bed, and the penetration of the injector into the pit from flush with the wall to its maximum insertion of 12.5 cm (5 inch) when it encountered the vertical central injector. No significant change in performance could be observed, and the injector was withdrawn so that the entry was flush with the pit wall (09.15). This prevented overheating. It remained in this position until the end of the run.

Flue gas was lined through to the tuyere, and the steam supply system purged in preparation for steam injection.

The regenerator remained in action throughout the day and no problems were encountered with the cyclone drains and fines returns system.

6.12.75 : Day 16

Steam injection was started shortly after midnight at 39 kg/hr (86 lb/hr). At this time, lean operation in a relatively shallow bed (98 cm) had stabilised. Flue gas recycle (through the plenum) was stopped to limit the temperature loss expected for the gasifier bed. The SO_2 level in the boiler off gas increased over a period of about half an hour from 440 to 560 ppm suggesting that the detrimental effect of steam on sulphur removal efficiency is time dependant. Steam was shut off at 04.07 and there was an immediate drop in SO_2 level to 280 ppm.

The air rate to the gasifier was reduced at 05.30 to run richer and when conditions were stable, steam injection was restarted at 40 kg/hr (88 lb/hr) at 06.44. The SO_2 level in the boiler flue gas increased to approximately 600 ppm, and dropped to 500 ppm when steam was shut off at 09.47.

It was clear that steam injection caused poor sulphur removal, and that the effect was greater under richer operation.

Following this experiment, conditions were lined out with a gasifier temperature of 955°C without flue gas injection, with a low bed velocity. These conditions prevailed until the end of the day.

No limestone had been added to the bed since midnight on Day 13, and the fines returns system had worked well throughout. The attrition rate for bed material was calculated at 2.2 kg/hr (4.8 lb/hr) average.

12.7.75 : Day 17

A main flame failure occurred at 05.10 due to seizure of the centre fuel pump. The gasifier bed was sulphated and a burn out started at 06.00 to clear carbon accumulations from the cyclone inlets and ducts.

At 15.00, the gasifier and regenerator were refluidised and it was necessary to rod out the solids transfer ducts to restore circulation.

The gasifier plenum inspection cover was removed at 23.45 and a small quantity of lime - 16 kg (35 lb) - removed showing that stone fall back through the distributor was minimal, particularly in view of the frequent bed slumps which had occurred. The nozzles themselves were cleared using a nitrogen supply, but no change in air flow rate or pressure drop was subsequently observed, indicating that no nozzle blockage had occurred.

8.12.75 : Day 18

Modifications were made to the flue gas recycle and gasifier air systems at 02.15 to arrange the blowers in series thus permitting the gasifier air to be boosted to 240 cfm, and a further small additional improvement was possible by optimising the boiler blower output.

A blockage in the flue gas recycle tuyere was cleared at 04.15 but no flue gas recycle was initiated into the gasifier bed.

Gasification was recommenced at 06.10 using a fuel supply through the right hand plenum injector only. A first time light up was obtained without any difficulty. The limestone feed system was started at 06.35 to replenish the bed which had dropped in level to 91 cm. Problems with this system persisted.

The transfer system seemed to be working satisfactorily, and the regenerator came on stream at 09.30.

From 13.00, the Bitumen supply system was being prepared for use. The Bitumen in the storage tank was heated to 180°C using the gas oil fired tunnel burners and steam was lined through the Bitumen ring main jacket and condensate drained. Bitumen circulation through the ring main was started at 15.00.

The twin Bitumen injection pumps were started up at 16.10 to purge the Bitumen delivery system through to the injector. Unfortunately, at this point the system was closed down temporarily and on restart, it was found that both pump filters were plugged with cold Bitumen and a filter by-pass had to be installed, trace heated and lagged.

9.12.75 : Day 19

The Bitumen delivery pumps were calibrated by delivering Bitumen through the supply lines to the right hand plenum injector. At this point, it was found that the pump delivery on maximum setting was about twice what was required, and moveover, the scale for indicating pump delivery was very coarse. Thus, flow rate could be set only very approximately according to the pump and it was necessary in practice to calculate the Bitumen flow from the boiler flue gas analysis.

Bitumen injection into the unit was started at 02.00 on the right hand plenum injector only, and by 02.30 the gasifier was running on 100% Bitumen. The Heavy Fuel Oil system was shut down.

Almost immediately, the boiler SO_2 reading increased to 1000 ppm, there was a decrease in boiler oxygen from 5% to 3.5%, and the regenerator CO_2 increased sharply with corresponding loss of SO_2 generation indicating over-rich operations. The Bitumen rate was reduced at 02.45 and a decrease in boiler SO_2 level to 800 ppm followed indicating approximately 70% sulphur removal. However, the regenerator performance did not recover, and attention had to be given to the sample lines which were plugged.

At 10.00, 61 kg (134 lb) of stone were withdrawn from the regenerator, and the stone feed system started to replenish the bed. The carbon level remained high despite further trimming of the Bitumen rate and the regenerator did not recover.

A disturbing problem arose when it was observed that the boiler water temperature was increasing steadily. An immediate shut down was not necessary but by 15.00 the temperature had built up to 120°C and a shut down became inevitable. The heat exchanger was the primary suspect for causing the temperature increase.

Preparations were started for a burn back · i.e. burning off the accumulated carbon deposits from the boiler end of the product gas ducting. It was initiated at 18.30 by closing the stack damper, thus causing a boiler air flow back into the gasifier and out through a vent provided in the gasifier lid. Steam was injected to cool the vent pipe whilst the burn back was in progress. The gasifier bed was slumped during this operation.

<u>10.12.75</u> : <u>Day 20</u>

The burn back continued until it became apparent that the carbon deposits in the cyclones were not being attacked. The stack damper was opened, the gasifier bed vent closed and a normal burn out started with intermittent combustion in the gasifier to reheat as necessary. This was completed at 08.00 and it was apparent by visual inspection through the sight glasses that the cyclone entries and internal surfaces were burned clean.

Whilst the burn out was in progress, the flue gas recycle filter bags were examined and found to be covered with a cake of damp fines about 1 cm thick. No solids draining was possible as the drain hopper was packed with wet fines. There were also several large leaks on the filter housing.

This equipment was given a thorough overhaul.

At 09.00 the bed was reheated and the regenerator refluidised after rodding from the top to break up stone and carbon accretions. The carbon burn in the regenerator was obvious when viewed through the regenerator overhead viewing port. The transfer ducts were also checked.

Gasification was initiated on Bitumen fuel at 20.30. A first time light up after 17 seconds was obtained. The regenerator came on stream at midnight.

The flue gas recycle filter was warmed up and at 23.00 flue gas was injected via the tuyere system at 50 m^3/hr (30 cfm). This was increased to 23.30, this was increased to 63 m^3/hr (40 cfm).

A leak was repaired in the boiler gas sampling system at 04.30 and this recurred at 05.30 requiring the moisture trap to be dismantled. Oxygen levels during this period were taken to be those pertaining when the blockages occurred since no fuel or air flow changes were made.

11.12.75 : Day 21

Bitumen gasification continued until 09.45 when a plant shut down was forced by a recurrence of excessive temperature in the boiler primary cooling system. Relatively few difficulties were experienced during the gasification period.

Flue gas recycle flow was increased to 76 m^3/hr (45 cfm) at 01.00 and to 85 m^3/hr (50 cfm) at 03.30. Throughout the period, lean operating conditions with a high gasifier bed temperature in a low bed applied.

Following the 09.45 shut down, the Bitumen system downstream of the pump was purged with gas oil and checks around the boiler cooling system were started. No obvious reason for the excessive temperature excursions could be determined. Provisions were made to improve the boiler pump delivery pressure monitoring system and a successful restart was made on Bitumen at 14.45.

By 20.00, the boiler water temperature was at 116°C and increasing. Pump performance was satisfactory both for the primary and secondary water circulation systems, and it was concluded that the problem lay with the heat exchanger. At 20.30, the plant was shut down again with the boiler water at 118.5°C.

12.12.75 : Day 22

The final experiment planned for the run was to establish in principle whether coal could be injected and gasified successfully. To enable tests to be conducted, a simple coal feed system was constructed using the flue gas recycle tuyere as the injector, and feeding coal via a weir system and fluidiser similar in operation to the system used for fines re-injection. The metered coal would be pneumatically conveyed into the gasifier bed. Above the weir system, coal was fed through a manually replenished lock hopper.

This temporary arrangement was considered sufficient to establish the necessary information as a preliminary step to designing and constructing a more complicated, automatic feed system.

The equipment was assembled and was ready for use at 06.30.

Bitumen gasification was started, again without difficulty, at 07.45, and the coal system primed with a supply of Illinois No.6. Initially, the particle size range was 1400 µ down but other particle size ranges were also available for testing.

Between 08.00 and 15.45, a variety of coal particle size ranges, and feed rates were tried with about 50% of the fuel input being provided by Bitumen.

Major difficulties were encountered with the simple coal feed system, particularly with respect to maintaining a uniform feed rate into the bed. However, the experiment was successful in demonstrating that coal could be injected and gasification could be maintained at a rate sufficient to maintain a stable flame in the boiler. Coal feed rates of approximately 97.5 kg (215 lb/hour) were estimated from the boiler gas analysis.

At 14.50, the coal system failed when a valve jammed and unsuccessful attempts were made to feed coal via the limestone injection system.

The unit was shut down finally at 15.45 and all systems secured for the cooling off stage.

TEST DATA

The major test results for periods when steady conditions prevailed are given in Tables C1-C11.

Table C12 summarises the gasification periods for the various fuels tested, and where possible identifies the reasons for shut down.

Fig. C34 is a chronological plot of unit performance during Run 10.

PERFORMANCE OF EQUIPMENT DURING RUN 10

Introduction

Reference has been made in Appendix A to the design basis, construction and materials of the CAFB pilot unit used during Run 10. Whilst comments on the inspection of the unit and equipment after Run 10 are provided later, the performance of major items of equipment is described below.

1. The Gasifier Distributor

This was designed with a central pit to provide protection for fuel injectors which could be inserted through the refractory wall into the central depression. Included were fluidising nozzles of a new design to minimise the fall back of the lime bed into the gasifier plenum, a phenomenon which tends to occur during the time the gasifier bed is being slumped.

The distributor with its associated nozzles proved to be very successful in all respects. Fuel injectors could be withdrawn and inserted very easily through any lime accumulations in the channels and excellent fluidisation was achieved with virtually no lime fall back with the new nozzle.

2. The Regenerator Distributor

Initially a top hat design was used but this was changed during the run to a three nozzle type due to poor fluidisation performance. In fact, both designs were probably quite satisfactory in operation, the problems encountered arising from an accumulation of deposit attached to the regenerator lower thermocouple and pressure tapping immediately above the distributor.

3. The Refractory Insulation

This was designed to minimise heat losses and escape of product gases from the gasifier, regenerator, cyclones and gas ducting.

The gasifier and regenerator refractory was constructed in layers of different refractory types (see Appendix A) and the fibrous layer immediately adjacent to the steel shell could be purged with nitrogen. In the event, the skin temperature was sufficiently low not to be uncomfortable to the touch and no serious gas leaks occurred.

The cyclones and product gas ducts leading to the boiler were lined with refractory but not of the same layered construction as the main vessels. No purge was provided. Thus they ran rather hotter than the gasifier regenerator shell, but were still not too hot to touch. No gas leaks were observed except initially round the cyclone lids. These were sealed with asbestos and refractory and any remaining small gas leaks were quickly eliminated by carbon lay down in service.

4. The Main Gasifier Cyclones and Drains

These were designed with an external snail entry to provide high efficiency of particle collection with large gas flows and low pressure drop. They performed well throughout the run in cleaning the product gas stream before the boiler except when problems were encountered with the draining of fines out of the cyclones into the re-injection system.

Here, one of the major difficulties was caused by a chunk trap, used to protect the re-injection system from blockage which allowed chunks to accumulate in such a position that they interfered with the discharge of the fines from the cyclones.

5. The Flue Gas Recycle Bag House Filter

This system proved very troublesome throughout the run. The major difficulty was found to be the accumulation of a damp cake of particulates on the fabric of the bag filter during the period when the system was being warmed up. This was sufficiently severe to cause serious restriction in the recirculation rate for flue gas due to the resultant increase in the pressure drop across the filter.

Problems were experienced in draining solids out of the filter housing, again because of moisture and a drain hopper of inappropriate design to allow discharge of the wet solids.

Serious leakage of gas from the housing of the filter also occurred throughout the run.

6. The Solids Transfer System

This proved troublesome only during the early part of the run until the causes of poor gasifier to regenerator bed transfer could be identified and eliminated.

Partial blockage of the transfer ducts, particularly the gasifier to regenerator occurred probably due to condensation during the early phases of the warm up and bed addition. This greatly reduced the transfer rate of bed material, and a further, but lesser problem was the slight reduction in duct cross sectional area when the regenerator and gasifier refractories expanded and moved relative to each other.

The difficulties were fairly easily overcome by rodding out the ducts until they eventually cleared, and by increasing the pressure of the nitrogen used as the transfer medium.

7. The Boiler Gas Sampling System

This was very troublesome throughout the run. It was very prone to springing leaks at the interconnections between the glass vessels used as the moisture knock out system, the condensate was difficult to drain except by completely dismantling the assembly, and the cooler in which the equipment was located occasionally iced up completely due to defective temperative control features.

8. The Limestone Feed Vibrating Table

This did not perform consistently throughout. The difficulty experienced was due to excessive damping of the vibrator by accumulations of limestone fines around the flexible supports, and packing underneath the table itself. Air jets were installed to reduce this and whilst an improvement was observed, the performance was still inadequate.

9. Gas Analysers

The Maihak analysers, in particular the boiler CO_2 analyser, required constant attention. The boiler CO_2 analyser eventually became unserviceable during the run.

The gas sampling limes to the analytical trains were difficult to trace and leaks proved very troublesome to identify and rectify.

INSPECTION OF UNIT AFTER RUN 10

Introduction

Run 10 was carried out with a completely refabricated unit and there was more than usual interest in the post run strip down and inspection to examine the refractory and other internals. Reference to the run log (see above) indicates that the refractory had been subjected to high temperature condition at up to 900°C for a total of approximately 700 hours, including the refractory curing period.

Throughout the operation, shutdowns caused cycling of temperatures from time to time which would tend to aggravate any tendencies of the refractory to crack and spall.

At the end of the run, the unit was shut down and cooled with a sulphided bed immediately following a period when a bitumen/coal mix was used to fuel the gasifier. No burn out was carried out.

Results of Inspection

Gasifier Lid

The only damage visible was to the insulating fibrous refractory immediately below the steel plate, see Fig. C1. This was caused when the asbestos and refractory packing were inserted to seal the lid.

Gasifier Bed

Fig. C2 shows the overhead view of the gasifier bed prior to its removal. The surface is littered with debris (refractory and asbestos rope) falling from the lid surrounds during dismantling.

The bulk of the gasifier bed was free of any agglomerates and was blackish-brown in colour due to the shut down being in the gasification mode, and the presence of ungasified coal.

Gasifier Vessel

Fig. C3 shows the internal condition of the gasifier after removal of the lime bed.

Clearly seen is the dark polished surface produced by the fluidised bed, extending to above the gasifier entry to the transfer duct to the regenerator.

Also shown is a build up of deposit above the gasifier distributor and covering several of the fluidising nozzles. The central pit fuel injector is covered with a carbonaceous deposit.

The end of the flue gas recyle tuyere is visible and it can be seen that this has been damaged and burnt during the run.

Gasifier Refractory

This was generally in excellent condition with only a few cracks of no major signficance visible. No spalling or other damage to the surface could be observed.

Fig. C4 shows the largest crack, between the gasifier and regenerator vessels, where it was widest across the sealing land of the gasifier lid. (Other minor cracks are shown in following photographs).

Cyclone Entries

Fig. C5 shows the cyclone entry ports. Again, some minor cracks had developed and heavy flaky carbon deposits had accummulated in both entries. Details of the right hand cyclone entry are shown in Figs. C6 and C8 and the left hand entry in Figs. C7 and C9. It is suspected that the crack visible in the right hand cyclone formed during the shut down as it is clean and extends through the carbon layer whereas in the left hand cyclone port the crack, which is plugged with carbon and lime, was probably formed whilst the gasifier was hot and in service.

Views looking down the ducts are shown in Figs. C10 (right hand) and C11 (left hand) showing the extent and the flaky nature of the carbon deposits laid down. Some of these would have peeled off the walls and roof of the ducts during the cooling off phase.

Transfer System Entry Port - Gasifier

Fig. C12 shows clearly an accumulation of deposit in the right hand side of the gasifier entry port to the transfer duct carrying material to the regenerator. A partial view of the duct can be seen behind this deposit.

The transfer line from the gasifier to the regenerator was partially blocked at a point approximately 30 cm (12 inch) below the port at the interface between the gasifier and regenerator refractory monoliths. The overlap was estimated to be approximately 0.3 mm (1/8") cold but could be considerably greater when the unit was hot, depending on the relative movement of the two monoliths. The lower discharge port into the regenerator was clear of deposits, but the rodding port into the transfer line was misaligned. These deficiencies undoubtedly contributed to the problems encountered with bed transfer during the run.

Gasifier Distributor

Fig. C13 gives a general view of the gasifier distributor after Run 10. Clearly visible is the channel through which the right hand fuel injector could be inserted, and the central injection point in the pit floor.

The refractory was in excellent condition generally, with only slight damage visible where flaking and minor cracking had occurred at the fuel injector entries, on the outside face of the distributor refractory, probably due to the relatively thin sectional area at this point. They can be seen more clearly in Figs. C14 and C15.

Fig. C16 shows a close-up of the gasifier distributor pit and the hole through which the right hand injector was inserted is visible. Also seen is the erosion which occurred due to the injected fuel on the pit wall, and the blocked holes in the fluidising air nozzles. Approximately 30% of the nozzle holes were plugged at the end of the run

Gasifier Plenum

This was in excellent condition but contained lime which had fallen back through the fluidising nozzles, and also quantities of tarry deposits caused by seepage of fuel down the central pit nozzle.

Regenerator

The regenerator refractory was in excellent condition with no evidence of serious cracking or spalling - Fig. C17.

Regenerator Distributor

Figs. C18 and C19 show the accumulations of material above the regenerator distributor. This is attributed to condensation in the cooler, lower regions of the regenerator during the warm-up. The lower portion of the regenerator is difficult to heat using hot gas circulation and condensed moisture tends to accumulate here. Entry of hot stone from the gasifier contacting the relatively damp walls above the regenerator plenum can then produce the accretions observed. These deposits were found to be anchored firmly by the lower pressure tapping and thermocouple fittings protruding into the regenerator above the plenum.

Regenerator to Gasifier Transfer Line

As for the gasifier, the regenerator transfer port was found to be partially blocked by deposit, and the rodding port was misaligned. There was no evidence of a displacement of the regenerator and gasifier refactory monoliths partly blocking the transfer duct.

Entries

A number of entries into the unit was found to be plugged at the end of the run:

Gasifier pit vertical central fuel injector.

Gasifier upper and lower drain ports.

Upper fuel injection points in the gasifier.

Limestone feed entry was partially plugged.

Regenerator lower pressure tapping.

Regenerator drain port.

Two thermocouples, viz at the gasifier lid, and at the product gas cylone cross duct lid were unserviceable.

Product Gas Ducts and Cylones

Figs. C20-23 show the components of the right hand cyclone after dismantling. Simlar pictorial evidence of the state of the left hand cyclone is shown in Figs. C24-27.

The cyclone entries and exits are characterised by heavy accumulations of flaky carbon deposits, plus some lime, notably at the cyclone inlets, though there was no evidence of restricted gas flow as indicated by the gasifier overhead space pressure. The silicon carbide off-take ducts were intact. The cyclones themselves, and the small collecting hoppers were virtually free of deposits except along the upper surfaces.

Cyclone Fines Recirculation

Fig. C28 shows the left hand fines drain lock hopper with its perforated plate chunk trap in place. This was found to be badly distorted and punctured at the end of the run, and virtually empty of any carbon chunks - a selection of those found is shown on the hopper flange. This chunk trap is thought to have been virtually inoperative, certainly towards the end of the run.

Fig. C29 shows the corresponding right hand lock hopper and trap, and the considerable quantities of chunks retained.

A further deficiency of the system was that the retained pieces of carbon accumulated on the perforated plate immediately above the drain point at the bottom of the lock hopper, thereby considerably hampering the discharge flow of solids.

Boiler Back

Figs. C30 and C31 show general view of the back of the boiler. Some accumulation of material is obvious but this was less than for previous runs. The first pass gas tubes were generally clear and contained only small quantities of dust.

Figs. C32 and C33 show the second pass gas tubes, again these were found to be unrestricted.

A total of 386 Kg (852 lb) of material was collected from the back of the boiler at the end of the run.

