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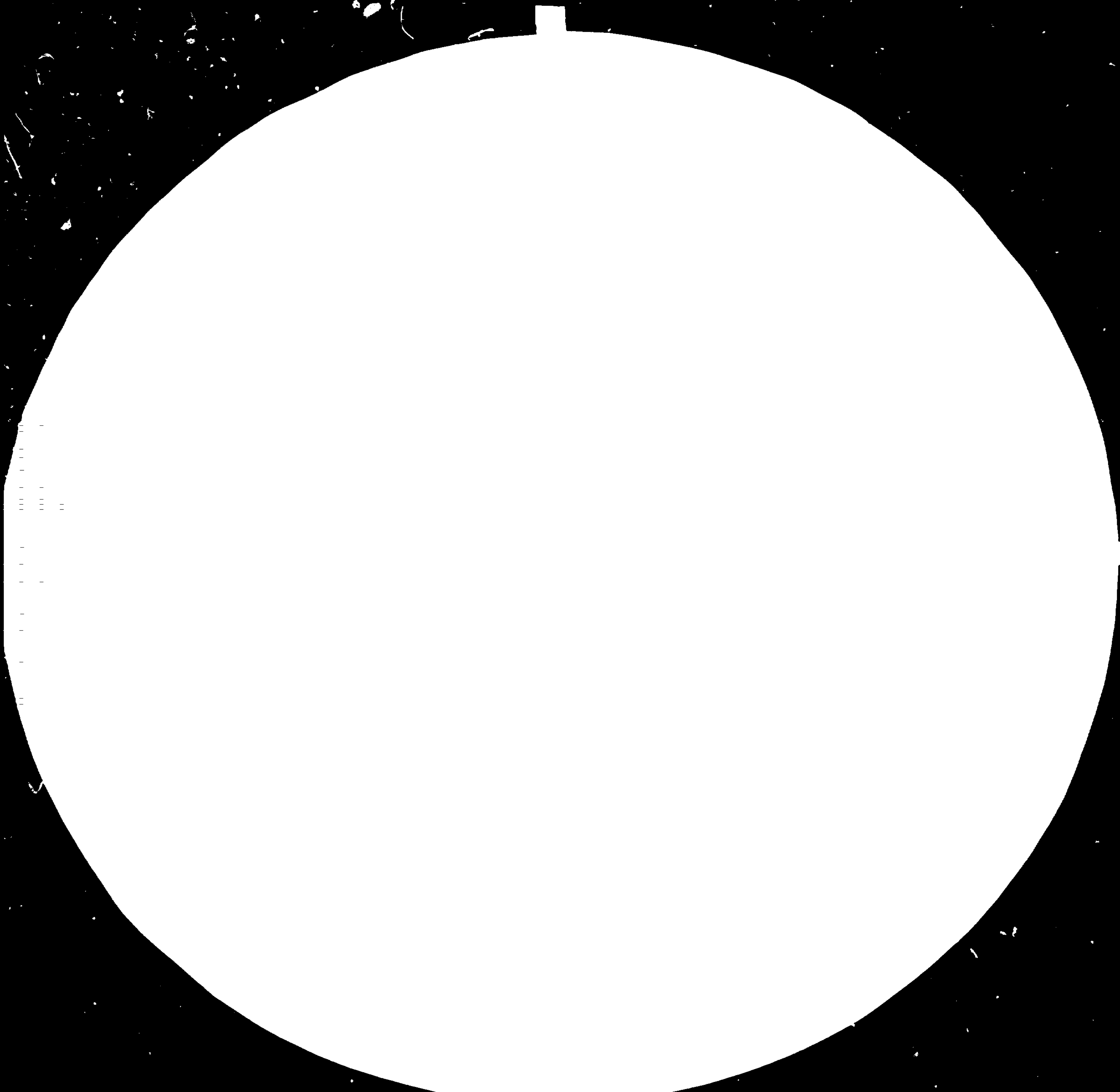

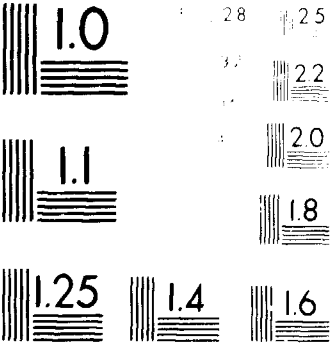

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NATIONAL BUREAU OF STANDARDS
STANDARD REFERENCE MATERIAL 1010a
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14352

RESTRICTED

December 1984
ENGLISH

→ Pendleton

COAL GASIFICATION.
DP/IND/80/004
INDIA.

Technical Report *

Mission 19 August to 3 September 1984

Prepared for the Government of India
by the United Nations Industrial Development Organization,
acting as executing agency for United Nations Development Programme

Based on the work of Brian Locke,
Consultant in Design and Operation of Large-Scale Plants

United Nations Industrial Development Organization
Vienna

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EXECUTIVE SUMMARY

The "Chemicals from Coal" project at the Regional Research Laboratory, Hyderabad, (RRL), has two broad objectives - to extend and improve the usage of Indian coals; and to produce chemicals and chemical feedstocks that would otherwise be made from petroleum. The two processes central to the project are gasification of coal to make a "synthesis gas" that is then purified and suitably adjusted in chemical composition: and catalytic synthesis from the gas, to make reaction products that can be processed to derive the chemicals and feedstocks required.

Of the available approaches to coal gasification, the Lurgi process had been chosen, and plant built and commissioned. Work had also begun on developing a catalyst for the synthesis. Ancillary processeses will be required and can be studied later on when sufficient progress has been made with the gasification and synthesis. The development of ancillary processes should cause no undue difficulties.

The two main components of the project follow on logically from the general coal processing and chemical programmes of RRL work thoughout several decades. In fact the gasifier was actually bought and delivered some twenty years ago, but owing then to shortage of resources was only installed recently. The chosen route is a good one, and the able teams of staff are well suited to handle the development.

The Consultant has discussed the plant and programmes with the teams concerned and has delivered lectures on important aspects, which themselves stimulated questions and disussion. He has made recommendations on both principles and details of the work, including means of putting them into practice.

They include matters such as safety, thermodynamics, chemistry, gasification engineering, the chemistry of catalysis, maintenance, plant layout, experimentation, recording, economic evaluation, programme coordination, university research, Fellowships abroad, and management succession at RRL. The recommendations have all been discussed with RRL staff and accepted with interest by them.

The need for technico-economics appraisals, and also for for industry/market studies, were discussed with the Director of RRL, Dr G Thyagarajan, who was enthusiastic about setting up a group to carry out such work. This can stimulate essential interplay between experimental, plant design, process selection, process costs, and marketing/commercial issues. It will provide useful bases from which RRL can present its work in the wider national context, and by which overall coal and chemical industry issues can figure in RRL planning. A technico-economics/market-appraisal group could begin with perhaps a chemist, an engineer, and an economist, with close contact with process industry and contractors or consulting engineers. Not only would it be useful in suggesting preferred economic directions of work, it would also help diffuse economic reality.

The combined facility at the Regional Research Laboratory at Hyderabad is probably the only one of its kind in the world. Accordingly it would be worth considering catering not only for India's needs, but also, where such work could be suitably scheduled, for the needs of other countries on a commercial revenue-earning consultancy basis.

There will be advantage in providing continuing guidance by Consultancy, to help RRL in coordinating the varied and interlocking components of the project. This way the best use could be made of the overall programme which should make a valuable contribution to the development of the coal and chemical industries in India.

CONTENTS

ABSTRACT

The pilot Lurgi high-pressure coal gasification plant had begun operations satisfactorily. It should become the basis for valuable work leading to the development of Indian processes for making chemicals synthetically from coal. The experimental work on synthesis and other process requirements will need to be kept in phase with the progress in gasification. An overall programme of experimental and technico-economic appraisal will need to be drawn up.

Discussions were held on relevant aspects of the work, and recommendations were made in respect of safety, gasification and synthesis, maintenance, instrumentation, experimentation, economic evaluation, programme planning, Fellowships abroad, further Consultancy and related issues.

The combined facility at the Regional Research Laboratory at Hyderabad is probably the only one of its kind in the world. Accordingly it would be worth considering catering not only for India's need, but also where such work could be suitably scheduled, for the needs of other countries on a commercial revenue-earning consultancy basis.

BACKGROUND

There have previously been Consultancy visits to RRL including those by

H. Jüngten - coal pyrolysis and gasification - February 1983

Lurgi staff - commissioning the pilot gasifier installation - October - December 1983

Alfred Sulima - commissioning a thermogravimetric analysis apparatus - November/December 1983

Karl H. Van Heek - UNIDO Chief Technical Adviser - supervisory visit - February 1984

E. Klose - gasification modelling - March 1984

B. Konrad - coal gasification - February/April 1984

Morton H. Dean - modelling of coal gasification reactions - June 1984

This UNIDO Consultancy has been undertaken to advise on both the component parts of, and also on the whole of the "Chemicals from Coal" project - and in particular to strengthen the capabilities of RRL for research and development work on high-pressure coal gasification.

OUTLINE OF THE CONSULTANCY

The Regional Research Laboratory at Hyderabad (RRL) is a substantial organization of some thirty years' standing, operating in advanced science and technology inter alia, in chemistry, coal processing, chemical engineering and pharmaceuticals.

The purpose of this consultancy was to strengthen the capabilities of RRL for research and development on coal gasification for chemicals production. RRL had made significant developments in processing coal at ambient pressure, and in other fields had operated small equipment at elevated pressures. However the Lurgi gasification pilot plant was large and complex in terms of high-pressure plant for a research organization, and also involved the use of oxygen, with its potential hazards.

The staff at RRL are of high quality, and some had acquired experience in other countries including work at commercial Lurgi gasification plants. Accordingly most of the necessary expertise was available, and indeed the whole concept of the work was not only sound, but fitted in well to the general scheme of work at RRL.

The purpose of this report is not to provide a view of the gasification and related work from the "outside", but to outline what transpired during the consultancy. This took the form of discussions with teams of RRL staff working on different aspects of subjects that, together, constituted the "chemical synthesis from coal" thinking. Therewere also lectures reviewing the main areas of interest to RRL.

Accounts of the discussions and lectures appear as Appendices. They contain both observations and recommendations by the Consultant, along with summaries of the work, and accounts of the plant and of activities, provided by RRL staff.

Some previous visits by UNIDO Consultants had been concerned with particular issues: this Consultancy considered the work as a whole, and how one section might best be fitted in with another.

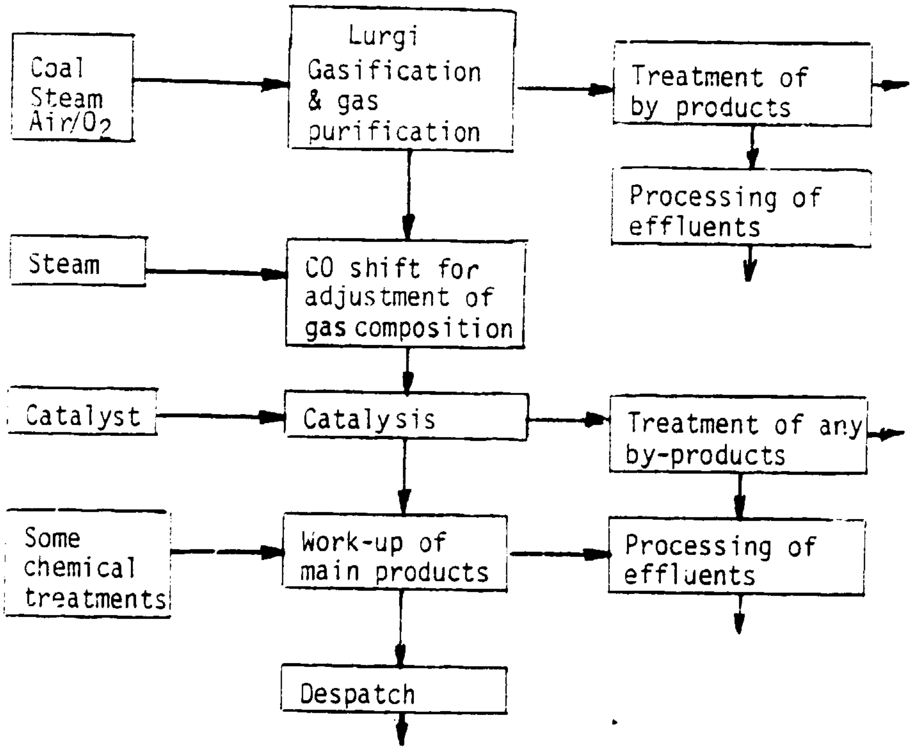
It also included examination of certain details that RRL staff felt to be important to them, and considered a wide range of issues, both of principle and as minutiae that arose during the disucussions. The subjects ranged from the chemistry of catalysed synthesis reactions, to techniques of gasifier operation, coal preparation, plant layout, process evaluation, and economic analysis.

The recommendations made by the Consultant should enable the work to progress in a satisfactory and ordered manner. This way it can best serve the main aim of the programme - which is to develop processes for synthesis chemicals, using coal as the raw material. The recommendations were all discussed with the appropriate RRL staff, and were well received. The staff were all very cooperative to secure the most useful outcome, and seemed dedicated to the project.

THE OVERALL "CHEMICALS FROM COAL" SCHEME

The pilot Lurgi high-pressure coal gasification plant is the first, and essential, stage in the overall route from coal to chemicals. Gas made in the gasifier will require adjustment of composition, catalytic synthesis, separation of any by-products, work-up of products to meet market specifications, and arrangements for despatch of products.

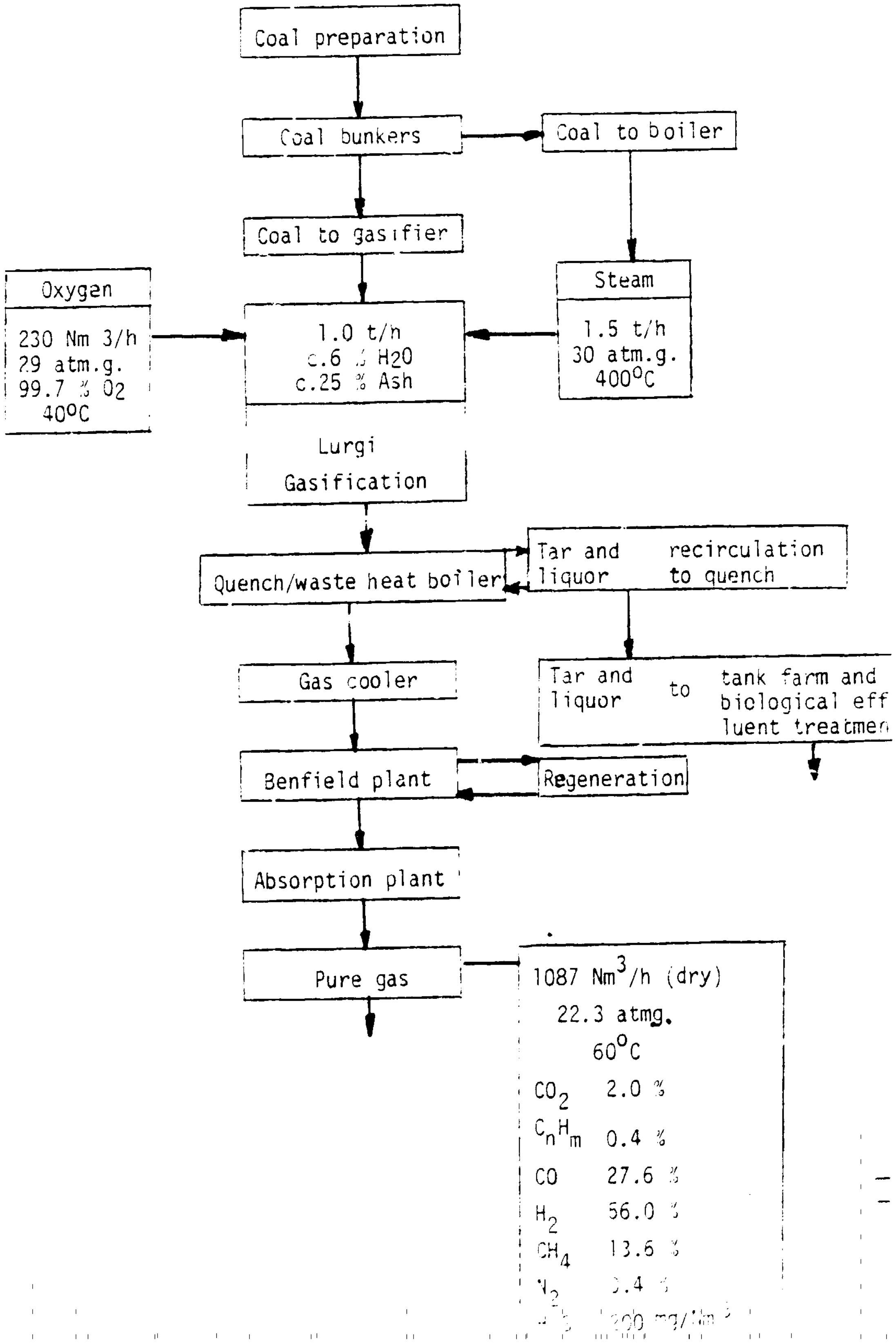
Following gasification, the processes needed may be based upon RRL (or other Indian) work, or brought in from other sources according to cost and timing. The general layout of a combined plant will be along the lines shown below.



LURGI COAL GASIFICATION

Simplified Process Scheme

(ex Lurgi - RRL Drg No. H-4-059)



THE LURGI GASIFIER INSTALLATION

CHRONOLOGY

Pre 1965 the decision was made to work on Lurgi gasification of Indian coals.

1965 The oxygen plant and the boiler were purchased from Lurgi.

1966 The Lurgi gasifier and ancillaries were received. The equipment was not installed, but was all left in its packing cases.

1975 RRL was"given a signal"to prepare for erection of the plant.

1977 Erection was begun by RRL, including various Indian components: design, fabrication, piping, testing etc. took place sequentially. The Benfield plant was fabricated in India from Lurgi drawings. The boiler and the oxygen plant were commissioned as also the coal preparation plant. Despite limitations in staff and facilities, all the work was done in-house. Precommissioning and pressure testing necessitated relapping the ash lock-hopper cone valve seating, and making rubber seals for the coal lock-hopper valves locally.

1981 The UNDP project began with assistance for instrument purchases; placing two staff in CSSR, one month each for training. In 1983 Mr Seshagiri Rao visited Usti gas works (CSSR) (14 gasifiers on stream). Then another man visited Resova (26 gasifiers).

1983 October - December - Lurgi staff commissioned the gasifier and the tank farm - but neither the Benfield plant nor the biological effluent oxidation plant.

1984 20 January - RRL operated the plant for two days to co-ordinate with visit of Dr May, Dr Maung and Dr van Heek, the chief Technical Officer of UNIDO. In March work began for calibration of the mass spectometers and the computer : also the Benfield plant was commissioned. The mass spectometer was commissioned by the makers. Gasifier operation was confined to air/steam reaction owing to low electricity voltages. (A 350 hp motor driving the oxygen plant air compressor fractured a coil owing to low voltage. It is now rewound and working).

17/8/84 the O_2 plant was commissioned at 15 bar at 80% and 90% of full load: and then at 20 bar and 90% of full full load.

23/8/84 H_2O/O_2 operation of the gasifier began.

24-27/8/84 the gasifier was operated on H_2O/O_2 at 20 bar, three shifts per day.

28/8/84 a power failure necessitated shutting down.

The gasifier was to start again from 1/9/84 to run for one week at 24 bar with oxygen/steam reaction.

(Note - power supply breakdowns from the public supply system are common, especially on Wednesdays when the steelworks power demand peaks. The supply authorities and other users are obliging, especially as regards rolling mill operation that uses the same supply; but there are still failures).

The biological oxidation plant is scheduled to be commissioned before the end of 1984

(Details supplied by RRL).

RECOMMENDATIONS

In general terms here - the relevant Appendices contain the details

General

The gasification and synthesis recommendations are made in relation to the stages of development and complexity of the various components of the work. For example, as the synthesis work expands, more detailed recommendations will be able to be made, analogous perhaps to those for gasification.

Recommendations have been made as to those aspects of both sections of work that require attention immediately, and those that apply particularly over the next year or so.

As experience is gained then programmes of experimentation can be made more formal and more detailed. Then, the overall project can be considered in the light of its purpose, especially on the basis of comparative economic evaluation.

Gasification

1) Safety, emergency shut-down, and planned maintenance procedures need to be instituted on formal bases. (Appendices 7,8)

2) Serious consideration should be given to improving the security, and maintaining the voltage of the electric power supply - whether as an in-house generator or as a dedicated section of the local supply system. (See Chronology).

3) Instrumentation should be provided for complete mass and heat balances , and positioning of the reaction zone within the gasifier. (Appendices 7,10)

4) A system of formal records should be set up right at the beginning of this long-term project. This should include not only operational data from test runs, but wear and corrosion measurements, and accounts of blockages and fouling, to quantify experience in time-related and throughput-related phenomena. (Appendix 7)

5) Pipework should be colour-coded according to contents, (Appendix 10)

6) The quench and the waste-heat boiler should be separated to aid

operational flexibility, and separators could usefully precede both the waste-heat boiler and the gas cooler (Appendix 10).

7) Enriched air instead of oxygen, and also the recycling of tar and liquor to the bed, would be worth considering at a suitable stage. (Appendix 8)

8) The experimental programme should progress logically through the caking range of appropriate Indian coals, so as to characterize them with regard to comparative process suitability. (Appendix 3)

9) The use of blends of coals might extend the range of availability of the process. Precarbonisation tests under pressure should be instituted. (Appendices 7,8)

Synthesis (see Appendices 4,5)

1) A recording system should be set up right at the beginning of this long term project. It should include a system of nomenclature for the catalysts investigated.

2) As well as iron catalysts that are cheap, but easily poisoned, cobalt/nickel and other catalyst systems should be examined.

3) Surface area blocking by non-reacting gases should be investigated.

4) Mechanics of poisoning, eg, by sulphur, should be investigated.

5) After an initial period, a specific programme of catalyst/synthesis research will need to be drawn up, so as to enable process requirements to be matched with those of the gasification plant.

The overall project

1) An overall programme should be drawn up to contain schedules of work for gasification, synthesis, gas cleaning, carbon monoxide shift, by product treatment, effluent treatment and disposal, product work-up etc. Such a programme will be more detailed for early years; less detailed further into the future. A budget will presumably be needed.

2) The overall project programme should be taken into the national plan at the right stage. Issues such as

   requirements for indigenous production of chemicals from coal

   coal availability for chemicals in relation to other demands

   characteristics (cost as well as scientifically) of available coals

   suitability of processes (including costs)

will need to be considered. Some form of cost/benefit appraisal will presumably be needed.

3) The project programme in the national plan will need to be related to the programmes of CFRI, BHEL etc.

4) A technico-economics group should be set up at RRL, not only to inject commercial issues into the planned directions and planned rates of progress of the components of the project. It will also make possible a better presentation of the RRL aims and objects in the context of national planning. A similar set of reasons would suggest the need for a market-orientated group (see 2 above). Both requirements might be met with one combined activity to cover all aspects of evaluation. (Appendices 4,9,11)

5) RRL in-house work could usefully be buttressed by university work in India, following practice in other countries. Research laboratories and universities can engender valuable synergy.

6) There is good work in universities and research organizations in many countries. RRL staff have already been to a number of countries on Fellowships concerned with scientific work of consequence to RRL. It would be useful for some of the younger RRL staff to make visits abroad to widen their experience in matters germane to the project.

7) There is need to ensure continuity in the technological leadership of relevant sections of work for this project (Appendix 11).

8) RRL possesses what is probably the only such "chemicals from coal" facility in the world involving a pilot Lurgi gasifier installation. RRL should, at the right time and to suit its own work schedule, offer a worldwide service in coal characterisation and consultancy in Lurgi coal gasification for synthesis. (Appendix 10).

9) The whole of this important project requires continuing review, and updating. The organization at RRL appears to be good and effective: specialist assistance to it will be worthwhile, preferably in the form of return visits on a regular basis, perhaps annually.

Matters where the Consultant may be able to assist progress

1) Assisting in obtaining obsolete gasifier spares.
2) Assisting in unearthing practical data on the biological oxidation of Lurgi gasifier effluent.
3) Making connexions useful for catalyst characterisation.
4) Seeking out sources of catalyst research data.
5) Helping in securing Fellowships in suitable universities and research organizations.
6) Keeping RRL informed as to literature on hazards and approaches to safety.
7) Periodically reviewing the quality of data from the experimentation, and investigating anomolous results.
8) Guiding the development of technico-economic and market evaluation.

Generally the concepts and recommendations in this report will need further development as progress is made with the programme. Repetition of this Consultancy would enable such development to be combined with a useful component of external assessment and continuous guidance.

APPENDICES

Appendix I
M. Maung/tug

21 February 1984

UNITED NATIONS

UNITED NATIONS INDUSTRIAL DEVELOPMENT ORGANIZATION

UNIDO

JOB DESCRIPTION

DP/IND/80/004/11-58/32.1.C.

Post title Consultant in Design and Operation of Large-Scale Plants (Coal Gasification)

Duration 15 days (about 10 days in Hyderabad, 1 day briefing or debriefing in Vienna, plus travel days)

Date required Around September 1984

Duty station Hyderabad, India

Purpose of project To strengthen the capabilities of the Regional Research Laboratory, Hyderabad (RRLH) for R+D work on coal gasification.

Duties The consultant will be assigned to RRLH. In consultation with the scientists and engineers of the RRLH, he will advise them on the following:

- Design and scale up philosophy of large-scale moving bed pressure gasification processes,
- Principles of techno-economic evaluation of large-scale gasification system for generation of medium Btu gas and synthesis gas,
- Guiding the experimental programme in the RRLH pilot plant for the above object,
- Gas cleaning and purification system,
- Raw gas conversion including shift conversion and methanation to convert the raw gas to town gas (medium Btu, SNG and $CO+H_2$ for synthesis purposes),
- FT synthesis including catalyst,
- Particle size fed to gasifier, utilization of 0-6 mm size, pelletization and conversion to formed coke,
- Slagging pressure gasification system.

Added at RRLH (
- Combined-cycle and power-generation applications of gasification
- Fluidized-bed gasification/combustion under pressure.

..../..

Applications and communications regarding this Job Description should be sent to:

Project Personnel Recruitment Section, Industrial Operations Division
UNIDO, VIENNA INTERNATIONAL CENTRE, P.O. [illegible] 300, [illegible], Austria

The consultant will submit a report to UNIDO concerning his findings and recommendations, particularly with regard to the experimental programme that is to be pursued by RRLH to fulfill the objectives of the project.

Qualifications

A mechanical engineer with considerable experience in designing and operation of gasification plants (pilot plants, process development unit and commercial plants), preferably based on fixed-bed gasification system. The expert should have experience in planning of experiments/runs, collection of data and evaluation of data for use in designing larger plants.

Language

English

Background information

A pilot plant for the complete gasification of coal under pressure with a capacity of 1400 Nm^3/hr of raw gas was imported from M/s Lurgi, West Germany. It has been erected at a site about 5 km from the main campus of RRLH. The gasifier is of 1 m. dia. The pilot plant has an oxygen plant (250 Nm^3/hr) supplied by M/s Linde AG, West Germany. A high pressure boiler supplied by Borsig AG to burn 0-6 mm of coal is available to generate steam. Coal crushing, screening and storing facilities have been created. The product from the gasifier will have the average composition as shown below:

	Raw gas	After hot potash wa
CO_2, Volume per cent	25.8	2.0
H_2S	0.2	-
C_nH_m	0.3	0.4
CO	24.0	32.0
H_2	39.6	52.2
CH_4	9.8	13.0
N_2	0.3	0.4
Calorific Value, Kcal/Nm^3	2800	3360

The raw gas will be subjected to hot potash wash for recoving carbon dioxide and hydrogen sulphide. After this, it will have the composition as shown above.

Condensation of byproducts and separation into tar, oil and liquor will also be made. The liquor will be subjected to biological oxidation. While coal of 6-20 mm will be used in gasifier, the under-sized coal (0-6 mm) will be used in the boiler. The plant is expected to be commissioned as soon as possible in accordance with the convenience of the consultant(s). Provision exists for adding units for shift conversion and finer purification.

It is also proposed to study air steam gasification in this plant for the generation of low Btu gas to be used in combined cycle power generation.

Appendix 2

Programme for Mr Brian Locke

RRL Coal Division

Date	Morning	Afternoon & Evening
24.8.1984		Visit to Coal Gasification pilot plant
	25.8.1984 to 26.8.1984 -	Holidays
27.8.1984	Discussion with the F.T. Synthesis Team (Committee Room)	Lecture 1 - Synfuels processes
28.8.1984	Discussion with Briquetting Group (formed coke) (Library Lecture Hall)	Visit to laboratory facilities TGA, CSU, etc.
29.8.1984	Discussion with Coal Gasification Team (pilot plant operational problems) (at pilot plant site)	Lecture 2 - Gasifier development and scale-up (Committee room) (eventually transferred to 31.8.84.)
31.8.1984	Discussion with Coal Gasification team (evaluation of the pilot plant data) (programme on coal gasification)(gas clean-up) (Committee Room)	Lecture 3 - Evaluation of gasification processes, under the auspices jointly of RRL and the Institution of Chemical Engineers (Committee room)
1.9.1984	Work on slagging gasification at British gas Corporation : fluidized combustion, power generation, plant instrumentation	

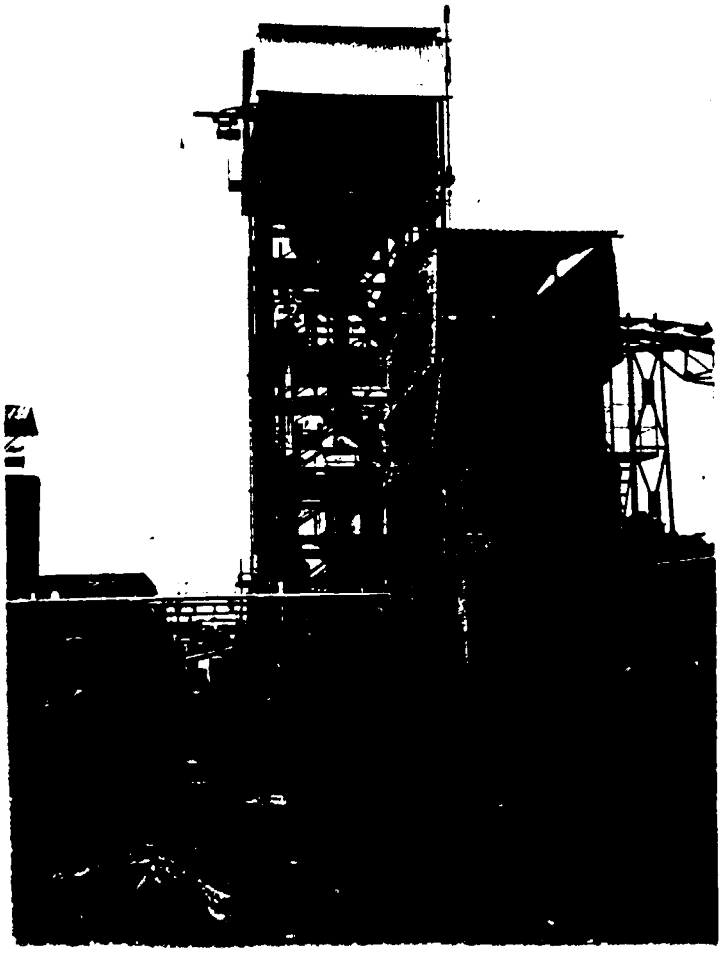
Appendix 3

The Lurgi Gasifier Installation

On 24 August 1984 the Lurgi Gasifier Installation was visited.

