STUDIES OF ADVANCED ELECTRIC POWER GENERATION TECHNIQUES AND COAL GASIFICATION

BASED ON THE USE OF HAT CREEK COAL

Prepared for

BRITISH COLUMBIA HYDRO AND POWER AUTHORITY

and

ENERGY, MINES AND RESOURCES CANADA

by

INTERCONTINENTAL ENGINEERING

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PREFACE

In 1975 B.C. Hydro and Energy, Mines and Resources Canada commissioned five studies to investigate potential uses of Hat Creek coal. Three of the studies were directed towards advanced high efficiency, clean methods of generating electric power, and alternatively, to producing synthetic natural gas, while a fourth examined the use of Hat Creek coal in the existing oil/gas fired Burrard plant.

The fifth study was assigned to a 'co-ordinating consultant' who was responsible for co-ordinating the work of the other four studies. The co-ordinating consultant was also directed to produce a summary report examining and comparing the results which were derived in the other studies. The summary report is included in Volume 1 of this report. The three studies examining advanced electric power generation and gasification are included in Volume 2 and the Burrard conversion study in Volume 3.

STUDY A — FLUIDIZED BED COMBUSTION STUDY

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1. SUMMARY

This report covers a study of proposed schemes for 2000 MW of power generation from the Hat Creek, British Columbia, sub-bituminous coal deposit using fluidized bed combustion technology. The technology is briefly described.

The Hat Creek coal characteristics are considered and found suitable for fluidized bed combustion subject to tests in experimental rigs.

A scheme using atmospheric pressure boilers with steam turbine generators and a scheme using pressurized boilers in a combined cycle with gas and steam turbine generators are chosen for detailed study. The unit sizes chosen are 648 MW and 623 MW respectively.

Plant layout and cycle drawings are presented and the stations are described. Present day capital costs are estimated at $435 per kW and $392 per kW for the atmospheric and pressurized schemes respectively. The corresponding power costs are estimated at 11.2 mills per kWh and 10.3 mills per kWh at 80% load factor.

For comparison purposes, capital costs and power costs based on alternative estimates of interest during construction are presented.

Construction shedules are included. The earliest in-service dates are assessed as 1983 for atmospheric units and 1988 for pressurized units.

The quantities of solid, liquid and gaseous effluents discharged are estimated and found to be within current provincial regulations. Thermal and water vapour discharge from the cooling towers are estimated.

The atmospheric scheme is considered feasible at present, although not without risk, but doubt is expressed concerning the pressurized scheme as a consequence of the early stage of development. The installation of pilot/demonstration plants is briefly considered.

No similar process being extensively developed was identified. Particulars are given of the Ignifluid process.

2. INTRODUCTION

This study was performed by *Engineering and Power Development Consultants* Limited of Marlowe House, Sidcup, Kent, DA15 7AU, England in association with Combustion Systems Limited of Kingsgate House, 66/74 Victoria Street, London, SW1E 6SL, England at the request of the British Columbia Hydro and Power Authority, Vancouver, British Columbia, Canada. The request was received via Intercontinental Engineering Limited of Vancouver, British Columbia, Canada, who were the co-ordinating consultant for this and other studies being performed concurrently.

The broad subject of the study was to consider the use of the Hat Creek coal deposit for electrical power generation by combustion in fluidized combustors. The full terms of reference for the study were as follows:-

TERMS OF REFERENCE

FOR

FLUIDIZED BED COMBUSTION STUDY

1. Provide engineering services to determine the feasibility and cost of a thermal generating station equipped with fluidized bed combustion furnaces. Consider appropriate unit sizes for a total installation up to 2000 MW of conventional or combined cycle thermal plant.

   The study will include a detailed review of:-

   a) Atmospheric fluidized combustion in combination with conventional steam turbines;

   b) Pressurized fluidized combustor furnaces in a combined cycle configuration with gas and steam turbines.

   Cost data associated with the pressurized cycle will be indicative because the state of development of this cycle precludes accurate estimation.

2. The study will incorporate a materials and energy balance for each of the main alternatives.

3. The study report will include the following information:-

   a) Comments on the feasibility of the alternatives considered;

   b) Statement on reasons for choosing the unit size used in the study.

4. The study will include a listing and a brief review of all known similar processes which are the subject of a major development effort, including, in particular,

the Ignifluid process. The review will incorporate statements on the schedule for development, the mechanism of the process, the possibility of the process becoming attractive commercially, and any special advantages and disadvantages.

5. Identify the possible environmental impacts of such a station in relation to accepted or assumed emission standards. This will include a flow balance for all gaseous, liquid and solid discharges. The site dependent environment impacts will be excluded.

6. The station would be located in the vicinity of Hat Creek and would be assumed to burn Hat Creek coal. At a later stage in the studies, data will be provided on East Kootenay coal. The study will incorporate a brief general analysis of the qualitative changes in the technical results and cost estimates in the study.

7. The work shall be in the form of engineering studies carried out utilizing published information and data from discussions with companies considered to be recognized authorities in the field having regard to present technology and possible technology in the future. In particular, the study will incorporate technical and cost data from Combustion Systems Ltd. (CSL).

8. Power cost estimates expressed in mills/kWh are to be calculated for range of capacity factors from 60% to the highest considered feasible for the schemes studied. Coal characteristics and costs will be provided by B.C. Hydro from existing data and, as study progresses, from sample tests. Capital cost estimates shall be broken down to clearly itemize the component costs.

9. Cost estimates shall be in September 1975 dollars and shall be broken down by years. Where possible, agreed common costs received from the co-ordinating consultant will be incorporated. The interest on capital and interest during construction shall be assumed at 10% but itemized in such a way that the effects of an alternative rate can easily be determined. The assumed plant lives will be agreed with B.C. Hydro.

10. Project schedules shall be prepared for the earliest in-service dates for various sizes and systems considered.

11. Prepare and submit a report in draft form by 30th September 1975 and in final form by 28th November 1975. In addition, progress reports will be made monthly of the results achieved, the costs incurred and the scheduling of future work and associated costs.

12. Provision shall be made for co-ordination of the work with other parallel studies which are to be undertaken of conventional thermal and coal gasification systems.

13. The study is to be controlled and co-ordinated by the Assistant General Manager, Engineering of B.C. Hydro and Power Authority or his appointee.

With regard to item 6 of the terms of reference, data on East Kootenay coal had not been received at the time of completion of this report.

3. BASE DATA AND ASSUMPTIONS

The base data used in the study was extracted from Issue 3 of the Integ document "Coal Gasification and Related Studies — Base Engineering and Cost Criteria", dated August 6th, 1975, as amended by Addendum 1 dated August 14th, 1975.

Generally any assumptions made were minor and are stated at the relevant points in this report. However, some more major assumptions are given below.

It has been assumed that the Hat Creek coal, when burned in a fluidized combustor will not exhibit abnormal characteristics in comparison with other coals that have been examined previously.

It has been assumed that development work on fluidized combustion will proceed and that its results will be as expected.

It is considered that there is substantial evidence that these two assumptions will be substantiated.

It has been assumed the discharge of liquid effluents from the station will not be permitted.

4. FLUIDIZED COMBUSTION FOR POWER GENERATION

The application of the technology being developed for the combustion of coal in a fluidized bed is expected to offer advantages over more conventional methods of coal combustion in the power generation field. Among the expected advantages are:-

a) Lower capital cost of plant.
b) More prefabrication of boiler giving improved quality control and shorter site construction time.
c) Less gas-side corrosion and fouling.
d) Reduced emission of oxides of sulphur and nitrogen.
e) Poor quality fuel can be burnt without difficulty.
f) Achievement of coal-fired combined gas turbine/steam turbine cycle with consequent high efficiency and low fuel cost element of the power cost.

The basic principle of the CSL system for the fluidized combustion of coal is that crushed coal is injected into and burnt in a fluidized bed of non-combustible material (Reference 1). The fluizided bed is formed by passing air upwards into the bed at a rate sufficient to fluidize the bed at the desired fluidizing velocity. The fluidizing air also serves to provide the air needed for combustion.

The process can take place either at approximately atmospheric pressure or at some higher pressure. A characteristic of the latter is that the combustor dimensions are very substantially reduced for the same heat output.

It is a feature of the system that the temperature of the bed is maintained in the range of 750 to 950°C (1382 to 1742°F). One important reason for avoiding a higher temperature is that ash softening temperatures should not be reached. In comparison with conventional coal combustion processes, bed temperatures in that range permit easier control of emission of oxides of sulphur and result in lower emission of oxides of nitrogen.

In order to maintain the bed at the desired temperature, heat is extracted from it by some means other than removal of the products of combustion. This can be effected advantageously by heat transfer surface both surrounding and within the bed. It is a feature of fluidized beds that high heat transfer coefficients are obtained to immersed surfaces. The heat transfer surface is normally used to generate steam or to heat air.

4.1 ATMOSPHERIC PRESSURE UNITS

Atmospheric pressure fluidized combustion units for coal-burning in power stations are expected to have many similarities to conventional power station boilers.

They would be working in steam cycles the same as those for conventional boilers and would therefore contain steam generation sections, superheaters and often reheaters.

Fans would be used to provide the fluidizing/combustion air. Regenerative air heaters and economisers would usually be economically justified.

Cyclones would be used for coarse grit and dust removal from the flue gases, with electrostatic precipitators for final clean-up before discharge to the stack.

The coal preparation and injection equipment would differ substantially from that used on pulverized coal-fired boilers. The coal would first be crushed to a size to suit the characteristics of the bed. This would usually be in the range of 1/8 inch - 0 inch to ¼ inch - 0 inch. Secondly the coal would be injected into the bed at a sufficient number of points for it to be distributed adequately by the turbulence of the bed so that it came into contact with sufficient air for combustion. This latter requirement would probably be best met by a pneumatic transport system.

Usually ash would be removed from the bed to prevent accumulation of bed material and this could be achieved by weirs at the desired top level of the bed.

The application of these principles to a power station boiler is described in 7.2 below.

4.2 PRESSURIZED UNITS

Important features of pressurized fluidized combustors for coal burning in power stations (References 2 and 3) are expected to be their small size and very high rates of heat release per unit volume.

It is anticipated that each complete combustor would be enclosed in a cylindrical pressure vessel which might be vertical or horizontal. Within the vessel, fluidized combustion beds would be mounted.

Coal and ash would enter and leave the vessel via pressure locks. Connections on the vessel would admit air for fluidizing/combustion and discharge flue gases. Combustors designed for steam raising would also have penetrations for steam/water pipes.

These units would normally supply hot flue gases to gas turbines, and high efficiency flue gas cleaning equipment operating at combustor pressure and temperature would be required to render the gases suitable for long-term gas turbine operation. Suitable gas cleaning equipment is currently being actively developed. This might take the form of multiple centrifugal dust separators and filter beds.

The application of these principles to a power station generating unit is described in 8.2 below.

4.3 CYCLES — GENERAL

Many cycles, both conventional and advanced have been proposed utilising the fluidized combustion of coal for power generation. A broad division can be made between cycles using atmospheric pressure combustion and those using pressurized combustion. Further sub-divisions can be made between those using steam turbine generators and those using a combination of the two, generally referred to as "combined cycles".

The pressurized fluidized combustion process depends upon the combustion/fluidizing air being compressed to the process pressure. The gaseous products of combustion and the excess air leave the process at this pressure. This leads naturally to the use of gas turbines in cycles employing pressurized fluidized combustion so that the combustion air can be compressed by the compressor of the gas turbine unit and the flue gases expanded in the gas turbine. Thus cycles using pressurized fluidized combustion generally include gas turbine plant.

The broad division between atmospheric and pressurized combustion is amplified in the detailed consideration of cycles that follows.

4.4 CYCLES FOR ATMOSPHERIC PRESSURE UNITS

Current designs use conventional steam cycles in the application of atmospheric pressure fluidized combustion. No special problems are envisaged in providing coal-fired atmospheric pressure fluidized combustion boilers suitable for use with steam turbine generators of the maximum size and steam conditions currently available. The fluidized combustion boilers can be regarded as the equivalent of the types of boilers currently used in power stations and could take their place, directly, in the same cycles.

4.5 CYCLES FOR PRESSURIZED UNITS

A variety of cycles have been proposed (references 2 and 3). These cycles include gas turbines to utilise the pressurized exhaust gases from the combustor outlet.

A fluidized combustor for burning coal requires heat to be removed from the bed in order to maintain the desired bed temperature (see 4.0 above). This can be effected by tubular heat transfer surface within and surrounding the bed. (Systems using large quantities of excess air to maintain bed temperature and thus avoiding the use of heat transfer surface are not considered in this report). The cooling medium may be water/steam, air or other fluids. This choice of cooling fluid can lead to a variety of proposed cycles.

The following are typical of the cycles proposed:-

a) Open cycle gas turbines in which the bed cooling medium is air supplied by the gas turbine unit compressor. Air from the compressor is split into two streams, one is used to provide the fluidizing/combustion air and the other is passed through tubes immersed in the bed. After cleaning of the flue gases, the two streams are mixed and expanded through the gas turbine. Waste heat may be recovered from the turbine exhaust gases.

b) A combined gas turbine/steam turbine cycle in which the bed cooling medium is steam/water which is used in a conventional reheat steam cycle with multi-stage regenerative feed heating. The combustion air is supplied by the gas turbine compressor. The cleaned combustor exhaust gases flow to the gas turbine which drives the compressor and a generator. Waste heat is recovered from the gas turbine exhaust gases by economisers integrated in the steam cycle feed heating system. Such a combined cycle is expected to have a higher efficiency than a conventional steam cycle.

The main line of development for large power outputs appears to be concentrated on cycles of these two types.

5. SUITABILITY OF HAT CREEK COAL

The suitability of the Hat Creek coal for combustion in a fluidized bed has been considered by Combustion Systems Limited by examining the coal analyses and other data supplied by B.C. Hydro and Power Authority.

This consideration has not revealed any characteristics of the coal that would preclude its use in fluidized combustors. However, as indicated in discussions prior to the commencement of this study, it would be necessary to undertake a series of experimental combustor studies using the anticipated coal blend before detailed design of a full-scale generating unit.

It has been assumed for the purposes of this study that the coal behaves "normally".

6. CHOICE OF SCHEMES FOR STUDY

6.1 GENERAL

The application of fluidized combustion to power generation has not yet reached a stage at which the optimum power plant design can be established from a basis of existing installations. In particular, there is insufficient evidence at this time to make a prior choice between atmospheric and pressurized fluidized combustion.

In view of these factors a basic decision was made, and incorporated in the Terms of Reference for the study, to include schemes for both atmospheric and pressurized fluidized combustion.

Some factors influencing the more detailed choice of schemes were the required power station capacity of up to 2000 MW, and the availability of data from previous studies. This latter factor was important because development of a cycle and equipment designs from the general principles of fluidized combustion technology was not possible within the required schedule and cost for the study.

It was considered that there were no factors that were likely to cause fluidized combustion plant to differ from conventional plant in that the specific cost reduces as unit size increases. It was therefore expected that the cost per kilowatt of fluidized combustion plant would become smaller as the unit size became larger provided the limits of existing technology were not exceeded.

6.2 ATMOSPHERIC PRESSURE SCHEME

There was little room for variation in choosing the atmospheric pressure fluidized combustion scheme because units that have been proposed are steam boilers capable of generating and reheating steam at the conditions commonly in use in power stations throughout the world. Bearing in mind the comparatively low cost of the coal to be used and the required power station output, a conventional reheat steam cycle was selected with steam conditions at the turbine stop valve of 2315 psia and 1050°F with reheat to 1050°F. A seven-stage regenerative feed heating scheme was selected, with a final feedwater temperature of 490°F.