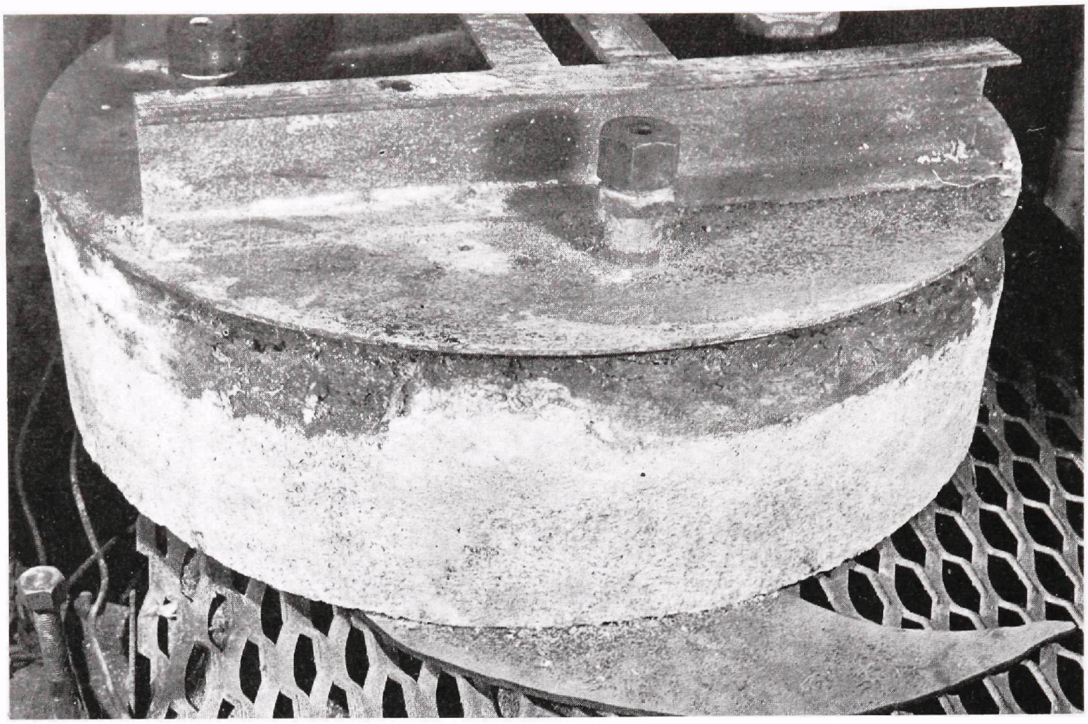


FIG. C1 GASIFIER LID

FIG. C2 OVERHEAD VIEW OF GASIFIER BEFORE BED REMOVAL

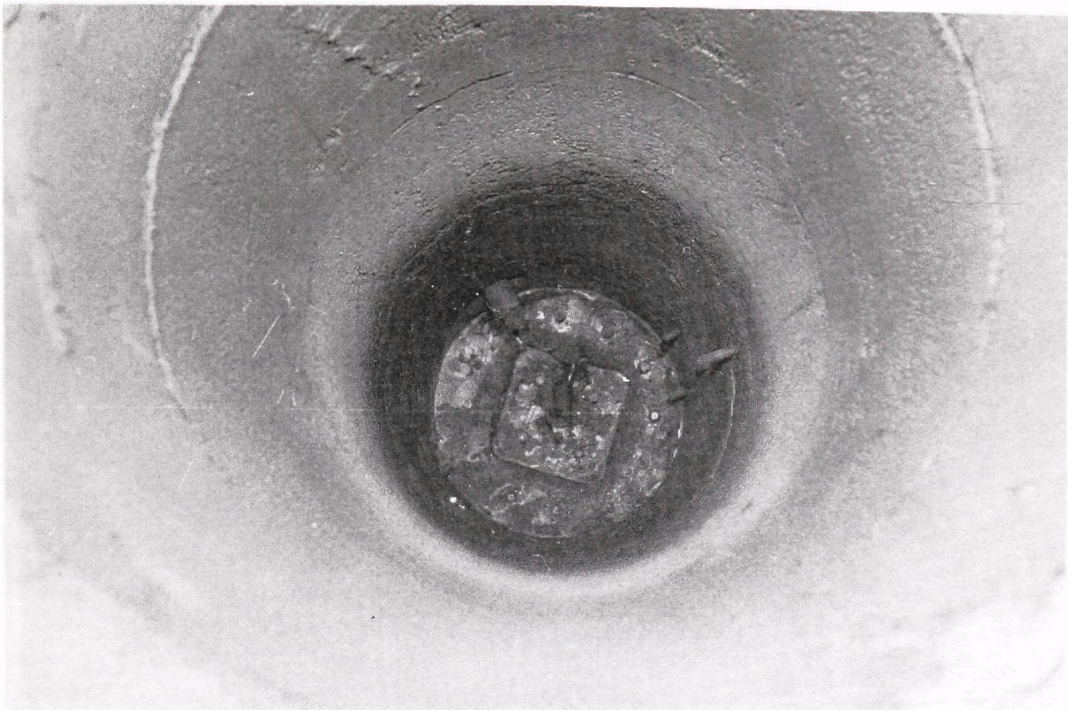


FIG. C3 OVERHEAD VIEW OF GASIFIER AFTER BED REMOVAL

FIG. C4 REFRACTORY CRACK BETWEEN GASIFIER AND REGENERATOR

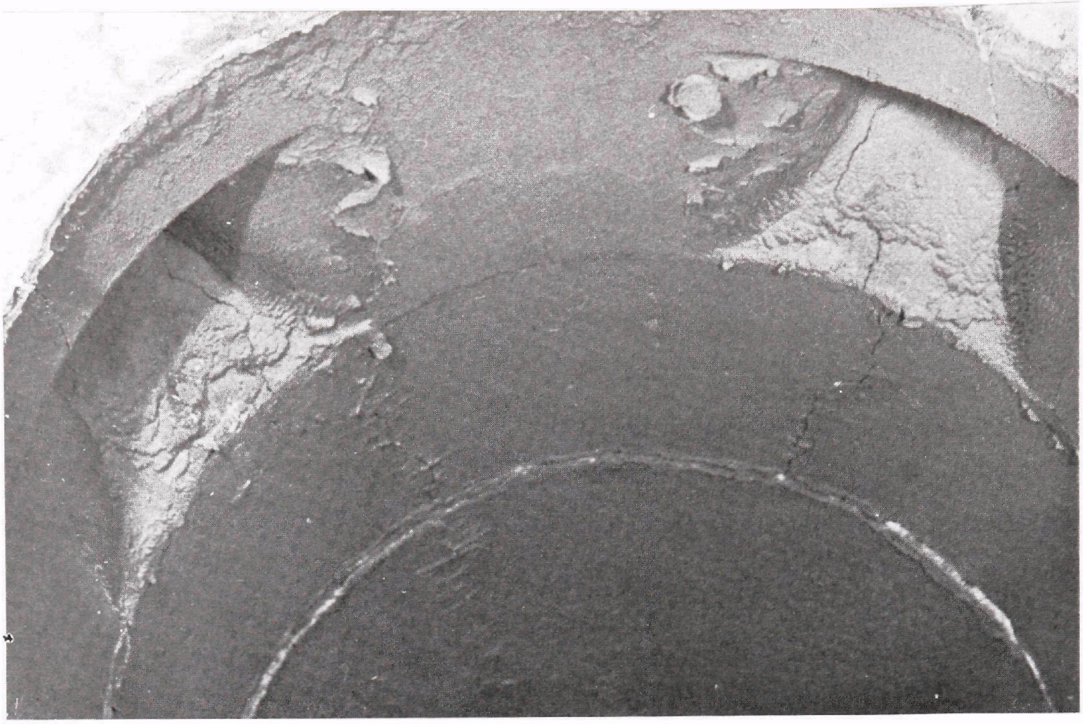


FIG. C5 CYCLONE ENTRY PORTS GENERAL VIEW

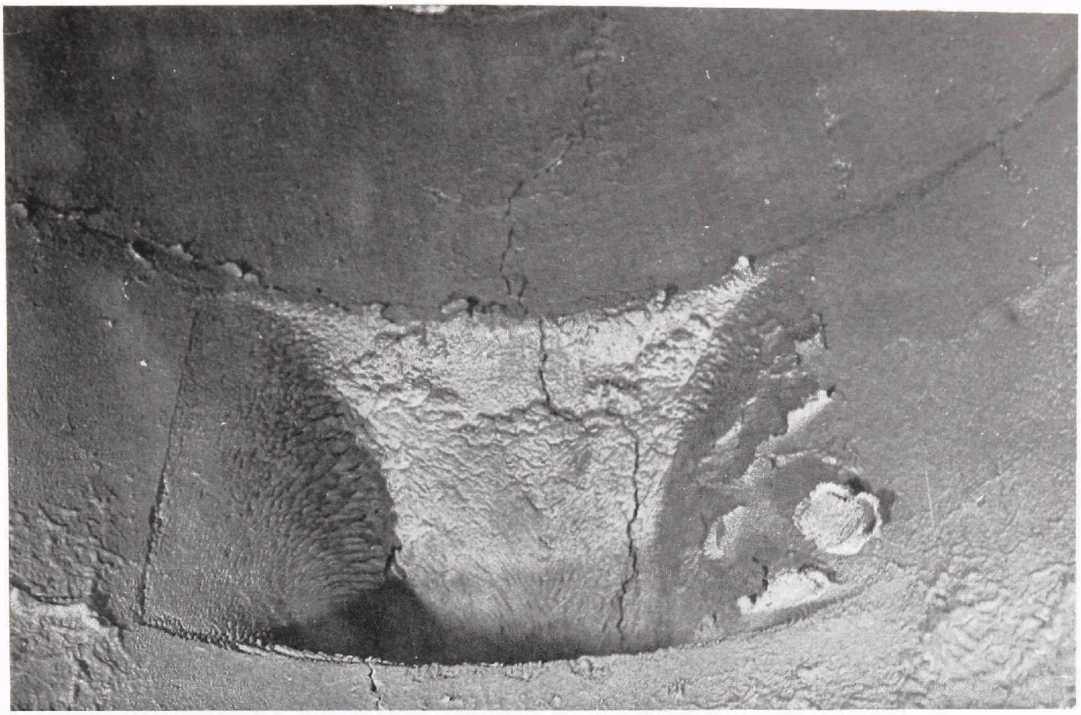


FIG. C6 RIGHT HAND CYCLONE ENTRY PORT

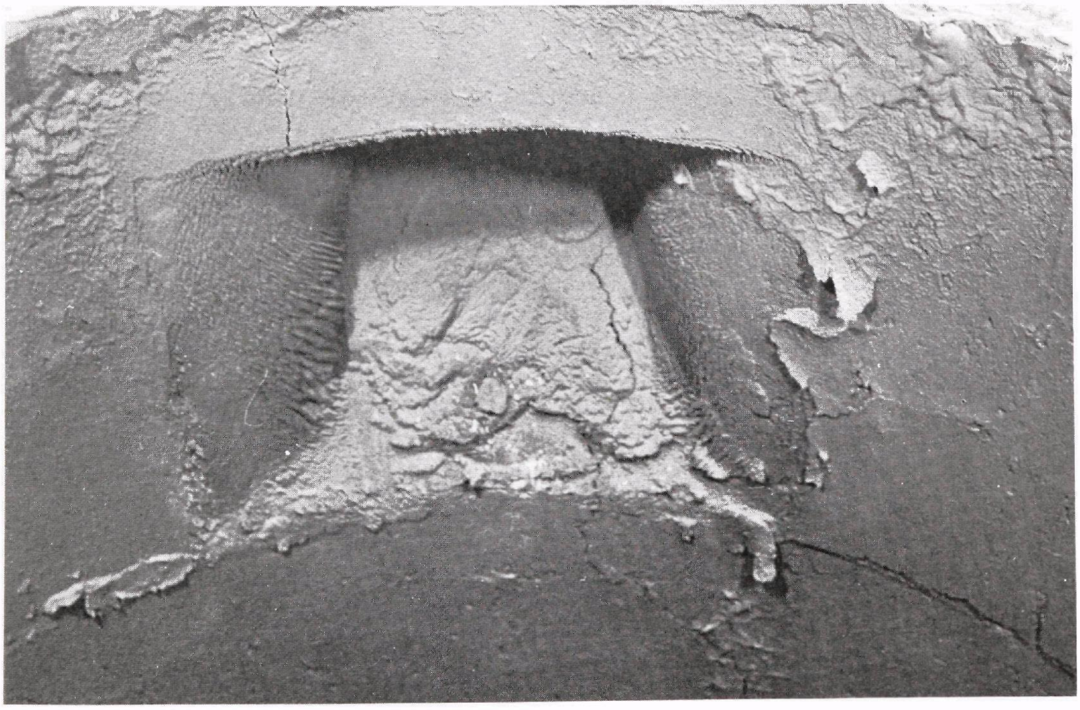


FIG. C7 LEFT HAND CYCLONE ENTRY PORT

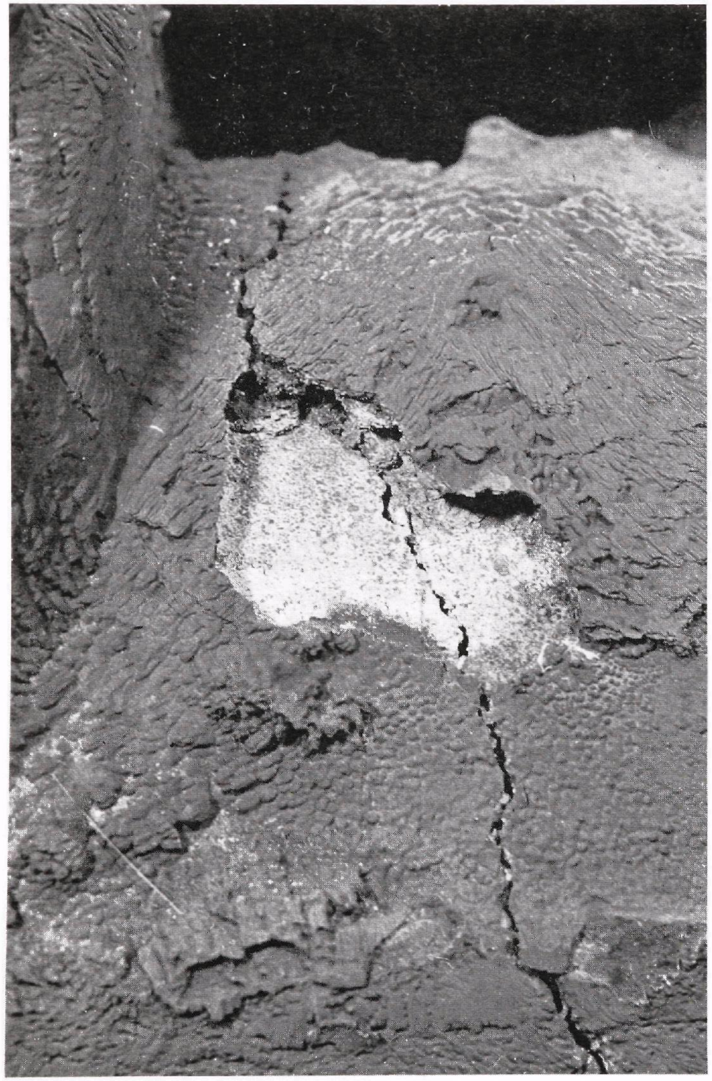


FIG. C8 RIGHT HAND CYCLONE ENTRY PORT SHOWING DEPTH OF CARBON DEPOSIT

FIG. C9 LEFT HAND CYCLONE ENTRY PORT SHOWING DEPTH OF CARBON DEPOSIT

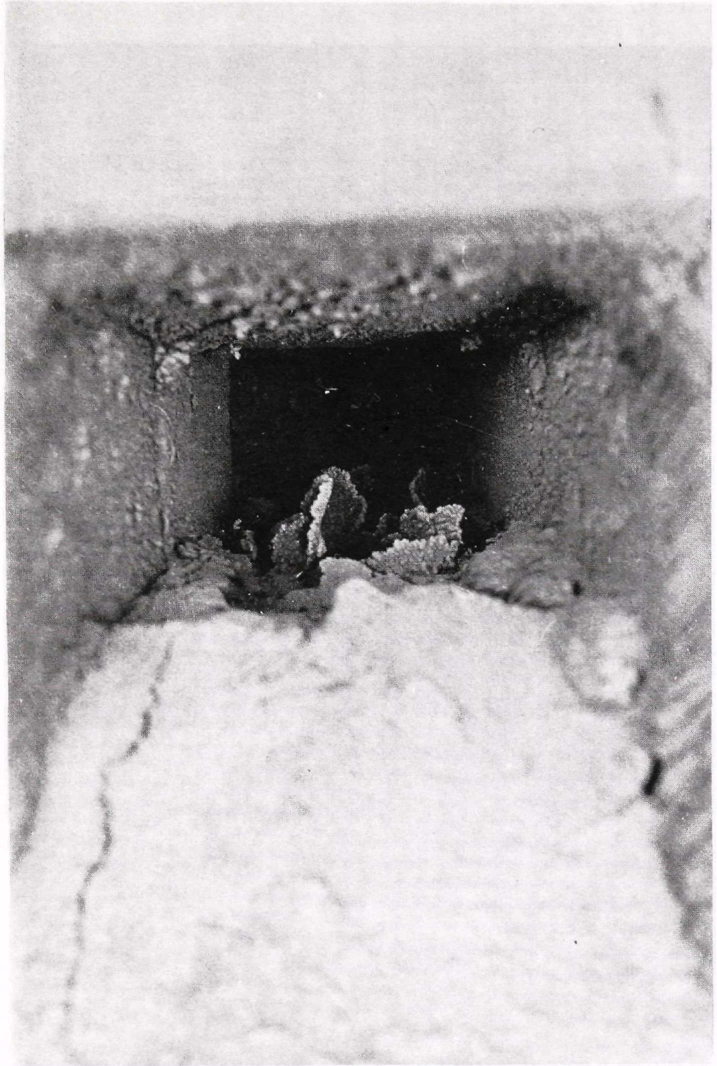


FIG. C10 RIGHT HAND CYCLONE DUCT

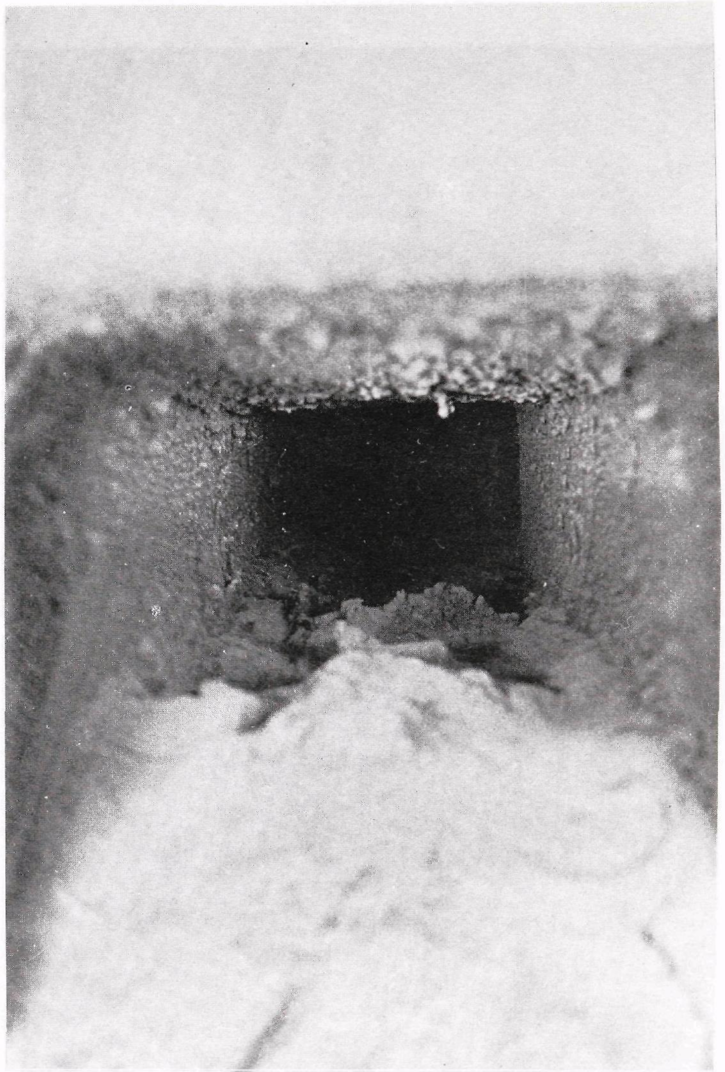


FIG. C11 LEFT HAND CYCLONE DUCT

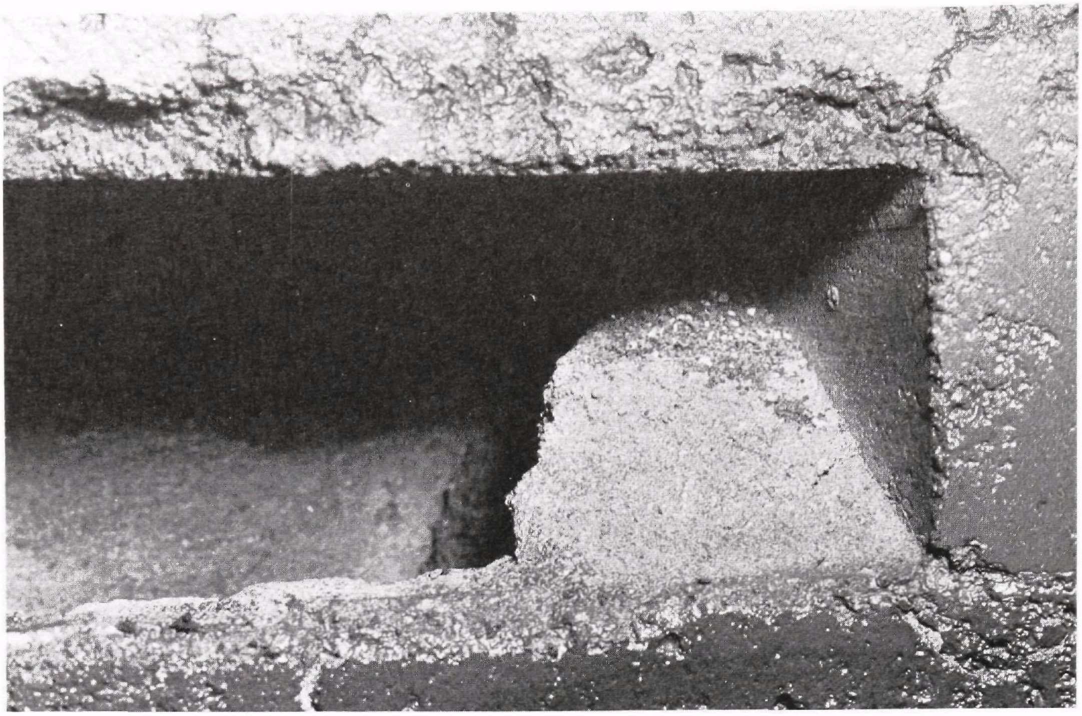


FIG. C12 GASIFIER TO REGENERATOR TRANSFER DUCT ENTRY IN GASIFIER

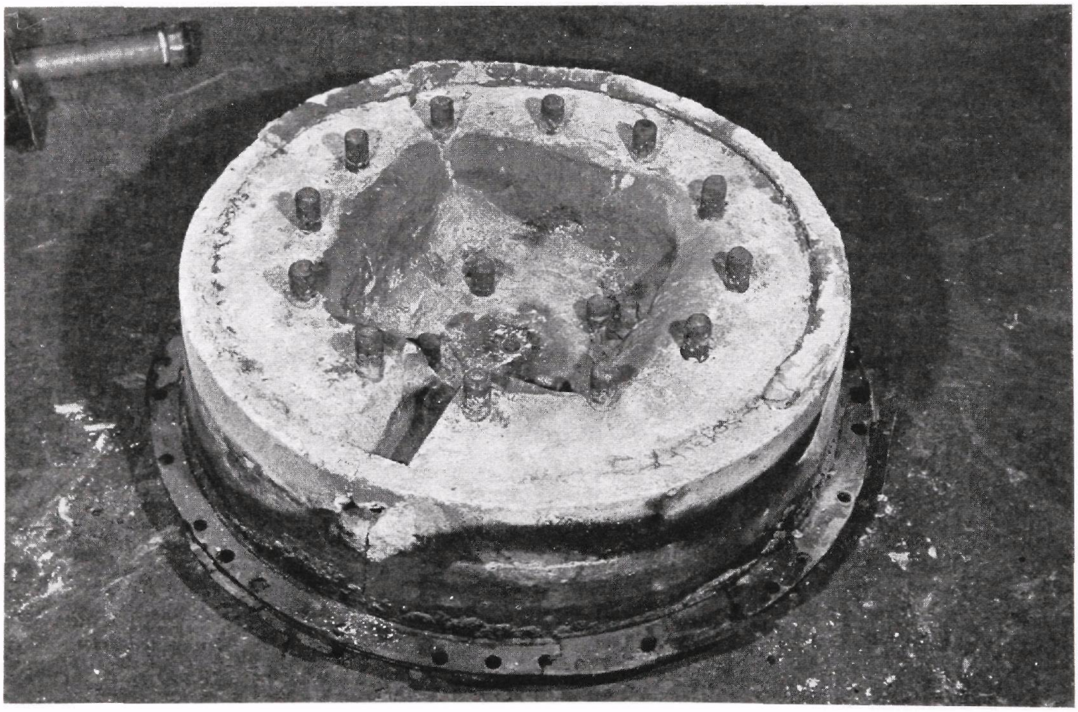


FIG. C13 GASIFIER AIR DISTRIBUTOR, GENERAL VIEW

FIG. C14 V-CHANNEL PLENUM FUEL INJECTOR ENTRY

FIG. C15 PLENUM FUEL INJECTOR ENTRY HOLE

FIG. C16 GASIFIER DISTRIBUTOR PIT SHOWING FUEL JET EROSION OF WALL AND PLUGGED NOZZLES

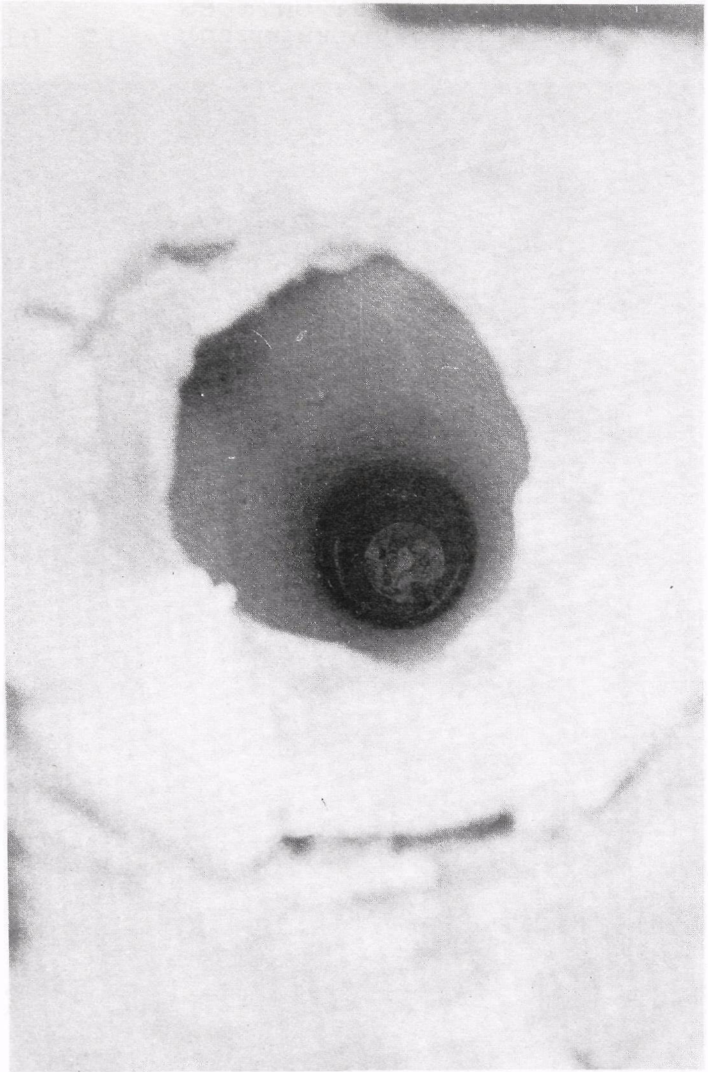


FIG. C17 REGENERATOR WITH THE DISTRIBUTOR IN PLACE

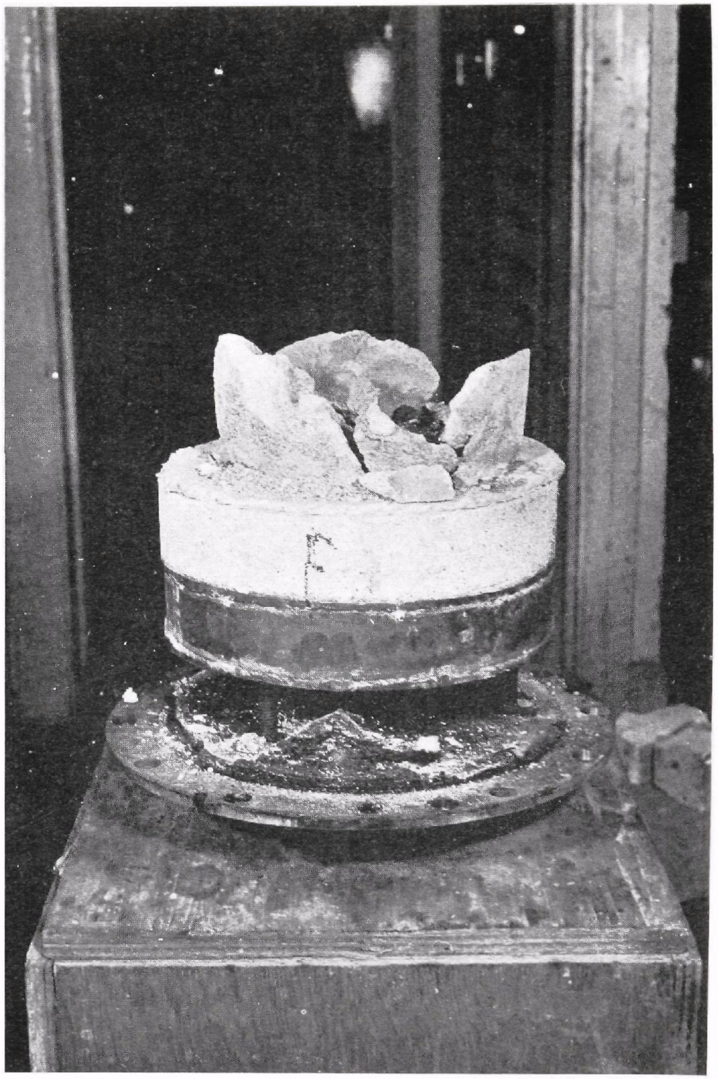


FIG. C18 REGENERATOR DISTRIBUTOR SHOWING LIME ACCUMULATION

FIG. C19 REGENERATOR DISTRIBUTOR SHOWING LIME ACCUMULATION

FIG. C20 RIGHT HAND CYCLONE EXIT SHOWING SILICON CARBIDE TUBE AND CARBON DEPOSITS

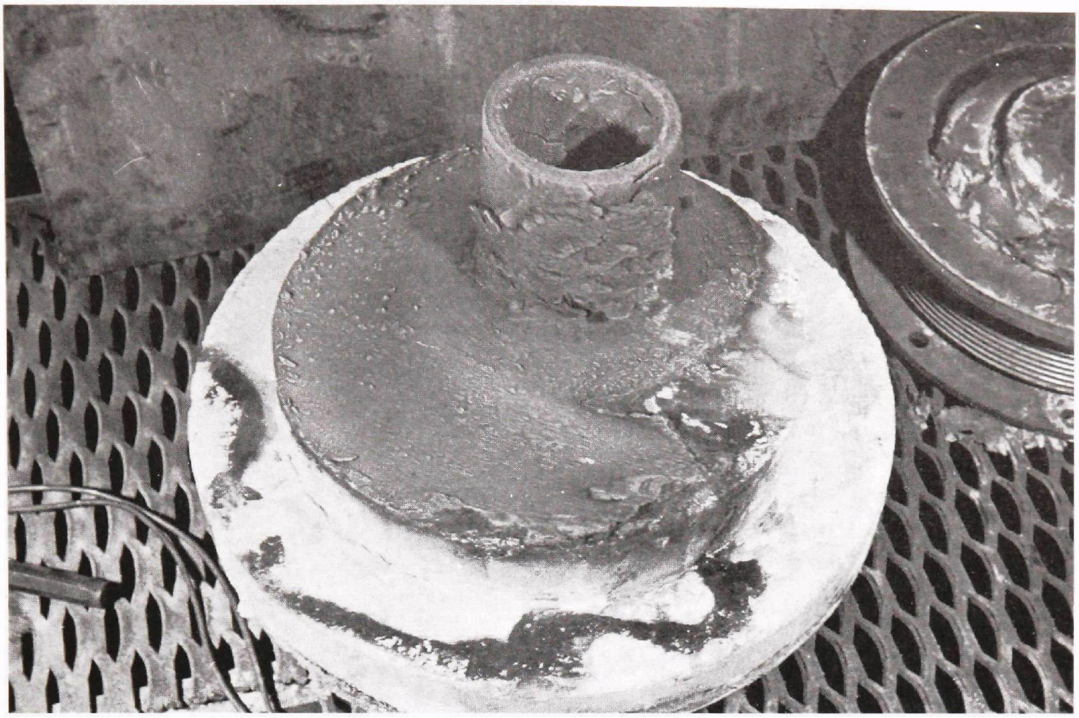


FIG. C21 RIGHT HAND CYCLONE LID

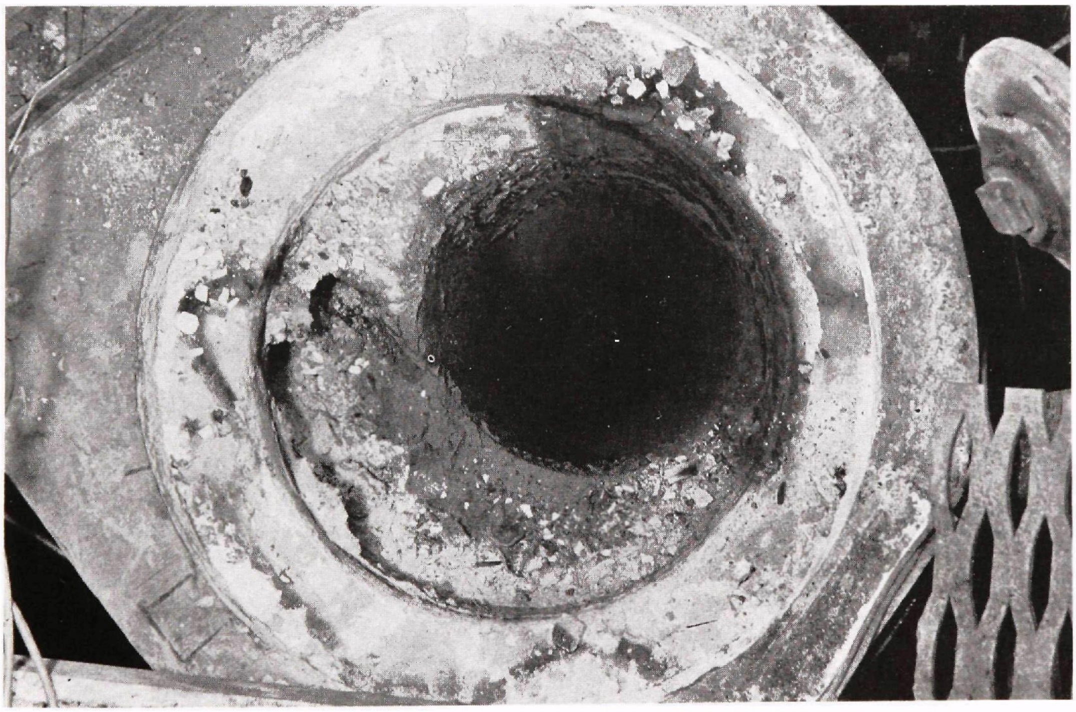


FIG. C22 RIGHT HAND CYCLONE ENTRY

FIG. C23 INTERNAL VIEW OF RIGHT HAND CYCLONE SHOWING COLLECTION POT

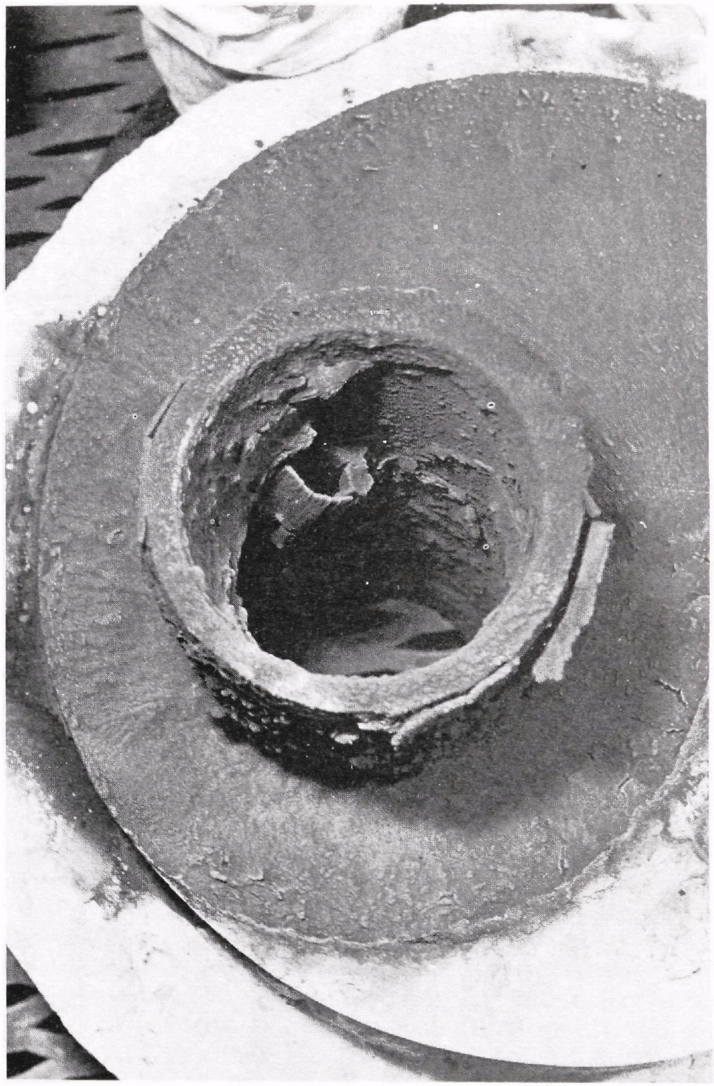


FIG. C24 LEFT HAND CYCLONE EXIT SHOWING SILICON CARBIDE TUBE AND CARBON DEPOSITS

FIG. C25 LEFT HAND CYCLONE LID

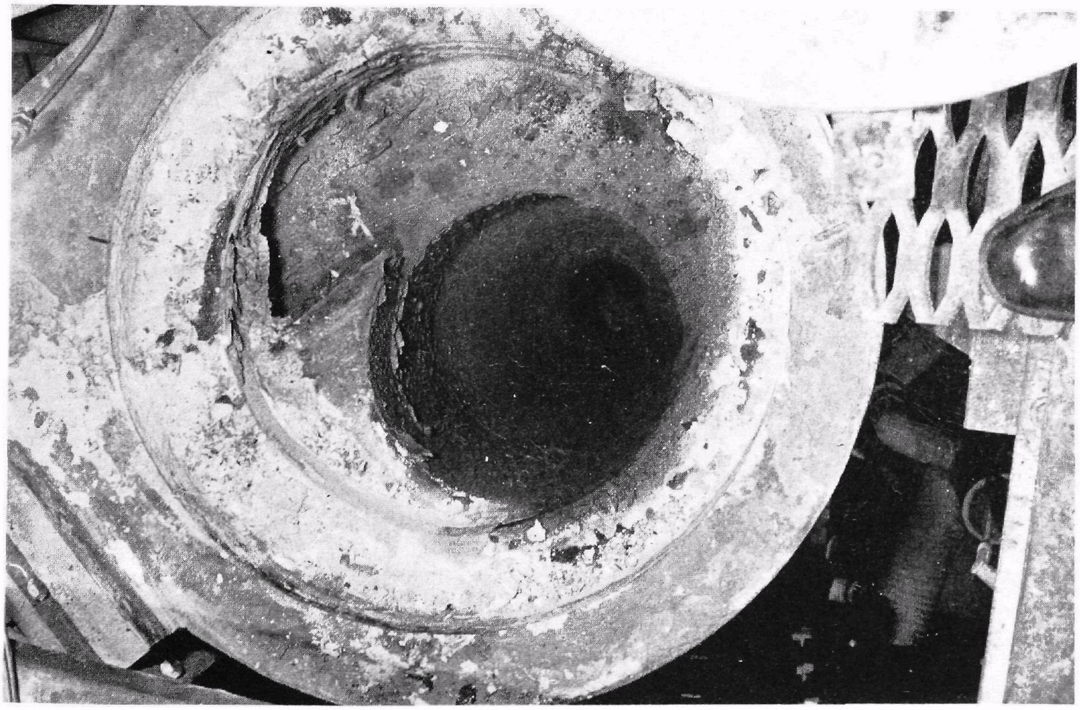


FIG. C26 LEFT HAND CYCLONE ENTRY

FIG. C27 INTERNAL VIEW OF LEFT HAND CYCLONE SHOWING COLLECTION POT

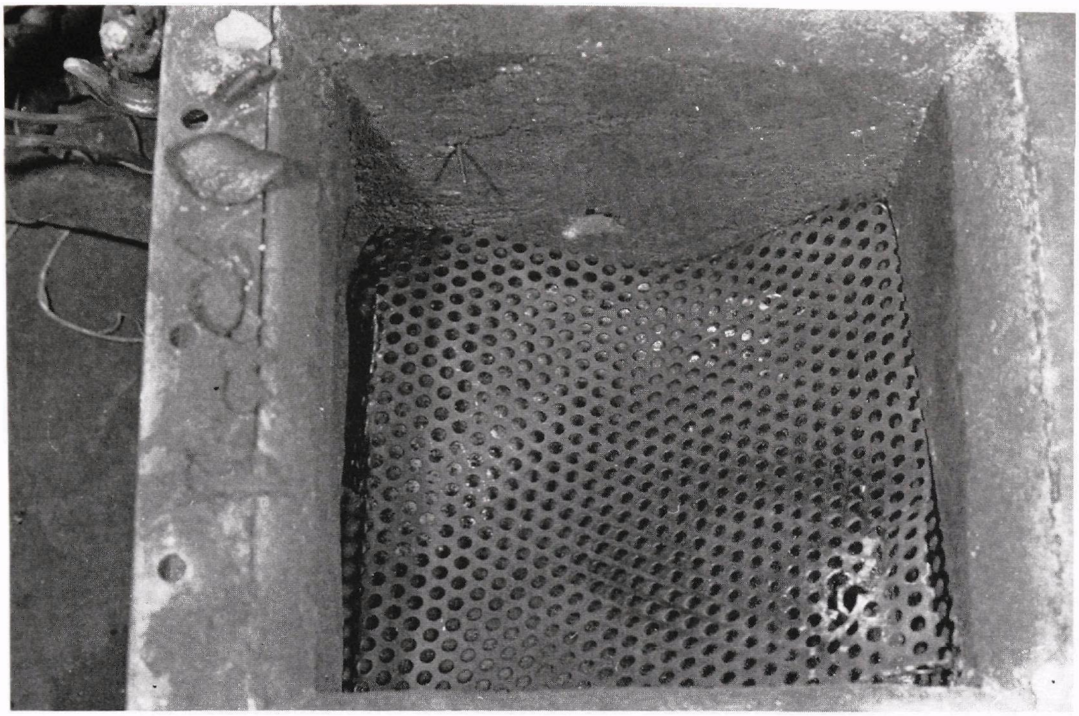


FIG. C28 LEFT HAND CYCLONE DRAIN HOPPER WITH PERFORATED PLATE CHUNK TRAP SHOWING DAMAGE

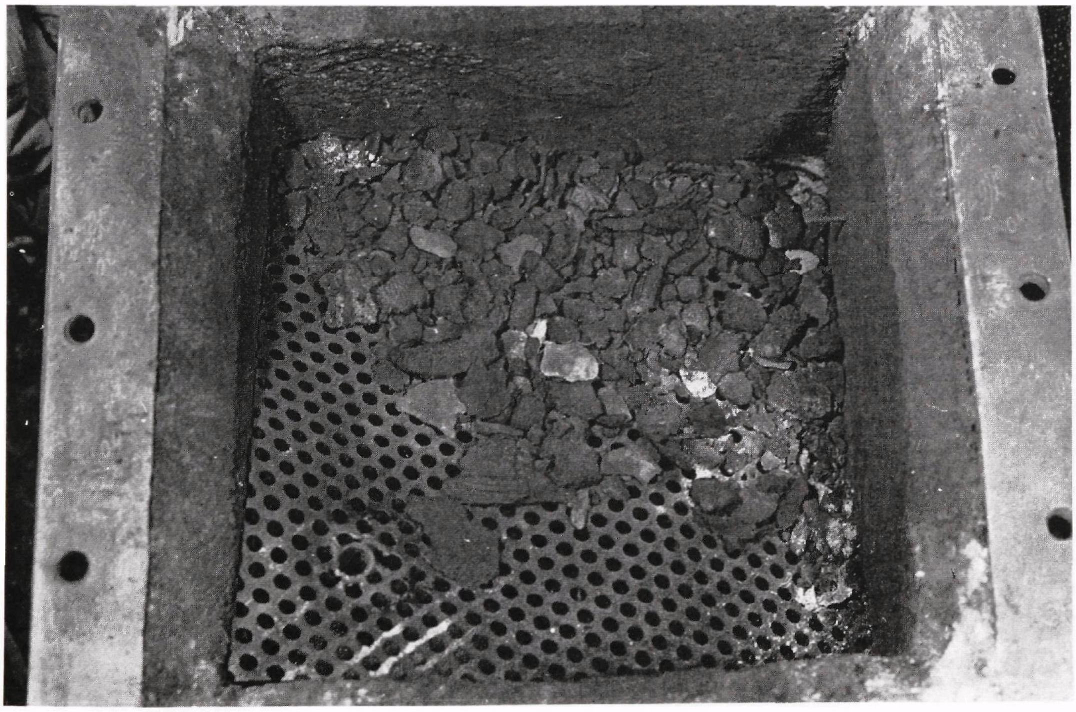


FIG. C29 RIGHT HAND CYCLONE DRAIN HOPPER WITH PERFORATED PLATE CHUNK TRAP AND RETAINED MATERIAL

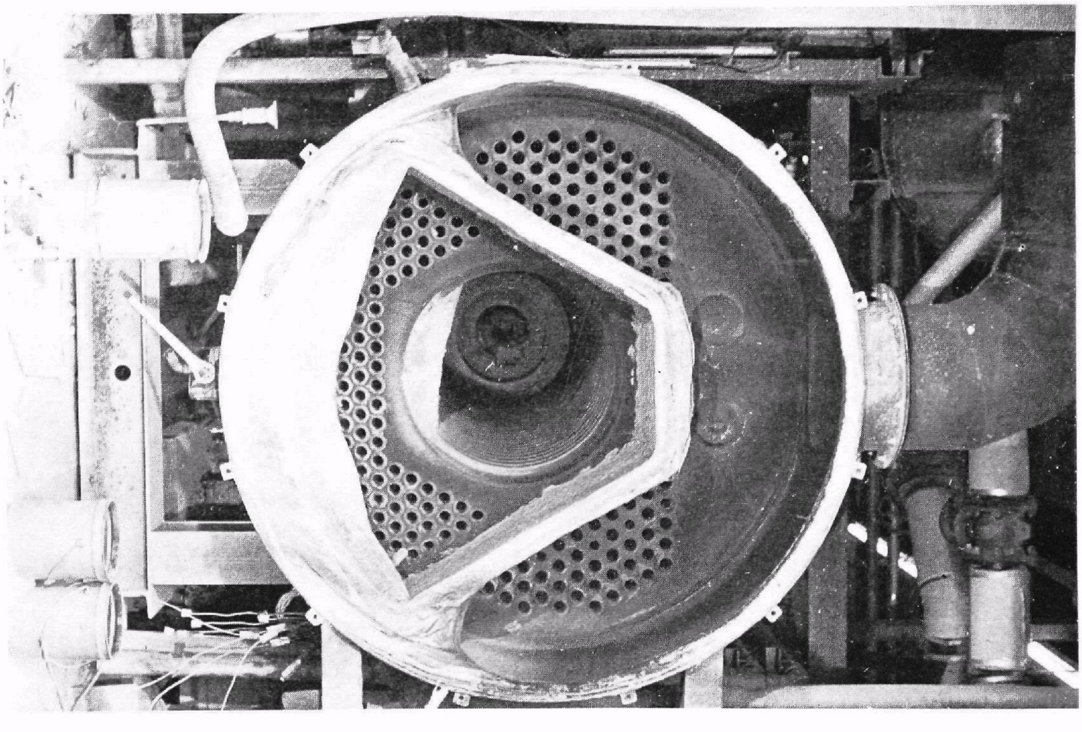


FIG. C30 BOILER BACK, GENERAL VIEW

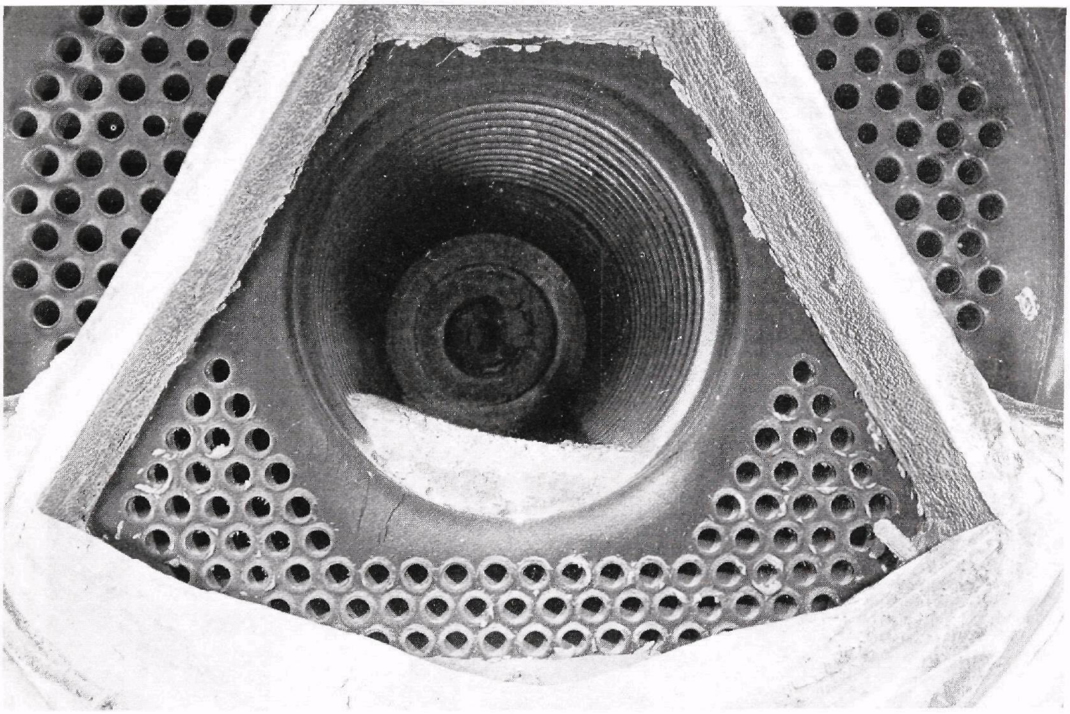


FIG. C31 BOILER BACK SHOWING CLOSE-UP OF FIRST PASS FIRE TUBE ENTRIES

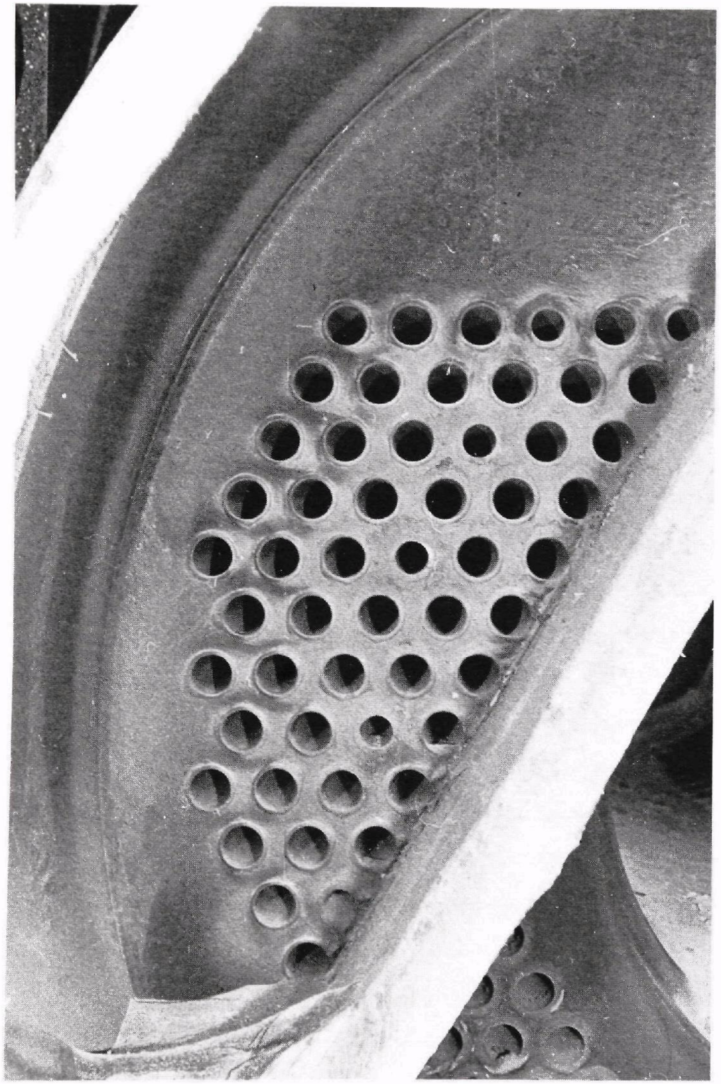


FIG. C32 BOILER BACK RIGHT HAND SIDE SHOWING SECOND PASS FIRE TUBE EXITS

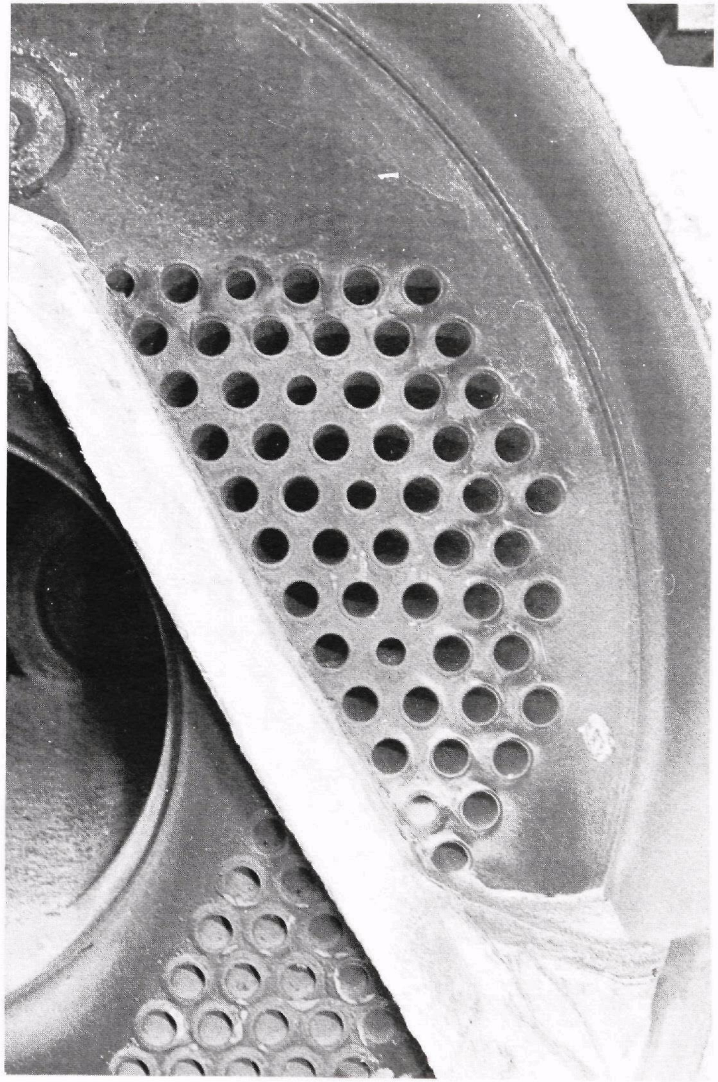


FIG. C33 BOILER BACK LEFT HAND SIDE SHOWING SECOND PASS FIRE TUBE EXITS

APPENDIX C: TABLE 1.

RUN 10: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	CYCLONES	FEED RATE KG/HR OIL	STONE
1.0130	945.	1020.	0.	255.	133.6	0.
1.0230	952.	1027.	0.	268.	133.6	4.5
1.0330	928.	1022.	0.	258.	133.6	22.7
1.0430	962.	1055.	0.	329.	133.6	10.0
1.0530	940.	1055.	0.	316.	133.6	53.3
1.0630	900.	1055.	0.	278.	133.6	67.1
1.0730	875.	1055.	0.	310.	133.6	34.9
1.0830	885.	1055.	0.	390.	133.6	44.5
1.0930	920.	1055.	0.	385.	133.6	0.
1.1030	922.	1055.	0.	387.	133.6	0.
1.1130	940.	1055.	0.	413.	133.6	0.
1.1230	942.	1055.	0.	410.	136.1	6.8
1.1330	920.	1052.	0.	390.	136.1	13.6
1.1430	905.	1055.	0.	345.	136.5	27.4
1.1530	900.	1045.	0.	428.	136.5	26.8
1.1630	905.	1053.	0.	381.	135.3	23.6
1.1730	905.	1055.	0.	443.	135.3	23.1
1.1830	907.	1055.	0.	425.	135.3	19.5
1.1930	920.	1055.	0.	320.	135.3	18.6
1.2030	910.	1052.	0.	315.	135.3	20.0
1.2130	918.	1048.	0.	315.	135.3	0.
1.2230	930.	1050.	0.	357.	135.3	5.9
1.2330	913.	1045.	0.	308.	135.3	24.9
2.0030	900.	1047.	0.	353.	135.3	33.1
2.0130	915.	1050.	0.	356.	135.3	0.
2.0230	928.	1052.	0.	362.	135.3	8.2
2.0330	940.	1055.	0.	316.	135.3	0.
2.0430	960.	1055.	0.	295.	135.3	3.4
2.0530	952.	1055.	0.	218.	139.4	8.4
2.0630	958.	1055.	0.	178.	139.4	5.9
2.0730	968.	1057.	0.	173.	139.4	3.6
2.0830	952.	1055.	0.	134.	139.4	2.3
2.0930	952.	1055.	0.	114.	139.4	0.
2.1030	955.	1056.	0.	82.	139.4	0.
SHUT DOWN AT 2.1030 FOR 6 HOURS						
2.1630	932.	1055.	0.	270.	143.5	3.2
2.1730	952.	1060.	0.	305.	143.9	0.5

APPENDIX C: TABLE 1.

RUN 10: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	CYCLONES	FEED RATE KG/HR OIL	STONE
2.1830	962.	1065.	0.	278.	143.5	0.
2.1930	960.	1072.	0.	263.	154.3	5.9
2.2030	955.	1072.	0.	258.	154.3	4.1
2.2130	955.	1075.	0.	250.	154.3	12.7
2.2230	950.	1075.	0.	265.	154.3	6.8
2.2330	960.	1075.	0.	250.	154.3	0.
SHUT DOWN AT 2.2330 FOR 54 HOURS						
5.0630	905.	1015.	0.	145.	112.2	7.5
5.0730	925.	1032.	0.	275.	112.2	6.1
5.0830	932.	1060.	0.	260.	112.2	4.1
5.0930	932.	1070.	0.	250.	119.2	11.1
5.1030	922.	1066.	0.	245.	122.1	7.0
5.1130	920.	1060.	0.	233.	122.1	5.0
5.1230	925.	1063.	0.	248.	122.1	3.6
5.1330	910.	1060.	0.	363.	122.1	1.8
5.1430	898.	1052.	0.	220.	132.8	5.9
5.1530	880.	1056.	0.	203.	132.8	20.0
5.1630	900.	1055.	0.	203.	132.8	16.8
5.1730	904.	1060.	0.	150.	129.5	11.8
5.1830	892.	1060.	0.	180.	129.5	10.4
5.1930	895.	1058.	0.	200.	129.5	11.8
5.2030	898.	1065.	0.	185.	129.5	12.2
5.2130	898.	1065.	0.	203.	129.5	12.7
5.2230	910.	1070.	0.	213.	129.5	6.8
5.2330	916.	1075.	0.	225.	129.5	6.6
6.0030	915.	1075.	0.	270.	129.5	8.6
6.0130	917.	1077.	0.	235.	129.5	7.9
6.0230	925.	1078.	0.	235.	129.5	1.8
6.0330	940.	1085.	0.	245.	129.5	1.1
6.0430	938.	1082.	20.	210.	116.7	1.1
6.0530	932.	1082.	35.	215.	101.9	3.2
6.0630	910.	1090.	50.	193.	101.9	4.5
6.0730	915.	1080.	58.	188.	101.9	5.0
6.0830	920.	1063.	55.	195.	101.9	3.2
6.0930	922.	1063.	50.	160.	101.9	4.8
6.1030	920.	1067.	50.	130.	103.5	19.3
6.1130	922.	1067.	45.	108.	103.5	17.5

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	CYCLONES	FEED RATE KG/HR OIL	STONE
6.1230	928.	1066.	45.	88.	104.3	6.4
6.1330	930.	1070.	42.	78.	104.3	4.8
6.1430	934.	1070.	45.	70.	110.5	3.9
SHUT DOWN AT 6.1430 FOR 134 HOURS						
12.0530	900.	1000.	0.	290.	142.3	0.
12.0630	918.	1025.	0.	286.	127.0	0.
12.0730	910.	1043.	0.	282.	136.5	23.8
12.0830	882.	1052.	0.	382.	135.7	45.4
12.0930	908.	1065.	0.	346.	142.3	0.
12.1030	938.	1076.	0.	322.	142.3	1.8
12.1130	920.	1082.	0.	359.	142.3	13.8
12.1230	936.	1087.	0.	314.	142.3	23.4
12.1330	914.	1085.	0.	401.	142.3	9.8
SHUT DOWN AT 12.1330 FOR 5 HOURS						
12.1830	905.	1058.	0.	393.	121.7	23.6
12.1930	903.	1054.	0.	455.	134.5	22.0
12.2030	895.	1047.	0.	447.	133.6	22.0
12.2130	890.	1037.	0.	445.	133.6	25.4
12.2230	888.	1035.	0.	460.	135.3	22.2
12.2330	900.	1035.	0.	455.	133.6	17.9
13.0030	906.	1045.	0.	430.	133.2	24.9
13.0130	914.	1050.	0.	330.	133.6	17.2
13.0230	884.	1018.	0.	505.	135.3	24.3
13.0330	880.	1028.	0.	500.	135.3	33.3
13.0430	864.	1032.	0.	455.	135.7	29.9
13.0530	898.	1038.	0.	460.	135.7	27.9
13.0630	938.	1054.	0.	555.	132.8	29.7
13.0730	910.	1047.	0.	525.	133.2	0.
13.0830	895.	1045.	70.	410.	134.5	13.2
13.0930	925.	1047.	0.	253.	131.6	3.2
13.1030	956.	1077.	0.	193.	131.6	0.
13.1130	958.	1084.	0.	203.	131.6	0.
13.1230	925.	1055.	0.	157.	131.6	0.
13.1330	915.	1060.	0.	140.	131.6	24.7

APPENDIX C: TABLE 1.
RUN 10: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	CYCLONES	FEED RATE KG/HR OIL	STONE
13.1430	912.	1062.	0.	170.	131.6	24.9
13.1530	908.	1054.	0.	195.	131.6	25.9
13.1630	905.	1058.	0.	255.	132.4	31.1
13.1730	902.	1062.	0.	210.	133.2	33.6
13.1830	900.	1064.	0.	193.	133.2	32.7
13.1930	905.	1055.	0.	330.	133.2	12.5
13.2030	893.	1044.	0.	475.	133.6	40.4
13.2130	892.	1035.	0.	440.	132.4	27.9
13.2230	897.	1046.	0.	448.	131.6	16.3
13.2330	898.	1058.	0.	460.	127.4	26.5
14.0030	910.	1068.	0.	440.	132.4	0.
14.0130	935.	1105.	0.	438.	131.2	0.
14.0230	942.	1082.	0.	433.	132.4	0.
14.0330	946.	1060.	0.	425.	132.8	0.
14.0430	948.	1100.	0.	463.	132.4	0.
14.0530	924.	1115.	70.	448.	124.6	0.
14.0630	934.	1110.	80.	450.	123.3	0.
14.0730	930.	1100.	80.	413.	122.1	0.
14.0830	927.	1102.	80.	393.	122.1	0.
14.0930	930.	1100.	75.	380.	122.1	0.
14.1030	935.	1065.	70.	368.	122.5	0.
14.1130	928.	1068.	70.	410.	122.1	0.
14.1230	885.	1075.	75.	428.	133.6	0.
14.1330	885.	1078.	80.	410.	137.3	0.
14.1430	888.	1062.	80.	400.	116.7	0.
14.1530	902.	1050.	90.	400.	127.9	0.
14.1630	916.	1055.	90.	398.	126.2	0.
14.1730	918.	1062.	90.	400.	124.6	0.
14.1830	920.	1045.	90.	380.	125.0	0.
14.1930	920.	1035.	85.	373.	124.1	0.
14.2030	920.	1034.	85.	355.	124.1	0.
14.2130	920.	1033.	85.	350.	124.1	0.
14.2230	922.	1040.	85.	338.	123.7	0.
14.2330	921.	1070.	85.	343.	124.1	0.
15.0030	924.	1070.	85.	338.	123.7	0.
15.0130	925.	1070.	85.	308.	124.1	0.
15.0230	932.	1072.	85.	325.	123.7	0.
15.0330	937.	1075.	85.	340.	123.7	0.
15.0430	920.	1068.	85.	323.	123.7	0.
15.0530	928.	1068.	85.	310.	122.9	0.

APPENDIX C: TABLE 1.
RUN 10: TEMPERATURES AND FEED RATES PAGE 5 OF 6

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	CYCLONES	FEED RATE KG/HR OIL	STONE
15.0630	921.	1068.	85.	303.	122.5	0.
15.0730	908.	1053.	85.	288.	121.7	0.
15.0830	901.	1048.	85.	265.	122.9	0.
15.0930	901.	1057.	75.	255.	120.8	0.
15.1030	900.	1055.	75.	255.	126.2	0.
15.1130	892.	1055.	80.	230.	122.9	0.
15.1230	896.	1050.	75.	220.	122.9	0.
15.1330	900.	1053.	40.	220.	125.8	0.
15.1430	901.	1062.	75.	225.	125.8	0.
15.1530	901.	1071.	90.	210.	116.3	0.
15.1630	905.	1078.	90.	210.	118.8	0.
15.1730	906.	1078.	85.	205.	128.3	0.
15.1830	907.	1079.	90.	205.	118.8	0.
15.1930	909.	1075.	85.	210.	122.1	0.
15.2030	917.	1079.	85.	210.	122.5	0.
15.2130	925.	1080.	85.	215.	124.1	0.
15.2230	927.	1084.	85.	225.	127.0	0.
15.2330	913.	1088.	85.	220.	122.1	0.
16.0030	918.	1095.	75.	190.	122.1	0.
16.0130	925.	1103.	0.	175.	122.5	0.
16.0230	912.	1110.	0.	170.	121.3	0.
16.0330	915.	1080.	0.	180.	121.7	0.
16.0430	891.	1050.	0.	170.	119.6	0.
16.0530	889.	1048.	0.	150.	123.7	0.
16.0630	878.	1045.	0.	135.	122.5	0.
16.0730	866.	1060.	0.	120.	122.5	0.
16.0830	862.	1060.	0.	110.	123.3	0.
16.0930	885.	1060.	0.	100.	122.5	0.
16.1030	927.	1063.	0.	95.	121.3	0.
16.1130	937.	1082.	0.	90.	120.0	0.
16.1230	945.	1093.	0.	90.	123.7	0.
16.1330	950.	1076.	0.	95.	125.0	0.
16.1430	955.	1078.	0.	95.	124.1	0.
16.1530	960.	1075.	0.	95.	123.7	0.
16.1630	958.	1070.	0.	103.	119.2	0.
16.1730	955.	1065.	0.	98.	123.3	0.
16.1830	955.	1063.	0.	103.	124.1	0.

APPENDIX C: TABLE 1.

RUN 10: TEMPERATURES AND FEED RATES

DAY.HOUR	TEMPERATURE, DEG. C. GASIFIER	REGEN.	RECYCLE	CYCLONES	FEED RATE KG/HR OIL	STONE

SHUT DOWN AT 16.1930 FOR 37 HOURS

DAY.HOUR	GASIFIER	REGEN.	RECYCLE	CYCLONES	OIL	STONE
18.0930	900.	1069.	0.	150.	129.5	35.8
18.1030	952.	1048.	0.	155.	129.5	0.
18.1130	947.	1050.	0.	185.	129.9	0.
18.1230	953.	1061.	0.	190.	129.1	0.
18.1330	952.	1067.	0.	182.	130.3	0.
18.1430	958.	1070.	0.	178.	130.3	0.
18.1530	957.	1070.	0.	180.	131.2	0.
18.1630	960.	1072.	0.	218.	129.9	0.
18.1730	958.	1068.	0.	313.	130.3	0.
18.1830	962.	1071.	0.	310.	129.9	0.

SHUT DOWN AT 18.1830 FOR 53 HOURS
CHANGE TO TJ MEDIUM VACUUM RESIDUUM FUEL

DAY.HOUR	GASIFIER	REGEN.	RECYCLE	CYCLONES	OIL	STONE
20.2330	952.	1050.	70.	163.	128.0	0.
21.0030	953.	1062.	70.	145.	129.3	0.
21.0130	956.	1074.	70.	158.	129.3	0.
21.0230	962.	1080.	70.	163.	129.3	0.
21.0330	975.	1082.	70.	148.	129.6	0.
21.0430	967.	1084.	70.	245.	128.5	0.
21.0530	976.	1086.	70.	300.	125.1	0.
21.0630	985.	1088.	70.	275.	122.7	0.
21.0730	980.	1095.	70.	305.	120.6	0.
21.0830	956.	1092.	70.	315.	118.0	0.
21.0930	947.	1087.	70.	320.	129.7	0.

SHUT DOWN AT 21.0930 FOR 6 HOURS

DAY.HOUR	GASIFIER	REGEN.	RECYCLE	CYCLONES	OIL	STONE
21.1530	968.	1100.	70.	268.	131.1	0.
21.1630	957.	1100.	70.	345.	128.7	0.
21.1730	959.	1100.	70.	350.	128.8	0.
21.1830	945.	1100.	70.	350.	128.7	0.
21.1930	941.	1100.	70.	300.	130.4	0.
21.2030	940.	1100.	70.	300.	130.4	0.

APPENDIX C: TABLE 2.