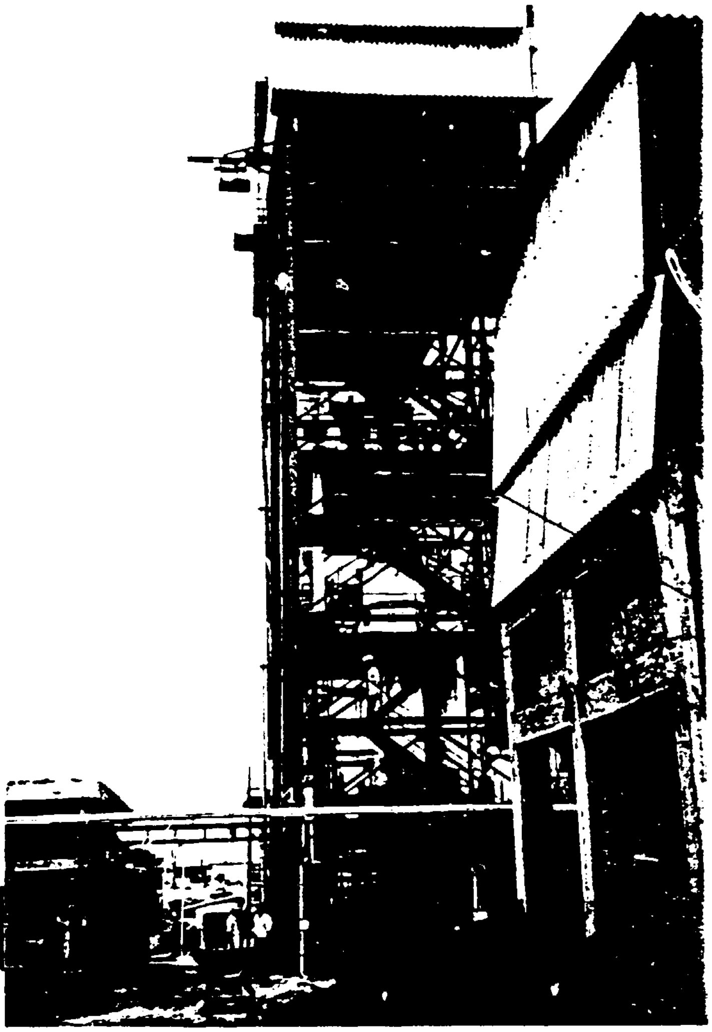
1) Photographs showing components of the Plant
2) Analyses of the principal Indian coals scheduled to be gasified
3) Plant layout and flow sheet.

COAL LOADING & CRUSHING

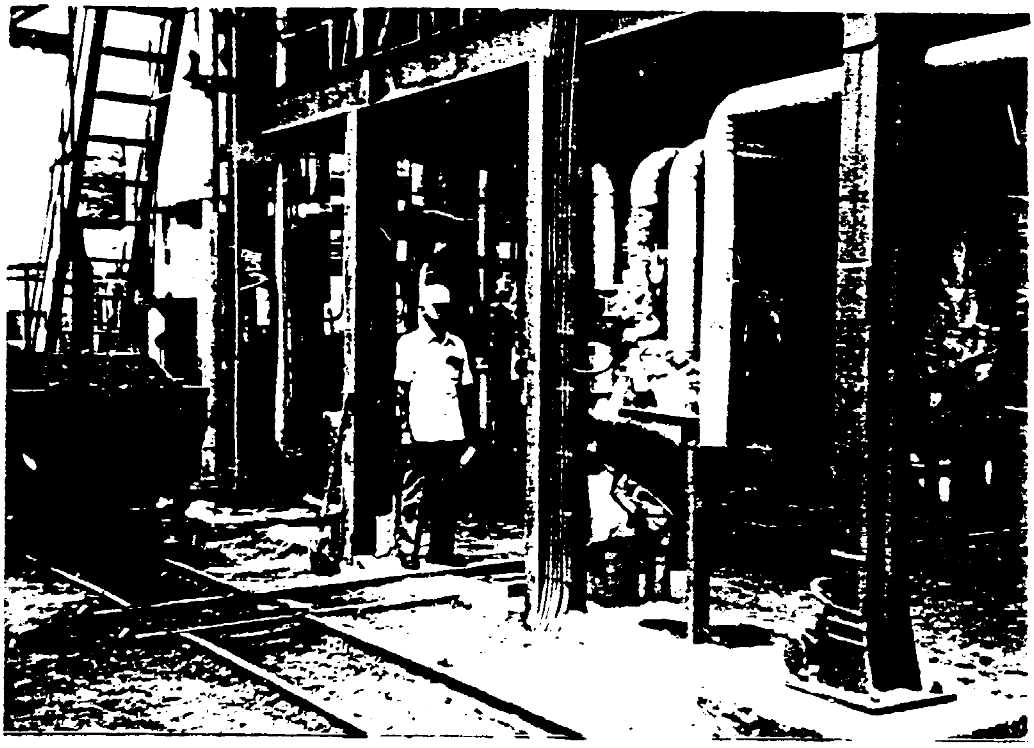


COAL BUNKERS

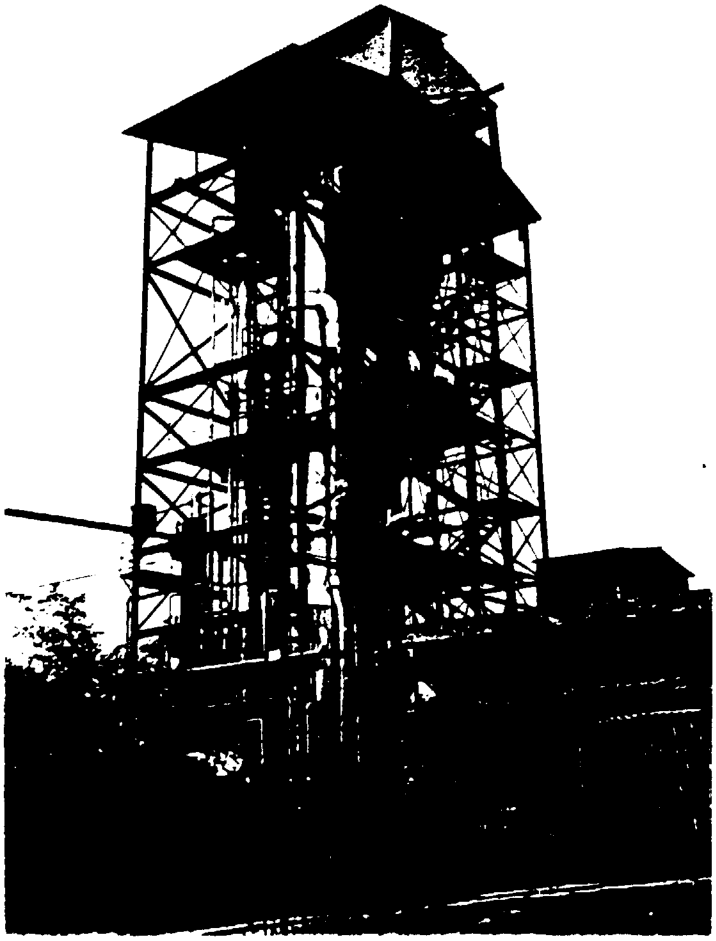
COAL TRUCK
TO GASIFIER



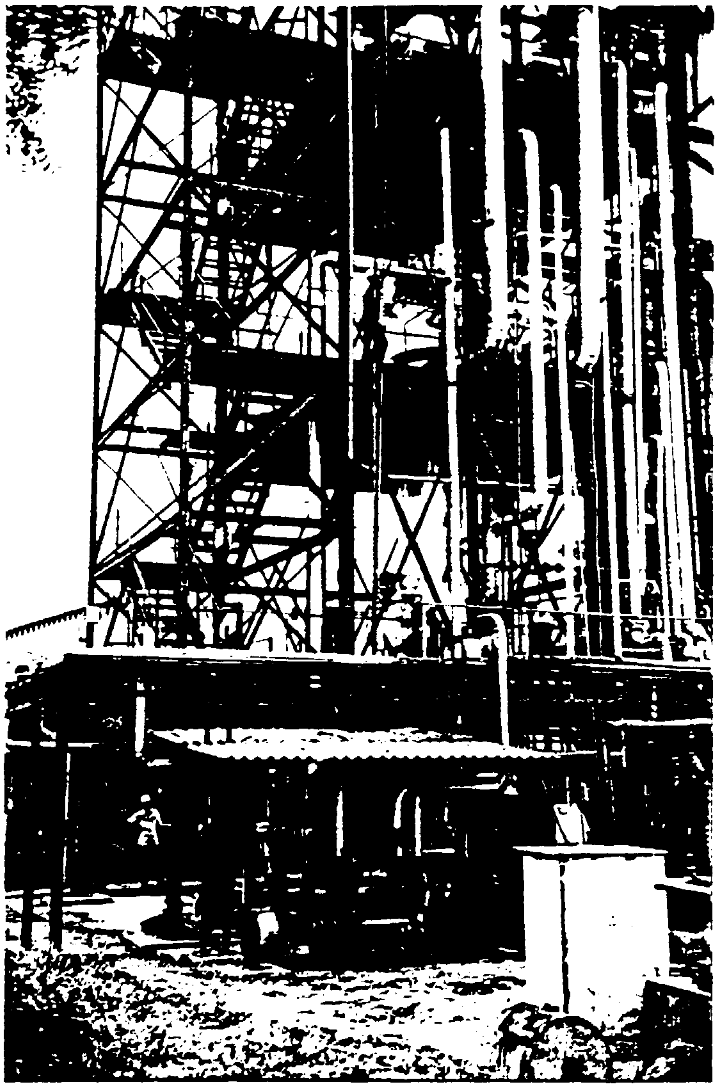
COAL TO BOILER



COAL TRUCK FOR GASIFIER (left)
ASH TRUCK FOR ASH (right)

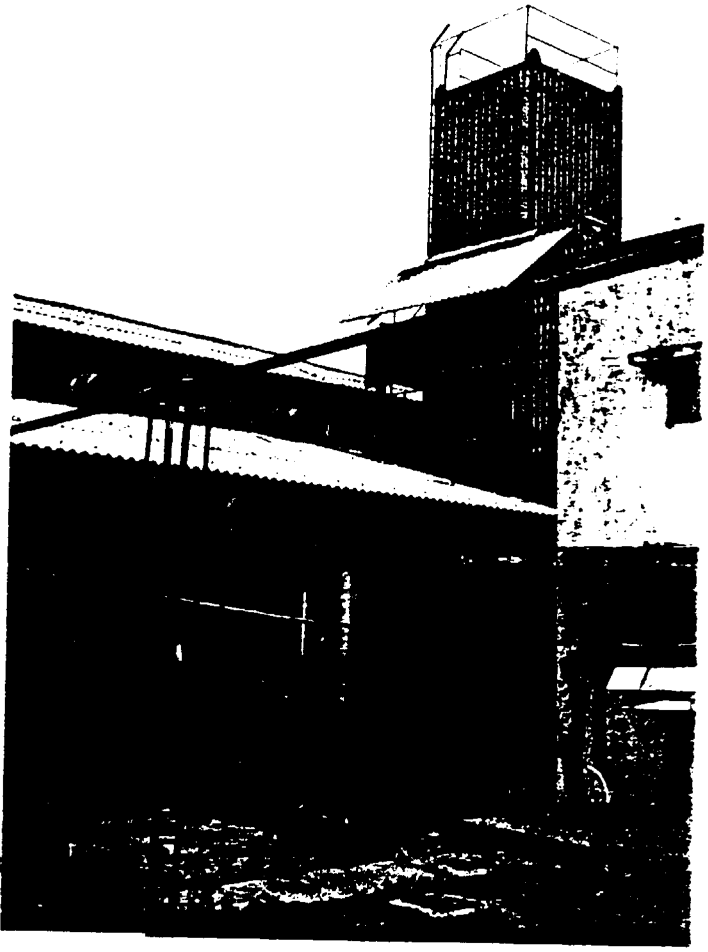


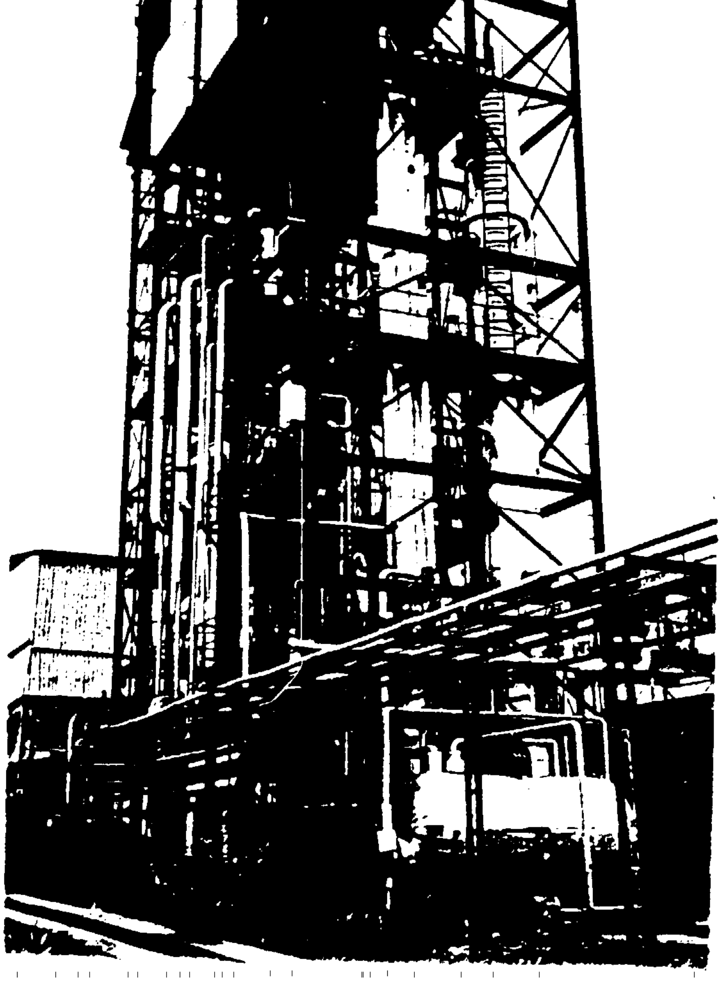
HOIST FOR COAL TRUCK
TO GASIFIER BUNKER (top right)



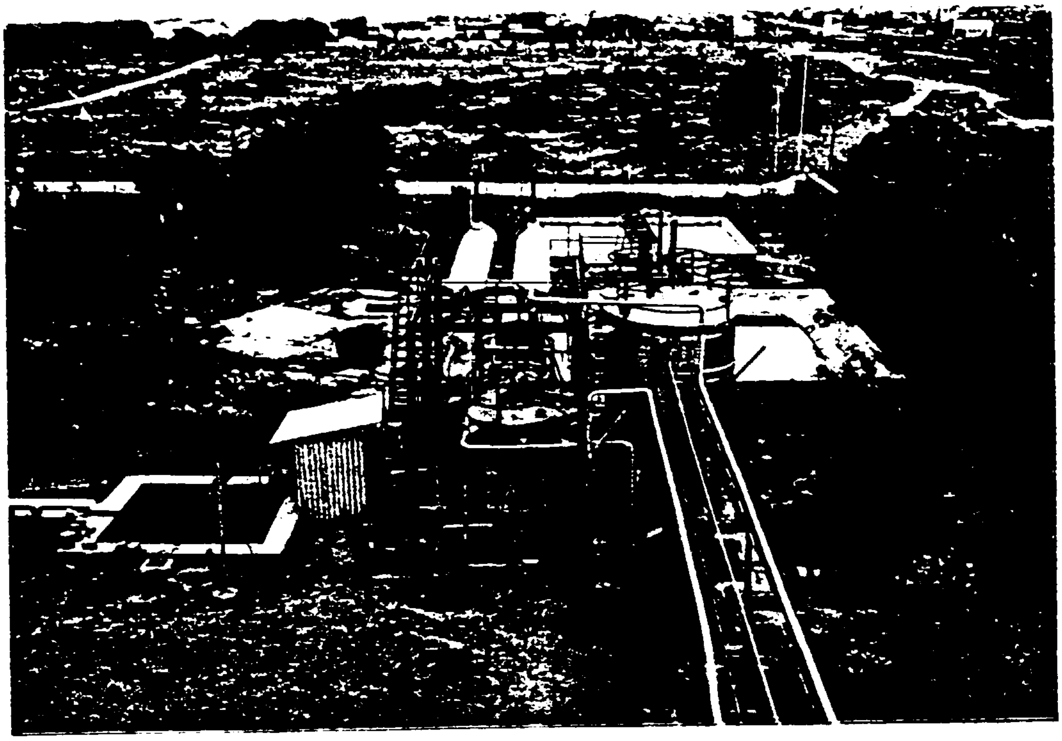
GASIFIER
(centre)

OXYGEN PLANT HOUSE





BENFIELD PLANT



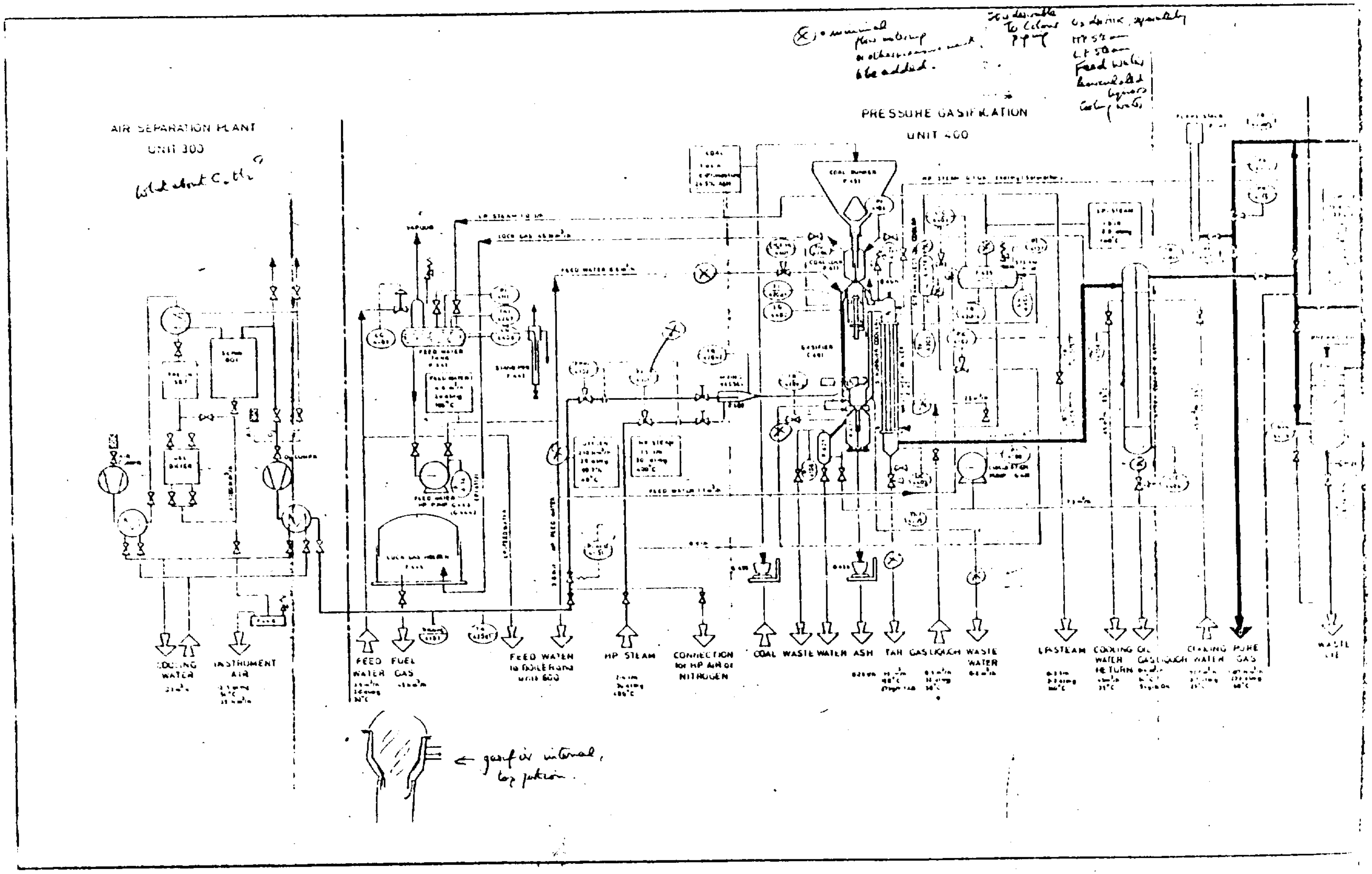
TANK FARM AND BIOLOGICAL OXIDATION PLANT

ASH DISPOSAL

The main coals to be gasified

Characteristics of Godavari Khani coals

Particulars of Analysis	GDK No.7A Incline 2 Seam Main Incline	GDK No.8A Incline 1 Seam	GDK No.10 Incline 3 Seam Bottom Sec. 171	GDK No. Open 4 Seam	GDK No. Open cast -I 3 Seam
1. Moisture at 60% RH and 40^{o}C	7.1	6.2	6.4	5.0	7.5
2. Proximate Analysis (As received basis).					
i) Moisture %	5.6	4.6	3.8	3.9	6.0
ii) Ash %	31.1	24.1	27.4	46.7	19.0
iii) Volatile Matter %	27.9	33.1	29.6	20.1	27.6
iv) Fixed carbon %	35.4	38.2	39.2	29.3	47.4
3. Ultimate Analysis					
i) Moisture %	5.56	4.58	3.75	3.93	5.97
ii) Ash %	31.15	24.12	27.38	46.73	19.06
iii) Carbon %	48.05	55.84	53.03	36.00	57.53
iv) Hydrogen %	2.97	3.76	3.37	2.19	3.46
v) Nitrogen %	1.18	1.08	0.97	0.79	1.13
vi) Sulphur %	0.40	0.72	0.50	0.24	0.58
vii) Oxygen % (by diff)	10.69	9.90	11.00	10.12	12.27
4. Ash Fusion Characteristics of Coal ash (under mildly reducing atmosphere)					
i) Initial deformation point (^{o}C)	1310	1270	1300	1325	1250
ii) Hemispherical point (^{o}C)	above 1400	1350	above 1400	above 1400	1355
iii) Flow point (^{o}C)	above 1400	1370	above 1400	above 1400	1375
5. Chemical Analysis of Coal ash					
i) Loss on ignition	0.17	0.76	0.84	0.37	1.05
ii) Silica as (SiO_2)	60.95	58.94	60.52	60.75	57.75
iii) Iron as (Fe_2O_3)	5.68	7.75	4.19	4.94	12.97
iv) Aluminium as (Al_2O_3)	29.71	27.59	27.44	30.32	25.18
v) Calcium as (CaO)	3.41	4.19	3.71	3.56	2.96
vi) Magnesium as (MgO)	-	0.55	0.78	-	-
vii) Sulphates as (SO_4)	-	-	2.48	-	-



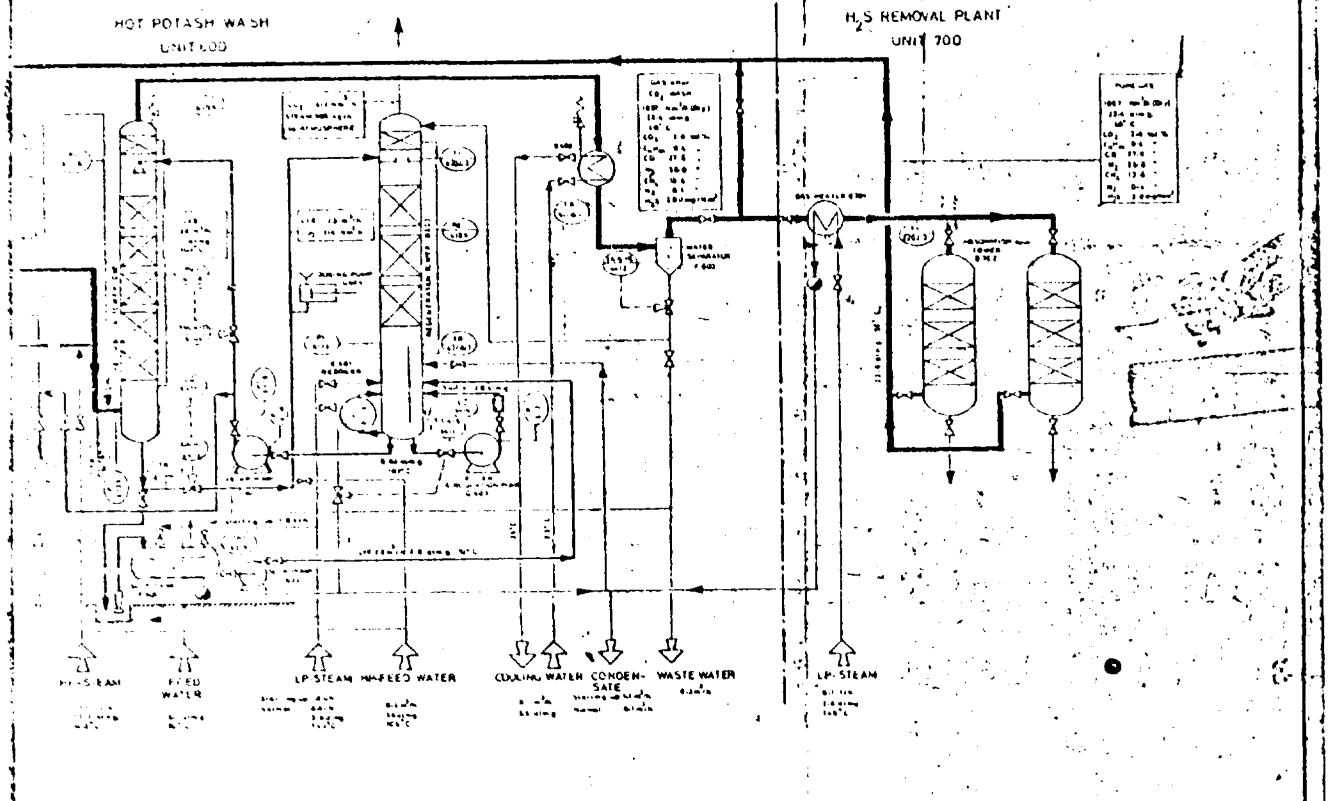
(X) = minimal flow indicator or other measurement to be added.

It is desirable to colour piping

as [illegible], separately
HP steam
LP steam
Feed water
[illegible]
liquors
Cooling water

What about C.W.?

← gasifier internal, top portion.



Reference. lurgi drawing No. 20-58-17350

REGIONAL RESEARCH LABORATORY
HYDERABAD – 9

FLOW SHEET OF THE
COAL GASIFICATION PLANT

CSIR

SCALE

DRG No H-4-05

Appendix 4

Discussion on Fischer-Tropsch Synthesis (27.8.84.am)

1. Aims of the entire Indian programme

A. Conversion of Synthesis Gas to make C_2 to C_4 olefines - RRL work

The need is for diesel fuel and kerosine - so work is concentrated on Fischer-Tropsch rather than direct hydrogenation as the demand is for paraffins rather than aromatics. RRL hopes to build a synthesis pilot plant if funds allow.

Commercial Fischer-Tropsch catalysts are not available, so RRL is

1/ preparing its own catalysts - work began early 1983

2/ screening their catalysts for close selectivity (to avoid undue product work-up). This involves testing 10 ml of catalyst in a mini-reactor, heated by a fluidized bed of hot sand. One particular Fe/Zeolite catalyst shows promise when tested with CO/H_2 -pure mixtures from cylinders.

3/ preparing repeated batches of catalyst for tests for repeatability and reproducibility

4/ preparing a 100 ml reactor which is being set up (with help from the Xytel Corporation of USA) for the early stages of scale-up to check geometry, kinetics etc

B. Another group in RRL is working on CH_3 OH synthesis with a second stage CH_3 OH reaction to make olefines

C. The Central Fuel Research Institute, Dhanbad (CFRI) is separately working on Fischer-Tropsch synthesis

This is for liquid fuel production - middle distillates - and the programme covers both catalysts and synthesis. (RRL and CFRI are both under CSIR so synergy is aimed-for between laboratories).

D. Bharat Heavy Electricals Ltd (BHEL) is working with a descending solid bed type gasifier for combined cycle power generation

Yet another BHEL group is building a fluidized gasifier for combined-cycle work, very near to the RRL pilot Lurgi plant - with which it will share steam and oxygen supplies.

3. This assignment - concerned with 1A above

A. The group at RRL had been in existence some eighteen months, and are working mainly with Fe catalyst - fused and otherwise - because it is cheap and previous results have been widely reported in general terms . The catalyst is hydrogen activated. Zeolite is the support, of diameter some 0.2 - 0.3 mm for controlled pore size. Approximately 2 gm catalyst are tested on each run. Pressures are used up to 20 bar and temperatures up to 500^0 C with up to 6000/h space velocity. (Note - details of pore-size determination are important)

60 - 70 gm C_2 - C_4 olefines can be obtained per nm^3 synthesis gas - which is similar to Ruhrchemie results - and samples have run up to 60 h continuously without C deposition.

B. RRL is interested in the possibility of a slurry reactor - as it has many advantages in use. The programme will include catalyst regeneration. and the use of promoters such as Ti, V, Mn, Cu.

C. UN has already provided a Hewlett-Packard gas chromatograph: butylene has been found to be a problem as it dissolves partly in the Na OH used to remove CO_2 from the sample. . Staff are still characterising the analytical equipment. It is possible in RRL to measure pore size and distribution, surface area etc. Facilities include infra red, e.s.r., Mossbauer spectrometry and Xray diffraction.

D. Facilities exist to make 2 - 4 kg of catalyst sample per batch in-house.

E. There are several Indian commercial firms that can make over 1 tonne per day of catalysts such as those for hydrodesulphurization, hydrogenation etc.

These catalysts are made entirely for the oil industry (UCIL collaborates with Girdler). CH_3OH catalyst is made in Bombay, also that for NH_3 synthesis.

F. RRL has already developed a cyclohexane catalyst, now licensed to UCIL - Girdler are making it - and it "works" slightly better than Girdler's own material.

G. There is as yet no work or plan to investigate surface blockage or or poisoning or use of Co/N catalysts. Commercial issues are not dominant yet. So this work is part of a logical scheme of seeing how best to use gasification.

H. CFRI are at work on liquid phase solvent extraction of coal, to make a liquid fuel for boiler use. RRL have a programme of super-critical solvent extraction of coal, from which they would examine any chemical uses of the extracts: residues would be gasified.