Further comments on the selection of this cycle are as follows:—

STEAM CYCLE

No cycle, other than a conventional steam cycle, was identified as having been developed for atmospheric fluidized combustion.

REHEAT CYCLE

A non-reheat cycle would be possible and might be economic for this low coal cost plant. However, data was not available from an earlier study and also it was not anticipated that the differences in power cost would be very significant.

STEAM PRESSURE

A pressure near 2315 psia is commonly used for reheat cycles of the output envisaged. It appeared unlikely that a reduction in pressure would reduce the power cost. It seemed even more unlikely, in view of the low coal cost, that a supercritical cycle would be economic.

STEAM AND REHEAT TEMPERATURE

It is appreciated that temperatures of 1000°F are more commonly used than 1050°F at present. However, there does not appear to be any technical objection to using 1050°F for coal-fired plant and in fact CEGB have standardised on this for their large coal-fired units. The previous study on which the design is based used a temperature of 1050°F and it was therefore preferable to use it for the present study. In view of this and also because a higher capital cost would probably be offset by a lower fuel cost element of the power cost, temperatures of 1050°F were used.

SEVEN-STAGE FEEDHEATING TO 490°F

A feed heating plant with fewer stages and/or a lower final feed temperature might be economic for this low coal cost plant but, for simplicity, a standard arrangement was used.

6.3 PRESSURIZED SCHEME

The gas turbine units currently available from the principal manufacturers could give an output of about 70 MW when used with fluidized bed coal combustion. It does not appear that units of any significantly larger size will be available in the next few years and a 2000 MW power station would therefore require about 30 units of the size available if power generation was solely by gas turbine generators.

The large amount of equipment that would be needed in a 2000 MW station producing power from gas turbine generators alone led to the rejection of this possibility in favour of a combined gas turbine/steam turbine cycle in which the ratio of steam turbine generator power to gas turbine generator power is typically 3 to 1 and a station of nearly 2000 MW requires only 3 steam turbine generators and 6 gas turbine generators.

Many forms of and refinements to the combined cycle are possible and it was necessary to lean heavily on the results of previous studies. Basically the cycle chosen has four pressurized fluidized combustors burning coal and supplying hot pressurized flue gas to two turbine generators. The gas turbines, in addition to driving generators, drive the compressors that supply fluidizing/combustion air to the combustors.

Heat is extracted from the fluidized beds by the steam cycle which has similar steam conditions to those described in 6.2 above for the atmospheric pressure unit. A single steam turbine generator is used with feedwater heating by steam extracted from the turbine, by the gas turbine intercoolers and by the gas turbine exhaust gases in economisers.

Similar comments on the choice of steam cycle and conditions apply to this scheme as described in 6.2 above. A higher final feedwater temperature was chosen to facilitate boiler and economiser design.

7. ATMOSPHERIC PRESSURE SCHEME

7.1 UNIT SIZE

The reasons for basing this study on previous more detailed studies have been set out in 6.1 above. These earlier studies therefore placed a limitation on the unit size on which this study is based. This consideration led to the choice of a 660 MW approximate gross unit size for the atmospheric pressure fluidized combustion scheme.

A unit size of 660 MW for non-nuclear plant is in line with current practice in the U.K. and in other countries. Larger units are in service in the United States. Proven technology is therefore available for units of this size for everything except the fluidized combustors. The combustors, being proposed on a more-or-less modular basis, do not present significantly different problems for 660 MW units compared with units of half that size or even less. In the present situation with the largest fluidized coal combustor in operation being of no more than a few MW capacity, there does not appear to be any reason to choose a unit smaller than 660 MW for a station of up to 2000 MW from this point of view.

The remaining, important factor in determining unit size is the stability of the electrical transmission system and its ability to withstand the sudden loss of the unit. This is outside the scope of the study but we understand that B.C. Hydro and Power Authority are currently contemplating the installation of 500 MW units and it seems reasonable to assume that if, as seems likely, fluidized combustion plant is not ordered for a few years, then the system will have grown sufficiently by that time to assimilate units of 660 MW capacity.

For the reasons described above, it was decided to base the study on atmospheric pressure fluidized combustion units of approximately 660 MW gross capacity. The final gross output adjusted to the site conditions was 648 MW. Three units giving 1944 MW gross were selected as the station capacity.

7.2 DESCRIPTION OF SCHEME

7.2.1 LAYOUT

The station comprises three nominal 660 MW (gross) coal-fired boiler/steam turbine generating units together with all associated equipment, buildings and civil works. A notional site plan is shown on drawing 15283-101-003. A plan and elevation of the boiler and turbine house plant arrangement are shown on drawings 15283-101-005 and -006 respectively.

7.2.2 CYCLE

The cycle proposed is a conventional reheat steam cycle; it is described in 6.2 above and shown on drawing 15283-101-001. The components of the cycle are described below. Particulars of the cycle are given in Table 1.

7.2.3 COAL HANDLING AND STORAGE

Coarse-crushed coal is received at the station boundary and directed by the coal handling system either to the coal storage pile or to the elevated bunkers in the boiler house. The bunkers have a capacity of about 5150 short tons per unit, equivalent to about 10 hours running at full load.

7.2.4 COAL PREPARATION AND FIRING

For each boiler unit, coal from the main boiler house bunkers is fed by gravity into three coal preparation units each comprising a proprietary coal dryer followed by crushing equipment. The coal is dried to facilitate pneumatic transport and is crushed to suit the design characteristics of the beds.

The coal dryers are currently envisaged to burn a small proportion of the feed coal as a heat source. Further consideration would be given to this during plant design and some other source of heat, such as the flue gases, might be used while the plant was on load if such a scheme proved satisfactory.

The moisture-laden air/gas from the three coal preparation units is ducted away to the main flue gas precipitator inlets via a small electrostatic precipitator which collects fine particles of coal that would otherwise be lost.

A system of conveyors and elevators transports the dried and crushed coal to a prepared coal bunker of about 200 tons capacity corresponding to about half an hour running at full load.

From the prepared coal bunker, conveyors transport the coal to three service bunkers of about 60 tons capacity each.

Coal from the service bunkers feeds through injectors into the pneumatic transport system and is conveyed by it through a branching pipe system to multiple coal inlets feeding coal upwards into the beds. The transportation air is supplied by motor driven compressors.

7.2.5 BOILER

The design of the atmospheric fluidized combustion boiler described below is now a few years old. However, Combustion Systems Limited consider that the basis of the design remains sound and, while it might now propose changes in detail in layout and components, the design is a fair representation of its present concept of a suitable, large, atmospheric boiler both as regards the size of the items and their overall relationship.

A sectional arrangement of the boiler is shown on drawing number 15283-101-009.

The boiler has three coal-burning fluidized combustion beds. It is designed for coal with a top size, as fired, of 3/16 inch and two of the beds (A and B) have a fluidizing velocity of 7.6 ft/sec which is consistent with the coal size chosen. The third bed (C), which is referred to as the "reheat" bed, since it contains the complete reheater, is used as a carbon burn-up cell by recycling fines to it from the other two beds in addition to the coal fed to it. The fluidizing velocity in the reheat bed is reduced to 4.6 ft/sec to avoid excessive carry-over of fines.

The proposed boiler has the three beds arranged side-by-side and while this leads to a boiler occupying a fairly large area, it minimizes the building height and avoids some engineering design difficulties that might be encountered with beds stacked one above another.

The beds are contained by membrane water walls of the conventional type widely used in the construction of combustion chambers for conventional boilers. Additional tubular heating surface is immersed in the beds. The water walls extend upwards to surround the complete combustion chambers and the convective heat transfer surface in the vertical passes above them.

A main object of the boiler design was to minimize site work by constructing the combustion chambers as a series of transportable modules for assembly on site without extensive use of skilled labour. This was achieved, and by using the same type of basic arrangement of containment for each bed, it proved possible for the boiler to be constructed from only two types of containment module, i.e. end sections and intermediate sections. Each section is 40 ft. long, 14 ft. wide and 15 ft. high. This is, in fact, too large to be transported by rail to Hat Creek but there would appear to be no problem in making each unit in two halves 40 ft. long by 7 ft. wide by 14 ft. 6 in. high (trimming 6 in. off the height) which is within the rail transport limits.

The modules are intended to be completely assembled in the factory with their tube nests fitted, the whole amounting to a shipping weight of about 60 tons.

On site, after construction of the air plenum chambers below the combustion chambers, the modules would be located side-by-side and joined by single-place junction welds between the water wall membranes.

Only minor departures have been necessary from conventional practice in the design of the pressure parts. The main difference is in the pitch of the tubes in the membrane walls. This has been increased, still keeping within design metal temperature limits, to avoid an excessive number of tubes in the bed containments which would otherwise result from their large plan area. Apart from this feature, the design of the tubing, insulation and casing, and support girthing are identical to conventional practice.

The length of the beds (112 ft. overall for Beds A and B) presents some problems of support for the horizontal membrane wall roof section, but satisfactory supporting steelwork arrangements have been devised.

Vanes at the exit from each combustion chamber deflect a proportion of the grit particles in the gases into a hopper for refiring in the bed. One surface of the hopper is not formed from membrane wall but is protected by refractory. This is the only refractory in the boiler, but its configuration is such that suitable support and provision for expansion can be made readily, and no undue maintenance problems should arise. Apart from this, all the containment surfaces exposed to flue gas at temperatures greater than 750°F are fully water cooled.

The proposed boiler has a single drum and is of the assisted circulation type with multiple wet-motor circulating pumps drawing water via downcomers from the drum and discharging to the various parallel circuits of the evaporation sections of the boiler. These sections are principally the membrane walls forming the containment but additional evaporative surface is immersed in bed A, the evaporator bed.

A comparatively high water circulation ratio of 7 is made necessary by the boiler configuration to maintain a minimum water velocity of 3.5 ft/sec in the horizontal tubes. This leads to the use of more circulating pumps than would be expected for a conventional boiler but does not present any special technical problem.

It is of interest to note that the evaporator bed tube nest passes approximately 50% of the total water flow, i.e. about as much as the total circulation in a conventional boiler. Although containing only 14% of the water circuit tube weight, it performs 70% of the heat transfer, which illustrates the improved rates of heat transfer possible in the fluidized bed.

The selection of the quantities of water passing through the various circuits has been made to keep the steam fraction in the risers from the bed tube nest below 20%.

The superheater is divided into primary and secondary sections. The primary superheater is formed by a part of the convection surface in the gas pass above beds A and B. The secondary superheater is immersed in bed B, the superheat bed.

The reheater is also divided into primary and secondary sections. The primary section is in the gas pass above bed C and the secondary section is immersed in bed C.

In order to realize the very high heat transfer rates in the superheater and reheater sections made possible by immersing tubes in the beds, it is necessary to accept some elevation of metal temperature. It is envisaged that metal temperatures up to 610°C in secondary superheater tubes and 620°C in the secondary reheater will be attained. The use of 12% chromium steel is proposed. Alternative austenitic materials could be used satisfactorily at some increase in cost. There is experimental evidence that erosion of immersed tubes will be insignificant.

The economiser is divided into low temperature and high temperature sections. The low temperature section is a single unit situated after the junction of the gas flows from the three beds. The high temperature section is divided into three parts situated in the gas pass above each bed. (See drawing 15283-101-009).

The fluidizing/combustion air is supplied by four motor driven forced draught fans operating in parallel. The air is heated by two regenerative air heaters and flows from them to plenum chambers beneath the beds. Each chamber is divided into four sections corresponding to sections of the bed divided off by division plates within the bed. The air supply to each section is controlled individually to assist in equalising air flows in the various sections and to allow sections to be shut off sequentially for part-load operation. The air enters the beds through the distribution plates at the base of the beds.

The flue gases from beds A and B, after leaving the convection banks at a temperature of about 750°F, enter high-efficiency grit collectors of the centrifugal type. The collected grit is refired in bed C. The gas from bed C enters a low efficiency dust collector from which the grits are not refired and which serves only to avoid an excessive dust content in the gases.

On leaving the dust collectors, the flue gases from the three beds enter a common duct and pass through the low temperature economiser and regenerative air heaters before entering the electrostatic precipitators.

The cleaned gases are ducted to a single three-flued stack serving the three boiler units.

7.2.6 ASH AND DUST HANDLING

The handling of "weir ash" from the fluidized beds in a power station in which the discharge of liquid effluents is prohibited presents difficulties.

The ash, which consists of soft unfused particles of up to the maximum coal feed size of 3/16 inch, is discharged continuously from multiple discharge points at the ends of the modules. The ash is at the bed temperature of about 850°C (1562°F).

Storage of the ash in hoppers integral with or adjacent to the boilers does not appear to be a very practical proposal. The ash discharge points are at the periphery of each bed and this, together with the general arrangement of the boiler, makes it difficult to provide hoppers beneath the beds. Hoppers might possibly be provided alongside the beds but the high ash content of coal would demand quite large hoppers for even a short period of storage and these would probably be detrimental to economic plant arrangement if they were to be positioned so that the ash could fall directly into them.

On the basis of this reasoning it was decided to provide for continuous ash removal from the vicinity of the boilers. It thus became necessary to select a suitable system to deal with this hot material.

A water sluicing system would be a convenient arrangement since it performs the dual functions of cooling and transporting the ash. However, it has the serious disadvantage for the Hat Creek site in that it requires large quantities of water and these become contaminated with dissolved and suspended solids from the ash. The water would be re-circulated and re-used but sooner or later it would become unfit for further use and would require extensive treatment to reduce the suspended solid content. Build-up of dissolved solids would also occur and this is an element in the general problem of dealing with solids build-up in water recirculated in a power station where liquid effluent is prohibited. Continuous treatment of a proportion of the ash sluicing water would be possible but this does not alter the overall problem in terms of the quantities of dissolved and suspended solids to be dealt with.

In view of these problems of effluent treatment, a system for water sluicing of the ash has been regarded only as a possible, but not preferred, solution and an alternative has been sought.

Pneumatic transport of the ash was considered but rejected on the grounds of high power consumption and no known economical system being available for clean-up of the very hot transportation air after use.

The ash handling organization of Babcock and Wilcox in London was approached with regard to the use of mechanical conveyors and the problem discussed with them. The possibility of cooling the ash and discharging it onto a conveyor belt was discussed but rejected for lack of a design of a suitable ash cooler. It did not appear possible to provide a water spray system that would be adequately controlled to avoid the discharge, at times, of dirty water which would add to effluent treatment problems.

Drag link conveyors were next discussed and while no designs were available to deal with ash at the bed temperature, there were systems incorporating water troughs which could be used. Such systems have been used for handling hot boiler ash in a number of installations on the continent of Europe.

It was decided to adopt this method and the proposed system is as follows.

Duplicate 100% duty drag link conveyors run in water troughs at each end of each boiler bed. The "weir ash" drops down refractory lined pipes into the water troughs where it is cooled and water evaporates. Make-up water is supplied via a level control valve to maintain trough water level. The troughs are enclosed and the vapour is vented through pipes for discharge to the stack. At the outlet end of the trough the ash is dragged by the links up a slope (perhaps 15 degrees) which is of sufficient length for most of the surplus water to drain back into the trough. The drag link conveyors discharge onto duplicate belt conveyors running the length of the boiler house. Drainage troughs installed below these conveyors direct any further water back to the drag link conveyor troughs.

The belt conveyors discharge into two elevated bunkers outside the boiler house, each bunker having sufficient capacity to hold about four hours worth of ash from the station when running at full load. The bunkers incorporate a drainage system to ensure that the ash leaving the station is dry enough to dump without water run-off problems.

The bunkers would be used alternately to allow adequate time for water to drain off. The drained water returns to the drag link conveyor troughs.

Grits from the reheater bed cyclones are discharged at a temperature of about 750°F and it is convenient to deal with these in the same ash handling system. A further pair of drag link conveyors are therefore included at each boiler for this purpose.