RUN 10: GAS FLOW RATES PAGE 1 OF 6

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
1.0130	314.	0.	4.5	31.6	2.6	1.04
1.0230	308.	0.	4.5	31.6	2.5	1.05
1.0330	332.	0.	4.5	30.6	2.4	1.01
1.0430	356.	0.	4.5	30.6	2.4	1.03
1.0530	359.	0.	4.5	37.5	2.4	1.25
1.0630	351.	0.	4.5	37.5	2.3	1.24
1.0730	334.	0.	4.5	37.8	2.3	1.26
1.0830	326.	0.	4.5	38.1	2.3	1.27
1.0930	343.	0.	4.5	41.0	2.4	1.36
1.1030	355.	0.	4.5	33.6	2.2	1.12
1.1130	356.	0.	4.5	36.9	2.0	1.22
1.1230	355.	0.	4.5	33.8	2.3	1.14
1.1330	345.	0.	4.5	27.2	2.2	0.93
1.1430	338.	0.	4.5	27.1	2.1	0.92
1.1530	335.	0.	4.5	26.8	2.3	0.91
1.1630	353.	0.	4.5	27.1	2.2	0.92
1.1730	346.	0.	4.5	25.8	2.4	0.89
1.1830	346.	0.	4.5	26.7	2.1	0.91
1.1930	346.	0.	4.5	26.4	2.3	0.90
1.2030	335.	0.	4.5	27.1	2.1	0.92
1.2130	317.	0.	4.5	26.7	2.2	0.91
1.2230	309.	0.	4.5	26.7	2.2	0.91
1.2330	311.	0.	4.5	26.7	2.2	0.91
2.0030	318.	0.	4.5	26.2	2.2	0.89
2.0130	308.	0.	4.5	25.9	2.2	0.88
2.0230	316.	0.	4.5	25.9	2.2	0.88
2.0330	317.	0.	4.5	26.4	2.2	0.90
2.0430	353.	0.	4.5	25.7	2.1	0.88
2.0530	327.	0.	4.5	26.3	2.2	0.90
2.0630	337.	0.	4.5	26.8	2.2	0.91
2.0730	347.	0.	4.5	26.8	2.2	0.91
2.0830	303.	0.	4.5	27.2	0.3	0.87
2.0930	300.	0.	4.5	28.7	2.3	0.98
2.1030	301.	0.	4.5	28.3	2.1	0.96
SHUT DOWN AT 2.1030 FOR 6 HOURS						
2.1630	283.	0.	4.5	29.0	2.2	0.98
2.1730	345.	0.	4.5	29.0	2.2	0.98

APPENDIX C: TABLE 2.
RUN 10: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
2.1830	346.	0.	4.5	29.3	1.8	0.98
2.1930	346.	0.	4.5	35.5	2.2	1.20
2.2030	346.	0.	4.5	39.9	2.2	1.33
2.2130	354.	0.	4.5	38.2	2.2	1.29
2.2230	354.	0.	4.5	38.2	2.2	1.28
2.2330	354.	0.	4.5	38.6	2.2	1.30
SHUT DOWN AT 2.2330 FOR 54 HOURS						
5.0630	234.	0.	4.5	26.2	3.1	0.89
5.0730	260.	0.	4.5	26.2	3.1	0.90
5.0830	376.	0.	4.5	27.1	2.5	0.93
5.0930	287.	0.	4.5	36.8	3.2	1.27
5.1030	286.	0.	4.5	36.4	3.2	1.25
5.1130	286.	0.	4.5	36.4	3.2	1.25
5.1230	286.	0.	4.5	25.0	3.2	0.89
5.1330	268.	0.	4.5	28.0	3.2	0.98
5.1430	268.	0.	4.5	37.2	3.2	1.27
5.1530	268.	0.	4.5	28.1	3.2	0.98
5.1630	285.	0.	4.5	38.7	3.2	1.31
5.1730	284.	0.	4.5	24.6	3.2	0.87
5.1830	266.	0.	4.5	28.3	3.2	0.99
5.1930	266.	0.	4.5	24.2	2.7	0.85
5.2030	284.	0.	4.5	25.2	2.8	0.89
5.2130	284.	0.	4.5	23.7	2.8	0.84
5.2230	291.	0.	4.5	31.2	2.6	1.08
5.2330	291.	0.	4.5	24.6	2.4	0.86
6.0030	291.	0.	4.5	30.2	2.8	1.05
6.0130	291.	0.	4.5	24.6	2.4	0.86
6.0230	291.	0.	4.5	31.1	3.0	1.09
6.0330	291.	0.	4.5	26.0	2.6	0.91
6.0430	239.	64.	4.5	31.3	2.8	1.09
6.0530	257.	102.	4.5	29.2	2.7	1.03
6.0630	257.	105.	4.5	29.1	3.1	1.04
6.0730	277.	104.	4.5	19.1	1.8	0.67
6.0830	296.	85.	4.5	28.7	3.6	1.03
6.0930	266.	75.	4.5	19.5	1.9	0.68
6.1030	265.	105.	4.5	22.3	2.6	0.79
6.1130	266.	86.	4.5	21.9	2.7	0.78

APPENDIX C: TABLE 2.
RUN 10: GAS FLOW RATES   PAGE 3 OF 6

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
6.1230	274.	56.	4.5	22.5	2.7	0.80
6.1330	282.	67.	4.5	21.9	2.7	0.78
6.1430	265.	86.	4.5	22.7	2.8	0.81
SHUT DOWN AT 6.1430 FOR 134 HOURS						
12.0530	294.	0.	4.6	37.0	4.0	1.23
12.0630	293.	0.	4.5	37.8	3.9	1.28
12.0730	393.	0.	4.5	35.8	4.5	1.25
12.0830	294.	0.	4.5	37.6	4.1	1.31
12.0930	288.	0.	4.5	38.0	3.7	1.32
12.1030	313.	0.	4.5	37.5	3.8	1.31
12.1130	331.	0.	4.5	35.4	3.7	1.25
12.1230	332.	0.	4.5	35.1	4.0	1.26
12.1330	332.	0.	4.5	35.1	4.0	1.25
SHUT DOWN AT 12.1330 FOR 5 HOURS						
12.1830	298.	0.	4.5	34.5	4.0	1.22
12.1930	346.	0.	4.5	35.1	3.5	1.22
12.2030	345.	0.	4.5	33.4	3.4	1.15
12.2130	335.	0.	4.5	28.6	2.4	0.96
12.2230	335.	0.	4.5	28.6	4.6	1.03
12.2330	352.	0.	4.5	28.3	2.4	0.95
13.0030	352.	0.	4.5	28.7	2.4	0.97
13.0130	353.	0.	4.5	28.6	2.3	0.97
13.0230	338.	0.	4.5	28.9	2.4	0.96
13.0330	339.	0.	4.5	28.5	2.5	0.96
13.0430	372.	0.	4.5	28.0	2.3	0.93
13.0530	405.	0.	4.5	26.7	2.2	0.90
13.0630	424.	0.	4.5	28.5	3.0	0.99
13.0730	406.	0.	4.5	28.5	3.0	0.98
13.0830	405.	140.	4.5	29.8	2.9	1.02
13.0930	318.	0.	4.5	35.5	3.0	1.20
13.1030	319.	0.	4.5	35.5	2.6	1.22
13.1130	318.	0.	4.5	35.5	2.9	1.23
13.1230	318.	0.	4.5	35.5	2.7	1.20
13.1330	318.	0.	4.5	35.5	2.7	1.20

APPENDIX C: TABLE 2.
RUN 10: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
13.1430	300.	0.	4.5	35.5	2.6	1.20
13.1530	299.	0.	4.6	35.5	1.9	1.17
13.1630	299.	0.	4.6	34.0	2.6	1.15
13.1730	325.	0.	4.6	33.5	2.3	1.13
13.1830	336.	0.	4.6	34.0	2.5	1.15
13.1930	331.	0.	4.6	34.0	2.3	1.14
13.2030	334.	0.	4.6	32.8	2.5	1.10
13.2130	334.	0.	4.5	30.8	2.2	1.02
13.2230	317.	0.	4.5	34.1	2.5	1.14
13.2330	334.	0.	4.5	36.1	2.3	1.21
14.0030	334.	0.	4.5	37.6	2.2	1.27
14.0130	333.	0.	4.5	36.8	2.2	1.28
14.0230	334.	0.	4.5	37.4	2.1	1.27
14.0330	333.	0.	4.6	37.8	2.3	1.27
14.0430	332.	0.	4.6	39.4	2.5	1.36
14.0530	334.	52.	4.6	39.9	2.2	1.38
14.0630	335.	54.	4.6	40.8	2.1	1.40
14.0730	318.	49.	4.6	41.6	2.1	1.42
14.0830	317.	50.	4.6	42.6	2.1	1.45
14.0930	318.	44.	4.6	43.5	2.1	1.48
14.1030	318.	45.	4.6	43.0	2.3	1.44
14.1130	318.	43.	4.6	44.1	2.7	1.49
14.1230	308.	46.	4.6	40.2	2.7	1.37
14.1330	299.	131.	4.6	47.2	2.2	1.58
14.1430	299.	131.	4.6	39.0	2.0	1.30
14.1530	342.	131.	4.6	34.5	2.4	1.15
14.1630	342.	112.	4.6	32.8	2.5	1.11
14.1730	343.	112.	4.6	32.8	2.6	1.12
14.1830	331.	102.	4.6	32.5	2.5	1.09
14.1930	332.	112.	4.6	33.7	2.8	1.13
14.2030	333.	112.	4.6	33.9	2.7	1.13
14.2130	333.	122.	4.6	33.6	2.8	1.13
14.2230	332.	122.	4.6	33.6	2.7	1.13
14.2330	333.	122.	4.6	31.6	2.6	1.09
15.0030	329.	122.	4.6	28.0	2.1	0.96
15.0130	334.	122.	4.6	28.0	2.2	0.96
15.0230	334.	122.	4.6	28.1	2.1	0.96
15.0330	340.	74.	4.6	28.2	2.1	0.97
15.0430	332.	64.	4.6	28.3	2.1	0.97
15.0530	332.	112.	4.6	28.0	2.1	0.96

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	FLUE GAS	PILOT PROPANE	REGENERATOR AIR	NITROGEN	REGEN. VELOCITY M/SEC
15.0630	339.	68.	4.6	28.5	2.1	0.97
15.0730	304.	121.	4.6	28.3	2.1	0.95
15.0830	298.	121.	4.6	20.0	2.1	0.69
15.0930	330.	112.	4.6	25.4	2.2	0.87
15.1030	346.	77.	4.6	33.6	1.9	1.12
15.1130	329.	116.	4.6	33.4	2.3	1.12
15.1230	330.	122.	4.6	33.9	2.2	1.13
15.1330	326.	121.	4.6	34.5	2.2	1.15
15.1430	325.	112.	4.6	32.6	2.0	1.09
15.1530	325.	122.	4.6	28.4	1.9	0.96
15.1630	325.	122.	4.6	27.7	1.9	0.95
15.1730	325.	112.	4.6	28.2	1.8	0.96
15.1830	326.	112.	4.6	26.5	1.8	0.91
15.1930	326.	112.	4.6	29.4	2.2	1.01
15.2030	327.	112.	4.6	24.8	1.8	0.85
15.2130	324.	103.	4.5	24.8	2.2	0.86
15.2230	325.	122.	4.5	27.7	2.1	0.96
15.2330	342.	112.	4.5	29.9	2.4	1.04
16.0030	322.	103.	4.5	27.7	2.2	0.97
16.0130	340.	0.	4.5	29.3	2.3	1.03
16.0230	339.	0.	4.5	26.2	2.1	0.93
16.0330	340.	0.	4.5	27.4	2.0	0.94
16.0430	305.	0.	4.5	27.4	2.1	0.92
16.0530	288.	0.	4.5	27.4	2.1	0.92
16.0630	321.	0.	4.5	27.4	2.4	0.93
16.0730	304.	0.	4.5	28.0	2.1	0.95
16.0830	288.	0.	4.5	27.6	2.1	0.94
16.0930	272.	0.	4.5	27.8	2.6	0.96
16.1030	291.	0.	4.5	28.0	2.3	0.96
16.1130	290.	0.	4.5	26.2	2.1	0.91
16.1230	289.	0.	4.5	26.9	2.0	0.94
16.1330	290.	0.	4.5	34.5	1.9	1.17
16.1430	288.	0.	4.5	40.1	2.7	1.37
16.1530	298.	0.	4.5	41.7	2.7	1.42
16.1630	434.	0.	4.6	41.2	2.2	1.38
16.1730	298.	0.	4.6	39.4	2.6	1.33
16.1830	314.	0.	4.6	42.0	2.5	1.40

APPENDIX C: TABLE 2.
RUN 10: GAS FLOW RATES

DAY.HOUR	GAS RATES M3/HR GASIFIER AIR	GASIFIER FLUE GAS	PILOT PROPANE	REGENERATOR AIR	REGENERATOR NITROGEN	REGEN. VELOCITY M/SEC
SHUT DOWN AT 16.1930 FOR 37 HOURS						
18.0930	311.	0.	4.5	29.7	1.5	0.99
18.1030	303.	0.	4.5	30.0	1.5	0.98
18.1130	308.	0.	4.5	29.5	1.8	0.98
18.1230	319.	0.	4.5	29.8	1.9	0.99
18.1330	319.	0.	4.5	29.6	1.9	0.99
18.1430	320.	0.	4.5	29.6	1.8	0.99
18.1530	317.	0.	4.5	29.9	1.8	1.00
18.1630	318.	0.	4.5	29.5	1.9	0.99
18.1730	320.	0.	4.5	29.6	1.9	0.99
18.1830	319.	0.	4.5	30.0	1.8	1.01
SHUT DOWN AT 18.1830 FOR 53 HOURS CHANGE TO TJ MEDIUM VACUUM RESIDUUM FUEL						
20.2330	328.	53.	4.6	16.1	1.4	0.55
21.0030	340.	61.	4.6	15.6	1.4	0.53
21.0130	346.	58.	4.6	15.3	0.9	0.51
21.0230	343.	58.	4.6	32.0	0.9	1.04
21.0330	345.	61.	4.6	31.5	1.2	1.04
21.0430	344.	67.	4.6	31.7	1.1	1.04
21.0530	343.	68.	4.6	31.8	1.0	1.05
21.0630	340.	66.	4.6	32.0	0.8	1.05
21.0730	347.	66.	4.6	32.0	0.7	1.05
21.0830	346.	63.	4.6	32.3	0.7	1.06
21.0930	344.	67.	4.5	31.8	0.8	1.04
SHUT DOWN AT 21.0930 FOR 6 HOURS						
21.1530	334.	79.	4.6	25.7	0.8	0.86
21.1630	332.	81.	4.6	25.7	0.8	0.86
21.1730	315.	121.	4.6	25.7	0.8	0.86
21.1830	294.	121.	4.6	25.7	0.8	0.86
21.1930	315.	121.	4.6	25.7	0.8	0.86
21.2030	315.	121.	4.6	27.2	0.8	0.91

APPENDIX C: TABLE 3.
RUN 10: PRESSURES

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
1.0130	2.8	4.7	10.2	1.05	13.9
1.0230	2.8	4.7	10.2	1.05	13.9
1.0330	2.9	4.7	10.2	1.05	13.9
1.0430	3.0	4.7	10.5	1.05	14.2
1.0530	3.0	4.7	10.5	1.05	14.2
1.0630	3.0	4.7	10.9	1.05	14.4
1.0730	2.9	4.7	11.7	1.05	15.2
1.0830	2.9	4.7	12.4	1.05	15.9
1.0930	2.9	4.5	12.7	1.05	15.9
1.1030	2.9	4.5	12.4	1.05	15.9
1.1130	2.9	4.7	12.2	1.05	15.9
1.1230	2.0	5.0	12.2	1.05	15.4
1.1330	1.9	4.6	12.2	1.05	15.4
1.1430	1.9	4.5	12.3	1.05	12.9
1.1530	1.9	4.7	12.4	1.05	12.9
1.1630	1.9	4.7	12.7	1.05	13.4
1.1730	1.9	4.5	12.7	1.05	13.4
1.1830	1.9	4.6	12.9	1.05	14.9
1.1930	1.9	4.7	12.9	1.05	14.9
1.2030	1.9	4.5	12.2	1.05	14.2
1.2130	1.8	4.0	12.7	1.05	14.4
1.2230	1.8	4.5	12.7	1.05	14.2
1.2330	1.8	4.5	12.7	1.05	14.4
2.0030	1.8	4.5	12.9	1.05	14.4
2.0130	1.8	4.5	13.2	1.05	14.7
2.0230	1.8	4.5	13.1	1.05	14.4
2.0330	1.8	4.5	12.9	1.05	14.4
2.0430	1.9	4.7	12.7	1.05	14.4
2.0530	2.0	4.7	12.2	1.05	14.4
2.0630	2.1	4.7	11.7	1.05	13.9
2.0730	2.1	4.7	11.3	1.05	13.7
2.0830	2.0	4.5	11.1	1.05	13.4
2.0930	1.9	4.5	11.1	1.05	12.9
2.1030	1.9	4.5	10.8	1.05	12.9

SHUT DOWN AT 2.1030 FOR 6 HOURS

2.1630	2.2	4.2	11.7	1.05	13.2
2.1730	2.5	5.0	11.7	1.05	13.2

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
2.1830	2.7	5.2	11.7	1.05	13.2
2.1930	2.8	5.2	11.4	1.05	12.9
2.2030	2.9	5.0	11.4	1.05	13.4
2.2130	2.9	5.0	11.4	1.05	13.4
2.2230	2.9	5.0	11.4	1.05	13.4
2.2330	2.9	4.7	11.3	1.05	13.4
SHUT DOWN AT 2.2330 FOR 54 HOURS					
5.0630	2.6	2.0	9.6	1.05	10.9
5.0730	2.6	2.2	10.0	1.05	10.9
5.0830	2.6	2.2	10.1	1.05	11.4
5.0930	2.6	2.5	10.1	1.05	11.2
5.1030	2.6	2.7	10.3	1.05	11.4
5.1130	2.6	2.5	10.2	1.05	11.9
5.1230	2.6	2.5	10.2	1.05	11.9
5.1330	2.6	2.6	10.1	1.05	12.2
5.1430	2.5	2.5	10.1	1.05	12.4
5.1530	2.6	2.7	10.2	1.05	10.9
5.1630	2.6	2.9	10.6	1.05	11.2
5.1730	2.7	2.9	10.7	1.05	11.4
5.1830	2.5	2.7	10.7	1.05	11.4
5.1930	2.2	2.7	10.8	1.05	10.9
5.2030	2.3	3.0	10.9	1.05	10.9
5.2130	2.3	2.7	11.1	1.05	11.2
5.2230	2.3	2.7	11.2	1.05	11.4
5.2330	2.5	3.0	11.2	1.05	11.4
6.0030	2.8	3.0	11.2	1.05	11.7
6.0130	2.9	3.0	11.2	1.05	11.7
6.0230	2.9	3.0	11.2	1.05	11.7
6.0330	2.9	3.0	11.2	1.05	11.7
6.0430	2.6	3.1	11.1	1.05	11.4
6.0530	1.9	3.2	11.1	1.05	11.2
6.0630	2.1	3.7	10.9	1.05	10.9
6.0730	2.1	4.0	10.7	1.05	10.9
6.0830	2.2	4.0	10.6	1.05	11.4
6.0930	2.5	4.0	10.5	1.05	11.4
6.1030	2.7	4.0	10.2	1.05	11.7
6.1130	2.7	3.7	10.2	1.05	11.7

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
6.1230	2.7	3.7	10.2	1.05	11.7
6.1330	2.8	3.7	10.1	1.05	11.7
6.1430	2.8	3.7	9.8	1.05	11.4
SHUT DOWN AT 6.1430 FOR 134 HOURS					
12.0530	2.5	4.0	10.2	1.05	12.4
12.0630	2.9	4.1	10.1	1.05	12.3
12.0730	2.8	4.2	10.1	1.05	12.3
12.0830	2.4	4.0	10.7	1.05	12.9
12.0930	2.3	3.7	11.7	1.05	13.2
12.1030	2.8	4.2	11.6	1.05	13.6
12.1130	2.9	4.2	11.6	1.05	13.8
12.1230	2.5	4.7	11.7	1.05	13.7
12.1330	2.7	4.5	11.7	1.05	13.7
SHUT DOWN AT 12.1330 FOR 5 HOURS					
12.1830	1.9	3.7	12.1	1.05	14.2
12.1930	1.9	4.2	12.2	1.05	13.9
12.2030	2.2	4.5	12.2	1.05	13.9
12.2130	2.3	4.2	12.4	1.05	14.2
12.2230	2.3	4.2	12.7	1.05	14.4
12.2330	2.3	4.2	12.8	1.05	14.4
13.0030	2.3	4.7	12.7	1.05	14.4
13.0130	2.4	4.7	12.3	1.05	14.2
13.0230	2.4	4.7	12.4	1.05	14.2
13.0330	2.4	4.0	13.3	1.05	14.9
13.0430	2.3	4.0	13.6	1.05	15.2
13.0530	2.3	4.0	14.1	1.05	15.4
13.0630	2.8	5.0	13.1	1.05	14.9
13.0730	2.9	6.0	12.4	1.05	14.4
13.0830	2.8	6.0	12.1	1.05	13.9
13.0930	2.5	6.0	11.3	1.05	12.9
13.1030	2.5	5.5	11.2	1.05	12.1
13.1130	2.5	5.2	10.9	1.05	11.7
13.1230	2.6	5.0	10.8	1.05	11.9
13.1330	2.6	4.7	10.9	1.05	12.3

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
13.1430	2.6	4.2	11.1	1.05	12.4
13.1530	2.4	4.0	11.3	1.05	12.3
13.1630	2.5	4.7	11.4	1.05	12.9
13.1730	2.6	4.5	11.4	1.05	12.9
13.1830	2.5	4.2	11.4	1.05	12.9
13.1930	2.4	4.2	11.4	1.05	13.4
13.2030	2.4	4.4	12.6	1.05	13.9
13.2130	2.3	4.2	12.9	1.05	14.3
13.2230	2.3	4.2	13.2	1.05	14.2
13.2330	2.3	4.0	13.4	1.05	14.3
14.0030	2.2	4.2	13.4	1.05	14.3
14.0130	2.1	4.2	13.7	1.05	14.3
14.0230	2.1	4.2	13.7	1.05	14.1
14.0330	2.1	4.2	13.6	1.05	13.9
14.0430	2.4	4.2	13.6	1.05	13.9
14.0530	2.4	4.5	13.3	1.05	13.9
14.0630	2.4	4.7	13.2	1.05	13.8
14.0730	2.4	4.7	13.1	1.05	13.6
14.0830	2.2	4.5	12.9	1.05	13.3
14.0930	2.2	4.5	12.9	1.05	13.4
14.1030	2.3	4.5	12.9	1.05	13.6
14.1130	2.3	4.7	12.8	1.05	13.4
14.1230	2.5	5.2	12.7	1.05	13.4
14.1330	2.5	5.5	12.7	1.05	13.4
14.1430	2.5	5.7	12.7	1.05	13.4
14.1530	2.4	6.0	12.7	1.05	14.1
14.1630	2.4	6.0	12.6	1.05	13.9
14.1730	2.2	2.0	12.6	1.05	14.2
14.1830	2.5	6.0	12.6	1.05	13.9
14.1930	2.5	5.7	12.2	1.05	13.6
14.2030	2.5	5.7	11.9	1.05	13.3
14.2130	2.5	5.7	11.9	1.05	13.3
14.2230	2.5	5.5	11.9	1.05	13.3
14.2330	2.4	5.5	11.7	1.05	13.3
15.0030	2.4	5.5	11.7	1.05	13.3
15.0130	2.6	5.7	11.4	1.05	13.1
15.0230	2.6	6.0	11.4	1.05	13.1
15.0330	2.6	6.0	11.4	1.05	13.1
15.0430	2.5	6.0	10.9	1.05	12.4
15.0530	2.4	6.0	11.4	1.05	12.9

APPENDIX C: TABLE 3.
RUN 10: PRESSURES PAGE 5 OF 6

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
15.0630	2.5	6.0	10.9	1.05	12.4
15.0730	2.4	6.0	11.2	1.05	12.7
15.0830	2.4	6.0	11.4	1.05	13.4
15.0930	2.4	6.2	10.9	1.05	12.3
15.1030	2.4	5.5	10.9	1.05	12.3
15.1130	2.3	5.0	10.9	1.05	12.3
15.1230	2.4	5.0	10.5	1.05	12.1
15.1330	2.4	5.0	10.5	1.05	12.1
15.1430	2.4	5.0	10.5	1.05	12.2
15.1530	2.4	5.0	10.5	1.05	12.2
15.1630	2.4	5.0	10.5	1.05	12.3
15.1730	2.4	5.0	10.5	1.05	12.3
15.1830	2.4	5.0	10.5	1.05	12.3
15.1930	2.4	5.0	10.5	1.05	12.3
15.2030	2.5	5.5	10.5	1.05	12.3
15.2130	2.5	5.5	10.3	1.05	12.2
15.2230	2.6	5.5	10.1	1.05	11.8
15.2330	2.4	5.0	10.1	1.05	11.8
16.0030	2.2	5.5	10.3	1.05	11.9
16.0130	2.2	5.0	10.1	1.05	11.7
16.0230	2.2	5.0	10.1	1.05	11.7
16.0330	2.2	5.0	10.1	1.05	11.7
16.0430	2.1	5.5	10.1	1.05	11.6
16.0530	2.0	6.0	10.1	1.05	11.4
16.0630	2.0	5.5	9.8	1.05	11.2
16.0730	2.0	5.5	9.8	1.05	11.2
16.0830	1.9	5.5	9.8	1.05	11.1
16.0930	1.9	5.0	10.0	1.05	11.2
16.1030	2.0	4.5	9.8	1.05	11.2
16.1130	2.0	5.0	9.6	1.05	11.1
16.1230	2.1	5.0	9.6	1.05	10.9
16.1330	2.1	5.5	9.6	1.05	10.9
16.1430	1.8	5.0	9.6	1.05	10.8
16.1530	2.1	5.5	9.6	1.05	10.7
16.1630	2.2	6.0	9.5	1.05	10.6
16.1730	2.7	5.5	9.5	1.05	10.6
16.1830	2.8	5.5	9.5	1.05	10.6

DAY.HOUR	GASIFIER P. KILOPASCALS GAS SPACE	DISTRIB. D.P.	BED D.P.	GASIFIER BED SP. GR.	REGEN. BED D.P.
SHUT DOWN AT 16.1930 FOR 37 HOURS					
18.0930	2.4	5.0	9.6	1.05	11.1
18.1030	2.4	5.0	10.1	1.05	11.3
18.1130	2.9	5.0	9.7	1.05	11.4
18.1230	3.0	5.0	9.7	1.05	11.4
18.1330	3.0	5.0	9.6	1.05	11.2
18.1430	3.0	5.0	9.6	1.05	11.3
18.1530	3.1	5.0	9.5	1.05	11.4
18.1630	3.0	5.0	9.5	1.05	11.4
18.1730	3.0	5.0	9.5	1.05	11.2
18.1830	3.1	5.0	9.5	1.05	11.2
SHUT DOWN AT 18.1830 FOR 53 HOURS CHANGE TO TJ MEDIUM VACUUM RESIDUUM FUEL					
20.2330	2.9	8.0	10.2	1.05	10.0
21.0030	3.0	8.0	9.7	1.05	10.0
21.0130	3.2	8.0	9.5	1.05	10.0
21.0230	3.2	8.0	9.5	1.05	10.0
21.0330	3.3	8.2	9.2	1.05	10.0
21.0430	3.3	8.2	9.0	1.05	10.0
21.0530	3.2	8.2	9.3	1.05	10.0
21.0630	3.2	8.2	9.3	1.05	10.0
21.0730	3.2	8.2	9.2	1.05	10.0
21.0830	3.2	8.2	9.3	1.05	10.0
21.0930	3.2	8.2	9.1	1.05	10.0
SHUT DOWN AT 21.0930 FOR 6 HOURS					
21.1530	2.8	8.5	9.0	1.05	10.0
21.1630	2.2	7.5	9.2	1.05	10.0
21.1730	2.6	7.2	9.3	1.05	10.0
21.1830	2.2	7.2	9.1	1.05	10.0
21.1930	2.8	7.2	9.1	1.05	10.0
21.2030	2.9	7.2	9.1	1.05	10.0

APPENDIX C: TABLE 4.

RUN 10: DESULPHURISATION PERFORMANCE PAGE 1 OF 6

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
1.0130	76.6	0.74	99.	22.0	0.	21.7	29.0
1.0230	78.9	0.73	99.	21.6	0.44	21.6	28.9
1.0330	83.2	0.77	99.	23.3	2.22	21.1	27.6
1.0430	80.6	0.85	101.	25.0	0.98	20.9	27.7
1.0530	78.4	0.84	101.	25.1	5.22	16.8	24.8
1.0630	82.5	0.80	106.	24.6	6.58	16.4	24.6
1.0730	85.5	0.74	113.	23.4	3.42	17.2	25.2
1.0830	85.2	0.73	120.	22.8	4.36	24.1	34.6
1.0930	80.1	0.79	123.	24.0	0.	45.5	75.2
1.1030	81.6	0.82	120.	24.9	0.	54.5	76.2
1.1130	80.6	0.84	118.	24.9	0.	51.4	81.9
1.1230	81.9	0.84	118.	24.4	0.65	56.7	84.4
1.1330	82.5	0.81	118.	23.7	1.31	55.2	61.6
1.1430	82.5	0.78	119.	23.2	2.63	52.1	56.7
1.1530	84.8	0.77	120.	23.0	2.57	60.9	61.5
1.1630	83.6	0.81	123.	24.4	2.28	58.0	62.0
1.1730	83.9	0.80	123.	24.0	2.24	61.0	63.8
1.1830	83.6	0.80	125.	24.0	1.89	62.2	67.1
1.1930	82.8	0.81	125.	23.9	1.80	66.6	70.9
1.2030	82.3	0.78	118.	23.2	1.93	71.8	77.2
1.2130	79.7	0.74	123.	21.9	0.	90.9	74.1
1.2230	80.2	0.73	123.	21.4	0.57	125.2	44.1
1.2330	81.5	0.72	123.	21.5	2.42	117.7	34.2
2.0030	82.5	0.73	125.	22.0	3.21	104.3	54.1
2.0130	83.1	0.72	128.	21.3	0.	103.0	69.4
2.0230	78.2	0.74	127.	21.9	0.79	109.6	52.1
2.0330	78.3	0.75	125.	21.9	0.	110.7	43.5
2.0430	81.6	0.85	123.	24.5	0.33	151.2	24.9
2.0530	81.2	0.78	118.	22.0	0.79	81.6	81.9
2.0630	78.6	0.81	113.	22.6	0.55	73.8	81.8
2.0730	75.9	0.84	110.	23.3	0.34	67.8	80.3
2.0830	73.0	0.73	107.	20.4	0.21	61.9	70.3
2.0930	74.4	0.72	107.	20.2	0.	68.7	77.2
2.1030	72.6	0.72	105.	20.2	0.	67.3	75.5
SHUT DOWN AT 2.1030 FOR 6 HOURS							
2.1630	67.6	0.67	113.	18.4	0.29	14.4	13.5
2.1730	65.7	0.82	113.	22.4	0.04	20.6	14.1

APPENDIX C: TABLE 4.
RUN 10: 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
2.1830	70.2	0.83	113.	22.5	0.	44.9	15.2
2.1930	72.6	0.83	111.	21.0	0.50	-	0.
2.2030	72.5	0.82	111.	21.0	0.35	-	0.
2.2130	73.0	0.84	111.	21.5	1.08	-	0.
2.2230	74.2	0.84	111.	21.5	0.58	-	0.
2.2330	73.8	0.85	110.	21.5	0.	-	0.
SHUT DOWN AT 2.2330 FOR 54 HOURS							
5.0630	73.2	0.54	93.	19.6	0.87	17.5	16.6
5.0730	74.1	0.61	96.	21.7	0.72	30.0	36.4
5.0830	70.8	0.88	97.	31.4	0.48	25.7	30.8
5.0930	68.7	0.67	97.	22.5	1.22	16.6	21.1
5.1030	68.8	0.67	100.	22.0	0.75	26.4	37.9
5.1130	66.4	0.66	99.	21.9	0.54	50.3	80.9
5.1230	66.3	0.67	99.	22.0	0.39	59.6	63.0
5.1330	65.4	0.62	97.	20.6	0.19	58.9	69.7
5.1430	67.4	0.61	97.	18.9	0.58	61.7	74.0
5.1530	68.6	0.60	99.	18.9	1.97	66.7	45.9
5.1630	74.4	0.65	102.	20.1	1.66	63.9	86.9
5.1730	74.3	0.65	104.	20.5	1.19	63.3	60.9
5.1830	75.3	0.60	104.	19.3	1.06	62.5	66.8
5.1930	80.5	0.61	105.	19.2	1.19	94.9	62.6
5.2030	82.0	0.65	106.	20.5	1.24	83.0	71.9
5.2130	80.3	0.65	107.	20.5	1.28	106.5	60.6
5.2230	77.3	0.67	108.	21.0	0.69	110.0	85.5
5.2330	75.3	0.67	108.	21.0	0.67	107.1	70.1
6.0030	78.0	0.67	108.	21.0	0.87	104.4	89.7
6.0130	79.1	0.67	108.	21.0	0.80	100.6	71.3
6.0230	78.7	0.68	108.	21.0	0.18	92.2	91.4
6.0330	72.3	0.69	108.	21.1	0.11	85.7	79.8
6.0430	73.3	0.73	107.	20.6	0.13	67.9	100.8
6.0530	71.1	0.87	107.	26.6	0.41	65.1	111.6
6.0630	70.8	0.86	106.	26.9	0.58	51.9	83.4
6.0730	67.4	0.90	104.	28.6	0.64	53.2	57.7
6.0830	66.2	0.90	102.	29.8	0.41	56.4	95.5
6.0930	70.8	0.81	101.	26.7	0.61	54.1	67.6
6.1030	71.4	0.88	99.	27.1	2.44	55.2	71.5
6.1130	73.3	0.83	99.	26.6	2.21	52.1	65.9

APPENDIX C: TABLE 4.

RUN 10: DESULPHURISATION PERFORMANCE PAGE 3 OF 6

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	72.9	0.78	99.	26.2	0.80	52.0	67.8
6.1330	72.3	0.83	97.	27.3	0.60	51.4	65.9
6.1430	68.0	0.84	95.	24.6	0.46	54.0	66.8
SHUT DOWN AT 6.1430 FOR 134 HOURS							
12.0530	70.8	0.67	99.	19.3	0.	-	0.
12.0630	72.0	0.68	97.	21.6	0.	-	0.
12.0730	71.9	0.90	97.	26.9	2.29	-	0.
12.0830	75.9	0.66	104.	20.3	4.38	-	0.
12.0930	74.3	0.66	113.	19.0	0.	-	0.
12.1030	80.7	0.74	112.	20.6	0.17	48.9	67.9
12.1130	81.8	0.77	112.	21.8	1.27	41.8	55.9
12.1230	81.1	0.78	113.	21.8	2.15	37.0	49.0
12.1330	80.1	0.77	113.	21.9	0.90	36.9	49.2
SHUT DOWN AT 12.1330 FOR 5 HOURS							
12.1830	77.8	0.69	117.	22.9	2.54	48.0	68.1
12.1930	80.8	0.80	118.	24.1	2.14	43.3	60.8
12.2030	81.6	0.79	118.	24.2	2.16	53.1	69.1
12.2130	81.2	0.76	120.	23.5	2.49	58.1	61.3
12.2230	81.5	0.76	123.	23.2	2.15	68.5	62.3
12.2330	82.3	0.81	124.	24.6	1.76	67.4	64.0
13.0030	83.6	0.81	123.	24.8	2.45	61.7	63.8
13.0130	84.0	0.82	119.	24.7	1.69	62.0	69.3
13.0230	84.9	0.76	120.	23.4	2.35	71.1	68.7
13.0330	84.0	0.76	129.	23.4	3.23	53.9	43.4
13.0430	79.9	0.82	131.	25.6	2.89	65.2	31.9
13.0530	80.7	0.92	136.	27.9	2.69	-	0.
13.0630	80.5	1.00	127.	29.9	2.93	-	0.
13.0730	80.1	0.93	120.	28.6	0.	-	0.
13.0830	81.3	1.26	117.	30.8	1.28	-	0.
13.0930	82.0	0.74	110.	22.6	0.32	-	0.
13.1030	82.5	0.76	108.	22.7	0.	-	0.
13.1130	83.5	0.76	106.	22.6	0.	-	0.
13.1230	84.5	0.74	105.	22.7	0.	-	0.
13.1330	85.7	0.73	106.	22.6	2.46	-	0.

APPENDIX C: TABLE 4.
RUN 10: 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
13.1430	85.1	0.69	107.	21.4	2.48	-	0.
13.1530	85.1	0.69	110.	21.3	2.57	62.4	71.2
13.1630	86.8	0.69	111.	21.1	3.07	59.6	74.8
13.1730	87.1	0.74	111.	22.9	3.30	55.6	70.2
13.1830	87.1	0.77	111.	23.6	3.21	58.3	76.3
13.1930	87.3	0.76	111.	23.2	1.23	60.1	76.7
13.2030	84.5	0.76	122.	23.4	3.96	93.9	81.5
13.2130	85.4	0.76	125.	23.6	2.76	62.3	60.5
13.2230	85.1	0.72	128.	22.6	1.63	85.5	55.7
13.2330	85.7	0.76	130.	24.6	2.73	88.0	76.5
14.0030	84.1	0.77	130.	23.6	0.	70.1	79.8
14.0130	84.4	0.79	133.	23.8	0.	80.9	69.7
14.0230	84.6	0.79	133.	23.6	0.	-	0.
14.0330	84.9	0.79	131.	23.4	0.	-	0.
14.0430	85.2	0.79	131.	23.5	0.	-	0.
14.0530	84.6	0.91	129.	26.3	0.	68.7	64.4
14.0630	87.1	0.93	128.	26.8	0.	92.0	114.5
14.0730	86.2	0.87	127.	25.5	0.	72.5	119.2
14.0830	84.5	0.87	125.	25.5	0.	45.2	73.0
14.0930	83.2	0.86	125.	25.4	0.	48.5	92.6
14.1030	81.6	0.86	125.	25.4	0.	44.4	76.2
14.1130	83.0	0.85	124.	25.4	0.	46.8	91.7
14.1230	77.6	0.81	123.	22.5	0.	48.1	77.1
14.1330	75.8	0.99	123.	22.8	0.	46.4	80.3
14.1430	74.6	1.00	123.	26.8	0.	44.0	71.8
14.1530	73.9	1.11	123.	27.4	0.	50.0	69.5
14.1630	73.4	1.07	122.	27.4	0.	49.4	67.3
14.1730	73.6	1.08	122.	27.7	0.	48.7	68.1
14.1830	74.0	1.03	122.	26.6	0.	48.6	63.0
14.1930	73.3	1.05	118.	27.1	0.	49.9	68.6
14.2030	74.1	1.06	116.	27.0	0.	49.0	68.8
14.2130	74.3	1.08	116.	27.2	0.	53.9	69.4
14.2230	74.0	1.08	116.	27.2	0.	56.0	74.3
14.2330	71.7	1.08	113.	27.2	0.	46.6	65.5
15.0030	71.0	1.07	113.	27.0	0.	53.2	64.7
15.0130	70.4	1.09	111.	27.3	0.	57.9	69.5
15.0230	70.5	1.09	111.	27.4	0.	53.9	66.9
15.0330	69.8	0.99	111.	27.2	0.	55.3	69.6
15.0430	71.3	0.93	106.	26.4	0.	56.5	72.3
15.0530	71.9	1.06	111.	27.6	0.	60.0	70.6

APPENDIX C: TABLE 4.
RUN 10: DESULPHURISATION PERFORMANCE PAGE 5 OF 6

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
15.0630	71.3	0.96	106.	27.1	0.	60.4	76.4
15.0730	69.7	1.00	108.	25.9	0.	60.2	67.8
15.0830	69.8	0.98	111.	25.1	0.	76.2	60.0
15.0930	69.3	1.03	106.	27.8	0.	63.5	60.9
15.1030	68.2	0.98	106.	27.2	0.	46.4	65.1
15.1130	68.1	1.03	106.	27.3	0.	46.1	59.8
15.1230	67.6	1.05	101.	27.2	0.	54.1	78.1
15.1330	67.3	1.04	101.	26.3	0.	53.1	75.6
15.1430	65.9	1.02	101.	26.1	0.	52.5	72.9
15.1530	67.0	1.05	101.	28.3	0.	53.5	69.1
15.1630	66.3	1.05	101.	27.8	0.	56.5	66.8
15.1730	66.4	1.03	101.	25.5	0.	59.0	69.1
15.1830	65.6	1.03	101.	27.6	0.	54.4	63.5
15.1930	65.1	1.03	101.	26.9	0.	62.4	79.3
15.2030	64.2	1.04	101.	26.9	0.	59.8	64.4
15.2130	65.7	1.02	100.	25.9	0.	56.9	61.9
15.2230	64.8	1.07	97.	25.8	0.	56.1	68.6
15.2330	63.9	1.07	97.	28.0	0.	59.5	82.2
16.0030	61.9	1.01	100.	26.2	0.	46.1	56.2
16.0130	56.3	0.79	97.	26.0	0.	48.4	66.2
16.0230	56.7	0.79	97.	26.2	0.	40.2	49.1
16.0330	58.3	0.79	97.	26.1	0.	31.7	38.1
16.0430	59.4	0.69	97.	23.9	0.	59.6	71.3
16.0530	72.2	0.65	97.	21.8	0.	61.7	61.8
16.0630	68.5	0.72	95.	24.6	0.	98.6	55.7
16.0730	61.1	0.68	95.	23.2	0.	55.0	54.5
16.0830	57.2	0.64	95.	21.8	0.	50.1	52.6
16.0930	55.6	0.62	96.	20.8	0.	46.4	49.8
16.1030	62.4	0.68	95.	22.5	0.	71.8	69.6
16.1130	63.3	0.69	93.	22.6	0.	107.2	88.1
16.1230	62.7	0.69	93.	21.9	0.	116.0	76.9
16.1330	63.5	0.69	93.	21.7	0.	–	0.
16.1430	64.2	0.69	93.	21.7	0.	–	0.
16.1530	62.2	0.72	93.	22.6	0.	–	0.
16.1630	58.1	1.04	91.	34.0	0.	75.9	53.3
16.1730	62.6	0.71	91.	22.6	0.	82.9	70.8
16.1830	63.5	0.75	91.	23.7	0.	75.6	80.1

APPENDIX C: TABLE 4.

RUN 10: 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
SHUT DOWN AT 16.1930 FOR 37 HOURS							
18.0930	85.4	0.71	93.	22.5	3.62	14.6	14.8
18.1030	84.8	0.72	97.	21.9	0.	64.9	77.8
18.1130	80.7	0.73	94.	22.2	0.	77.6	83.7
18.1230	76.0	0.76	94.	23.1	0.	69.1	73.0
18.1330	71.2	0.76	93.	22.9	0.	64.6	82.9
18.1430	67.9	0.76	93.	23.0	0.	55.0	72.9
18.1530	65.9	0.76	91.	22.7	0.	57.3	77.2
18.1630	65.3	0.76	91.	23.0	0.	51.5	68.1
18.1730	63.0	0.76	91.	23.0	0.	48.6	63.6
18.1830	63.4	0.76	91.	23.0	0.	48.3	64.4
SHUT DOWN AT 18.1830 FOR 53 HOURS CHANGE TO TJ MEDIUM VACUUM RESIDUUM FUEL							
20.2330	78.7	0.92	99.	25.2	0.	46.5	7.6
21.0030	76.7	0.97	94.	26.0	0.	72.1	33.1
21.0130	71.9	0.97	91.	26.3	0.	80.1	33.7
21.0230	69.2	0.97	91.	26.1	0.	64.5	73.9
21.0330	65.7	0.99	89.	26.2	0.	66.2	83.2
21.0430	62.0	1.00	87.	26.6	0.	62.7	78.2
21.0530	59.9	1.01	90.	27.3	0.	60.6	76.5
21.0630	59.4	1.00	90.	27.6	0.	57.2	72.3
21.0730	59.5	1.01	89.	28.7	0.	55.2	72.7
21.0830	59.3	0.99	90.	29.4	0.	52.9	70.9
21.0930	63.1	0.98	88.	26.3	0.	53.8	61.1
SHUT DOWN AT 21.0930 FOR 6 HOURS							
21.1530	58.6	1.01	87.	26.0	0.	55.3	15.4
21.1630	62.0	1.01	89.	26.0	0.	22.1	10.6
21.1730	63.7	1.07	90.	25.1	0.	-3.0	0.0
21.1830	64.0	1.01	88.	23.6	0.	47.2	42.3
21.1930	60.9	1.05	88.	24.7	0.	52.6	41.7
21.2030	58.8	1.05	88.	24.9	0.	64.2	36.9

APPENDIX C: TABLE 5.

RUN 10: 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
1.0130	-	-	8.3	-	0.20	1.4	2.4	21.0	21.0	0.	0.
1.0230	-	-	8.3	-	0.20	1.4	2.4	21.0	21.0	0.	0.
1.0330	9.5	14.0	8.7	151.	0.20	1.0	2.4	21.0	21.0	0.	0.
1.0430	9.0	14.0	9.1	181.	0.20	0.9	2.4	21.0	21.0	0.	0.
1.0530	-	-	8.2	-	1.80	0.9	1.8	21.0	21.0	0.	0.
1.0630	-	-	8.2	-	1.40	0.9	1.8	21.0	21.0	0.	0.
1.0730	9.0	14.0	9.0	136.	1.00	1.9	1.8	21.0	21.0	0.	0.
1.0830	9.0	14.0	9.0	139.	0.50	3.2	2.4	21.0	21.0	0.	0.
1.0930	-	-	8.3	-	0.	3.6	4.8	21.0	21.0	0.	0.
1.1030	8.0	14.0	9.8	186.	0.	3.4	5.8	21.0	21.0	0.	0.
1.1130	9.0	14.0	9.0	181.	0.	2.7	5.8	21.0	21.0	0.	0.
1.1230	4.0	14.0	12.9	241.	0.	2.2	6.5	21.0	21.0	0.	0.
1.1330	3.5	14.0	13.2	240.	0.	3.3	5.8	21.0	21.0	0.	0.
1.1430	3.4	14.0	13.3	241.	0.	3.4	5.4	21.0	21.0	0.	0.
1.1530	3.4	14.0	13.3	210.	0.	4.7	5.8	21.0	21.0	0.	0.
1.1630	5.0	14.0	12.1	206.	0.	4.1	5.8	21.0	21.0	0.	0.
1.1730	5.4	14.0	11.8	197.	0.	3.7	6.2	21.0	21.0	0.	0.
1.1830	4.1	14.0	12.7	217.	0.	3.9	6.4	21.0	21.0	0.	0.
1.1930	4.0	14.0	12.8	229.	0.	4.1	6.7	21.0	21.0	0.	0.
1.2030	4.0	14.0	12.8	236.	0.	4.5	7.1	21.0	21.0	0.	0.
1.2130	3.8	14.0	13.0	274.	0.	8.2	6.5	21.0	21.0	0.	0.
1.2230	3.6	14.0	13.1	272.	0.	13.7	3.7	21.0	21.0	0.	0.
1.2330	3.1	14.0	13.4	262.	0.	14.3	2.9	21.0	21.0	0.	0.
2.0030	3.4	14.0	13.2	244.	0.	11.6	4.8	21.0	21.0	0.	0.
2.0130	3.7	14.0	13.1	230.	0.	9.8	6.2	21.0	21.0	0.	0.
2.0230	3.8	14.0	13.0	296.	0.	12.1	4.6	21.0	21.0	0.	0.
2.0330	5.0	14.0	12.1	274.	0.	13.2	3.7	21.0	21.0	0.	0.
2.0430	4.2	14.0	12.7	244.	0.	15.7	2.2	21.0	21.0	0.	0.

2.0530	4.5	14.0	12.5	244.	0.	5.0	7.8	21.0	21.0	0.	0.
2.0630	4.2	14.0	12.7	282.	0.	3.6	7.8	21.0	21.0	0.	0.
2.0730	3.9	14.0	12.9	323.	0.	2.2	7.8	21.0	21.0	0.	0.
2.0830	3.7	14.0	13.1	365.	0.	3.2	7.2	21.0	21.0	0.	0.
2.0930	4.1	14.0	12.8	339.	0.	4.2	6.9	21.0	21.0	0.	0.
2.1030	4.1	14.0	12.8	364.	0.	4.1	6.9	21.0	21.0	0.	0.

SHUT DOWN AT 2.1030 FOR 6 HOURS

2.1630	4.8	14.0	12.3	413.	0.	4.7	1.3	21.0	21.0	0.	0.
2.1730	5.2	14.0	12.0	428.	0.	8.6	1.3	21.0	21.0	0.	0.
2.1830	4.0	14.0	12.8	403.	0.	13.7	1.3	21.0	21.0	0.	0.
2.1930	2.8	14.0	13.9	394.	0.	0.2	0.	21.0	21.0	0.	0.
2.2030	2.6	14.0	14.0	399.	20.00	0.2	0.	21.0	21.0	0.	0.
2.2130	3.0	14.0	13.7	383.	21.00	0.	0.	21.0	21.0	0.	0.
2.2230	2.9	14.0	13.7	368.	21.00	0.	0.	21.0	21.0	0.	0.
2.2330	2.9	14.0	13.7	374.	21.00	0.	0.	21.0	21.0	0.	0.

SHUT DOWN AT 2.2330 FOR 54 HOURS

5.0630	3.5	14.0	13.1	360.	5.00	1.7	1.3	21.0	21.0	0.	0.
5.0730	5.3	14.0	11.8	312.	0.50	2.8	2.9	21.0	21.0	0.	0.
5.0830	5.2	14.0	10.8	322.	3.00	1.2	2.4	21.0	21.0	0.15	0.
5.0930	3.9	14.0	12.9	413.	6.00	0.9	1.3	21.0	21.0	0.	0.
5.1030	3.3	14.0	13.3	428.	3.50	1.2	2.4	21.0	21.0	0.	0.
5.1130	6.0	14.0	11.3	391.	1.00	2.7	5.1	21.0	21.0	0.	0.
5.1230	4.2	14.0	12.6	438.	0.80	3.7	5.4	21.0	21.0	0.	0.
5.1330	3.5	14.0	13.2	470.	0.70	3.9	5.4	21.0	21.0	0.	0.
5.1430	4.0	14.0	12.8	434.	0.80	6.3	4.8	21.0	21.0	0.	0.
5.1530	5.0	14.0	12.0	395.	1.00	8.6	3.7	21.0	21.0	0.	0.
5.1630	5.0	14.0	12.0	322.	0.50	5.6	5.4	21.0	21.0	0.	0.
5.1730	5.5	14.0	11.7	311.	0.30	4.5	5.6	21.0	21.0	0.	0.
5.1830	5.8	14.0	11.4	293.	0.70	4.7	5.4	21.0	21.0	0.	0.
5.1930	4.5	14.0	12.4	252.	0.20	9.2	5.6	21.0	21.0	0.	0.

DAY.HOUR	F L U E G A S O2 %	CO2 VOL % ANAL	CO2 VOL % CALC	SO2 PPM	REGENERATOR GAS O2 %	CO2 %	SO2 %	GASIFIER INLET GAS O2 VOL % ANAL	O2 VOL % CALC	CO2 VOL % ANAL	CO2 VOL % CALC
5.2030	4.7	14.0	12.2	229.	0.20	6.8	6.4	21.0	21.0	0.	0.
5.2130	4.7	14.0	12.2	252.	0.20	10.3	5.4	21.0	21.0	0.	0.
5.2230	5.2	14.0	11.9	282.	0.10	10.5	6.0	21.0	21.0	0.	0.
5.2330	5.0	14.0	12.0	309.	0.10	9.8	6.2	21.0	21.0	0.	0.
6.0030	5.7	14.0	11.5	264.	0.20	9.2	6.5	21.0	21.0	0.	0.
6.0130	5.0	14.0	12.0	261.	0.20	9.0	6.4	21.0	21.0	0.	0.
6.0230	5.3	14.0	11.8	261.	0.40	7.8	6.5	21.0	21.0	0.	0.
6.0330	7.0	14.0	10.5	302.	0.30	6.5	6.9	21.0	21.0	0.	0.
6.0430	6.3	14.0	11.0	302.	0.50	3.4	6.9	17.2	17.8	1.55	3.09
6.0530	7.2	14.0	10.2	302.	0.60	1.9	7.2	18.0	16.9	1.62	4.12
6.0630	7.5	14.0	10.0	299.	1.00	1.9	5.4	15.0	16.9	3.33	4.20
6.0730	7.5	14.0	10.0	333.	1.00	1.6	5.8	15.5	17.2	2.98	3.96
6.0830	7.5	14.0	10.0	345.	0.40	1.6	6.2	16.0	17.9	2.64	3.24
6.0930	7.3	14.0	10.1	302.	0.40	1.4	6.2	16.4	17.8	2.32	3.30
6.1030	7.3	14.0	10.1	297.	0.60	1.2	6.2	16.8	16.9	2.98	4.20
6.1130	7.4	14.0	10.0	275.	0.50	1.2	5.8	17.2	17.5	2.00	3.64
6.1230	7.3	14.0	10.2	281.	0.50	1.2	5.8	17.3	18.5	1.93	2.55
6.1330	7.5	14.0	10.0	283.	0.50	1.0	5.8	17.5	18.3	1.93	2.80
6.1430	6.8	14.0	10.5	346.	0.60	1.2	6.0	17.4	17.3	1.85	3.63
SHUT DOWN AT 6.1430 FOR 134 HOURS											
12.0530	6.0	14.0	11.4	342.	14.00	0.2	0.	21.0	21.0	0.	0.
12.0630	5.3	14.0	11.9	343.	10.30	0.7	0.	21.0	21.0	0.	0.
12.0730	5.5	14.0	11.0	322.	9.50	1.4	0.	21.0	21.0	0.	0.
12.0830	4.8	14.0	12.2	306.	9.00	1.9	0.	21.0	21.0	0.	0.
12.0930	5.1	14.0	12.1	322.	9.50	3.7	0.	21.0	21.0	0.	0.
12.1030	6.4	14.0	11.0	222.	0.50	3.4	4.8	21.0	21.0	0.	0.
12.1130	6.2	14.0	11.2	212.	0.80	2.5	4.2	21.0	21.0	0.	0.
12.1230	6.4	14.0	11.0	217.	1.20	1.9	3.7	21.0	21.0	0.	0.
12.1330	6.8	14.0	10.7	222.	1.00	2.1	3.7	21.0	21.0	0.	0.