Recommendations - Catalysis

1) A nomenclature system will be required for the different catalysts examined.
2) It would be useful to consider the Kolbel approach, and Co/Ni catalysts for specificity.
3) Work should be done on surface area blockage by non-reacting gases.
4) Work should be done on catalyst poisoning, eg by Sulphur.
5) It would be useful to set up a small techno-economic studies group (perhaps a chemist, an engineer, and an economist) that could begin, perhaps, with the comparative economics of catalytic conversion, poisoning/purification etc. and then point out the more economic directions for research and development, and produce costings for concievable products.

   This would :

   a) introduce process costing into RRL coal process thinking
   b) accustom the different disciplines to work together
   c) be useful in suggesting preferred economic directions of work
   d) be useful when discussions start with chemical industry, plant builders, government etc.
   e) build up a body of process cost data

6) Market(ing) thinking should be introduced, so as to help guide the research programme towards products that will be needed when the processes are in operation.

Possibilities where the Consultant may be able to assist progress

Making connexions useful for Catalyst characterisation
Seeking out sources of catalyst research data
Guiding the development of technico-economic and market evaluation

Those present at the Fischer Tropsch Synthesis Discussion

K. Seshagiri Rao	Deputy Director in charge of coal gasification
Dr Y V Subba Rao	Separation processes in coal tar processing, currently engaged in in super critical gas extraction of coal
Dr G Murali Dhar	F-T catalyst preparation and characterizations
Dr M Janardana Rao	F-T Synthesis - catalyst preparation and establishing process conditions to produce C_2 - C_4 olefines
K B S Prasad	1 Modelling of a Lurgi gasifier 2 Modelling of char gasification 3 Modelling of secondary pyrolysis 4 Modelling of briquette reactor 5 Deactivation of catalyst
B Madhusudhan	1 Coal gasification 2 Mathematical modelling of fixed bed gasifiers 3 Coal-gas cleaning
M M Mallikarjunan	1 Modelling of vapour phase catalytic reactions 2 Kinetics of char gasification 3 Modelling of pressure gasifier 4 Modelling catalytic reactors undergoing catalyst deactivation
K Radha Krishnan	F-T catalyst preparation and establishing condition to produce C_2 - C_4 olefines by slurry reactor

V Durga Kumari	1 F-T catalyst preparation characterization and evaluation 2 Cracking of methanol to $C_2 - C_4$ olefines 3 F-T kinetics
A Sudhakar Rao	Hydroprocessing of coal-derived liquids
S N Reddy	1 Coal gasification of fixed bed and fluidized bed 2 Gas clean-up
K V L N Prakash	Preparation and screening of F-T catalysts for the production of $C_2 - C_4$ olefines
K Venkateswarlu	Catalyst preparation and screening for the hydrogenation of carbon monoxide and hydrogen (F-T synthesis) for the production of lower olefines
A Satyanarayana	F-T catalyst preparation, screening and operation of coal gasifier
B S Narayana Rao	F-T synthesis - project leader

Appendix 5

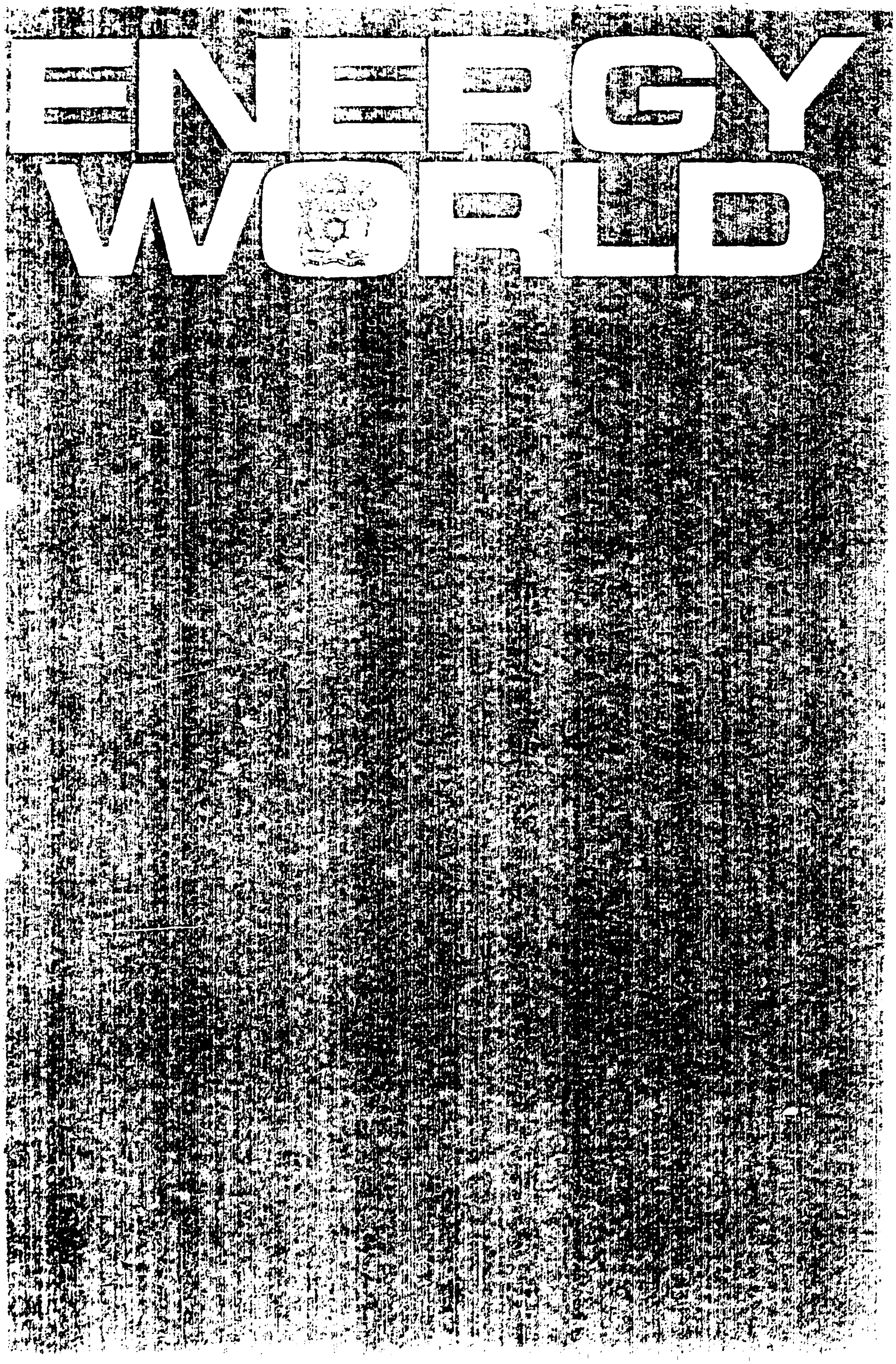
Lecture on Synfuels Processes (27 August pm)

The lecture covered the history and scope of work to derive liquid fuels and chemicals from coal, worldwide. It included an outline of British work on Fischer-Tropsch synthesis from the 1950's onwards - and gave a brief account of the Britsih Gas Corporation slagging gasification ork at Westfield.

The attached paper by the Consultant (from Energy World, the bulletin of the Institute of Energy, February 1982) covers much of the lecture material.

Other papers from which the British Government, British Gas and British Coal Utilisation Research Association work was described, are listed in Appendix 12 - documents given to RRL by the Consultant.

ENERGY WORLD



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Published monthly by the Institute of Energy
18 Devonshire Street. London W1N 2AU
telephone: 01-580 7124
Editor Christopher Payne *Assistant Editor* Joan V Deakin BA
The Institute of Energy is a Corporation Member of the Council of Engineering Institutions
The Institute of Energy is in association with:
The American Society of Mechanical Engineers
Canadian Institute of Energy
L'Institut Français de l'Energie (Paris)
The Fuel Society of Japan (Tokyo)
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Contents: February 1982

Institute of Energy *March meetings*

North-Eastern

2 Mar. A mass and energy balance on a glass tank, by Dr J H Harker and Dr J H Backhurst. Newcastle Polytechnic, Ellison Building at 1815 h (refreshments in the refectory before meeting).

Scottish

2 Mar. Heat pipes, by D A Reay. Royal Scottish Automobile Club, 11 Blythswood Square, Glasgow at 1830 h (tea and sandwiches 1800 h). Joint meeting with CIBS.

Midland

4 Mar. The use of oxygen in high temperature combustion processes, by A O Dyer. University of Aston in Birmingham at 1900 h.

Yorkshire

9 Mar. One day conference on *Generation of power from nuclear energy*. CEGB North Eastern Division HQ, Beckwith Knowle, Harrogate.

East Midland

11 Mar. Experiences with pipeline charging and coal preheating at Redcar, by Dr W A Honeyball. BCRA Research Centre, Wingerworth, Chesterfield at 1900 h (refreshments after lecture). Joint meeting with COMA.

S Wales and W of England

15 Mar. Thermal insulation. Illustrated talk and works tour. Rockwool Ltd, Rhiwceiliog, Pencoed, Bridgend at 1430 h.

London and Home Counties

17 Mar. Tea at 1710 h, AGM at 1730 h. Alternative energy sources, by Dr D T Swift-Hook at 1800 h.

North-Western/Merseyside

18 Mar. Funding of oil explorations and development, by J H MacDonald. Thornton Research Centre, Pool Lane, Ince at 1700 for 1730 h. Joint meeting with Stanlow branch, Institute of Petroleum.

Midland

19 Mar. Social evening. West Midlands Gas HQ, Solihull.

North-Eastern

22 Mar. The state of the art in coal-fired ships, by J B Riksheim. Neville Hall, Newcastle at 1800 h (refreshments before meeting). Joint meeting with NE Institution of Engineers and Shipbuilders.

Merseyside sub-section

25 Mar. Heat recovery system, by J Edwards. Manweb Head Office, Chester at 1800 for 1830 h. Joint meeting with CIBS.

Northern Ireland

25 Mar. Tidal energy from Strangford Loch, by Dr L McClements. Ashby Lecture Theatre, Belfast at 1900 h.

East Midland

26 Mar. Annual dinner. The George Hotel, Nottingham.

Yorkshire

27 Mar. Dinner dance at Cairn Hotel, Harrogate at 1915 for 1945 h.

31 Mar. AGM. BSC, Sheffield Laboratories, Swinden House, Rotherham at 1400 h.

Feedstocks and fuels from coal: The need for process changes*

Brian Locke CEng FInstE FIChemE†

Origins

Feedstocks and fuels from coal are not new, and from the beginning of coal processing their characteristics have had to be matched against, and reconciled with, the needs of their users. The Romans used coal in Britain 2000 years ago, and grumbles about pollution have echoed through the millenia, even to today. Gaseous fuel first appeared in 1691 (Clayton) and was commercialized from 1802 (Boulton and Watt). Tar and oil were distilled as primary products of coal carbonization in 1782 (Dundonald); and the first gasification of powdered coal for power generation was in 1806 (Niepce).

Synthesis

Modern synthetic processes derive from Sabatier & Senderens (1902) via BASF (1913), Fischer (1923), and Fischer & Tropsch (1926) – leading to the first plant in 1937 using ambient pressure synthesis gas derived from coke oven gas. The original commercial purpose was for synthesis of lubricating oils, later changing to fuels as the primary object. Koppers, Didier and others experimented with versions of gasification; Winkler used a fluidized gasifier; and, in 1938 Lurgi operated their O_2/steam gasifier at 8-10 ata at Holten. An interesting problem then was that increased pressure reduced the gasoline output, while increasing diesel fractions and waxes. The products were treated by catalytic and thermal cracking, aromatization and polymerization.

Hydrogenation

Fischer-Tropsch synthesis depends first on reducing steam by coal (using the heat of partial oxidation of more coal to CO) so as to add the hydrogen to the carbon (monoxide) by catalytic synthesis. The opposite route to the same end – to increase the H/C ratio – but by direct reduction of coal, began with Berthelot (1870). Bergius (1913) reduced coal with hydrogen at 100 ata, followed by Fischer jointly with Tropsch (1917), and with Schrader (1912), the first Bergius plant being installed in 1926 at Leuna to treat coals and also tars and oils from carbonization.

Work was also done in the USA, France, Italy, Japan and Korea, for example, before World War 2, on both synthesis and hydrogenation as well as the production of liquid fuels and chemical feedstocks from coal carbonization and shale distillation. British work on hydrogenation began in 1923, followed by a pilot plant in 1927; and ICI built a commercial plant in 1935. There were also several Fischer-Tropsch experimental units in Britain from 1935 onwards; and over 150 processes were considered up to the war. Most UK liquid fuels, however, were oils from overseas.

UK home production of liquid fuels at the beginning of World War 2 was as shown below:

Process	*Gasoline tpa*	*Heavy oils, creosote, diesel oil tpa*
High temperature coal carbonization	214 000	357 000
Low temperature coal carbonization	4000	10 000
Shale oil distillation	23 000	68 000
Hydrogenation of coal and creosote	141 000	–
Domestic oil wells	–	1000
Total	382 000	436 500

In Germany in 1939 where foreign oil supplies were a problem there were ten hydrogenation, nine synthesis plants with capacities of 1 300 000 and 600 000 tpa respectively.

Solvent extraction

Deriving from analysis techniques, solvent extraction of coal probably began in 1918 (Fischer and Gluud) and, until the war in 1939, practically every conceivable solvent was tried on coals, peats, lignites etc, in many countries around the world. The extracts were of higher, and the residues of lower H/C ratio than were the feed coals, and ranges of unsaturated aromatic, and some aliphatic compounds were found, resins etc. However, there was never any sign of a commercial process (except for electrode manufacture) until after the war with work particularly in the USA, and in Britain first at BCURA, and then by NCB.

Fitting products to markets

Throughout the period up to the war the major problem in coal processing was that of fitting products to markets, economically. Synthesis produced good diesel fuel, and olefines suitable for polymerization to lubricants; but poor gasoline and a range of waxes requiring individual treatments to substitute for existing products. Hydrogenation produced good gasoline, but no diesel or lubricating feedstocks. Both produced by-product gases. Generally, with the need to develop purification and by-product work-up streams where the primary processing was quite overshadowed by the ancillary plant – and the need to fit new products into established markets – it was cheaper and easier in most places to go on using petroleum – despite the difficulties of war. The pressures are different today, but we have similar problems, and must overcome them.

After 1945 there was some work on synthesis, hydrogenation etc, in the USA, USSR, Germany, Britain etc, although (except in South Africa) process economics could not compete against falling oil prices. The UK government continued to run Fischer-Tropsch development begun before September 1939 including work on mill-scale, Co, and other catalysts. The government's 'oil from coal' project of the 1950s provided for high CO/H_2 ratio synthesis gas and the slurry synthesis with Kolbel catalyst, based on slagging gasifier gas. The slagging gasifiers so successfully developed by the government under contract at Solihull and at

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*From a paper given to the European seminar on *Feedstocks and fuels from coal*, held in London, 24 September 1981

†Senior Partner, Cadogan Consultants, London

Table 1 Some basic properties of coals

Coal rank code	Carbon (dmmf) (per cent)	Hydrogen (dmmf) (per cent)	Volatile matter (dmmf) (per cent)	Calorific value (dmmf) (Btu/lb)	Moisture (ad) (per cent)	Gray-King coke type	BS Swelling number	Main uses*
100	92-95	2.8-3.9	2-9	15300-15850	0.9-2.9	A	0	Closed stoves, horticulture, malting, boilers
201	91-93.5	3.7-4.3	9-13.5	15500-15950	0.6-2.0	A-C	0-4	Closed stoves, boilers
202-4	90.5-93	4.1-4.8	13.5-19.5	15600-15950	0.5-1.3	B-G8	1->9	Boilers, blending for coking
301	87-91.5	4.5-5.4	19.5-32	15300-16000	0.4-2.0	G4 and over	6->9	Blast-furnace and foundry coke
302 303	86.5-91.5	4.4-5.4	19.5-32	14900-15900	0.7-4.0	A-G3	0-9	Boilers, carbonization (CRC 302 only)
400	84.5-89.5	5.1-5.8	32-40	15200-15900	0.9-2.5	G9 and over	5½-9	Blast-furnace coke, gas
500	83.5-88.5	5.1-5.8	32-40	14900-15700	1.0-4.5	G5-G8	3½-9	Blast-furnace coke, gas
600	83-88	5.1-5.8	34-42	14800-15600	1.5-6.0	G1-G4	3-8	Blast-furnace coke, boilers, open fires, gas
700	82-87.5	5.0-5.8	34-42	14550-15400	2.0-8.5	E-G	1-6½	Boilers, gas producers, blending for carbonization, open fires
800	80-86	4.9-5.7	34-45	14050-15050	3.0-14.0	C-D	½-4½	Boilers, open fires, gas producers
900	78.5-84	4.9-5.6	34-45	13600-14800	4.0-14.5	A-B	0-2	Boilers, open fires

Dmmf – dry, mineral-matter-free basis; ad – air-dried basis. 1 Btu/lb equals 2.326 kJ/kg.

*In assessing the suitability of a coal for a specific use (eg coke manufacture), account has usually also to be taken of other properties (eg as and sulphur). 'Closed stoves' and 'open fires' are domestic appliances.

(By courtesy of the National Coal Board)

Table 2 Some characteristics mainly related to ash

Description	*Ash (per cent)*	*Sulphur (per cent)*	*Carbon dioxide (per cent)*	*Chlorine (per cent)*	*Phosphorus (per cent)*	*Arsenic (ppm) As_2O_3*	*Hemisphere temperature of ash (°C) in reducing atmosphere*
Very high	15.1 and over	4.01 and over	2.51 and over	0.61 and over	0.071 and over	71 and over	1460 and over
High	10.1-15.0	2.51-4.00	1.01-0.60	0.31-0.60	0.031-0.070	31-70	1360-1450
Moderately high	7.6-10.0	2.01-2.50	–	–	–	–	–
Moderate	5.1-7.5	1.51-2.00	0.51-1.00	0.16-0.30	0.011-0.030	11-30	1210-1350
Moderately low	–	1.01-1.50	–	–	–	–	–
Low	2.6-5.0	0.51-1.00	0.21-0.50	0.06-0.15	0.0051-0.010	4-10	1110-1200
Very low	2.5 and under	0.50 and under	0.20 and under	0.05 and under	0.0050 and under	3 and under	1100 and under

Notes– (1) The steps between adjacent ranges indicate the degree of approximation to which each property should be reported. The figure are 'on air-dried coal'.

(2) Individual coalfields may tend to have their own characteristics. Thus, in Scotland and South Wales where most coals hav sulphur contents of under 1.5%, 1.9%, for example, would be 'high' rather than 'moderate'.

(By courtesy of the National Coal Board)

Leatherhead were the forerunners of the British Gas slagging gasifier at Westfield.

Some practical issues involved today

In the past, coals used for processing have, of course, been specially selected to produce optimum results. However, the tonnages so processed have been relatively small compared with the demands likely to be made upon coal as oil becomes relatively higher in price. Much wider ranges of coal will have to be considered in future.

Coal, generally, contains:

Useful components	Hydrocarbons (H_2% 0-6%)	
Also, and in varying amounts	Oxygen Sulphur Nitrogen	both organic and inorganic
	Ash minerals	– some of which decompose when processed.
	Water	

Only the hydrocarbons are of real value. All the other components, even though they may represent some value as by-products, are really a nuisance – either directly, or, as a result of the processing, that may produce compounds such as cyanides or thiocyanates against whose toxicity special measures have to be taken. In some parts of the world the hydrocarbons may comprise less than half the weight of the coal as mined: in some areas the nature of the coal is such that even coal washing may leave substantial proportions of minerals in the factions that can be sent forward to processing – Brazil, or India, for example.

Before the post-World War 2 oil boom the principal processes involved in treating coal chemically used different means of performing the three main operations required

(a) *separating the hydrocarbon 'compound' in coal into constituent fractions*
(b) *adding more hydrogen chemically*
(c) *adapting the resulting compounds for sale as required by the markets*

Let us examine the implications. The means were:

Carbonization: separated everything volatilizable, often with some mild hydrogenation – then separated the products for onward treatment usually in different works;

Synthesis: split the initial hydrocarbon, either in two stages (via coke) or in one stage directly, into $CO + H_2$: synthesized; and processed the products;

Hydrogenation: reacted the coal directly with hydrogen (itself often derived from coal, via coke); and processed the products.

The above are all heavily independent on carbonization initially – which is something we should not forget after a mere couple of decades of relatively cheap oil.

The coal processing industry of the future will be very different from that in the past – and the present carbonization industry in the world should really be regarded as at an interim, rather than an ultimate stage. During the oil boom energy demand has risen, and so have the unit sizes not only of processing plants, but also the demands of customers for fuels and feedstocks. And social and legal requirements have advanced, too.

Accordingly, in future, coal processing plants will be different in the following respects:

☐ *larger throughput:* for larger markets
☐ *larger unit size:* to reduce replication of units
☐ *higher pressures, where possible:* to reduce capital charges
☐ *higher temperatures – in many cases:* to improve efficiency
☐ *reduced pollution:* gaseous, liquid, particulate, noise etc, in view of advancing social and legal requirements.
☐ *less manual labour:* in line with social trends
☐ *less labour:* for economic reasons
☐ *more on-line optimization and process control:* for economics; and also as product specifications tighten
☐ *more amenity-oriented:* in line with social trends

Mere scale-up, or replication of 'old' processes will not suit the future.

There will also be the inevitable problems caused by such trends, as seen in the ammonia industry for example. The advantage of single product specification and large output, so great while a long-term contract is running, becomes a disadvantage when it runs out and flexibility of economic output and of product range become desirable – but difficult. Some of these issues, along with insufficient coal/oil price differentials, have been behind the soft-pedalling of major oil-from-coal schemes recently.

Accordingly when, in a decade or two, coal processing plants come to be built in rising numbers, the processes themselves and the equipment are likely to have developed much further than those we have at present.

Relating products to markets

Markets tend to have product specifications that vary only slightly: coals can vary greatly, not only from place to place, but even from one seam to another in the same mine. And of the processes available all not only have different characteristics, but are capable of varying outputs of different products, according to the operational conditions chosen.

When the oil industry was evolving it could make its progress step by step, each on a reasonably small scale as the prices of its products fell – as a result of optimization of plants and outputs in relation to the markets it was largely creating. But the new coal processing industry will not be able to enter markets with speciality products having good economic margins: it will have to 'go big' from the start, and with narrow economic margins (most of which are still negative today). This makes planning difficult.

Coal supplies

Tables 1 and 2 show how basic, and ash-related properties vary in British coals, which are perhaps the best classified in the world – as well as being of higher 'quality' than much coal around the world that will be processed over coming decades.

Products in markets to be served by coal-based feedstocks and fuels

(a) *Gases:* Tables 3 and 4 cover some of the requirements for gases in industry, and for pipe-line purposes. Table 5 outlines the uses for gases from coal processing.
(b) *Liquids:* Tables 6 and 7 cover typical specifications of the liquid fuels from the oil industry and from coal carbonization to which markets have become accustomed. Table 8 is a digest of some of the requirements for transport fuels. Table 9 outlines some of the uses for liquids from coal processing.
(c) *Solids:* The main uses for solids from coal process-

Table 3 Some industrial gas requirements

Industry	*Method of utilization of gas*	*Scale of operations Quantity of gas or equivalent used per unit of industrial plant (therms/d)*	*Quality of gas required*	*Pressure*
Glass	Heating glass furnaces	Up to 20 000	CV 280 to 340 Btu/s.ft^3 Relatively constant composition Sulphur 12 to 20 gr/100 ft^3	Not less than 1 lb/in^2, but could be 100 to 150 lb/in^2
Iron and steel	Heating coke ovens	Up to 100 000	100 to 500 Btu/s.ft^3 unpurified but free from dust	Atmospheric
	Furnace heating (excluding integrated steel-coke oven plants)	From 50 000 upwards	300 Btu/s.ft^3 or above Low sulphur content	High pressure could be advantageous
Chemical	Synthesis of ammonia	100 000 upwards	Very low concentration of methane and sulphur For ammonia synthesis; mixture of N_2, CO, CO_2 and H_2 before CO conversion to required composition	About 250 atm
	Synthesis of alcohols and HCN		For alcohol production etc: CO, CO_2 and H_2 before CO conversion to required composition	
Ceramic	(1) Heavy clay products (2) Refractories	5000 to 20 000	Unpurified gas free from dust Low sulphur advantageous	Atmospheric
	(3) Fine ceramics	1500 to 3000	Purified gas. Low sulphur content	
Coal	Heating coke ovens	Up to 75 000	Unpurified. CV 100 to 500 Btu/s.ft^3	Atmospheric
	Brick kilns, coal drying	5000 to 10 000		

Table 4 Some town and natural gas characteristics

		Height-flame-speed gases					*Low-flame-speed gases (natural gas)*	
		Coal gas	*Coal-gas/CWG mixture*	*Cyclic-reformer gas*	*Continuous-reformer gas*		*A North Sea gas*	*Dutch Slochteren*
					CRG-enriched	*GRH-enriched*		
Composition CO_2	% vol	2.0	4.2	5.0	14.2	13.3	0.6	0.9
O_2		0.5	0.6	0.4	nil	nil	—	—
C_nH_m		3.5	3.8	7.4	nil	0.4	—	—
CO		7.5	14.8	20.3	3.3	5.5	—	—
H_2		51.8	48.7	49.8	48.6	55.9	—	—
CH_4		27.0	*20.0	*11.5	33.7	17.9	90.0	81.7
C_2H_6		1.2	—	—	—	7.0	5.1	2.7
C_3H_8 and higher		—	—	—	—	—	1.6	0.7
N_2		6.5	6.6	5.6	0.2	—	2.7	14.0
Total		100.0	100.0	100.0	100.0	100.0	100.0	100.0
Specific gravity (air = 1.0)		0.422	0.482	0.550	0.467	0.470	0.618	0.643
Calorific value	Btu/ft^{3}stp(sat)	550	500	500	500	500	1033	881
	Btu/ft^{3}stp(dry)	559.8	508.9	508.9	508.9	508.9	1051	897
	MJ/m^3(st)	20.88	18.98	18.98	18.98	18.98	39.20	33.44
Wobbe number (UK)		846	720	674	730	730	1315	1100
	(SI)	32.12	27.33	25.59	27.71	27.71	49.92	41.76
Weaver flame-speed factor 'S' (hydrogen = 100)		43.5	40.2	40.4	35.1	39.9	14.1	13.0

*Includes higher paraffins

ing are likely to include:

- Cokes: metallurgical reductions and melting eg iron and steel
- 'Desmoked' coals and derivatives: domestic heating appliances
- Residues as fuels: boiler and domestic fuels to replace high S coals
- Residues as carbon: to replace petroleum coke in anodes for aluminium etc, refining
- Residues as feedstock: to replace coking coals in formed coke for the steel and phosphorus industries
- Slag and ash: for use as aggregates, or in cement, or foamed or cast according to analysis and local demand.

(d) *Product adjustment:* The main adjustments to products to serve markets will be made during the processing and blending, but in some cases external additives may be required.
Examples are:
for town gas or SNG: odorants; Wobbe Index control; emissivity control
for gasoline· Anti knock-TEL (in small quantity) or MTBE (more modern)

Some of the principal processes for coal

Carbonization: Conventional – low, medium and high temperature
COED
Toscoal
NCB fluidized carbonization
Flash pyrolysis (and hydrogenation/pyrolysis)
Gasification: Lurgi Dry Ash (conventional)
Slagging Lurgi (BGC)
Koppers Totzek
Texaco/GVV
Winkler
Shell-Koppers
Cogas
CO_2 Acceptor
Synthane
Syntheses: Fischer-Tropsch – conventional
Fischer-Tropsch – high CO (from slagging gasification (BGC))
CH_4
CH_3 OH
CH_3 OH to gasoline
Hydrogenation: Coalcon
H Coal
Catalytic hydrocracking
Synthoil
Hydrogenation using non-fossil-derived hydrogen, and also using nuclear heat
Solvent refining, extraction and hydrogenation: SRC 1 and 2
SCSE
LSE
Donor solvent
Consol synthetic

Some ancillaries to coal processes

Synthesis gas composition adjustment and purification:
CO shift
Benfield and
Stretford developments
Rectisol
Giammarco-Vetrocoke etc

Effluent treatment: Neutralization, precipitation, biological oxidation (from phenols, cyanides, thiocyanates etc) etc

Product work-up: Cracking, hydrocracking, hydrotreating, reforming, hydrodenitrogenation (from indo. quinolines etc)

What the processes produce

Gases: Table 10 lists principal characteristics of most of the major gas-making processes that can be regarded as operable on a commercial scale. Oil gasification processes are included alongside those for coal and coke gasification and carbonization, for completeness. The range of properties of the product gases will be seen to be almost as great as that for coals.

Table 11 lists some of the requirements for hydrogenation and synthesis gases, in comparison with (a) the way aliphatic synthesis products may vary, and (b) the way Fischer-Tropsch synthesis products may vary according to process conditions, catalyst etc.

The selection of gasification processes for gas making or for synthesis, and the selection of synthesis details should be relatively straightforward, based on the economics of the market place.

Liquids: there are perhaps 100 potential processes under consideration or experimentation or pilot development around the world.

These may produce useful liquids from coal, via gasification (by fixed, entrained and fluidized bed; and in one, two, or three stages) and synthesis, or hydrogenation (including solvent, donor solvent and catalytic) or carbonization (several techniques including multi-stage), or solvent extraction; and intermixtures of these different techniques, including potential product work-

Table 5 Typical uses for coal-based gaseous feedstocks and fuels

Gaseous Feedstock	*Industry*	*Uses*
H_2	Food	Oil and fat hydrogenation
	Oil	Hydrocracking Hydrodesulphurization
	Speciality	Fuel Cells – weather stations eg
	Transport (future)	Fuel (if storage techniques improve) Fuel Cells
	Coal Processing (future)	Hydrogenation processes
	Iron & Steel (future)	Direct reduction
H_2 + CO	Chemical	Ammonia synthesis
	Chemical	Methanol synthesis
	Gas	Town gas components
	Gas (future)	Methane synthesis for SNG
	Industry generally	Burner & furnace fuels in ceramics, glass, tobacco, food, textiles, leather, rubber, ferrous and non-ferrous metals, paint drying
	Agriculture	CO_2 enrichment in glass houses etc
	Coal Feedstocks & Fuels	Fischer-Tropsch &c syntheses
H_2 + CO + CH_4	Gas (future)	possibly as substitute for high-flame-speed town gas

up stages. The range of possible products is vast. There are also shale and tar-sand processes, some of which can be sensibly adapted to coal, for example, Toscoal. It is interesting to go into detail as to which particular product needs which particular work-up to suit it for its market.

Each process has factors to commend it, yet experience suggests that only a handful will actually be built and operated on a sizeable scale. Presumably the protagonists will continue each to extol the advantages of each particular process or route until market demand selects winners. And this will be lengthy owing to the large costs of a coal processing plant.

Solids: there should not be much problem over the solids produced, either as main products or as by-products. Products such as cokes on the one hand are tailor-made to market requirements; on the other, burnable residues will tend to be more-easily-usable than the original coal – and so should be saleable directly. Other by-products, such as asphalts, should generally be adaptable to markets; and ashes, or slags, should be able to be dumped without undue difficulty, or sold where demand exists.