Disposal of ash from the bunkers could either be by conveyor or by wheeled transport.

The water in the drag link conveyor troughs will require treatment in the same way as that used in an ash sluicing system. However, the quantities involved are much smaller and the quality could probably be allowed to become considerably worse before replacement. The quantities would probably be small enough to be removed from the site by tanker.

A conventional pneumatic system is proposed for handling the dust collected in hoppers below the electrostatic precipitators. This system would be operated periodically to transfer the dust from the hoppers to elevated dust silos adjacent to the ash bunkers. The air used for transportation is cleaned up prior to discharge to the atmosphere, the final stage of clean-up utilising fabric bag filters.

A problem that might arise in handling the ash, grits and dust concerns the nature of the ash. It is possible that the addition of water to the refuse might result in hardening as the material dried. Some experimental work would be necessary to assess this problem.

7.2.7 TURBINE GENERATOR PLANT

A conventional 660 MW reheat steam turbine generator is used. This is a 3600 rev/min tandem compound unit with four exhaust flows. The exhaust blade length would be chosen to provide the required steam passing capacity and to provide the optimum balance between capital cost of the turbine condenser and cooling system, and the operating cost, to give the minimum power cost.

The condenser is of the surface type with tubes currently assumed to be of admiralty brass. This would be reconsidered during design to take into account the proposed quality of water circulating in the cooling system. If the concentration of dissolved solids in the cooling water was allowed to build up to unusually high levels to minimize the necessary blowdown, then careful consideration of the tube material would be required.

Motor driven extraction pumps pass condensate through low pressure surface heaters to an elevated direct-contact deaerating heater with storage tank.

Duplicate 50% duty variable speed motor driven boiler feed pumps draw water from the deaerator storage tank and discharge it to the boiler economiser inlet via high pressure feed heaters and boiler feed water regulating valves.

It might prove slightly more economical to provide a single 100% duty auxiliary steam turbine driven boiler feed pump with a 30% motor driven pump for starting. However, for simplicity, motor driven pumps only were included in the study.

7.2.8 WATER TREATMENT PLANT

A condensate polishing plant of sufficient capacity to treat about 10% of the full load condensate flow has been included for the condensate system of each turbine generator. This plant is intended to assist in control of feedwater quality and to expedite clean-up of the system on commissioning and after shut-downs.

Demineralising plant is included for treatment of river water to produce make-up water for the steam cycle. The capacity of the plant is about 2½% of the total steam generation.

By combining the polishing plant with the make-up demineralising plant, capital cost savings have been achieved and increased polishing capacity is available for an individual turbine generator if required.

The station steam cycle make-up water requirements during normal full load running have been assessed as 1½% of the total steam generation. This comprises 1% to replace boiler blowdown and ½% to replace miscellaneous losses.

7.2.9 ELECTRICAL EQUIPMENT

The generator is a 3-phase 60 Hz unit generating at the manufacturer's standard voltage for the rating. The rotor is hydrogen cooled and the stator is water and hydrogen cooled. Alternative cooling systems would probably be available from some manufacturers.

The station auxiliary power systems are conventional with isolated phase bus ducts connecting each generator directly to its main generator step-up transformer which feeds the 500 KV transmission system via a circuit breaker, and to its unit step-down transformer supplying station auxiliary power at 13.8 KV. A station start-up transformer supplies power to the 13.8 KV system from the 500 KV transmission system in the event of internal power being unavailable.

Further step-down transformers supply all but the largest auxiliary loads. A battery, with chargers and an inverter system provides power for control and for emergency shut-down of the station if all auxiliary supplies are lost.

A power supply for the remote river water pumping station is provided from the 13.8 KV system.

The estimated auxiliary power consumption at full load is given in Table 7.

7.2.10 COOLING WATER SYSTEM

The condenser cooling water system comprises wet cooling towers placed above ponds from which motor-driven cooling water pumps take their suction. The cooling water is piped to the turbine house, where it passes through the condenser tubes and is returned to the cooling towers for cooling and re-use.

Water lost from the system by evaporation from the cooling towers is made up by river water drawn from the holding pond and by treated liquid effluents from the remainder of the plant.

The concentration of dissolved solids in the cooling water would be maintained in balance at the desired level by blowdown of water from the system.

7.2.11 LIQUID EFFLUENT TREATMENT

The liquid effluent treatment systems have been designed to produce water of suitable quality for further use in the main cooling water system. The systems comprise: a biological sewage treatment plant; sumps with provision for dosing for pH control of effluents from water treatment plants, condensate polishing plants, boiler blow down and boiler chemical cleaning effluents; and separating tanks and equipment for removal of oil, dust and dirt from drain water collected both within and outside the buildings.

Dissolved solids in the river water used for make-up to the power station systems would eventually be discharged in the water blown down from the cooling towers.

If a water sluicing system was used for ash handling, an expensive installation to control the concentration of suspended solids in the re-circulated sluicing water would also be needed.

7.2.12 CIVIL WORKS

The oivil works follow conventional power station practice. The main difference is in the low boiler height that has to be accommodated in the boiler house.

The buildings have concrete foundations and structural steel frames with steel cladding. Intermediate floors are also steel.

The coal bunkers in the boiler house are steel. The external ash and dust bunkers are concrete.

7.2.13 START-UP, SHUT-DOWN AND CONTROL

To start the boiler up it is necessary to raise the temperature of the bed material to about 450°C (842°F) at which temperature coal will ignite when injected into the bed and combustion become self-sustaining. This pre-heating is effected by oil burners. A system of oil storage tanks, pumps, pipework and controls is included.

Shut-down is effected by "slumping" the bed i.e. shutting off the supply of fluidizing/combustion air (and the coal) and allowing the bed to settle on the air distributor plate. The insulating properties of the bed material serve to retain the heat in the bed and avoid overheating of the air plenum and other parts adjacent to the bed. The retention of heat also permits rapid restarting after a shut down of several hours or more, if this is required.

Emergency shut-down on loss of auxiliary electrical supplies is a situation that has not yet received a great deal of study and would require detailed consideration in the engineering design phase of a project. The problem lies in the loss of the boiler circulating pumps and the boiler feed pumps while a large quantity of heat remains in the bed.

Detailed study would be required of the rate of heat flow out of the bed and its absorption by the boiler. It is confidently expected that the heat flow leaving the bed will be so slow that no special features will be required to maintain the boiler feed supply or to assist boiler circulation. It is possible, however, that a feed and/or a boiler circulation pump should be arranged for drive by steam directly from the boiler in emergency.

Boiler control during normal operation is effected by means of fuel and air flow controls with the limitation that air flow cannot be varied too greatly because of the fluidization characteristics of the bed material. Because of this, the bed is compartmented so that the coal and air flows may be shut-off from a compartment to effect a corresponding load reduction.

7.3 CAPITAL COST

The capital cost of the plant was estimated and the total cost and its breakdown are given in Table 9.

All the principal costs were derived as appropriate from budget quotations given by equipment manufacturers, from previous studies updated by allowing for inflation or, in the case of civil works, from an estimate made by a quantity surveyor based on outline drawings and rates applicable to the Hat Creek site.

Cash flow and interest during construction were calculated in accordance with the base data for the study. Inflated cash flow was also calculated in accordance with the base data. The cash flows are shown in Table 11.

The distinction between the mechanical, electrical and civil costs is not exact and, in particular, a certain amount of mechanical work has necessarily been included in the civil works.

Costs in pounds sterling have been converted to Canadian dollars at the rate of 2.20 Canadian dollars per pound sterling.

Contingencies have been added to all plant costs in accordance with the base data i.e. at 10 per cent for budget estimates of well defined items and at 15 per cent for the remainder. It is considered, however, that a somewhat higher contingency would be more appropriate for the boiler plant.

For purposes of comparison with other studies, the station cost was also calculated on the basis of an alternative estimate of the total interest during construction of 26.6 per cent of the capital cost. The alternative costs are shown in Table 16.

7.4 POWER COST

The plant performance was estimated and is shown on the flow diagram (see drawing No. 15283-101-012). Corresponding material and heat balances are given in Tables 3 and 5.

The power costs at load factors of 60, 70 and 80 per cent were calculated and are given in Table 13. Account was taken of fuel cost, capital charges and the various operating costs identified in the base data.

The capital charges were based on an interest rate of 10 per cent per annum as specified in the terms of reference and the allowance of 0.369 per cent per annum for depreciation was arrived at on the basis of a sinking fund with the same interest rate of 10 per cent per annum and a plant life of 35 years.

Power costs based on a station cost including the alternative estimate of interest during construction referred to in 7.3 are shown in Table 18.

7.5 ENVIRONMENTAL IMPACT

The environmental impact of the station was studied from the point of view of solid liquid and gaseous effluents. Thermal and noise pollution were also briefly considered. Effluents are indicated on the flow diagram (drawing No. 15283-101-012).

a) Solid effluents. The major solid effluent from the station is the ash and dust remaining after combustion of the coal. When the station is running at full load, burning coal with the maximum anticipated ash content of 31%, the quantities produced daily are approximately 5500 short tons of ash and grits and 5525 short tons of dust. The ash and grits will probably be soft particles up to 3/16 inch. It is envisaged that both ash and dust will be removed from the site in a damp condition to minimize spillage and wind-blown nuisance. B.C. Hydro have stated that they do not envisage any problem in disposing of these wastes.

In the event that coal of a higher sulphur content than that studied is required to be burnt, it is envisaged that crushed limestone will be added to the fluidized combustion beds and will be discharged with the ash partly converted to calcium sulphate. The quantities would be about 1070 short tons per day when burning a 1% sulphur coal and 2300 short tons/per day when burning 2% sulphur coal.

Particles emission from the stack is dealt with below.

Blown coal dust might be a source of nuisance and it is therefore envisaged that the coal conveyors would be enclosed. It is assumed that the coal stockpile will not require protection.

Other solid wastes are considered to be of a minor nature and can readily be removed from the site for disposal if necessary. These wastes include:—

Domestic refuse.
Sewage sludge.
Worn-out parts and sweepings and other material usually collected in bins.
Water and effluent treatment plant solid wastes.

b) Liquid effluents. With a few exceptions, all liquid wastes are treated to render them suitable for further use in the main cooling water system.

Waste water from the submerged drag-link conveyors used for ash handling could be cleaned up for discharge to the main cooling water system or could be taken off site by tanker vehicle.

Waste lubricating oil might be burnt in the boilers or shipped off site.

c) Gaseous effluents. The flue gas temperature is 300°F and the gases are discharged at a height of 1000 ft. above ground level.

Approximately 50% of the ash in the coal is expected to reach the electrostatic precipitator inlets of the proposed boiler. Thus for coal with an ash content of 31%, the highest envisaged in the study, an ash quantity of:

$$2000 \times \frac{31}{100} \times \frac{50}{100} = 310 \text{ lb per short ton of coal reaches the precipitator inlet.}$$

Now to meet the British Columbia Pollution Control Board Level guideline of a maximum total particulates of 5lb per ton of coal, a precipitator efficiency of:

$$\frac{100 \times (310 - 5)}{310} = 98.4\% \text{ is necessary.}$$

The base data provided for the study by B.C. Hydro indicates that a maximum organic plus pyritic sulphur content of the coal of 0.38% can be expected and that the whole of this sulphur should be assumed burnt to sulphur dioxide.

Now 32lb of sulphur combines with 32lb of oxygen to produce 64lb of sulphur dioxide. Thus the sulphur dioxide discharged per short ton of coal is:

$$2000 \times \frac{0.38}{100} \times \frac{64}{32} = 15.2\text{lb}$$

This is well inside the Level A guideline of 20lb per ton of coal.

The sulphur dioxide quantity calculated above is considered pessimistic for a fluidized combustion boiler. The temperature level in the bed, and the turbulence, lead to satisfactory conditions for the sulphur to react with the calcium oxide present naturally in the coal ash and result in the flue gas sulphur dioxide content being reduced.

The use of coals of a higher sulphur content would be possible by addition of limestone or dolomite to the fluidized beds without exceeding any emission standards for sulphur dioxide.

Assuming that limestone of satisfactory reactivity is available, the quantities required to keep within the Level A guideline would be approximately 70 lb of limestone per ton of coal for a 1% sulphur coal and 150 lb per ton for a 2% sulphur coal. In view of the local availability of large quantities of limestone it is not proposed that spent limestone discharged from the bed should be regenerated. It also appears unlikely that sulphur recovery from the residue would be economic. In making this statement it is assumed that B.C. Hydro and Power Authority could make suitable arrangements to dispose of the residue.

Very little additional equipment would be necessary if limestone was to be used. It is envisaged that additional small limestone bunkers would be situated in the boiler house adjacent to the main coal bunkers. These bunkers would be fed from a stockpile in the yard, by the coal handling system, during an interval in coaling. The limestone would be fed from the bunkers directly into the coal preparation equipment where it would mix with the coal and be crushed to the same size as the coal. It would be injected into the bed with the coal by the pneumatic coal injection system.

Emission of oxides of nitrogen has been measured on fluidized combustion test rigs; for atmospheric pressure rigs emissions corresponding to 7 to 18 pounds per short ton of Hat Creek coal burnt have been found. These are within the British Columbia Level A guideline of 27 pounds per ton. The United States EPA level corresponds to about 13 pounds per ton and evidence from some of the larger experimental rigs indicates that it should be possible to keep within this limit when operating with about 3% of excess oxygen.

The discharge of trace elements from the fuel with the flue gases will occur due to vaporization of these substances. Some experimental work on this subject has been reported and it appears that the magnitude of this problem will be less than for pulverised coal fired boilers due to the lower combustion temperature (Reference 4).

d) Other environmental effects. The cooling towers will emit heat in the form of warm air and water vapour. The rate of heat discharge is approximately 2480 MW or 8460 Million Btu/h. The quantity of water vapour discharged is approximately 8460 klb/h. These figures refer to full load operation of the whole station.

No abnormal noise problem is anticipated and it has been assumed that no special sound insulation is needed.

7.6 CONSTRUCTION SCHEDULE

A proposed schedule is given on drawing 15283-101-014. The schedule is based on the assumption that this would not be the first of the type ordered and that the manufacturer therefore had a developed design available.

There does not appear to be any reason for the overall manufacture and erection programme for fluidized combustion boilers to be longer than conventional boilers. It is likely to be quicker due to the anticipated shorter site construction time.

The delivery periods for plant vary somewhat with demand and for simplicity it has been assumed that the deliveries obtained require orders to be placed at the time site preparation begins. An overall schedule of five years would be typical for the period from start of site preparation to full commercial operation of the first unit.

In determining the earliest in-service date, it would at present be necessary to assume that this was the first station of the type and consequently a longer schedule would be needed for the boiler, together with an engineering phase prior to the boiler maker starting his design. An engineering phase of one year is suggested prior to finalising details for boiler design. A boiler design phase of one year would commence 6 months after the start of the engineering phase. Thus a period of 18 months would be added at the start of the schedule for engineering and design. Assuming an additional year added to the period for boiler manufacture, erection and commissioning, the total schedule would be 7½ years giving an earliest in service date of April 1983 if work started immediately; refer to drawing 15283-101-015.

7.7 FEASIBILITY

Apart from the boiler, almost all the plant is conventional and there can be little doubt concerning its feasibility.

The boiler, in particular the fluid bed, its containment and the coal preparation and injection systems are untried in power station practice. The various parts of these are discussed below.

Large gas-fired fluidized beds have been in commercial use for a number of years for ore roasting and others are in service for burning waste products. Experimental rigs burning coal have also been in operation for over ten years. A small boiler (about 45,000 lb/hr steam production) has recently been commissioned and a 300,000 lb/hr boiler is due to be commissioned early in 1976. It is claimed that increasing the area of the bed in plan does not present any problems in maintaining an effective bed and that the only problems likely to arise would be in balancing coal and air flows throughout the bed.