SHUT DOWN AT 12.1330 FOR 5 HOURS

12.1830	6.5	14.0	10.8	249.	2.10	2.5	4.4	21.0	21.0	0.	0.		
12.1930	4.9	14.0	12.0	240.	0.50	2.8	4.4	21.0	21.0	0.	0.		
12.2030	4.8	14.0	12.1	232.	0.10	4.1	5.1	21.0	21.0	0.	0.		
12.2130	5.4	14.0	11.7	229.	0.	5.2	5.3	21.0	21.0	0.	0.		
12.2230	5.4	14.0	11.7	226.	0.	6.7	4.9	21.0	21.0	0.	0.		
12.2330	6.0	14.0	11.2	208.	0.	6.7	5.4	21.0	21.0	0.	0.		
13.0030	6.0	14.0	11.2	192.	0.	5.6	5.4	21.0	21.0	0.	0.		
13.0130	6.0	14.0	11.2	186.	0.	4.5	6.0	21.0	21.0	0.	0.		
13.0230	6.0	14.0	11.2	176.	0.	6.7	5.8	21.0	21.0	0.	0.		
13.0330	6.5	14.0	10.8	181.	0.	8.0	3.7	21.0	21.0	0.	0.		
13.0430	6.5	14.0	10.8	230.	0.	11.9	2.7	21.0	21.0	0.	0.		
13.0530	6.5	14.0	10.8	222.	0.	17.0	0.	21.0	21.0	0.	0.		
13.0630	6.5	14.0	10.8	222.	0.	10.0	0.	21.0	21.0	0.	0.		
13.0730	6.5	14.0	10.9	224.	4.00	0.9	0.	21.0	21.0	0.	0.		
13.0830	6.0	14.0	11.2	218.	21.00	1.0	0.	18.0	17.1	1.85	3.66		
13.0930	5.9	14.0	11.3	211.	0.	0.	0.	21.0	21.0	0.	0.		
13.1030	5.7	14.0	11.5	207.	0.	0.	0.	21.0	21.0	0.	0.		
13.1130	5.6	14.0	11.6	197.	0.	0.	0.	21.0	21.0	0.	0.		
13.1230	5.4	14.0	11.7	187.	0.	0.	0.	21.0	21.0	0.	0.		
13.1330	5.2	14.0	11.8	174.	0.	0.	0.	21.0	21.0	0.	0.		
13.1430	5.0	14.0	12.0	186.	0.	0.	0.	21.0	21.0	0.	0.		
13.1530	4.8	14.0	12.1	189.	0.	7.4	4.9	21.0	21.0	0.	0.		
13.1630	4.5	14.0	12.3	171.	0.	5.3	5.4	21.0	21.0	0.	0.		
13.1730	4.5	14.0	12.3	167.	0.20	4.7	5.3	21.0	21.0	0.	0.		
13.1830	4.5	14.0	12.3	167.	0.10	4.5	5.6	21.0	21.0	0.	0.		
13.1930	4.5	14.0	12.4	164.	0.	5.2	5.6	21.0	21.0	0.	0.		
13.2030	4.8	14.0	12.1	198.	0.	9.6	5.8	21.0	21.0	0.	0.		
13.2130	4.6	14.0	12.3	188.	0.	7.4	4.8	21.0	21.0	0.	0.		

APPENDIX C: TABLE 5.

RUN 10: GAS COMPOSITIONS

	F L U E G A S				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
13.2230	5.1	14.0	11.9	187.	0.	11.9	3.7	21.0	21.0	0.	0.
13.2330	5.6	14.0	11.5	173.	0.	10.7	4.8	21.0	21.0	0.	0.
14.0030	7.2	14.0	10.3	171.	0.	8.2	5.1	21.0	21.0	0.	0.
14.0130	7.0	14.0	10.5	171.	0.	10.7	4.4	21.0	21.0	0.	0.
14.0230	6.8	14.0	10.6	172.	0.	15.4	0.	21.0	21.0	0.	0.
14.0330	6.5	14.0	10.9	172.	0.	13.7	0.	21.0	21.0	0.	0.
14.0430	6.0	14.0	11.2	174.	0.	12.1	0.	21.0	21.0	0.	0.
14.0530	6.8	14.0	10.6	170.	0.	10.7	3.7	19.0	19.0	0.46	1.97
14.0630	7.3	14.0	10.2	137.	0.40	9.0	6.2	19.0	19.0	0.55	2.04
14.0730	6.8	14.0	10.6	151.	0.40	5.8	6.5	19.0	19.0	1.40	1.97
14.0830	7.0	14.0	10.4	167.	0.40	5.5	4.0	19.0	19.0	1.55	2.00
14.0930	7.0	14.0	10.5	181.	0.40	3.4	5.1	19.2	19.2	1.40	1.80
14.1030	7.2	14.0	10.3	196.	0.40	4.7	4.2	19.2	19.2	1.40	1.83
14.1130	6.8	14.0	10.6	186.	0.40	3.2	4.9	19.2	19.2	1.32	1.77
14.1230	6.4	14.0	10.9	254.	0.40	3.4	4.9	19.0	19.0	1.85	1.92
14.1330	6.0	14.0	11.2	282.	1.00	3.7	4.6	16.4	16.3	2.98	4.42
14.1430	6.0	14.0	11.1	292.	1.00	4.1	4.2	16.4	16.3	2.98	4.42
14.1530	5.5	14.0	11.5	312.	0.10	4.2	4.9	16.6	16.6	2.64	3.97
14.1630	5.5	14.0	11.5	317.	0.10	3.9	4.9	17.6	17.1	2.48	3.53
14.1730	5.3	14.0	11.7	320.	0.20	3.6	4.9	17.6	17.0	2.08	3.53
14.1830	5.2	14.0	11.8	317.	0.20	4.5	4.6	17.6	17.2	2.08	3.37
14.1930	5.5	14.0	11.5	319.	0.30	4.2	4.8	17.6	17.0	2.08	3.61
14.2030	5.2	14.0	11.7	316.	0.40	3.9	4.8	17.5	16.9	2.08	3.61
14.2130	5.2	14.0	11.7	313.	0.90	4.7	4.8	17.5	16.7	2.00	3.83
14.2230	5.2	14.0	11.7	317.	1.10	4.2	5.1	17.5	16.7	2.00	3.83
14.2330	5.2	14.0	11.8	345.	0.40	2.7	4.9	17.6	16.7	2.00	3.83
15.0030	5.2	14.0	11.7	353.	0.20	3.6	5.4	17.6	16.7	2.00	3.83
15.0130	5.2	14.0	11.7	361.	0.20	3.9	5.8	17.8	16.7	2.00	3.83

15.0230	5.2	14.0	11.7	359.	0.20	3.3	5.6	18.0	16.7	1.85	3.83
15.0330	6.0	14.0	11.2	350.	0.30	3.2	5.8	18.2	18.2	1.76	2.61
15.0430	6.0	14.0	11.2	331.	0.50	2.8	6.0	18.5	18.5	1.62	2.33
15.0530	6.1	14.0	11.1	322.	0.80	3.9	5.8	18.0	17.1	1.85	3.62
15.0630	5.5	14.0	11.6	343.	1.00	3.0	6.2	18.3	18.3	1.69	2.44
15.0730	6.0	14.0	11.1	350.	1.40	4.2	5.4	17.3	16.5	2.32	4.19
15.0830	6.0	14.0	11.1	349.	2.00	4.1	6.5	17.2	16.5	2.32	4.19
15.0930	5.8	14.0	11.3	360.	2.30	4.2	5.3	17.3	17.1	2.32	3.61
15.1030	5.9	14.0	11.3	372.	0.70	3.3	4.8	18.2	18.2	1.76	2.60
15.1130	5.8	14.0	11.3	375.	0.60	4.7	4.2	16.9	17.0	2.56	3.70
15.1230	5.0	14.0	11.9	400.	0.30	3.9	5.4	17.1	16.6	2.48	3.83
15.1330	5.1	14.0	11.8	403.	0.30	4.1	5.3	17.0	16.7	2.48	3.82
15.1430	5.1	14.0	11.8	419.	0.40	3.4	5.4	17.2	16.9	2.48	3.61
15.1530	5.1	14.0	11.8	403.	0.30	3.7	5.4	17.1	16.7	2.40	3.83
15.1630	5.1	14.0	11.8	412.	0.40	4.4	5.4	17.2	16.7	2.48	3.83
15.1730	5.0	14.0	11.9	417.	0.30	3.9	6.0	17.2	16.9	2.40	3.60
15.1830	5.0	14.0	11.9	424.	0.30	3.9	5.4	17.1	16.9	2.48	3.60
15.1930	5.2	14.0	11.7	425.	0.30	4.1	6.2	17.2	16.9	2.40	3.61
15.2030	5.2	14.0	11.8	437.	0.30	3.9	6.0	16.8	16.9	2.16	3.61
15.2130	4.5	14.0	12.3	438.	0.40	3.3	5.8	16.0	17.0	2.40	3.37
15.2230	4.8	14.0	12.1	441.	0.20	3.0	6.0	16.0	16.6	2.64	3.82
15.2330	4.8	14.0	12.1	453.	0.20	3.0	6.4	18.0	17.0	2.16	3.45
16.0030	4.4	14.0	12.4	489.	1.00	2.5	4.8	18.0	17.0	2.48	3.37
16.0130	4.2	14.0	12.7	567.	0.80	1.9	5.3	21.0	21.0	0.	0.
16.0230	4.2	14.0	12.7	561.	1.20	1.4	4.4	21.0	21.0	0.	0.
16.0330	4.2	14.0	12.7	541.	1.80	1.4	3.3	21.0	21.0	0.	0.
16.0430	4.0	14.0	12.8	534.	0.	4.5	5.8	21.0	21.0	0.	0.
16.0530	3.8	14.0	13.0	373.	0.	6.5	5.1	21.0	21.0	0.	0.
16.0630	3.8	14.0	12.9	423.	0.	11.9	4.2	21.0	21.0	0.	0.
16.0730	3.0	14.0	13.5	544.	0.	6.8	4.4	21.0	21.0	0.	0.
16.0830	2.8	14.0	13.7	605.	0.20	5.5	4.4	21.0	21.0	0.	0.
16.0930	2.8	14.0	13.7	625.	0.40	5.0	4.0	21.0	21.0	0.	0.
16.1030	3.6	14.0	13.1	507.	0.80	7.0	5.3	21.0	21.0	0.	0.
16.1130	3.4	14.0	13.2	501.	0.90	8.6	6.9	21.0	21.0	0.	0.
16.1230	3.8	14.0	12.9	500.	1.00	10.3	6.0	21.0	21.0	0.	0.
16.1330	3.7	14.0	12.9	497.	1.00	15.7	0.	21.0	21.0	0.	0.

APPENDIX C: TABLE 5.

RUN 10: 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
16.1430	3.3	14.0	13.2	501.	1.00	16.3	0.	21.0	21.0	0.	0.
16.1530	3.7	14.0	12.9	516.	0.30	14.6	0.	21.0	21.0	0.	0.
16.1630	3.2	14.0	12.2	538.	0.60	12.6	2.7	21.0	21.0	0.	0.
16.1730	2.6	14.0	13.8	540.	0.50	11.2	3.8	21.0	21.0	0.	0.
16.1830	3.0	14.0	13.5	514.	0.50	10.0	4.2	21.0	21.0	0.	0.
16.1930	2.8	14.0	13.7	590.	0.80	6.7	4.8	21.0	21.0	0.	0.
SHUT DOWN AT 16.1930 FOR 37 HOURS											
18.0930	5.1	14.0	11.9	182.	2.80	2.8	1.3	21.0	21.0	0.	0.
18.1030	5.2	14.0	11.9	187.	0.	5.0	6.4	21.0	21.0	0.	0.
18.1130	4.9	14.0	12.1	242.	0.	6.5	6.7	21.0	21.0	0.	0.
18.1230	4.8	14.0	12.2	302.	0.	6.7	5.8	21.0	21.0	0.	0.
18.1330	4.8	14.0	12.2	363.	0.	3.6	6.9	21.0	21.0	0.	0.
18.1430	4.8	14.0	12.2	403.	0.	2.7	6.2	21.0	21.0	0.	0.
18.1530	4.7	14.0	12.3	431.	0.	2.5	6.5	21.0	21.0	0.	0.
18.1630	4.4	14.0	12.5	446.	0.	2.5	5.8	21.0	21.0	0.	0.
18.1730	5.0	14.0	12.0	458.	0.	2.5	5.4	21.0	21.0	0.	0.
18.1830	5.0	14.0	12.0	453.	0.	2.5	5.4	21.0	21.0	0.	0.
SHUT DOWN AT 18.1830 FOR 53 HOURS CHANGE TO TJ MEDIUM VACUUM RESIDUUM FUEL											
20.2330	6.2	14.0	11.2	310.	0.	13.5	1.3	18.8	18.8	0.	2.08
21.0030	6.0	14.0	11.3	342.	0.	6.0	6.2	18.6	18.6	0.	2.24
21.0130	6.0	14.0	11.3	413.	0.	7.2	6.5	18.7	18.7	0.	2.15
21.0230	6.0	14.0	11.4	453.	0.2	3.0	7.4	18.7	18.7	0.	2.15
21.0330	6.1	14.0	11.3	501.	0.2	1.0	8.5	18.6	18.6	0.	2.26

5

21.0430	6.5	14.0	11.0	539.	0.4	1.0	8.0	18.5	18.5	0.	2.41
21.0530	6.8	14.0	10.7	557.	0.7	1.0	7.6	18.5	18.5	0.	2.46
21.0630	7.0	14.0	10.6	555.	1.0	1.0	7.1	18.6	18.6	0.	2.40
21.0730	7.0	14.0	10.6	553.	1.0	0.4	7.1	18.6	18.6	0.	2.40
21.0830	7.8	14.0	10.0	523.	1.2	0.4	6.7	18.8	18.8	0.	2.33
21.0930	6.0	14.0	11.4	543.	1.0	1.8	6.3	18.4	18.4	0.	2.43

SHUT DOWN AT 21.0930 FOR 6 HOURS

21.1530	8.0	14.0	9.8	528.	15.0	0.4	1.8	18.4	18.4	0.	2.80
21.1630	6.7	14.0	10.8	533.	11.0	0.2	1.3	18.1	18.1	0.	2.84
21.1730	5.3	14.0	11.8	559.	7.0	0.2	0.0	17.7	16.5	0.	3.99
21.1830	5.3	14.0	11.8	554.	1.5	1.4	5.4	17.5	16.3	0.	4.18
21.1930	5.3	14.0	11.8	604.	0.5	4.4	5.3	17.7	16.5	0.	3.99
21.2030	5.7	14.0	11.5	624.	0.5	9.2	4.2	17.8	16.6	0.	4.00

APPENDIX C: TABLE 6.

RUN 10: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
1.0130	-	-	-	-	-	0.	6.7	-6.7
1.0230	-	-	-	-	-	2.7	13.4	-10.7
1.0330	0.107	0.018	0.030	0.008	0.052	16.3	20.1	-3.7
1.0430	0.214	0.039	0.059	0.015	0.101	22.3	26.6	-4.3
1.0530	-	-	-	-	-	54.3	33.4	20.9
1.0630	-	-	-	-	-	94.6	39.9	54.7
1.0730	0.322	0.054	0.086	0.020	0.161	115.5	44.1	71.4
1.0830	0.430	0.070	0.124	0.022	0.214	142.2	46.3	95.9
1.0930	-	-	-	-	-	142.2	48.4	93.8
1.1030	0.537	0.090	0.205	0.025	0.217	142.2	50.6	91.6
1.1130	0.644	0.111	0.293	0.027	0.213	142.2	52.7	89.5
1.1230	0.753	0.130	0.384	0.030	0.208	146.3	54.8	91.5
1.1330	0.862	0.149	0.452	0.033	0.228	154.4	56.9	97.6
1.1430	0.972	0.169	0.514	0.036	0.254	170.9	59.0	111.9
1.1530	1.081	0.185	0.581	0.038	0.276	187.0	61.1	125.9
1.1630	1.190	0.203	0.649	0.042	0.297	201.1	63.5	137.6
1.1730	1.299	0.221	0.718	0.048	0.312	215.0	68.5	146.5
1.1830	1.408	0.238	0.791	0.051	0.327	226.7	70.3	156.4
1.1930	1.516	0.257	0.868	0.052	0.339	237.9	71.0	166.9
1.2030	1.625	0.276	0.952	0.053	0.344	249.8	71.6	178.2
1.2130	1.733	0.298	1.032	0.054	0.349	249.8	72.3	177.5
1.2230	1.841	0.320	1.080	0.055	0.387	253.4	73.0	180.4
1.2330	1.950	0.340	1.117	0.058	0.435	268.3	75.4	192.9
2.0030	2.059	0.359	1.176	0.059	0.465	288.2	76.1	212.1
2.0130	2.167	0.377	1.251	0.060	0.479	288.2	76.7	211.5
2.0230	2.275	0.401	1.307	0.061	0.506	293.1	77.3	215.8
2.0330	2.383	0.424	1.354	0.062	0.543	293.1	77.9	215.2
2.0430	2.491	0.444	1.381	0.062	0.604	295.2	78.2	217.0

2.0530	2.603	0.465	1.473	0.066	0.600	300.2	80.3	219.9
2.0630	2.714	0.489	1.564	0.066	0.596	303.7	80.6	223.2
2.0730	2.826	0.516	1.653	0.066	0.590	305.9	80.8	225.1
2.0830	2.937	0.546	1.732	0.067	0.593	307.3	81.1	226.2
2.0930	3.048	0.574	1.818	0.067	0.589	307.3	81.3	225.9
2.1030	3.160	0.605	1.902	0.068	0.586	307.3	81.6	225.7

SHUT DOWN AT 2.1030 FOR 6 HOURS

2.1630	3.275	0.642	1.917	0.068	0.647	309.2	81.8	227.3
2.1730	3.389	0.681	1.933	0.073	0.702	309.4	84.2	225.2
2.1830	3.504	0.716	1.951	0.076	0.762	309.4	85.7	223.7
2.1930	3.627	0.749	1.951	0.081	0.846	313.0	88.5	224.5
2.2030	3.751	0.783	1.951	0.086	0.931	315.4	91.3	224.2
2.2130	3.874	0.817	1.951	0.091	1.016	323.0	94.0	229.0
2.2230	3.998	0.848	1.951	0.096	1.102	327.1	96.8	230.4
2.2330	4.121	0.881	1.951	0.101	1.188	327.1	99.5	227.6

SHUT DOWN AT 2.2330 FOR 54 HOURS

5.0630	4.211	0.905	1.966	0.107	1.234	331.6	102.3	229.3
5.0730	4.300	0.928	1.998	0.112	1.262	335.3	105.1	230.2
5.0830	4.390	0.954	2.026	0.117	1.293	337.7	107.8	229.9
5.0930	4.486	0.984	2.046	0.122	1.333	344.4	110.6	233.8
5.1030	4.583	1.015	2.083	0.127	1.358	348.6	113.3	235.3
5.1130	4.681	1.047	2.162	0.135	1.337	351.6	117.6	234.0
5.1230	4.779	1.080	2.224	0.139	1.336	353.8	119.7	234.1
5.1330	4.876	1.114	2.292	0.142	1.329	354.9	121.1	233.8
5.1430	4.982	1.149	2.370	0.145	1.319	358.4	122.4	236.0
5.1530	5.089	1.182	2.419	0.148	1.340	370.4	123.8	246.6
5.1630	5.196	1.209	2.512	0.151	1.324	380.5	125.2	255.3
5.1730	5.299	1.236	2.575	0.160	1.329	387.5	130.5	257.1
5.1830	5.403	1.262	2.644	0.162	1.335	393.8	131.7	262.1
5.1930	5.507	1.282	2.709	0.163	1.352	400.9	132.6	268.3

APPENDIX C: TABLE 6.

RUN 10: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS					EQUIVALENT BURNT STONE KILOGRAMS		
	IN	FLUE	REGEN	FINES	IN-OUT	FEED	REMOVED	IN-OUT
5.2030	5.611	1.301	2.784	0.165	1.361	408.2	133.5	274.7
5.2130	5.714	1.321	2.847	0.167	1.380	415.9	134.4	281.5
5.2230	5.818	1.345	2.935	0.168	1.370	419.9	135.3	284.7
5.2330	5.922	1.370	3.008	0.174	1.370	423.9	138.2	285.6
6.0030	6.025	1.393	3.101	0.176	1.356	429.1	139.3	289.8
6.0130	6.129	1.415	3.175	0.178	1.362	433.8	140.4	293.4
6.0230	6.233	1.437	3.269	0.180	1.347	434.9	141.5	293.4
6.0330	6.336	1.465	3.352	0.182	1.337	435.6	142.6	292.9
6.0430	6.430	1.490	3.446	0.184	1.309	436.3	143.8	292.5
6.0530	6.512	1.514	3.538	0.191	1.269	438.2	147.7	290.4
6.0630	6.594	1.538	3.606	0.193	1.256	440.9	148.8	292.1
6.0730	6.676	1.565	3.654	0.196	1.262	443.9	149.9	294.0
6.0830	6.758	1.593	3.732	0.198	1.235	445.8	150.9	294.8
6.0930	6.840	1.616	3.783	0.201	1.239	448.6	152.0	296.6
6.1030	6.924	1.641	3.843	0.203	1.237	460.2	153.1	307.1
6.1130	7.008	1.663	3.899	0.208	1.238	470.7	156.1	314.6
6.1230	7.091	1.686	3.955	0.213	1.237	474.5	162.1	312.4
6.1330	7.175	1.709	4.011	0.222	1.234	477.4	173.0	304.4
6.1430	7.264	1.737	4.070	0.230	1.226	479.7	183.9	295.8
SHUT DOWN AT 6.1430 FOR 134 HOURS								
12.0530	7.378	1.770	4.070	0.239	1.299	479.7	194.8	284.9
12.0630	7.479	1.799	4.070	0.275	1.335	479.7	229.3	250.4
12.0730	7.589	1.830	4.070	0.375	1.314	494.0	312.2	181.8
12.0830	7.699	1.856	4.070	0.381	1.392	521.2	316.1	205.1
12.0930	7.812	1.885	4.070	0.386	1.470	521.2	319.9	201.3
12.1030	7.926	1.907	4.147	0.392	1.479	522.3	323.8	198.5
12.1130	8.040	1.928	4.211	0.399	1.502	530.6	329.4	201.2
12.1230	8.155	1.950	4.267	0.403	1.535	544.6	331.9	212.7

12.1330	8.268	1.973	4.323	0.405	1.568	550.4	333.1	217.3

SHUT DOWN AT 12.1330 FOR 5 HOURS

12.1830	8.366	1.994	4.390	0.407	1.576	564.6	334.3	230.3
12.1930	8.474	2.015	4.455	0.410	1.594	577.8	337.5	240.3
12.2030	8.582	2.035	4.530	0.414	1.604	591.0	339.9	251.0
12.2130	8.689	2.055	4.596	0.421	1.618	606.2	345.4	260.9
12.2230	8.798	2.075	4.663	0.426	1.634	619.6	349.0	270.6
12.2330	8.905	2.094	4.732	0.429	1.650	630.3	351.0	279.3
13.0030	9.012	2.112	4.800	0.432	1.669	645.3	352.5	292.7
13.0130	9.120	2.129	4.875	0.434	1.682	655.6	354.1	301.6
13.0230	9.228	2.145	4.949	0.436	1.698	670.2	355.6	314.6
13.0330	9.337	2.163	4.997	0.439	1.740	690.2	357.1	333.1
13.0430	9.447	2.185	5.031	0.441	1.790	708.1	358.6	349.5
13.0530	9.556	2.206	5.031	0.447	1.871	724.9	363.7	361.2
13.0630	9.663	2.227	5.031	0.450	1.955	742.7	365.3	377.4
13.0730	9.769	2.248	5.031	0.453	2.037	742.7	367.2	375.6
13.0830	9.878	2.268	5.031	0.543	2.036	750.6	440.8	309.8
13.0930	9.983	2.287	5.031	0.546	2.119	752.5	442.6	309.9
13.1030	10.088	2.305	5.031	0.548	2.203	752.5	444.5	308.0
13.1130	10.193	2.323	5.031	0.553	2.286	752.5	448.4	304.1
13.1230	10.298	2.339	5.031	0.556	2.372	752.5	449.8	302.7
13.1330	10.404	2.354	5.031	0.557	2.461	767.3	450.9	316.4
13.1430	10.510	2.370	5.031	0.608	2.500	782.3	495.0	287.3
13.1530	10.616	2.386	5.107	0.610	2.513	797.8	496.1	301.8
13.1630	10.722	2.400	5.187	0.611	2.525	816.5	497.1	319.3
13.1730	10.830	2.414	5.262	0.615	2.539	836.6	500.2	336.4
13.1830	10.937	2.428	5.344	0.616	2.549	856.2	501.2	355.0
13.1930	11.044	2.441	5.426	0.622	2.555	863.7	504.8	358.9
13.2030	11.152	2.458	5.514	0.628	2.553	887.9	508.9	379.0
13.2130	11.258	2.473	5.578	0.633	2.574	904.6	513.0	391.7

APPENDIX C: TABLE 6.

RUN 10: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
13.2230	11.364	2.489	5.637	0.638	2.599	914.4	517.0	397.4
13.2330	11.467	2.504	5.716	0.645	2.602	930.4	523.3	407.1
14.0030	11.572	2.521	5.800	0.649	2.602	930.4	526.5	403.9
14.0130	11.677	2.537	5.873	0.652	2.614	930.4	529.0	401.4
14.0230	11.783	2.553	5.873	0.656	2.701	930.4	531.4	398.9
14.0330	11.889	2.569	5.873	0.659	2.787	930.4	533.9	396.5
14.0430	11.994	2.585	5.873	0.662	2.874	930.4	536.3	394.0
14.0530	12.094	2.600	5.939	0.666	2.889	930.4	538.8	391.6
14.0630	12.193	2.613	6.052	0.673	2.855	930.4	543.9	386.5
14.0730	12.290	2.626	6.169	0.677	2.819	930.4	546.3	384.0
14.0830	12.388	2.642	6.240	0.680	2.827	930.4	548.5	381.9
14.0930	12.486	2.658	6.330	0.683	2.814	930.4	550.6	379.7
14.1030	12.584	2.676	6.405	0.686	2.817	930.4	552.8	377.6
14.1130	12.681	2.693	6.495	0.691	2.803	930.4	556.9	373.4
14.1230	12.788	2.717	6.577	0.694	2.801	930.4	558.8	371.6
14.1330	12.899	2;743	6.666	0.696	2.794	930.4	560.3	370.1
14.1430	12.993	2.767	6.733	0.698	2.794	930.4	561.8	368.5
14.1530	13.096	2.794	6.805	0.700	2.797	930.4	563.4	367.0
14.1630	13.197	2.821	6.873	0.702	2.801	930.4	564.9	365.5
14.1730	13.297	2.848	6.941	0.706	2.802	930.4	568.5	361.9
14.1830	13.398	2.874	7.005	0.708	2.811	930.4	570.1	360.3
14.1930	13.498	2.900	7.073	0.710	2.814	930.4	571.8	358.6
14.2030	13.597	2.926	7.142	0.713	2.817	930.4	573.5	356.9
14.2130	13.697	2.952	7.211	0.715	2.819	930.4	575.1	355.3
14.2230	13.797	2.978	7.285	0.717	2.817	930.4	576.7	353.6
14.2330	13.897	3.006	7.351	0.722	2.818	930.4	580.5	349.8
15.0030	13.997	3.035	7.415	0.724	2.823	930.4	581.8	348.6
15.0130	14.097	3.065	7.485	0.725	2.822	930.4	582.7	347.6

15.0230	14.196	3.094	7.551	0.726	2.825	930.4	583.6	346.7
15.0330	14.296	3.124	7.620	0.728	2.824	930.4	584.6	345.8
15.0430	14.395	3.152	7.692	0.729	2.821	930.4	585.5	344.9
15.0530	14.494	3.180	7.762	0.733	2.819	930.4	588.6	341.7
15.0630	14.592	3.208	7.837	0.734	2.812	930.4	589.7	340.7
15.0730	14.690	3.238	7.904	0.736	2.813	930.4	590.8	339.5
15.0830	14.789	3.268	7.963	0.737	2.821	930.4	592.0	338.4
15.0930	14.886	3.298	8.022	0.739	2.827	930.4	593.1	337.2
15.1030	14.988	3.330	8.088	0.740	2.829	930.4	594.3	336.1
15.1130	15.087	3.362	8.147	0.745	2.833	930.4	598.1	332.3
15.1230	15.186	3.394	8.225	0.746	2.821	930.4	599.0	331.4
15.1330	15.287	3.427	8.302	0.747	2.812	930.4	599.6	330.8
15.1430	15.389	3.461	8.375	0.748	2.804	930.4	600.2	330.1
15.1530	15.483	3.492	8.440	0.748	2.801	930.4	600.8	329.5
15.1630	15.578	3.525	8.504	0.749	2.800	930.4	601.5	328.9
15.1730	15.682	3.559	8.576	0.753	2.794	930.4	604.3	326.1
15.1830	15.778	3.592	8.637	0.754	2.795	930.4	605.0	325.4
15.1930	15.876	3.627	8.715	0.755	2.780	930.4	605.7	324.6
15.2030	15.975	3.662	8.778	0.756	2.778	930.4	606.5	323.9
15.2130	16.075	3.697	8.840	0.757	2.781	930.4	607.3	323.1
15.2230	16.177	3.733	8.910	0.758	2.776	930.4	608.0	322.3
15.2330	16.276	3.768	8.991	0.762	2.754	930.4	611.4	319.0
16.0030	16.374	3.806	9.047	0.764	2.758	930.4	612.0	318.3
16.0130	16.472	3.848	9.112	0.765	2.747	930.4	612.5	317.9
16.0230	16.569	3.890	9.159	0.767	2.753	930.4	612.9	317.4
16.0330	16.666	3.931	9.196	0.768	2.771	930.4	613.4	317.0
16.0430	16.762	3.970	9.264	0.770	2.758	930.4	613.9	316.5
16.0530	16.860	3.997	9.325	0.781	2.757	930.4	622.0	308.4
16.0630	16.958	4.028	9.380	0.782	2.769	930.4	622.4	308.0
16.0730	17.056	4.066	9.433	0.782	2.775	930.4	622.7	307.6
16.0830	17.155	4.108	9.485	0.783	2.779	930.4	623.0	307.3
16.0930	17.252	4.152	9.534	0.783	2.784	930.4	623.4	307.0
16.1030	17.349	4.188	9.600	0;784	2.777	930.4	623.7	306.7
16.1130	17.445	4.223	9.685	0.784	2.753	930.4	624.0	306.4
16.1230	17.544	4.260	9.761	0.785	2.738	930.4	624.3	306.1
16.1330	17.644	4.296	9.761	0.785	2.801	930.4	624.6	305.7

APPENDIX C: TABLE 6.

RUN 10: SULPHUR AND STONE CUMULATIVE BALANCE.

DAY.HOUR	TOTAL SULPHUR KILOMOLS IN	FLUE	REGEN	FINES	IN-OUT	EQUIVALENT BURNT STONE KILOGRAMS FEED	REMOVED	IN-OUT
16.1430	17.743	4.332	9.761	0.786	2.864	930.4	624.9	305.4
16.1530	17.842	4.369	9.761	0.786	2.925	930.4	625.3	305.1
16.1630	17.937	4.409	9.811	0.862	2.854	930.4	678.0	252.3
16.1730	18.035	4.446	9.881	0.865	2.843	930.4	680.5	249.8
16.1830	18.134	4.482	9.960	0.868	2.824	930.4	681.8	248.6
16.1930	18.232	4.523	10.038	0.872	2.799	930.4	683.9	246.5
SHUT DOWN AT 16.1930 FOR 37 HOURS								
18.0930	18.337	4.538	10.053	0.876	2.869	951.9	686.1	265.8
18.1030	18.440	4.554	10.134	0.881	2.872	951.9	688.2	263.6
18.1130	18.544	4.574	10.221	0.888	2.861	951.9	692.6	259.3
18.1230	18.647	4.599	10.296	0.888	2.864	951.9	694.4	257.5
18.1330	18.751	4.629	10.382	0.888	2.852	951.9	695.3	256.5
18.1430	18.855	4.662	10.458	0.888	2.847	951.9	696.3	255.6
18.1530	18.960	4.698	10.539	0.888	2.835	951.9	697.2	254.6
18.1630	19.063	4.734	10.609	0.888	2.832	951.9	698.2	253.7
18.1730	19.168	4.772	10.676	0.891	2.829	951.9	701.3	250.5
18.1830	19.271	4.810	10.742	0.891	2.828	951.9	701.8	250.0
SHUT DOWN AT 18.1830 FOR 53 HOURS CHANGE TO TJ MEDIUM VACUUM RESIDUUM FUEL								
20.2330	0.129	0.027	0.010	0.004	0.088	0.	2.2	-2.2
21.0030	0.259	0.058	0.053	0.007	0.141	0.	4.4	-4.4
21.0130	0.389	0.094	0.097	0.011	0.187	0.	6.6	-6.6
21.0230	0.520	0.134	0.193	0.015	0.177	0.	8.8	-8.8
21.0330	0.651	0.179	0.302	0.019	0.151	0.	11.0	-11.0
21.0430	0.780	0.229	0.403	0.022	0.126	0.	13.2	-13.2

21.0530	0.907	0.279	0.500	0.026	0.101	0.	15.3	-15.3
21.0630	1.031	0.330	0.590	0.030	0.081	0.	17.5	-17.5
21.0730	1.153	0.379	0.678	0.034	0.061	0.	19.6	-19.6
21.0830	1.272	0.428	0.763	0.037	0.044	0.	21.7	-21.7
21.0930	1.403	0.476	0.843	0.041	0.043	0.	23.8	-23.8

SHUT DOWN AT 21.0930 FOR 6 HOURS

21.1530	1.535	0.531	0.863	0.045	0.096	0.	25.9	-25.9
21.1630	1.666	0.580	0.877	0.048	0.159	0.	28.0	-28.0
21.1730	1.796	0.628	0.877	0.057	0.235	0.	32.2	-32.2
21.1830	1.927	0.675	0.933	0.057	0.263	0.	32.3	-32.3
21.1930	2.059	0.727	0.988	0.057	0.288	0.	32.4	-32.4
21.2030	2.192	0.781	1.037	0.057	0.317	0.	32.4	-32.4

APPENDIX C: TABLE 7

SOLIDS REMOVED DURING RUN 10, KG. (RAW DATA)

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	BOILER BACK	BOILER FLUE	CYCLONES
1.0700	-	-	-	-	44.00	-
1.1600	-	-	-	-	19.05	-
1.1800	-	2.27	-	-	5.90	-
2.0040	0.91	0.91	-	4.54	3.63	-
2.0330	-	-	-	0.91	-	-
2.0600	0.91	0.91	-	-	1.81	-
2.1615	-	-	1.81	4.99	10.43	-
2.1800	0.91	0.91	0.91	0.45	1.81	0.45
3.0615	2.04	1.81	-	-	-	-
3.1130	-	-	2.04	-	-	-
5.0500	-	-	-	-	1.81	0.45
5.0600	-	-	-	-	1.81	0.45
5.1200	0.68	0.68	-	28.12	10.43	0.23
5.1800	2.27	1.59	-	4.54	5.90	0.23
5.2359	0.45	0.45	1.81	1.81	4.08	1.36
6.0600	1.36	0.91	0.91	0.91	6.35	0.68
6.1200	0.91	0.91	-	3.63	1.81	0.23
6.1800	0.91	0.91	-	-	1.81	-
7.1220	0.91	0.91	-	2.27	1.81	-
7.2345	-	-	-	-	-	37.19
8.0500	-	55.79	-	-	-	-
8.0805	-	21.77	-	-	-	-
8.1200	54.43	-	-	-	-	-
12.0700	0.91	0.91	-	54.43	1.81	22.68
12.0815	-	-	-	-	-	78.47
12.1200	0.91	0.91	4.99	7.71	12.70	-
12.1415	-	-	-	-	-	7.94
12.1805	-	-	0.91	4.99	19.05	-
12.2000	0.91	0.91	0.45	1.81	3.18	0.23
12.2245	0.91	0.91	-	3.63	7.26	-
13.0610	0.91	0.91	-	3.63	8.16	1.81
13.0915	-	-	-	-	-	73.94
13.1200	0.91	0.91	1.36	6.80	4.54	0.23
13.1540	-	-	-	-	-	44.45
13.1820	-	-	-	-	-	23.59
13.1840	0.91	0.91	0.45	4.54	2.72	0.23
13.2359	0.91	0.91	0.91	9.07	13.61	0.45
14.0720	0.91	0.71	0.45	5.90	13.61	0.91
14.1200	0.91	0.91	0.45	6.80	3.63	0.23
14.1800	0.91	0.91	0.45	5.44	4.08	0.23

APPENDIX C: TABLE 7

SOLIDS REMOVED DURING RUN 10, KG. (RAW DATA) PAGE 2 OF 2

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	BOILER BACK	BOILER FLUE	CYCLONES
14.2359	0.91	0.91	0.45	3.63	7.26	0.45
15.0600	0.91	0.91	0.23	3.18	3.18	0.45
15.1200	1.36	1.36	-	4.76	3.18	-
15.1800	0.91	0.91	0.23	-	1.81	0.45
15.2359	0.91	1.36	0.45	4.99	2.72	0.45
16.0600	1.04	0.23	2.72	0.45	0.45	7.31
16.1700	27.22	-	-	-	-	29.03
16.1800	0.91	0.91	-	2.72	1.81	0.45
17.0300	36.27	-	-	-	-	-
18.0045	15.88	-	-	-	-	-
18.0830	-	26.76	-	-	-	18.14
18.1200	0.91	0.91	0.23	8.16	5.44	0.45
18.1800	0.91	0.91	0.23	2.72	2.72	0.45
19.0715	-	-	-	-	7.48	-
19.1030	-	60.78	-	-	-	-
19.1200	0.91	37.65	4.08	3.18	4.54	0.45
20.0245	-	-	-	-	13.61	-
20.1200	-	-	-	-	4.54	-
21.1215	-	-	-	-	13.61	-
21.1800	0.91	0.91	2.27	28.12	4.54	0.45
22.1200	0.91	0.91	-	-	1.81	0.45

APPENDIX C: TABLE 8

ANALYSIS OF SOLIDS REMOVED DURING RUN 10

C A R B O N WT.%

DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	BOILER BACK	BOILER FLUE	CYCLONES
1.0700	-	-	-	-	0.89	-
2.0040	0.96	0.28	-	0.60	9.24	-
2.0600	0.17	0.10	-	1.22	-	-
2.1800	1.82	0.59	17.80	1.12	25.89	6.84
3.0615	0.88	1.72	-	-	-	-
5.0600	-	-	-	-	-	11.09
5.1200	0.17	0.18	-	-	-	-
5.1800	0.17	0.05	-	3.23	30.10	-
5.2359	0.47	0.05	2.37	-	-	9.98
6.0600	0.20	0.05	2.43	-	-	5.42
6.1200	0.15	0.05	-	0.90	-	6.96
6.1800	0.05	0.05	-	-	-	-
12.0700	0.10	0.05	-	0.66	6.34	-
12.1200	0.12	0.05	0.76	-	-	-
12.1805	-	-	-	-	2.49	-
12.1200	0.10	0.05	-	0.55	9.42	-
12.2245	0.05	0.05	-	1.97	4.80	-
13.0610	0.14	0.14	-	-	-	0.96
13.1200	0.35	0.05	-	-	-	-
13.1840	0.26	0.05	-	-	2.70	-
13.2359	0.85	0.12	1.73	0.60	4.60	2.00
14.0720	0.61	0.07	-	-	4.14	-
14.1200	0.05	0.20	-	-	-	-
14.1800	0.05	0.05	-	0.74	1.59	-
14.2359	0.05	0.13	3.74	0.48	15.80	1.86
15.0600	0.42	0.14	-	-	-	0.64
15.1200	0.14	0.05	-	-	-	-
15.1800	0.14	0.05	-	-	-	1.16
15.2359	0.13	0.05	13.40	0.72	25.80	5.34
16.0600	0.49	0.08	-	-	-	10.50
16.1200	0.05	0.10	-	-	-	-
16.1800	0.16	0.12	-	1.60	32.70	8.44
18.1200	0.28	0.06	0.75	2.08	28.30	6.12
18.1800	0.05	0.05	-	-	-	6.11
19.1030	-	3.16	-	-	-	-
19.1200	6.23	1.84	30.60	3.59	54.70	17.50
21.1800	0.57	0.05	4.74	2.97	40.50	4.36
22.1200	1.19	0.06	-	-	-	6.07

APPENDIX C: TABLE 9

ANALYSIS OF SOLIDS REMOVED DURING RUN 10

DAY.HOUR	S U L P H A T E GAS'R	S U L P H U R REGEN	WT% REGEN CYCLONE	BOILER BACK	BOILER FLUE	CYCLONES
1.0700	-	-	-	-	2.68	-
2.0040	0.01	0.44	-	1.57	1.17	-
2.0600	0.01	0.47	-	1.24	-	-
2.1800	0.08	0.01	2.78	2.66	1.05	0.01
3.0615	0.01	0.07	-	-	-	-
5.0600	-	-	-	-	-	1.01
5.1200	0.01	1.06	-	-	-	-
5.1800	0.01	0.97	-	1.43	2.05	-
5.2359	0.01	0.24	5.79	-	-	0.01
6.0600	0.01	0.86	6.22	-	-	0.01
6.1200	0.01	0.99	-	2.97	-	0.01
6.1800	0.01	0.71	-	-	-	-
12.0700	0.01	1.01	-	1.72	3.61	-
12.1200	0.01	1.42	-	-	-	-
12.1805	-	-	-	-	2.04	-
12.2000	0.01	0.95	4.76	1.92	2.06	-
12.2245	0.01	0.38	-	1.25	2.20	-
13.0610	0.01	0.01	-	-	-	0.01
13.1200	0.01	0.43	-	-	-	-
13.1840	0.01	0.82	-	-	1.67	-
13.2359	0.01	0.01	2.96	2.01	2.15	0.01
14.0720	0.13	0.01	-	-	2.08	-
14.1200	0.01	0.01	-	-	-	-
14.1800	0.01	0.01	-	1.55	2.93	-
14.2359	0.01	0.50	2.51	2.07	2.01	0.01
15.0600	0.01	0.61	-	-	-	0.01
15.1200	0.13	0.75	-	-	-	-
15.1800	0.04	0.78	-	-	-	0.01
15.2359	0.01	0.51	4.39	1.78	2.56	0.01
16.0600	0.22	0.41	-	-	-	0.12
16.1200	0.01	0.01	-	-	-	-
16.1800	0.01	0.01	-	2.05	1.35	0.01
18.1200	0.01	0.01	8.40	2.88	3.43	0.55
18.1800	0.01	0.01	-	-	-	0.01
19.1030	-	0.01	-	-	-	-
19.1200	0.01	0.01	0.76	3.20	1.59	0.42
21.1800	0.06	1.51	1.12	1.98	1.85	1.81
22.1200	0.14	0.35	-	-	-	0.66

APPENDIX C: TABLE 10

ANALYSIS OF SOLIDS REMOVED DURING RUN 10

	TOTAL	SULPHUR	WT.%			
DAY.HOUR	GAS'R	REGEN	REGEN CYCLONE	BOILER BACK	BOILER FLUE	CYCLONES
1.0700	-	-	-	-	3.10	-
2.0040	4.45	3.47	-	3.49	4.68	-
2.0600	5.44	4.01	-	4.11	-	-
2.1800	6.82	6.49	8.04	4.89	3.56	4.71
3.0615	7.11	6.90	-	-	-	-
5.0600	-	-	-	-	-	5.96
5.1200	5.43	4.47	-	-	-	-
5.1800	4.88	3.99	-	4.89	5.34	-
5.2359	5.43	3.93	8.40	-	-	5.28
6.0600	5.07	3.90	8.85	-	-	5.54
6.1200	4.34	3.53	-	5.57	-	4.10
6.1800	4.34	3.83	-	-	-	-
12.0700	2.22	1.64	-	1.99	4.55	-
12.1200	2.86	1.78	10.20	-	-	-
12.1805	-	-	-	-	5.01	-
12.1200	2.28	1.60	-	3.02	3.99	-
12.2245	2.64	1.99	-	4.76	4.06	-
13.0610	3.49	2.63	-	-	-	3.76
13.1200	3.37	2.55	-	-	-	-
13.1840	3.14	1.98	-	-	4.63	-
13.2359	2.88	2.03	8.02	3.96	2.99	3.29
14.0720	4.64	3.23	-	-	4.77	-
14.1200	3.29	2.86	-	-	-	-
14.1800	3.02	2.78	-	3.78	3.77	-
14.2359	3.49	3.17	5.84	3.86	3.69	4.59
15.0600	3.51	2.85	-	-	-	3.45
15.1200	3.87	3.13	-	-	-	-
15.1800	3.77	3.02	-	-	-	3.49
15.2359	3.61	2.83	7.59	2.87	4.39	3.37
16.0600	3.48	2.93	-	-	-	3.70
16.1200	4.02	2.72	-	-	-	-
16.1800	4.26	3.97	-	4.13	4.54	4.43
18.1200	3.44	2.86	10.20	4.56	5.98	7.55
18.1800	3.23	2.49	-	-	-	6.65
19.1030	-	4.78	-	-	-	-
19.1200	5.04	4.83	7.86	5.38	5.41	6.09
21.1800	7.65	5.99	3.45	5.17	4.16	3.83
22.1200	6.23	6.16	-	-	-	7.03

APPENDIX C : TABLE 11

RUN 10 - LIMESTONE FEED PARTICLE SIZE DISTRIBUTION

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.					
50734	.0000	.4	12.1	9.1	20.0	22.3	27.9	5.9	1.2	1.0
50735	.0000	.5	17.8	12.0	24.9	26.8	16.3	1.1	.1	.4
50735	.0000	.4	15.6	12.3	25.6	25.7	17.8	1.5	.4	.6
50742	1.2359	.9	16.8	12.5	26.9	13.0	25.5	3.0	.4	1.1
50750	2.0600	1.5	13.4	9.8	21.5	25.1	23.8	3.4	.3	1.1
50754	2.1800	.3	10.1	8.7	20.0	24.2	28.6	5.8	1.1	1.3
50765	5.1200	.1	6.1	7.2	20.0	31.8	30.8	3.0	.1	.9
50775	5.1800	.3	12.6	10.3	23.6	26.7	22.5	2.7	.5	.8
50781	6.0001	.3	9.7	7.9	25.7	24.3	25.7	4.4	.9	1.1
50790	6.0600	.3	12.8	10.1	21.0	23.7	25.9	4.2	.8	1.0
50828	6.1200	.3	7.9	7.0	24.4	22.9	29.0	6.1	1.1	1.4
50847	12.1200	.3	18.7	8.7	19.5	22.1	24.5	4.4	.6	1.2
50851	12.2000	.3	10.3	8.4	18.5	22.1	29.3	7.2	1.7	2.2
50861	12.2359	.2	11.5	9.8	21.3	24.0	26.9	4.4	.7	1.4
50877	13.1200	.4	11.9	9.1	19.7	22.7	27.9	5.7	1.2	1.5
50887	13.1830	.2	13.7	10.8	23.1	24.5	23.8	2.8	.4	.9
50889	13.2359	.4	12.0	9.4	20.2	22.4	27.8	4.9	.6	2.4
50959	15.2359	1.1	11.2	9.0	19.0	21.9	28.1	6.2	1.0	2.5
50988	18.1230	.4	3.9	9.1	20.5	24.2	30.7	6.9	1.6	2.8
50998	18.1800	.4	12.3	9.6	20.6	22.3	26.6	5.1	1.0	1.9
51018	19.1200	.4	12.4	9.8	20.6	23.4	26.6	4.5	1.0	1.3