Table 6 Approximate specifications of typical oil fuels (in 1980)

Properties	*Standard grade kerosine*	*Gas, oil, diesel oil*	*Light fuel oil*	*Medium fuel oil*	*Heavy fuel oil*
Class	*SG*	*CI/SH*	*SG*	*SG*	*SG*
Viscosity kinematic (cSt) at					
40°C min	1.0	1.5			
40°C max	2.0	5.5			
82°C max			12.5	30	70
Type of oil	*Distillate*	*Distillate*	*Residual*	*Residual*	*Residual*
CV–Gross Specific Energy					
MJ/kg	46.5	45.4	43.3	43.0	42.8
MJ/litre	36.7	38.0	40.1	40.6	41.1
Btu/lb approx	20,100	19,500	18,700	18,500	18,400
Therms/gallon approx	1.58	1.64	1.74	1.74	1.77
Sulphur content %wt	0.2	0.8*	3.2	3.5	3.5
Mean specific ht 0-100°C	0.5	0.49	0.46	0.45	0.45
Relative density at 15°C kg/l	0.79	0.840	0.93	0.93	0.98
Min. handling temp °C	Ambient	Ambient	10	30	45
Min. storage temp °C	Ambient	Ambient	10	25	35
Atomizing temperature °C	–	Ambient	60-70	80-105	115-125
Flash point (closed) °C min	38	55	66	66	66

(to be updated)
*0.5% from 1 October 1980

Table 7 Coal tar fuels

CTF	*Pumping temperature minimum*		*Density at 100°F lb/gal*	*Calorific values*		
				Gross: Btu/lb	*Net: Btu/lb*	*Net: Btu/gal at atomizing temperature*
50	*Atmospheric*		9.5 to 10.5	16 500 to 17 500	15 900 to 16 650	166 000 ± 10 000
	°F	°C				
100	90	32	9.5 to 10.5	16 500 to 17 500	15 900 to 16 650	162 700 ± 9800
200	85	29	19.9 to 12.1	16 250 to 16 900	15 710 to 16 250	166 200 ± 10 000
250	140	60	10.9 to 12.1	16 200 to 16 700	15 670 to 16 050	176 000 ± 10 600
300	180	82	11.4 to 12.6	16 050 to 16 550	15 550 to 15 940	196 700 ± 11 800
400	270	132	11.9 to 13.1	15 800 to 16 300	15 370 to 15 750	177 400 ± 10 600

Notes–1. The range given for densities and calorific values embrace coal tar fuels derived from the various sources of crude tar.
2. Net CV = 0.75 × (Gross CV + 4700). Btu/lb.
3. The CTF grade number indicates the temperature (°F) usually considered suitable for satisfactory atomization:
4. Individual suppliers determine their own prices in negotiating with the buyers, but in general the prices of Coal Tar Fuels are comparable with the prices of petroleum fuel oils on a calorific value basis.
5. BS 1469 covers grades, properties, sampling and tests.
(By courtesy of The Association of Tar Distillers).

1 lb/gal ≃ 99.8 kg m^{-3}. 1 Btu/lb ≃ 2.33 10^3 J kg^{-1} 100 000 Btu/gal ≃ 23.3 MJ/litre.

Reconciling issues

Liquid fuels and feedstocks will need the most effort in matching market requirements, because they have ranges of constituents, mostly in proportions different from those to which the markets have become adjusted.

It may be convenient to consider Synthoil as just one example of the issues in product work-up. The most attractive targets would be diesel and jet fuels and gasoline. Accordingly, the multi-ring aromatics would all need selective hydrocracking – either to branched- or to cyclo-paraffins. In turn this would make it desirable to develop special catalysts, probably of the zeolite or perovskite types, and possibly in conjunction with tungsten or molybdenum sulphides. Phenols and nitrogen- and sulphur-containing compounds would need to be hydrotreated. Even for use as fuel oil the carbon-deposition propensities would have to be considerably reduced.

Generally, with coal-based liquids, it is the aromatic components that require the work-up, which means adding additional hydrogen to the derived molecules beyond what was possible in the original processing of the coal.

Transport fuels

In parallel with the efforts of coal processors to match products to market needs the users, too, will be making some efforts – in two main ways. Perhaps transport engines are the most sensitive to fuel specification change so they provide a convenient example. Recent developments, such as stratified charge and straight-through combustion chambers have been made in the ordinary course of engine evolution, and come fortunately at a time when there is already some move down the gasoline octane number scale on public opinion and general economics grounds. Such developments should help the assimilation of new gasolines.

Then, too, there are the developments made especially to widen the range of fuels – such as the introduction of gasohol that has already gone a little too fast for government-adjusted economics in Brazil recently. There is experiment once more with cylinder methane for motor cars. In London there have been, for years, 1000 taxicabs fuelled by lpg. And there has even been a re-examination (abroad) of the possibilities of reviving the UK war-time vehicle gas-producer.

So the climate is such that different gasolines would be accepted – but they would have to be economic, and preferably intermiscible with current gasolines.

Methanol offers an especially interesting opportunity for transport. Not only can it be acceptable up to perhaps 15% in conventional gasoline: it can also be used on its own in engines of about 13:1 compression ratio. These, of course, would need to be made. Methanol is being seriously considered for transport fleets where the savings can outweigh the inconvenience of special filling stations.

The point is that fuel-users are continually developing anyway, so their requirements are evolving. It is up to those who are evolving coal processes to work hand-in-hand with the future users of their products, so that both types of evolution go in the same direction.

Table 8 Transport fuels

CH_3OH	S% as synthesized is lower than pollution regulations require.	Use directly in c13:1 compression ratio engine: or blend with other gasoline components. Water dilution must be prevented
Diesel-light engines	S c 0.1% or less. N according to local pollution regulations.	Low viscosity, pour points etc. Flash point especially important. Boiling range 325-550°F. Higher % of paraffinic components
Diesel-heavy engines	S c 1% or less. N according to local pollution regulations.	Higher viscosity, pour points, flash point; and storage stability important. Boiling range 350-800°F. Higher % of aromatic components
Gas turbine-land	S c 1% or less. H minimum 11.3%. Low metal contents in ppm. N according to local pollution regulations.	Carbon residue low (under 1%). Other characteristics as for diesel fuels
Jet engines	S low, mercaptans below 30 ppm. Olefines below 5%. Polycyclics below 5%. CV 42.9 MJ/kg minimum. Low metal contents.	Some (small %) monocyclic aromatics, but aromatics generally below 20%. Paraffins and naphthenes desirable
Gasoline engine	Octane number balance. S 0.1% or less. N according to local pollution regulations.	Octane number control by aromatics, and branched paraffins. Previously enhanced by TEL, in future by MTBE

The way ahead

The costs of coal-processing plants will be in the range of hundreds of millions of pounds; market preparation will be involved, and process selection will be difficult in view of the apparent surplus of some options and the need for a range of developments in work-up for some of the products. There will be need for further developments in effluent treatments, coal handling, catalysis and poisoning, and, possibly pretreatments of coals. There will also, presumably, be new chemical processing routes with new feedstocks (such as methanol) to

Table 9 Typical liquid coal-based feedstocks and fuels

Process producing liquid feedstocks	*Components*
Synthesis	Motor fuels (including methanol) Alcohols Olefines (for PVC, PE etc) Ketones Aldehydes etc
Solvent refining	Motor fuels Boiler fuels Naphtha-type fractions Aromatics – and for chemicals
Hydrogenation	Motor and aircraft fuels Higher paraffins Lubricants Solvents

(*Note* – it should be cheaper to burn coal, or solvent refining residue, in large boilers than to manufacture special liquid fuel for this purpose. The combustion processes would include fluidized combustion either at ambient pressure for simple cycles or under elevated pressure for combined cycles. It is a matter of some uncertainty whether gasification cycles solely for steam boiler use will eventually be competitively economic even though they may be so in special cases in the short term.)

replace those based on ethylene and acetylene in the past.

The way developments have been supported suggests that the real power lies in the market. Each time a need crystallizes the customer will descend *deus ex machina*, and pick the then chosen process from the wide range available. However, the coal liquids so far produced by most of the processes under development cannot be used directly to replace oil products as now used. They all require considerable work-up to suit their markets, as a result of process thinking, and starting by considering new ways of handling coal.

But there is another approach. Simplicity can be worth more than complexity, or fascinatingly involved process flow sheets; and elegance is usually easier t handle in practice than sophistication. A produc oriented approach to process selection has man advantages, at least partly on the ground that whateve the concept it becomes more complicated in th flowsheet, which becomes more complicated in the plan layout, which becomes more complicated in operatio which becomes more complicated in the product work up, which makes the products more expensive in th catalogue, which leads to lower return on capital, and t commercial doldrums.

Table 10 Some typical fuel and feedstock gases

1 Btu/scf equals 0.0323 J cm⁻³	CV Btu/scf	SG Air =1	CO_2 %	CO %	H_2 %	CH_4 %	Other combustibles %[1]
Coal Gasification[2]							
Producer, 1 Stage	160	0.9	4	30	12	2.5	0.2
Producer, 2 Stage	170	0.8	4	30	15	3	0.3
Winkler, air	100	0.9	10	15	15	0.5	—
Winkler, O_2	230	0.75	28	30	40	1	—
Flesch-Demag	145	0.9	5	30	15	—	—
Ruhrgas, air	100	0.9	5	20	10	—	—
Ruhrgas, O_2	250	0.8	20	50	30	—	—
Rummel, 1 shaft O_2	280	0.7	14	45	40	1	—
Rummel, 2 shaft, air	270	0.6	8	23	46	3	0.5
Lurgi, O_2, dry ash	290	0.7	25	18	40	10	0.5
Lurgi, O_2, slagging	340	0.8	6	59	26	7	0.2
Koppers-Totzek, O_2	270	0.7	12	50	35	—	—
Lurgi plus GRH	500	0.5	14	5	53	21	6
Shell-Koppers	310	0.7	1	65	26	—	—
Texaco-GVV	270	0.7	12	46	38	—	—
Coke Gasification[2]							
Producer gas	130	0.9	5	30	11	0.5	—
Blue Water gas	295	0.5	5	40	50	1	—
Flesch-Demag	145	0.9	5	30	15	—	—
Lurgi, O_2, dry ash	260	0.7	25	23	45	2	—
Lurgi, O_2, slagging	290	0.7	5	63	26	1	—
Blast furnace gas	90	1.0	11	27	2	—	—
CWG gas oil	500	0.7	5	30	36	8	11
Coal Carbonization[1]							
Low temperature	890	0.6	4	4	30	50	6
CVR steamed	475	0.5	4	18	50	20	2
Coke oven gas	525	0.4	2	7	54	28	3
Horizontal report	560	0.4	2	8	52	30	4
CVR, no steaming	540	0.4	3	9	54	25	3
Oil Gasification							
(a) *Cyclic*							
Onia-Gegi, 950 sec oil	500	0.6	11	20	50	11	5
Segas, 950 sec oil	500	0.6	11	16	48	14	6
Jones, steam, heavy oil	520	0.4	3	16	49	23	3
Hall, steam, heavy oil	1000	0.5	5	2	18	30	30
(b) *Continuous*							
Shell O_2, heavy oil	310	0.6	4	47	49	0.3	—
Texaco O_2, heavy oil	300	0.6	6	54	39	0.1	—
GLC, air, gas oil	510	1.0	3	6	2	12	17
Hercules, steam, CH_4	500	0.5	5	7	42	33	—
ICI, steam, naphtha	300	0.4	20	5	69	6	—
ICI, plus CRG	500	0.5	14	3	49	34	—
ICI plus GRH	500	0.5	13	6	56	18	7
ICI, 'Town Gas'	500	0.5	3	14	34	34	—
Catalytic Rich Gas	840	0.5	1	1	21	77	—

Notes—1. Includes 'unsats', hydrocarbons, aromatics etc.

2. Coal means UK black coal, and coke, as made from such. Brown coals and the reactive semi-cokes made from them will tend to yield gases lower in CO_2 and higher in CH_4 and other combustibles.

(*Note*—Analyses depend on reaction conditions, and on the extent of CO shift, and CO_2 removal. Figures are indicative.)

It is likely that the simplest route from coal to produc will have the ultimate economic advantage when *all* th costs are put together. The multi-product routes ma find it hard to compete. Cogas for medium CV gas slagging (fixed bed, or suspension) gasification for hig CO synthesis for high CV gases (eg methane), o chemical feedstocks such as methanol, may well b cheaper overall than most of the routes that produc mixed liquid products directly.

The fundamental advantage of the single product synthesis is that market need can be supplied directly wit minimal work-up. Processes that make a variety o products need to work-up and place each produc separately on the market – and the result can easily b that several of the products realize prices belo economic costs. A two-product industry, such as cokin and gas-making, gives trouble enough since optimu demands for the two products hardly ever coincide. single, controlled product can be much better to handl than the ranges of products of different specificatio from so many of the coal-using processes now bein developed.

The conclusion from this may, perhaps, be surprisin It is *not* that all process routes should be abandone except the synthesis ones. It is that, whatever the mos elegant approach to handling and processing coal, th approach to marketing should be similarly simple and, possible, elegant.

For example, super-critical solvent extraction ma well turn out to have decided advantages despit probably producing a range of possible products.

In the coal-processing industry of the future there wi be not one preferred process, but a spread of types o plant and route.

There is no quick answer to process selection; and o prices are not yet low enough in relation to coal prices t make the much more expensive coal processin economically attractive. But coal's time will come whe coal prices are realistic and coal processing system avoid all unnecessary processing costs. The questio will be 'what processing is both necessary and sufficie to satisfy market needs?' Then the question of adaptati to market requirements will arise. It should simply be matter of minimizing the overall cost of doing the job - firing the boiler or furnace, powering the motor car o truck or plane or train, or making the plastic or fertilize etc. User and feedstock producer can work togethe when the coal processor convinces the user that re costs have to minimized. Incidentally, the appropriat national and international standard specifications an codes may then have to be adjusted.

Conclusion

Coal processing in the future will be quite different fro what it has been in the past: and so will be the fuels an feedstocks produced. There will be some parallels wit the oil industry as it emerged, with different crudes fro different oil fields (and with different impurities-) – eac possible product seeking the markets as they emerge.

Table 11 Some typical hydrogenation and synthesis gases

		Hydrogen for hydrocracking		*Hydrogen for edible oil hydro-genation*	*Ammonia synthesis gas*	*Methanol synthesis gas*	*Hydrogen for electronics etc (small scale)*
		From partial combustions	*From steam reforming*				
Composition							
H_2	mol %	98.4	96.5	99.5	74.2	68.5	>99.999
CO		—	—	—	—	14.6	
CO_2		—	—	—	—	13.0	Purified
CH_4		1.2	3.5	0.5	0.7	3.9	by
N_2		0.2	nil	nil	24.8	—	palladium
Air		0.2	nil	nil	0.3	—	diffusion
Total		100.0	100.0	100.0	100.0	100.0	—
Trace impurities							
CO_2	ppm mol	<20	<20	—	<5		—
CO	ppm mol			<10			
Sulphur	ppm w/w	<2	<2	1	<2	<2	—

Synthesis of hydrocarbons on supported Group 8 catalysts

Temperature (°C)	*Pressure (psig)*	*GHSV (h^{-1})*	*Catalyst composition*		*C efficiency (%)*				
			%Rh	*%Fe*	*Methane*	*Methanol*	*Acetaldehyde*	*Ethanol*	*Acetic acid*
300	1000	4000	5.0	—	37	0	24.0	16	20.0
300	1000	3900	—	1.0	69	12	0.3	10	0.2
300	1000	2700	2.5	1.35	51	19	0.6	23	0.6
300	1000	3000	2.5	0.68	44	20	1.0	30	1.3
325	2500	11000	2.5	—	41.0	1.6	5.2		49.0
325	2500	9500	2.5	1.35	39.0	13.0	37.0		4.0
325	2500	9300	2.5	0.135	5.2	2.3	13.0		31.0
325	2500	8700	2.5	0.135	46.0	1.4	15.0		36.0

Fischer-Tropsch synthesis with various catalysts

	Catalyst no		
	I [a]	*II* [b]	*III* [c]
Process conditions			
Temperature (°C)	190	190-210	315
Space velocity (per hour)	100	250	700
H_2/CO ratio	2:1	1.3:1	2:1
Pressure (atm)	1	20	18
Reactor	Packed	Packed	Fluidized
Recycle ratio	—	10	2
Product composition (%)			
C_1	13	12	9
C_2	2	6	8
C_3-C_4	9	25	30
C_5-C_8	25	22	32
C_9-C_{16}	35	17	15
C_{17}-C_{19}	6	18	6
C_{20+}	10		

(a) 100 Co:18ThO_2:100 kieselguhr.
(b) Fe_3O_4-Al_2O_3-K_2O (fused).
(c) Fe-oxide-K_2CO_3 (reduced).

and developed, and largely outside the coal and carbonization fields: but coal processing now will have to fit into the structure that oil has already laid out. It will have to face tight specifications, pollution controls, and small profit margins.

Coal in future will need to substitute for natural gas as well as for oil, as the worked deposits of both progress along the path of increasing cost of extraction. Similar factors, too, will affect coal costs, as pollution, amenity considerations of sites, movements of labour etc all have their effects. But the size of the long-term coal reserves makes it certain that coal will in future supply carbon for much chemistry, and initially, too, for endothermic reaction heat – though moving towards energy from renewable and from nuclear resources to supply driving forces for endothermic reactions.

So coal processing will be in transition for some decades while the economic need to develop resources balances itself with the need to supply evolving market demands, and the processes themselves develop and evolve to form an entirely new industry.

The need is clear; the path can be derived in the light of technological possibilities and market needs; good planning will be both important and difficult; the costs will be enormous and much greater than mere process development; and the timing will be determined by the

LETTERS TO THE EDITOR

Tracing those engineers forced into emigration by the Nazis, 1933-45

The . ssociation of German Engineers (VDI) has embarked upon research into the fate of former colleagues who were forced to emigrate under the Nazi regime, with the intention of informing the general public about what happened to them and to their families. Under the chairmanship of Dr Karl-Heinz Ludwig, professor in history at the University of Bremen, a committee has been formed, whose membership also comprises former victims of Nazi persecution.

Names and addresses sought

A first step must be to try to obtain the names and addresses of engineers who emigrated – either directly from them, or from others who can help locate them. Any information at all will be greatly appreciated and this would be followed up by a questionnaire about conditions at that period and subsequent effects of emigration. This will help as a basis for an account of the lives of our colleagues during those dark days. No names, however, will be mentioned without permission. Communications, please, to be addressed to:

Dr Koenig, Technik-Geschichte, VD
Postfach 1139, 4000 Düsseldorf, W
Germany.

W Goldstein (Senior Fellow)

Waste heat recovery

I believe that fellow members of the Institute of Energy who work in the area of waste heat recovery, will be interested to know that the famous book by Prof H Hausen is being published in English for the first time, early this year.*

The book, *Heat transfer in co-current, counter-current and cross flow* was first published in German in 1950 by Springer Verlag. Unofficial English translations of parts of this book were prepared and circulated among some people working in the field, but these were never published. In 1976, Prof Hausen published a thoroughly revised 2nd edition in German which has now been translated by Mrs M S Sayer; I have been responsible for the editing of th English translation.

It may be of interest to note it h been necessary to include translations a number of sections from the 1 edition. Reference is made to these Prof Hausen in the 1976 version ar thus it seemed to the translator ar myself that these should be included appendices to the translation.

I understand a translation of the 19 edition has now appeared in Russian.

A J WILLMOTT PhD (*Member*)
Dept of Computer Science
University of York

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*The English version of the 2nd edition being published by McGraw Hill – Ed.

Feedstocks and fuels from coal *cont*

result of politically unforecastable stresses. But the job will have to be done.

There will be advantage in simplicity of approach – to the chemical engineering of the processes, to the fitting of the products to the markets, and to the economics. It will be the economics that unites chemical engineers and customers in the search for the most useful ways of satisfying the needs for fuels and feedstocks.

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The Institute of Energy

Future events reminder

20 April – **Annual Luncheon**
Inn on the Park, W1
See this issue for ticket application form

27 May – **The Council/Sections' Affairs meeting**
followed by
AGM
(at 18 Devonshire St, W1N 2AU

16/17 Nov – Two day conference and exhibition at Kensington Conference Centre
Coal – new industrial opportunities

(Watch future issues of *Energy World* for programme and registration details)

Appendix 6

Discussion with the Briquetting Group on Formed Coke - (28.8.84 am)

Dr Agrawal described how : 75% of Indian coal is non-coking,
volatiles 20-30%
fixed carbon 30-50%
ash 20-50%
moisture up to 20%
caking index under 4

: Some 10% of the coals are "coking" - and are washed down to 12 - 20% ash so the coke still contains 25% ash. 15% are medium-caking

: Coke requirements for steel are rising and the processes are of low efficiency - 900 kg/1 tonne of steel. Imports from Australia are useful but expensive - costing 4 times the price of Indian coals

: Apparently the Indian coals are not well characterised as regards washability

: Much Indian coal is destined for domestic use, where the fuels' products of combustion permeate the room because the appliances have no flues; open braziers are common

: A serious problem in India was shortage of coking coal and possibilities arising in discussion included:

a) import highly-caking coal to blend with local non-caking coal

b) briquette coals with binder and then carbonise

c) Lignite briquetting in Assam

d) Separating petrographic constituents of coal: eg. one sample of 27% ash when size graded has led to a + 3mm fraction of 24% ash and a -3mm fraction of 30% ash. This suggested a possibility of fusain or durain separation. It would be interesting to examine a -1mm fraction. If it proved possible to separate fusain and durain from clarain and vitrain, then the latter might prove useful for coking. Also, ash may reduce coke strength .

e) It was just possible that solvent refining might concentrate a coking fraction of coal.

Discussion also included:

i) increasing output from coking plants - and UK work on size consist control, oiling and stamp-charging was mentioned

ii) briquetting - and UK work on mechanical and hydraulic reciprocating presses, rotating-table presses and screw extruders was outlined; also dish peletisers

iii) binders - and UK experience with pitch, bitumen, sulphite lye and starch was discussed

iv) formed coke - and UK work on the FMC, and BBF cokes, and that by Formed Coke Ltd was summarized

v) "smoke-removing" processes - and UK developments such as the Phurnacite and the fluidized carbonisation and mild-heat-treatment processes were compared and contrasted with regard to the Inichar, Lurgi/Ruhrgas and other processes

The Consultant's (joint) papers on "Coals for Carbonisation Processes" and (in draft) "Indigenous Rural Power Centres" were given to RRL.

Appendix 7

Discussion with Coal Gasification Team

(29.8.84. am also 31.8.84. am)

Practicalities of experimental operation of the gasifier and its ancillaries were discussed with the team. The members had acquired good and useful experience both during the commissioning of the plant and in other places (GDR, CSSR, FRG, UK, USA etc). However, although RRL staff had operated many pilot plants on a similar scale, it was recommended that, in view of the complexities, unknowns and safety implications, there should be a rigorous approach to operation and recording. Oxygen, for example, could be a major hazard. Safety, whether as risk from instruments suspended on light pipework (instead of being secured), or from tripping hazards on gantries, or from over-travel on hoists or swinging loads, or from reliance on control systems (instead of intrinsic safety), or perhaps hazardous mixing in pipework or reverse flows, or misplacement of the gasifier reaction zone, or by-passing in the bed - should be a specific concern, and the responsibility of one person.

The Lurgi Operation and Maintenance Manual (which also contains an account of commissioning and start-up over three months) covers the installation well. The recommendations made during this visit concern principles, and the programme should be regarded as in harmony with, and additional to those points made by Lurgi.

Gasifier - Practicalities

The following recommendations were made

Mass balances and Heat balances are both important for experimental rigour — instrumentation should be arranged so that these balances can be prepared for every unit - and for the whole installation - so as to check upon matching between the units.

Safety should be in the charge of one person - every item should be considered from a safety point of view, and interaction too. eg. is there any possibility of oxygen passing into steam lines, or upwards into coal, in event of, say, failure of the steam system? Safety procedures should be drawn up, and rehearsed systematically and frequently.

Planned Maintenance should also be in charge of one person and a schedule drawn up and carried out.

Monitoring of gasifier performance is useful - criteria for assessment include - eg.

- Coal feed rate kg/h
- O_2 and H_2O flow rates (also a safety matter) kg/h
- O_2/H_2O ratios (also a safety matter)
- Coal conversion % also kg/1000 MJ gas CV
- Steam decomposition % also kg/1000 MJ gas CV
- Oxygen usage kg 1/1000 MJ gas CV
- Hot and cold gas efficiencies gas CV as % of thermal input - at gasifier outlet (including tars and liquors) and after gas-cooler.
- Waste heat as % of thermal input

Emergency shut-down and crisis procedure

- there will be different techniques worked out according to cause - eg. failure of power, O_2, H_2O or steam supplies, fracture of pipes etc; and a technique for voltage drop (as it can drop from 440v to 360v or so) or perhaps an air supply to keep the gasifier going in the event of interruption of O_2 supply. Emergency shut-downs should be practised.

Wear and corrosion measurements should be recorded from these early days - gasifier internals, lock-hopper stellite valve faces, grate mechanism, Benfield plant etc etc.

Blockages and fouling will occur, and these should be both alleviated and recorded, along with observations on the effectiveness of measures tried.

Recording system - it is desirable to set up a comprehensive recording system right from the beginning. As one detail, a number of changes have been made to the plant already: the flow sheet should be kept up to date.

Power Supply - any experimentation where the power supply can fall, say, to 360v (as does, apparently, happen) will be difficult, and might on occasions be dangerous. Power cuts, too, will create problems. Emergency power supplies would be highly desirable - and their design would need to be carefully worked out.

Minimum requirements for Heat and Mass Balances

Samples - coal, incremental samples aggregated, coned and quartered, air dried(recorded); and despatched to the laboratory for CV, proximate and ultimate analysis

- ash - sampling as for coal, but analysis mainly for % carbon content and for sintering/clinkering propensities
- gas etc - as recorded at appropriate points to suit the units involved
- tars and liquors - sample points at selected places, incremental samples aggregated, and laboratory containers despatched for analysis and CV

Flow rates

- coal (by weight over duration of test)
- ash (by weight over duration of test)
- air, O_2, steam to gasifier
- water into (or steam out from) jacket and waste heat boiler
- gas(es) at appropriate points to suit the units involved

- tars and liquors, at appropriate points to suit the units involved
- lock-hopper gas (measure holder)

Temperatures

- ambient
- beneath coal lock
- gasifier outlet
- ash lock
- steam/oxygen (or air) mixture
- at different heights above the bed (gas samples too) for analysis of bed/reaction conditions

Pressures

- everywhere there is steam, air, oxygen, gas or lock gas, flow and temperature measurement

Discussion with Coal Gasification Team 29.8.84.

Present

K Seshagiri Rao	Coal Gasification
T S R Anjaneyulu	Mechanical Engineer
G Ravindranath	Mechanical Engineer
M A Khaleelakmal	Chemical and Mechanical Engineer
M S Subrahmanyam	Mechanical Engineer
M M Mallikarjunan	Chemical Engineer
K B S Prasad	Chemical Engineer
K B S Saisabu	Fluidbed Gasification
K H V Prasad	Thermo Gravimetric Kinetic Studies
B Madhusudhan	Chemical Engineer
J V Prasad	Mechanical Engineer
L N Kota	Coal Gasification
S N Reddy	Coal Gasification
K V R S Musty	Coal Gasification
A Satyanarayana	Chemical Engineer

Appendix 8

Discussion and Lecture Headings on Gasifier Development and Scale-up - (31.8.84. am)

General

The plant includes, eg. coal receipt, preparation and handling; O_2 and steam production; receipt, handling and usage of other raw materials; gasification; waste-heat usage; by-products; gas cleaning; synthesis; recycling; product work-up; storage and despatch; effluents; management; legalities.

Issues for the process as a whole -

Flexibility, effluents re legislation, provision for unit failure, intermediate storage, interpretation of energy inputs and outputs, optimisation - computer-aided-flowsheet-design, hazards, controls philosophy, safety, emergency shut-down, failure of utilities, capital costs and optimisation, provision for monitoring.

Issues concerning process units -

Models of variables - and confirmation, check tests elsewhere or even on rigs for single values. Transparent models for visual demonstration and measurement, Radio active tracers for bed position measurements. Size-related issues - eg. heat transfer, surface/volume factors, pressure drops, flow, static similarity, dynamic similarity.

Performance-related issues - eg. mass throughput, impurities build-up (amount vs proportion) eg. C_2H_2 in O_2 plant, or CN in gasifier recycles, erosion, catalyst poisoning.

Time-related issues - eg. fouling of surfaces, sludges, corrosion, wear, packing flexibilities.

Issues deriving from interactions of variables -

Hunting and resonances of flows, piping system design, effects on maintenance, safety schedules, spares stocks, operating shifts and labour, management, performance monitoring, continuing process improvement.

Recommendations made to RRL

Gasifier etc.

1) It is essential to have a safety discipline and also emergency shut down procedures.
2) A planned maintenance scheme should be instituted.
3) The use of enriched air should be considered, to compare the economics with those using "pure" oxygen.
4) Recycling of tars and liquors should be considered.
5) Blending coals may be useful to extend the range.

A Typical Gasifier Programme

1) Continue the present mode of operating, incorporating the recommendations made herein - particularly as regards safety, maintenance, instrumentation, heat and mass balances.
2) Taking one coal, study the reaction zone geometry at different H_2O/O_2 proportions and different flow rates. The coal should preferably be at one end of the acceptable range of caking characteristics - say the non-caking end.
3) Then repeat 2) with several other coals throughout the range of acceptable caking properties. This will indicate how far it will be possible to operate on Indian coals without resort to stirring. Precarbonisation tests will be useful.
4) As a result of 3) it will be possible to decide whether or not to install a stirrer.
5) There may be opportunity to offer a gasification testing service to other organizations and countries. Useful revenue could be earned this way: and such initial work could lead to consultancy services.

Possiblities where the Consultant may be able to make useful connexions in Britain and elsewhere

1) Obtaining obsolete gasifier spares (a problem for RRL)
2) Experience with biological oxidation of Lurgi gasifier effluent.
3) Possible Fellowships in Britain - Universities (eg Sheffield or Imperial College) or British Gas

Those present at the Discussion and Lecture on Gasification Development and Scale Up 31.8.84.