Pneumatic transport of crushed material is established technology and if properly applied, should not give rise to difficulty with the coal injection into the bed.

The coal dryers would be proprietary units, the coal crushing equipment would employ a proven type of crusher. The coal dust separating cyclones would be of straight forward design. The coal dust electrostatic precipitator is unusual but no particular problems are envisaged.

The shape of the boiler membrane wall containment and its construction in modules is unusual. However the construction appears to be straightforward.

Possibly the area in which most doubt exists is in the control of the boiler during start-up, shut-down and load changing. Features are proposed to enable satisfactory control to be achieved but these remain to be proven in practice.

Consideration of all the above factors leads to the conclusion that the scheme is feasible, although it should be recognized that the process is untried for power generation on a large scale and the risks involved in the initial application of the technology are probably greater than those that would usually be taken in the provision of large capacity generating plant.

8. PRESSURIZED SCHEME

8.1 UNIT SIZE

Reasoning similar to that used for the atmospheric unit (see 7.1) was applied in determining the unit size for the pressurized unit, but with the additional criterion that the gas turbine units selected should be of existing design. This was considered desirable so that the power plant studied would be capable of being implemented without being dependent upon future gas turbine development.

In order to obtain the lowest cost per kilowatt, a gas turbine of the largest size currently available was selected. This unit size is about 70 MW. To facilitate the use of data from previous studies, the Stal-Laval GT 120 gas turbine generator unit was used as the basis for the study. Stal-Laval have confirmed that this unit is considered suitable for use with fluidized coal combustors as far as can be ascertained on the basis of present experimental work.

Having selected the gas turbine, the choice of unit size then depends principally upon the number of gas turbines per unit. For the type of combined cycle contemplated, two gas turbines resulted in a gross unit output of about 623 MW. This size was adopted for the study.

The number of units in the station could have been 4, giving a gross output of 2492 MW, but 3 units giving 1869 MW gross was selected to remain within the terms of reference for the study which refer to "a total installation up to 2000 MW".

8.2 DESCRIPTION OF SCHEME

8.2.1 LAYOUT

The station comprises three 623 MW (gross) generating units together with all associated equipment, buildings and civil works. A nominal site plan is shown on drawing 15283-101-004. A plan and elevation of the boiler and turbine house plant arrangement are shown on drawings 15283-101-007 and 008.

Each generating unit comprises four boiler modules associated with two gas turbine units and one steam turbine unit.

8.2.2 CYCLE

The cycle is shown on drawing 15283-101-002 and particulars of the cycle are given in Table 2. Gas turbine driven air compressors draw in air from outside the building through silencers and filters. On leaving the low pressure compressor the air is cooled in an intercooler firstly by condensate from the steam turbine condenser and secondly by river water to the main cooling water system. From the intercooler, the air is compressed in the high pressure compressor whence it enters a connection on the boiler module where it serves as fluidizing/combustion air.

Each boiler module contains four fluidized beds fired with crushed coal from the coal preparation and injection plant.

Hot flue gases leave the boiler module and pass through the gas cleaning equipment prior to expanding through the HP and LP gas turbines driving the HP and LP compressors.

On leaving the LP turbine, the gases expand through the separate power turbine, driving the generator. Exhaust gases reach the stack via high and low temperature economisers.

Steam generated and reheated in the heating surface in and around the fluid beds is used by the steam turbine generator. Exhaust steam is condensed in surface condensers. Approximately half the condensate is heated by a seven-stage feed heating train using steam extracted from the turbine. The remainder, after use in the gas turbine intercooler, is heated by the gas turbine exhaust gas in the low temperature economiser. At this point both condensate flows mix and pass through the high temperature economiser prior to being fed to the boiler modules.

8.2.3 COAL HANDLING AND STORAGE

The plant is the same as that described under 7.2.3 for the atmospheric unit except that the bunker capacity is reduced to about 4600 short tons per unit, but still giving about 10 hours running at full load.

8.2.4 COAL PREPARATION AND FIRING

The system of main coal bunkers, coal dryers, crushers, coal dust precipitation and conveyors and elevators feeding a prepared coal bunker of about 200 ton capacity is the same as that described in 7.2.4 for the atmospheric unit but with the crushed coal size adjusted to suit the characteristics of the particular pressurized unit fluid beds chosen.

Conveyors transport the coal from the prepared coal bunker to the pressurized coal feeding system by which it is pressurized and fed to the beds.

The pressurized coal feeding system for each 623 MW generating unit comprises sixteen identical units (one for each of the four beds in each of the four boiler modules). A single unit is shown on drawing 15283-101-011. It consists of a coal bin, storage injector, primary injector with a feeder outlet for each of the four injection nozzles in a bed. The storage vessel is intermittently filled from the coal bin and raised to the same pressure as the primary injector before coal is discharged by gravity into the primary injector. The primary injector contains four locally-fluidized off-takes which discharge coal into conveying lines to the boiler. Injection or conveying air is added to the conveying lines to avoid settling-out of particles.

Changes in coal feed rate are effected by —

(a) varying the pressure in the primary injector relative to that in the boiler (this is a relatively slow process) and

(b) varying the quantity of conveying air. This has an almost immediate effect on coal feed rate.

The whole filling and feeding sequence is automatically controlled.

8.2.5 BOILER

BOILER DESIGN

The boiler consists of four identical modules, one of which is shown diagrammatically in drawing 15283-101-010. The module is contained in a pressure shell some 12 ft. diameter and 100 ft. high. Each module consists of four fluidized beds, each providing a separate function:— pre-evaporator, two beds for the superheater and one for reheat. Water/steam flows by forced circulation through the system. After passing through the pre-evaporator bed, the water enters the evaporator tubes which form the "water" walls of the beds and extend over the full height of the module. Combustion air (from the gas turbine compressors) enters near the base of the pressure shell and flows between the water walls and the pressure shell. The amount of air entering each bed is controlled by dampers situated below the individual distributor plates. The hot gases from each bed are collected in a common duct and leave the pressure shell to enter the dust separation unit.

Each bed has an area of 8 ft. by 8 ft. and a depth of up to 12 ft. — sufficient to accommodate the required heat transfer surface. The fluidizing velocity is approximately 5 ft/s. The combination of moderate fluidizing velocity, deep bed and 25% excess air results in a combustion efficiency greater than 99% so that a carbon burn-up cell is unnecessary.

The parameters quoted above are based upon CSL experience which includes operation of their pilot-plant at fluidizing velocities of about 3 ft/s with bed depths of up to 5 ft. i.e. similar gas residence times to the proposed combined cycle plant. The next phase of the experimental programme will investigate velocities up to 10 ft/s and bed depths up to 8 ft.

GAS CLEANING

The bulk of the solids elutriated from the bed must be removed in order to protect the gas turbine blading from erosion and deposition. The whole subject of gas cleaning is now under intensive development throughout the world with major advances being made in the area of filter beds. It is likely that much improved performance will be available by the time any large plant is built. For the present purpose two stages of cyclone-type dust collectors are used. The first stage consists of four (per module) large cyclones of conventional type and the second stage consists of a new type of cyclone — "Aerodyne" — which has an improved performance compared with the conventional type. All the cyclones are contained in a pressure vessel 12 ft. diameter and 30 ft. long.

The performance of the gas cleaning equipment cannot be predicted accurately until more is known about the friability of the ash of the Hat Creek coal. It is expected, however, that the particulate emissions would be less than 3 lb/ton of coal (i.e. well within the pollution limits) with a maximum particle size of about 5 microns.

8.2.6 ASH AND DUST HANDLING

The problems of handling hot ash without producing large liquid effluents which are described in 7.2.6 for the atmospheric units are also encountered on the pressurized units. In respect of temperature of the weir ash the situation is, however, a little easier since the ash is cooled within the boiler module before being discharged. It is estimated that the ash will leave the modules at about 390°F.

Dust from the flue gas cleaning equipment is discharged continuously and its temperature will be approximately 525°F.

Approximately equal quantities of weir ash and flue dust will be discharged.

Due to the large quantities, it is not proposed that ash and dust hoppers should be provided in the boiler house and a continuous ash and dust removal system is envisaged.

A similar analysis of the problem to that made for the atmospheric units and presented in 7.2.6 led to the same solution and the proposed system has a pair of 100% duty drag link conveyors running in water troughs and transporting the ash and dust from one boiler module to a pair of 100% duty belt conveyors. The remaining features of the system are the same as those proposed for the atmospheric units.

8.2.7 TURBINE GENERATOR PLANT

GAS TURBINE GENERATOR

The design of the cycle is based upon the use of two Stal-Laval GT 120 gas turbines. These are industrial gas turbines which normally operate on distillate oil or natural gas. Some 20 sets are now in operation in various parts of the world. In its standard form a GT 120 is rated at 70 MW when operating at sea level with an ambient temperature of 41°F and a turbine inlet temperature of 1470°F. For the Hat Creek application at 3000 ft. above sea level, with an ambient temperature of 38°F, turbine inlet temperature of 1470°F and 25% excess air, the output has been estimated to be 73.8 MW.

The GT 120 is normally fired by separate distillate oil-fired combustors. For the proposed application these combustors would be replaced by the fluid -bed boilers. Thus the modifications necessary to the standard unit would be minimal. However, it is recommended that two standard oil-fired combustors be purchased so that both gas turbines can be tested following site installation, or for subsequent emergency use.

The proposed turbine is arranged in two lines with the compressors and their driving turbines co-axial and two power turbines, back to back, driving the generator in a separate line parallel to the compressors.

STEAM TURBINE GENERATOR

With a few exceptions, the information given in 7.2.7 concerning the atmospheric unit also applies to the pressurized unit. The changes are as follows:

The unit has a generator output of 476 MW.

Approximately 50% of the condensate passes through the gas turbine intercooler and low temperature economiser instead of through the feed heating plant.

The boiler feed water pumping plant comprises four pumping units:—

A steam turbine driven unit for 100% of the feed heating plant flow.

A motor driven starting unit for 30% of the feed heating plant flow.

A steam turbine driven unit for 100% of the intercooler and LT economiser flow.

A motor driven starting unit for 30% of the intercooler and LT economiser flow.

This multiplicity of feed pumps is not an arrangement that is favoured; it arises from the limitations of this study and the need to make use of earlier work. It is anticipated that a re-assessment of the cycle in the engineering design phase of a project would result in a simpler feed pumping system.

8.2.8 WATER TREATMENT PLANT

The boilers, being entirely of the once-through type, operate without blowdown for control of water purity; the purity of the feedwater is therefore vital. A typical recommendation is that the dissolved solids content should not exceed 0.1 ppm.

In view of this requirement, a 100% duty condensate polishing plant is included in the condensate system of each turbine generator.

Demineralising plant is included to produce make-up water for the steam cycle from river water. The capacity of the plant is about 2½% of the total steam generation.

The station steam cycle make-up water requirements during normal full load running have been assessed as ½% of the total steam generation. As there is no boiler blowdown, the whole quantity is accounted for by miscellaneous losses.

8.2.9 ELECTRICAL EQUIPMENT

The electrical equipment for the pressurized units is generally the same as that for the atmospheric units as described in 7.2.9.

The generators of the gas turbine units have their own step-up transformers for connection to the 500 KV system.

8.2.10 COOLING WATER SYSTEM

The cooling water system is as described under 7.2.10 for the atmospheric unit.

In addition to use of cooling water by the steam turbine generator condensers, cooling system make-up water is used for the second stage of cooling in the gas turbine intercoolers.

8.2.11 LIQUID EFFLUENT TREATMENT

The systems included for the atmospheric units, as described in 7.2.11, also apply to the pressurized units with the exception of the boiler blowdown which is not applicable to the once-through boilers.

8.2.12 CIVIL WORKS

The civil works for the pressurized scheme differ from those for the atmospheric scheme in the main building arrangement. Separate gas turbine and steam turbine houses are included on either side of the boiler house. Apart from this difference, the general construction features are the same.

Climatic conditions at the site probably preclude the use of an outdoor installation although this might be used elsewhere. The proposed design has buildings for all plant.

8.2.13 START-UP, SHUT-DOWN AND CONTROL

A variable speed electric starting motor of about 1000 KW capacity is used to drive the HP compressor motor up to around 2000 rpm, at which speed the HP compressor provides sufficient airflow, at the appropriate pressure, to fluidize the cold bed. At this point, normal procedure would be to burn gas in the bed. But since gas is unlikely to be available at the Hat Creek site, oil burners firing onto the bed surface are used — one oil burner per bed.

As the gas temperature rises, the HP rotor speed increases, thus increasing the air flow through the system and raising the cycle pressure. Eventually self-sustaining speed is reached when the starting motor is automatically disengaged.

When the bed temperature reaches c. 850°F, coal is fed into the bed, gradually replacing the oil fuel.

Load changing is too complex a subject to be studied in detail within the present remit. However, the general principles will be similar to other applications which have been studied. A demand for load reduction is met by a reduction in coal feed rate. This reduces the bed temperature and hence reduces both steam and gas turbine output. The fall in gas turbine load causes a reduction in both air flow and pressure — thus maintaining the fluidizing velocity approximately constant. This proceeds until the bed temperature is c. 1400°F, the minimum at which combustion is satisfactory. Further reduction in load is accomplished by by-passing air directly from the compressor to the turbine. In this way, load reductions to about 50% can be accomplished. Further reductions in load are then carried out by shutting down one gas turbine with its boiler modules.

The control of a combined cycle is an area which is still under development. One aspect which is peculiar to a supercharged cycle, however, is the need to protect the power gas turbine in the event of an electrical trip-out, bearing in mind the huge amount of stored energy in the boiler modules. In such an event, this energy has to be vented to atmosphere rapidly. It is worth noting, however, that the Stal-Laval unit is particularly suited to this eventuality, since venting can be carried out between the compressor turbines and the power turbines, where the gas temperature is relatively low.

8.3 CAPITAL COST

The capital costs and other financial data were derived in the same manner as described for the atmospheric scheme in 7.3. A breakdown of the capital costs is given in Table 10 and cash flows in Table 12.

Due to the earlier stage of development, the costs for this scheme should only be regarded as indicative.

As noted in 7.3 we consider that it would have been preferable to allow a contingency rather greater than the 15 per cent stipulated in the base data for the boiler modules and their gas cleaning modules.

For purposes of comparison with other studies, the station cost was also calculated on the basis of an alternative estimate of the total interest during construction of 21.0 per cent of the capital cost. The alternative costs are shown in Table 17.

8.4 POWER COST

The performance was estimated and is shown on the flow diagram (drawing No. 15283-101-013). Corresponding materials and heat balances are given in Tables 4 and 6.

The power cost was estimated as described in 7.4 for the atmospheric scheme and is given in Table 14.

Power costs based on a station cost including the alternative estimate of interest during construction referred to in 8.3 are shown in Table 19.

8.5 ENVIRONMENTAL IMPACT

The environmental impact of the station is generally the same as described in 7.5 for the atmospheric units. The differences are described below.

The maximum limit of particulate matter in the flue gas, to which the gas cleaning equipment must conform to avoid deterioration of the gas turbine, implies a stack emission well within the Level A Guideline of 5 lb. per ton of coal. Electrostatic precipitators are not, therefore, required.

Regarding emission of sulphur oxides,the predicted level is the same as for the atmospheric units. For higher sulphur coals, although dolomite is the preferred additive for pressurized combustion, limestone can also be used satisfactorily. It would be injected with the coal in the same manner as for the atmospheric units.

Measurement of nitrogen oxides in the gases from experimental pressurized fluid combustors has indicated that very low levels can be expected. An emission of 2.6 lb. per ton of coal is predicted, compared with the Level A Guideline of 27 lb. per ton.