APPENDIX C : TABLE 12
RUN 10 - GASIFIER BED PARTICLE SIZE DISTRIBUTION
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.					
50736	.0000	.3	9.0	12.4	28.7	28.7	20.5	.3	.0	.0
50737	1.2359	.2	6.3	9.2	23.7	29.6	30.4	.5	.0	.2
50743	2.0600	.2	6.3	9.4	24.8	30.6	28.5	.1	.0	.0
50752	2.1200	.0	3.8	6.4	21.4	32.5	34.6	.9	.0	.4
50761	2.1800	2.7	7.5	10.2	25.4	28.8	25.1	.2	.0	.1
50764	3.0615	.1	5.6	9.5	25.9	31.0	27.7	.1	.0	.1
50766	5.1200	.1	5.6	9.3	25.2	30.3	29.0	.2	.0	.2
50778	5.1800	.2	5.5	9.4	25.4	30.5	28.2	.3	.0	.4
50782	6.0001	.2	5.1	8.8	24.3	29.8	30.3	.3	.0	1.2
50791	6.0600	.2	5.6	2.4	27.3	32.9	31.0	.3	.0	.3
50798	6.1200	.1	5.5	9.4	26.1	30.2	27.7	.3	.0	.7
50834	6.1800	.2	5.6	9.7	26.7	32.1	25.5	.2	.0	.2
50840	12.0700	.1	6.3	10.4	26.1	28.6	27.1	1.0	.0	.4
50841	12.1200	.2	7.4	11.3	27.2	28.5	24.7	.5	.0	.2
50854	12.2000	.1	7.0	10.9	26.2	27.7	25.9	1.3	.2	.7
50856	12.2359	.1	6.9	10.7	25.3	27.7	27.6	1.1	.1	.5
50863	13.0600	.3	8.7	12.4	27.6	27.6	22.7	.3	.0	.3
50878	13.1200	.1	7.0	11.1	28.4	29.7	23.3	.3	.0	.1
50885	13.1830	.2	9.1	12.8	29.5	28.5	19.8	.1	.0	.1
50890	13.2359	.3	6.6	9.9	25.2	27.5	27.5	2.0	.0	1.0
50899	14.0600	.0	5.7	9.7	25.3	29.0	28.7	1.1	.0	.5
50913	14.1200	.3	6.6	11.0	26.9	28.8	25.8	.5	.0	.1
50921	14.1800	.4	6.4	10.7	26.8	28.9	26.1	.5	.0	.2
50926	14.2359	.3	6.1	10.5	26.7	29.0	27.0	.3	.0	.1
50933	15.0600	.3	5.6	10.0	26.9	30.1	26.8	.3	.0	.1
50943	15.1200	.2	3.7	7.6	24.2	32.0	32.0	.2	.0	.0
50949	15.1800	.0	4.6	9.3	26.4	31.8	27.9	.0	.0	.0
50957	15.2359	.3	5.5	10.6	28.0	30.5	24.9	.1	.0	.1
50966	16.0600	.3	6.2	11.5	29.2	30.1	22.6	.1	.0	.0
50975	16.1130	1.4	6.8	11.1	27.7	28.9	23.8	.1	.0	.1
50980	16.1800	.2	5.2	10.6	28.6	30.9	24.4	.1	.0	.1
50990	18.1200	.2	4.3	8.8	26.6	31.1	28.7	.2	.0	.2
51001	18.1800	.3	2.9	7.1	24.8	32.2	32.2	.3	.0	.3
51009	19.1200	.2	3.5	7.5	25.2	31.6	31.3	.4	.0	.2
51020	21.1800	.1	2.4	6.7	25.7	33.6	30.9	.4	.0	.3
51027	22.1200	1.7	6.0	9.3	27.3	29.2	25.5	.6	.0	.4

APPENDIX C : TABLE 13

RUN 10 - REGENERATOR BED PARTICLE SIZE DISTRIBUTION

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.								
50738	1.2359	.0	3.9	7.0	22.3	30.8	33.7	1.8	.0	.5
50744	2.0600	.0	5.5	8.9	25.0	31.2	29.3	.2	.0	.0
50760	2.1800	.2	5.2	8.9	25.3	31.4	28.8	.2	.0	.0
50763	3.0615	3.0	7.0	10.3	25.9	28.5	25.1	.2	.0	.1
50767	5.1200	.2	4.6	7.8	22.9	30.4	33.7	.3	.0	.2
50777	5.1800	.1	4.6	8.0	23.8	30.3	31.8	.9	.2	.2
50783	6.0001	.3	5.2	8.5	23.8	30.7	30.7	.5	.0	.3
50792	6.0600	.1	4.7	8.3	23.7	30.8	31.7	.4	.0	.1
50799	6.1200	.1	4.7	8.6	24.3	30.9	31.1	.1	.0	.1
50835	6.1800	.2	5.4	8.9	26.5	32.5	26.6	.0	.0	.0
50839	12.0700	.2	6.7	10.2	26.5	28.7	26.8	.8	.0	.2
50842	12.1200	.2	7.3	11.3	27.0	28.4	25.0	.6	.0	.2
50855	12.2000	.2	6.6	10.5	25.8	28.2	27.4	1.1	.0	.2
50857	12.2359	.1	7.0	10.7	26.6	28.3	26.5	.7	.0	.1
50864	13.0600	.1	7.4	11.4	26.7	27.7	25.7	.7	.0	.1
50879	13.1200	.2	8.2	12.2	29.0	28.9	21.2	.1	.0	.2
50886	13.1830	.1	7.2	11.2	28.6	30.0	22.6	.1	.0	.1
50891	13.2359	.2	7.3	10.4	26.9	28.4	26.3	.6	.0	.0
50900	14.0600	.2	6.1	9.9	26.4	28.9	27.9	.6	.0	.0
50912	14.1200	.2	6.6	11.1	27.2	29.1	25.4	.2	.0	.1
50920	14.1800	.1	6.5	11.1	27.4	28.9	25.5	.3	.0	.1
50927	14.2359	.2	5.9	10.4	26.7	29.8	26.7	.3	.0	.0
50935	15.0600	.2	6.4	10.8	27.5	29.3	25.5	.2	.0	.0
50944	15.1200	.2	5.5	10.3	27.5	30.2	26.3	.1	.0	.0
50950	15.1800	.1	5.4	10.5	28.2	30.2	25.5	.1	.0	.0
50958	15.2359	.1	4.9	10.2	27.8	31.0	25.7	.1	.0	.1
50967	16.0600	.2	5.0	9.8	27.2	30.7	26.9	.1	.0	.0
50976	16.1130	.2	4.9	9.6	27.4	30.8	26.9	.2	.0	.2
50981	16.1800	.2	4.7	9.9	27.5	31.0	26.4	.1	.0	.0
50991	18.1200	.2	3.3	7.6	24.5	30.9	32.8	.5	.0	.1
51002	18.1800	.2	3.3	7.7	24.8	30.9	32.0	.7	.0	.2
51008	19.1030	.2	2.7	6.7	22.7	31.6	35.3	.5	.0	.2
51011	19.1200	.4	3.5	7.5	23.9	30.5	33.4	.7	.0	.1
51021	21.1800	.1	3.0	7.6	27.2	32.9	28.8	.3	.0	.0
51028	22.1200	.4	2.1	5.3	22.5	31.4	37.9	.4	.0	.2

APPENDIX C : TABLE 14
RUN 10 - GASIFIER CYCLONE DRAIN PARTICLE SIZE DISTRIBUTION
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.					
50759	2.1800	.3	3.5	5.0	14.8	19.2	34.4	8.8	2.8	11.0
50784	6.0001	.0	.6	1.4	4.5	8.5	34.2	35.2	13.9	1.7
50793	6.0600	.0	.6	1.4	5.0	9.5	38.5	19.8	6.0	19.3
50830	6.1500	.0	1.4	2.6	10.1	21.1	49.4	2.8	1.2	11.3
50866	13.0600	.0	2.5	4.2	11.8	16.5	39.5	10.9	3.4	11.2
50893	13.2359	.0	1.6	4.8	9.5	19.0	49.2	9.5	.0	6.3
50925	14.2359	1.6	3.2	4.8	12.9	17.7	38.7	11.3	.0	9.7
50936	15.0600	.0	1.6	4.9	13.1	19.7	41.0	8.2	.0	11.5
50954	15.1800	.0	2.1	3.1	11.3	17.5	39.2	8.2	.0	18.6
50960	15.2359	.0	1.7	3.5	9.8	16.8	35.8	5.8	1.2	25.4
50968	16.0600	.0	1.0	1.0	7.0	13.0	34.0	8.0	1.0	35.0
50986	16.1800	.0	6.8	13.6	29.5	27.3	15.9	2.3	.0	4.5
50933	18.1200	.0	.0	2.9	5.9	8.8	36.8	23.5	.0	22.1
51003	18.1800	.0	.0	2.0	4.0	6.0	22.0	28.0	.0	38.0
51023	21.1800	.6	.6	1.2	5.5	9.8	39.6	18.3	5.5	18.9
51029	22.1200	.0	.7	.7	2.9	5.8	22.3	21.6	10.1	36.0

APPENDIX C

TABLE 12

RUN 10 : SUMMARY OF GASIFICATION PERIODS

FUEL	FROM DAY	FROM TIME	TO DAY	TO TIME	HOURS GASIFICATIONS	CAUSE OF SHUT DOWN
Heavy Fuel Oil (HFO)	0	23.30	4	06.00	78.5	Cold fuel
HFO	5	04.30	6	20.00	39.5	Deliberate
HFO	7	03.00	7	04.30	1.5	Cold fuel
HFO	7	07.00	7	10.00	3.0	Not identified*
HFO	7	15.00	7	17.30	2.5	Not identified*
HFO	7	19.00	7	21.00	2.0	Not identified*
HFO	12	05.00	12	14.00	9.0	Not identified
HFO	12	17.30	17	05.00	107.5	Pump Seizure Fuel
HFO	18	06.00	19	02.00	20.0	
Bitumen	19	02.00	19	15.00	13.0	Boiler over temperature
Bitumen	20	21.00	21	10.00	13.0	Boiler over temperature
Bitumen	21	14.30	21	20.30	6.0	Boiler over temperature
Bitumen	22	08.00	22	10.30	2.5	
50% Bitumen 50% Coal	22	10.00	22	16.00	7.5	Boiler over-temperature
				TOTAL	305.5	

NOTE: Total hours available for gasification 520 hours

Service factor overall 59%

* Probably boiler air adjustments causing pulsating flame triggering fire eye system.

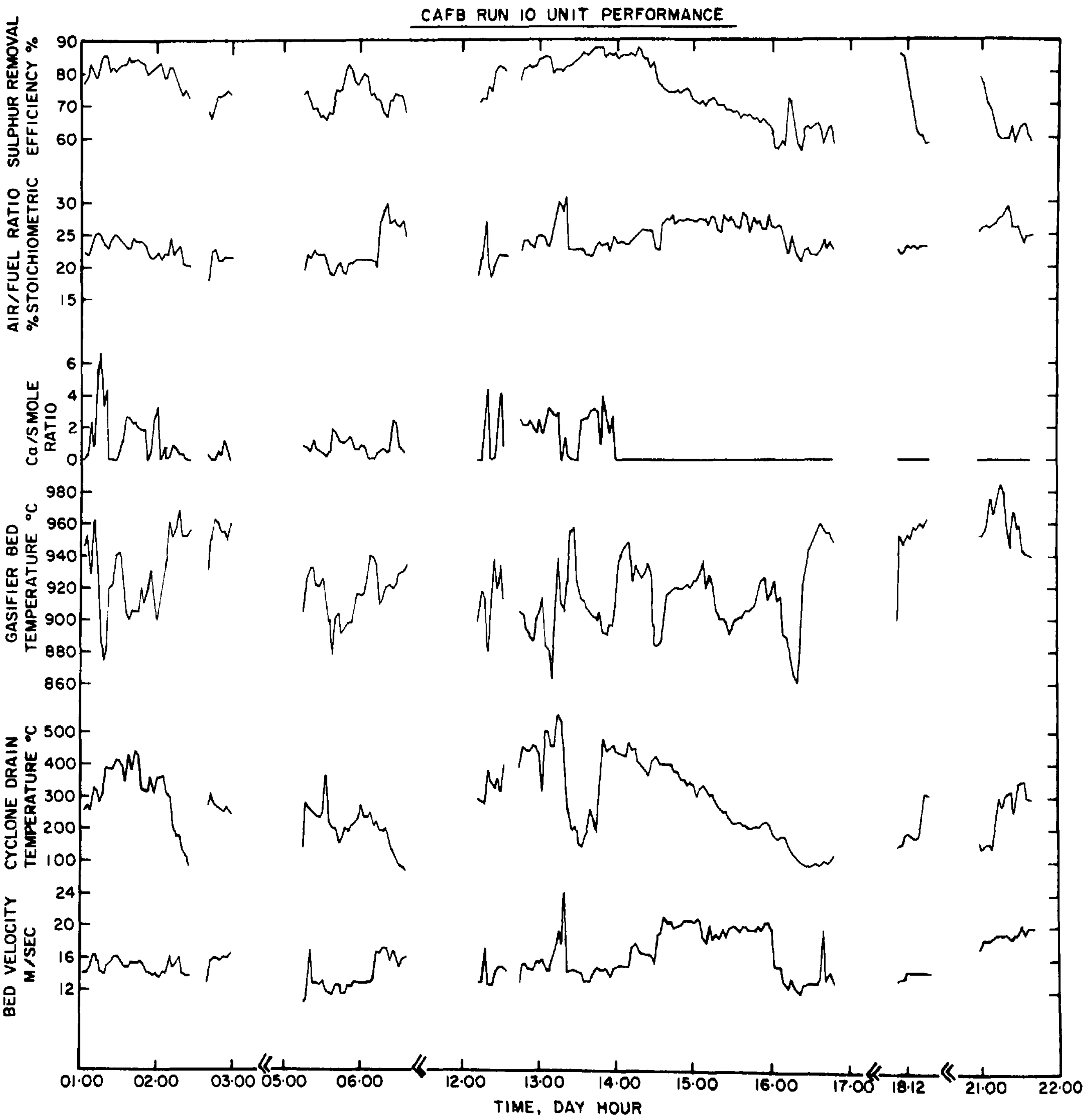

CAFB RUN 10 UNIT PERFORMANCE
SULPHUR REMOVAL EFFICIENCY %
90
80
70
60
AIR/FUEL RATIO %STOICHIOMETRIC
30
25
20
15
Ca/S MOLE RATIO
6
4
2
0
GASIFIER BED TEMPERATURE °C
980
960
940
920
900
880
860
CYCLONE DRAIN TEMPERATURE °C
500
400
300
200
100
BED VELOCITY M/SEC
24
20
16
12
01·00
02·00
03·00
05·00
06·00
12·00
13·00
14·00
15·00
16·00
17·00
18·12
21·00
22·00
TIME, DAY HOUR

FIG. C34

APPENDIX D

CAFB OPERATORS MANUAL

* Note: This appendix comprises the working instructions used during Run 10. The contents (page 240) listing referes to the numbers on page headings.

CONTENTS

CONTENTS (continued)

I. CAFB PILOT PLANT

INTRODUCTION

A continuous CAFB gasifier pilot plant has been constructed at the Esso Research Centre to provide a demonstration of the CAFB process under continuous operating conditions and to provide a means for studying those operational variables which cannot be measured in batch reactors. Features of the pilot plant are summarised here.

FLOW PLAN AND LAYOUT

Figure 1 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 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 2 shows the layout of the pilot plant equipment within its building. 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 in the control room. Electrical instrumentation and manometers are mounted in a separate control cabinet in the control room. Gasifier blowers are located in a separate blower house outside the main building, and the cooling tower is mounted on the roof.

GASIFIER - REGENERATOR UNIT

The gasifier and regenerator reactors are cylindrical cavities in a refractory concrete block which is insulated and enclosed in a steel casing as shown in figure 3 which is a vertical section through the gasifier and regenerator axes. The gasifier is 23" Dia. at the base, flares to 26" at $32\frac{1}{2}$" from the bottom and then flares more gradually to 28" Dia. over the next 95" giving an overall inside height of $127\frac{1}{2}$". The axis of the regenerator is parallel to that of the gasifier and 27" from it. Diameter of the regenerator is 7.5" at the base and flares to 10" Dia. at 43" level. Above that level the regenerator bore is constant at 10" over 89" of height, giving a total height of 132". Tops of both reactor cavities are level; bottom of the regenerator is $4\frac{1}{2}$" lower than that of the gasifier.

The block contains other cavities which make up the cyclone inlets and transfer lines through which solids circulate between gasifier and regenerator. It also contains a vertical ceramic paper separator which divides it into two separate blocks which can expand independently.

The gasifier air distributor is circular and contains 16 nozzles. The outer 12 nozzles (which are arranged in a circle about the centre) have two rows of horizontal holes (8 holes and 6 holes)) the inner 4 nozzles arranged in a square about the centre are located in a depression in the middle of the distributor and have 22 holes in three rows. The holes in the outer nozzles are 0.12" diameter, whereas the inner nozzle holes are 0.1285" diameter.

PROCESS CONTROL

Figure 4 is a diagram of the pilot plant instrumentation system. Automatic contro. boxes are used to regulate regenerator temperature and regenerator bed level, and gasifier bed level. A packaged pressurisation system maintains constant boiler cooling water pressure and temperature. All other systems are manually controlled by the process operator. Dashed lines in Figure 4 show the indicators and control valves normally used by the operator. Manometers indicate pressures and pressure differences in most applications. In four instances pressure differences are also detected by pneumatic delta pressure cells and transmitted to recorders. Pressure switches also are employed in several locations to operate warning lights for abnormal conditions.

A detailed schematic diagram of the layout of the manometers, differential pressure cells and pressure switches is given in figures 5 and 6.

FUEL SYSTEMS

(a) Heavy Fuel Oil

A schematic diagram of the heavy fuel oil system is shown in figure 7.

(b) Kerosene

Kerosene is used as a warm up fuel and is stored in a 500 gallon tank located below ground level at the back of building 3F close to the blower house. A hand lift pump transfers the fuel to two 50 gallon drums located above the tank.

Inside building 3F the kerosene passes through a fuseable firevalve before joining the heavy fuel oil line as shown in figure 7.

(c) Propane

Propane is stored in a bulk tank located between building 3F and central stores. The offtake line passes underground to a valve on the outside of building 3F. It then branches, one line goes to the CABF pilot plant, whereas the other goes to the boiler house.

(d) Bitumen

A schematic diagram of the bitumen fuel system is shown in figure 8.

Ministatic Trailer

A detailed drawing of the bitumen trailer is shown in figure 9.

The original 'LISTER' twin-cylinder diesel engine has been replaced by a 10 H.P. 3ph AC electric motor (d) rotating at 1440 rpm, tow bar end of the trailer.

The output shaft from this motor is provided with two ribbed pulleys; one drives the rotary vane compressor at 1000 rpm, whilst the other drives a hydraulic pump (b) at about motor speed. A hydraulic control valve (19) selects pump delivery on reverse or neutral positions. This pump energises a hydraulic motor (c) which is connected to the Barclay Kellett gear pump (a). The latter is fitted internally in the bitumen tank and can either circulate the hot bitumen inside the tank or discharge through the swivel pipe (41) through the 3-way valve (38). Lever (39) at positions CIRCULATE or DELIVERY respectively (see figure 9)

Details

a) The Barclay Kellet gear pump (36) is type 33 x 8T has a capacity of 4000 g.p.h. and 50 p.s.i.

b) The hydraulic gear pump (16) is a Dowty type GP2/85AU.

c) The hydraulic motor (17) is an ADAN Hydraulic Orbit Motor, type OMP-28.

d) BROOKS AC Motor, Frame C254, Serial No. L87207 10 H.P., 1440 rpm, 400/440 volts.

Items 2, 3, and 4 are mounted undercover at the tow bar (front) end of the trailer, with the electric motor contactor adjacent on the trailer chassis. (See figure 10)

The following equipment is provided on the rear end of the trailer (refer to figure 11).

i) A dial type thermometer (30) 50 - 400 °F, which has electrical contacts and a pointer which can be set to the maximum bitumen temperature required. When this predetermined temperature is attained the contacts close and operate the solenoid valve (42), cutting the air pressure to the burners (one per hematite tube) and to the diaphragm cut off valve (12), automatically shutting off the oil supply to the burners.

ii) A tank capacity gauge (29) with dial indication of 1000-6000 gallons in increments of 500 gallons. Normal minimum reading is 1000 gallons, which is the safe level to cover the heating tubes.

iii) A Mowbrey magnetic switch (31) which operates the solenoid valve (42) on the air line and shuts off the burners when the tank contents fall to about 6" above the heating tubes.

When the bitumen predetermined maximum temperature is reached the thermometer (30) contacts will close, energising the solenoid valve (42) on the air supply and exhausting the air to atmosphere. Thus the oil cut-off valve to each burner is activated and the burner, deprived of both air and oil, is extinguished.

The solenoid valve is also energised by the Mowbrey float switch operating on a low bitumen level condition. The above sequence of events follows.

Details

29) Tank Capacity Gauge

Supplied by Bayham, Basingstoke.

30) Thermometer

Supplied by Municipal Appliances, manufacturer of the trailer unit 12 Volts D.C.

31) Magnetic Switch

Mobrey Type S01/F02 No. 6605 12 volts D.C.

42) Solenoid Valve

Trist Lucifer Cat. No. 321B 05. 220 p.s.i. - 9/16" orifice, 12 volts DC.

Air Compressor Drwg No. B2598(3) (Fig 19)

This is a rotary vane compressor, suitable for pressures of 3 - 15 p.s.i., and operates normally at 1000 rpm.

Oil Burning Equipment

a) Oil Cut-Off Valve - Print B1027 (Fig 20)
   (see page 3)

b) Atomiser - Print 1123 (Fig 21)
   (see page 3)

c) Flame Failure Thermomstat 1 B1018 (Fig 29) (Samuel Lee-Bapty, Type S, Max. Temp. 1200°F)
   (see page 3)

d) Fuel Supply to the Burners
   (see page 3)

(e) Gas Oil

Gas oil is supplied to replenish the Bitumen Trailer Heating Burner storage tank and to provide gas oil purging on the Bitumen piping system. (see figure 12).

The road tanker is provided with four compartments numbered 1,2,3, and 4 from the front. The capacity of each is 500 Imperial Gallons except No. 4, which will contain only 300 Imperial gallons. Each compartment is provided with a quick-opening foot valve (G1) actuated by a handwheel accessible from the catwalk on top of the tank. Also, each compartment's outlet pipe terminates in a spring loaded, normally closed, quick acting valve (G2) operated by a special key which can retain the valve in the open position.

The first three compartments have been manifolded with a Worcester valve (G3) on the outlet. A flexible hose connects this to the rigid $\frac{3}{4}$" B.S. black pipe located in a shallow trench across the hardstanding to the electrically driven gear pump (P), whose starting gear is mounted on the adjacent wall of Building 3F. From the pump, pipework is laid beside the kerb to the portable trailer unit (P.T.U.). Also a tee branch and valve (G4) provides a purge to inside the boiler house. Thus, gas oil from the road tanker can be pumped to either the P.T.U. or to the purge points on the bitumen lines feeding the gasifier.

The riser pipe on the P.T.U. has two valved flexible branches. Valve (G5), when open will fill the 50 gallon gas oil service tank for supplying the P.T.U. burners. Valves (G6 & 7) will provide purge gas oil to the ring main and the standby pump manifolding.

Fines Return System see Fig 13

THE STONE HANDLING SYSTEM

Limestone is fed to the gasifier through a gravity feed line from a pressurised weight hopper. A vibrator is used to control the rate of stone addition. A ground level hopper is charged from bags of stone. A pneumatic transfer line moves the stone from the ground level hopper to an upper hopper from which stone is periodically dropped into the weigh hopper. The upper hopper acts an an air lock to avoid depressurising the weigh hopper.

N_2 SYSTEM TO TRANSFER PULSERS

A schematic diagram is given in figure 14.

MAIN BURNER DESCRIPTION

The standard oil burner of the 10 million Btu/hr boiler was replaced with an experimental burner designed to handle the hot gasifier product.

A general arrangement of the gasified fuel burner is shown in figure 15. It will be seen that the air required for complete combustion is introduced in 3 stages. Roughly about 10% of the air enters a crude injector (1) and is pre-mixed with the gas. Further premix may be introduced at point (2) and the remainder of the air is fed tangentially into a swirl chamber and emerges from an annular nozzle which is concentric with the gas nozzle. The design of this burner was purely empirical and was based on recommended pressure drops at the nozzle. The main gas duct is sized to give a flow velocity of about 60ft/second, the gas nozzle gives a pressure drop in the region of 3" w.g., and the air nozzle gives a pressure drop of about 4-5"w.g.

A pilot light fired by propane gas is used for lighting purposes.

ANALYTICAL SAMPLING AND INSTRUMENTS

On stream analysers monitor the compositions of the key gas streams. Table I lists these analysers and their applications.

Figure 16 gives a flow diagram of the gas analysing equipment whereas figure 17 shows the detailed sampling of the boiler flue gas.

TABLE I

CAFB Pilot Plant Gas Analysers

Gas Stream	Component	Analyser	Operating Principle	Range
Air-Flue Gas Mix to	O_2	Servomex OP 250	Paramagnetic	0-25% by vol
Gasifier Plenum	CO_2	Maihak Unor 6	Infra Red	0-10% by vol
Boiler Flue Gas Sampled at fire	O_2	Servomex OA 137	Paramagnetic	0-5% by vol
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	Wösthoff	Electrical conductivity of H_2O_2 - SO_2 reactor products in solution	0-1000 ppm
	SO_2	Hartman & Brown	Infra Red	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

BOILER AND PRESSURIZATION SYSTEM

A handbook containing details of the boiler and pressurization unit will be kept in the crew room. If problems occur with this equipment which is not immediately apparent call for help.

SAFETY ALARM SYSTEMS

1. Types of Action

The installation is protected by a number of alarm circuits, in some cases the consequence of an alarm is automatic where immediate safety is concerned, but many others are warnings of some changing condition where there is time for corrective action.

Action A.

This action is an automatic plant shut down and produces the following actions:-

- Fire valves close on oil feed and return lines at entrance to building.
- Oil circulation pump stops.
- Gasifier control panel shuts down everything apart from main air blower on the boiler, primary and secondary cooling circuits of the boiler cooling system blowers on 3rd stage regenerator boost and pressure control (regen off gas).
- Interior and exterior bells ring at 3F, which may be silenced by a mute button on the auxiliary panel in the air lock passageway.
- Red light shows on the auxiliary panel and also on the gasifier control panel warning light for the auxiliary panel.

Action B.

- Alarm light shows on gasifier control panel and rings a bell on the panel which may be muted for that particular alarm by a switch located above that alarm light. Automatic gasifier shut down is not possible with action B.

Action C.

- Alarm light shows on gasifier control panel which can be linked to a gasifier shut down by selecting the switch on the panel to "Automatic Shut Down Mode". This alarm will ring a bell unless muted.

Action D.

- Alarm light shows on gasifier control panel - cannot ring a bell or cause an automatic gasifier shut down.

Action E.

- Alarm light on auxiliary panel and light on control panel.
- No other action.

Action F.

- Horn type warning sounded in 3F, 3A and the grinding room.

Action G.

- Low pressure in the pressurisation unit causes a unique operation - namely - a bell on pressurisation unit, light on gasifier control panel, auxiliary panel warning light, and automatic shut down of all equipment controlled from main panel but not air to burner, cooling pumps or shut down of fire valves or oil circulation pump.

2. Alarm Sources

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.

2.1 The Boiler and its Cooling System

The water in the boiler is pressurised to about 48 psi and is pumped through a heat exchanger. The secondary side of the heat exchanger is cooled by an evaporative cooler on the roof of the building.

Primary Circuit Protection

- The pressurisation unit has a low pressure warning set at 40 psi.

Action G

- High water temperature in the boiler water - set at 245°F.

Action B

Secondary Circuit Protection

- Lack of cooling water flow is detected by a differential pressure switch across feed and return lines to cooler. This switch is alarmed if the pump is switched on and the differential pressure is less than 3 psi approx.

Action A

- High water temperature to cooler - a mechanical reset overtemperature alarm set in the vertical leg from the heat exchanger - operates at 200°F. The reset button is 10 ft up the vertical leg.

Action A

2.2 The Gasifier and Regenerator

The gasifier has a variety of temperature alarms and high or low pressure alarms.

- High temperature in the gasifier bed - usually set to 950°C - set, shown and alarmed from Guardian controller.

Action B

- o High temperature in the regenerator bed - usually set to 1100°C - set, shown and alarmed on Leeds and Northups recorder.

Action B

- o Gasifier distributor low pressure drop - set to 5" w.g. by pressure switch - letter E. (Under control panel.)

Action D

- o Regenerator distributor high pressure drop - set to 10" w.g. by pressure switch - letter D. (Under control panel.)

Action D

- o Regenerator low bed level - set to 10" w.g. by pressure switch - letter F. (Under control panel.)

Action D

- o Regenerator high bed level - set to 30" w.g. by pressure switch - letter C.

Action D

- o Pressure rise in gasifier gas space - set to 24" by pressure switch G.

Action B

2.3 Main Burner and Pilot Burner

Main Burner

The main burner flame can be scanned by 2 detectors, one at each end of the boiler and failure of both will cause alarm C.

Pilot Burner

The pilot burner will only light up if there is gas pressure to the pilot and air pressure to the experimental burner plenum.

If the fireye which scans the pilot does not see sufficient flame it will cause the pilot to lock out and show as an alarm light. Failure of gas pressure or plenum pressure will have the same result. It is not possible to check action of fireye without attempting start up.

Action B

2.4 General Alarms

- o Bitumen trailer pump failure - Action B
- o N_2 supply failure - Action B
- o Fire detector - a fusible link above the boiler set to melt at 155°F will cause

Action A

This link needs to be replaced after operation and because it may not be convenient to isolate the power at that moment a bypass electrical switch has been fitted on the auxiliary panel as a temporary procedure until the link can be replaced. Replacement must be done as soon as possible.

- <u>Sump level</u>

  If the level in the sump rises to about 1/8" of floor level a light will show on the auxiliary control panel, the main panel will not be alarmed and no bells will ring.

  <u>Action E</u>

- <u>Emergency Stop Buttons</u>

  There are four emergency stop buttons located:

  - Close to the sliding window in cubicle
  - Main door at 3A end of laboratory
  - Main door at main stores end of laboratory
  - Adjacent to ladder on side of the pit

  <u>Action A</u>

  These must be reset by turning knob and allowing knob to spring back.

- <u>Emergency Stop Button On Main Control Panel</u>

  Located in centre of gasifier control panel and shuts down all items controlled from this panel but does not shut down fire valves, cooler pumps, air to burner, or oil circulating pump regenerator boost or pressure control blowers. Reset by turning ring and allowing button to return.

  The fire valves and oil circulating pump can be shut down by then depressing the emergency stop button by the sliding window.

  i.e. <u>Action A</u>

- <u>Call for Assistance</u>

  There are four buttons located on each wall of the pit with a further button on the cubicle wall at the top of the pit steps. These sound horns in 3A, 3F and the grinding room.

- <u>Fuel Shut Off Valves</u>

  In the event of an emergency shut down where there is any possibility of fire the propane gas and kerosene must be isolated at their external valves. The propane valve is situated at the external corner of the building adjacent to the supply feeder from the propane line. The kerosene valve is located by the barrel stand adjacent to the semi buried storage tank.

Summary of Pilot Plant Alarm System

	Source of Alarm	Indication: Auxiliary Panel	Indication: Main Control Panel
1.	Failure of water circulating pump or lack of water in secondary cooling circuit	Red Light and bells	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 high temperature"
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 light and bell
6.	Regenerator high temperature	None	Regenerator high temperature warning light and bell
7.	Gasifier low pressure across distributor	None	Gasifier low pressure distributor warning light
8.	Regenerator high pressure across distributor	None	Regenerator blocked distributor warning light
9.	Regenerator low bed level	None	Regenerator low bed level warning light
10.	Regenerator high bed level	None	Regenerator high bed level warning light
11.	Pressure rise in gas space	None	Downstream pressure rise warning light and bell
12.	Experimental Burner & Pilot Failure of: Main Flame	None	Main flame failure warning light and bell
	Pilot Flame	None	Pilot flame warning light & bell
13.	Bitumen Trailer pump failure	None	Warning light and bell
14.	N_2 supply failure	None	Warning light and bell
15.	Fire detector	Red light and bells	Red light titled "Auxiliary Panel"
16.	Sump level	Red light	Nothing

Summary of Pilot Plant Alarm System (cont'd)

Source of Alarm	Indication: Auxiliary Panel	Indication: Main Control Panel
17. Emergency stop buttons in building	Red light and bells	Red light titled "Auxiliary Panel"
18. Emergency stop button	None	None

LOCATION OF ELECTRICAL COMPONENTS AND FUSES. (see Fig 18)

TABLE I

FUSE BOX	MAP REFERENCE	LOCATION
A	CONTROL ROOM	TO THE RIGHT OF PANEL (WHEN FACING PANEL)
B	CONTROL ROOM	BEHIND PANEL AT THE RIGHT (WHEN FACING PANEL)
C	C/4	HIGH ON WALL IN NORTH EAST CORNER
D	C/4	ABOVE FUSE BOX C IN NORTH EAST CORNER
E	C/3	TO RIGHT OF GASIFIER UNIT (NORTH SIDE), NEAR BACK OF BOILER

Fuse Labelling Convention

Typical Label - A/1/ABC

(a) First letter gives fusebox number

(b) Second letter gives fuse vertical column number (from the left)

(c) Third letter (or set of letters) gives the phase:

A refers to top horizontal row,

B the next one down and so on.

Table 2

FUNCTION	LOCATION	ISOLATOR LOCATION	FUSE LOCATION
BLOWERS:			
First stage main air and flue gas blower to Gasifier	In outside shed	To right of panel in Control Room	A/3/ABC
Second stage main air and flue gas blower to Gasifier	In outside shed	To the right of panel (when facing panel) in control room	A/1/ABC
Air blower to regenerator (first stage)	A/2 To left of boiler, beside boiler	A/2 By the blower	E/3/ABC
Air blower to Regenerator (second stage)	A/2 To left hand side of boiler, beside boiler	To right of Panel in Control Room	A/6/ABC
Main air blower to Regenerator (third stage)	A/2 To left hand side of boiler, beside boiler	To right of panel in control room	A/2/ABC
Main air blower to main boiler flame	A/2 To left of boiler, near boiler front	A/2 Above the blower	E/2/ABC
Flue gas recycle blower	B/2 To right of boiler and above it	C/2 To right of boiler, midway along boiler side	D/1/ABC
Blower for back pressure on Gasifier to regenerator balance valve	C/3 To right of Gasifier unit, nearer wall	C/3 By the blower	C/1/ABC

Table 2 (Continued)

FUNCTION	LOCATION	ISOLATOR LOCATION	FUSE LOCATION
COMPRESSORS:			
Single Compton transfer compressor Regenerator to Gasifier	C/2 To right hand side of boiler	To right of panel in Control Room	A/7/ABC
Double Compton transfer compressor Gasifier to Regen- ator	C/2 To right hand side of boiler, beside boiler	To right of panel in Control room	A/4/ABC
PUMPS:			
Fuel pump No. 1 (Left)	In Control Room to left (when facing panel)	On right side of panel in Control room	B/1/ABC
Fuel Pump No. 2 (Centre)	In Control Room to left (when facing panel)	To right of panel in Control Room	A/5/ABC
Fuel pump No. 3 (Right)	In Control Room to left (when facing panel)	On right side of panel in Control room	B/2/ABC
Oil Circulating Pump	C/2 To right of boiler, near wall	C/2 To right of pump	C/6/ABC
Secondary oil take-off supply pump	B/2 To left of boiler, under boiler	C/3 At the top of fuse box E, to the right of Gasifier	E/1/ABC
Compton gas sampling pump	At back of Control Room near office door	Above the pump	C/2/ABC

Table 2 (Continued)

FUNCTION	LOCATION	ISOLATOR LOCATION	FUSE LOCATION
Primary boiler water Pump	A/2 To left of boiler, near boiler front	A/2 By pump on the wall	C/4/ABC
Secondary water cooling pump for for boiler	A/2 To left of boiler, near boiler front	A/2 By pump on the wall	C/3/ABC
Water cooling pump	A/4 in Gasifier pit by ladder	A/4 In Gasifier pit by ladder	D/4/C
Stack water washer pump	C/1 Outside north west corner	C/2 To right of boiler, midway along boiler and on wall	D/2/ABC
Submersible pump for removal of water in Gasifier	C/3 To right of Gasifier pit in corner	None	C/12/C
HEATERS:			
Fuel outflow immersion heater tank No. 1 (right hand side when looking at it)	Fuel storage outside	None (Need to take fuse out)	C/11/ABC
Fuel outflow immersion heater tank No. 2 (Centre)	Fuel storage outside	None (Need to take fuse out)	C/10/ABC
Fuel outflow immersion heater tank No. 3 (left looking at it)	Fuel storage outside	None (need to take fuse out)	C/9/ABC

Table 2 (Continued)

FUNCTION	LOCATION	ISOLATOR LOCATION	FUSE LOCATION
Oil immersion heater for supply oil	A/2 To left of boiler, near boiler back	C/3 At the top of fuse box E, to right of Gasifier	E/4/AB
Outside trace heating for Oil	On fuel supply pipes	None	C/13/C
Inside trace heating for Oil	From oil primary circuit to metering pumps	In Control room to the left (when facing panel)	C/13/C
Cooling Tower immersion heater	A/2 On roof	None	C/14/C
FANS:			
Cooling Tower Fan	On roof	A/1 To left of boiler, near boiler front	C/5/ABC
Cubicle purge fan for warm air in control room	In roof of control room	In passage to control room	C/16/A
Two speed exhaust fan	B/3 In roof	In passage to Control room	C/8/ABC
Extract Fan	B/3 In roof	In passage to control room	C/8/ABC
Dust extract fan on west wall	A/1 On west wall in south west corner	A/1 On west wall in south west corner	C/18/ABC

Table 2 (Continued)

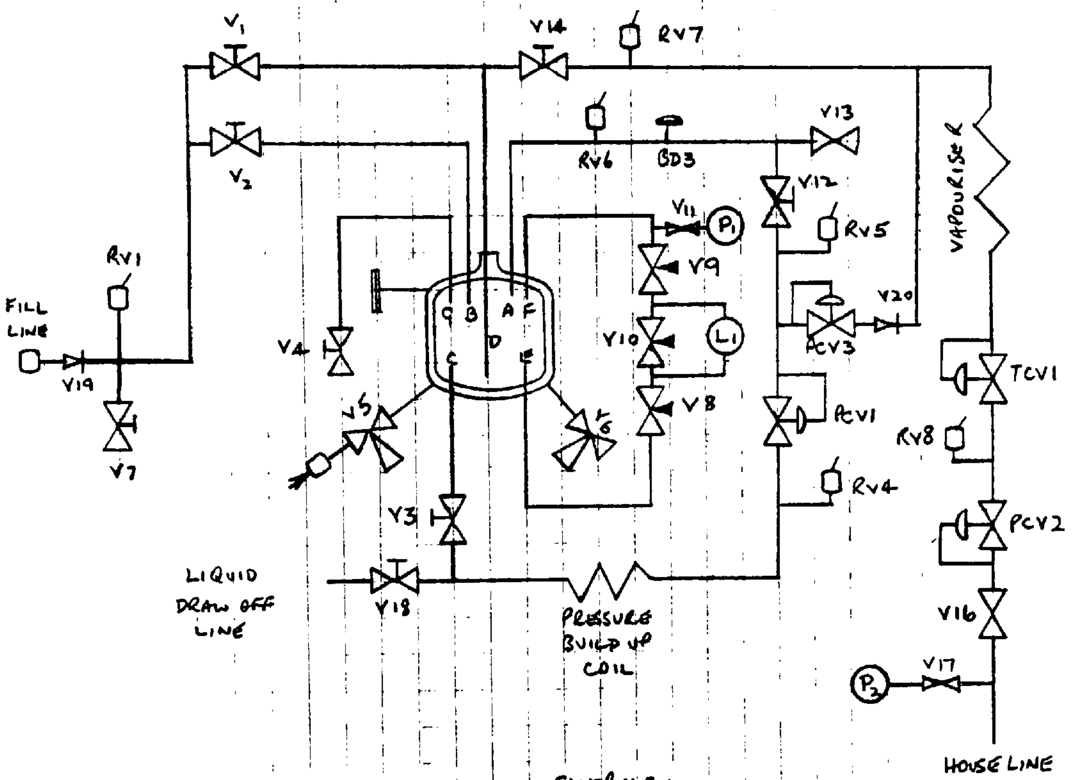
FUNCTION	LOCATION	ISOLATOR LOCATION	FUSE LOCATION
LIGHTS, etc:			
Relay Box and emergency horns	In passage to control room	None	C/13/A
Spur Box and 4 x 13 amp Sockets in Gasifier pit	B/3 In Gasifier pit	None	C/16/B
Spur for transformer for 12V lamp in Gasifier pit	A/3 In Gasifier pit	Switch by the transformer A/3	C/16/B
Large flood light outside office door	Outside office door	Outside by door to office	C/16/C
Outside flood light on stack	On chimney stack outside	A/1 On wall in south west corner	D/3/B
Light at bottom of stack	On chimney stack outside	C/1 On wall in north west corner	D/3/C
Fluorescent lights in pit	B/3 In Gasifier pit	A/4 by ladder in Gasifier pit	D/3/C
Outside floodlight on west wall	On outside west wall	A/1 on wall in south west corner	D/3/B

Table 2 (Continued)

FUNCTION	LOCATION	ISOLATOR LOCATION	FUSE LOCATION
OTHERS:			
Boiler Pressurisation Unit	A/2 To left of boiler, near Control room	A/2 Inside the pressurisation unit	C/7/ABC
Panel Supply	In Control Room	To right of Panel in Control room	A/8/A
"Fire Eye" for propane warm up burner to Gasifier	A/3 To left of Gasifier (pit grid level)	In Control room to left (when facing panel)	A/8/B
"Fire Eye" for propane pilot to main burner	B/3 At back of boiler near main flame	In Control room to left (when facing panel)	A/8/B
Motorised valve for main air supply to Gasifier	In outside shed	In control room to left (when facing panel)	A/8/B
Motorised valve for flue gas supply to Gasifier	In outside shed	In control room to left (when facing panel)	A/8/B
Stone feed system	C/3 By crane upright	C/3 By crane upright	C/17/ABC
Crane	C/3 In north east corner by the Gasifier unit	C/3 On crane upright	C/17/ABC
NOT IN USE	None	None	(A/8/C (B/3/ABC (
NOT IN USE	None	None	D/4/AB

20A

LIQUID NITROGEN TANK SYSTEM



LEGEND:

NOZZLES

- A VAPOUR FROM PRESSURE COIL
- B TOP FILL LINE
- C LIQUID WITHDRAWAL
- D DIP TUBE + BOTTOM FILL
- E LOWER LIQUID LEVEL
- F UPPER LIQUID LEVEL
- G FULL LEVEL LINE

SYMBOLS:

- EXTENDED SPINDLE VALVE
- NORMAL SPINDLE
- CHECK VALVE
- THERMOCOUPLE
- P PRESSURE
- L LEVEL
- CONTROL VALVE
- NEEDLE VALVE
- VACUUM VALVE
- RELIEF VALVE
- BURSTING DISC
- BLOW-OFF DISC

VALVES

- V1 BOTTOM FILL
- V2 TOP FILL
- V3 LIQUID SHUTOFF
- V4 FULL TRY COCK
- V5 VACUUM GAUGE SHUTOFF
- V6 VACUUM LINE SHUTOFF
- V7 FILL LINE DRAIN
- V8 LEVEL GAUGE LIQUID SHUT OFF
- V9 LEVEL GAUGE VAPOUR SHUT OFF
- V10 LEVEL GAUGE NORMALISING VALVE
- V11 PRESSURE SHUT OFF
- V12 VAPOUR SHUT OFF
- V13 VAPOUR VENT
- V14 LIQUID FEED TO HOUSE LINE
- V16 HOUSE LINE REGULATOR SHUTOFF
- V17 HOUSE LINE PRESSURE GAUGE SHUTOFF

VALVES (CONT'D)

V_{18} LIQUID DRAW OFF

CHECK VALVES

V_{19} FILL CONNECTION CHECK VALVE
V_{20} ECONOMISER CIRCUIT CHECK VALVE

GAUGES

P_1 TANK PRESSURE
P_2 HOUSE LINE PRESSURE

L_1 TANK LIQUID LEVEL

CONTROL VALVES

TCV1 HOUSE LINE TEMPERATURE CONTROL VALVE

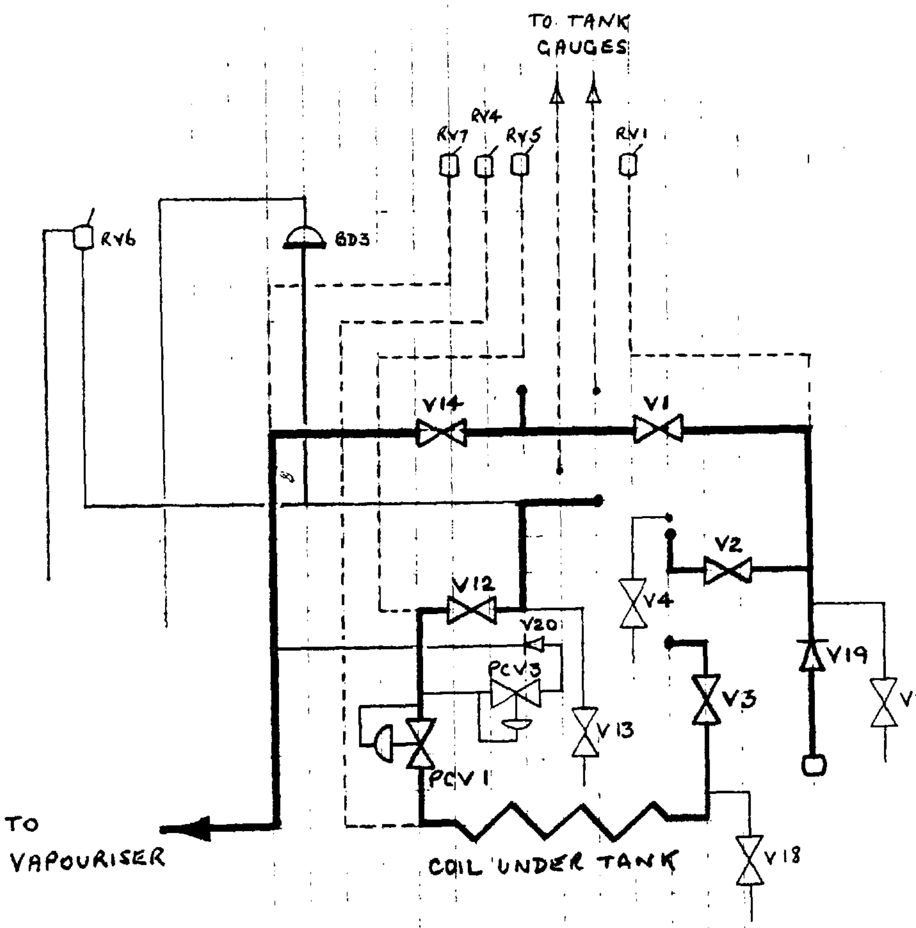
PCV_1 TANK PRESSURE BUILDUP REGULATOR
PCV_2 HOUSE LINE PRESSURE BUILDUP REGULATOR FROM VAPOURISER
PCV_3 ECONOMISER CIRCUIT PRESSURE REGULATOR FROM TANK VAPOUR SPACE

SAFETY VALVES

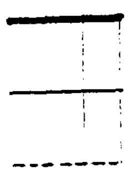
RV1 CHARGING LINE RELIEF VALVE
RV4 PRESSURE BUILDUP COIL RELIEF VALVE
RV5 " " LINE " "
RV6 INNER TANK RELIEF VALVE
RV7 VAPOURISER " "
RV8 TEMPERATURE CONTROL VALVE RELIEF VALVE

BD3 BURSTING DISC FOR TANK CONTENTS

LAYOUT OF LIQUID NITROGEN TANK LINES AND CONTROLS



LEGEND:-



MAJOR LINES IN 1" COPPER, LIKELY TO BE HEAVILY COATED WITH ICE.
OTHER LINES IN 1" AND ½" COPPER, NOT USUALLY COATED WITH ICE.
LINES IN ¼" COPPER TO MINOR RELIEF VALVES AND TANK CONTENTS GAUGES.