1.	K Seshagiri Rao	Coal Gasification
2.	Mrs Aziz - Mirza	Coal Division
3.	B S Narayana Rao	F T Synthesis
4.	Y V Subba Rao	Gas Extraction
5.	M S Subrahmanyam	Coal Gasification
6.	G Ravimoranath	Coal Gasification
7.	M A Khaleel Akmal	Coal Gasification
8.	M Ehran	Coal Gasification
9.	A M E Rahman	Briquetting
10.	C S N Prasad	Briquetting
11.	A Krishna Reddy	Design & Development
12.	A Satyanarayama	Coal Gasification
13.	D A K Sivazi	Chemical Engineer
14.	A. ?	Oils and Fats
15.	H Shripathi Rao	Oils and Fats
16.	IV K Dutt	Oils and Fats
17.	S A Hamud	Coal Engineer
18.	T S R Anjaneyulu	Engineer
19.	B Madhusudhan	Chemical Engineer in Coal Gasification

20.	T Krishradu	Coal Gasification Programme
21.	J V Prasad	Mechanical Engineer
22.	D Rambabu	Data Processing
23.	S N Reddy	Coal Gasification
24.	M M Mallihajunan	Coal
25.	K V R S Musty	Coal Gasification

Appendix 9

Discussion and Lecture Headings on Evaluation of Gasification Processes

Evaluation involves looking ahead and asking questions so as to decide upon criteria and priorities. Three main components of evaluation, in logical order, are -

a) markets, coals and suitability of processes - gasification and subsequent (and also preliminary) processes

b) principles and practicalities - industry infrastructure, environmental issues, reliability and risk, complexity. (This leads to a short list of processes)

c) economics

a) Markets, coals and process suitability

Consider how markets for products may change over the years of planning and building (5?, 10?, 15?) and operating (20?, 30?) the plant. If development is involved (5?, 10? years) then its aims may change during its course. Coal supplies, too, will change. Some processes are (probably) towards the end of their evolution and can offer much design and operational experience. Others may be able to offer more potential for improvement, but with a shorter history of success. Then, processes need to fit together to make a flowsheet - do unit outputs and inputs match easily as regards analysis, temperature, pressure? Are by-products worthwhile? Are some units very critical and sensitive? Is one type of unit more suitable than other types? (eg which type of oxygen plant will be most suitable - gaseous, liquid or pressure-swing?)

b) Principles and Practicalities

Is the plant to be part of a more general scheme - eg of using a particular fraction of coal, or to make products critical to other works? Is the scale critical, and should there be a few large or many small units? Should the plant be near the coal or the user of its products? (eg if C_2 - C_4 olefines are made, who will use them? Can the coal be used directly, or does it require special pre-processing? Rank ash, size and impurities are important in process evaluation and selection. What deliberate plant redundancy is required - to continue output in event of failure (or maintenance) of a unit - or for safety reasons? What are the replacement and maintenance or regeneration intervals for components, units, catalysts, absorbents etc. What are the gaseous and liquid effluent implications? manpower needs? top grade management needs?

Also what is the potential for process improvement during operation?

These issues lead to notional plant design, and to the bases for economic evaluation. A "feel" will be obtained for the types of process most likely to suit the coal, the final product, and the circumstances.

A short list of processes can now be drawn up, comprising the best choices from the wide range of possible processes

c) Economics

Previous considerations will have led to a short list. It will probably consist of some of (in the case of the RRL project) : Lurgi (dry ash); BGC Slagging Lurgi; Winkler; and Koppers-Totzek that have operated over sufficiently large tonnages of coal to be considered conventional. Texaco G V V , Shell-Koppers and Cogas can be considered reasonably well developed: and there are many other ingenious (and some not so practical) processes on which much effort has been spent, and of which some may become commercial in future. However, the more conventional the process, the better the economic data - capital and operating cost - and the less the risk.

A convenient approach to economic evaluation is to list for each process considered, for a given notional works output:

Basic Data

Gas (or product) output per unit or section
MJ/hour, or tonnes/day ______

Number of units or sections ______

Works output (the same for each process considered)
MJ/hour, or tonnes/day ______

Site area for works Hectares ______

Any pecularities or requirements regarding the site ______

Cost of fuel (including transport and preparation
elsewhere) per unit weight - Rs per tonne ______

Capital cost of works Rs ======

Process Economic Data

Cost of fuel (gasifier only) - Rs per tonne of output ______

Cost of oxygen (including power) - " " " ______
Cost of steam (including water) - " " " ______
Cost of power (excluding O_2 plant) - " " " ______
Cost of water (excluding boiler water) - " " " ______
Cost of catalysts, bought-in items, chemicals - " ______
Cost of labour (including maintenance) - " " " ______
Capital charges - " " " ______
Gross cost of gasification/operation - " " " ______
Less credits for steam sold elsewhere - " " " ______
by-products " " - " " " ______
off gases " " - " " " ______
Net cost of gasification/operation - " " " ======

Many of these figures may involve lengthy calculation, so computer aids may be useful.

Then comes the soul-searching - how practically real are the figures? Is there any bias or personal preference? Is there any tendency to keep figures low to suit a known budget?

Finally, possible errors must be considered. Each final process cost has evolved from adding and subtracting figures, each of individual "tolerance" according to the accuracy (and relevance) of its component information. Accordingly, the difference between competing process costs should be considered carefully in the light of the likely tolerance of the figures being compared.

A practical example

An example of economic and technological evaluation of gasification processes, based on the Lurgi process, was given in the form of an outline of a joint exercise in Britain by the (then) Gas Council and the National Coal Board. The Joint Lurgi Study Group provided three contractors with common data and specifications (as described below). The three contractors (given the description 'A', 'B' and 'C') produced designs and costs, that the Joint Study Group adjusted to achieve true comparability. Further to illustrate the method, the process designs are also summarized below, along with the Study Group's comparisons of sites; coal data; the "site cost" of gas-making; biological treatment of effluent and its costs; process data; and the costs of steam and power, investment and lean gas production.

Economic and technological evaluation of different Lurgi gasification process flowsheets - for the same site and coal

Examples from the report of the Joint Lurgi Study Group

appointed by the Gas Council (UK)
and the National Coal Board (UK)
October 1963

Beginning with the initial notional cost of gas related to a particular site, Desford, the Lurgi Study Group examined sites in some detail (Moorgreen, Desford and Newstead) From these Newstead was selected as the best - see Tables I and II. Then, it provided the same basic gasification and other data, and the same ground rules, to the three contractors (A,B & C) - Tables III and IV. The schemes of the three contractors were then evaluated, - Tables V and VI, and compared as summarized in Tables VII to XII.

Table I

Analytical Data

Colliery		*Moorgreen*	*Desford*	*Newstead*
Moisture	%	5.6	11.0	10.7
Ash	%	19.0	18.4	24.2
Volatile matter	%	29.1	29.0	25.5
Carbon dioxide	%	0.5	1.2	—
Fixed carbon	%	45.8	40.4	39.6
		100.0	100.0	100.0
Sulphur	%	1.7	1.5	0.5
Chlorine	%	0.44	0.23	0.27
Calorific Value	B.t.u. lb.	10,692	9,504	8,827
Ash Fusion	°C.	1,279	1,190	1,400
Ash Fusibility:	comment	Satisfactory	Low	Satisfactory
Coke Residue (various tests):	comment	Weakly caked: sintered: frothy, expanded, particles little changed.	Sandy: weakly swollen, many unchanged particles, loose.	Weakly caked: sintered: many unchanged particles, loose conglomerates.
Swelling and caking properties:	comment	Less satisfactory than alternative coals. Two constituent seams not recommended.	Satisfactory except for Merrylees component.	Satisfactory.

Table II

Gasification Data and Cost Index

		Moorgreen Composite 1965	*Desford Composite 1965*	*Newstead Composite 1965*
Comparison of Gasification Data (Based on lean gas leaving the gasifier)				
Gas	therms per gasifier per hour	2003	2349	2098
Gas	therms per ton of coal (dry ash-free basis)	250	242	245
Coal	lb. (dry ash-free) per therm gas	8.96	9.24	9.16
Oxygen	cu. ft. per therm gas	53.1	51.6	52.6
Steam	lb. per therm gas	16.4	17.0	12.4
Gasification Cost Index (Desford 100 units)				
Coal		60.2	59.1	58.9
Oxygen		17.8	16.9	17.2
Steam		18.8	19.5	14.3
Capital charges on Lurgi gasifier		5.2	4.5	3.9
Gasification Cost Index		102.0	100.0	94.3

The Newstead site was selected

Table III

Biological Treatment of Lurgi Effluent Liquor from Dorsten

		Without CO Shift		*With CO Shift*	
		I	II*	III	IV*
		Before Treatment	*After Treatment*	*Before Treatment*	*After Treatment*
Phenols	p.p.m.	260/390	14/24	230/260	11
Ammonia	,,	1250/1600	60/140	1650/1700	50/70
Thiocyanates	,,	160/230	Nil/3	86/97	Nil/2
Permanganate value	,,	600/700	64/76	450/500	31/43
Biological Oxygen demand	,,		24/25		12
Suspended solids	,,	—	60 approx.	—	60 approx.
Retention time	hrs.		15		7

*Including removal of fixed ammonia by distillation with lime.

Table IV

A Summary of Design Study Specification

1. **GENERAL**

The plant was to be designed to produce purified lean gas at minimum cost using adequately established technical means under guaranteeable conditions. The plant output was to correspond to the gasification of a nominal 750,000 tons of (specified) coal per annum.

2. **SITE**

The site between Newstead and Linby collieries was specified, and its technical and engineering characteristics were enumerated in a site report issued to contractors.

3. **DESIGN**

a) Standby provisions were to be such as to suit the requirements of a plant to operate as the first of a national scheme of gas production. A minimum of three process streams was specified.

b) **Load factor.** 86.5% was specified. In addition estimates were also to be given of lean gas costs at an average load factor of 70%.

c) **Gas Quality.** The specification was given for a lean gas which on enrichment by various methods would give a satisfactory town gas.

d) **Processes.** The gasification was to be based on data supplied by the Lurgi Company; and the carbon monoxide conversion was to be by the raw gas shift based on the work of the Lurgi and Ruhrgas Companies. Liquid effluent was to be treated under conditions specified by the Gas Council and the National Coal Board, and after treatment to be discharged to the River Trent in accordance with the requirements of the Trent River Board. Gaseous effluents were to satisfy the Alkali Inspectorate, whose probable requirements were provided. Steam and power were to be generated on site as far as contractors calculated was reasonable. Additional power could be taken from the grid if required. Water was to be taken from the River Trent.

4. **COSTINGS**

Capital estimates were to be accurate to within 7½% on the total figure. The cost of lean gas was to be given in pence per therm along with the contractors' estimate of its accuracy. The method of steam and power costing, and the rates for capital charges, labour and maintenance, were specified.

Table IV (Continued)

.The following rates were specified for by-products:–

a) Tar and oil—as crude tar at 10d. per gallon.

b) Benzole —as crude benzole at 10d. per gallon.

c) Ammonia —as 25% concentrated liquor at 3s. 6d. per unit
(later amended to 2s. 6d. per unit).

d) Sulphur —as elemental sulphur at 180s. per ton.

e) Phenol —as crude phenol at £45 per ton.

For the purpose of the exercise, all coal was to be charged at 3.25 pence per therm irrespective of size.

5. COAL SUPPLY AND ASH DISPOSAL

Coal supply was to be taken from an N.C.B. conveyor run to the plant boundary. Contractors would be responsible for receipt, handling, crushing and grading as necessary, of the coal, and for providing both strategic and seasonal stocks. Ash was to be disposed to the adjacent colliery spoil heaps.

6. STANDARDS

Standards of design and construction, battery limits, the extent of ancillaries, and obligations and conditions of contract were all specified.

7. PRESENTATION OF REPORT

Details were given of the categories into which the capital and the operating costs were to be itemised. Certain specimen costs were to be given and also the costs of removing carbon dioxide to varying final concentrations. Heat and material balances, gas analyses and drawings required, and the forms of process description were all specified.

8. DATA PROVIDED TO CONTRACTORS

Contractors were given details of the make-up of the coal supply, delivery of raw coal and its screening analysis; also its probable behaviour on crushing, screening and reclamation from stock. Full analytical data and gasifier performance data of the Lurgi Company were also given. A process specification for biological oxidation was provided along with analytical details of a typical treated effluent and the requirements of the Trent River Board.

Table V

Data provided in Contractors' Design Studies

Item		*Scheme 'A'*	*Scheme 'B'*	*Scheme 'C'*
Purified lean gas production— Million therms per annum		119.6	117.0	120.3
No. of Gasifiers (inc. standby)		8	8	8
Type of Oxygen Plant		British Oxygen Co. Low Pressure	Linde Low Pressure	H.R.I. Low Pressure
Method of Carbon Dioxide and Hydrogen Sulphide Removal		Benfield Process and Tower Purifier	Adip Process*	Vetrocoke Processes
Coal Gasified	tons year	774,000	776,600	779,500
Coal used on boilers	tons/year	150,000	217,200	204,800
Steam produced by boiler plant	lb hour	311,000	450,000	390,000
Power produced (including mechanical equivalent)	MW	12	25	24.6
High Pressure Steam used for Gasification	lb/hour	196,907	196,500	196,889
Oxygen used for gasification (95%)	lb hour	72,720	73,362	72,617
No. of men employed		205	201	250

By-Products Recovered

Sulphur	tons/year	2,057	3,055	3,250
Phenol	tons year	2,791	3,468	2,652
Ammonia	tons year	7,425	6,985	6,259
Tar	million galls/year	3.379	3.032	6.403
Wash Oil	million galls/year	1.956	2.173	
Benzole	million galls year	3.085	3.424	1.403

Approximate Analysis of Purified Lean Gas

	%
CO_2	13.9—14.3
CO	4.8—5.1
CH_4	10.9—11.3
H_2	66.0—68.7
C_nH_m	0.0—0.3
N_2	1.0—3.5

CV—nominally 350 B.t.u. s.c.f.

*Including a simple tower absorber for final trace removal of hydrogen sulphide.

Table VI

Summary of Contractors' Process Designs

	Scheme 'A'	*Scheme 'B'*	*Scheme 'C'*
1. Oxygen plant	3 units × 275 tons/day. B.O.C. low pressure. Electrically-driven air-compressors.	3 units × 253 tons/day. Linde low pressure. Electrically-driven air-compressors.	2 units × 390 tons/day. H.R.I. low pressure. Steam-driven air-compressors.
2. Main boiler plant	3 @ 175,000 lb./hr. 1675 p.s.i.g. 950°F.	3 @ 225,000 lb./hr. 2400 p.s.i.g. 950°F.	3 @ 190,000 lb./hr. 1500 p.s.i.g. 950°F.
3. Gasifiers	6 working: 2 standby: 3.2m. diam.		
4. Benzole removal	Absorption by oil wash.	Absorption by oil wash.	See 8 below.
5. Carbon monoxide removal	Raw gas conversion Lurgi process-improved design.	Raw gas conversion Lurgi process-modified design.	Raw gas conversion Lurgi design.
6. Hydrogen sulphide removal	Benfield process. (Potassium carbonate solution.)	Adip process. (Organic solvent.)	Vetrocoke H_2S process. (Sodium carbonate/sodium arsenite solution.)
7. Carbon dioxide removal			Vetrocoke CO_2 process. (Potassium carbonate/potassium arsenite solution.)
8. Benzole removal	See 4 above.	See 4 above.	Absorption by oil wash.
9. Final purification	Iron oxide towers (Bischoff system).	Simple iron and zinc oxide tower (trace removal).	Integral with H_2S removal above.
10. Odorisation	Tetrahydrothiophene		
11. Drying	Activated Alumina regenerated by hot gas.	Glycol-regenerated by distillation.	Refrigeration and Glycol. Glycol regenerted by distillation.
12. Sulphur recovery	Modified Claus Kiln – from acid gas.	Vetrocoke low pressure process-from acid gas.	Integral with hydrogen sulphide removal process.
13. Phenol recovery	Phenosolvan process		
14. Ammonia recovery	Distillation of free and fixed ammonia		
15. Effluent treatment	Biological oxidation as specified by Study Group		

Table VII

Cost of steam and power— pence per therm of lean gas

	Scheme 'A'	*Scheme 'B'*	*Scheme 'C'*
Cost of purchased electricity	0.96	0.12	0.03
Cost of boiler coal	0.76	1.12	1.04
Capital charges and other costs	0.96	1.24	0.95
	2.68	2.48	2.02

Table VIII

Costs of Biological Treatment of Lurgi Plant Effluent
(expressed in pence per therm of lean gas made)

	Capital Charges	*Operating Costs*	*Total Running Costs*
Scheme 'A'	0.066	0.065	0.131
Scheme 'B'	0.035	0.030	0.065
Scheme 'C'	0.046	0.034	0.080

Table IX

Total Investment

	Scheme 'A' £000	*Scheme 'B'* £000	*Scheme 'C'* £000
Plant capital cost	15,805	17,784	15,548
Interest on capital during construction	865	970	851
Total capital cost of works	16,670	18,754	16,399
Working Capital	456	418	413
Total Investment	17,126	19,172	16,812

Table X

Cost of Lean Gas Production

	Scheme 'A' d. therm	Scheme 'B' d. therm	Scheme 'C' d./therm
Coal for gasification	4.130	4.206	4.096
Coal for steam and power generation	0.761	1.127	1.039
Purchased electricity	0.963	0.116	0.030*
Other raw materials	0.310	0.385	0.273
Operating labour	0.303	0.300	0.383
Maintenance	0.931	1.061	0.943
Capital charges	2.849	3.267	2.783
	10.247	10.462	9.547
Credit for by-products	1.179	1.294	1.101
	9.068	9.168	8.446

*Power consumed at pumping station at River Trent.

Table XI

Effect of Load Factor upon Lean Gas Cost

Load Factor	*Scheme 'A'*		*Scheme 'B'*		*Scheme 'C'*	
93% Fuel etc.	4.99		4.54		4.34	
Capital etc.	3.80		4.31		3.82	
Total		8.79		8.85		8.16
86.5% Fuel etc.	4.99		4.54		4.34	
Capital etc.	4.08		4.63		4.11	
Total		9.07		9.17		8.45
70% Fuel etc.	4.99		4.54		4.34	
Capital etc.	5.05		5.72		5.08	
Total		10.04		10.26		9.42
50% Fuel etc.	4.99		4.54		4.34	
Capital etc.	7.06		8.01		7.11	
Total		12.05		12.55		11.45

Table XII

Cost of Gas Production - Comparison with Desford Estimate

	Lean Gas *d. therm*
Desford Scheme as reported to Select Committee	9.61
Desford Scheme revised*	10.78
Newstead Scheme 'A'	9.07
Newstead Scheme 'B'	9.17
Newstead Scheme 'C'	8.45

The estimates in this Table assume a load factor of 86.5%.

*Revised to take account of cost data used in the present studies.

Appendix 10

Summarizing Discussions
-on subjects under consideration sequentially during
the week up to 1 September, 1984

1) Pressurised Slagging Gasification. The work of the UK government oil-from-coal programme of the 1950's was further outlined. This had involved slagging fixed-bed type gasifiers at the Gas Council Midlands Research Station at Solihull and at the British Gas Utilisation Research Association (BCURA) at Leatherhead. The former establishment was supplied with a 1 m dia gasifier bought from Lurgi who had been unable to operate succesfully under slagging conditions: it was adapted and continually modified until a system was derived that worked satisfactorily.

The BCURA gasifier followed experiments in slagging at ambient pressure. Various configurations of hearth and slag tap were used, and many succesful running hours were logged.

The programme eventually led up to the present-day slagging Lurgi gasifier plant at Westfield, where large commercial scale trials can be carried out.

Technical papers, and an outline of British Gas Corporation consultancy services, were given to RRL. RRL were also shown original reports and documents from UK experimental and pilot plant work. (Appendix 12)

Note - The Godavari Khani coals intended to be used for gasification and of which analyses were given (Appendix 3) would not be suitable for slagging gasification. The slags would be too viscous. With high ash contents and the need for substantial additions of fluxes the efficiency advantages of slagging would be mitigated by slag heat loss, increased maintenance, and reduced operational reliability.

The criterion used is Silica Ratio - the ratio of the silica content of the ash to the sum of the contents of (silica plus iron, calcium and magnesium oxides) which can be expressed as a percentage. From this the slag viscosity can be estimated for different flow temperatures. 250 poises is about the maximum for practicable slag flow, which is the viscosity at which slag can flow under gravity. 50-100 poises is a commonly-used range. The higher the 250 poise temperature the greater is the quantity of fluxes that have to be used to bring the slag-tap operating temperature within the range practicable for refractory life and operational control.

Godavari Khani Coals

	GDK 7A	GDK 8A	GDK 10	GDK open 4 seam	GDK open cast I 3 seam
Silica Ratio	87%	82½%	87½%	87¾%	78½%
Estimated temperature for a slag viscosity of 250 poises	1680°C	1580°C	1700°C	1700°C	1520°C

2) Fluidized Combustion. The history of fluidized combustion in UK and elsewhere was outlined, including circulating as well as conventional deep and shallow beds, and operation at both ambient and elevated pressure.

Some papers were left with RRL. This subject, interesting in itself, was of peripheral interest, and can be continued in greater depth on another occasion.(See pp 77-110 and Appendix 12)

3) Power Generation. Various cycles involving gasification techniques and also fluidized combustion, were discussed. This topic could be continued on another occasion.

4) Gasifier Instrumentation and Operating and Layout Matters -

a) General instrumentation would be improved if flow measurements were provided for

   i) oxygen and steam flow separately, into the gasifier - deriving from the flow ratio controller (FRC 4301)
   ii) feed water to the gasifier jacket
   iii) steam to the ash lock-hopper (F421)
   iv) steam into (during start-up) and from the L P steam drums (E438 and F435)
   v) liquor recirculation to the quench (on top of E431)
   vi) tar from the waste heat boiler (E431)
   vii) blowdown (if any) from the gasifier jacket (h.p.).

These are noted on the RRL flowsheet. As other sections of the plant came into use, e.g. the Benfield plant, then they could require similar consideration for instrumentation. (See Appendix 3)

b) Pipework should be colour-coded according to contents, particularly

oxygen
air
h.p. steam
l.p. steam
feed water
recirculated liquors
cooling water.

These points are noted on the RRL flowsheet.(Appendix 3)

c) Layout of equipment was such that present difficulties with tar fog in the gas were likely to continue with the present arrangement of a contiguous quench and waste-heat boiler. Although it should be possible to derive a set of operating conditions where tar and liquors would separate properly, such conditions would not be able to be maintained when experimenting with different regimes and different coals.

It would be worth considering separating the waste heat boiler from the quench; and also installing separators between quench and waste heat boiler, and also before the gas cooler. Then the plant would be more flexible in operation as the different functions would be performed separately.

d) The position of the reaction zone could not be judged in detail at present, the only bases being gas outlet temperature, gas analysis, and ash condition in relation to grate speed etc. The Dent/Gas Council approach to measuring temperatures and gas analyses was described. Conditions at different positions vertically on the gasifier axis could be measured, using small bore austenitic gas sample tubes and small diameter mineral insulated thermocouples of different lengths.

The Hoy/BCURA technique of passing instrumentation tubing into a gasifier was also described. An appropriately thick gasifier flange, (or an extra ring of flange i.d. and o.d.) could have radial holes for instrumentation unions alternately with flange bolt holes. Then the sample tubes and thermocouples (appropriately stayed inside the gasifier, which had no stirrer,) could be assembled and connected to the union halves inside the vessel. The external instrumentation connexions could be made to the external half unions. This way the instrumentation neither interfered with nor was complicated by the offering-up and bolting of the top dish section of the gasifier.

With such instrumentation it was possible to follow the passage of the reaction zone up and down the axis of the the gasifier - and to learn how to keep its location under control during operation. The sample tubes and thermocouples in the lowest positions might burn away, but that was of small importance. The tubes and thermocouples at higher positions would give warning of rise in height of the reaction zone - and appropriate operating measures could be taken.

The rate of beds descent may be measured, as distinct from merely being calculated, with the use of a radioactive tracer and detectors at different heights.

5) The Gasifier Plant Generally Basically the plant, although newly operating, was twenty years old. In the interim many changes have occurred. Instrumentation and controls have considerably developed. Also, research and developement facilities elsewhere for the Lurgi process, have fallen in disuse. These points suggest two consequences.

a) The instrumentation is rather out of date, although it works, of course. There will be a shortage of certain spares, which it would be worth trying to collect.

b) RRL Hyderabad probably now has the only small experimental Lurgi installation in the world. Other Lurgi plants are larger, and so more costly to operate on a pilot basis. As interest in coal processing is likely to increase worldwide, RRL could usefully consider offering its facility on a commercial basis. Suitably managed, the RRL installation could not only do useful work for Indian interests; it could also earn revenue by characterising offshore customers' coals as a preliminary to their spending large sums of money for testing in larger plants on a commercial scale. RRL Consultancy on gasification could be well worthwhile.

Two papers on fluidized combustion follow.

"Fluidized Combustion of Fossil Fuels" By A.G.Roberts, H.R. Hoy, H.G. Lunn & H.B. Locke. Coal Processing Technology, Volume Two, 1975.

"Status of Pressurised Fluidised Bed Technology Today - A Review" - Engineering Foundation Conference Combustion of Tomorrows Fuels - II. By H. R. Hoy, J. E. Stantan, A. G. Roberts

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COAL PROCESSING TECHNOLOGY

FLUIDIZED COMBUSTION OF FOSSIL FUELS

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and

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VOLUME 2 1975 PAGES 34-39

Fluidized Combustion of Fossil Fuels

Highly efficient and operating at low temperature, the technique reduces boiler investment, improves environmental acceptability of exhaust gases and is especially useful for burning of coal.

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and
H.G. Lunn and H.B. Locke
Combustion Systems Ltd.
London, England

Fluidization, a widely accepted technique for bringing about heat and mass transfer in the chemical and processing industries, has more recently been extended to cover the combustion of fossil fuels in fluidized beds of inert particulate matter.

In fluidized combustion, high efficiency and high rates of heat adsorption can be obtained at low combustion temperatures. The consequent advantages of this technique compared to conventional combustion plant include: reduction in boiler size and cost; reduced NO_X emission; effective SO_2 removal with low capital and operating costs; the possibility of passing the combustion gases from low-grade fossil fuels through gas turbines; and particularly in the case of coal, reduced fuel preparation costs together with a widening of the range of solid fuels that can be fired with satisfactory results.

A great deal has been written and published (*1, 2, 3, 4, 5*) amplifying the merits of fluidized combustion and dealing with the attendant phenomena in detail. Consequently this discussion will be limited to reviewing the application of pressurized fluidized combustion to electric power generation.

The present impetus to develop novel power generating systems arises from two main sources; firstly, the continuing environmental lobby, in certain areas, demanding low SO_2, NO_X and solids emissions; and secondly, the overwhelming desire since the "energy crisis" of 1973/74 for energy importing countries to become self-sufficient. This latter statement for most of Western Europe and North America implies, regardless of the status of national nuclear programs, an ongoing commitment to power generation from fossil fuels. As these fuels become more difficult to win and consequently more expensive, it will become increasingly desirable to combust the low grade solid fuels or "the bottom of the barrel."

Work on fluidized combustion started in earnest in the U.K. in 1964. The depth and scope of the R&D progra has continually expanded with major contributions to th technology being made by the National Coal Board (NCB BCURA Ltd. and British Petroleum Ltd. (BP). The prese state of the art in the UK is derived from over 20,000 hr. operating experience on a variety of pilot plant and co bustion rigs, and a substantial proportion of the expend ture to date (both capital and man-hours) has been directe towards the investigation of pressurized fluidized combu tion.

In 1972, the National Coal Board and British Petroleu together with the National Research Development Cor (NRDC) formed Combustion Systems Ltd. (CSL). Th object of CSL is to commercially exploit the know-how its parent companies in the field of fluidized combustio and to continue with the development of the technology

The number of possible thermodynamic cycles is large and only a few will be described. Broadly speaking, the can be classified into two groups. In one the working flu is air and/or combustion products; in the other there ar two working fluids—steam and combustion products.

Thermodynamic cycles where air is working fluid

Open-cycle gas turbine. This is the simplest possib cycle in which all the working fluid is used as fluidizing a and is heated to turbine inlet temperature in passin through the bed. Although this cycle represents the ult mate in simplicity, it does not utilize one of the main fea tures of a fluidized bed namely, the high heat transfe rates to surfaces immersed in the bed - and all the workin fluid has to be cleaned before it reaches the turbine.

"Air heater" cycle. This is a derivative of the simple cycle and is shown in Figure 1. Air from the compressor split into separate streams and most of it (up to approx mately two-thirds) passes through tubes immersed in th fluidized bed, where it is heated to a temperature approach

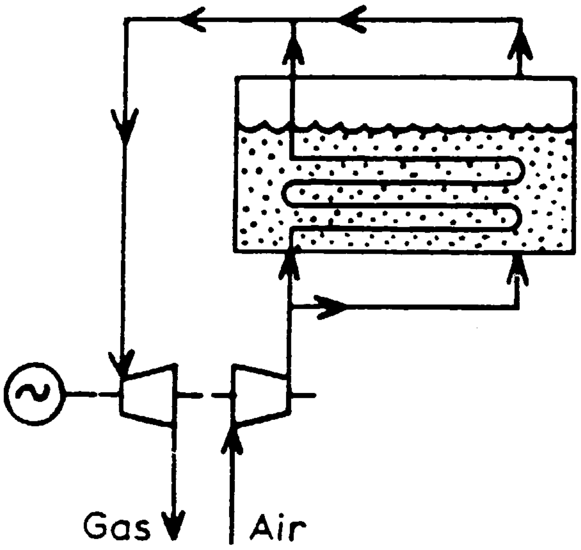

Figure 1. Simple "air-heater" cycle. Part of air is used as combustion air; the remainder is heated in tubes in fluidized bed.

ing the combustion bed temperature before being mixed with the products of combustion from the remaining air stream. Such a cycle requires a smaller combustor than the simple open-cycle, with a smaller quantity of gas requiring cleaning. It also follows that combustion is carried out at a lower excess air than in the simple open-cycle.

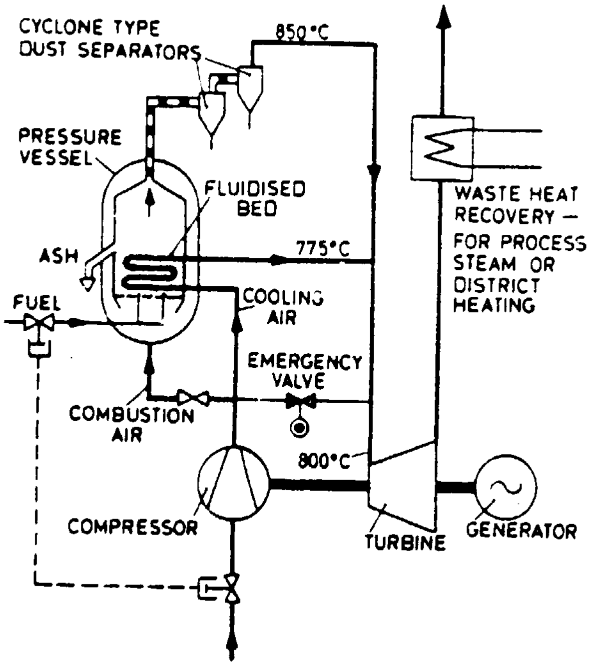

Figure 2. 66-MWe air heater cycle with waste heat recovery. Air is compressed in a two-stage compressor with intercoolers, giving a thermal output of 118 MW for district heating, in addition to 66 MW electrical power.

A scheme for a 66-MWe airheater cycle, based on an existing well-proven industrial gas turbine, is being launched by associates of CSL. A diagrammatic arrangement of the cycle is shown in Figure 2. The scheme incorporates twin fluidized combustors each with a gross heat input of 130-MW and operating at 16 atm. The bulk of the sensible heat remaining in the exhaust gases after the turbine is recovered in a district heating boiler giving a heat output of 118-MW in addition to the electrical output of 66-MW, with an overall thermal efficiency of 72% and a power generating efficiency of 26% (27.2% if the district heating boiler is omitted). Estimates indicate that the capital cost of this plant would be only 65% of the capital cost of an equivalent coal-fired steam plant with back pressure turbine and stack gas cleaning.