Heat and water vapour emission from the cooling towers is marginally less than the atmospheric units.

Noise from the gas turbine air intakes, if unsilenced, would be objectionable and silencers are therefore included.

8.6 CONSTRUCTION SCHEDULE

In general terms, the schedule described in 7.6 for the atmospheric units would be applicable. Some saving might be made on the site work for boiler erection but the larger number of pieces of equipment that it would be necessary to install would probably result in the overall schedule being generally unchanged.

It is the opinion of Combustion Systems Limited that a further 5 years is required for the development of pressurized fluidized combustion combined cycle plant to the point at which a large-scale commercial plant could be ordered. This, together with the 7½ year schedule indicated in 7.6 for the first station of the type, implies an earliest in-service date of 1988. See drawing 15283-101-016.

8.7 FEASIBILITY

Important areas of the plant would be novel, in particular the fluidized combustion boiler modules.

The boiler modules employ very high rates of heat release. This suggests that careful engineering design will be necessary to contain and absorb the heat without difficulties. Sophisticated controls will also be necessary.

The coal preparation equipment is the same as that dealt with in 7.2.4 for the atmospheric units and was a combination of more-or-less proven equipment. The coal pressurizing feeders are based on an established design. The pneumatic coal transport system is based on established technology.

The flue gas cleaning equipment employs a new type of cyclone and may also incorporate a newly developed filter bed. In view of these features and of the use of a pressurized arrangement, the flue gas cleaning modules can be regarded as novel. The feasibility of the station is dependent upon the gas cleaning equipment being effective because, if it is not, severe problems with the gas turbine can be expected.

The gas turbines are of proven design for distillate oil or natural gas burning. Their use with fluidized coal combustors is novel.

The remainder of the plant is conventional and well-proven.

It is considered that the feasibility of this scheme is not yet assured. The continuation of development over the next few years, including pilot or demonstration plants, should enable a much better judgement of feasibility to be made later.

9. COMPARISON OF SCHEMES

A comparison of the main technical features of the two schemes studied is shown in Table 15.

The small difference in unit size and station output is due to the need to adopt, where possible, existing technology in relation to standard size equipment. This particularly relates to gas turbine plant for the pressurized scheme.

Auxiliary power consumption is higher for the atmospheric scheme, due mainly to the use of forced draught fans, electric feed pumps and boiler circulating pumps. Electric power consumption could be reduced by the adoption of turbine driven feed pumps.

A comparison of plant costs indicates an advantage for the pressurized scheme both in capital cost and power cost per unit output.

Environmental considerations are similar for both schemes showing a slight advantage for the pressurized scheme in relation to particulate and NOx emissions from the stack, although particulate emission from the atmospheric units could be reduced by use of more efficient electrostatic precipitators.

Water vapour from the cooling tower is of the same order for both schemes.

The atmospheric boiler scheme could be constructed about 5 years earlier than the pressurized scheme which requires considerable detailed development work.

10. PILOT/DEMONSTRATION PLANTS

Any commitment to the construction of a large power station at Hat Creek employing fluidized combustion technology would be justified more easily if a pilot/demonstration plant were in operation or, at least, in an advanced stage.

If B.C. Hydro and Power Authority decided to proceed with such a plant themselves, various alternatives might be considered.

For an atmospheric pressure unit, a simple boiler — turbine unit with an output of 120 MW might be considered. The boiler could be of similar modular construction to the 660 MW unit studied. An indicative price for a fluidized combustion boiler only of this size would be thirteen million Canadian dollars excluding the manufacturer's development costs.

A simple pressurized unit could consist of a pressurized fluidized combustion hot gas generator supplying a gas turbine in an open cycle. An indicative price for a complete unit of about 70 MW installed as an additional unit at an existing power station site would be about twenty-eight million Canadian dollars including the manufacturers development costs.

A further alternative for a pressurized unit could consist of a combined cycle unit of about 300 MW. This would be a half-size version of the 623 MW unit studied, and would comprise two boiler modules, one gas turbine generator and one steam turbine generator. An indicative plant cost excluding B.C. Hydro and Power Authority overheads and interest during construction would be one hundred and twenty-five million dollars.

11. SIMILAR PROCESSES

The terms of reference require similar processes which are the subject of a major development effort to be studied.

It is assumed that "similar processes" alludes to other processes for the direct combustion of coal. In this sense, it was not possible to identify any similar process which is currently the subject of a major development effort.

With regard to fluidized combustion itself, a great deal of the major development is work which Combustion Systems Limited have performed or with which they are associated. The principal work in which they are not concerned is the atmospheric boiler development being undertaken by Pope, Evans and Robbins Incorporated and Foster Wheeler Corporation in the United States.

The Ignifluid process is specifically referred to in the terms of reference and this was studied. It emerged, however, that this process is not currently the subject of a major development effort, although some work is in hand to develop its use for coal gasification.

Another aspect of the subject of similar processes is the improvement of conventional processes of direct combustion. It was considered relevant to study the improvement of pulverised coal firing.

The information obtained is given below.

11.1 POPE, EVANS AND ROBBINS/FOSTER WHEELER

These organisations are constructing a 300,000 lb/h atmospheric fluidized combustion boiler at Rivesville power station, West Virginia, U.S.A. (Reference 9). It is currently expected to go into service in July, 1976.

As this development covers equipment comparable to the Combustion Systems Limited development described in this report, the study of the work of these companies was not pursued.

11.2 FIVES — CAIL BABCOCK

This company have installed boilers using the Ignifluid process for the direct combustion of coal and are continuing development of this process. A visit was made to Paris to discuss the process with Fives — Cail Babcock.

11.2.1 THE IGNIFLUID PROCESS

Coal, crushed to a maximum size of about one inch, is injected into a fluidized bed.

The Ignigluid bed is much more a bed of burning coal in a reducing atmosphere than a bed of ash with a small proportion of burning coal and sufficient excess air for complete combustion as proposed by CSL. Combustion in the Ignifluid process is completed by injection of secondary air above the bed.

The Ignifluid bed is designed to operate at a higher temperature than the CSL bed and therefore does not contain immersed heat transfer surface to diminish the temperature.

The bed temperature for the Ignifluid process is chosen, on the basis of the coal ash characteristics, to ensure that the ash particles sinter and form clinkers which descend, by gravity, to the bottom of the bed and are removed by a narrow inclined travelling grate. This generally results in a bed temperature of about 1200°C.

By contrast, the ash in the fluidized bed of CSL does not form clinkers and remains in the bed which is usually at a temperature of about 850°C. The ash particles remain small and constitute the vast majority of the bed material.

The construction of an Ignifluid boiler appears fairly conventional with an Ignifluid combustor at the bottom of a water-cooled furnace and conventional gas passes. Flue gas cleaning equipment, including electrostatic precipitators, removes grit and dust from the gases. The whole quantity collected is reinjected into the bed and thus all ash eventually leaves the bed over the back of the grate in the form of clinker.

The environmental impact of Ignifluid boilers is not greatly different from pulverized coal boilers. Particulate emission can be controlled to desired levels by use of electrostatic precipitators. Experiments have been conducted on the addition of limestone to the bed for sulphur dioxide control but the results are now viewed with caution by Fives — Cail Babcock and no claim is made for substantial sulphur dioxide control. Quite good results for the reduction of emission of oxides of nitrogen in comparison with conventional boilers have been obtained but the levels obtained appear to be similar to those that can be achieved by pulverized coal boilers designed for reduced emission. The whole of the boiler solid refuse appears as coarse ash and in some circumstances this more readily saleable material might be advantageous unless a market for the dust produced by pulverized coal boilers is available.

11.2.2 DEVELOPMENT

Fives — Cail Babcock has manufactured 24 Ignifluid furnaces up to now and some others have been installed by their licensees.

The two largest in service are of approximately 250,000 lb/h capacity. These have been in service since 1968 and together supply a 60 MW steam turbine generator. The plant is in Morocco.

The largest boiler on order is a unit of approximately 350,000 lb/h. This is due to go in service in Vietnam in 1977.

Fives — Cail Babcock do not appear currently to be developing the process for larger capacity boilers although they have previously tendered reheat units of 150 MW (e) capacity each. They remain willing to take an order immediately for a boiler of 100 MW and perhaps up to about 150 MW. They do not see difficulty in taking such an order on a purely commercial basis provided the coal characteristics or other requirements are not unusual.

Beyond this stage, the pattern of development is less clear. Fives — Cail Babcock are confident that the Ignifluid process has the potential to be developed to boiler sizes of 500 MW (e) and greater. They appear to consider however, that any such development should be preceded by operation of a boiler of 100 to 150 MW (e).

On the basis of the above information, it is likely to be at least 10 years before a 2000 MW power station using the Ignifluid process could be in service unless a multiplicity of small units were installed.

Development of the Ignifluid process is at present directed towards coal gasification.

11.2.3 COAL CHARACTERISTICS

Most Ignifluid furnaces have been used for burning anthracite, often of high ash content. However, bituminous coals are used on some installations. Sub-bituminous coals have not been burnt commercially but tests have not indicated any difficulty.

Fives — Cail Babcock initial reaction to the available information concerning Hat Creek coal was that it would not be difficult to burn in an Ignifluid furnace. An important consideration however, is the ash characteristics and if B.C. Hydro decide to continue consideration of the Ignifluid process, it is recommended that a two kilogram representative coal sample should be sent to Fives — Cail Babcock for testing.

11.2.4 ADVANTAGES AND DISADVANTAGES

The advantages claimed for the Ignifluid process are: low capital cost, low auxiliary power requirements, flexibility with regard to coal quality and ability to burn poor quality anthracite without use of oil.

The disadvantages of the process for a 2000 MW power station might be: the need to develop larger grates, maintenance of grates and some refractory is required (however, grate maintenance costs have not been high) and efficiency may be 1% less than a pulverized fuel boiler due to higher carbon in ash loss.

11.3 COMBUSTION ENGINEERING

The development by Combustion Engineering of improved processes for the direct combustion of coal was discussed with Mr. D.K. Whish of their Montreal office and was also studied by reference to a number of technical papers published by Combustion Engineering (see references below).

The information released by Combustion Engineering indicates that their development work is concentrated upon the improvement of conventional combustion systems rather than the development of any new system.

Their main development effort appears to have been in the treatment of flue gases for sulphur dioxide removal (References 10, 11 and 12). This does not appear to be relevant to the Hat Creek coal which is of low sulphur content.

Other work has been concerned with reducing the formation of oxides of nitrogen in the combustion of coal in pulverized fuel boilers. (Reference 14). No schedule for this work was submitted and it appears to be a continual process of measurement on operating units and the incorporation of design modifications on new units. Increased understanding of the mechanism of the formation of oxides of nitrogen has been gained and has permitted changes in design and operation which have resulted in reduction of emission. The reductions have been achieved principally by lowering the combustion temperature by admitting a larger proportion of the combustion air as overfire air with consequently less air entering with the coal. Combustion Engineering claim that such measures used in their tangentially-fired furnaces enable them to design units to comply with EPA emission standards.

Another aspect of Combustion Engineering development is in the field of coal-pulverizers with the object of developing reliable and economical designs of sufficient capacity to handle the coal quantities required for large boilers burning low — Btu coal. (Reference 15).

The developments described do not lead to a new system for the direct combustion of coal but continue the development of pulverized coal firing. In the areas of emission of oxides of nitrogen and of pulverizer development, these are relevant to the Hat Creek coal deposit if conventional pulverized coal boilers are used.

11.4 CLARKE CHAPMAN LTD.

This Company have stated that they are making a development effort with the fluidized combustion of coal but that they are regretfully unable to release any information at the present time.

They also state that they are not developing any new processes for the combustion of coal other than fluidized combustion.

11.5 OTHER MANUFACTURERS

Some other manufacturers were approached to enquire if they were able to release any information concerning their development efforts but no results were obtained. Those approached included the Babcock Organization both in the U.K. and North America and the Riley Stoker Corporation, U.S.A.

12. CONCLUDING REMARKS

12.1 RESULTS OF STUDY

The study was completed in accordance with the terms of reference. The following paragraphs indicate the work done.

The study has determined the capital cost of thermal power stations using Hat Creek coal by the fluidized combustion process in its atmospheric and pressurized forms. These costs and their breakdown are given for the two schemes in Tables 9 and 10.

Comments on the feasibility of the schemes are given in 7.7 and 8.7.

Materials and energy balances for the two schemes have been calculated and are given on drawing Nos. 15283-101-012 and 15283-101-013. The balances are set out in Tables 3, 4, 5 and 6.

Unit sizes for the schemes were determined and comments on the choice are given in 7.1 and 8.1.

Consideration was given to identifying similar processes which are the subject of a major development effort. Details are given in Section 11.

The environmental impact of each scheme was studied and is described in 7.5 and 8.5

Due to lack of data concerning the properties of East Kootenay coal, no study could be made of changes to the schemes that this might require.

In performing the study, part of the technical and cost data was provided by Combustion Systems Limited.

Power cost estimates were prepared for the schemes and are given in Tables 13 and 14.

Project schedules were prepared for the schemes and are presented on drawings 15283-101-014, 15283-101-015 and 15283-101-016.

For the purposes of comparison with other studies, capital and power costs were also calculated on the basis of alternative estimates of interest during construction. These results are presented in Tables 16, 17, 18 and 19.

12.2 FUTURE ACTION

If the British Columbia Hydro and Power Authority decide to proceed further with the consideration of fluidized combustion for the Hat Creek coal, it is recommended that an early step should be the initiation of tests on the coal in the experimental fluidized combustors of Combustion Systems Limited.

Further action would depend upon consideration of the information in this report concerning fluidized combustion applications, in relation to the electricity generation requirements of B.C. Hydro and Power Authority.

If it was decided to proceed with a prototype/demonstration plant, considerable study and negotiation with potential contractors and collaborators would be necessary. Such work would need to be started immediately if the data obtained was to be of value in engineering later commercial units if these were required without undue delay.