CAFB PILOT PLANT FLOW PLAN

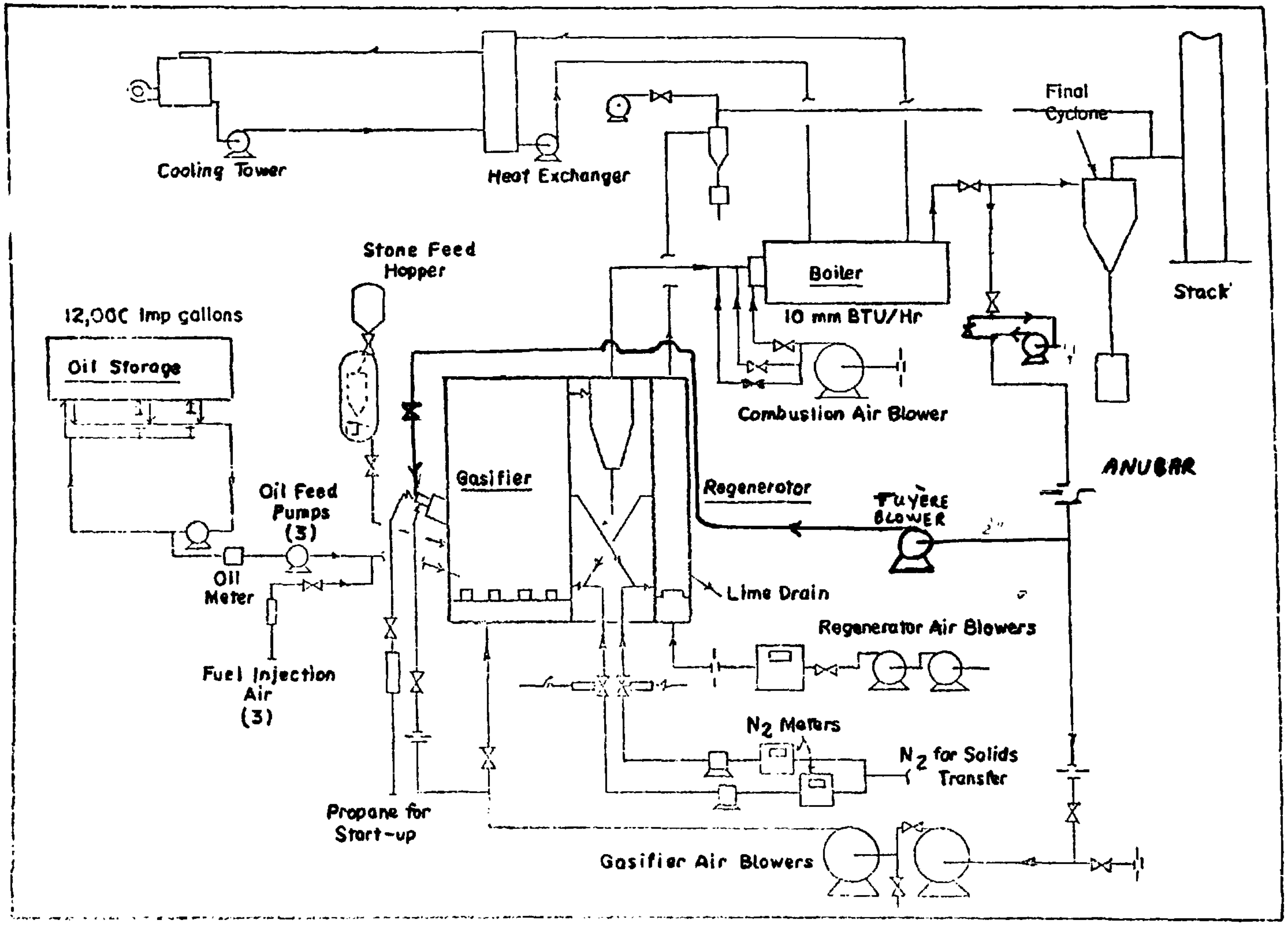
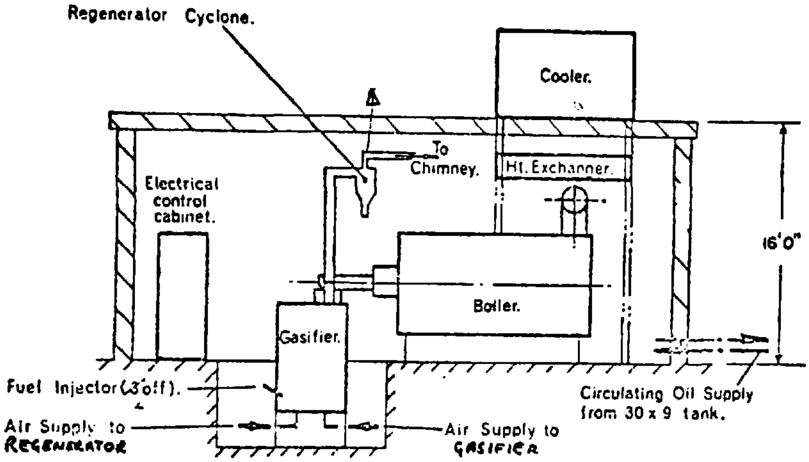
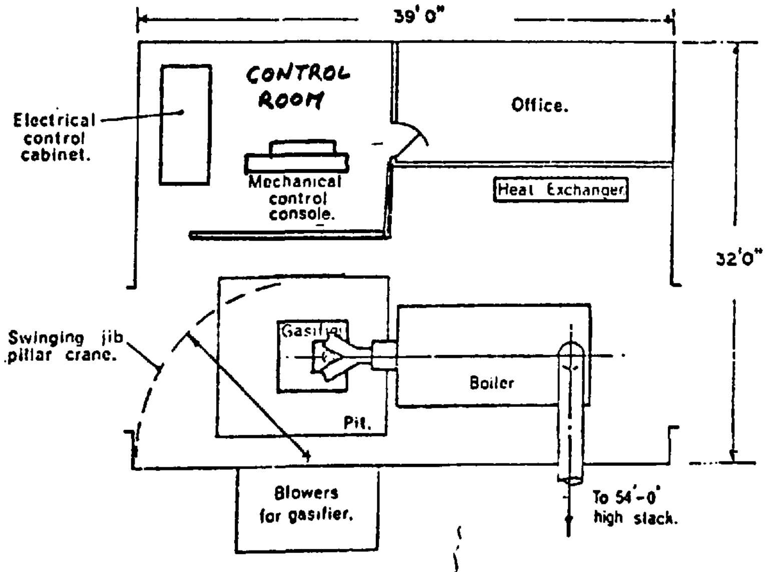


FIG. 1



SECTION.



PLAN.

GENERAL PLANT LAYOUT

FIG. 3

LAYOUT OF GASIFIER - REGENERATOR (MK II)

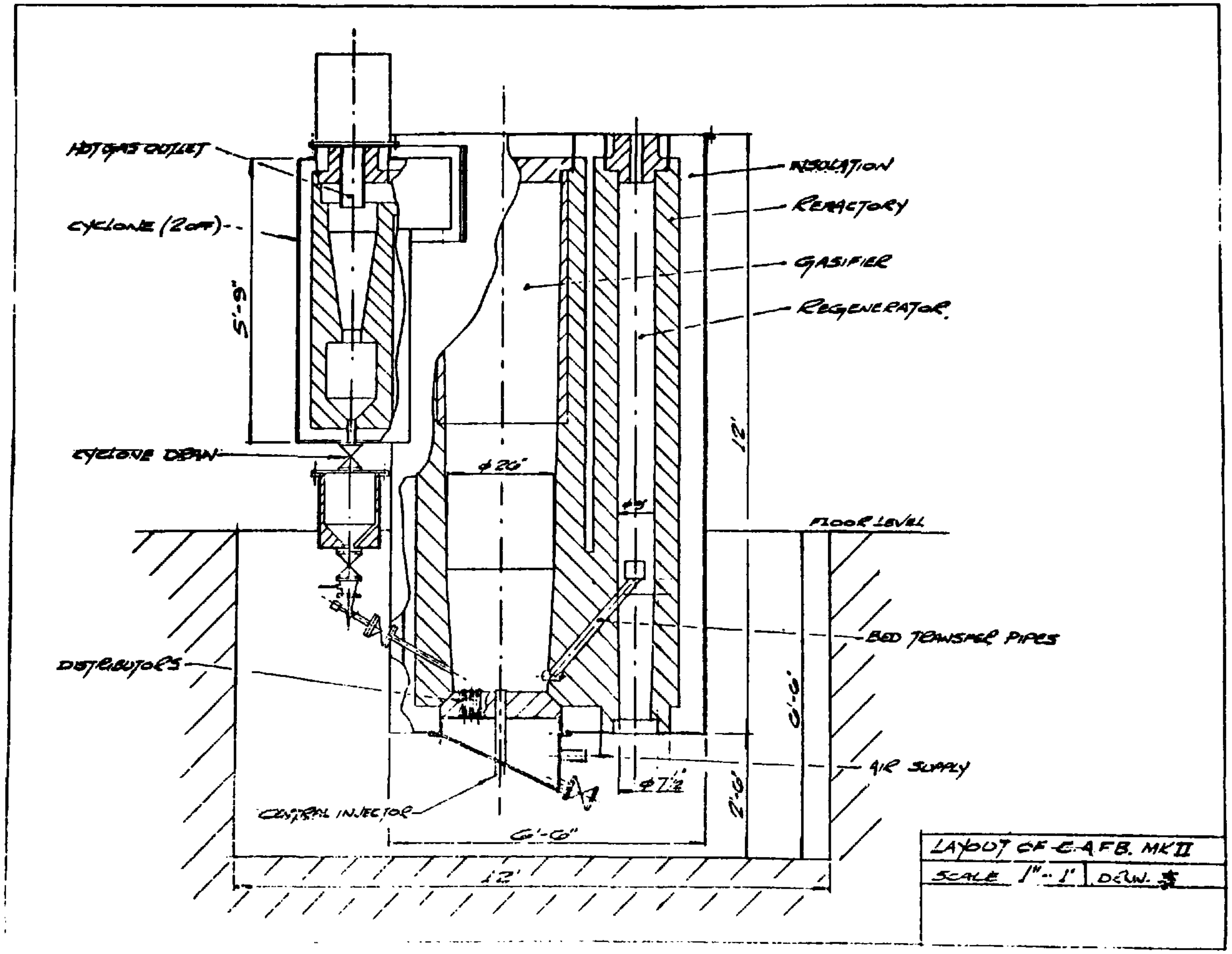


Fig 3

CAFB Pilot Plant Instrumentation Flow Plan

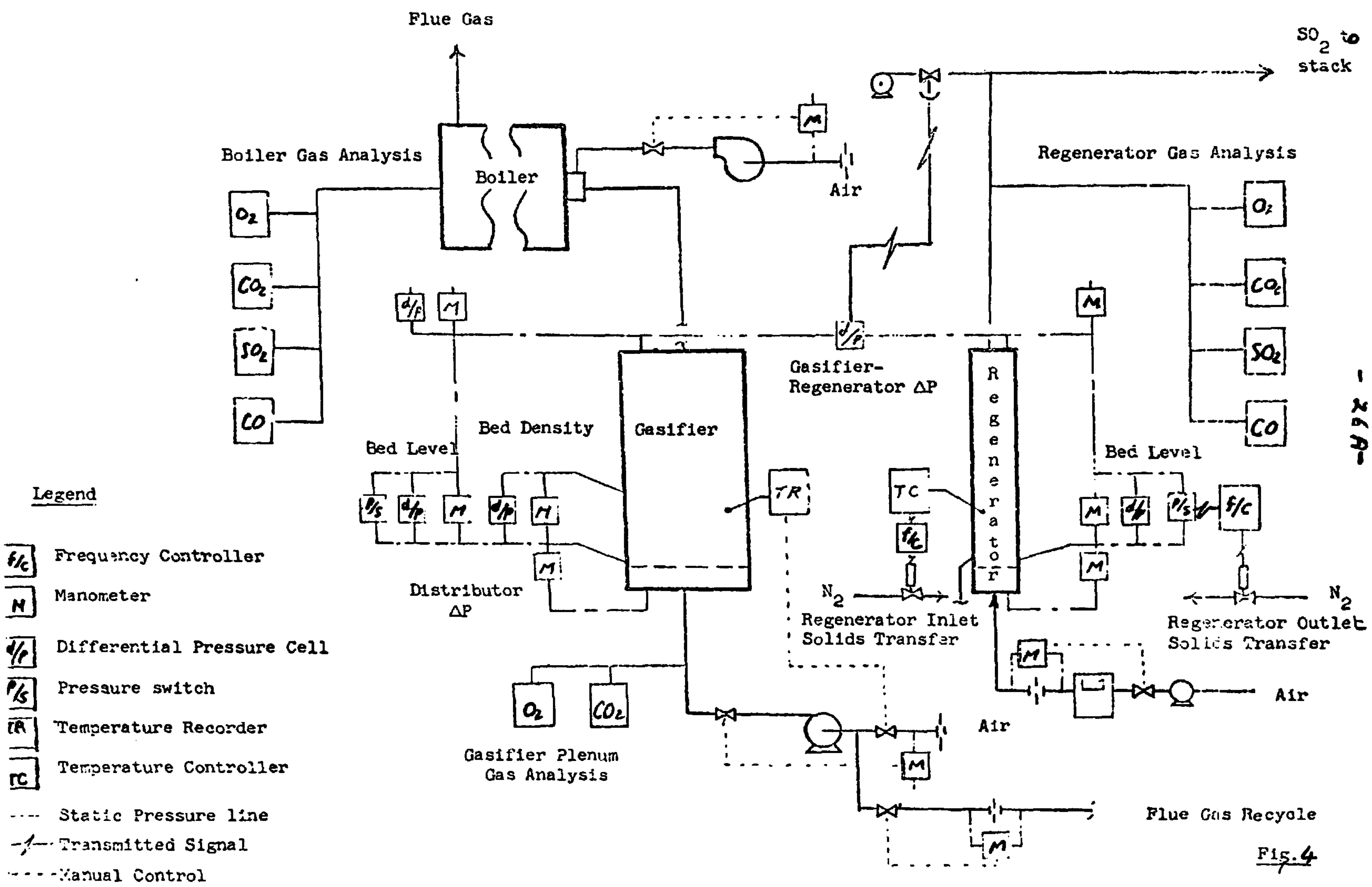
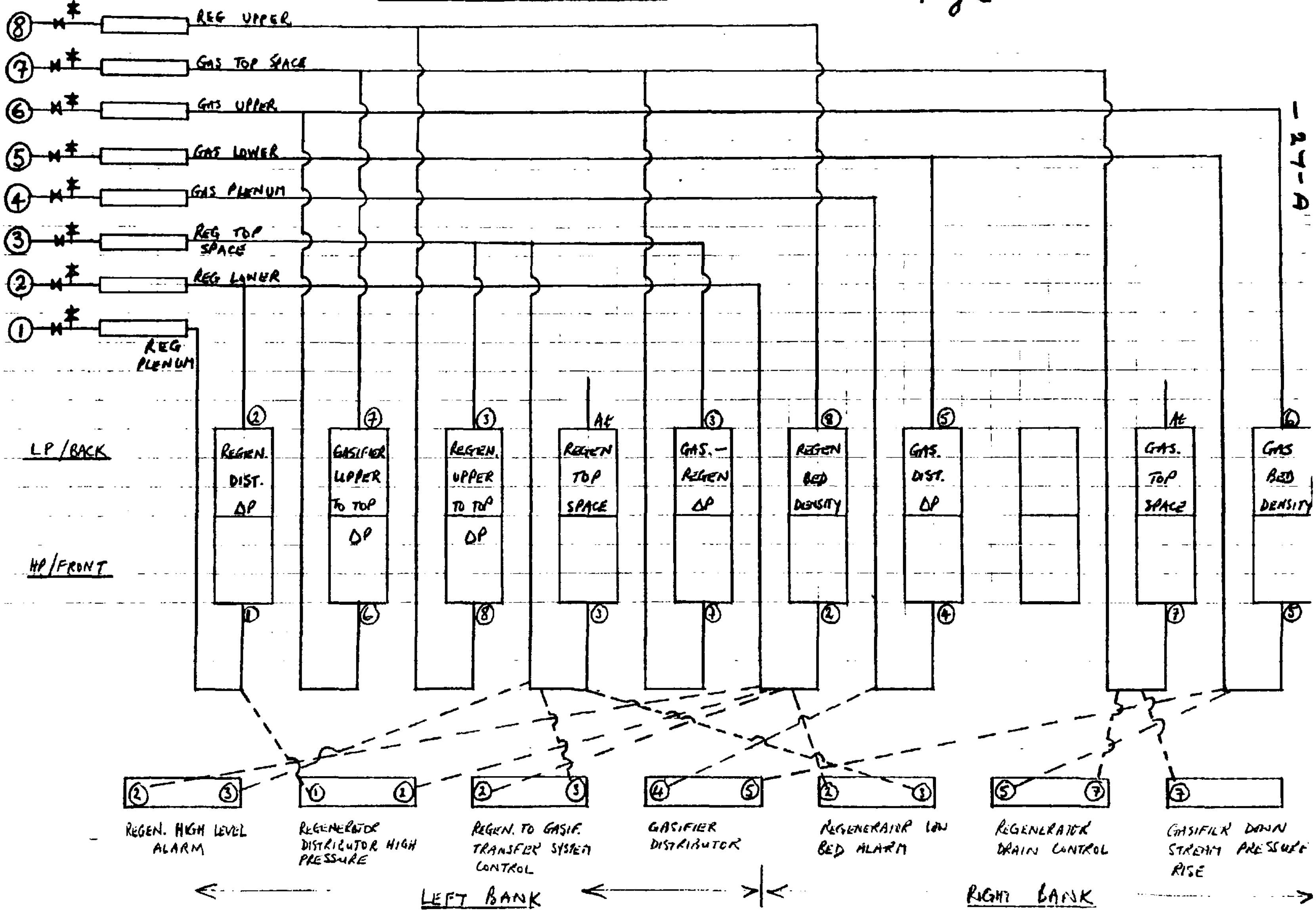


Fig. 4

MANOMETER CONNECTIONS.

Fig 5



1-27-D

PRESSURE CONNECTIONS

- 289 -

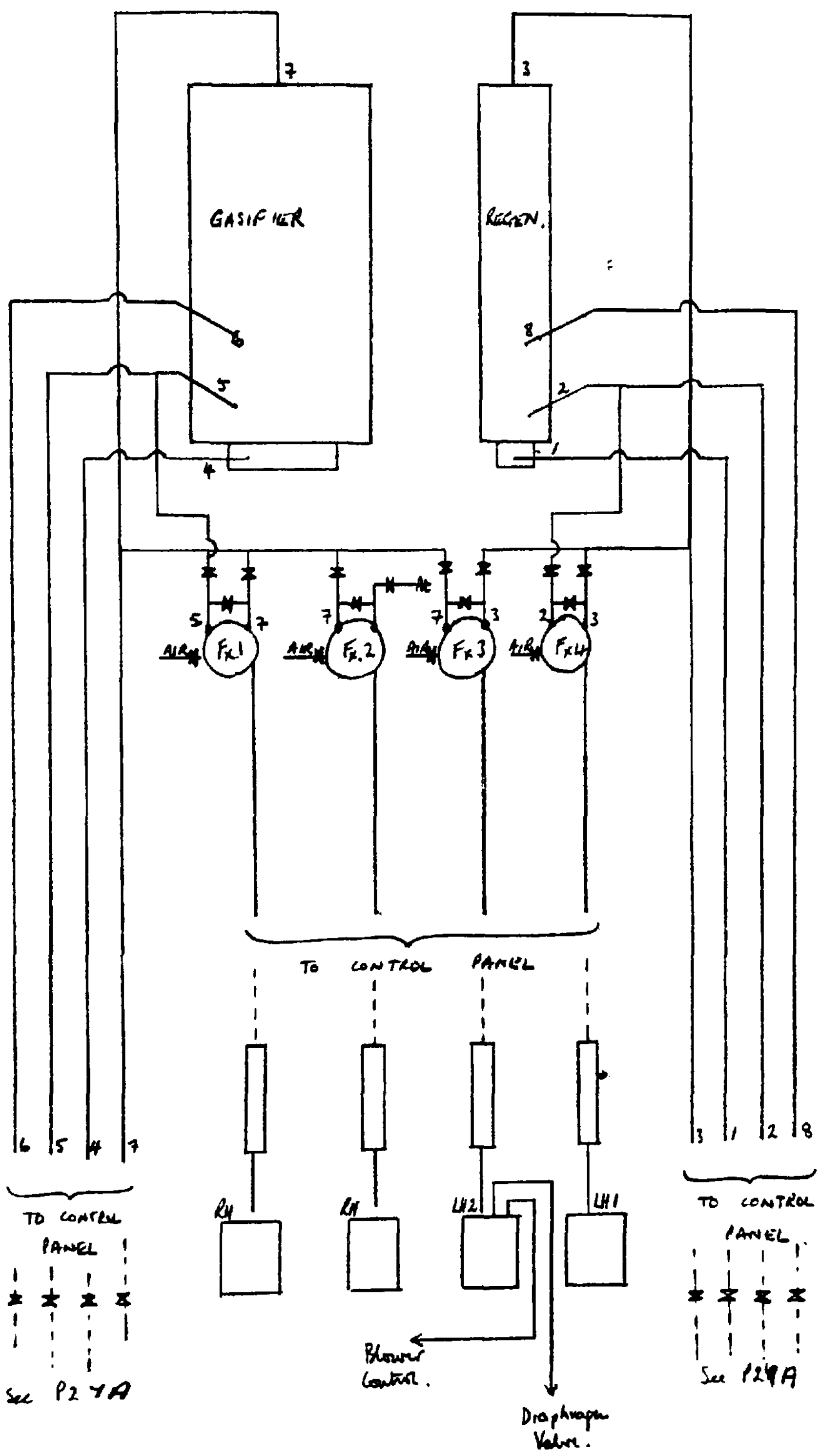


FIG 6

FUEL OIL SYSTEM

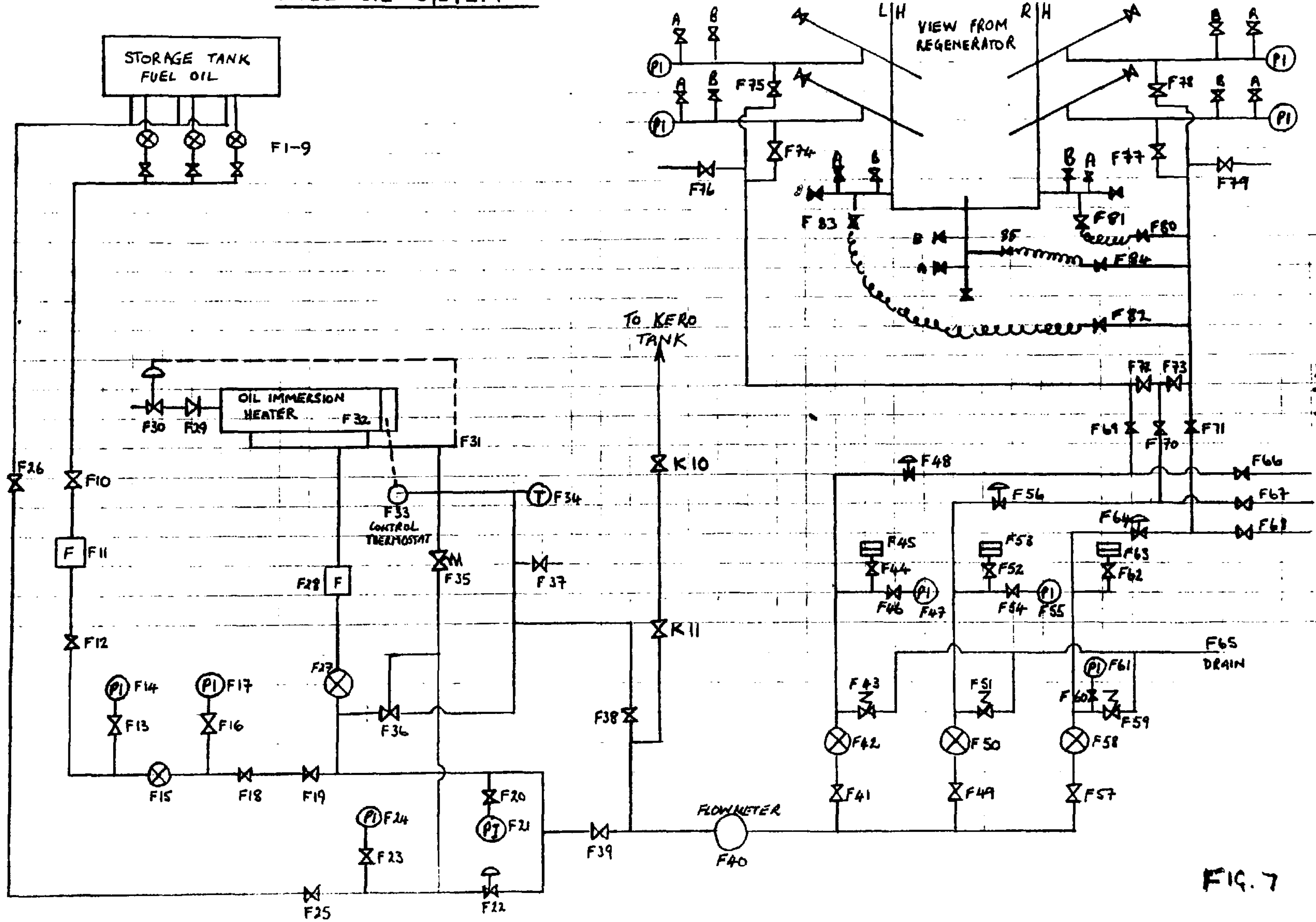
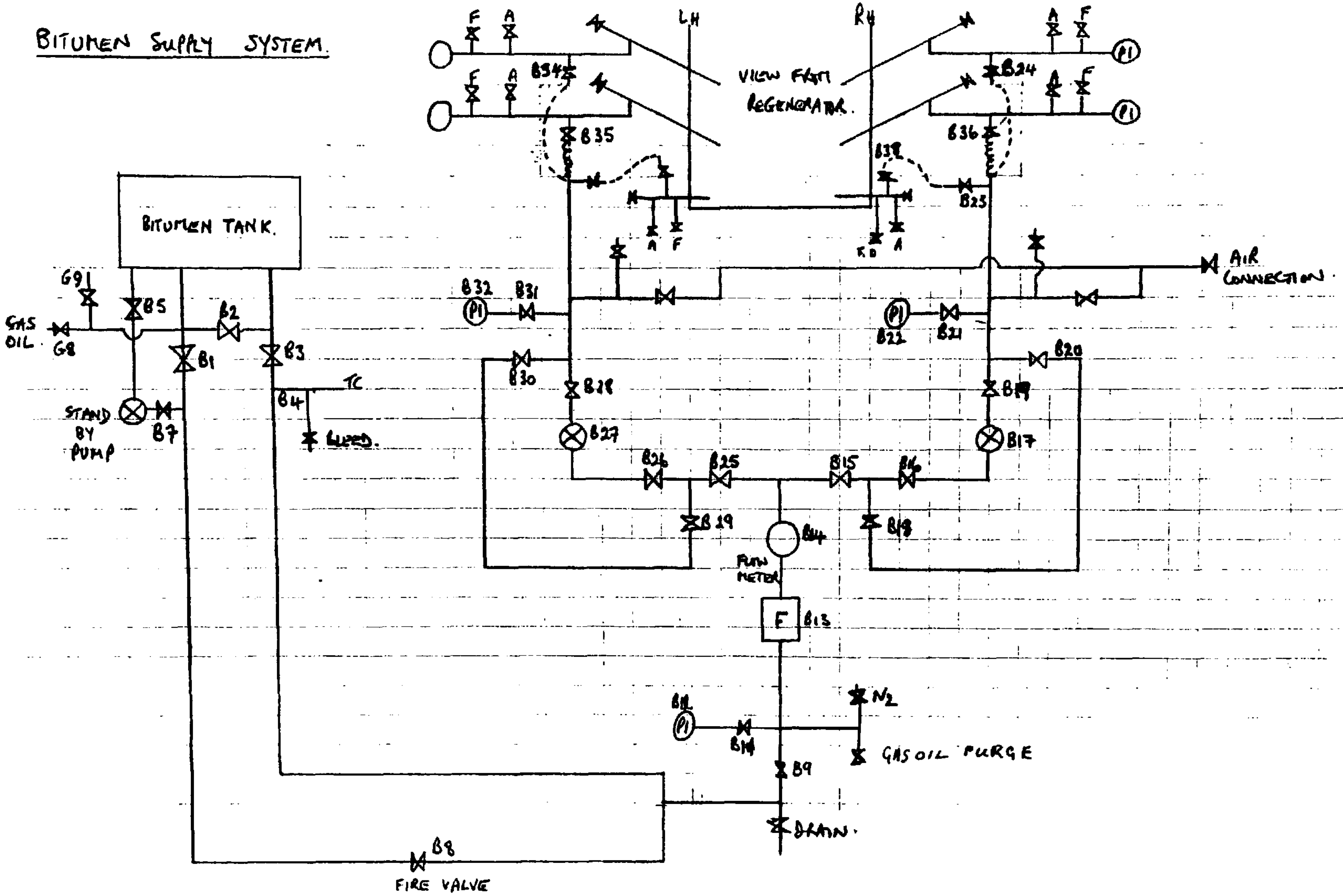


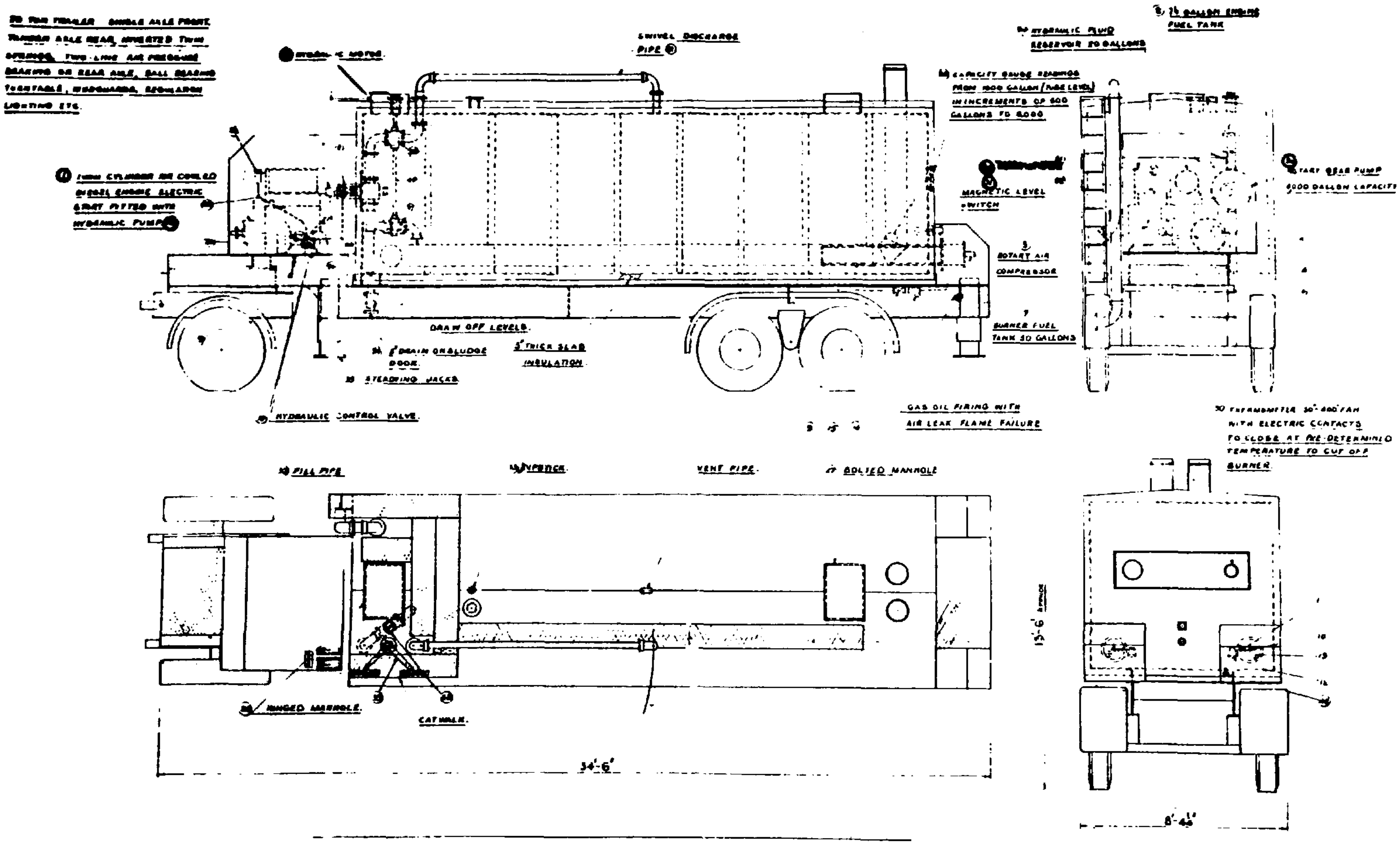
FIG. 7

BITUMEN SUPPLY SYSTEM.



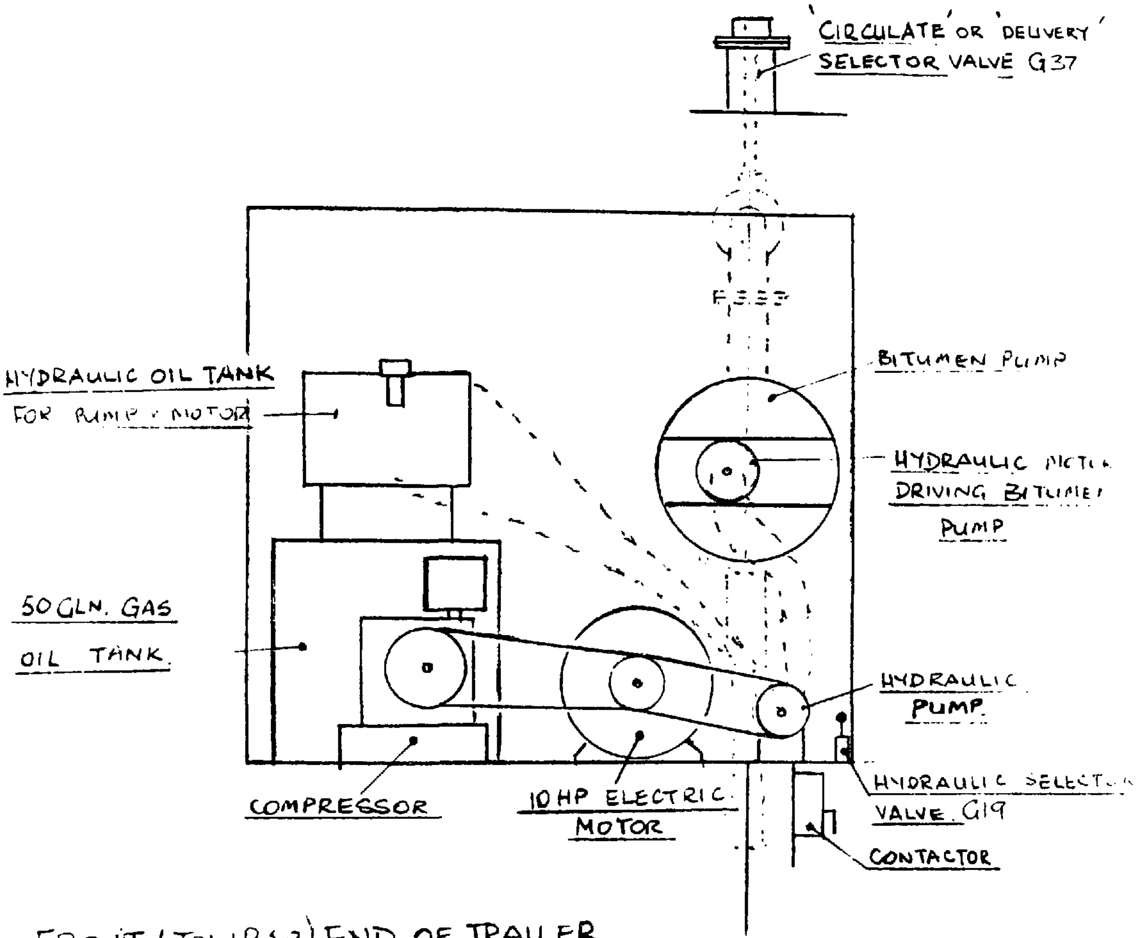
NOTE. ---- TEMPORARY CONNECTIONS (FLEXIBLE) IF REQUIRED

FIG 8

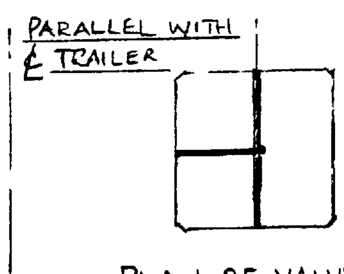


ARRANGEMENT OF TRAILER MOUNTED 6000 GALLON OIL FIRED 'MUNISTATIC' TANK.

7169



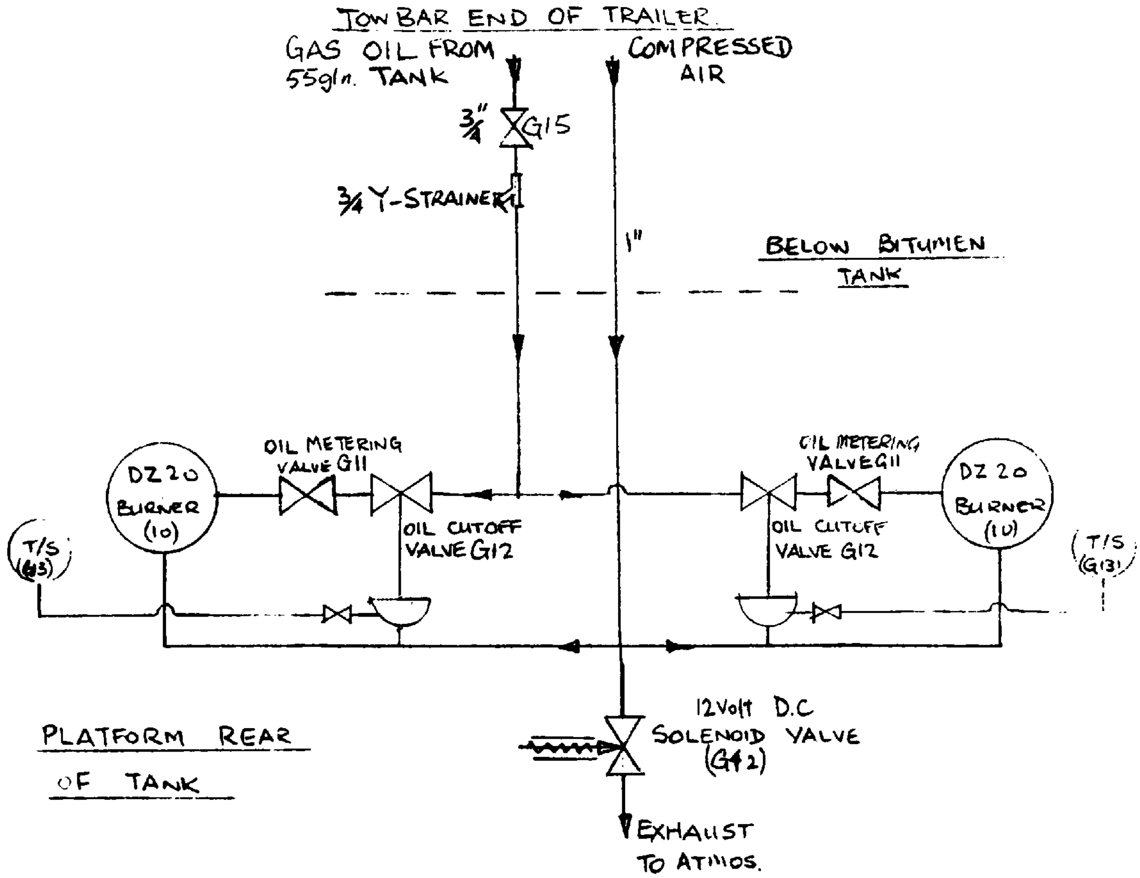
FRONT (TOWBAR) END OF TRAILER
SHOWING DRIVE MOTOR & ANCILLARIES.



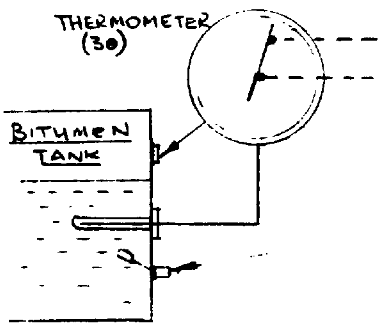
PLAN OF VALVE SPINDLE
IN 'DELIVERY' POSITION

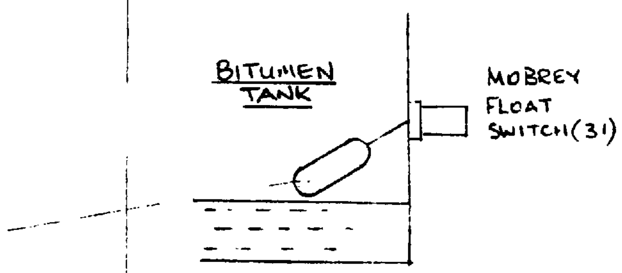
F 10

GAS OIL & COMPRESSED AIR SUPPLIES TO BITUMEN HEATING



SOLENOID VALVE ENERGISED BY:



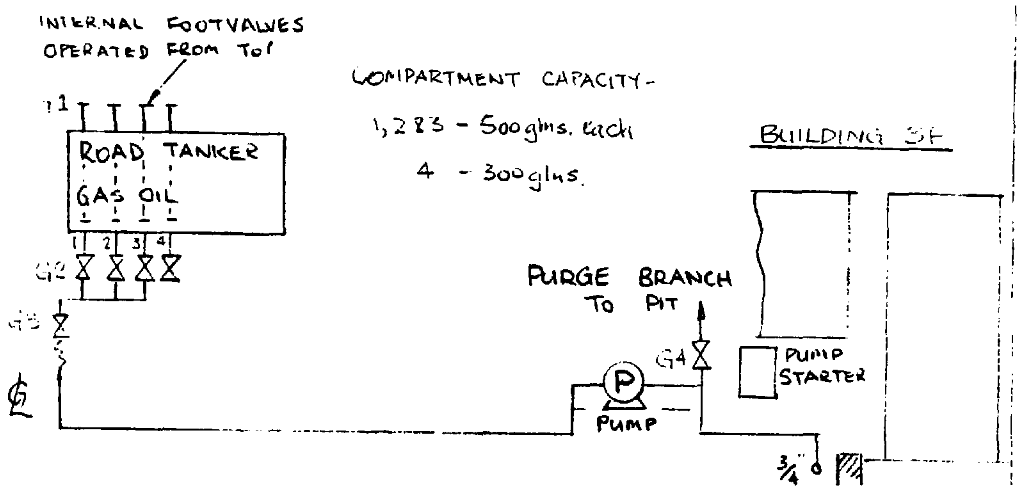


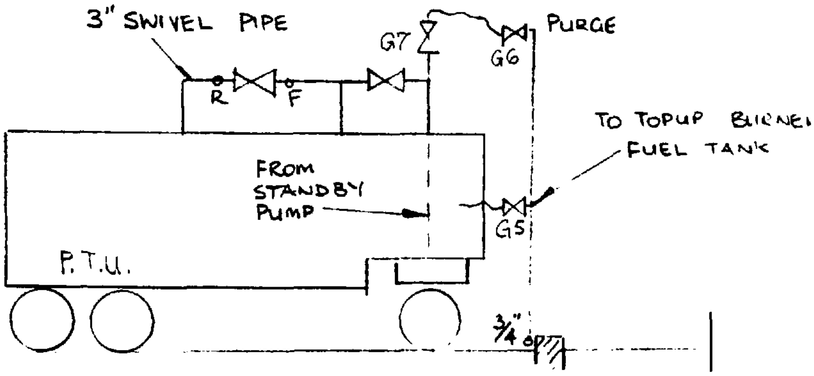
1. HIGH BITUMEN TEMPERATURE CLOSING THERMOMETER CONTACTS

2. LOW BITUMEN LEVEL OPERATING MOBREY FLOAT SWITCH CONTACTS

1.& 2 WILL OPEN SOLENOID VALVE THUS EXHAUSTING AIR TO ATMOSPHERE, CLOSING VALVE G12 (ie NO FUEL- NO AIR SITUATION)
TWO CONDITIONS FOR BURNER FAILURE.

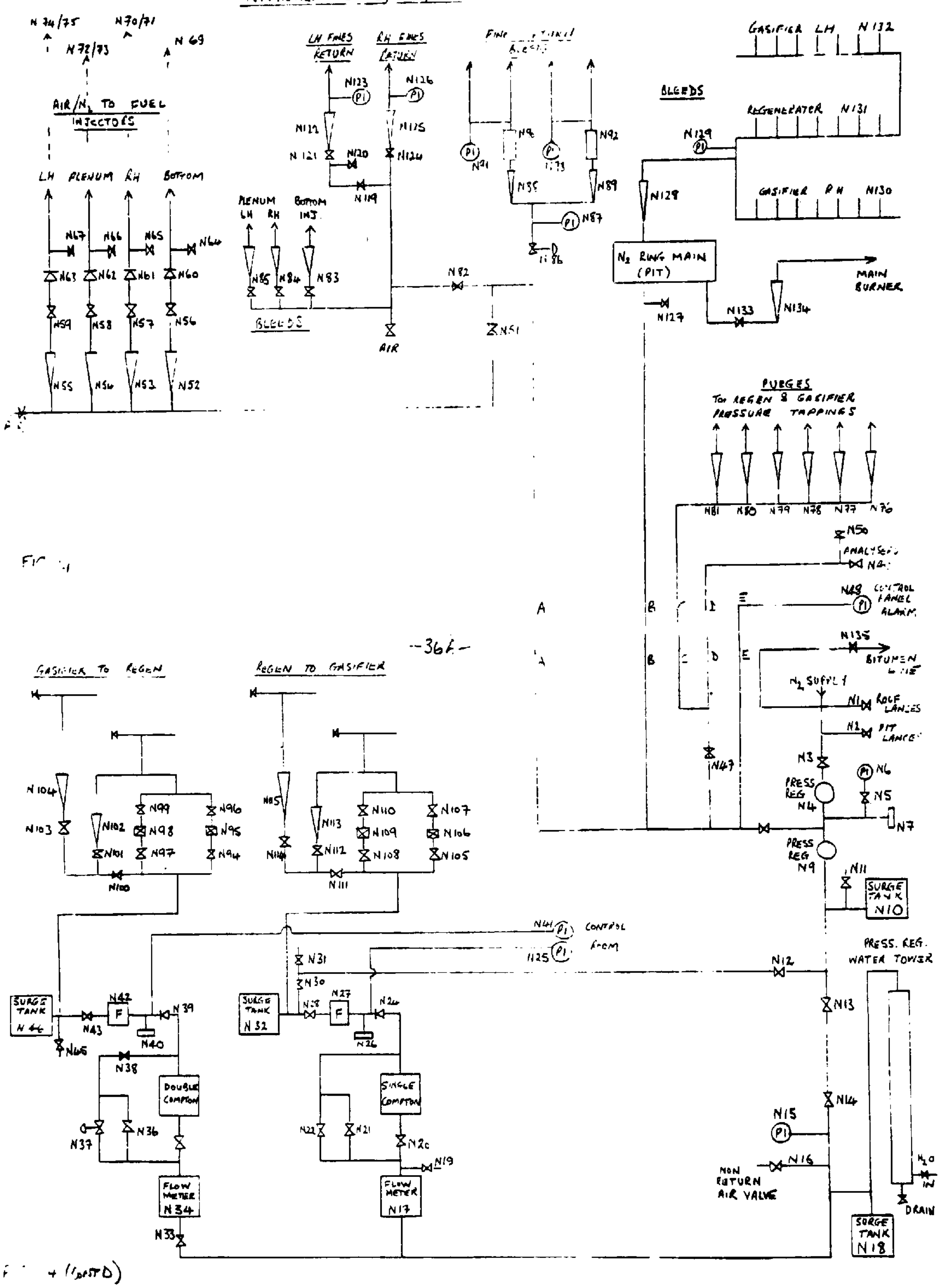
FIG. 11



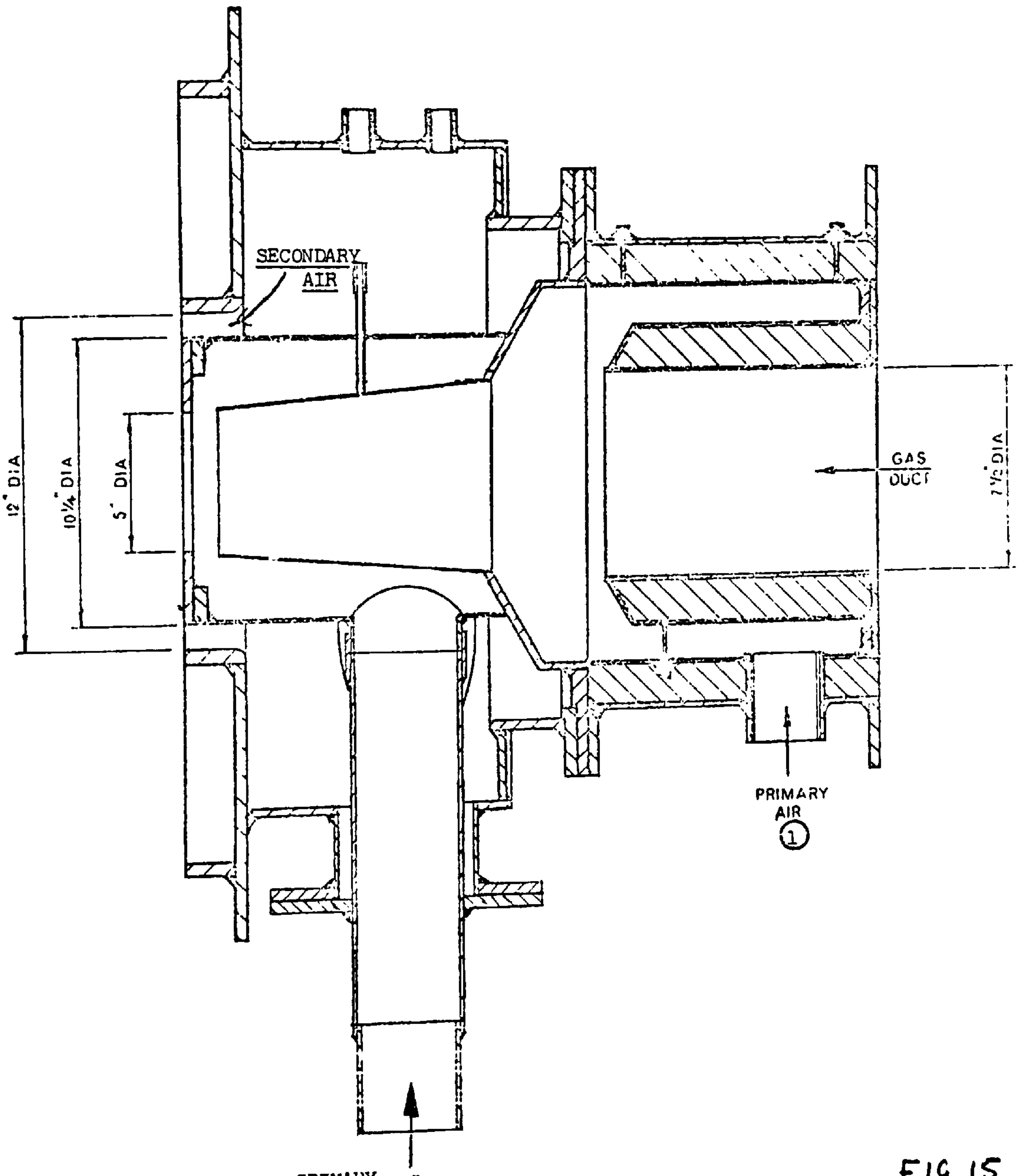


GAS OIL SUPPLY FOR BURNER FUEL TANK
AND PURGING POINTS.