Closed cycle gas turbine. In this cycle, the air heated in the tubes forms part of a closed-cycle unit. Generally such units operate at a lower pressure ratio than conventional gas turbines, but at a significantly higher pressure level. This leads to substantial increases in the heat transfer coefficient between the air and the tube wall and hence to a reduction in the amount of heat exchange surface required. The combustor can either be pressurized, in which case the combustion gases pass through an open-cycle gas turbine, or it

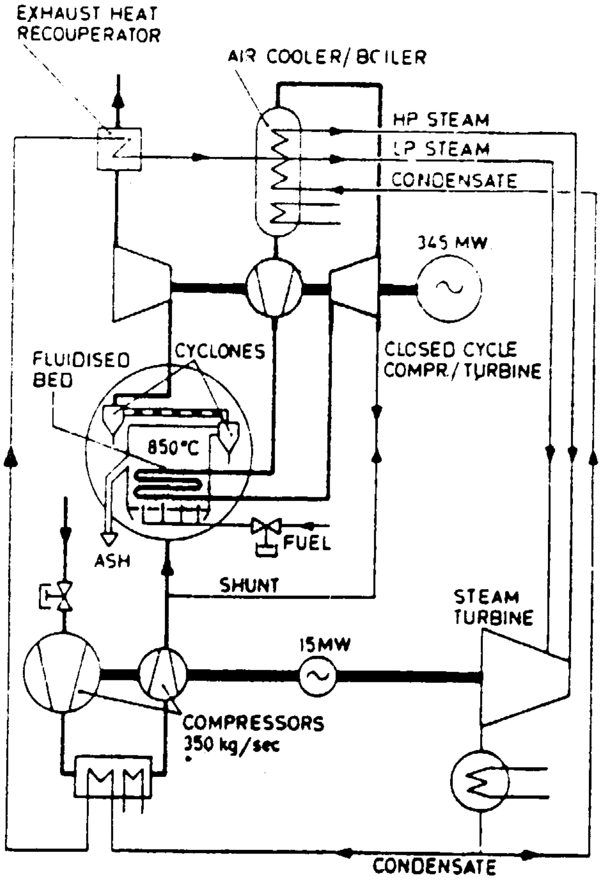

Figure 3. Flow diagram for a 360-MW composite plant with a pressurized fluid-bed air heater. The same general arrangement can be used for a district heating plant capable of 260 MW electrical output and 500 MW heat output.

can operate at atmospheric pressure.

Complex systems can be envisaged using closed cycle gas turbines. The example shown in Figure 3 gives an electrical output of 360-MW at an efficiency of 40 to 42%. Waste heat is recovered from the closed cycle loop in a pressurized cooler boiler with the steam turbine being used to drive the compressors for the open cycle gas turbine.

Apart from the closed cycle compressor and turbine, all the rotating components are commercially available.

The high heat transfer coefficient to tubes immersed in fluid beds together with uniform heating, open up a wide range of possibilities. Other fluids such as potassium vapor or helium could be used in the closed cycle loop but we are limiting the scope of this article to cycles that may be exploited with a minimum of development effort. Fraas has discussed the potential use of potassium vapor cycles elsewhere. (6)

In an authoritative study carried out in mid-1973, the capital costs of a 660-MW pressurized fluidized combustion combined cycle system were shown to be 17% less than those for a conventional P.F. system. Similar reductions (i.e. 17½%) in the comparative cost of electric power were also indicated.

This type of cycle is illustrated in Figure 4. Although there is some scope for variation in the way in which the steam generating surface is disposed, there are fewer possible cycle variations than with the air cycles. This scheme is basically a high efficiency cycle and the main scope for ingenuity lies in transferring as much heat as possible from the gas turbine exhaust back into the steam cycle.

There is some freedom of choice in the operating parameters, however. Although it is appreciated that an increase in gas turbine inlet temperature will increase the cycle efficiency, it is perhaps less immediately obvious that operating with higher excess air (up to, say, 100%) also increases the efficiency as well as altering the ratio of power output from the steam turbine and gas turbine. This is illustrated in Table 1.

Of course, operating the fluidized bed at high excess air increases the size of the bed per kw of total output and thus increases the containment and gas cleaning costs. On the other hand, the relative amount of steam generating surface in the combustor is considerable reduced, the combustion situation is more flexible, corrosion of tubes in the bed or of turbine blading is even less likely, and certain control problems which are discussed in greater detail below are made easier.

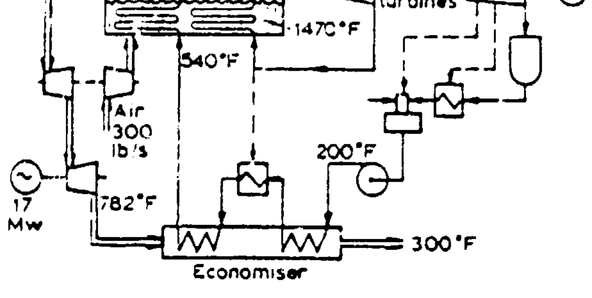

Figure 4. Combined steam-gas turbine cycle of 140 MWe output.

Table 1. Typical cycle efficiencies at different operating conditions

Gas turbine inlet temperature	1600°F		1700°F	
Excess air %	20	100	20	100
Cycle efficiency, %	40	41½	40½	42½
Gas turbine output Total output, %	20	30	22	32

Effect on fluidized bed dimensions by cycle type

At this point, it is useful to discuss the effect of the type of cycle on the physical dimensions of the fluidized bed. In pressurized fluidized combustion the very high volumetric heat release rates that are feasible mean that the depth of bed is controlled by the need to accommodate the heat transfer surface rather than by combustion considerations—even though heat transfer coefficients to the surface are in the range 60 to 70 Btu./sq.ft.hr.°F. The heat release rate—and consequently the depth of bed—is directly proportional to the operating pressure and almost proportional to the fluidizing velocity.

For a given heat transfer duty, much more heat transfer surface is required for the airheater than for the steam generator. This arises partly because the air has to be heated to a much higher temperature than the steam—hence the driving force (7) is lower—and partly because the internal heat transfer coefficient between the tube and the heated fluid is lower with air than it is with steam. (This is necessary if high pressure drops in the airheater tubes are to be avoided.)

Consequently the volume of bed required in an airheater is substantially higher than that in a steam generator. If it is accepted that bed depths should not exceed 8 to 10 ft. (this is at the limit of present experience) then airheaters are limited to maximum fluidizing velocities of, say, 2 to 4 ft./sec., whereas steam generators can employ velocities up to, say, 10 ft./sec.

Changing load in a supercharged combined cycle power plant is more complex than in an air cycle plant. The load range which can be attained is primarily dependent on the temperature range over which a fluidized combustion bed can be operated. On present indications this is likely to be from 1,600°F down to 1,350 or 1,400°F. At higher temperatures the volatilization of deleterious alkali metals is likely to be excessive, and at lower temperatures (the exact lower limit will depend partly on the choice of excess air) the combustion efficiency will be too low.

The variation in heat input to tubing immersed in the bed (and hence steam turbine output) which may be achieved by varying bed temperature is somewhat greater than the variation in temperature difference between bed and tube, because a significant part of the heat transfer is due to radiation; so that in effect the heat transfer coefficient falls with bed temperature. Even so, the turn-down which can be achieved in a single bed is likely to be limited to 60 to 65% of full load. Over this range there would seem to be no major technical problems. Changing the fuel input will alter the bed temperature and therefore: 1) the steam turbine output; 2) the gas turbine entry temperature and its power output; and 3) in two shaft machines, the compressor airflow and pressure.

These gas turbine characteristics mean that the fluidizing velocity and excess air remain substantially constant over the load range (hence the part-load efficiency of the cycle remains high).

However, to reduce the load below about 60 to 65% will present problems. One solution is to compartment the bed so that successive areas of the bed can be slumped. Detailed analysis of this approach reveals a number of facets: firstly, the compartmented bed sections have to be of differing sizes; secondly, the system is limited to one in which the excess air is greater than about 60% at full load.

The following analysis will illustrate this latter point. Consider a system where the bed is in four compartmented sections, A,B,C, and D. Further consider the situation when only bed section A is activated and B,C, and D are slumped. As the load is increased and B,C, and D are successively reactivated, the mass flow through bed section A, and hence the heat input, increases (as the compressor output and pressure increase). However, the heat that can be extracted from the compartment A into the steam cycle remains more or less constant. Thus the only way in which the bed temperature can be maintained below 1,600°F is by increasing the excess air. At full plant load, bed section A will be operating at an excess air well above 200% and only bed section D can operate at low excess air.

Thirdly, every time a bed section is slumped (this will normally happen when all the bed temperatures will be at their minimum), the temperature in the remaining beds must be increased rapidly from minimum to maximum to accomplish a smooth load change. CSL has given considerable thought to the ways in which this step change in individual beds can be accomplished smoothly.

A potential alternative to compartmented beds is to alter the height of the fluidized bed and thus submerge more or less of the tube surface in the bed. This has the effect of altering the average heat transfer coefficient to the tube surface because the heat transfer coefficient to tubes submerged in the bed is many times greater than to tubes exposed in the freeboard. Model investigations have shown that material can be transferred to and from the bed at a sufficiently rapid and controlled rate to meet rates of load change of 5% per minute.

A radically different approach to the problem of load changing is to carry out the functions of combustion and heat transfer in two separate beds with means for pumping hot ash from one to the other. A number of variations o this basic idea are possible, one of which is shown in Figur 5.

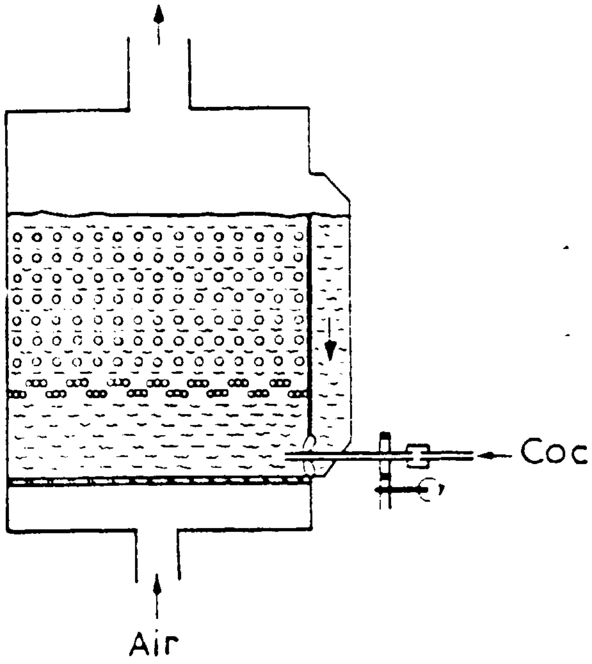

Figure 5. Diagrammatic arrangement of circulating bed Fluidized bed is separated into a lower combustion bed an an upper heat-transfer bed. Output is controlled by circu lating the bed using mechanical (as shown) or pneumati means.

Fuel is introduced into the first bed which operat throughout at temperatures within the limits for good com bustion. The combustion gases pass through the upper be which contains steam generating tubes; and the beds a separated by a horizontal 'one way' screen which allows as to move upwards only. Ash overflowing from the top bed pumped at a controlled rate back to the lower bed. Th temperature of the gases leaving the top bed and passing the gas turbine depend on the circulation rate, as does th amount of heat transferred to the steam cycle. Theoretica studies have been made of this type of combustor, but n experimental work as yet.

As mentioned previously, load changing is accomplishe more simply in airheater cycles. Because the temperatur differences between bed and heated fluid are much les than in steam generating units, the effect of changing be temperature is much greater. This is helped by the fact th as the load is reduced, the pressure of the heated fluid reduced, which still further reduces the heat transfer to th fluid. It is anticipated that none of the measures describe earlier in this section would be necessary in airheater cycle

State of the development today

In all cycles involving pressurized combustion, the ga turbine power is obtained by expanding a gas consisting entirely, or in part, of the products of combustion of coal. The outstanding question is whether or not the turbine materials will suffer from corrosion, erosion or deposition.

It was largely to obtain some initial experience in this field that a pilot plant was built at BCURA Ltd. in the UK. It operates at about 6 atm., and the fluidized bed can be either 4 x 2 ft. or 3 x 2 ft. in cross section. The exhaust gases, after cleaning in a two-stage cyclone arrangement, pass through a segment of the nozzle guide vane from a marine gas turbine. Downstream from this "cascade," at a point where the Mach No. is about 0.6, are two target rods or pins.

Over a period of about 1,000 hr. operation at varying operating conditions with a number of coals, but with bed temperatures averaging 1,470°F and cascade temperatures of 1,450°F, an X40 cascade and Nimonic 75 pins remained substantially free from corrosion. Sodium and potassium present as aerosol in the gases was about 1 ppm (i.e. about 1% of the alkali input).

Subsequent operation for 200 hr. at higher temperatures (1,750°F bed temperature, up to 1,700°F cascade temperature) produced some sulfidation attack in an X40 cascade.

No sulfidation occurred in an EPK55 pin, but longer operating periods are necessary before any firm conclusions are drawn. The alkali aerosol content of the exhaust gases was about 7 ppm. at these higher temperatures.

At no stage have there been any signs of erosion. This was to be expected in view of the soft nature of the exhaust dust (due to the relatively low combustion temperatures) and its small size (typically 95% less than 10 micron and 55% less than 5 micron with a dust concentration of about 200 ppm.).

Corrosion of tube materials in the bed has been examined both in the pressurized combustor and in a special combustor operating at atmospheric pressure and designed for continuous operation. Data accumulated over many thousands of hours have confirmed that corrosion of typical boiler tube materials is less than in conventional coal-fired boilers. Data for metals operating at the high temperatures corresponding to airheater cycles are less extensive, but suggest that corrosion rates would be acceptable.

Although with present materials we would hesitate to suggest operating a gas turbine at a temperature higher than, say, 1,560°F from a conventional fluidized bed, arrangements might be developed to give higher temperatures.

For example, part of the bed can be arranged as a fluidized devolatilizer into which the coal is fed. Gases from this unit would be cooled and cleaned thoroughly before being burned with additional combustion air. The char from the devolatilizer would be fed into a conventional fluidized bed combustor which operated at a maximum of, say, 1,560°F. The two hot streams would then be mixed to give a turbine inlet temperature of 1,800 to 1,900°F.

Other experience with the pressurized combustor is summarized below. Further details have been reported by Hoy and Roberts. (7)

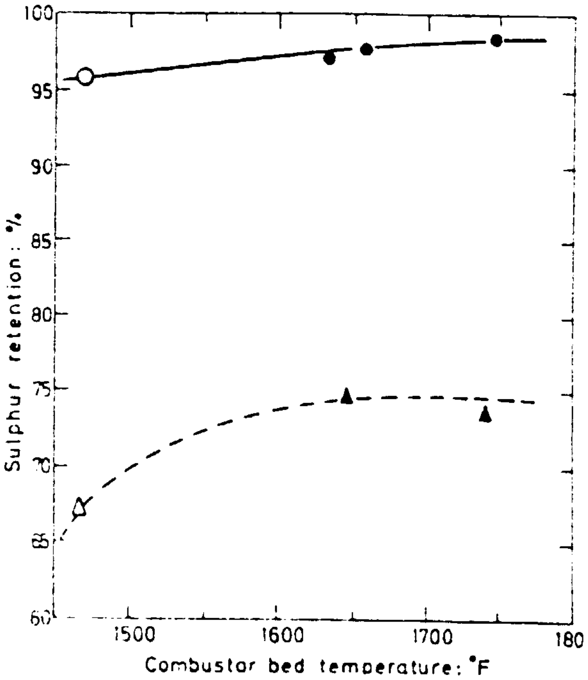

Figure 6. Effect of combustor temperature on sulfur retention in pressurized combustor. Upper curve: dolomite addition; Ca/S mol ratio is 2. Lower curve: limestone addition; Ca/S mol ratio is 2.

Combustion efficiency: Because of the deep beds and dense tube packing inherent in pressurized applications, gas bubbles are broken up before they grow large: and there is therefore better gas-solids contacting than at atmospheric pressure. Depending on the choice of operating parameters it is possible to achieve high combustion efficiencies (greater than 99%) without the need for recycling of elutriated material.

Sulfur retention. Results from the combustor confirm that high sulfur retention can be obtained when dolomite is added to the bed. Results summarized in Figure 6 show that over any likely working range, contrary to atmospheric pressure experience, there is no optimum temperature for sulfur retention. Although on a mole ratio basis, limestone is less effective than dolomite, on a weight basis there is little to choose between the two.

NO_x emission: Figure 7 shows data for NO_x emission as a function of excess air. It will be seen that even at high excess air values the NO_x emission is only about 200 ppm.

Plans for the future

The present state of the art relating to pressurized fluidized combustion and its application to power generation have been outlined in general terms. The logical next step is to build and operate a realistically-sized demonstration plant. Combustion Systems Ltd. is working towards this end, to a great extent in co-operation with various sponsoring agencies, contractors and manufacturers.

CSL will continue to develop the technology of fluidized

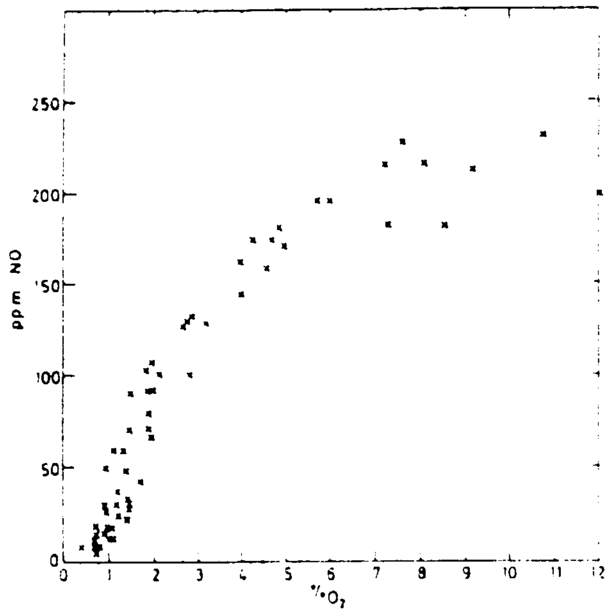

Figure 7. NO_x emissions from pressurized combustor.

combustion using all the resources available to it, and in the near future a new program of work will be started on the BCURA Pressure Rig. So far this plant has operated at low velocities (2-3 ft./sec.) with relatively shallow beds (4-5 ft.) and it is envisaged that work over the next two years wi extend this operating range to cover fluidizing velocities o up to 10 ft./sec. and bed depths of up to 8 ft. In paralle with this work, a number of control problems, along wit their possible solutions, will be investigated.

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COMBUSTION OF TOMORROWS FUELS - II

Status of Pressurised Fluidised Bed Technology Today - A Review

by

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ABSTRACT

The first ever pressurised fluidised bed combustor (PFBC) came into operation at the laboratories of the British Coal Utilisation Research Association at Leatherhead in 1969. This operated at pressures up to 6 bar and ultimately had a thermal input of 4.5 MW. In the ensuing years other laboratory scale combustors came into operation with thermal inputs ranging from 0.4 MW to about 2.4 MW and with working pressures up to 20 bar. Scale-up of the technology began in 1980 when the 60 MW(t) PFBC test facility sponsored equally by the Governments of the United States of America, The United Kingdom and the Federal Republic of Germany came into operation at Grimethorpe in the U.K. The scaling-up process was augmented in 1982 when the 15 MW(t) Component Test Facility (CTF) built in collaboration between ASEA-ATOM, Stal-Laval/Deutsche Babcock, and the Sydkraft power utility came into operation.

Comprehensive data on combustion, hot gas clean-up and environmental aspects of the technology have been accumulated and design studies have shown that combined cycle power plant based on PFBC have the potential to provide electric power at lower busbar costs than competing advanced power generating technologies.

The paper provides an overview on the development of the technology, the technical advances made, and those in the pipeline aimed at making the system even more attractive.

Status of Pressurised Fluidised Bed Technology Today - A Review

Introduction

The motivation to develop fluidised-bed combustion systems has been its potential for:

(i) burning a wide range of fuels with minimal sensitivity to changes in fuel characteristics

(ii) reducing the emission of sulphur and nitrogen oxides at low capital cost and with low thermal penalties

(iii) reducing the size and capital cost of steam-generating or airheating equipment by virtue of (a) the higher rates of heat transfer achieved with tubes immersed in fluidised beds and (b) elimination of the need for large furnaces in which to cool molten ash particles by radiation before they enter the convection passes.

The motivation to develop pressurised fluidised-bed combustion (PFBC) was to add to the foregoing the potential advantages of (i) more compact plant capable of a greater degree of shop fabrication (thus reducing capital cost) and (ii) higher power generating efficiency by burning coal directly in combined gas and steam turbine driven power generating plant.

This paper outlines the work that has been done so far in developing PFBC and mentions the work that is in progress or that is needed to sustain or enhance its preeminence over rival technologies.

Combined Cycle Power Plant

Gas turbines convert heat into power at higher temperatures than are practicable in conventional power plant using steam turbines. Steam turbines, however, reject heat at a lower temperature so that combining gas turbines with steam turbines widens the temperature range over which heat is converted to power, thereby increasing power generating efficiency. Large combined cycle power plant are already in operation burning clean fuels such as natural gas and LPG. For the system to find wide application however the source of heat in the combustion gases expanded through the gas turbine needs to be coal.

Fluidised-bed combustion under pressure offers the possibility of burning coal directly without incurring the losses inherent in first gasifying or liquifying the coal to remove the ash. The relatively low temperature (850-900°C) in the fluidised bed combustion system is considered to be the key factor. It minimises the extent to which constituents likely to cause corrosion and fouling of gas turbine blades are released from the ash. The temperature is also below that at which ash particles vitrify and become more erosive.

Many types of combined cycle based on PFBC have been proposed. The design and costing studies carried out to date (e.g. Refs. (1) and (2)) have shown the Steam Cycle (Fig. 1) to be the most likely to be both technically and commercially successful.

It is envisaged that operating pressures will be in the range 12-17 bar, bed temperatures around 850°C-900°C and that up to about 30 percent of the power sent out will be generated by the gas turbine.

The potential benefits from adopting the PFBC steam cycle as compared with conventional pulverised coal-fired power plant having wet scrubbers to reduce SO_2 emission have been estimated (Ref. 2) for USA conditions to be,

(i) 3-4 percentage points increase in power generating efficiency

(ii) about 13 percent lower capital costs, and

(iii) a reduction of around 10 percent in the busbar cost of electric power.

The relative importance of the savings in operating and capital costs will of course vary from country to country. Where coal is expensive (e.g. in the UK) high efficiency will be given priority over reduction in capital costs.

A number of issues have yet to be resolved however before installation of large (e.g. 200 MW(e) upwards) units can be envisaged with confidence. As might be expected the main uncertainty lies in the long-term durability and aerodynamic performance of the gas turbine blades, i.e. can PFBC combustion gases, after treatment in practicable designs of hot gas clean-up equipment, be expanded through gas turbines without causing unacceptable rates of fouling, erosion, or corrosion?

The answer to this key question can only be obtained by operating industrial designs of gas turbine on PFBC combustion gases. The large expenditure involved has so far precluded this. Tests have been carried out in which static cascades of blades or small rotating elements have been exposed to combustion gases (Refs. 3, 4, 5). The results have generally been satisfactory but the underlying doubt has led to proposals (Ref. 6) being made to operate the first PFBC systems as pressurised boilers i.e. the Turbocharged boiler (Fig. 2). The gas turbine would operate at relatively low inlet temperatures (e.g. 400°C-500°C) and only sufficient power would be recovered from the combustion gases to drive the air compressor. At these lower temperatures it would be simple to achieve efficient operation of the gas clean-up system and the possibility of corrosive attack on turbine components would be minimised. There would however be a loss of about two percentage points in power generating efficiency.

The main market for the PFBC combined-cycle is likely to be for plant in the size range 500 MWe to 660 MWe - although turbo-charged boilers may be used in smaller sized units, e.g. c. 100 MW(e). When connected to a major distribution network combined cycle plant may not be required to operate much outside the range 80% mcr to 100% mcr with the ability to accommodate load changes of 3-5% per minute. In other applications the ability to operate over the range 30% mcr to 100% mcr and to accommodate load changes of 3%-5% mcr per minute over that range could be important and two shift operation may also be required.

Origins of the Technology

The potential benefits of the PFBC were first appreciated by the late Douglas Elliott whilst at the UK Central Electricity Generating Board (CEGB) and staff at the British Coal Utilisation Research Association* in late 1967, and the first-ever pressurised fluidised-bed combustor came into operation at Leatherhead in 1969. Work on PFBC at Leatherhead has, since 1972, mainly been funded by the United States Department of Energy, by its predecessors, and by organisations either directly or indirectly associated with it.

Other organisations that have had a significant involvement in the development of the technology at Leatherhead or for which test work has been carried out include Stal-Laval (now ASEA-PFBC), General Electric (USA), American Electric Power, The Electric Power Research Institute (USA), Fluidised Combustion Contractors Ltd. (FCCL), British Columbia Hydro, Swedish NE Board, Curtiss-Wright and IEA Grimethorpe.

Major contributors to the development of the technology elsewhere include Exxon who began work in 1973, Curtiss-Wright (1977), IEA Grimethorpe (1980), General Electric-Malta Rig (1981), New York University (1983) and ASEA-Stal-Laval Malmo Rig (1981).

Several design studies for PFBC power plant in the size range 200-660 MW(e) have been carried out or are in progress, e.g.

(i) Westinghouse (Ref. 7), for the USA Environmental Protection Agency in 1971-73:

(ii) GE (Ref. 8) for the US DOE in 1981

(iii) Burns and Roe (Ref. 1) for EPRI in 1980

(iv) Gilbert Associates (Ref. 2) for US DOE (1982)

(v) Coal Processing Consultants (Ref. 9) for BC Hydro (1982)

(vi) Brown Boveri in association with Foster Wheeler and Lurgi/ Combustion Engineering (Ref. 10) for EPRI (1983).

Operating Conditions Explored

The test rigs that have produced or are producing significant data for PFBC systems are listed in Table 1 together with the upper limits of some of the operating conditions explored in them.

Coals from the USA, UK, West Germany and Canada have been burned. These have mainly been high volatile bituminous or semi-bituminous coals with ash contents ranging from 7% to 44%.

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* Since 1977 the NCB Coal Utilisation Research Laboratory

The coals have mainly been fed dry (near zero surface moisture) with a top size in the range 3-6 mm depending upon the fluidising velocity. Pressurisation has usually been effected using a lock hopper system from which in most recent installations the outflow of coal has been controlled by rotary valves. The coal is then transported in air to the combustors.

Tests have also been carried out in the 20 atm rig at Leatherhead (Fig. 3) in which coal-water mixtures containing 70% coal with a top size of 3 mm have been burned at pressures in the range 10-16 bar. This method of feeding eliminates coal drying, simplifies pressurisation, and eliminates the need for inert gas for pressurising lock hoppers.

There is however a reduction in power generating efficiency (Fig. 4). The reduction (Ref. 11) depends, as might be expected, on the water content of the slurry, and also upon the contribution the gas turbine makes to the total power output (at constant turbine inlet temperature this mainly depends upon the excess air value). Where the gas turbine power is about 30% of the total the estimated loss is about 0.8 percentage point increasing to about 1½ percentage points for the pressurised boiler cycle where the gas turbine generates only sufficient power to drive the air compressor.

An alternative approach to reducing the cost of feeding coal - namely eliminating the drying stage and increasing the top size to 25 mm - has also been tried (Ref. 12).

The SO_2 sorbents used have been limestones (raw, and precalcined) and dolomites. These have normally been fed dry crushed to have a top size appropriate to the fluidising velocity but in some tests in the Leatherhead 20 atm rig the sorbent has been fed as coal-water mixture.

Although fluidising velocities as high as 3.7 m/s have been used the majority of the testwork has been at fluidising velocities of 1.5 m/s or lower and bed depths for the majority of tests have been sufficient to provide gas residence times of two seconds or longer.

Excess air values of about 30% or above have predominated since from the point of view of cycle efficiency high excess air is not a disadvantage in combined cycle plant and it simplifies attaining the required rates of load change. The greater interest in reducing capital cost now applying in some countries has resulted in more effort being directed to operation at lower excess air values.

Bed temperatures have been in the range 720-950°C but because of the need to minimise alkali release the upper limit has usually been in the range 850-900°C.

Both tapered and parallel bed combustor geometries have been tested. So far as the steam cycle is concerned, most of the experience has been with horizontal or near horizontal tubes with packing densities in the range from about 7 m^2/m^3 bed to about 19 m^2/m^3.

The fouling, erosion and corrosion propensities of the combustion gases after cleaning in cyclone type dust collectors have been assessed by passing them through cascades of turbine blades/airfoils, or small rotating blade assemblies.

The Performance Obtained

Important issues include: combustion and sulphur retention efficiencies, NO_x emission, SO_3 formation and its effect in limiting stack temperatures, alkali concentrations in the combustion gases entering the turbine, the ability to load-follow, and factors that affect availability such as tube bank metal wastage.

The main findings are summarised below:

Combustion Efficiency: Departure from complete burn-out of the combustion material fed to the combustor is almost entirely due to incompletely-burned material being elutriated from the bed. With but few exceptions this is captured by the first-stage cyclones. Combustion efficiencies close to, or above 99% can be expected where the bed temperature is above about 850°C, excess air is above about 30%, bed residence time is 3 seconds or longer, and where the coal being burned is high-volatile bituminous or semi-bituminous. Combustion efficiencies have been somewhat higher at 16 bar, than at 10 bar and below (Ref. 13).

As might be expected lower combustion efficiencies apply when low-volatile coals are burned and also at low loads. Reduction of load to 30%-40% mcr, for example, would involve operation (a) with reduced bed height (e.g. giving a bed residence time around 1 second); (b) with reduced bed temperature (e.g. c. 750°C) and (c) with reduced pressure. Unless corrective measures are taken, e.g. by recycling primary cyclones fines to the combustor, combustion efficiency is likely to decline from 99% to around 95% (Ref. 4).

High combustion efficiencies (99%+) have been obtained when burning coal-water slurries and efficiency appears to decline less rapidly with drop in bed temperature than when the coal is fed dry - see Fig. 5 (Ref. 6). The slurry tends to form globules of carbonaceous material as it leaves the nozzle. These probably incorporate the fines in such a way as to increase their residence time in the bed.

Some exploratory tests in which the fluidising air was admitted in two stages with the objective of reducing NO_x emission showed significantly lower combustion efficiencies when operating at 10 bar pressure, particularly at excess air values below about 25%, see Fig. 6, (Ref. 16).

Scaling-up from the small rigs (e.g. 300 mm x 300 mm) to the largest pressurised fluidised bed combustor (e.g 2 m x 2 m) has not so far resulted in significant deviations from the combustion efficiency values that would have been anticipated from a correlation developed at an earlier stage in the investigations. There is, however, some uncertainty as to the effect on performance of coal feed nozzle spacing and this is unlikely to be resolved until much larger-sized plant comes into operation. (More details of the larger scale operation will be presented in a later paper).

Sulphur Capture: The ability to meet emission regulations at an acceptable cost for SO_2 sorbent is of paramount importance to the success of PFBC. Reliable predictions of sulphur capture by limestone or dolomite added to the bed still relies heavily upon carrying out test work under operating conditons closely approximating to those that will apply in commercial plant.