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TABLE 1
CYCLE PARTICULARS ATMOSPHERIC SCHEME

STATION

Station output — gross	MW	1944
— net	MW	1780
Number of units		3
Number of boilers per unit		1
Number of steam turbine generators per unit		1

BOILER

Rated output	klb/h	4221
Type	Atmospheric pressure fluidized combustion assisted circulation reheat.	
Bed Material		coal ash
Drum pressure	psia	2700
Superheater outlet — pressure	psia	2415
— temperature	°F	1055
Feed water inlet temperature	°F	490
Reheater inlet — pressure	psia	613
— temperature	°F	683
Reheater outlet — pressure	psia	583
— temperature	°F	1051
Reheater steam flow	klb/h	3780
Number of circulating pumps		5
Number of fluidized combustion beds		3
Number of sections per bed		4
Approximate depth of beds	ft	3
Coal injection system type		Pneumatic
Forced draught fans — number per boiler		4
— type of driver		Electric Motor
— head	in.W.G.	42
Primary grit and dust collectors		Cyclonic type
Secondary grit and dust collectors		Electrostatic precipitators
Combustion air heaters — type		Rotary regenerative
— air outlet temperature	°F	500
Exit flue gas temperature	°F	300
Approximate heating surfaces:		
— economiser	sq. ft	170000
— evaporation	sq.ft	193000
— superheater	sq.ft	108500
— reheater	sq.ft	38600

STEAM TURBINE GENERATOR

Generator gross electrical output	MW	648
Speed	rev/min	3600
Turbine configuration		TC4F
Number of cylinders		4

Steam stop valve — pressure	psia	2315
— temperature	°F	1050
Reheated steam — pressure	psia	575
— temperature	°F	1050
Number of steam extractions		7
Nominal exhaust pressure	in.Hg abs	2.5

CONDENSER

Number of shells		2
Number of water flows		1
Cooling water flow	USgpm	255000
Nominal cooling water inlet temperature	°F	70

FEED HEATING PLANT

Number of stages of feed heating		7
Type of heaters		6 surface 1 direct - contact deaerating
Final feedwater temperature	°F	490

BOILER FEED PUMPS

Number of pumps per unit		2 x 50%
Pump rated outlet	klb/h	2440
Discharge pressure	psia	3000
Type of driver		Electric motor

TABLE 2
CYCLE PARTICULARS PRESSURIZED SCHEME

STATION

Station output — gross	MW	1870
— net	MW	1821
Number of units		3
Number of boiler modules per unit		4
Number of gas turbine generators per unit		2
Number of steam turbine generators per unit		1

BOILER (All data is per boiler of 4 modules)

Rated output	klb/h	2884
Type		Pressurized, fluidized combustion once-through reheat
Superheater outlet — pressure	psia	2415
— temperature	°F	1055
Feed water inlet temperature	°F	540

Reheater inlet	— pressure	psia	560
	— temperature	°F	678
Reheater outlet	— pressure	psia	530
	— temperature	°F	1051
Reheater steam flow		klb/h	2654
Number of fluidized combustion beds			4 x 4
Number of sections per bed			1
Approximate depth of beds		ft	11
Coal injection system type			Pneumatic
Grit and dust collection			Two-stage cyclonic or better
Air inlet temperature to modules		°F	365
Gas exit temperature to stack		°F	300
Pressure within boiler casing		psig	182
Approximate heating surfaces			
— economiser (in modules)		sq.ft.	13,000
— evaporation		sq.ft.	18,000
— superheater		sq. ft.	28,000
— reheater		sq.ft.	13,000

GAS TURBINE GENERATOR (All data is per gas turbine generator)

Number of lines	2
Composition of first line	LP compressor driven by LP turbine and HP compressor driven by HP turbine.
Composition of second line	Double flow power turbine driving generator

TABLE 3
ATMOSPHERIC SCHEME

Materials Balances in klb/h
Totals for 3 Units at Full Load
(Figures are given to nearest 0.1klb/h for balancing purposes only. Expected accuracy of individual figures is generally better than ± 1%)

COAL/AIR/GAS/ASH ETC.

Coal from bunkers	2856.0
Air into F.D. fans	16792.8
	19648.8
Gas to stack	18900.6
Dust to stack	6.0
Ash from beds	163.8
Grits from cyclones	207.0
Dust from precipitators	371.4
	19648.8

STEAM/WATER IN STEAM CYCLE

Superheated steam to turbine generator	12600.0
Miscellaneous steam losses	63.0
Reheated steam to turbine generator	11340.0
Boiler blowdown	126.0
	24129.0
Steam returned as feed water	12600.0
Make-up	189.0
Steam to reheater	11340.0
	24129.0

COOLING SYSTEM WATER

To condensers and auxiliaries	402828.0
Evaporation from cooling towers	8459.4
Blowdown	1827.9
	413115.3
From condensers	382686.0
From auxiliary coolers	20142.0
Make-up	10021.5
Treated effluents	265.8
	413115.3

RIVER WATER

From river	10397.7
To cooling towers	10021.5
To steam cycle make-up	189.0
To ash cooling	47.4
To domestic use	139.8
	10397.7

TABLE 4
PRESSURIZED SCHEME

Materials Balance in klb/h
Totals for 3 Units at Full Load
(Figures are given to nearest o.1 klb/h for balancing purposes only. Expected accuracy of individual figures is generally better than ± 1%)

COAL/AIR/GAS/ASH ETC.

Coal from bunkers	2687.7
Air to Compressors	15852.3
Air to Dryers	600.9
	19140.9
Gas to Stack	17161.2
Ash from Beds	322.8
Grits from Cyclones	330.6
Gas from Dryer	1116.9
Gas Turbine losses	209.4
	19140.9

STEAM/WATER IN STEAM CYCLE

Superheated steam to turbine generator	8652.9
Misc Steam losses	43.2
Reheated steam to turbine generator	7961.1
	16657.2
Steam returned as feed water	8652.9
Make-up	43.2
Steam to reheaters	7961.1
	16657.2

COOLING SYSTEM WATER

To condensers & auxiliaries	341506.5
Evaporation from cooling towers	7376.4
Blowdown	1477.5
	350360.4
From condensers & auxiliaries	341506.5
From intercoolers	8714.1
Effluents	139.8
	350360.4

RIVER WATER

From river	8939.4
To steam cycle make-up	43.2
To domestic use	139.8
To intercoolers	8714.1
To ash cooling	42.3
	8939.4

TABLE 5
ATMOSPHERIC SCHEME

Heat Balance — Totals for
3 Units at Full Load
(All figures are MBtu/h)

Heat in coal from bunkers		18283.8
Boiler losses	— gas	2450.1
	— carbon in ash	438.9
	— sensible heat in ash	47.4
	— radiation and unaccounted	71.4
Heat losses	— boiler blowdown	95.4
	— miscellaneous steam losses	94.2
	— from pipes etc.	8.7
Heat to cooling system		8459.4
Heat converted to electricity		6633.0
Less heat in make-up		14.7
		18283.8

TABLE 6
PRESSURIZED SCHEME

Heat Balance — Total fro
3 Units at Full Load
(All figures are MBtu/h)

Heat in coal from bunkers		17206.8
Heat in air to gas turbine above 32°F		96.0
		17302.8
Heat losses	— Coal dryer	747.9
	— Stack	2091.6
	— Ash & Grits	220.5
	— Boiler radiation etc.	138.3
	— Misc. steam losses	64.65
	— Pipe line losses	6.3
	— Gas turbine misc.	18.0
	— Cooling system	7635.75
Heat converted to electricity (steam)		4869.9
Heat converted to electricity (gas)		1509.9
		17302.8

TABLE 7
ATMOSPHERIC BOILERS

Auxiliary Power Consumption
for 3-Boiler Turbine Units

	M.W.
BOILERS	
F.D. Fans	34.0
Coal and Ash Handling Plant	1.0
Coal Preparation and Injection	10.0
Feed Pumps	54.0
Circulating Pumps	19.0
Miscellaneous	3.0
Total for Boilers	121.0
TURBINES	
Turbine Plant Auxiliaries	32.0
River Water Pumps	11.0
Total for Turbines	43.0
Total for Station	164.0
% of Station Output	8.42

TABLE 8
PRESSURIZED BOILERS.

Auxiliary Power Consumption
for 3-Boiler Turbine Units

	M.W.
BOILERS	
Coal and Ash Handling Plant	0.9
Coal Prepration and Injection	9.0
Miscellaneous	3.0
Total for Boiler Plant	12.9
TURBINES	
Turbine Plant Auxiliaries	26.0
River Water Pumps	9.9
Total for Turbines	35.9
Total for Station	48.8
% of Station Output	2.6

TABLE 9
CAPITAL COSTS

Atmospheric Scheme
(Figures are thousands of Canadian dollars)

CAPITAL COST BREAKDOWN

MECHANICAL		
Fuel delivery storage and handling	15,755	
Boiler Plant including precipitators boiler feed pumps and water treatment	181,003	
Steam turbine generating plant including condensing and feed heating plant	97,812	
High pressure pipework and valves	14,664	
Low pressure pipework systems	5,790	
Ash and dust plant	16,445	
Miscellaneous mechanical equipment	5,401	
Total Mechanical		336,870
ELECTRICAL		
Transformers	10,065	
Switchgear	7,079	
Cabling and bus trunking	5,212	
Instruments and controls	3,491	
Miscellaneous electrical equipment	1,363	
Total Electrical		27,210
CIVIL		
Site preparation	7,188	
Camp accommodation	17,917	
Temporary works and site services	8,050	
Piling, excavation and backfilling	1,876	
Rail spur	9,343	
Main Building	41,935	
Other plant buildings	1,566	
Condenser cooling water system including cooling towers	22,520	
River water works and holding pond	29,670	
Stack	9,660	
Offices, Workshop, stores, gatehouse	632	
Total Civil		150,357
PLANT TOTAL		514,437

CAPITAL COST SUMMARY AND INDIRECT COSTS

Plant costs — mechanical	336,870
— electrical	27,210
— civil	150,357
— total	514,437
Engineering including procurement and construction supervision (8%)	41,155
Land	100
Sub-Total	555,692
Corporate overhead (5%)	27,785
Sub-Total	583,477
Interest during construction	190,921
Total station cost at September 30, 1975	774,398
Total station cost inflated to 1983 in-service date, including interest during construction.	1,219,346

TABLE 10
CAPITAL COSTS

PRESSURIZED SCHEME
(Figures are thousands of Canadian Dollars)

CAPITAL COST BREAKDOWN

MECHANICAL

Fuel delivery storage and handling	14,354	
Boiler plant including dust collectors boiler feed pumps and water treatment	74,191	
Steam turbine generating plant including condensing and feed heating plant	72,938	
Gas turbine generating plant	81,336	
High pressure pipework and valves	16,130	
Low pressure pipework systems	5,385	
Ash and dust plant	13,156	
Miscellaneous mechanical equipment	5,401	
Total Mechanical		282,891

ELECTRICAL

Transformers	15,368	
Switchgear	8,379	
Cabling and bus trunking	7,089	
Instruments and controls	5,237	
Miscellaneous electrical equipment	1,826	
Total Electrical		37,899

CIVIL

Site preparation	7,188	
Camp accommodation	17,917	
Temporary Works and site services	8,050	
Piling excavation and backfilling	1,876	
Rail spur	9,343	
Main building	51,187	
Other plant buildings	1,098	
Condenser cooling water system including cooling towers	19,828	
River water works and holding pond	29,670	
Stack	9,660	
Offices, workshop, stores, gatehouse	632	
Total Civil		156,449
PLANT TOTAL		477,239

CAPITAL COST SUMMARY AND INDIRECT COSTS

Plant costs — mechanical	282,891
— electrical	37,899
— civil	156,449
— total	477,239
Engineering including procurement and construction supervision (8%)	38,179
Land	100
Sub-Total	515,518
Corporate overhead (5%)	25,776
Sub-Total	541,294
Interest during construction	177,473
Total station cost at September 30, 1975	718,767
Total station cost inflated to 1988 in-service date, including interest during construction.	1,445,365

TABLE 11
CASH FLOW ATMOSPHERIC SCHEME
(Figures are thousands of Canadian dollars)
UNINFLATED

YEAR	1	2	3	4	5	6	TOTALS
Land	100	-	-	-	-	-	100
Civil Works	37,589	37,589	37,589	37,590	-	-	150.357
Mechanical and Electrical Works	91,020	91,020	91,020	36,408	36,408	18,204	364.080
Engineering, Construction Management and Corporate Overheads.	17,239	17,234	17,234	9,916	4,879	2,438	68,940
Sub-Total	145,948	145,843	145,843	83,914	41,287	20,642	583,477
Interest during Construction	7,297	22,617	39,463	54,897	66,647		190,921
Cash Flow (uninflated)	153,245	168,460	185,306	138,811	107,934	20,642	774,398

INFLATED

YEAR	1979	1980	1981	1982	1983	1984	TOTALS
Inflation from September 30, 1975	46%	54%	61%	69%	78%	87%	
Inflated Cash Flow excluding interest during construction	213,084	224,598	234,807	141,815	73,491	38,600	926,395
Interest during construction	10,654	33,604	59,934	84,759	104,000		292.951
Inflated Cash Flow	223,738	258,202	294,741	226,574	177,491	38,600	1,219,346

TABLE 12
CASH FLOW PRESSURIZED SCHEME
(Figures are thousands of Canadian dollars)

UNINFLATED

	1	2	3	4	5	6	TOTALS
Land	100						100
Civil Works	39,112	39,112	39,112	39,113			156,449
Mechanical and Electrical Works	80,198	80,198	80,198	32,079	32,079	16,038	320,790
Engineering, Construction Management and Corporate Overheads	16,000	15,985	15,985	9,538	4,298	2,149	63,955
Sub-total	135,410	135,295	135,295	80,730	36,377	18,187	541,294
Interest during construction	6,770	20,983	36,611	51,073	62,036	-	177,473
Cash Flow (uninflated)	142,180	156,278	171,906	131,803	98,413	18,187	718,767

INFLATED

YEAR	1984	1985	1986	1987	1988	1989	TOTALS
inflation from September 30 1975	87%	96%	106%	116%	127%	138%	
Inflated cash flow excluding interest during construction	253,217	265,178	278,708	174,377	82,576	432,85	1,097,341
Interest during construction	12,661	39,847	71,026	100,782	123,708		348,024
Inflated Cash Flow	265,878	305,025	349,734	275,159	206,284	432,85	1,445,365

TABLE 13
POWER COST ESTIMATE ATMOSPHERIC SCHEME
Capital cost (including interest during construction) $ CAN. 774,398,00 (uninflated)

FIXED CHARGES	COST PER YEAR % OF CAPITAL COST	COST MILLION CANADIAN DOLLARS PER YEAR
Operation and Maintenance	1.45	11.228
Administration and General	0.3625	2.807
Insurance	0.25	1.935
Interim Replacement	0.35	2.710
Taxes	1.00	7.743
Interest on Capital	10.00	77.439
Depreciation	0.369	2.857
TOTAL	13.7815	106.719

LOAD FACTOR	%	60	70	80
Coal Cost	M$/yr	22.516	26.270	30.022
Start-up Oil Cost	M$/yr	0.022	0.022	0.022
Total Annual Cost	M$/yr	129.257	133.011	136.763
Units Sent Out per year	GWh	9355.7	10915.0	12474.2
Cost per Unit	$/kWh	0.0138	0.0122	0.0109
	mills/kWh	13.8	12.2	10.9
Variable Maintenance Cost	mills/kWh	0.3	0.3	0.3
Total Power Cost per unit Sent Out	mills/kWh	14.1	12.5	11.2

TABLE 14
POWER COST ESTIMATE PRESSURIZED SCHEME
Capital cost (including interest during construction) $ CAN. 718,767,000 (uninflated)

FIXED CHARGES	COST PER YEAR % OF CAPITAL COST	COST MILLION CANADIAN DOLLARS PER YEAR		
Operation and Maintenance	1.308		9.401	
Administration and General	0.327		2.350	
Insurance	0.25		1.796	
Interim Replacement	0.35		2.515	
Taxes	1.0		7.187	
Interest on Capital	10.0		71.876	
Depreciation	0.369		2.652	
TOTAL	13.604		97.777	
LOAD FACTOR	%	60	70	80
Coal Cost	M$/yr	21.190	24.721	28.253
Start-up Oil Cost	M$/yr	0.022	0.022	0.022
Total Annual Cost	M$/yr	118.989	122.520	126.056
Units Sent Out per year	GWh	9571.2	11166.4	12761.6
Cost per Unit	$/kWh	0.0124	0.0109	0.0098
	mills/kWh	12.4	10.9	9.8
Variable Maintenance Cost	mills/kWh	0.48	0.48	0.48
Total Power Cost	mills/kWh	12.9	11.4	10.3