FIG 12


NITROGEN SUPPLY SYSTEM
AIR/N_2 TO FUEL INJECTORS
LH PLENUM RH BOTTOM
LH FINES RETURN
RH FINES RETURN
PLENUM LH RH
BOTTOM INJ.
BLEEDS
AIR
GASIFIER LH N132
REGENERATOR N131
GASIFIER RH N130
BLEEDS
N_2 RING MAIN (PIT)
MAIN BURNER
PURGES
TO REGEN & GASIFIER PRESSURE TAPPINGS
ANALYSERS
CONTROL PANEL ALARM
BITUMEN LINE
N_2 SUPPLY
ROOF LANCES
PIT LANCES
PRESS REG N4
PRESS REG N9
SURGE TANK N10
PRESS. REG. WATER TOWER
GASIFIER TO REGEN
REGEN TO GASIFIER
CONTROL
FROM
SURGE TANK N44
SURGE TANK N32
DOUBLE COMPTON
SINGLE COMPTON
FLOW METER N34
FLOW METER N17
NON RETURN AIR VALVE
H_2O IN
DRAIN
SURGE TANK N18

MAIN GAS BURNER
(PHASE 2)



DWG. No. AE.4965

FIG 15

CAFB PILOT PLANT

FLOW DIAGRAM FOR GAS ANALYSING EQUIPMENT.

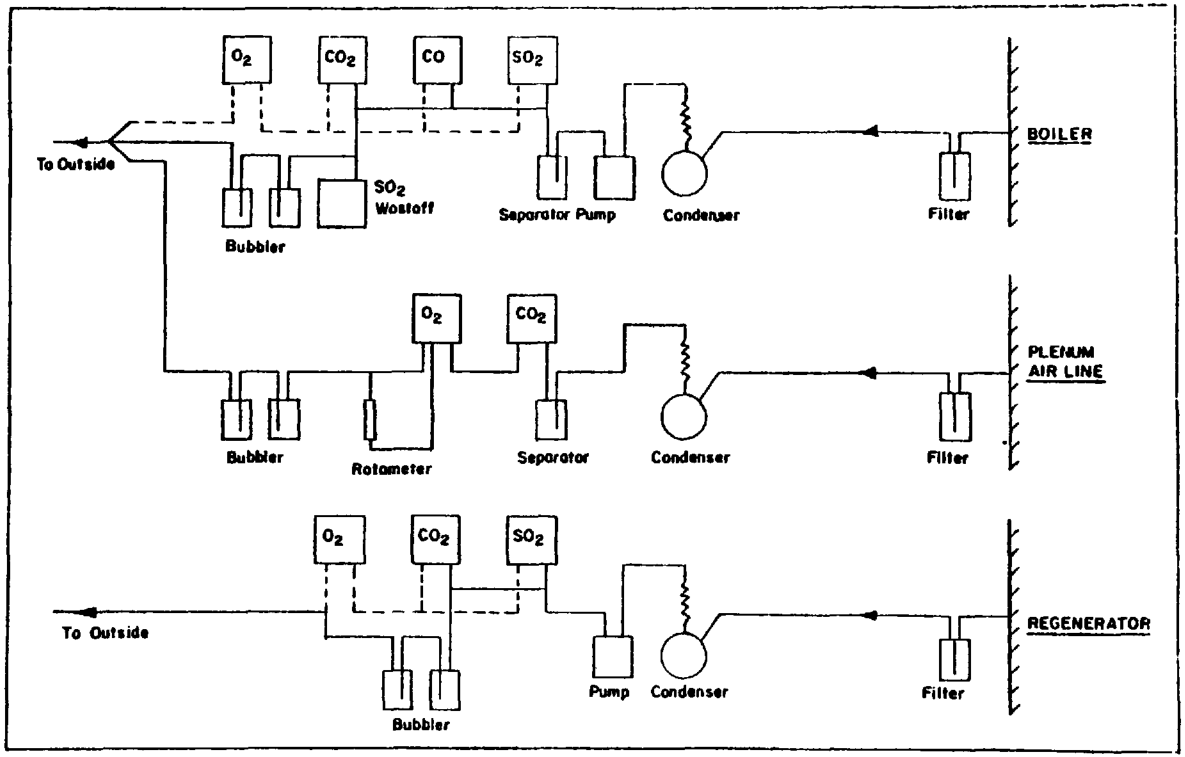
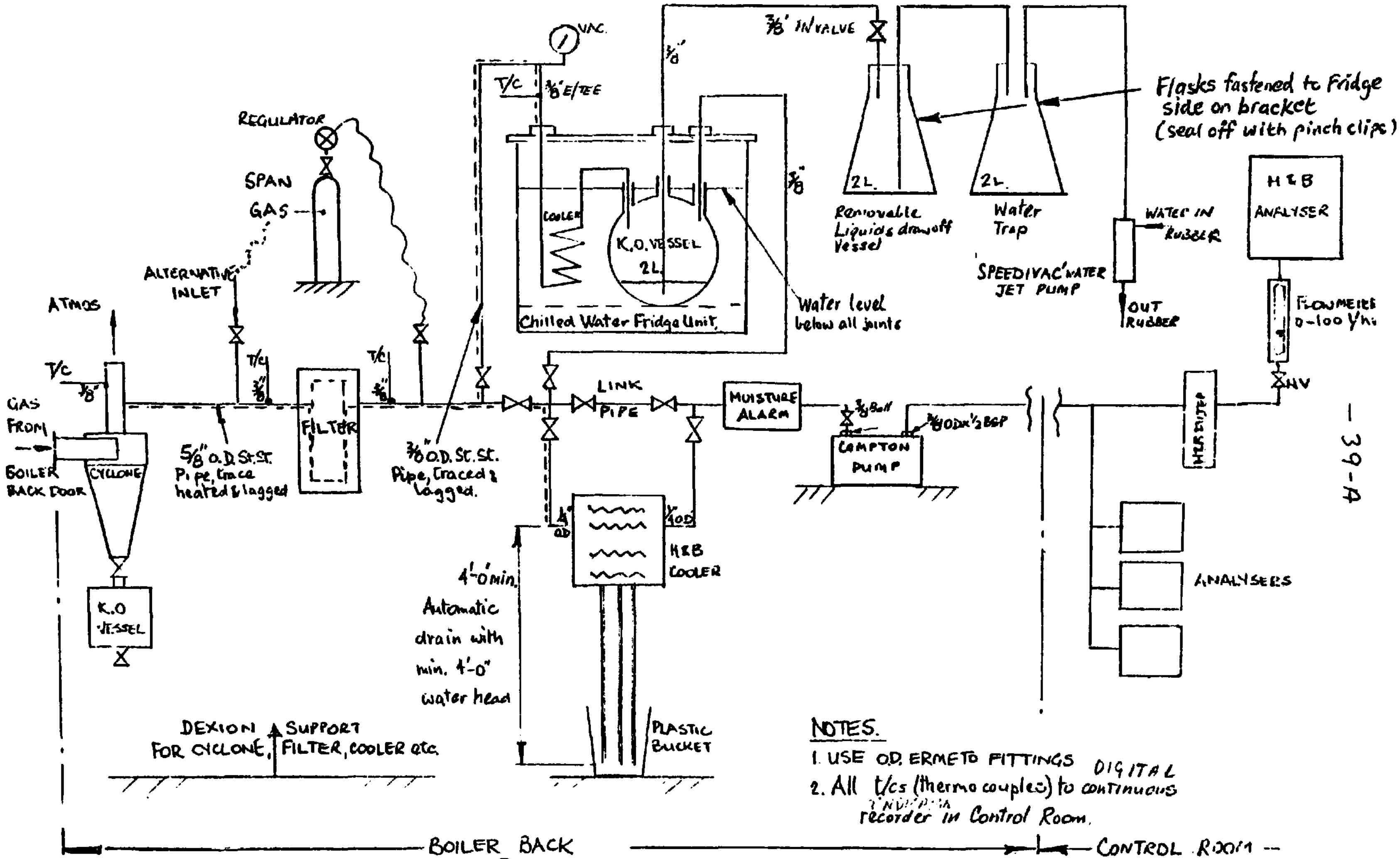


Figure 16



— FLOW DIAGRAM. Flue gas sampling system —

FIG 17

GENERAL PLANT LAYOUT

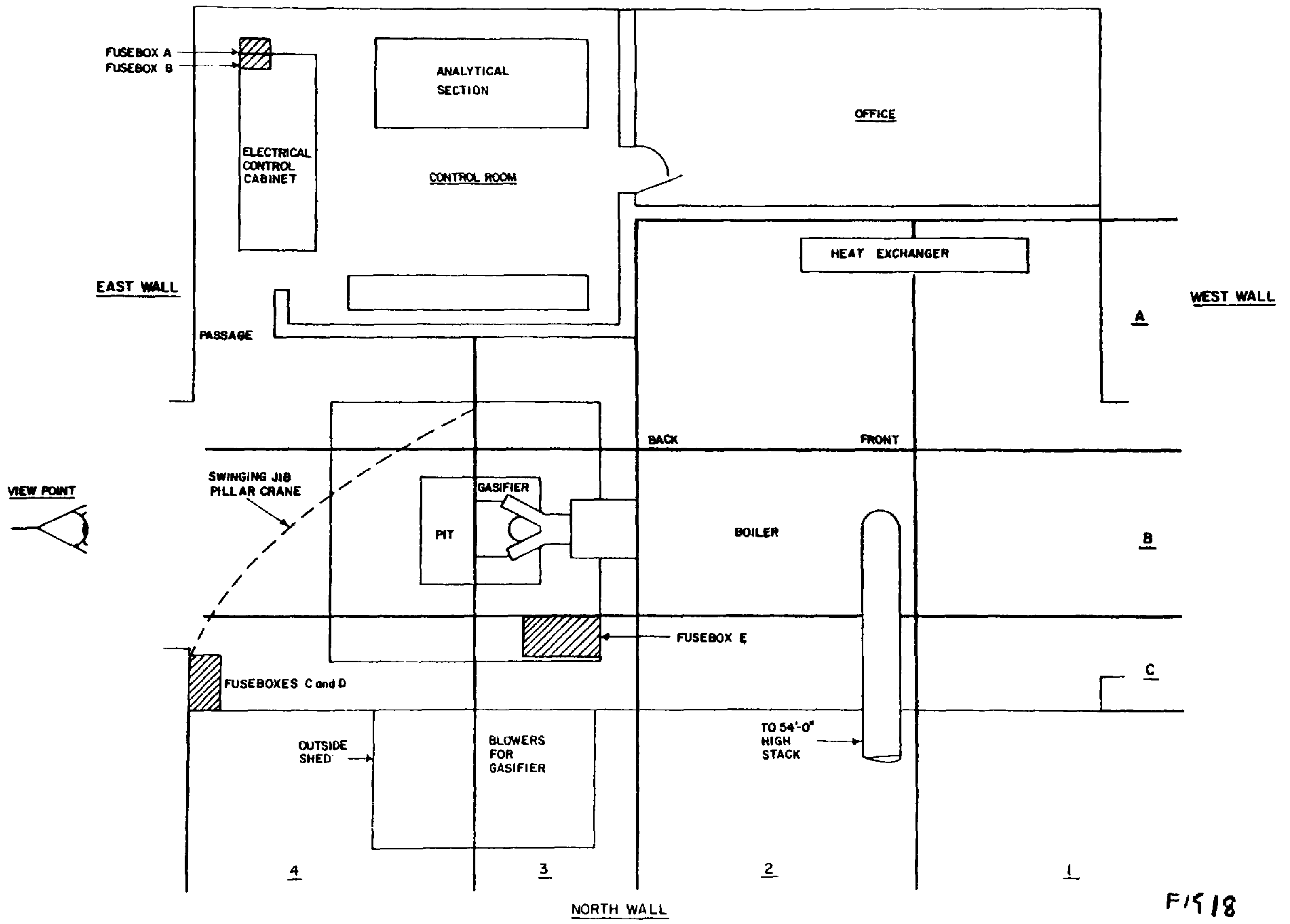


FIG 18

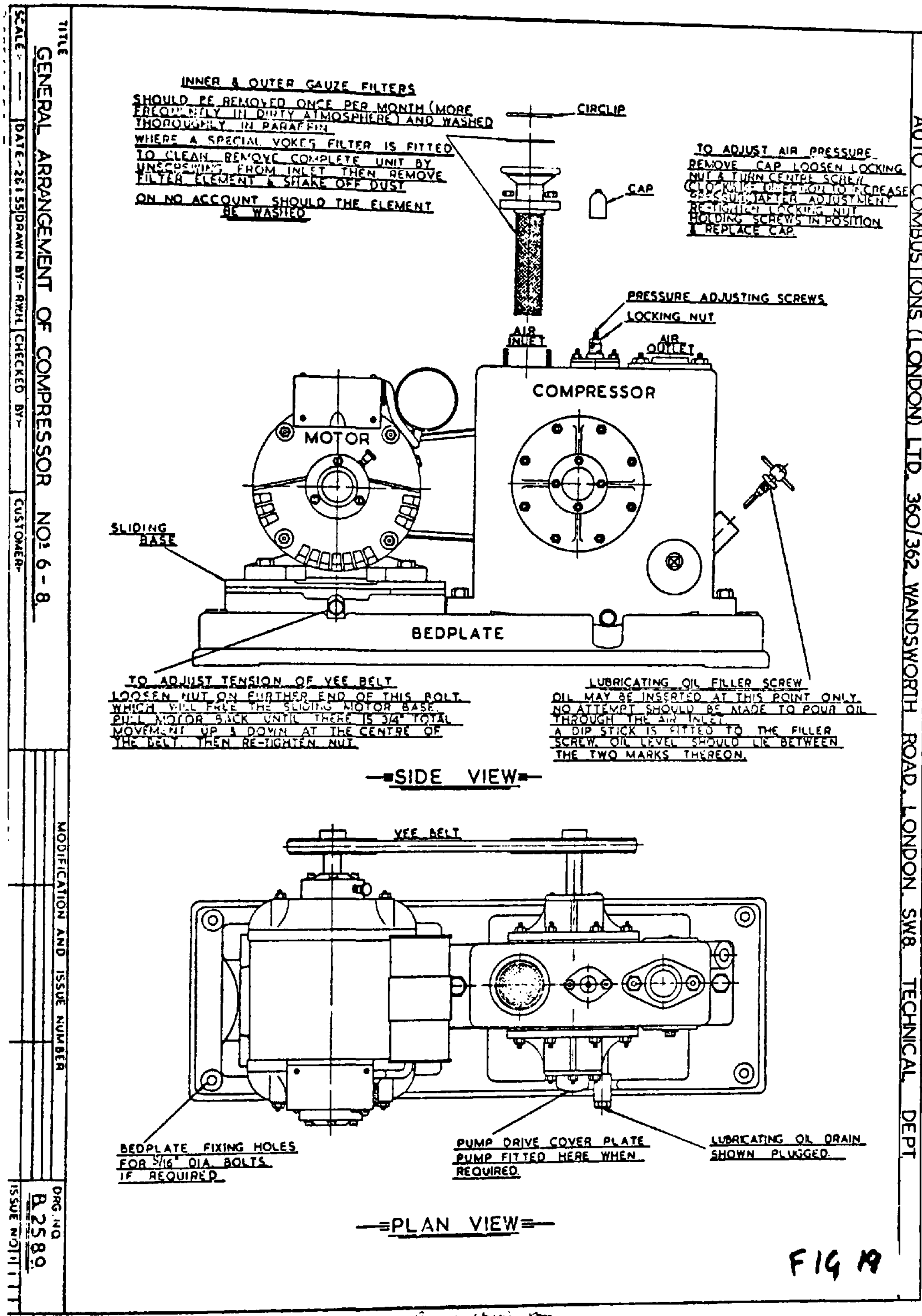


FIG 19

2 cylinder LISTER 10 HP @ 1500 rpm.

AUTO COMBUSTIONS (LONDON) LTD., 360/364, WANDSWORTH ROAD, S.W.8 ENG. TECHNICAL DEPT.

"B"

1. OIL SPINDLE.
2. SPINDLE BUTTON.
3. FLANGED BELLOWS WITH GASKET.
4. AIR DIAPHRAGM.
5. CHEESE-HEADED DIAPHRAGM SPINDLE (CONTACTING BELLOWS.)
6. AIR VENTS.
7. BOTTOM CASTING.
8. TOP CASTING.
9. THERMOSTAT CONNECTION (AIR LEAK TYPE.)
10. AIR UNION.
11. AIR CONNECTION FROM MAIN AIR LINE.
12. THROTTLE DISC (HELD BY AIR UNION AND LIMITING AIR SUPPLY TO CUT-OFF VALVE.)
13. TOP CAP.
14. SPRING (ON SPINDLE.)
15. SOFT OIL SEAT (IN SPINDLE.)
16. DIAPHRAGM DISC.
17. TOP CAP WASHER.

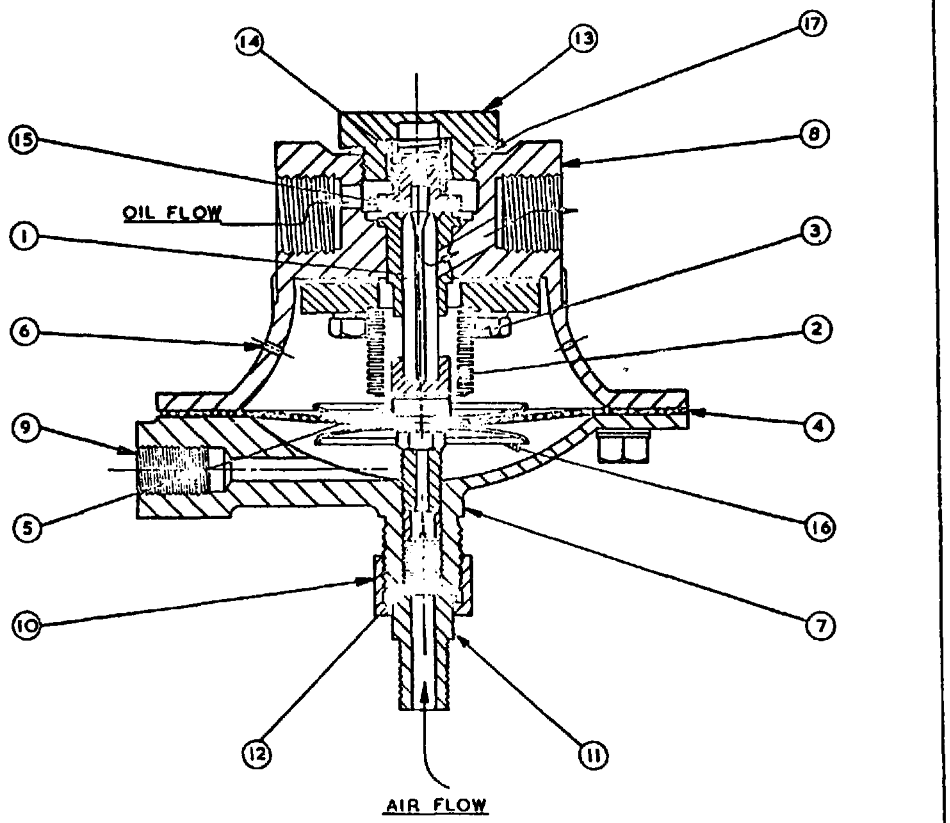


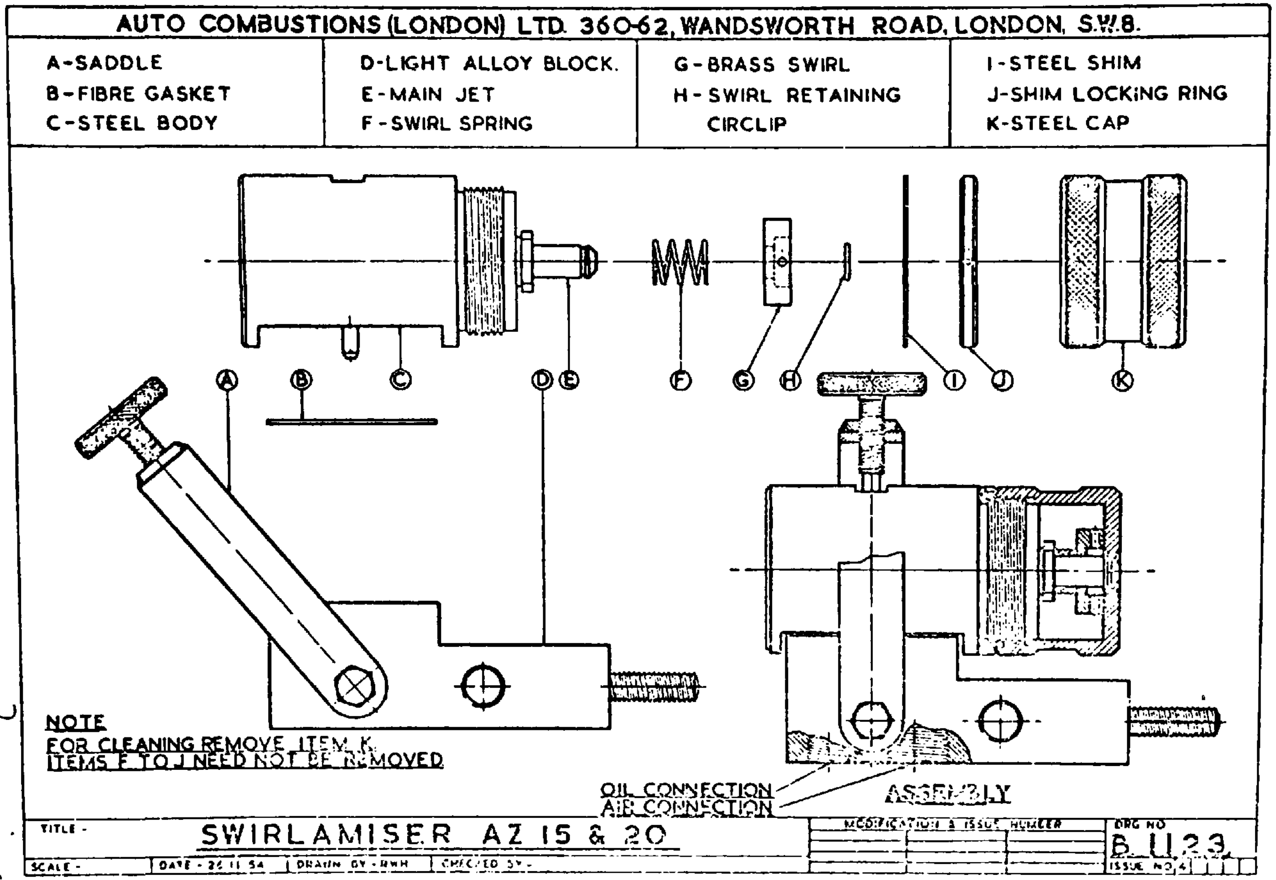
Fig 20

ASK

GENERAL ASSEMBLY OF CUT-OFF VALVE.

B.1027.

15-6-60 P.A.H.



Max^m Capacity of burner AZ 15 2 gal/hr } economy air pressure from
AZ 20 4 gal/hr } compressor 10
7/10 lbs/□"

Fig 21

- 284 -

- 447 -

AUTO COMBUSTIONS (LONDON) LTD

OPERATION & SETTING OF S TYPE THERMOSTAT

'B'

EXPLANATORY

COLD POSITION. THE OPERATING BAR ① IS FORCED OFF THE BALL VALVE BY THE OPERATING ROD ③ THUS ALLOWING AIR TO ESCAPE FROM THE AIR OPERATED CUT-OFF VALVE, VIA THE PIPE LINE ⑥ SHUTTING OFF OIL TO BURNER

FOR THE PURPOSE OF STARTING THE ISOLATING COCK ON PIPE LINE ⑥ SHOULD BE IN THE OFF POSITION.

HOT POSITION THE OPERATING ROD ③ CONTRACTS FROM THE OPERATING BAR ① CLOSING THE BALL VALVE ⑤ THUS ALLOWING AIR PRESSURE TO OPERATE CUT-OFF VALVE. THE ISOLATING COCK SHOULD BE TURNED TO THE ON POSITION AFTER THE FIRE HAS BEEN BURNING FOR 15 MINUTES THUS MAKING PLANT SAFE SHOULD A FLAME FAILURE OCCUR.

SETTING FOR USE AS FLAME FAILURE

TURN OFF ISOLATING AIR COCK SITUATED ON PIPE LINE ⑥ AND RUN PLANT ON SMALL PILOT FIRE FOR 20 MINUTES WITH ISOLATING AIR COCK STILL IN THE OFF POSITION ROTATE THE ADJUSTING SCREW ②(ANTI-CLOCKWISE TO OPEN VALVE ⑤ CLOCKWISE TO CLOSE) UNTIL VALVE IS OPEN THEN ROTATE THE ADJUSTING SCREW IN A CLOCKWISE DIRECTION UNTIL BALL OF VALVE IS JUST SEATING NOW TURN ISOLATING COCK TO THE ON POSITION - BURNER SHOULD STAY ALIGHT IF BALL VALVE ⑤ IS CORRECTLY ADJUSTED

OPERATION

IN THE EVENT OF FLAME FAILURE - EXPANDING TUBE ④ COOLS DOWN OPERATING ROD ③ PUSHES ON OPERATING BAR ① ALLOWING BALL VALVE ⑤ TO OPEN RELEASING AIR FROM CUT-OFF VALVE AND SHUTTING OFF OIL TO BURNER

SETTING FOR USE AS STEP-IN-START

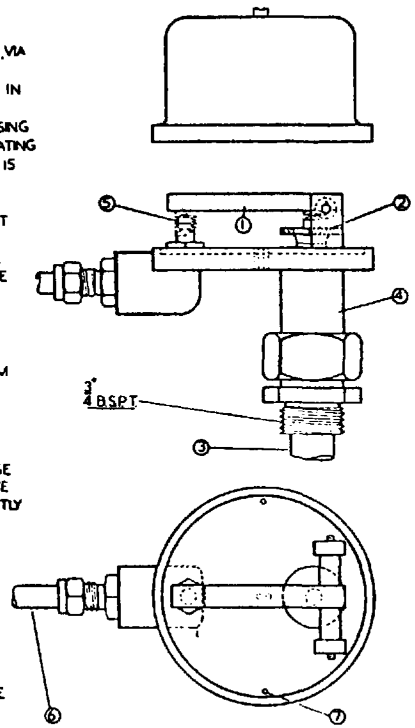
TURN OFF MAIN METERING VALVE AND RUN PLANT ON PILOT FIRE FOR ONE OR TWO MINUTES ROTATE ADJUSTING SCREW ②(ANTI-CLOCKWISE TO OPEN VALVE ⑤ CLOCKWISE TO CLOSE) UNTIL VALVE ⑤ IS OPEN, THEN ROTATE THE ADJUSTING SCREW IN A CLOCKWISE DIRECTION UNTIL BALL OF VALVE ⑤ IS JUST SEATING - NOW TURN ON MAIN METERING VALVE AND ADJUST TO FULL FIRE THIS OPERATION CAN BE CARRIED OUT IF VALVE ⑤ IS CORRECTLY ADJUSTED

OPERATION

PLANT STARTS ON SMALL FIRE AND BALL VALVE ⑤ CLOSE AFTER ABOUT ONE MINUTE THUS ALOWING AIR PRESSURE TO OPERATE CUT-OFF VALVE AND TO SUPPLY OIL FEED TO THE PRE-SET METERING VALVE TO GIVE A FULL FIRE

DATA:

① OPERATING BAR ② ADJUSTING SCREW
③ OPERATING ROD ④ EXPANDING TUBE
⑤ BALL VALVE ⑥ COPPER PIPE TO CUT-OFF VALVE
⑦ AIR VENT (4 OFF)



TITLE: SETTING INSTRUCTIONS FOR S TYPE THERMOSTAT	MODIFICATION & ISSUE NUMBER	DRG No: B.1018.
SCALE — DATE 15 7 53 DRAWN BY CHKD BY CUSTOMER		ISSUE No 1

Fig 22

II OPERATING PROCEDURES

Safety

There are many possible hazards in the CAFB operation, and it is essential that every person works carefully and safely. The plant must be operated by a three man team unless it is a warming up operation when two men operation is permissible provided the activities are limited to recording and controlling the temperature.

If it becomes necessary to enter the pit at any time there must be two men available at the top of the pit to render assistance should the man in the pit get into difficulties. Two sets of breathing masks fed from bottles of breathing air must be available, one for use by the man in the pit, and one for the man standing by at the top of the pit.

Whilst every precaution has been taken to reduce the dust in the laboratory area by local dust extraction ducts it may be necessary to wear respirators when in this area when certain actions are carried out. Please ensure that you change the filters regularly and wash out the respirator after use.

There are also three positive pressure respirators which are powered by rechargeable batteries fixed to the belt which also houses the blower and filter unit. These are available for general use, but people must thoroughly wash the mask (as instructions) after use, and ensure that the battery is recharged.

Spare primary and main filters are available.

Ear protection is available either as ear cups or Bilsholm wool plugs - individuals may find that the use of this protection reduces fatigue after long periods of time in the noisy area.

Eye protection must be worn at all times and in some cases full goggles or face shields must be used if there is a danger of hot or dusty material being thrown or dropping onto the face.

Asbestos gloves are available and particularly must be worn when draining hot lime samples from the unit or any other operation involving hot material. Light weight gloves are available for general use.

A safety helmet has been purchased for each person, and it is important that these are always worn when in the gasifier area. If they are too bulky for some operations there are some hard hats available for general use in such circumstances.

In time of trouble assistance may be called by use of the alarm buttons which sound outside the laboratory in 3A and the stone storage garage.

Fire

There are three powder fire extinguishers - one in the control cubicle and two in the gasifier area. Be familiar with their location and method of operation, but remember that their operation time is short and they will only deal with small fires. Also available are two portable CO_2 extinguishers

- one in the crew room and one in the gasifier area near the control room door.

There is a CO_2 hose reel on the wall immediately outside the main sliding door - there are two spare full CO_2 cylinders on the wall adjacent. The main reel must be immediately recharged with the two new bottles after it has been used.

There is a fire hydrant close to Featherbed Lane near the main door. The hose will be laid out as a safety precaution with an adjustable spray nozzle/single jet fitting located on the end.

Fuel and N_2 Supply Pre-run Checks

(a) Heavy Fuel Oil

- 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 five 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) Bitumen System Preparation

- Ensure that the weight of the trailer is completely taken on the four static jacks (35) provided, to relieve the weight on the tyres (see Figure 9).
- Check
  - (a) that the 10 HP motor is connected to the electrical mains through suitable control gear.
  - (b) lubricant level in the rotary compressor (3) on the dip stick.
  - (c) 35 seconds gas oil tank level for supplying the burners (after filling always ensure that the tank cap is replaced and tightened). Fuel tank capacity - 50 gallons.
  - (d) that the hydraulic system oil reservoir (20) is topped up (Teresso 43) to line in the bottom of the filling filter and maintained at that level. Reservoir capacity - 20 gallons.
- BEFORE CHARGING THE STORAGE TANK WITH HOT BITUMEN check that the drain valve (34) on the sludge door beneath the tank is CLOSED.
- Ensure that the hydraulic selector valve (19) is placed in the NEUTRAL position on electric motor start up. The hydraulic pump driving the Barclay Kellett pump must only be operated when the bitumen has reached 270°F temperature, and must NEVER be operated whilst the bitumen is cold.

(c) Operating the Bitumen Heat-up Burners (10)

1. Having completed the checks under "Bitumen System Preparation" section, start the 10 HP Brook electric motor. This will rotate the compressor AND the hydraulic motor. Make sure the selector valve (19) is in NEUTRAL.

2. Allow the compressed air pressure to build up according to the gauge mounted above the compressor.

3. Adjust the pressure between 7 & 8 p.s.i. (see Figure 9). Air will now be supplied to the burners. Burner operating pressure 5-10 p.s.i., according to fire required.

4. Fully open oil shut off valve (15) in the oil line from the fuel tank, situated under the bitumen tank at the burner end.

5. Open oil metering valves (11) and light burners with a suitable open flame torch.

6. Regulate valve (11) until black smoke appears at the chimney, and then gradually close valve (11) until only a blue haze appears at the chimney.

   Maximum fuel efficiency is thus obtained.

7. Set the flame failure thermostats (13) as instructed on print B1018 under heading "Setting for use as flame failure".

8. When the bitumen maximum temperature has been attained and the burners extinguished automatically through the closing of the thermometer contacts, the oil metering (11) and oil shut off (15) valves should be closed.

(d) Operating the Bitumen Trailer

The swivel pipe (41) has been provided with $1\frac{1}{2}$" flow and return branches for the ring main (described later) and a flow control valve between the two branches. This f.c. valve must be fully open on start up, and only closed during operation sufficiently to divert a flow of hot bitumen through the ring main. It must never be closed whilst the electric motor is running.

The Barclay Kellett discharge pump (36) is controlled by the hydraulic selector valve (19) which will be left in the DELIVERY position of ring main pressurisation. This pump is fitted with a two-way suction pipe to allow hot bitumen to be drawn off just above the burner level through the open valve (37) or to empty the tank within $\frac{3}{4}$" of the bottom (valve 37 now closed). To divert the hot bitumen through the swivel pipe and thus pressurising the ring main, the swivel pipe lever must be in the DELIVERY position.

Summary of valve positions for pressurisation of ring main -

(19) Hydraulic selector valve - DELIVERY

(37) Swivel pipe valve - DELIVERY

(e) Operating the Bitumen Ring Main

The $1\frac{1}{2}$" ring main from the swivel pipe on the P.T.U. to inside the North end of the boiler house is steam jacketted and lagged to maintain the high temperature of the heated bitumen (350°F). For an emergency, and actuated by the collapse of a fusible link located above the boiler burner, a fire valve is fitted in the bitumen flow line and is located outside and above the control room, (B9, Figure 8).

To charge the ring main with hot bitumen:-
(assuming that the bitumen is at the desired temperature, that the bitumen pump on the trailer is circulating the hot bitumen, and valve B9 is closed)

- Open fully the $1\frac{1}{2}$" valve B1 on the swivel pipe, and slowly operate the 3" valve B2 (MUST NEVER BE COMPLETELY CLOSED) until hot bitumen returns to the swivel pipe. This can be observed through the bleed valve B4. Care must be taken during this operation not to overload the trailer bitumen pump. In the event of this pump failing the standby bitumen pump must be brought into operation.

(f) Standby Bitumen Pump Operation

- Switch on the trace heating on the 3" manifolding.
- Rope-start the Petter diesel engine and allow to warm up for 5 minutes.
- Open the 3" stop valve B5.
- Engage the pump drive clutch.

Should the failure of the trailer bitumen pump also mean that the compressed air to the burners has been cut off then:

- Engage the compressor drive clutch to restore air pressure.
- Re-light the burners after air pressure has been restored.

Both these pipe systems can be purged with gas oil through valve G8 with the gas oil supply in operation as per Figure 12.

The trailer bitumen pump should be put back into use as soon as the repairs/servicing are complete.

(g) Bitumen Pump Calibration

A 1" valved spur (valve B9, Figure 8) is taken from the $1\frac{1}{2}$" ring main to supply hot bitumen through a filter and flow meter to each of two Plenty pumps. These have vanes rotating in a variable orifice to vary the rate of flow.

To supply hot bitumen to the Plenty pumps for pump calibration:

- Switch on electrical trace heating to raise the temperature of all the metal hardware, i.e. pipes, filter, meter and pump casings, as near as possible to the hot bitumen temperatures.

- Check that all the systems valves are closed, then:
- Open valves B9, B12, B15 and B16.
- Start up motor on right hand pump.
- Disconnect $\frac{1}{2}$" supply line at injector and have tared 7 lb. empty tin handy.
- Open valve B19 and collect throughput of bitumen in the tin and time with stop watch.
- Avoid spillage and have dry powder extinguisher available.
- Purge the system through by opening valve G12 to admit gas oil; close valve.
- Reconnect $\frac{1}{2}$" line at injector.
- Close all valves.
- Repeat, opening valves B9, B12, B25, B26 and starting left hand pump.
- Disconnect line to L.H. injector, open valve B28 and calibrate pump.
- Purge with gas oil, reconnect $\frac{1}{2}$" line and close all valves.
- Ensure that no bitumen is present in the system which will solidify when the trace heating cools.

(h) Gas Oil Supply to Service Tank and Purge Lines

To replenish gas oil in 50 gln service tank for the trailer's burners -

- Open foot valve G1 (hand wheel near tanker catwalk) (see Figure 12).
- Open springloaded valve G2, with special key.
- Open Worcester valve (key parallel to pipe).
- Go over to pump position, close the isolator and start the pump motor (make sure valve G4 is tightly closed).
- Mount the P.T.U. platform, towbar end of the trailer.
- Open G5 on the end of the flexible hose and fill the service tank.
- When full, close G5 and stow away flexible safely.

To provide gas oil purge to the bitumen supply to the injectors system:

- With G1, G2, G3, valves open start the pump motor.
- Open G4, gas oil will now flow to the purge point, through the flexible hose, valve G.

(i) Kerosene

- Stored in 500 gallon 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. Check fire valve has not actuated.

(j) Nitrogen

- Stored in liquid N_2 tank. Check that there is sufficient for anticipated usage. Air Products (Bracknell) must be alerted to make daily "topping up" visits during the operational period.
- Check all valves are turned off at the bleed locations on gasifier before opening valve on manifold.

(k) Propane

- Check valve on tank and valve outside building 3F is open.

Boiler and Systems Pre-run Checks

- Check water level in pressurisation unit header tank is at rubber band marker.
- Make up water supply open to tank.
- Pressurisation unit on: check N_2 pressure is approximately 48 p.s.i.
- On start up bell will ring. Cancel bell. N_2 off at cylinder.
- Air bled from boiler cooling circuit.
- Valves open from water reservoir to cooling tower located on first floor of water tower (valve is labelled).
- Notify site services of soft water usage.
- Start boiler circulating pump.
- Start cooling tower circulating pump; hold cut-out for 15 seconds.
- Power on to cooling tower fan.
- Check automatic mixing valve working (temperature setting 160°F).
- Open cooling system purge. Check for flow in drain pit by blower house.
- Switch boiler panel on.
- Check fuel oil circulation in ring main.
- Check fuel oil circulation in boiler circuit.
- Check line up of three way valve for flow through boiler circuit heater.
- Check water pressure in boiler is approximately 48 p.s.i.
- Check sample lines are correctly installed in flue.
- Boiler door shut, held on four bolts.
- Ensure flue damper fully open; (manufacturer markings are not correct).

- Fridge out of the way.

Gasifier Pre-run Checks

- The gasifier is assumed cold and empty of bed.
- Check belts O.K.
- First stage blower.
- Second stage blower.
- Regenerator blowers.
- Extract blower.
- Vacuum cleaner for stone feed.
- Lubrication
- Fuel circulating pump.
- Fuel metering pumps (3).
- First stage blower.
- Second stage blower.
- Recycle blower.
- Regenerator blowers (3).
- Main burner blower.
- Extract blower.

- Tuyere blower.
- Cubicle blower.
- Cooling tower water circulating pump.
- Boiler water circulating pump.
- Oil circulating pump under boiler.
- Bitumen circulating pumps (2).
- Bitumen metering pumps (2).
- Analysers
- Sample lines.
- Calibration.
- Pumps working.
- Filters and traps.

Gasifier Warm Up

- Check action of alarms.
- Alarm switch to "all alarms show".
- Bell switches to MUTE.
- Auto shutdown switches to inactive.
- Main panel power on.
- Instrument air on - 20 p.s.i.
- Instruments on and working - charts O.K.
- N_2 supply to ring main - 40 p.s.i.
- N_2 to pressure tap bleeds - 2.5 CFH each.
- All manometers full.
- D/P cells zeroed.
- Regenerator blower on - 5 CFM flow.
- Check water level in Compton supply seal - 9 FT.
- Line up to Comptons. Check operation. Set to 3 p.s.i.
- Switch off Comptons, line up N_2 via by-pass which avoids meters, i.e. open valve 98 and 115 - close valves 97 and 114.
- Adjust N_2 to 3 p.s.i.
- Pocket agitator N_2 flow to 4 CFH.
- Injector N_2 flow to 0.3 CFM.
- Start pulsers and check action.
- Check solenoid action and line-up to main (M).
- Regenerator temp control on manual.
- Regenerator control switch behind panel to "off".

Pre-Heat Burner Light Up

- Close valve in line from stone feed hopper to gasifier.
- Open line to low range orifice for gasifier air.
- Ensure high range orifice inlet blanked.
- Ensure:- Propane gas line to burner open (three valves, on propane tank, outside and inside Building 3F).
  - Igniter and flame eye connected.
  - Visual sight glass clean and tight.
  - Purge to flame eye connected (4 CFH).
  - Burner air valve open and throttle shut.
- Start both gasifier air blowers, set to 130 CFM.

- Ensure flue gas recycle valve closed.
- Ensure burn out connection shut.
- Close plenum air valve and open burner throttles.
- Compare burner air flow indicator and gasifier air flow indicator with Anubar.
- Check zero setting of both air flow gauges.
- Open plenum air valve and set throttle to marking 1.5.
- Adjust burner air to 30 CFM.
- Check regenerator air - 5 CFM.
- Air to fuel lower side injectors - 5 CFM each.
- Check centre fuel injector retracted and N_2 purge to sleeve to 0.7 CFM, and small air flow to injector.
- Upper side injectors retracted and 4 CFH air bleed injected.
- Plenum fuel injector extracted and 1 CFM air injected.
- Start boiler burner blower - 200 CFM.
- Air flow to stone feed - 90 L/Min.
- N_2 flow to each fines return systems - 4 CFM.
- 4 CFH N_2 bleeds to fines returns lines.
- Start regenerator pressure control blower, and set G/R ΔP controller to zero.
- Check that propane pressure reads 30 p.s.i.
- Turn on electric power to start up burner control.
- Open pilot propane cock in the control room.
- Close cock slightly on air to pilot to reduce pressure.

- Activate start up burner control and go to propane throttle by gasifier.
- Adjust propane flow to 40 CFH as soon as propane solenoid opens.
- Shut off pilot cock in the control room when white dot visible on the controller, and adjust burner air flow to 40 CFM.
- If flame does not lock on, allow one minute purge before starting again - from "open pilot propane cock in the control room", reset burner air flow to 30 CFM.
- Adjust back pressure to send hot gas to regenerator by increasing G/R ΔP but avoid excess temperature in R to G return pocket as measured by temporary thermocouple.
- Control the temperature on return pocket by adjusting G/R pressure balance.

- Increase flow of regenerator air if necessary.
- Monitor temperatures and adjust gas and air flows to follow temperature schedule of 15°C per hour.
- Maintain air/fuel ratio in the burner of 1 CFM/1 CFH and about 100 CFM air flow through the plenum.
- Sequence for increasing the propane firing rate is:
  - Increase total air (motor valve).
  - Increase propane rate.
  - Adjust burner air.
  - Open plenum air throttle as required.

Pre-Heat Burner Flame Out

- Flame out: in case of flame out allow one minute purge time, identify cause of failure and correct eg dirty fire eye then follow "pre-heat burner light up" from the line "open pilot propane cock in the control room". When re-lit increase propane and air (including plenum air) to the value before flame out and continue raising temperature.
- Observe for hot spots on the shell - correct by stopping leaks with fibrefrax or asbestos and sairset and by water spray.

Heavy Fuel Oil Pump Calibration

- Close fuel valves to injectors.
- Open sampling valves.
- Verify trace heating on and pipes hot.
- Line oil through to the metering pumps and calibrate.
- Check operation of 80 p.s.i. overflow valve.
- Line up fuel to centre injector.
- Turn off fuel oil, turn on kero and line through to kero barrel, disconnect oil line to centre injector in the pit and clear through any F.O. into a bucket with kero (avoid spillage and have dry powder cylinder available). Reconnect oil line to centre injector.

Change From Propane to Kerosene

- When gasifier temperature reaches 700°C raise centre injector to 3" and begin kero injection at very low rate with 5 CFM of air.
- Gradually replace gas flow with kero while increasing plenum air flow to 250 CFM and lowering propane to 50 CFH.

- Line up air supply to the fines return control panel.
- Adjust air to fines return pulser to 3 p.s.i.
- Set bleed below $\frac{3}{4}$" Audco on each fines return pipe to 4 CFH and open the Audco valve.
- Set N_2 to each fines return injector to 4 CFM.
- Set timers on fines return control panels to 5 min. cycle and switch on.
- N_2 rate to fines return injectors and timers on the fines return panel may need adjustment once the system settles.

Starting Stone Feed

- Record weight of limestone supply and fill ground level feed hopper.
- Close main valve and balance valve between stone hoppers.
- Start motor on transfer vacuum line.
- Open valve on vacuum line and valve on fill line and suck stone up to overhead hopper.
- Close fill line and vacuum line valves.
- Equalise pressure between stone feed hoppers by opening balance valve.
- Open big valve between hoppers and drop lime into lower hopper.
- Check that weigh cell is working - record weight change.
- Close main and balance valves between hoppers.
- Check N_2 pressure to injectors is 3 p.s.i.
- Open N_2 to meters.
- Check bleeds to injectors and agitators (0.3 CFM, 4 CFH respectively).
- Set bleed N_2 flows to pressure tappings (2.5 CFH).
- Pressurise outer casing of gasifier to 10" WG and note flow rate.
- Check manometer fluid levels.
- Increase regenerator air flow to 15 CFM.
- Adjust pressure balance to stop G to R gas flow.
- Check regenerator to gasifier injector bleed is at 0.3 CFM.
- Check that temperature at G to R pocket is below 1,000°C.
- Start rapper stack cyclone.
- Do following to start stone feed:
  - Start 1 CFM N_2 to stone hopper vessel.

- Open valve between hopper and gasifier.
- Start vibrator feeding lime at 50 lb./hr.
- Check that lime is being fluidised.
- Increase lime rate as conditions permit - keep bed temperature above 850°C.

- Check that the cyclone drains and fines return system operates satisfactorily:
  - Visually from tops of cyclones.
  - Visually through tops of boxes.
  - Observe valve opening and closing sequence.
  - Check that valves are opening and closing fully.
- As bed depth increases to near 5" you may need to increase the propane pressure by screwing down the propane pressure reducer. Count the number of turns and record in log book.
- When bed depth reaches 5" with good kero combustion.
- Close main valve in propane line to burner.
- Turn off power to start up burner control.
- Close air valve adjacent to burner, 2" ball valve.
- Start 4 CFM air to burner purge.
- Stop combustion air to start up burner.
- Open plenum throttle and adjust plenum air to 170 CFM using motor valve.
- Adjust kerosene rate to maintain 870°C - note rate in log.
- Adjust regenerator pressure to 3-4" below gasifier.
- Adjust pulsers to obtain good solids circulation.
- Continue addition of limestone to specified level.
- Check regenerator drain operation.
- Check cyclone drain system as specified above.
- Timing of the fines return cycle may need changing.