Performance is influenced by many factors and their interactions: notably, residence time in the bed and freeboard, bed temperature, reactivity of the sorbent (particularly accessibility of the pore-structure), particle size, and oxygen concentration at the bottom of the bed. Correlations have been evolved that take these into account reasonably well and from which the sulphur capture performance in a large combustor can be predicted with a reasonably high degree of confidence using the results of tests carried out in a small combustor.

Some guidance as to the performance of a sorbent in a combustor can also be obtained by carrying out tests on a small sample of sorbent in pressurised thermogravimetric analysers and using the Westinghouse (Ref. 7) model (with suitable adjustments) to interpolate the data.

The choice of sorbent will, in most instances, be based on local availability because the cost of transport is normally the major factor in its cost at a plant. High calcium utilisation is important; the degree of utilisation affects the quantity of material required, the quantity of material to be disposed of, and the cost of treating effluent waters from the disposal site in order to avoid contamination of ground water and streams. There is consequently a need to review the sulphur capture performance periodically to see to what extent this might be improved.

On the basis of the calcium/sulphur ratio, dolomite is about twice as effective as the best limestones and 90% sulphur capture can be expected with a Ca/S mol ratio of 1.5 under typical PFBC operating conditions (but on a mass ratio basis there is little difference between the performances attainable with good limestones and dolomites).

Until comparatively recently sulphur capture performance was superior as regards both the Ca/S ratio and mass ratio to that achieved by AFBCs using limestones. Aided by tall freeboards and fines recycle, 90% sulphur capture has recently been achieved in AFBCs at Ca/S molar ratios of about 2.0 using limestone, and sulphur captures of 90% are claimed for circulating beds operating at Ca/S molar ratios of as low as 1.2. There is scope however for improving PFBC sulphur capture performance, e.g. by (a) using low (app. 0.9 m/s) fluidising velocity.; this results in a smaller bed particle size, (b) using deep beds and hence long residence times, and (c) by recycling fines captured by the primary gas clean-up stage.

Sulphur capture may also be improved by as much as 30% by adopting two-stage combustion, Fig. 7 (Ref. 18). The engineering problems involved in supplying air at two levels in a large bed are, however, formidable and from this point of view it is probable that alternative solutions will be explored first.

Sulphur Trioxide: Very little SO_3 is likely to be present in the combustion gases leaving a fluidised bed to which sufficient sorbent has been added to capture 90% of the sulphur. Oxidation of SO_2 to SO_3 in the gases downstream of the bed will however proceed (Ref. 19) to an extent which depends upon factors such as, (a) operating pressure, (b) excess air, (c) residence time, (d) presence of catalytic materials, and (e) the presence or otherwise of the unreacted sorbent. Usually on the small test rigs, less than 5% of the total sulphur oxides emitted has been present as SO_3.

The acid dewpoint temperature increases with increase in SO_3 and water vapour concentratiuons, being higher for example for CWM than for dry coal, see Fig. 8. With operation at 16 bar burning dry coal with sufficient sorbent to ensure 90% sulphur retention, SO_3 concentrations of 3-8 ppm can be expected. Unless corrective measures are taken, e.g. by ensuring that finely-sized sorbent is present in the gas path, the exhaust gas temperature from PFBC systems would need to be about 5°C to 20°C higher than from non-pressurised systems in order to avoid acid condensation.

NO_x emission: NO_x emissions from PFBC's operated under conditions of commercial interest are generally well within current regulatory requirements, see Fig. 9. Since there are already more stringent requirements in some locations, and there is a possibility of these becoming more widespread, tests have been carried out to see whether a worthwhile reduction might be made by two-stage air admission.

The reduction obtained in these tests (Ref. 18) was marginal and would not justify the cost and complication involved in overcoming the considerable engineering problems involved in providing an additional air distributor.

Heat Transfer: Heat transfer rates have been shown to increase with pressure. It will be seen from Fig. 10 (Ref. 13), that by increasing the pressure from 10 bar to 20 bar, the average convective heat transfer coefficient (to water cooled tubes) was increased from about 280 to about 340 W/m^2K. The heat transfer data obtained from the Grimethorpe combustor are reported in another paper to this session. The data appear generally to have been in line with predictions from data from the small rigs.

Some indication has also been obtained (Ref. 20) of the rate of heat transfer to surfaces in the splash zone when a bed level is below the top of the tube bank - as would apply in operation at part load. It will be seen from Fig. 11 that in the tests the splash zone extended for about 600 mm above the nominal top of the bed.

Elutriation: Gas-borne material leaving the bed includes particulates from the coal and sorbent feeds of elutriable size; breakdown products from both; incompletely reacted coal particles; and material 'splashed' out of the bed due to eruption of bubbles. The extent to which the latter enter the hot gas-cleaning system depends upon factors such as freeboard height, geometry and gas velocity.

For a USA coal widely used in the test work in small rigs, typically about 20 percent of the material withdrawn from the bed would be coal ash and the remainder 'dolomite'. The UK coals however have been associated with a larger proportion of minestone and this has been the predominent constituent of the material withdrawn from the bed.

Significantly more material has been elutriated from large (e.g. 2 m x 2m) than from small combustors (e.g. 0.9 m x 0.6 m) operated under otherwise similar conditions. The larger cross-section more probably allows full growth of bubble size and energy and hence 'splash' effects are increased. There also appears to be greater breakdown of sorbent particles in the larger beds. The nett effect appears to be to about double the amount of material elutriated compared with 'small' bed operation. It is thought that loss of material by splashing could be significantly reduced by having much more closely spaced tubing at the top of the tube bank than in the remainder.

The relative amounts of sorbent and coal ash elutriated and the amounts retained in the bed can be calculated from analytical data with a reasonable degree of confidence, but it is less certain that elutriation can be correlated with operating conditions with an acceptable degree of accuracy.

Alkali Emission: Alkalis, particularly in association with sulphur compounds, are considered to be the principal sources of corrodents at high temperatures. Alkalis enter the PFBC system in the coal ash and in the sorbent. Mainly they leave the system via the excess material discharged from the bed and in the dust discharged from the cyclones. About 5% will typically be present in the combustion gases and gas-borne particles of dust after the cyclones. The potentially condensible concentration of sodium and potassium in the gases passing through the turbine amounts to less than 1 ppm of each, but these concentrations are much in excess of what is deemed to be acceptable in oil-fired gas turbines. The presence of fine particulate matter together with the lower temperatures and the corrosion resistant materials envisaged for the coal-fired plant can be expected to provide some compensation.

The indications from the cascade tests mentioned previously (4, 5) were that provided bed temperatures are not higher than about 850-880°C materials/coatings may be available to provide satisfactorily long blade life. Some of the issues involved are discussed in reference 21. By restricting the operating temperature to the above levels it has also been shown that dust deposited on blades can be readily removable using conventional on-line cleaning methods.

The over-riding need is however to expand PFBC gases through a turbine with industrial style blading in order to have a true measure of the problems that might need to be faced.

Engineering Development

High availability will be a major factor in the adoption of PFBC systems. The beneficial effects of higher power generating efficiency and lower capital cost are rapidly lost if the availability of the plant does not match that of current systems. The status of development of the various components of a PFBC system can be summarised, in the authors' opinion, as follows:

Solids Preparation, Pressurisation and Feeding: Experience has confirmed the need to remove surface moisture to ensure reliable feeding of solids in pneumatic systems. The lock hopper, despite the high energy loss incurred and the present requirement of inert gas for pressurising, is currently the only proven means for introducing dry coal into a pressurised system. Rotary valves are now available for controlling the rate of feed from pressurised hoppers into pneumatic transport systems. These provide much more satisfactory control of flow than any solely pneumatic system and also involve lower feeder hopper pressures.

Stable coal-water mixtures containing 70% of coal with a top size of 3 mm can be produced without the aid of additives, and these can be reliably pumped into combustors operating at high pressures (e.g. 16 bar). The sorbent can also be incorporated into a coal-water mixture, albeit with an increase in the water input to the combustor and correspondingly higher loss of efficiency. The indications are that commercially-available preparation and pumping equipment should be adequate to reliably and economically meet the requirements of large PFBC plant.

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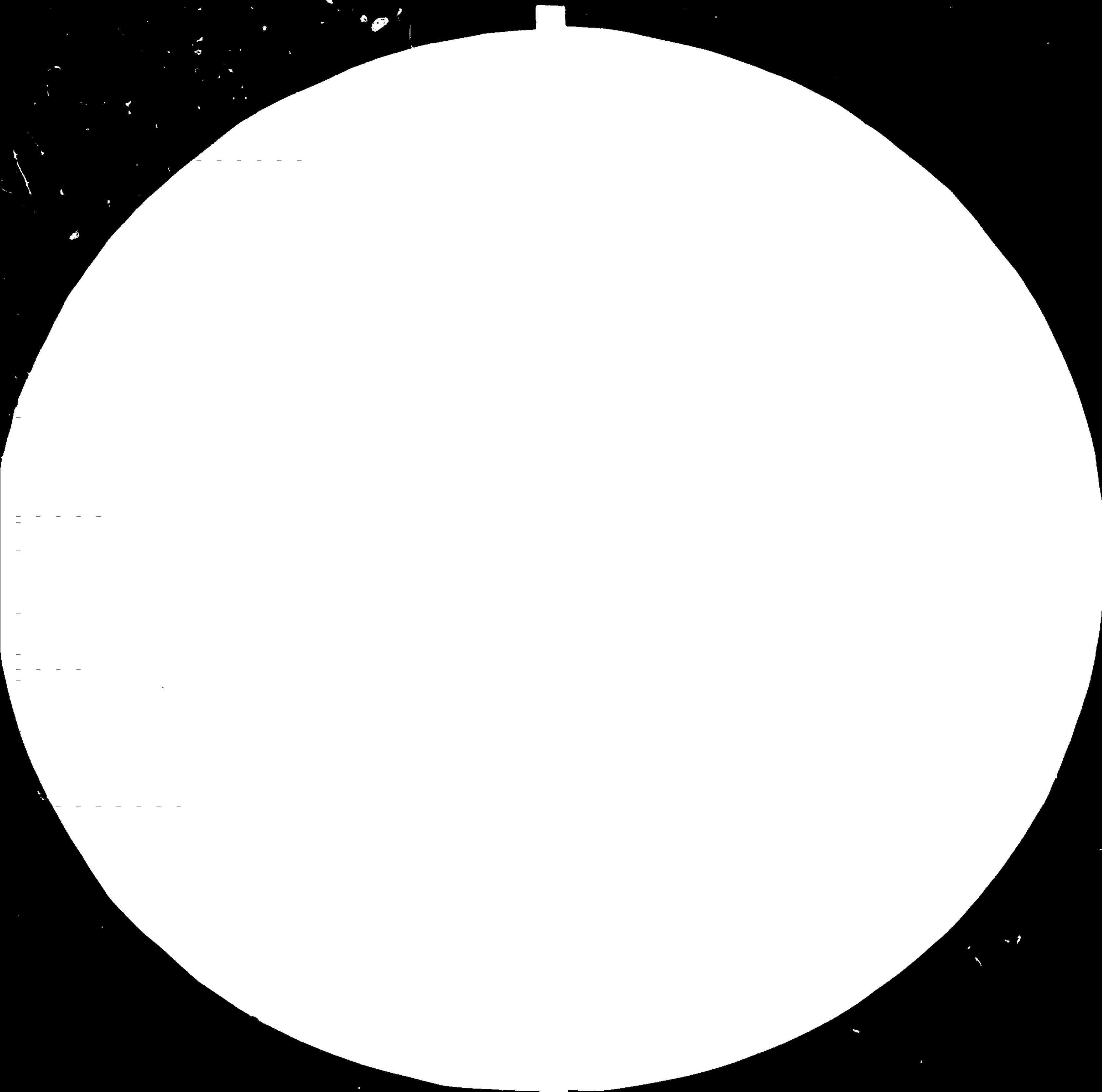

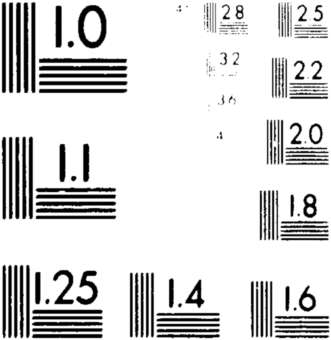

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There is however scope for developments to (i) enable run-of-mine coal to be used without the need for preparation other than to reduce the top size to about 25 mm and (ii) to evolve an alternative to the lock hopper for pressurising that will have lower energy losses and obviate the need for using inert gas for pressurising the contents.

Combustor Design

In the authors' view the trends will be:

(i) to operate at relatively low fluidising velocity (e.g. 1 m/s) and with deep beds (e.g. 3 m upwards) to enhance combustion and sulphur capture performance. Low velocity is also desirable to minimise tube bank metal wastage.

(ii) to use combustors with tapered beds in order to simplify air distributor design, and to simplify distribution of coal and the removal of excess material from the bed, and

(iii) to use change of bed level as well as change of bed temperatures as the means for changing load.

Differences of approach are likely to be in the distribution of heat transfer surface and in the design of the combustor containment. The main alternative as regards heat transfer surface is whether the functions of evaporation, super-heating and reheating are carried out in separate beds, each with separate control on firing rate, or whether they are confined into single beds.

The alternatives as regards combustor containment are (i) the large sphere designed to contain both the combustor and the bulk of the hot gas cleaning equipment and (ii) separate cylindrical vessels for the combustor and the gas cleaning system. The latter is suitable for prefabrication off site. The choice is likely to be site-dependent.

Further developments are considered to be needed, (i) to evolve tube bank geometries and means for protecting tubes that will result in lower rates of metal wastage, and (ii) to enable run-of-mine coal to be burned without accumulation of shale particles in the bed that can lead to defluidisation.

Control

Control of load in a system involving steam and gas turbine plant together with a combustion system having significant inertia is inevitably complicated. This has received attention in the test work particularly at Leatherhead and Grimethorpe. The data obtained on system response to changes in firing rate have contributed to the formulation of designs for control systems over which there is now considerable confidence.

Hot Gas Clean-Up

Cyclones appear to be capable, when operating as designed, of removing particles from the gas stream that are larger than 10 microns and the great majority of particles larger than 5 microns. It is conceivable that three stages of cyclones might be sufficient to reduce the dust loading acceptable to a large industrial gas turbine where relative velocities are low. It is most unlikely, however, that the dust loading will approach that needed to meet the USA EPA emission limitation of c. 30 ppm and hence the turbine would need to be followed by an electrostatic precipitator or a bag-house filter.

The elimination of the need for this large and expensive equipment by the development of particulate removal equipment suitable for installation ahead of the turbine is an important objective. Several systems are under development. Reliable equipment for on-line measurement of dust concentration to give warning of mal-function of the clean-up systems is urgently required.

Solids Removal and Depressurising

The principal means for depressurising dust/ash is the lock hopper. Developments in recent years have led to valves that will seal in hot and dusty environments. The valves are expensive and maintenance requirements can be severe, consequently an approach now being adopted is to cool the solids before they enter the lock system so as to reduce the duty and hence capital cost of the valves. Some skill will be required, however, in the designing of means for cooling the fine dust captured downsteam of the primary collectors.

Summary

Data from both the small and from larger-scale test facilities are now available to aid the design of PFBC systems.

The are areas of the technology where the data base is not as comprehensive as could be desired and hence additional flexibility will need to be built into the design of plant and/or a more cautious approach will need to be adopted. These areas include:

(i) Combustion efficiency - at low loads or with low volatile coals: there is uncertainty as to the extent of fines recycle needed to achieve or to maintain a high combustion efficiency and as to the effect recycle will have on the quantity and size distribution of the material to be removed from the combustion gases.

(ii) Sulphur retention efficiency - the effectiveness of recycling in enhancing sulphur retention needs to be fully quantified and as for (i) the consequences as far as hot gas clean-up is concerned needs to be assessed. There is also still a need for a simple laboratory test that will enable the potential effectiveness of a sorbent to be reliably quantified.

(iii) Elutriation - the data available so far do not enable reliable predictions to be made of the quantity or of the size distribution of the material that will be elutriated. Some skill is therefore needed in interpreting the data. The data also do not take account of recycling.

(iv) Tube bank design - the current indications are that from the point of view of minimising metal wastage, fluidising velocities at the top of the bed should not be higher than about 1 m/s. It may however be possible to operate with higher fluidising velocities at lower levels in a tube bank in a tapered bed without incurring unacceptably high rates of metal loss. Further development is needed to establish how best to arrange the tubing over the full depth of a bed whilst at the same time avoiding metal wastage even with the most aggressive coal ashes.

(v) Sintering of bed material - some mixtures of limestone and coal ashes show tendencies to form sintered agglomerates that can lead to defluidisation. The tendency can be reduced by operating at higher fluidising velocity (e.g. 1.5 m/s) in the lower part of the combustor - a feature of a tapered bed - and avoiding high bed temperatures (e.g. above about 880°C).

(vi) Control of load - control of load in a bubbling-bed system in most instances will involve change of bed level. The feasibility of changing bed level has been shown but the ability to follow the changes in load demand experienced in utility power systems over prolonged periods has yet to be demonstrated.

(vii) Hot Gas Clean-Up - the adequancy or otherwise of hot gas clean-up equipment to meet the requirements of a gas turbine in a combined cycle can only be established by carrying out tests in which the cleaned gases are expanded through a gas turbine having blading representative of commercial designs.

The aim should be to satisfy both the turbine and the EPA by clean-up equipment installed ahead of the turbine. The early development and application of the pressurised boiler cycle will provide a valuable means for establishing PFBC technology in the power industry and for obtaining experience with advanced designs of gas clean-up equipment under less arduous conditions than in the combined cycle.

In Conclusion:

It is justifiable to have confidence that economic solutions will be found to the remaining technical problems which need to be overcome before the PFBC combined cycle power plant is in commercial operation. The new programme at the UK Grimethorpe Test Facility can be expected to make a major contribution to this end.

Future Developments

Higher power generating efficiencies are potentially feasible by adopting the developments in steam pressures and temperatures envisaged for conventional pulverised coal-fired plant. With stream pressure of c. 300 bar and double reheat giving steam temperatures of around 620°C the currently envisaged efficiency about 40% could increase to about 42.5%

Acknowledgment

The views expressed in this paper are those of the authors and do not necessarily reflect those of the Sponsors of the projects with which they have been associated.

Table 1. Pilot Scale PFBC Rigs 150mm or Larger

Name	Rating MW(t) (Max.)	Pressure (Max)	Bed Plan Area m^2	Depth m (Max.)	Temp ^{o}C (Max)	Fluidising velocity m/s (Max)	Status
Argonne National Laboratory USA	0.15	6	0.02	0.9	980	1.8	A
NASA Lewis Laboratory USA	0.5	8	0.04	2.4	870	2.1	B
Exxon Eng. USA	1.7	10	0.07	c.3.6	980	2.1	B
New York University USA	7	7	0.46	3.6	950	2.4	A
General Electric Malta. USA	1.6	10	0.07	1.6	950	0.9	B
Curtiss Wright (Airheater) USA	4.9	6	0.66	4.9	900	0.8	C
NCB CURL 1 U.K.	6	6	0.56 – 0.88	2.8	950	2.1	C
NCB CURL 2 UK	0.3	6	0.07	1.2	950	3.7	C
NCB CURL 3 U.K.	1.2	20	0.09	3.1	950	1.5	C
IEA Grimethorpe U.K.	60	12	4	4.5	950	2.5	A
ASEA PFBC Sweden	15	16	2	app.3.7	900	0.9	A

Status: A - Operational
B - Not in use or Mothballed
C - Dismantled or being dismantled

Note: NCB-CURL Rig 3 being reassembled at NCB Stoke Orchard Laboratory.

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3. Hoy, H.R. & Roberts, A.G. "Investigations on the Leatherhead pressurised facility", 6th International Conference on Fluidized Bed Combustion, Atlanta, April 1980.

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7. Keairns, D.L., et al, Westinghouse 1973. "Evaluation of the fluidised bed combustion process". Final Report, July 1971 - May 1973. EPA Report EPA-650/2-73-048A.

8. General Electric Co., March 1981. "PFB coal-fired combined cycle development program. Final Report. US DOE Contract AC21-76ET10377: NTIS PC A08/MFA01.

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10. EPRI Journal, November 1983.

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16. 20 Atmosphere Rig PFBC Test Results, Test 13. "Investigations into the use of two-stage air admission", Work performed for US DOE under Contract AC21-80MC-14129, Document DOE/MC/14129-1501 (DE 84000229).

17. Ulerich, N.H., Newby, R.A., Keairns, D.L., "Calcium-based sorbent desulphurisation in pressurised fluidised bed combustion power plants", EPRI Report, No. CS-1847, May 1981.

18. 20 Atmosphere Rig. Investigations into the use of two-stage air admission, US DOE Report DOE/MC/14129-1501.

19. Burdett, N.A., Stantan, J.E., "The control of sulphur emissions from PFBC systems", Institute of Energy Conference at Leatherhead on effectiveness of fuel additives. Paper to Session 4, 4-5 October 1983.

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Fig. 1 STEAM-COOLED PFB COMBINED CYCLE

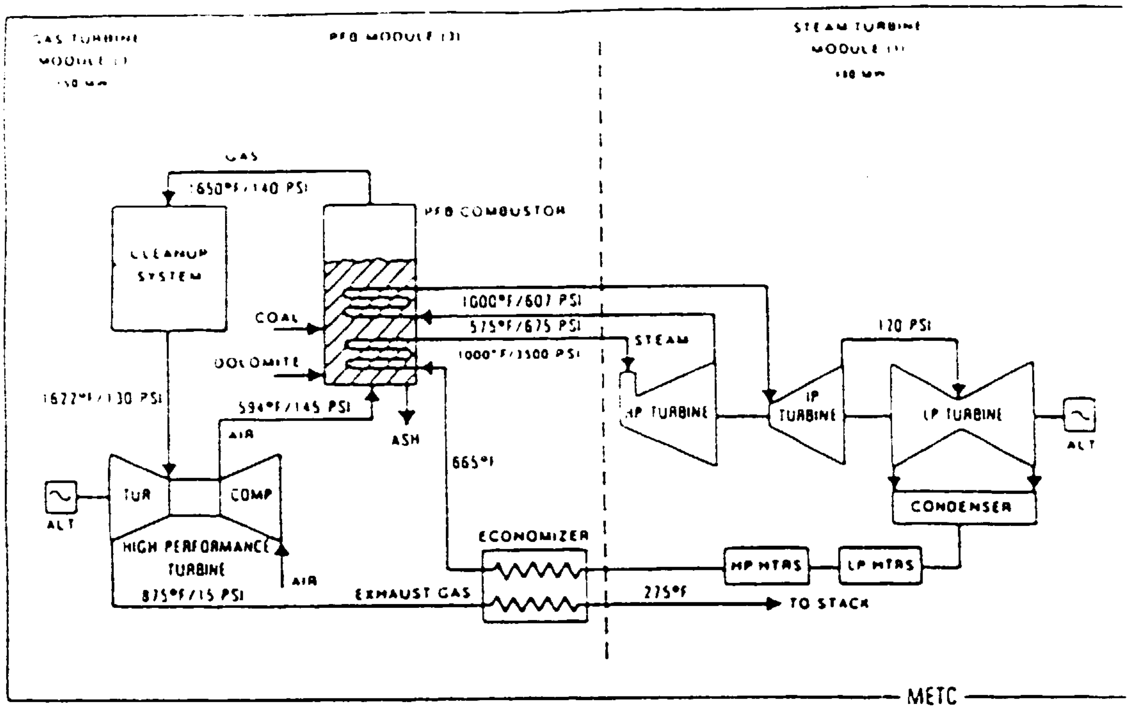

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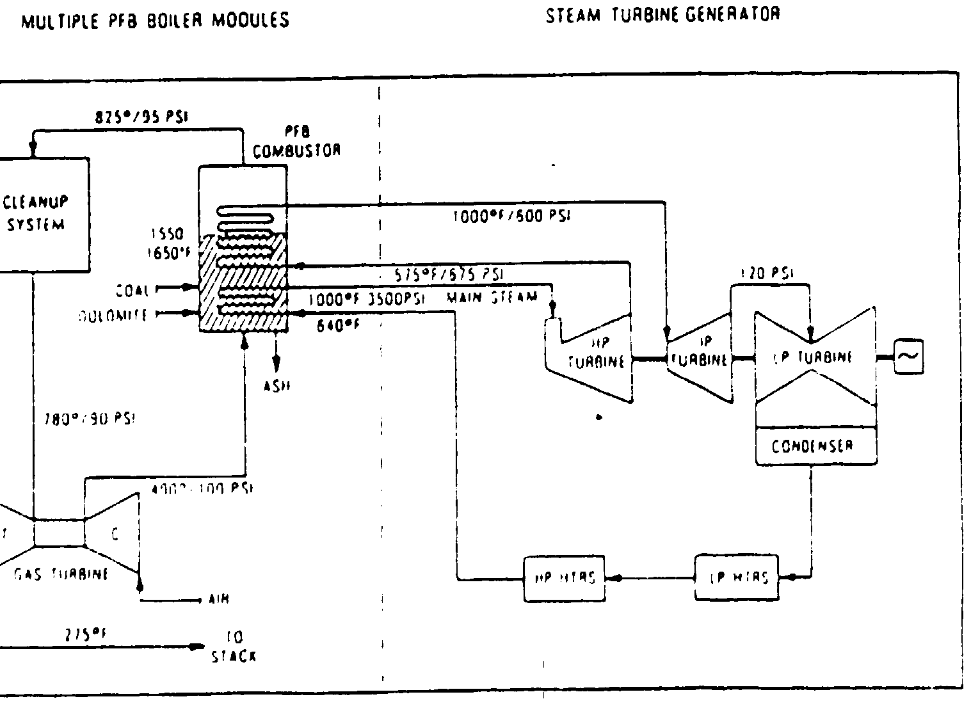
Fig. 2 Turbocharged PFB Simple Cycle



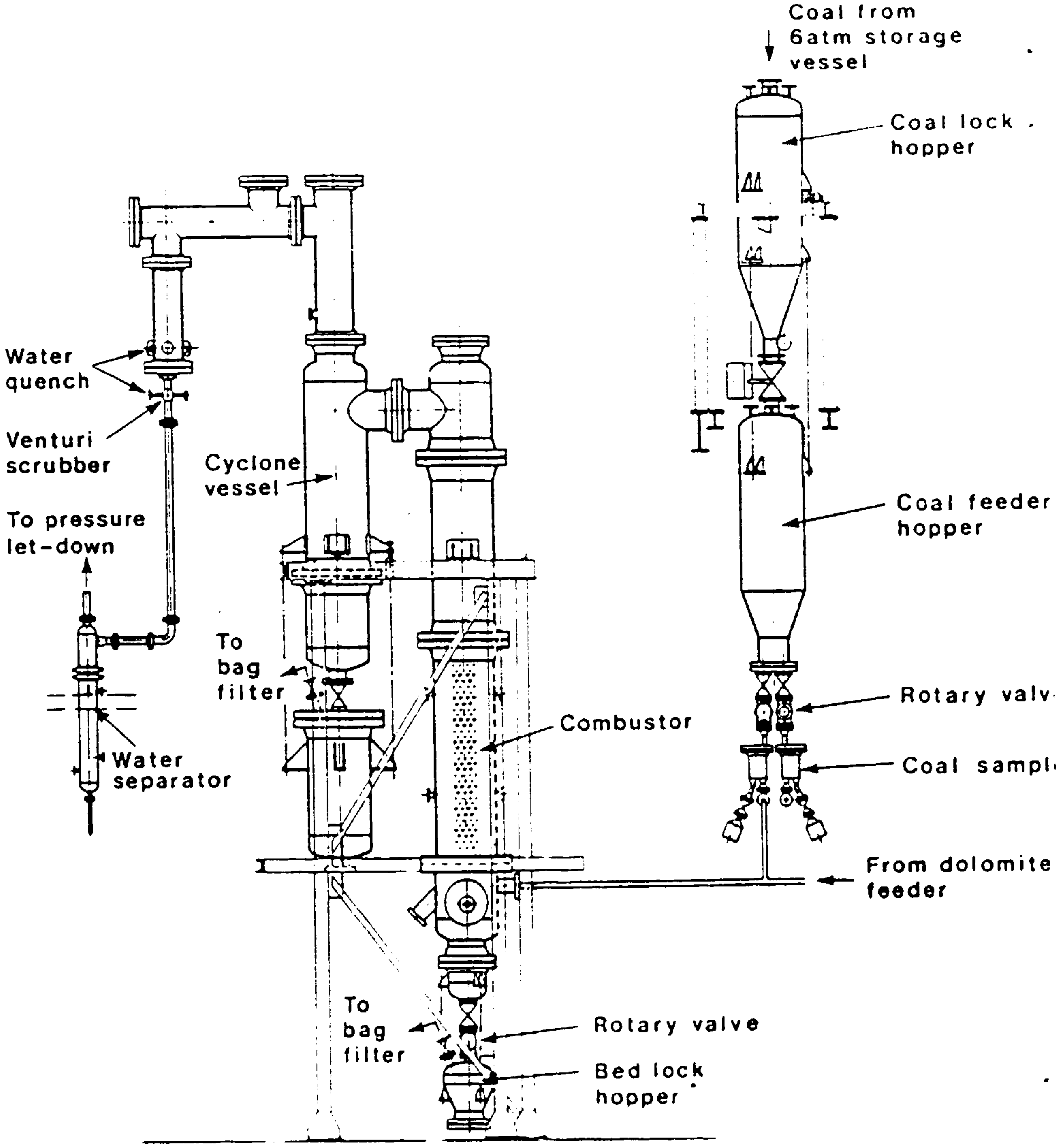
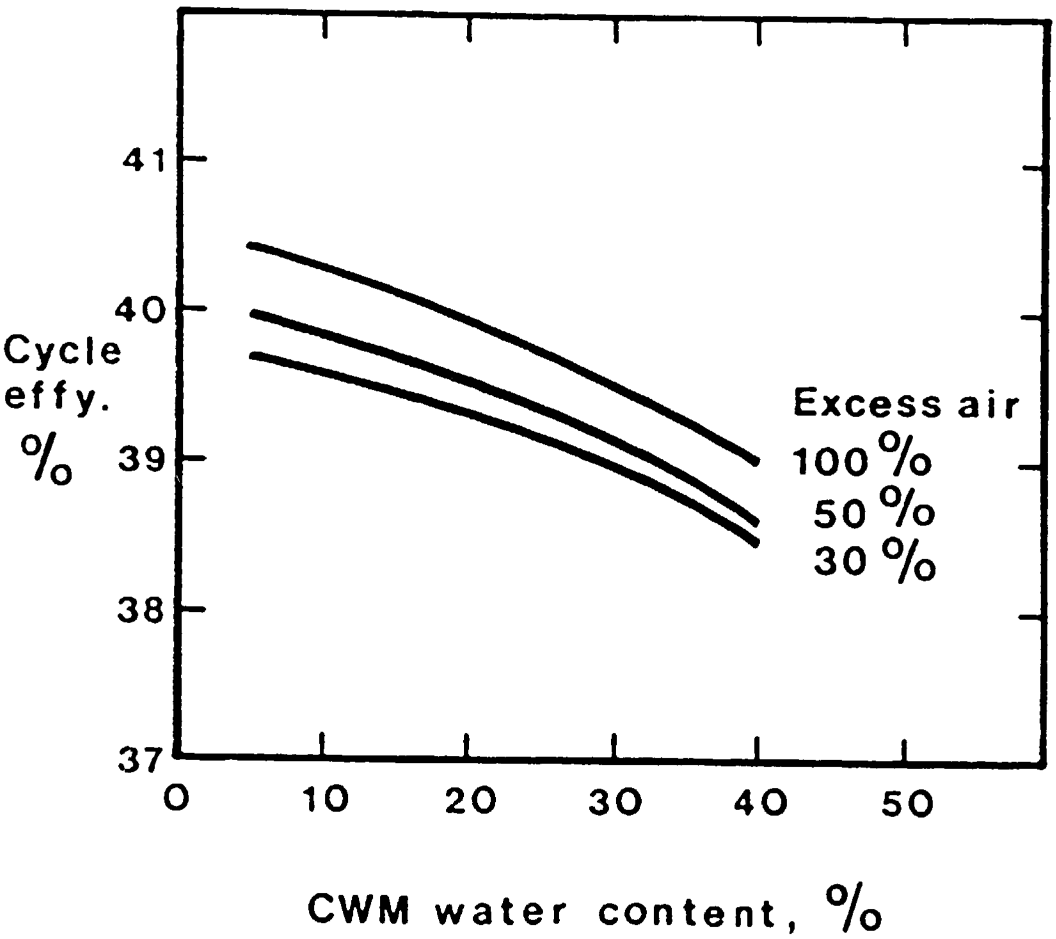
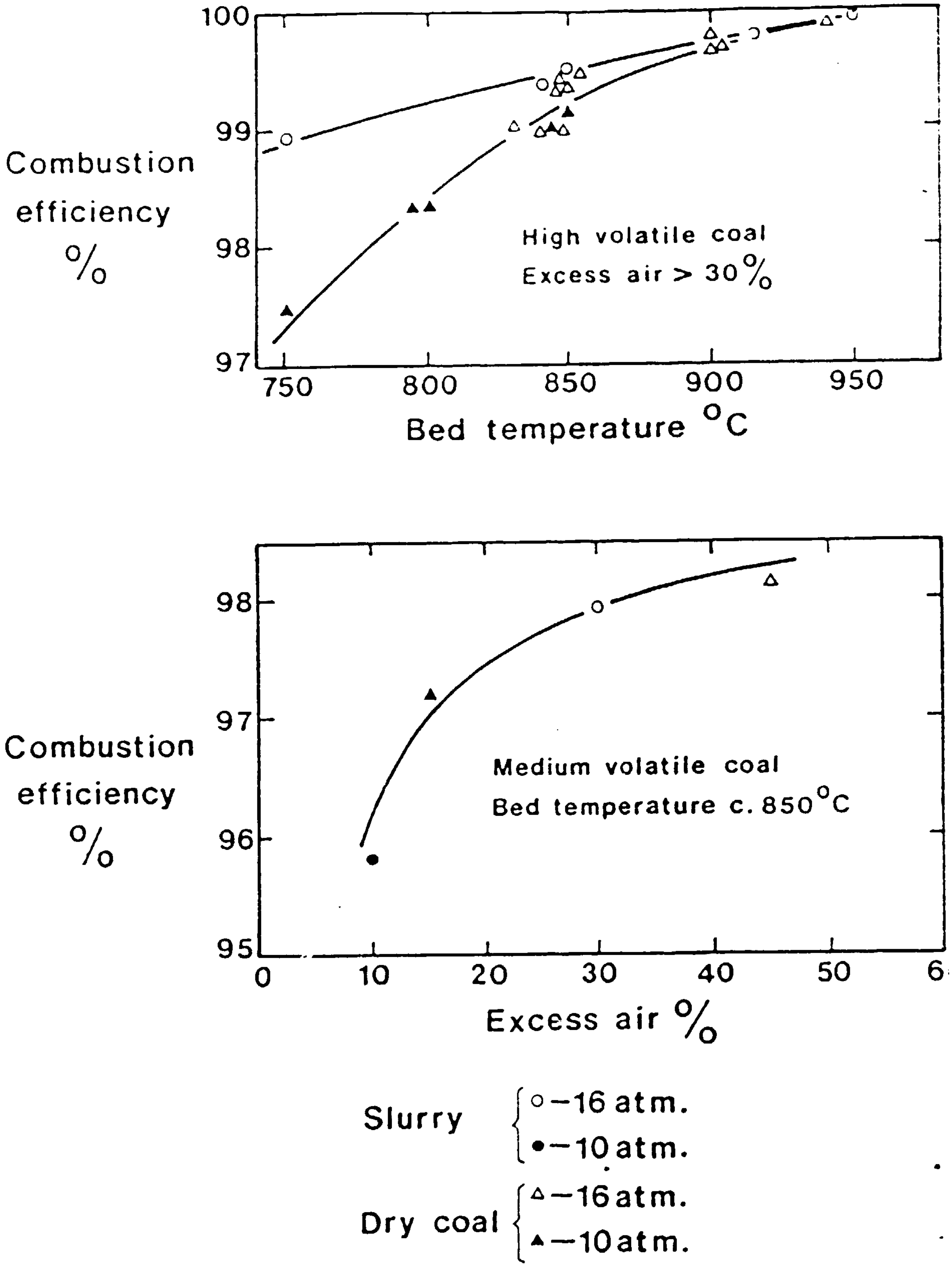


Fig. 3. 20 atm Combustor



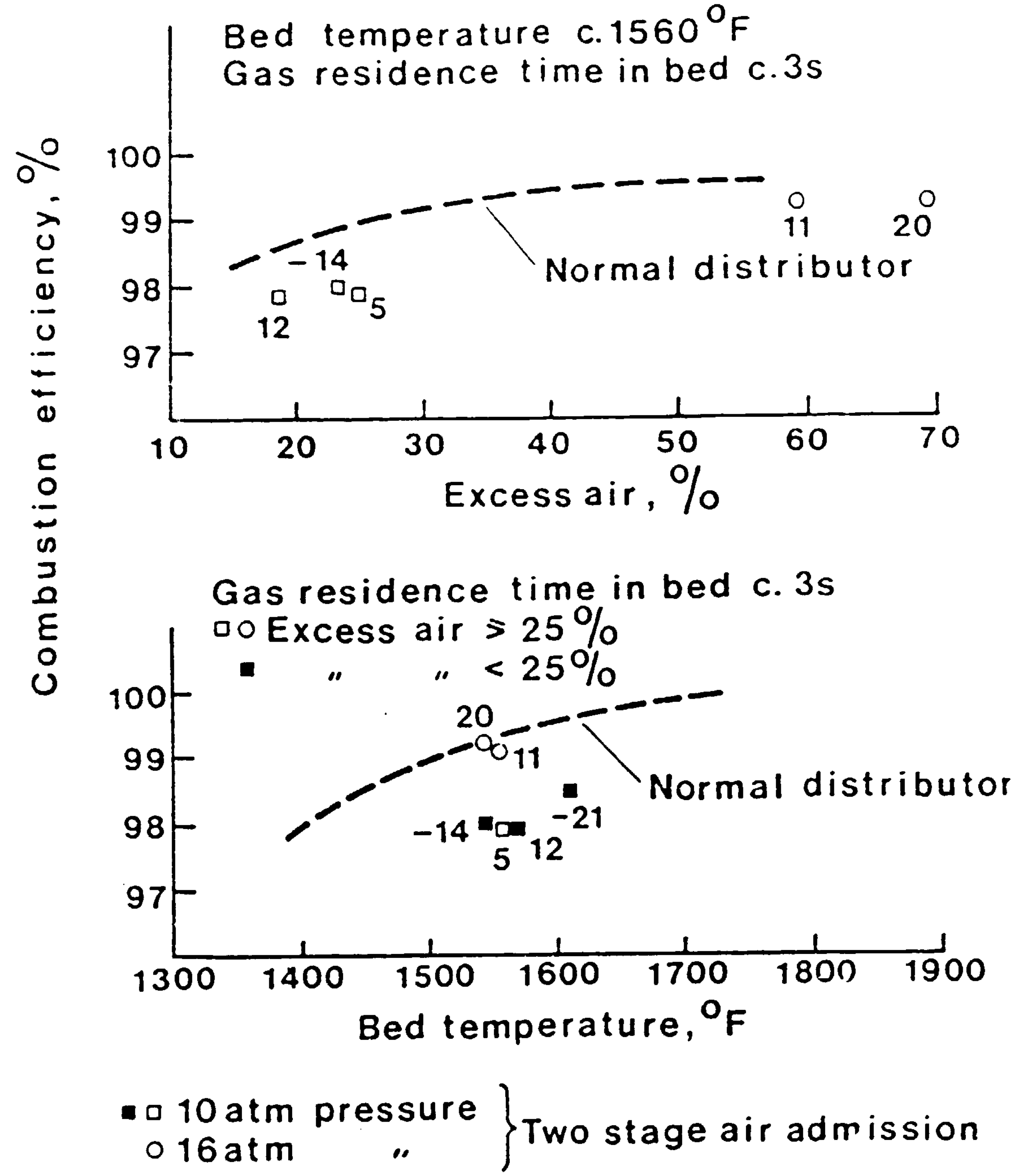
10 bar combined cycle PFBC plant
10 % ash in coal
Stack temperature : 280° F

Fig.4 Effect of fuel water content on cycle efficiency



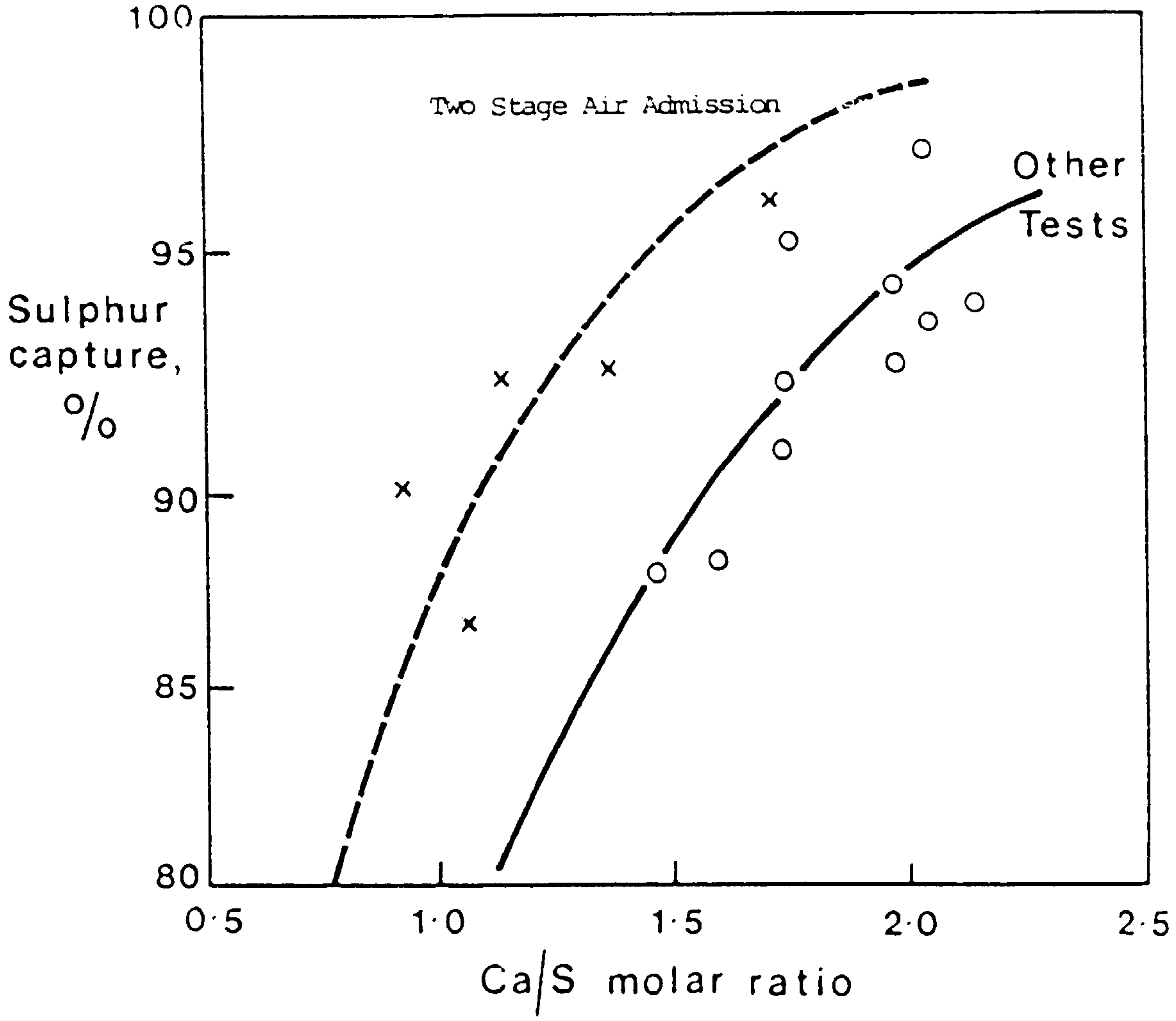
Gas residence time in bed c. 3secs.

Fig 5 Combustion efficiency. Comparisons between slurry and dry coal



Numbers alongside points are nominal excess air at a level below the 2nd stage air admission

Fig. 6 Effect of two stage air admission on combustion efficiency.



Data normalised to 3 sec gas residence
Bed temperature 1540 – 1560 °F
Mean sorbent particle size 600 – 700 μm

Fig. 7 Sulphur capture data

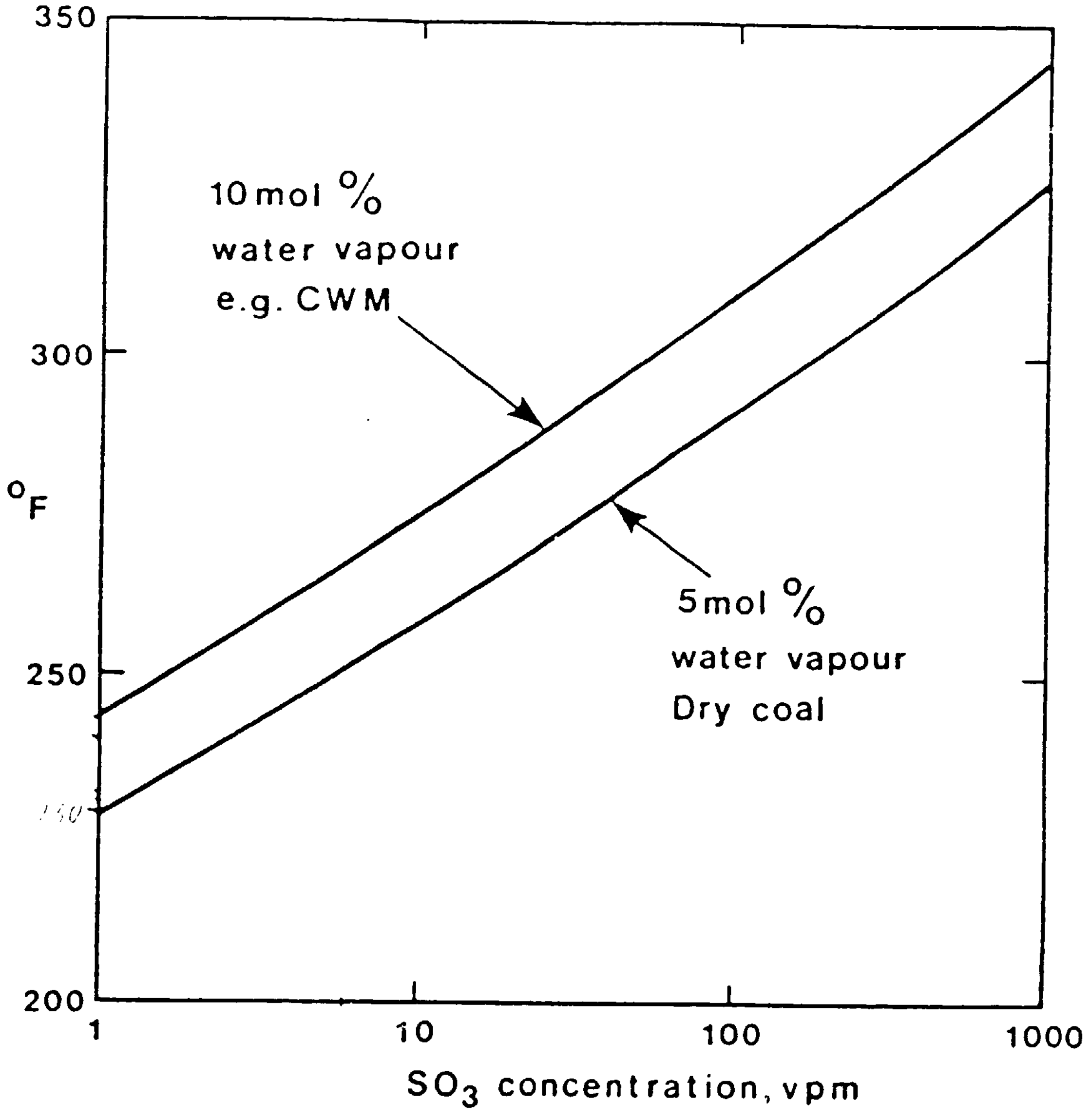
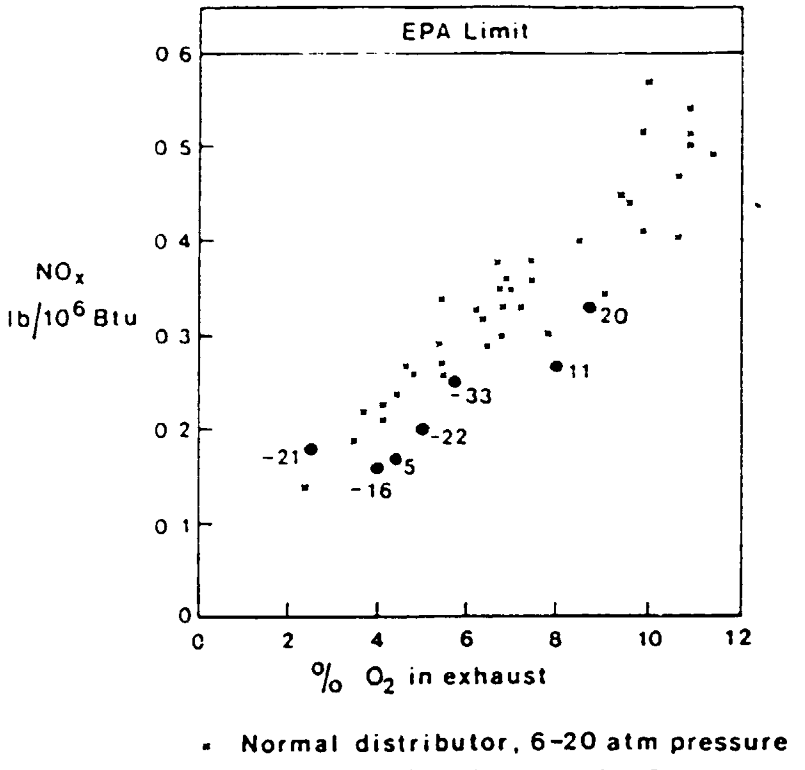


Fig.8 Effects of SO_3 and water vapour concentrations on the acid dew point



Numbers alongside points are nominal excess air at a level below the 2nd stage air admission.

Fig.9 Comparison of NO_x emissions. Two-stage air distributor and normal distributor

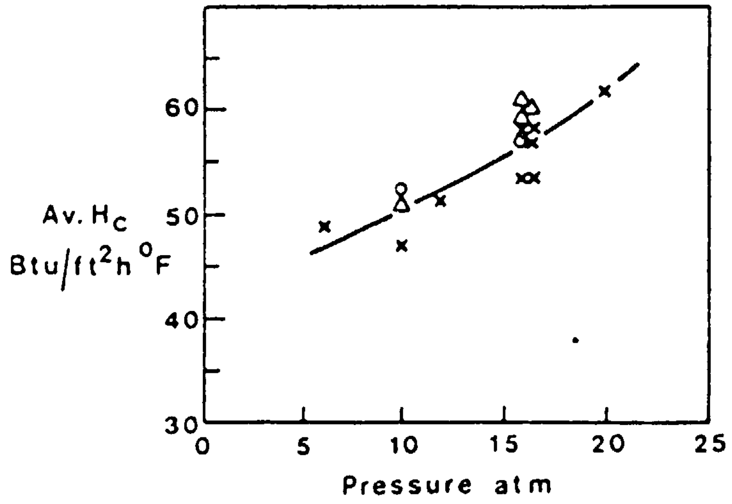


Fig.10 Effect of pressure on the average convective heat transfer coefficient

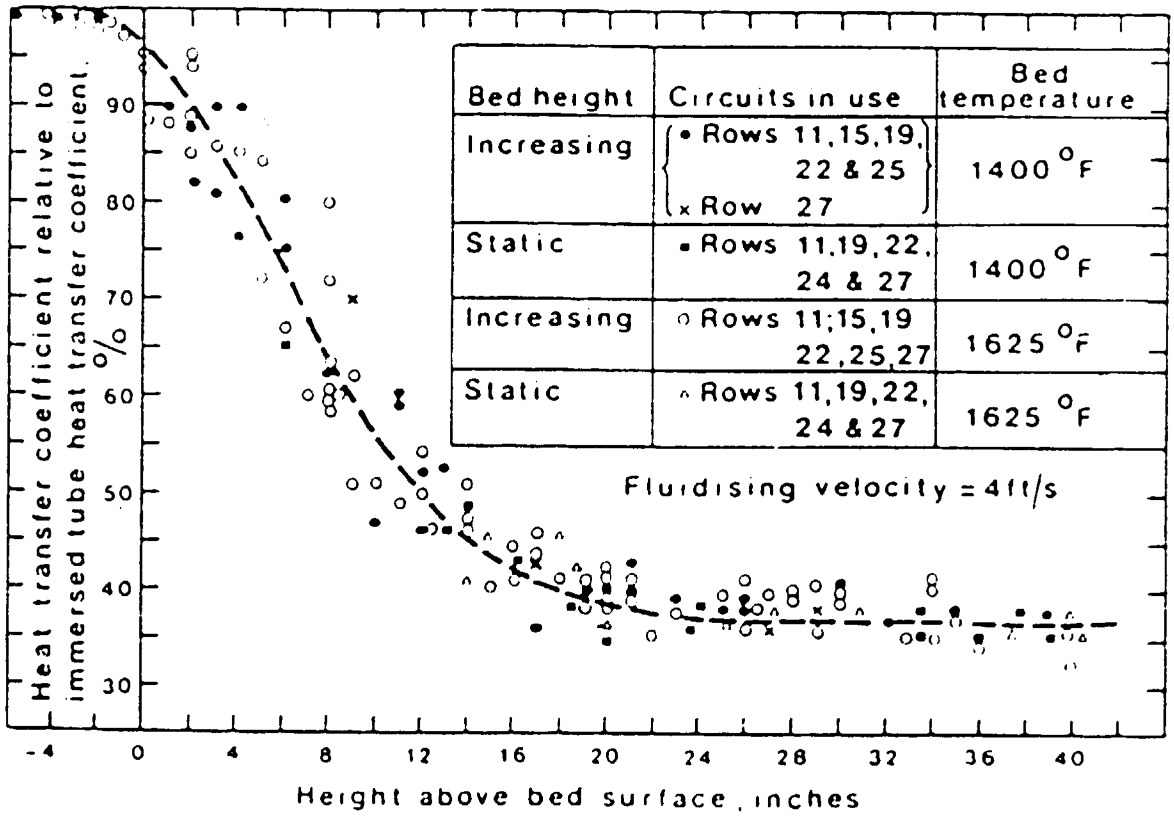


Fig.11 Heat transfer in the "splash" zone

Appendix 11

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31 August, 1984

as from Cadogan Consultants
37 Cadogan Street
London SW 3 2PR
UK

To
Dr.M. Maung,
IDO/Chem/IO
UNIDO, Vienna International Centre,
A-1400,
Vienna,
Austria

India Gasification Project:

I have tried to telephone, as you asked, but the switch board could not make connection. When we met in Vienna on 20th August 1984 you briefed me most interestingly on the Indian RRLH coal gasification and synthesis project. The report is under way. You also asked me to let you know about a number of points, both of principle and of detail, so here are the points:

1. The replacement of Dr.R.Vaidyeswaran, Director of the Coal Division of RRLH on his retirement (after July, 1986):

Dr. G. Thyagarajan, the RRL Director is concerned about this. Also Mr. Seshagiri Rao, Deputy Director, Coal Division and Mr. B.S.Narayana Rao, Assistant Director, are shortly to retire in 1986 and 1984 respectively (So too is the Director of Chemical Engineering). This is a common problem of new laboratories that recruited similarly - aged young staff, when they opened. Trawls are being made throughout India. There are expatriate Indians of leadership ability and initiative, but local salaries cannot attract them nor can local educational and domestic conditions, now they are used to Western ways.

...2...

This work must have a leader, and the RRLH Director is excellent. He would encourage the right Coal Division Director but first the right man (in his active 40's) must be found, and others to buttress him.

2. Relationship between the RRLH programme and the National Plan:

The relationship is one-sided apparently. RRLH thinking is generally in sympathy with the National Plan, but does not form part of it (so they say). The RRLH work is of course, part of the general programme of CSIR. I have recommended that RRLH form a small techno-economic study group that should (in time) help keep process aspirations in line with economic reality and also build up a useful body of expertise and data.

Also, no process (on a scale relevant to coal gasification and synthesis) could, from RRLH now, meet a market demand in under 10 years - more like 15 or more. So I have recommended a marketing study to consider markets for products from today, over the next couple of decades. Also, it may well be necessary for India to 'buy in' some technology to add to what it develops itself.

RRLH gasification work is good, but at too early a stage for the National Plan to rely on it.

3. How much work has been done on the plant?

Good, useful work, still trouble-shooting and training - only now on oxygen (previously air). Waste-heat-boiler water-levels surge; grate drive seals leak; valve spindles seize. There are problems with the tar/liquid condensation and circulation. These are common "teething troubles" and there will be more. The synthesis catalyst work is good, but has as far to go as gasifier characterisation of coals.

What is needed now is a programme and an increasing intellectual/commercial discipline. RRLH seems to be capable of providing the right background and environment - and has had successes in the past. Some "catalysis" of the staff during this coming difficult period of changing leaders, would be of great benefit.

4. The purpose of the gasification/synthesis work - support?

This seems to be largely self-propelled by RRLH, to make use of its expertise and to keep level-pegging with CFRI and other initiatives (such as those of BHEL). This healthy competent competition is probably very good. But RRLH will probably value some external influence to appear at intervals and help keep things moving forward. They will get the best support when their results justify it. This will take time.

5. RRLH-University Bridges:

There are many, it seems, but on a personal level, not as regards research work. There are probably many minor, but useful research contracts that RRLH could place with universities. The 3-year Ph.D time table will have to be reckoned-with, providing yet another reason for deriving a programme of work bringing in research, progress elsewhere, techno-economic studies, and marketing realities, as well as all the interlocking aspects of the gasification, catalyst and synthesis work that will be necessary. At present the work is (apparently) organised ad hoc.

6. Use of gasifier ash:

There is no real consideration, now of the use of ash. I have outlined some possible uses - but at the present stage of the work, ash is of low priority. As disposal is, of course, a problem throughout coal-using industry throughout India.

7. **Fellowships:**

Some should be useful - see report

8. **Need for a return visit:**

I am making some recommendations for activity and the Director of RRLH is keen on the proposal to set up a small (part time) techno-economic unit. A return visit in perhaps six or twelve months would be useful to help keep the programme moving, and to ensure planned rather than ad hoc activity. After that some regular arrangement could be decided. The objectives of the work are worthwhile and it will be worth making sure that it makes real progress. The first priority is to operate the gasifier satisfactorily.

9. **The general usefulness of the work:**

This work could lead to a major advance in India. It could also provide a base from which to compare bought-in processes from abroad. The competence is available, and one can see useful indigenous processes being developed for use by Indian industry. Encouragement will be needed, as the programme must keep pace with what is being done elsewhere in the world.

I hope this may be helpful. Perhaps we can talk about the implications when you have had an opportunity to consider my report.

I shall look forward to hearing from you,

With kind regards,

signed

Brian Locke

Appendix 12

Literature given to RRL by the Consultant

A Papers by others

Hebden D, Edge R F and Foley K W	"Investigations with a Small Pressure Gasifier" Gas Council, Research Communication G C 14 : November 1954
Ricketts T S	"The Westfield High Pressure Coal Gasification Plant" I Gas E Publication 567 : June 1960
Ricketts T S	"The Operation of the Westfield Lurgi Plant etc" I Gas E Publication 633 : May 1963
Harvey F	"High Pressure at Solihull" I Gas E, Midland Section : 5 March 1965
Elgin D C & Perks H R	"Results of Trials of American Coals in Lurgi Pressure-Gasification Plant at Westfield,Scotland" British Gas
Elgin D C	"Gasification of US Coals at Westfield, Scotland" I Gas E Communication 946 : November 1974
Jones J E M	Commissioner, Summary Report (Full scale tests of Kentucky No 9 Coal in a Lurgi Gasifier at Sasol One) Bureau of Energy Research and Development : August 1982 Kentucky

Roeger A and
Jones J E M "Testing Kentucky Coal to set Design Criteria for a Lurgi Gasification Plant" 5th Annual Energy Conservation Technology Conference and Exhibition, Houston, Texas : April 1983

British Gas - International Consultancy Service - Brochure

B Papers by the Consultant

Potter N & Locke H B "Coal Supplies for Carbonisation Processes" The Mining Engineer : Vol 125, June 1966

Locke B & Lunn H G "Advances in Pollution-Free Heat and Power Generation made possible by Developments in Fluidized Combustion Technology" Achema, Frankfurt, Germany : June 1976

Locke B & Schomburgh I "The Introduction of New Technology into Systems and Plants" I Chem E Jubilee Symposium, Imperial College, London April 1982

Locke B "Energy Flux in Chemical Engineering Design" I Chem E Symposium Series No 78 : October 1982

Locke B, Choudhury R & Lecamwasam D C D "Indigenous Rural Power Centres" draft for Institution of Mechanical Engineers, London, April 1985

C Papers in this Report

Roberts A G, Hoy H R, Lunn H G, & Locke H B, "Fluidized Combustion of Fossil Fuels" Coal Processing Technology, Vol 2, 1975

Locke B, "Feedstocks and Fuels from Coal : The Need for Process Changes" Energy World : February 1982

Hoy H R, Stantan J E, Roberts A G, "Status of Pressurized Fluidized Bed Technology Toady - A Review". Engineering Foundation Conference, Combustion of Tomorrow's Fuels, London, October 1975

D Manuscripts shown to RRL

Original manuscripts concerning British Government contract work done by the British Coal Utilisation Research Association were discussed, but, of course, could not be given to RRL.

They covered subjects such as
slagging gasifier design,
calculating performance data,
tuyere raceway geometry,
slag analysis/viscosity relationships,
carbonisation under pressure, and
monthly progress reports;
for example.

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