Boiler Clean Out

- Line up equipment and manpower to clean out boiler:
  - Rake.
  - Breathing apparatus.
  - Door bolts.
  - Spanners.
- Stop stone feed vibrator.
- Close valve between stone feed and gasifier.
- Stop N_2 feed to stone hopper.

- Raise gasifier temperature to 950°C.
- Perform following operations in quick sequence to interrupt combustion and hold temperature.
  - Stop fuel pumps and close fuel line valves.
  - Switch off gasifier blowers.
  - Close hand valve in air line to plenum.
  - Stop regenerator air blower.
  - Turn bed transfer pulsers to 10/2.
  - Reduce air to all fuel injectors protruding into bed to 1 CFM.
  - Stop air to stone feed.
  - Stop N_2 to fines return.
  - Stop cyclone drain controllers.
  - Stop regenerator P control blower.
- Perform following operations rapidly to prevent

  too much temperature loss in gasifier bed.
  - Clean solids from boiler tubes.
  - Remake boiler door seal with $\frac{3}{4}$" rope + bostick and close door.
  - Check that main flue damper is open.
  - Check flue gas recycle valves are closed.
- When all is ready, and before bed temperature falls below 700°C resume combustion by following steps in quick order:
  - Start air to protruding fuel injectors - 5 CFM each.
  - Start blowers and open air to plenum, check fluidisation.
  - Open fuel line valves and start pump at rate for steady 870°C (see log).
  - Start air to stone feed.
  - Start N_2 to fines return.
  - Start air to regenerator and adjust to 15 CFM.
  - Start regenerator P control blower.
  - Start cyclone drain controllers.
  - Increase pulser rates to 5/1.
- Raise gasifier temperature to 870°C.
- Resume stone addition:
  - N_2 to stone hopper.
  - Open feed valve.
  - Start vibrator.

<u>Start Main Flame Pilot Light</u>

(a) <u>Main Flame Out</u>

- Clean fire eye and start purge air in control room (60 l/min.).

- Set main burner air to 600 CFM.
- Line up all three propane valves to pilot burner.
- Check propane pressure at 30 p.s.i.
- Open air valve to pilot burner (set at 12 CFM, pitot 0.1).
- Turn on power to pilot burner control.
- Start control sequence.
- When propane opens check flow is 1.9 CFM.
- When lit check stability by inspecting.
- Increase burner air to 800 CFM.
- Check pilot stability.

(b) Main Flame On

- Clean fire eye and check purge at 60 l/min.
- Line up all three propane valves to pilot burner.
- Check propane pressure at 30 p.s.i.
- Open air valve to pilot burner (set at 12 CFM, pitot 0.1).
- Check power on to pilot burner control.
- Start control sequence, but hold in cut out button until propane pressure gauge needle flicks to maximum position and release button.
- Check propane flow is 1.9 CFM.
- Reset pilot burner alarm and cancel mute.

Change From Kerosene Combustion to Fuel Oil Combustion

- Activate alarm bells - except main flame.
- Check cooling system - levels, pressures, flows + set bleed, make up valve in water tower open.
- Check that hand valve (Pit) in the flue gas recycle line is closed and F.G.R. orifice gives zero reading.
- Line up valves from boiler flue through F.G.R. blower and filter to blower house inlet.
- Start flue gas recycle blower.
- With motor valve fully closed, open hand valve (Pit) in F.G.R. line.
- Open flue gas recycle motor valve to give 30 CFM.
- Set plenum air supply to 170 CFM.
- Adjust kero feed rate to give steady 870°C.
- Check that trace heating to oil feed lines is on and hot.
- Close kero valve and open fuel oil valve. Leave pump running and adjust back pressure to 40 p.s.i.
- Stop stone addition:
  - Stop vibrator.

- Close feed valve.
- Stop N_2 to hopper.

- Start air bleed to flame fire eye at boiler back end.
- Check operation of boiler main flame fire eyes (2).
- Re-install and check isolating valve open for rear door fire eye and tapping clear of lime.
- Check rear door main flame fire eye light shows on boiler control panel.
- Switch off Middle and R.H. metering pumps and adjust L.H. pump to give steady 870°C for an hour. Record all flows and temperatures and any relevant comments in the log book.

Change Over From Combustion to Gasification

- Ensure that combustion is steady on L.H. pump only, Middle and R.H. off.
- Set Middle pump to 56 lb./hr.
- Set R.H. pump to 146 lb./hr.
- Stop L.H. pump and start Middle pump. Adjust back pressure to 40 p.s.i. if necessary.
- Control bed temperature by intermittent operation of Middle pump - estimated 20% on 80% off.
- Set L.H. pump to 146 lb./hr.

To Gasify

- Verify boiler pilot alarm and check presence of flame visually.
- Verify all other alarms and operation of main flame fire eyes (which respond to torch light).
- Set high temperature bed alarm to 1,000°C.
- Main flame failure and auto shut-down inoperative.
- Switch Middle pump off and allow bed temperature to fall to 850°C.
- Switch Middle pump on to purge boiler with inerts (10 sec.).
- When bed temperature rises to 870°C do the following simultaneously and rapidly:
    - Switch on L.H. and R.H. pumps.
    - Start stop watch.

  8 sec. - reset main flame failure alarm.

    - If M.F.F. alarm cancelled switch off M.F.F. alarm mute and activate auto shut down.

      Proceed to "on gasification".
    - If M.F.F. alarm did not cancel do as follows:

  9 sec. - reset M.F.F. alarm:

    - If M.F.F. alarm cancelled, do as at 8 sec.

- If M.F.F. alarm not cancelled do as follows:

10 sec. - reset M.F.F. alarm:

- If M.F.F. alarm cancelled do as at 8 sec.
- If M.F.F. alarm not cancelled do as follows:

11 sec. - reset M.F.F. alarm:

- If M.F.F. alarm cancelled do as at 8 sec.
- If M.F.F. alarm not cancelled:

  Switch off L.H. and R.H. pumps.

  Reset L.H. pump to steady combustion rate (approximately 30 lbs./hr.).

  Switch off Middle pump before gasifier temperature reaches 1,000°C but not less than 10 secs. after L.H. and R.H. pumps off.

  Switch on L.H. pump.

  Line out on combustion at 870°C.

Attempt to find out what went wrong, rectify if necessary and repeat from "change over from combustion to gasification".

On Gasification

- Switch off Middle pump.
- Reset high temperature bed alarm to 950°C.
- Adjust metering pumps' pressure to 40 p.s.i. (if necessary).
- Check for leaks on gasifier and seal if necessary with fibrefrax and sairset.
- Resume stone addition:
  - N_2 to hopper.
  - Open valve between hopper and gasifier.
  - Start vibrator.
- Adjust flue gas recycle and air rates to obtain specified temperature and gas velocity.
- Adjust stone circulation rate to obtain desired regenerator temperature.
- Put regenerator temperature control on automatic:
  - At controller.
  - Behind panel switch to "on".
- Check cooling system operation; set cooler control to 160°F.
- Bring conditions to specified level.

Trouble-Shooting on Steady State Conditions

- Do not change any control settings unless one or more of the listed deviations occur.

- Ensure:- preventive maintenance procedures are carried out.

  Watch
  - O_2 level in the flue gas.
  - Gasifier Bed Temperature.
  - Regenerator bed temperature.
  - Gasifier top space pressure.
  - Fuel delivery pressures.
  - Fines return system.
  - Regenerator CO_2 and SO_2.
  - Compton pressures.
  - Boiler flue gas exit temperature.
  - Gasifier bed depth.

- Enter in the log book, details of any incidents which occur, together with the time of occurrence and the remedial actions which were taken.

- Possible Incidents due to Malfunction

1. Bed Temperature Starts to Rise Rapidly

(a) Check O_2 level in flue gas. If this is rising then the fuel is not getting through to the gasifier.

(b) If O_2 level is normal check the stone feed and see that this is functioning properly. If not try adjusting the feed rate, and other remedial actions such as refilling the hopper.

(c) If the O_2 is rising then check the fuel pump delivery pressures. If these are high (80) then fuel may be by-passing the injectors and blowing out into the dump can. In this case check the dump pipe outlet for oil flow and adjust the fuel pressure control valve to bring the pressures both to 40 p.s.i.g.

(d) If opening the fuel valve will not lower the pressure then either the fuel injector is blocked or else there may be emulsion in the fuel. Try switching the fuel to the alternate injectors. If this does not ease the trouble then switch back to the previous fuel tank and call for help. If the fuel shortage persists then eventually there will be a flame-out and the unit will shut down automatically. Initiate emergency shut down procedure.

(e) If the fuel pressure is low with rising bed temperature then the fuel tank may be empty. Check this and swap tanks if necessary.

2. Regenerator Temperature Rises Rapidly

The regenerator temperature actuates an alarm at 1,100°C and a nitrogen quench if valve is open. If it suddenly rises the chances are that the bed transfer system is inoperative.

(a) First check the Compton pressures, these should be about 3 p.s.i.g. If they are low then the Comptons will have stopped; switch to the alternative nitrogen supply.

(b) If on automatic control, check the rate at which the gasifier to regenerator pulse is working. If this is very rapid then it will be inoperative. Switch to manual and adjust to a reasonable rate, i.e. greater than 5/1, which brings the temperature back into line.

(c) If the temperature is excessive and must be reduced quickly, use the regenerator pressure control valve to swing the pressure differential between the gasifier and regenerator. This will rapidly exchange bed material and will temporarily bring the regenerator temperature down.

(d) If manual control does not work, and the gas supply is satisfactory then there may be a blockage in the gas injectors. Go down into the pit, check and rod out injectors, bleeds and transfer lines, if necessary.

3. Fines Return N_2 Pressure Rising

This indicates a blocked fines return pipe. Switch off the corresponding control panel and rod out the pipe.

4. Cyclones Not Draining Properly

This is indicated by cyclone drain temperature dropping steadily, and can be verified by a visual check through sight glasses in the top of the cyclone and on the box. Check valve operation; if not satisfactory then rectify faulty valve. If blocked, rod out. If the blockage appears to be in the pulser then turn off the control panel, and turn off N_2 to injector and shut off the $\frac{3}{4}$" Audco valve. Rod through the valve from below the puffer. As a last resort dismantle the puffer, clean out and replace.

5. Regenerator CO_2 Increases and SO_2 Decreases

This indicates that too much carbon is accumulating on the bed material. The remedy is to increase the air and flue gas supply to the gasifier to maintain temperature. Do not change the fuel rate because this is related to the stone feed rate.

6. Gas Space Pressure Increases

Prepare for planned shut down and decoke when the gas space pressure approaches 20" WG.

7. Automatic Shut Down

Initiate emergency shut down on gasification procedure.

8. Boiler Flue Gas Temperature Rises

If this temperature rises rapidly then the brickwork arch which seals against the boiler door will have developed a leak. Call for help when the temperature exceeds 500°F.

9. Back Up Pump on the Pressurisation System Runs Continuously

Call for help.

10. Main Flame Pilot Light Goes Out

This will actuate the alarm bell. Try to relight the pilot using 'start main flame pilot light (b)' procedure, and if unsuccessful call for help.

11. Failure of Bitumen Trailer Pump

An alarm bell will actuate in control room. The procedure for bringing a standby bitumen pump into operation is given in this manual. Do this immediately and arrange for trailer pump repair.

12. Nitrogen Supply Failure

An alarm bell will actuate in control room. The transfer system/fines return system/nitrogen bleeds will fail. The Comptons will automatically switch off. The regenerator temperature initially will rise rapidly and you may need to activate the N_2 quench, but eventually the regenerator temperature will fall when carbon has been removed and stone sulphated.

Transfer nitrogen bleeds to cylinder supply. Identify cause of fault and rectify immediately if possible. If it is not possible to remedy fault within five minutes initiate planned shut down without sulphation. Note the N_2 plenum purge will need to be coupled to N_2 cylinder supply.

13. Flue Gas Recycle Filter Problems

The pressure drop through the bag filter will be recorded on an hourly basis (see data sheet 4). Monitor this pressure difference and take action if:-

(a) pressure drop is zero.

(b) pressure drop is greater than 6" H_2O.

Condition (a) will occur if one or more of the bags come loose or are punctured. Isolate system and replace with spare bags. If in doubt call for help.

Condition (b) will occur if the bags are blocking. Try clearing them using the N_2 back flow rapper (several quick 4 second blips) located on the right hand side of the filter. Open drain valve and remove material deposited.

Any other problems with this system which cannot be dealt with immediately call for help.

14. Liquid Nitrogen System Failures

This system is fully automatic, and provided that it is used properly, and is filled in good time, should require no attention from CAFB shift personnel.

However, problems have occurred on almost every run, and the following notes should help to avoid calling out Air Products unnecessarily, or may keep us going until they arrive:

(a) Loading

1. See That Air Products Do Arrive As Scheduled

Sometimes they miss us out as we are a small outlet compared with Harwell. Call Air Products on if no delivery by noon. Usually a replacement delivery can be arranged within three hours of our call.

2. See That Air Products Do Deliver When They Arrive

Drivers have been known to go away if either their electrical offloading pump or our electric supply is faulty. If our supply is off check isolating switch on wall behind N_2 tank is on, and that fuses (in workshop - fuse box labelled) are O.K. Spare fuses should be alongside fuse box.

N.B. No fuses in outside box - it is only an isolator.

3. See That Air Products Load Properly

Proper procedure is to load via valve V_1 or V_2. If loading via V_1 the full trycock, V_4 <u>must</u> be vented to check no overfill. Always check this trycock after a delivery. Run until gas is vented. Proper procedure is to load at a moderate rate. Loading too fast may ice up lines and freeze valve spindles, resulting in valves which do not seat properly or may seize shut.

(b) Misuse of System

1. See That Other Site Users Have Not Made Mistakes

This is most likely during normal working hours. Errors which have been made include:-

- Loading Dewars via V_7 and V_1 - bad practice as it can freeze loading valves and Air Products have to waste time unfreezing them - not a way to get their co-operation on deliveries.

- Loading Dewars via V_{18}, which is correct, but closing V_{18} and V_3 afterwards - this loses tank pressure fast!

2. See That We Do Not Make Mistakes

If we overload the demand circuit then the vapouriser (vertical finned tubes) will fail to warm the nitrogen sufficiently and TCV1 will close, reducing supply, and our pressure will fall. Keep a watch on the finned tubes and clean off ice and/or reduce N_2 demand. (Note that Pilot Plant is on same circuit - are they using a lot/have they got a leak?)

3. See That We Do Not Damage Anything

When we are running, the N_2 system is stretched to its limit and many lines are heavily iced up. If you wish to remove ice to get at valves, or just to check what you are doing be careful what you use to knock off ice, and where you hit - there are six small copper lines to a set of relief valves and to the liquid level/pressure gauges behind the display panel, and these are, at best, silver soldered; some may even be soft soldered. They are all rather fragile to impact at -186°C, and a rupture of one of these joints could shut us down.

(c) Actions If Pressure At CAFB Is Low

1. Look for large leaks in CAFB and Pilot Plant.

2. Check liquid N_2 pressure.

   If Normal 150 p.s.i. - check vapouriser is not frozen over. If it is, remove ice, reduce CAFB demand on tank (check demand by rest of site by throttling valve and listening to change in noise. If this demand is high, something is probably faulty, since rest of site demand is usually one-seventh of CAFB on run). If no obvious over-demand TCV1 may be faulty. Check V_{16} is wide open. If it is and still low pressure at CAFB PCV2 may be faulty. CALL AIR PRODUCTS.

   If Low - check V_2, V_4 shut tight. V_3, V_{12}, V_{14}, V_{16} wide open. If still low pressure, PCV1 may be failed shut. Try warming it - if no improvement CALL AIR PRODUCTS and prepare for imminent gasifier shut down for many hours, i.e. bed sulphation, burn out and standby on combustion.

(d) Actions If Pressure At Tank Is High (Over 180 p.s.i.)

Check V_{13} is shut tight. If it is then PCV1 is failed in open position. Try warming it - if no response control tank pressure by throttling on V_3 - this will be tricky to do and will require a constant watch on tank pressure, but can be kept up indefinitely. CALL AIR PRODUCTS FOR ADVICE. If they insist on removal/servicing of PCV1 this need not shut us down provided tank pressure is raised to maximum recommended by Air Products before block valves V_3 and V_{12} are shut by Air Products to remove PCV1, but we must then reduce demand to a minimum as tank pressure will fall during servicing.

(e) Actions If Tank Pressure Is Very High (Over 250 p.s.i.)

This is approaching the bursting disc pressure limit, so reduce tank pressure to below 200 p.s.i. immediately using V_4. If the high pressure was reached suddenly then the liquid nitrogen will still be cool, and pressure will reduce quickly with little loss of tank contents. However, if the liquid nitrogen has warmed up to equilibrium with 250+ p.s.i. then a significant loss of tank contents will be necessary to reach a stable pressure below 200 p.s.i. IF THE TANK CONTENTS ARE ALREADY LOW CALL AIR PRODUCTS FOR AN EARLY DELIVERY. If not possible, watch situation closely and be prepared to shut down if nitrogen runs out. Look for faults as in (d) above.

(f) Actions If Disc Bursts Or Relief Valve (RV6) Lifts

If bursting disc blows, shut down gasifier, sulphate, burn out and go to standby on full combustion. CALL AIR PRODUCTS. If Air Products response is slow, try to rectify tank system. A spare bursting disc is kept in a glass fronted case on the wall behind the tank. Close V_3 and V_{12} while replacing disc.

If RV6 lifted, try to reseat after tank is below 200 p.s.i. RV6 may have frozen open, so warm up if necessary. Find fault as in (d) above and rectify.

PLANNED SHUT DOWN ON GASIFICATION

- Possible reasons:

    (a) Prior to decoke.

    (b) Gasifier maintenance, e.g. removal of lime from plenum.

    (c) Boiler Back End Clean Out.

Two alternatives:

(a) with sulphation.

(b) without sulphation.

- If the reason for the shut down necessitates an activity in which the bed is exposed to the open air, then it is necessary to sulphate the bed to avoid formation of CO gas in the pit.

(a) With Sulphation

- Reduce gasifier temperature to below 850°C by increasing flue gas recycle rate.

- Switch regenerator temperature control to manual - switch behind panel to 'off'.

- Increase pulse rate to reduce regenerator temperature below 1,000°C.

- CHECK auto valve on cooler is functioning and pressure in pressurisation unit is O.K. (If auto valve not working be prepared for loss of pressure in pressurisation unit, e.g. arrange for back-up (hand) pump.)

- Shut vent from boiler sample cyclone with plug.

- Disconnect sample lines to boiler SO_2 analyser, and clip off so that other analysers can monitor boiler flue without sucking air.

- Shut off auxiliary air bleeds to gasifier -

    (a) Air bleed to lid if present.

    (b) Air to preheat burner purge.

    (c) Switch fuel injectors from air to N_2, set at 2 CFM (open N_2 FIRST).

    (d) Air to stone feed.

- Close stone feed valve.

- Close N_2 to stone feed hopper.
- Switch off stone feed vibrator.
- Close needle valve on air meter manometer to prevent blowing.
- Open flue gas orifice by-pass valve.
- Switch automatic shut down to 'inoperative' and mute main flame and pilot flame alarms.

Supervisor

- Switch off fuel pumps.
- Close gasifier air motor valve.
- Open FGR motor valve to give 140 to 280 CFM as measured on gasifier air flow Anubar.

Man A

- As soon as M.F.F. light on, close throttles on main and premix burner air lines.

Man B

- As soon as M.F.F. light on, close fuel valves in control room, go to blower house and block off open air inlet.

- Control bed temperature to 900-950°C by regulating flow of air to main burner (use first primary air throttle). Oxygen concentration in flue gas recycle of $\sim$10 vol.% is required (measured on plenum air O_2 meter).
- Sulphation is complete ($\sim$ 20 minutes duration) when oxygen levels increase and bed temperatures falls.
- When complete -

  (a) Shut off gasifier blowers.

  (b) Close plenum valve.

  (c) Close FGR valve in pit.

  (d) Reduce regenerator air flow to 3 CFM.

  (e) Close first primary air throttle to main burner.

  (f) Switch off fines return system.

  (g) N_2 to fines injectors to 1 CFM each.

  (h) Open needle valve on air meter manometer.

(b) <u>Without Sulphation</u>

- Increase gasifier temperature to 930°C by shutting off flue gas recycle and/or reducing fuel supply to gasifier.
- Close FGR in pit.
- Shut vent from boiler sample line with plug.
- Disconnect sample lines to boiler SO_2 analyser, and clip off so that other analysers can monitor boiler flue without sucking air.
- Shut off auxiliary air bleeds to gasifier -
  - (a) Air bleed to lid if present.
  - (b) Air to preheat burner purge.
  - (c) Switch fuel injectors from air to N_2 set at 2 CFM (OPEN N_2 FIRST).
  - (d) Air to stone feed.
- Close stone feed valve.
- Close N_2 to stone feed hopper.
- Switch off stone feed vibrator.
- Mute main flame failure alarm.
- Switch off fuel pumps.
- Check blowers and pumps are off.
- Close regenerator blower outlet valve.
- Close plenum throttle.
- Close plenum valve tight.
- Open plenum throttle.
- Turn on plenum N_2 purge at 1 CFM.
- Switch off regenerator blower and fines return system.
- N_2 to fines returns to 1 CFM each.
- Close fuel valves.

<u>Emergency Shut Down on Gasification</u>

- When alarms ring and main flame failure shows:
  - Check blowers and pumps are off.
  - Close regenerator blower outlet valve and burner purge.
  - Close plenum throttle.
  - Close plenum valve tight.
  - Open plenum throttle.
  - Check plenum N_2 purge at 1 CFM.
  - Close air to stone feed.
  - Close stone feed valve.
  - Close N_2 to stone feed hopper.

- Switch off vibrator.
- Change fuel injector purges to N_2 and set at 2 CFM.
- Close fuel valves.
- Close off any air supply, e.g. to lid.
- Switch off regenerator blower and fines return systems.
- N_2 to fines return to 1 CFM each.
- Close FGR valve in pit.

The gasifier is now safe to leave slumped for at least one hour to investigate cause for shut down and rectify, or call for help. If the shut down is likely to last for more than 15 minutes purge through fuel oil lines with kerosene, or if on bitumen purge through on gas oil (see "Gas Oil Supply to Service Tank and Purge Lines").

Probable Causes for Shut Down

- Failure of 12V stabilised power supply to main board.
- Interruption of mains supply/fire eyes dirty/not connected/eye ports blocked.
- Pressurisation unit lost pressure - pump failure/leak/non-ret valves.
- Failure of fuel supply - tank empty/filters blocked/fire valves dropped/ water-oil emulsion.

Carbon Burn Out

- The procedure for carbon burn out is the same whether the shut down was planned (with or without sulphation) or emergency.
- Put plug on air inlet to gasifier blower.
- Put cap on outlet from sample cyclone.
- Close throttle on main and premix burner air lines.
- Restart gasifier blowers.
- Open FGR in pit.
- Close flue gas recycle by-pass valve.
- Open FGR motor valve.
- Open valves in gas line to gasifier lid and start small flow initially, building to a flow of 100 CFM flue gas, holding thermocouple in duct to below 1,000°C.
- Check main air blowers running, and adjust flow by opening premix throttle to give O_2 content of $\sim$7 vol.% through lid. (Measure in flue gas O_2 meter.)
- Monitor thermocouples downstream of bed and adjust air flow to maintain temperature below 1,000°C.
- Increase gas flow through lid to 150 CFM towards end of burn-out while still maintaining downstream temperatures below 1,000°C.
- Purge fuel lines to injectors with kerosene by opening drains or disconnecting lines in pit. Run kero into a keg and have a dry powder extinguisher at the ready. Close fuel valves and reconnect fuel injectors.

- Burn out is complete when CO_2 in boiler flue gas begins falling.
- During burn out monitor bed temperature (T2).
- If temperature falls below 700°C and bed is sulphated, change to kerosene combustion:-
  - Close off air/FGR to lid, and air bleed via burner premix throttle.
  - Close FGR motor valve.
  - Set one fuel pump to deliver 56 lbs./hr.
  - Check trace heating is on and hot.
  - Remove plug from air inlet to gasifier blowers.
  - Open plenum valve and adjust to 170 CFM by opening air motor valve.
  - Start fuel pump and open fuel valves to injectors.
  - Continue combustion till bed temperature rises to 850°C.
  - Maintain watch on downstream thermocouples while bed is being heated up.
  - At 850°C - shut off fuel pump and close fuel valve.
  - Shut off plenum valve and close air motor valve.
  - Open FGR motor valve.
  - Continue decoke from "Carbon Burn Out Procedure".
- If temperature falls to 700°C, and bed is sulphided:-
  - Close off air/FGR to lid, and air bleed via burner premix throttle.
  - Close FGR motor valve.
  - Remove plug from air inlet to gasifier blowers.
  - Open plenum valve.
  - Adjust air flow to 170 CFM using air motor valve.
  - Continue until temperature rises to 850°C.
  - Maintain a watch on downstream thermocouples.
  - At 850°C close plenum valve.
  - Open FGR motor valve.
  - Continue decoke from "Carbon Burn Out Procedure".
- If re-heat of sulphided bed fails to reach 850°C, and temperature starts to fall because bed carbon is depleted and sulphide oxidised to sulphate, resume under "Temperature fall during decoke with sulphated bed".

Completion of Carbon Burn Out

- When CO_2 level in flue gas starts to fall, indicating completion of burn out:-
  - Close air plus FGR to lid.

- Close FGR valve in pit.
- Close FGR motor valve.
- Remove plug from air inlet to gasifier blower.
- Remove plug from sample cyclone.
- Increase N_2 to fines return to 4 CFM each.

Restart After Shut Down

(a) From Sulphated Bed after Carbon Burn Out

- After "Completion of Carbon Burn Out" procedure:-
- Reconnect SO_2 analyser to sample system, zero and check calibration.
- Resume kerosene combustion using as a check list the procedure under "Boiler Clean Out" after the boiler door has been closed.
- Go to procedure for "Change from Kerosene Combustion to Fuel Oil Combustion", but omit final step of holding 870°C for one hour.
- Go to procedures for "Change Over from Combustion to Gasification" and "On Gasification".
- Resume test programme.

(b) From Sulphided Bed After Carbon Burn Out.

- After "Completion of Carbon Burn Out" procedure:-
- Reconnect SO_2 analyser to sample system, zero and check calibration.
- Check main burner pilot or re-establish, see "Start main flame pilot light".
- Activate alarm bells except main flame.
- Check cooling system levels/pressures/flows/bleed.
- Check FGR blower on.
- Check FGR valve in pit is closed, and close FGR motor valve.
- Check trace heating is on and hot.
- Verify operation of main flame detectors.
- Check boiler pilot alarm and verify visually.
- Verify all other alarms except main flame.
- Set high temperature alarm on gasifier to 1,000°C.
- Main flame alarm muted and auto shut down inoperative.
- Set fuel pumps to deliver 292 lbs./hr. total. Arrange flows such that equal volumes flow down each injector with all pumps on.
- Open plenum valve.
- Adjust air to 170 CFM.
- Quickly open FGR valve to 30 CFM.
- Start air to regenerator and adjust to 15 CFM.

- With sulphided bed allow gasifier temperature to rise to 850°C.
- At 850°C simultaneously switch on two pumps totalling 292 lbs./hr. and start stop watch.
- At 8 sec. reset main flame failure alarm (MFFA).
- If MFFA cancels switch off mute and arm auto shut down.
- Proceed to "On Gasification".
- At 9 sec. reset MFFA and repeat as above.
- At 10 sec. reset MFFA and repeat as above.
- At 11 sec. reset MFFA and repeat as above.
- If MFFA has not cancelled at 11 sec. switch off pumps immediately.
- At 21 sec. switch off main air blowers and proceed as for "Restart After Shut Down".
- Investigate causes for failure to ignite, rectify and repeat.

(c) From Sulphided Bed Without Carbon Burn Out

- If emergency shut down has been effected find the cause of shut down and rectify.
- Check fuel oil fire valves are open, and oil circulating pump is operating.
- Check red light is out on auxiliary panel.
- Check main burner pilot or re-establish, see "Start Main Flame Pilot Light".
- Activate alarm bells except main flame.
- Check cooling system levels/pressures/flows/bleed.
- Check FGR blower on.
- Set N_2 to fines return 4 CFM each.
- Check FGR valve in pit is closed, and close FGR motor valve.
- Check trace heating is on and hot.
- Verify operation of main flame detectors.
- Check boiler pilot alarm and verify visually.
- Verify all other alarms except main flame.
- Set high temperature alarm on gasifier to 1,000°C.
- Main flame alarm muted and auto shut down inoperative.
- Set L.H. and R.H. fuel pumps to deliver 292 lbs./hr. total. Arrange flows such that equal volumes flow down each injector.
- Open plenum valve.
- Adjust air to 170 CFM.
- Quickly open FGR valve to 30 CFM.
- Start air to regenerator and adjust to 15 CFM.
- Allow gasifier temperature to rise to 850°C.

- At 850°C simultaneously switch on two pumps totalling 292 lbs./hr. and start stop watch.
- At 8 sec. reset main flame failure alarm (MFFA).
- If MFFA cancels switch off mute and arm auto shut down.
- Proceed to "On Gasification".
- At 9 sec. reset MFFA and repeat as above.
- At 10 sec. reset MFFA and repeat as above.
- At 11 sec. reset MFFA and repeat as above.
- If MFFA has not cancelled at 11 sec. switch off two larger pumps immediately.
- At 21 sec. switch off main air blowers and proceed as for "Restart After Shut Down".
- Investigate causes for failure to ignite, rectify and repeat.

Changeover From Heavy Fuel Oil to Bitumen

It is assumed that one Right Hand lower side and one Left Hand lower side injector are in operation on heavy fuel oil supplied equally by two metering pumps (refer to Figure 8). The change to bitumen should be carried out on one injector at a time. The bitumen system is assumed flushed with gas oil.

- Note and record fuel oil delivery values on both pumps.
- Set main burner air to give 5% O_2 in flue gas.
- Raise gasifier high temperature alarm to 1,000°C.
- Ensure gasifier temperature not lower than 890°C.
- Select to change over to R.H. injector first (for instance).
- Reduce heavy fuel supply to this injector by 35 lbs./hr. (approximately 10% total fuel) - reduce on one pump only (hereafter called pump A).
- Set R.H. bitumen (Plenty) pump to meter 70 lbs./hr. bitumen.
- Line up valves to inject bitumen to R.H. injector down to final bitumen to injector valve (B36).
- Open bitumen valve (B36) and quickly switch on R.H. bitumen pump.
- Monitor gasifier temperature and flue gas O_2 levels - these will decrease and level out at new value.
- Hold for 5 minutes to ensure bitumen is through to gasifier (N.B. an increase in gasifier temperature after the decrease means a bitumen blockage).
- Back off heavy fuel oil on pump A by 70 lbs./hr.
- Raise bitumen flow to 140 lbs./hr. and then switch off pump A and close valves at injector and in control room.

- Increase R.H. bitumen pump delivery to level originally metered by pump A.
- Repeat same procedure for second injector.
- Monitor flue gas oxygen and adjust if necessary to give stable flame.
- Purge fuel oil pumps and lines with kerosene. Have dry powder fire extinguisher to hand during purge procedure, and avoid spillages in pit.

Changeover From Bitumen to Heavy Fuel Oil

The procedure is similar to the changeover from heavy fuel oil to bitumen except that after the bitumen fuel valves have been closed the bitumen line must be flushed with gas oil (see "Gas Oil Supply to Service Tank and Purge Lines"). Have fire extinguisher to hand during purge procedure and avoid spillages in pit.

Sampling of Hot Bed Material

- Wearing of asbestos gloves is recommended for this operation.
- Ensure that both bleeds are at 2.5 CFH.
- Open sample pot exit valve and drain leg valve, and drain bed material into 28 lb. keg until red hot bed material appears - rod out drain leg if necessary.
- Close exit valve and fill sample pot with hot bed material ($\sim$ 4 lbs.) observe through sight glass.
- Leave N_2 purge and leave sample to cool ($\sim$ 30 mins.).

Rodding Out

- Wearing of asbestos gloves and a face-mask is necessary, and in some instances, for example rodding from the top of the cyclones, full breathing apparatus should be used.
- Turn on N_2 to rodding lance and set at high flow rate. N.B. too high a N_2 flow may trip the safety switch on the Compton compressors.
- Remove plug, insert and position N_2 lance in rodding port before opening valve.
- After rodding withdraw lance beyond valve, but not fully out and close valve.
- Care should be taken when rodding fuel injectors, fines return pipes or manometer tappings where a significant positive gas pressure exists on the other side of the blockage. A cloth wrapped (like a scarf) round the neck inside the overalls will save neck burns. In the case of regenerator tappings remember the gas stream will contain 8-10 vol.% SO_2.

Injection of Toxic Gases

This will only be carried out by a qualified person fully familiar with toxicity of handled materials, and with safe handling and test procedures. For Run 10, Z. Kowszun is designated as sole person qualifying.

Injection shall take place only when another shift supervisor is present and observing the following precautions.

(a) Both sliding doors to gasifier area open.

(b) Access of personnel not involved in test to gasifier area must be prevented by posting sentries.

(c) Full breathing apparatus should be worn by tester and back-up man.

(d) All equipment should be leak tested before use.

(e) On the spot analysis for toxic materials of atmosphere in main pilot plant area and control room must be carried out throughout test period with a Drager.

(f) Samples and any necessary readings outside the control room will be taken by the nominated person only.

Preventive Maintenance and Sampling During Run

The following action items should be carried out on a regular basis throughout the run as indicated below.

- Blow out pressure tapping lines and zero manometers (every shift).
- Zero and calibrate analytical instruments/change filters (day shift only).
- Rod out flue gas sample line - change filter on flue sample line and regenerator sample line, and empty knock-out vessel (every shift).
- Clean fire eyes on boiler (every shift)
  - Check if red light is on at boiler control panel, showing rear door fire eye is seeing a flame. If light is on, remove and clean front fire eye first, then clean rear door fire eye. If light is off, clean rear door fire eye first, then front fire eye.

    Always check red light is on at boiler panel after cleaning rear door fire eye.

    Failure to follow this sequence for cleaning fire eyes may result in an automatic shut down.
- Oil dust extractor fan bearings located in roof and grease boiler water cooling pump, and boiler water circulating pump (2 grease points each) (every shift).
- Check blowers for belt wear (a) blower house (2)
(b) regenerator (3)
(c) cooling fan on roof
(hourly - see data sheet 4)
- FGR filter maintenance.
  - At least once a shift or more often if pressure drop on FGR filter gauge approaches 6" H_2O, clean bags by activating N_2 back flow rapper (several quick 4 seconds blips). The valve is located on the right hand side of the filter unit. Open drain valve at bottom of filter and remove deposited material.
- Every six hours take gasifier and regenerator bed samples and gasifier cyclone (L.H.).
- At 12.00 and 24.00 take gasifier, regenerator, regenerator cyclone, boiler back and boiler flue, gasifier cyclones (2) samples.
- On bitumen trailer unit:-

  Compressor

  - Examine compressor/motor/hydraulic pump drive belts for deterioration WITH THE MOTOR ISOLATING SWITCH OFF NOT LONGER THAN 2 MINUTES (day shift only).
  - Maintain compressor lube oil level no higher than top level of the dipstick (day shift only).
  - Check that all pipe joints are tight and do not leak air (every shift).

  Lubricants

  Hydraulic system - Teresso 43

  Rotary vane compressors - Estor HD 30

  Petter engine on standby pump - Esso 20W/50

APPENDIX E

ANALYSIS OF LIMESTONE AND FUELS : RUN 10

Limestone

The USA (ex Grove Limestone Company, Stephens City, Virgina) limestone BCR 1359 was used throughout Run 10. Inspections of samples taken during the run are given in Table E1.

Heavy Fuel Oil

The Heavy Fuel Oil used throughout Run 10 was a single batch produced from Venezuelan crudes at Amuay refinery. Inspections of samples taken during Run 10 are given in Table E2.

Residuum

The residuum used was produced at Fawley refinery from TJ 102 Medium crude. It is the vacuum pipestill bottoms produced when normal atmospheric pipestill bottoms are further distilled under vacuum, Table E3.

Illinois No.6 Coal

The very limited test work at the end of Run 10 on coal injection and gasification was conducted entirely on Illinois No.6 coal. Specific quality checks were not conducted at the time on each size distribution used. The typical inspection of the material from which the samples were taken is given in Table E4.

APPENDIX E

TABLE 1

ANALYSIS OF LIMESTONE, BCR 1359 USED FOR RUN 10

Sample Number Time (D.H.)	50750 2.0600	50828 6.1200	50887 13.1200	50988 18.1200	51018 19.1200	Mean
CaO wt %	58.2	57.8	59.9	58.7	58.7	58.7
MgO wt %	0.53	0.50	0.48	0.51	0.51	0.51
SiO_2 wt %	0.83	0.86	1.08	1.30	0.86	0.99
Al_2O_3 wt %	0.26	0.24	0.24	0.34	0.27	0.27
CO_2 wt %	40.8	38.5	39.7	39.8	38.5	39.5
Fe_2O_3 ppm	529	410	485	862	615	580
Na_2O ppm	77	101	99	92	92	92
S wt %	<0.04	<0.04	<0.04	0.29	<0.04	-
V ppm	35	68	34	76	134	69
Ni ppm	34	39	23	47	38	36

APPENDIX E

TABLE 2

ANALYSIS OF FUEL OIL USED FOR RUN 10

Sample Number Time (D.H.)	50762 2.0600	50789 6.0600	50852 12.2000	50898 13.0000	50932 14.0000	50987 16.1800	Mean
C wt %	85.45	85.95	85.38	85.66	85.33	85.54	85.55
H wt %	10.90	10.87	10.92	10.77	10.77	10.82	10.84
S wt %	2.48	2.55	2.58	2.59	2.60	2.58	2.56
V ppm	-	-	-	-	375	340	338
Na ppm	-	-	-	-	47	47	47
Fe ppm	-	-	-	-	24	30	27
Ni ppm	-	-	-	-	41	43	42
SG at 60°F	0.9567	0.9595	0.9583	0.9591	0.9591	0.9595	0.9687
Conradson Carbon	8.47	10.31	9.89	10.83	10.58	10.00	10.00
Asphaltenes	6.27	6.23	6.76	6.28	6.61	6.14	6.38
Water %	Nil	Trace	Nil	Nil	Nil	Nil	Nil

APPENDIX E

TABLE 3

ANALYSIS OF TJ 102 MEDIUM VACUUM BOTTOMS

Sample Number	50912	50713	Mean
C wt %	85.75	85.91	85.83
H wt %	10.51	10.40	10.46
S wt %	3.18	3.22	3.20
N wt %	0.55	0.55	0.55
V ppm	630	625	628
Ni ppm	78	78	78
SG	1.0134	1.0146	1.0140
Conradson Carbon	7.59	9.79	8.69
Asphaltenes	9.83	10.00	9.92

APPENDIX E

TABLE 4

ANALYSIS OF ILLINOIS No. 6 COAL

	Illinois No. 6
Particle Size Distribution	up to 0.32 cm (1/8 inch)
Carbon wt %	67.70
Hydrogen wt %	4.62
Sulphur wt %	2.52
Nitrogen wt %	1.59
Oxygen wt %	7.92
Ash wt %	11.39
Moisture wt %	4.26

APPENDIX F

MODIFICATIONS TO THE CAFB PILOT UNIT IN PREPARATION FOR COAL GASIFICATION TESTS

Summary

A number of modifications were made to the pilot unit arising from either improvements necessary based on experience during Run 10, in anticipation of the planned test programme on coal gasification, or as a direct request from Foster Wheeler Energy Corporation to test equipment to assist in the design of a 20 MW demonstration unit at San Benito, Texas. The major areas where modifications were made were:-

a) A new cyclone drain and fines returns was installed.

b) A new lime/coal feed system was designed and built.

c) A number of coal injection needles were provided for testing.

d) Provision was made for flue gas recycle via tuyeres and the flue gas recycle was redesigned.

e) A gas oil injection system was installed for regenerator warm-up.

f) New instrumentation for gas analysis was installed.

g) A new gasifier air blower was fitted.

h) Detailed changes were made to the fuel oil, pressure monitoring, solids transfer, nitrogen and compressed air systems.

Details of these changes are provided below.

a) Cyclone Drain and Fines Returns System

During Run 10, major difficulties were experienced in operating the cyclone drain and fines returns systems. Whilst improvements in design were indentified which undoubtedly would have eleminated some of the problems, these were not implemented since it was reqested that a system be tested to assist in the design of the Foster Wheeler Energy Corporation 20 MW demonstration unit at San Benito in Texas. The design basis for the new system is summarised below.

During Run 10, all the fines were re-injected into the gasifier. For operations on coal, a new problem of ash build up as the system appears and the fines return system had to be capable of fractionating the fines, re-injecting the coarse fraction and rejecting the fine, ash fraction.

A second major requirement was that the cyclone drain and thus the fines pick up should be below the gasifier distributor so that a lift system would be required. This consideration was a direct consequence of limitations of space at the San Benito test site.

A final consideration was that large chunks of material liable to break loose from the cyclone walls during burn outs could be removed from the system.

With these major considerations in mind, a valveless system was designed - see Figure 1.

The fines collected in the two main gasifier cyclones drain via a Y-shaped duct into a hopper where the contents could be fluidised with pulsed nitrogen supply through radially arranged nozzles. The fluidised solids could thus flow over a weir into an eductor, the rate being controlled by the frequency and duration of the nitrogen pulse. From the eductor, the solids are circulated to two cyclones in series on a recirculating air stream. The cyclones fractionate the solids, the first cyclone being relatively inefficient so as to remove only the coarse fraction and the second cyclone being highly efficient to remove the fine fraction from the system. The coarse fraction is re-injected into the gasifier bed.

The operation of the hopper fluidising system is triggered by a temperature sensor in the cyclone drain leg which activates the pulsers when the solids levels in the legs increases. Conversely, when the solids levels drops below the temperature senor, the pulsers are switched off. A solids seal can thus be maintained between the cyclones and the fines eductor/re-injection systems.

Because of the configuration of the hopper, large chunks of material could be expected to remain in the bottom and not be fluidised. They could then be removed via a discharge valve periodically.

Provisions were incorporated around the system to cool the gas stream prior to recirculation to the blower, and to bleed nitrogen into the gas stream if required for temperature control when the fines include hot fine carbon particles.

b) Coal and Lime Feed System

Extensive modifications were carried out on the lime feed equipment in order to increase its capacity to utilise it as a coal feed system. A schematic of the new arrangement is shown in Figure 2.

The system is fully automotic with control being via two micro-switches incorporated in a modified chart recorder used also to measure the solids feed rate. The chart recorder scale was directly matched to the weigh cell so that the micro-switches when triggered by the movement of the recorder pen could be directly set to correspond to suitable upper and lower levels of solids in the weigh hopper.

Thus, when coal or lime was fed from the weigh cell into the gasifier, the decreasing level eventually triggered the low level micro swich. This caused valve V3 to open whilst V1 and V2 closed. Flow restrictors in the compressed air lines delayed the opening of V3 until V1 and V2 were closed in order to prevent depressurisation of the weigh hopper through the lock hopper. With V3 open, coal or lime could fall into the weigh hopper until the high level micro-switch was triggered when V3 closed and V1 and V2 opened, again with a delay on V1 and V2 opening to limit pressure surges.

The closure of V3 triggered off a third micro-swich controlling the vacuum lift switch which then replenished the top lock hopper until switched off by a high level detector. Thus system ensured that the vacuum lift system could not be actuated with V3 open.

Time controls were included in the system so that alarms were activated if the lock hopper high level was not triggered in a certain time (showing the dispenser required refilling), if the solids replenishment of the weigh hopper took too long (V3 plugged or empty lock hopper), or if V3 failed to close within a specified time. All of these times can be adjusted manually for convenience within limits.

Some detailed modifications to this system were also carried out. The injection system below the weigh hopper and metering vibrating table was redesigned in order to accommodate injection needles of different materials penetrating the gasifier bed for tests on their suitability for incorporation in the design of the San Benito demonstration unit. Tests were planned for two silicon carbide needles (self bonded, and recrystallised types) and two stainless steel needles protected with either hard or soft refractories. As a consequence, the purge and bleed systems for air and nitrogen around the solids feed system were modified substantially.

Flue Gas Recycle System

The bag filter and housing employed during Run 10 was replaced with a high efficiency cyclone and the dust loading of the flue gas was reduced by locating the take off point downstream of the main stack cyclone and knock out vessels. The recycled flue gas was fed directly into the fluidised bed via four 1 11/16 inch diameter Firebird Blue stainless steel tubes.

Analytical System

A completely new analytical system for the gas streams was installed, only the Wosthoff SO_2 analyser being retained from the existing equipment. The new equipment was supplied by Hartmann and Braun and comprised the necessary gas cooling and pumping components and the infra.red analysers.

An additional sampeling point for the stack SO_2 gas was installed to provide total SO_2 level (regenerator + boiler emission) as a back up for calculating sulphur balance to that provided by the separate boiler and regenerator samples. This was installed well above the regenerator gas inlet to the stack so that a representative level of SO_2 could be determined.

A schematic of the new system is shown in Fig.3.

Gas Oil Burner for Regenerator

Occasional difficulties are experienced in reheating the regenerator bed following prolonged shutdown. To overcome this problem a gas oil metering and delivery system was linked to a removeable injector inserted through the lower regenerator pressure tapping to inject up to 150 ml/hour gas oil when a fast reheat was required.

Bed Transfer System

The rodding out ports into the entry boxes in the two reactors were enlarged and the horizontal pulser injectors were calibrated so that their penetration could be adjusted and measured accurately.

Fuel Oil Injection system

The side injectors used during Run 10 were removed and replaced with flue gas recycle entry tuyeres. Thus, only the two horizontal injectors onto the distributor pit were available for fuel oil injection and the pipe work and valving arrangements could be simplified.

Pressure Monitoring system

Manometers used to monitor pressures around the gasifier and regenerator were replaced by differential pressure gauges.

Gasifier Fluidising Air Blower

A new Godfrey positive displacement blower was installed to provide fluidising air for the gasifier.

Minor Modifications

A large number of minor modifications were made to the nitrogen and air supply system to incorporate the additional needs of the new fines returns system and the modified limestone/coal feed system, solids transfer system and flue gas recycle system.

The kerosine delivery system to the gasifier was stripped out as it was no longer required for the operation of the unit.

Thermocouples and pressure tappings were refurbished and replaced where necessary.

A large number of items were inspected, and overhauled as a matter of routine. The major item investigated was the limestone/coal feed metering vibrating table which had proved to be so troublesome during Run 10. Whilst no mechanical or electrical deficiency in the mechanism could be found, a possible cause of its poor performance was thought to be due to dust accumulating around and under the vibrator itself. Steps were taken to improve this by inserting a pad of foam rubber under the vibrator to prevent ingress of dust.

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

1. REPORT NO. EPA-600/7-79-066	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil--Fourth Phase		5. REPORT DATE February 1979
		6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) A.W. Ramsden and Z. Kowszun		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Esso Research Centre Abingdon, Oxfordshire OX136AE England		10. PROGRAM ELEMENT NO. 1AB013; ROAP 21ADD-BE
		11. CONTRACT/GRANT NO. 68-02-1479
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Final; 6/74 - 12/76
		14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES IERL-RTP project officer is Samuel L. Rakes, MD-61, 919/541-2825. EPA-600/2-76-248 and EPA-650/2-74-109 are earlier related reports.

16. ABSTRACT The report gives results of Phase 4 of a study on the CAFB process for gasification/desulfurization of liquid and solid fuels in a bed of hot lime. A new pilot unit was designed and constructed, incorporating such novel features as: a new fluidizing air distributor, high-flow/low-pressure-drop cyclones, and improved refractory construction. Conclusions include: (1) confirmation of the process description by a statistically derived equation; (2) bed age has a significant effect on desulfurization efficiency; (3) heavy residua can be gasified and desulfurized; (4) solid fuels show potential as feedstocks; (5) trace element retention depends on stone replenishment rate (a rate sufficient to maintain bed depth--0.2 molar--is adequate to maintain trace element capture and sulfur removal performance); (6) satisfactory performance of the redesigned pilot unit; (7) a carbon burn-back procedure was not completely effective in cleaning the cyclones; (8) confirmation that coal and Texas lignite are suitable feedstocks for the CAFB and that ash accumulation and fusion do not appear to be limiting; and (9) Texas limestone is satisfactory as bed material if it is available in a suitable particle size range. Remaining work includes performance tests and evaluation.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Fuel Oil	Pollution Control	13B	11H,21D
Fluidized Bed Processors		Stationary Sources	13I,07A	
Gasification	Residual Oils	CAFB Process	13H	
Limestone	Heavy Oils	Chemically Active Fluid Bed	08G	
Desulfurization	Coal		07D	
Trace Elements	Lignite	Fluidized Lime Bed	06A,06P	11B
Design	Refractories			

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 340
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (9-73)