TABLE 15
COMPARISON BETWEEN ATMOSPHERIC & PRESSURIZED UNITS

TYPE OF PLANT		ATMOSPHERIC	PRESSURIZED
TECHNICAL DATA			
Station Output Gross	MW	1944	1870
Station Output Net	MW	1780	1821
Auxiliary Power Consumption	MW	164	49
Auxiliary Power % of Station Output		8.4	2.6
No. of Units		3	3
Unit Size (Gross)	MW	648	623
Steam Turbine Generator Output	MW	648	476
Gas Turbine Generator Output	MW	-	147
Gas Pressure in Boilers	psig	Atmospheric	182
Coal Quantity per Unit at Full Load	klb/h	952	896
Heat Rate (coal/U.S.O.)	Btu/kWh	10270	9450
Station efficiency (U.S.O.)	%	33.2	36.1
COSTS			
Capital cost (including I.d.c.)	M$CAN.	774.398	718.767
Annual Charges	M$CAN.	106.719	97.777
Annual Fuel Cost 60% load	M$CAN.	22.538	21.212
(coal + start-up oil) 70% load	M$CAN.	26.292	24.743
80% load	M$CAN.	30.044	28.275
Units sent out per Year 60%	GWh	9355.7	9571.2
70%	GWh	10915.0	11166.4
80%	GWh	12474.2	12761.6
Total power cost 60%	mills/kWh	14.1	12.9
70%	mills/kWh	12.5	11.4
80%	mills/kWh	11.2	10.3
Capital cost per kW of S.O. capacity (including I.d.c.)	$CAN.	435	395
ENVIRONMENTAL CONSIDERATIONS			
Ash Quantity (per unit)	klb/h	123.6	107.6
Dust Quantity (per unit)	klb/h	123.8	110.2
Type of Gas Cleaning Equipment		Precipitator	Multi-Cyclones
Stack Emission — particulates per ton of coal	lb/ton	5	2.5
— SO_2	lb/ton	15	15
— No_x	lb/ton	7 to 18	2.6
Cooling tower evaporation	klb/h	2819.8	2458.8
River water supply	klb/h	10397.7	8939.4
TYPE OF PLANT			
CONSTRUCTION			
Earliest date for Order		Oct. 1975	Oct. 1980
Time Scale for Design & Construction	Years	7½	7½
Earliest date in Service		April 1983	April 1988

TABLE 16
ALTERNATIVE ESTIMATE OF INTEREST DURING CONSTRUCTION

CAPITAL COSTS ATMOSPHERIC SCHEME

Interest during construction — percentage of capital cost		26.6
Total plant costs, engineering including procurement and construction supervision (8%), land and corporate overhead (5%) — see Table 9.	$ CAN	583,477,000
Interest during construction	$ CAN	155,205,000
Total station cost at September 30, 1975	$ CAN	738,682,000
Capital cost per kW of S.O. capacity (including interest during construction).	$CAN	415

TABLE 17
ALTERNATIVE ESTIMATE OF INTEREST DURING CONSTRUCTION

CAPITAL COSTS PRESSURIZED SCHEME

Interest during construction — percentage of capital cost		21.0
Total plant costs, engineering including procurement and construction supervision (8%), land and corporate overhead (5%) — see Table 10.	$ CAN	541,294,000
Interest during construction	$ CAN	113,672,000
Total station cost at September 30, 1975	$ CAN	654,966,000
Capital cost per kW of S.O. capacity (including interest during construction).	$ CAN	360

TABLE 18
ALTERNATE ESTIMATE OF INTEREST DURING CONSTRUCTION
POWER COST ESTIMATE ATMOSPHERIC SCHEME
Capital cost (including interest during construction) $ CAN. 738,682,000 (uninflated)

FIXED CHARGES	COST PER YEAR % OF CAPITAL COST	COST MILLION CANADIAN DOLLARS PER YEAR
Operation and Maintenance	1.45	10.711
Administration and General	0.3625	2.678
Insurance	0.25	1.847
interim Replacement	0.35	2.585
Taxes	1.00	7.387
Interest on Capital	10.00	73.868
Depreciation	0.369	2.726
TOTAL	13.7815	101.802

LOAD FACTOR	%	60	70	80
Coal Cost	M$/yr	22.516	26.270	30.022
Start-up Oil Cost	M$/yr	0.022	0.022	0.022
Total Annual Cost	M$/yr	124.340	128.094	131.846
Units Sent Out per year	GWh	9355.7	10915.0	12474.2
Cost per Unit	$/kWh	0.0133	0.0117	0.0106
	mills/kWh	13.3	11.7	10.6
Variable Maintenance Cost	mills/kWh	0.3	0.3	0.3
Total Power Cost per unit Sent Out	mills/kWh	13.6	12.0	10.9

TABLE 19
ALTERNATIVE ESTIMATE OF INTEREST DURING CONSTRUCTION
POWER COST ESTIMATE PRESSURIZED SCHEME
Capital cost (including interest during construction) $ CAN. 654,966,000 (uninflated)

FIXED CHARGES	COST PER YEAR % OF CAPITAL COST	COST MILLION CANADIAN DOLLARS PER YEAR
Operation and Maintenance	1.308	8.567
Administration and General	0.327	2.142
Insurance	0.25	1.637
Interim Replacement	0.35	2.292
Taxes	1.0	6.550
Interest on Capital	10.0	65.497
Depreciation	0.369	2.417
TOTAL	13.604	89.102

LOAD FACTOR	%	60	70	80
Coal Cost	M$/yr	21.190	24.721	28.253
Start-up Oil Cost	M$/yr	0.022	0.022	0.022
Total Annual Cost	M$/yr	110.314	113.845	117.377
Units Sent Out per year	GWh	9571.2	11166.4	12761.6
Cost per Unit	$/kWh	0.0115	0.0102	0.0092
	mills/kWh	11.5	10.2	9.2
Variable Maintenance Cost	mills/kWh	0.48	0.48	0.48
Total Power Cost	mills/kWh	12.0	10.7	9.7

APPENDIX 1

FURTHER DETAILS OF FOSTER WHEELER'S ACTIVITIES IN FLUIDIZED BED COMBUSTION.

GENERAL

Foster Wheeler are associated with Pope Evans and Robbins (PER) of New York, NY, USA in the Fluidized Bed Combustion Company. In 1967 PER, under contract from the United States Office of Coal Research, built a 5000 lb/h coal-fired, atmospheric fluidized combustion boiler and operated it for several years (Reference 9).

Subsequent to this work, further funding was provided to PER and Foster Wheeler by the Office of Coal Research (now part of the Energy Research and Development Administration) for the 300,000 lb/h boiler referred to in 11.1. This unit is a coal-fired. atmospheric fluidized combustion boiler intended to supply steam at 1350 psig 925°F to existing steam turbine plant at the Rivesville Power Station of the monongahela Power Company (Allegheny Power System) West Virginia, U.S.A.

The date originally planned for the commissioning of this boiler has not been achieved and it is currently expected to start up in mid 1976 (Reference 16). The boiler is now complete and the balance of plant equipment is presently being completed. System check-outs are planned for May-June with initial firing in July.

Foster Wheeler and PER have made a conceptual design of an 800 MW (e) coal-fired atmospheric fluidized combustion boiler comprising four identical 200 MW (e) modules and anticipate that operation of the 300,000 lb/h Rivesville unit will provide the detailed information necessary to complete the design and permit the fabrication and erection of one of the 200 MW (e) modules referred to.

Foster Wheeler anticipate that they would be able to accept an order, on normal commercial terms, for a utility fluidized bed steam generator in early 1977 after several months operation of the Rivesville plant.

Apart from the Rivesville unit, we are not aware of any orders for fluidized combustion boilers already received by Foster Wheeler, although they are active in proposals to and discussion with ERDA and prospective Clients.

Foster Wheeler are negotiating with ERDA concerning a substantial development effort on pressurized fluidized combustion.

800 MW (E) BOILER

The 800 MW (e) boiler proposed by Foster Wheeler differs from the 660 MW (e) unit proposed by Combustion Systems Limited principally in the arrangement and in the fuel injection equipment.

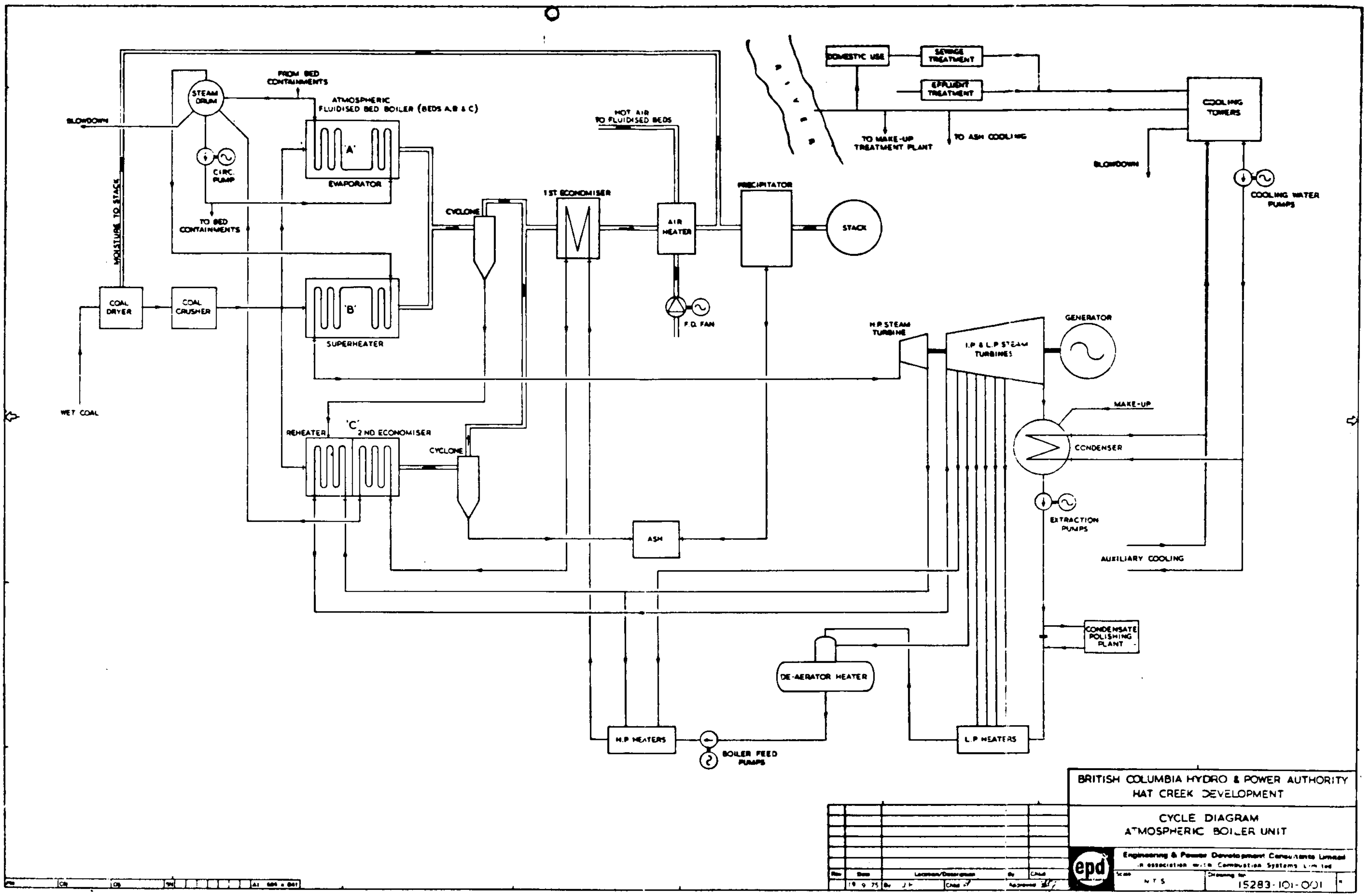
The Foster Wheeler boiler is arranged in four 200 MW (e) modules, each of which may be operated or shut down independently of the others. The modules contain several cells stacked vertically; each cell comprises a fluidized bed with its associated heat transfer surface. One of the cells in each module is a carbon burn-up cell operating at a lower fluidizing velocity and at a higher temperature than the remainder.

Coal is injected into the beds pneumatically in a downwards direction by multiple injection pipes rather than upwards as in the CSL design.

In other respects the boiler designs of the two organisations appear similar although rather few details are available of the Foster Wheeler proposals.

ADDITIONAL REFERENCE

16. Fluidized bed combustion, clean power from high sulfur — low grade fuels. R.L. Gamble and F.R. Warshany. Annual Conference of the South Eastern Electric Exchange, Bal Harbour, Florida. April 17 - 18, 1975.


FROM BED CONTAINMENTS
STEAM DRUM
ATMOSPHERIC FLUIDISED BED BOILER (BEDS A,B & C)
BLOWDOWN
'A'
CIRC. PUMP
EVAPORATOR
MOISTURE TO STACK
TO BED CONTAINMENTS
HOT AIR TO FLUIDISED BEDS
CYCLONE
1ST ECONOMISER
AIR HEATER
PRECIPITATOR
STACK
RIVER
DOMESTIC USE
SEWAGE TREATMENT
EFFLUENT TREATMENT
COOLING TOWERS
TO MAKE-UP TREATMENT PLANT
TO ASH COOLING
BLOWDOWN
COOLING WATER PUMPS
COAL DRYER
COAL CRUSHER
'B'
SUPERHEATER
F.D. FAN
H.P. STEAM TURBINE
I.P & L.P STEAM TURBINES
GENERATOR
WET COAL
MAKE-UP
REHEATER
'C'
2ND ECONOMISER
CYCLONE
CONDENSER
EXTRACTION PUMPS
ASH
AUXILIARY COOLING
CONDENSATE POLISHING PLANT
DE-AERATOR HEATER
H.P HEATERS
BOILER FEED PUMPS
L.P HEATERS
BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT
CYCLE DIAGRAM
ATMOSPHERIC BOILER UNIT
epd
Engineering & Power Development Consultants Limited
in association with Combustion Systems Limited
Scale
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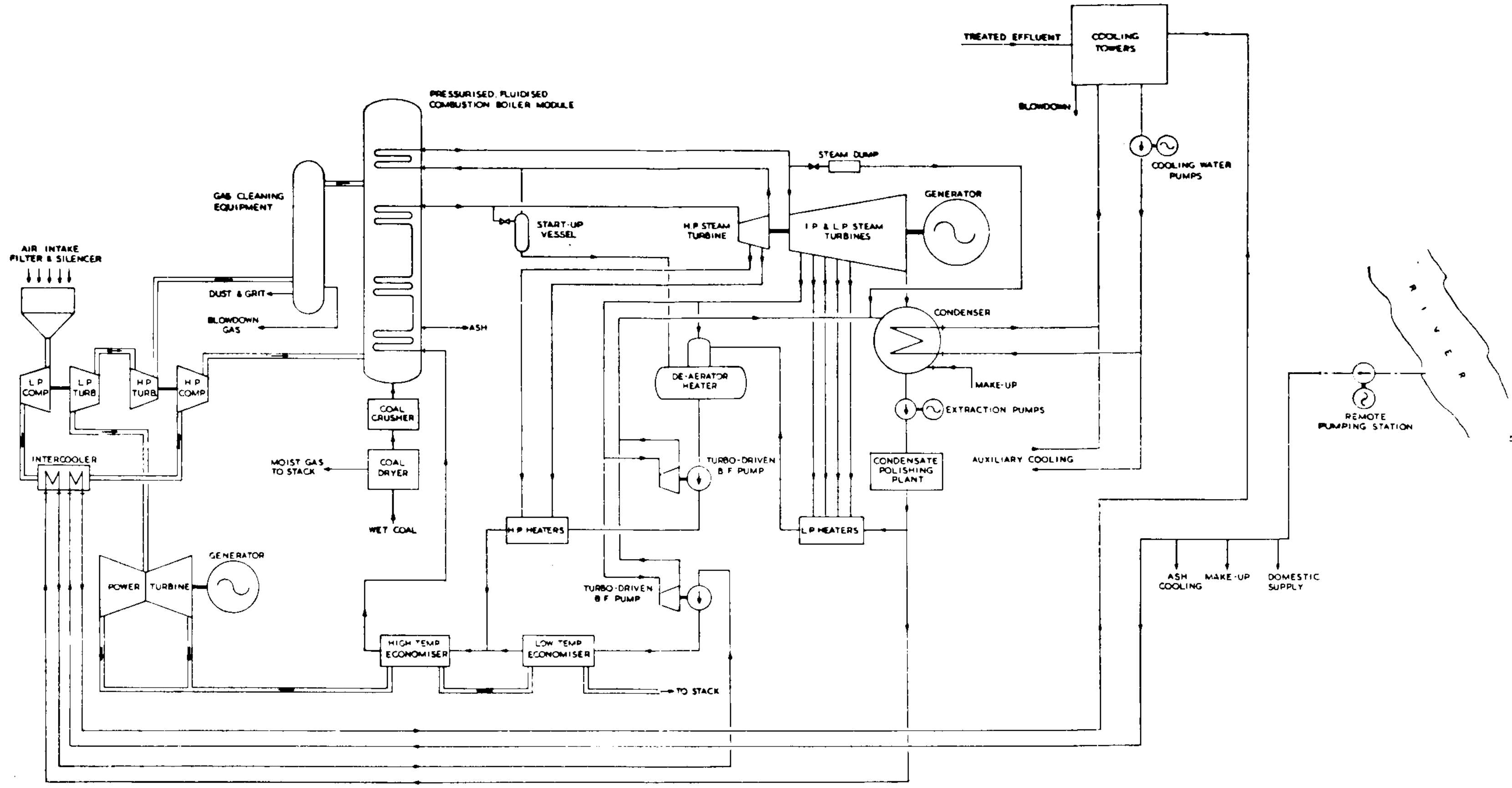
Drawing No
15283-101-001
19 9 75
By J.P



BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

CYCLE DIAGRAM
COMBINED CYCLE UNIT

Engineering & Power Development Consultants Limited
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Rev	Date	Location/Description	By	Chkd
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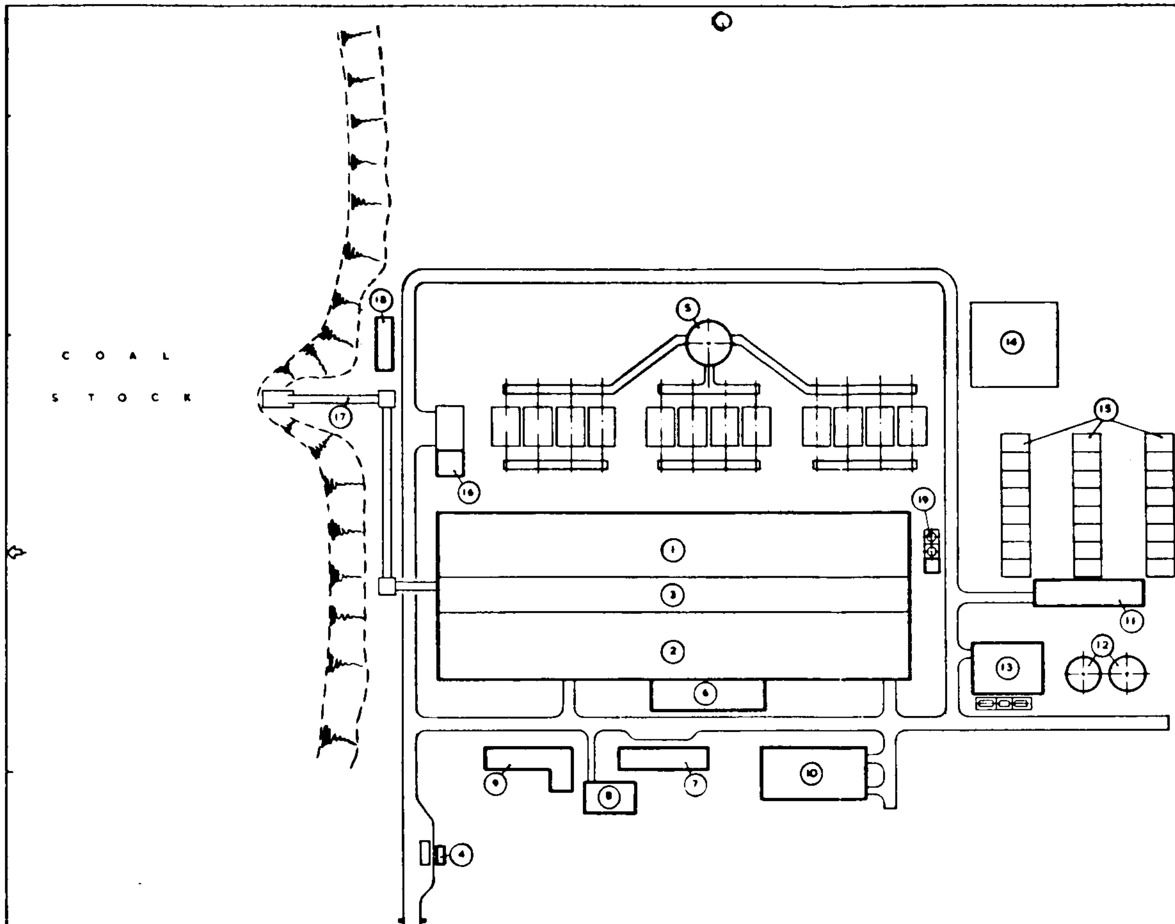
Scale: N T S

Drawing No: 15283-101-002



1. BOILER HOUSE
2. TURBINE HOUSE
3. HEATERS & SWITCHGEAR BAY
4. GATE HOUSE & WEIGHBRIDGE
5. STACK
6. CONTROL BLOCK
7. ADMIN. BUILDING
8. CANTEEN
9. WELFARE & AMENITIES
10. WORKSHOP & STORES
11. C.W. PUMPHOUSE
12. R.F.W. STORAGE
13. WATER TREATMENT PLANT
14. EFFLUENT TREATMENT PLANT
15. COOLING TOWERS
16. ASH PLANT & STORAGE
17. COAL CONVEYOR
18. COAL HANDLING CONTROL BUILDING
19. START-UP FUEL OIL STORAGE & PUMPHOUSE

BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

FLUIDISED BED COMBUSTION STUDY
POWER STATION – SITE PLAN
ATMOSPHERIC UNITS

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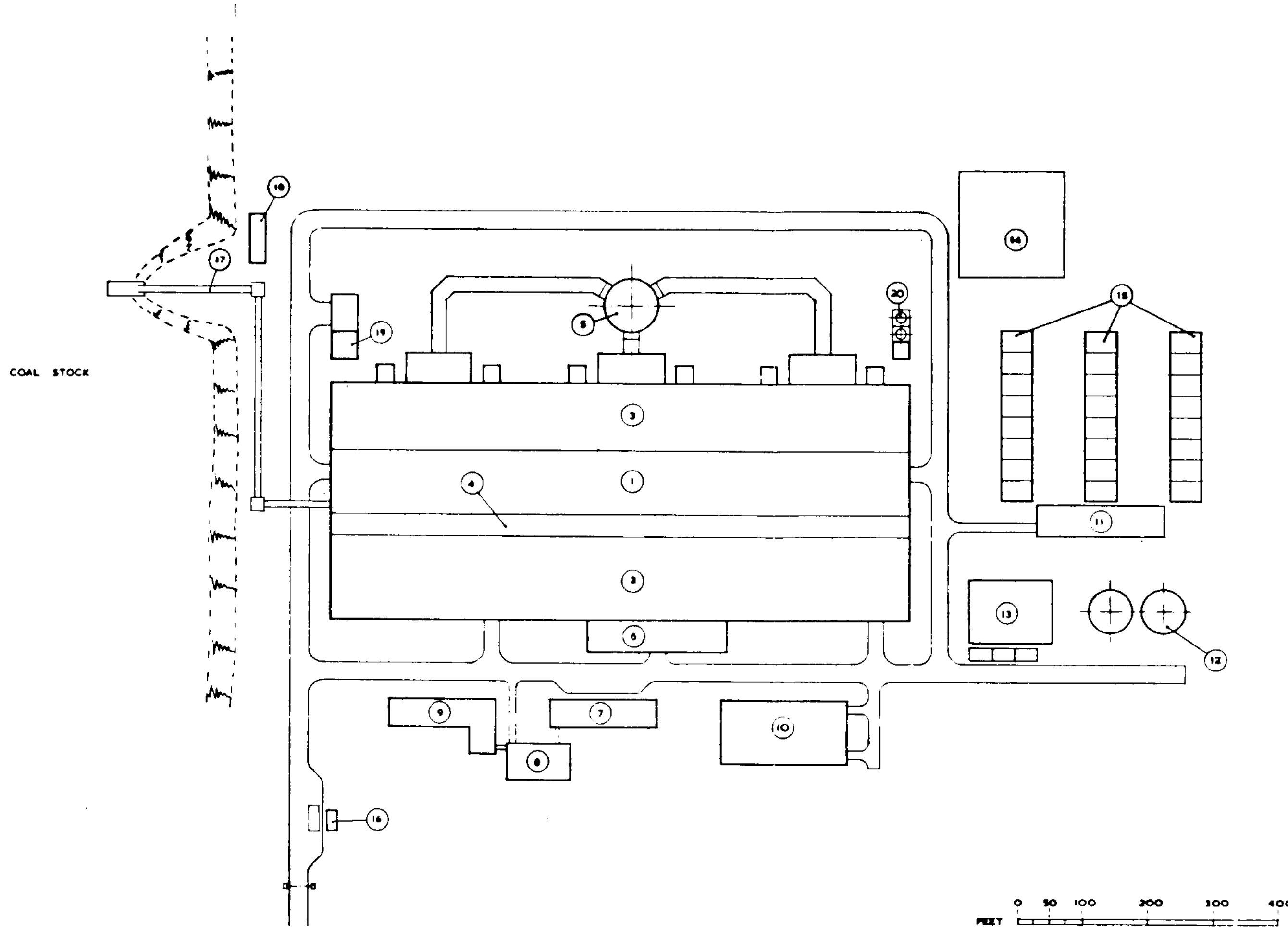
Scale 1:1250 | Drawing No 15283-101-003 | A

Date 19.9.75 | By R.E.H.



KEY :-

1. BOILER HOUSE
2. STEAM TURBINE HOUSE
3. GAS TURBINE HOUSE
4. HEATERS & SWITCHGEAR BAY
5. STACK
6. CONTROL BLOCK
7. ADMIN. BUILDING
8. CANTEEN
9. WELFARE & AMENITIES
10. WORKSHOP & STORES
11. CW PUMP HOUSE
12. R.F.W. STORAGE
13. WATER TREATMENT PLANT
14. EFFLUENT TREATMENT PLANT
15. COOLING TOWERS
16. GATE HOUSE & WEIGHBRIDGE
17. COAL CONVEYORS
18. COAL HANDLING CONTROL BUILDING
19. ASH PLANT & STORAGE
20. START-UP FUEL OIL STORAGE & PUMPHOUSE

BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

FLUIDISED BED COMBUSTION STUDY
POWER STATION SITE PLAN
PRESSURISED UNITS

Engineering & Power Development Consultants Limited
in association with Combustion Systems Limited

Rev	Date	Location/Description	By	Chkd
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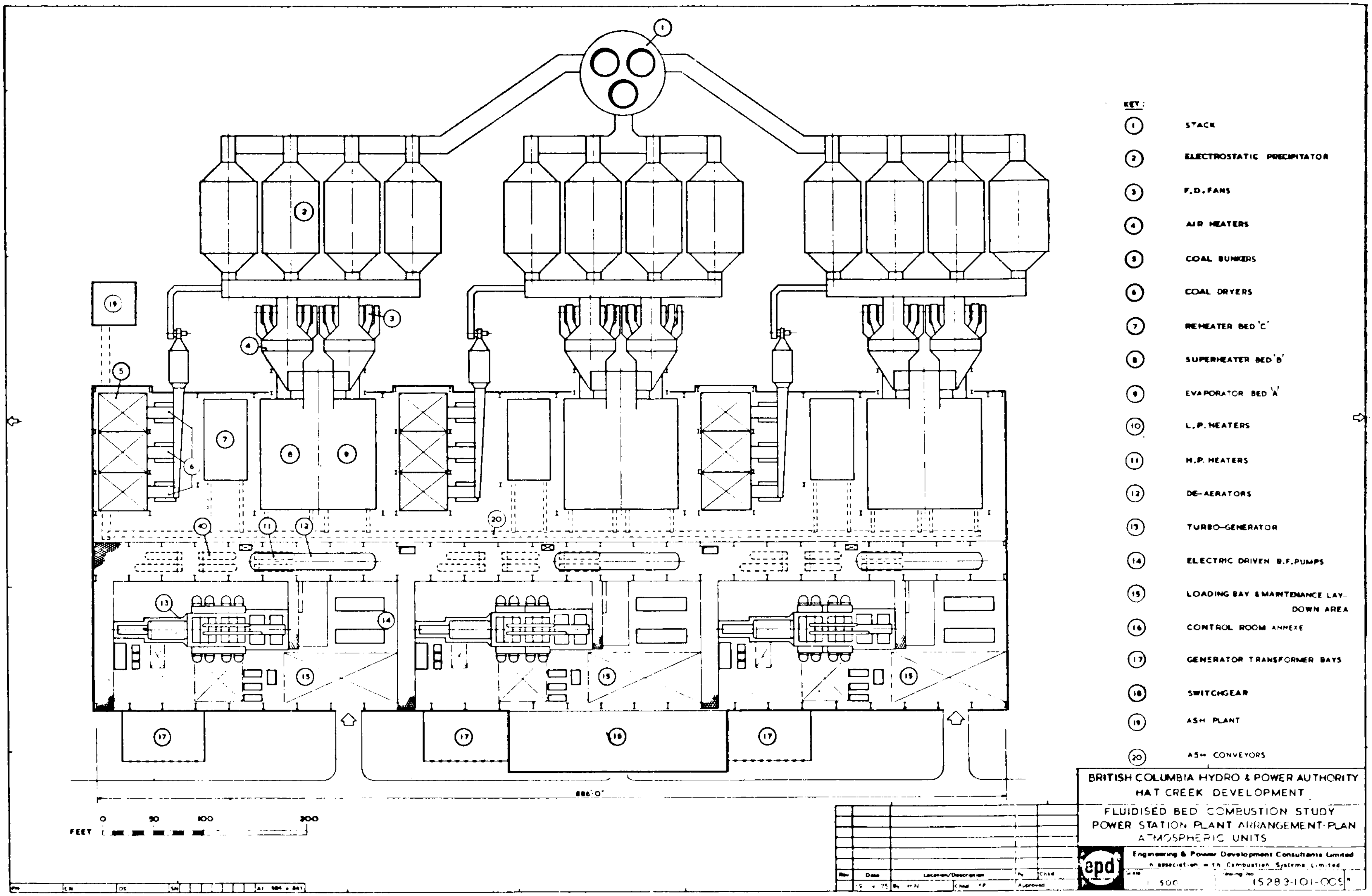
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2 ELECTROSTATIC PRECIPITATOR
3 F.D. FANS
4 AIR HEATERS
5 COAL BUNKERS
6 COAL DRYERS
7 REHEATER BED 'C'
8 SUPERHEATER BED 'B'
9 EVAPORATOR BED 'A'
10 L.P. HEATERS
11 H.P. HEATERS
12 DE-AERATORS
13 TURBO-GENERATOR
14 ELECTRIC DRIVEN B.F. PUMPS
15 LOADING BAY & MAINTENANCE LAY-DOWN AREA
16 CONTROL ROOM ANNEXE
17 GENERATOR TRANSFORMER BAYS
18 SWITCHGEAR
19 ASH PLANT
20 ASH CONVEYORS
886'0"
FEET 0 50 100 300
BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT
FLUIDISED BED COMBUSTION STUDY
POWER STATION PLANT ARRANGEMENT-PLAN
ATMOSPHERIC UNITS
Engineering & Power Development Consultants Limited
in association with Combustion Systems Limited
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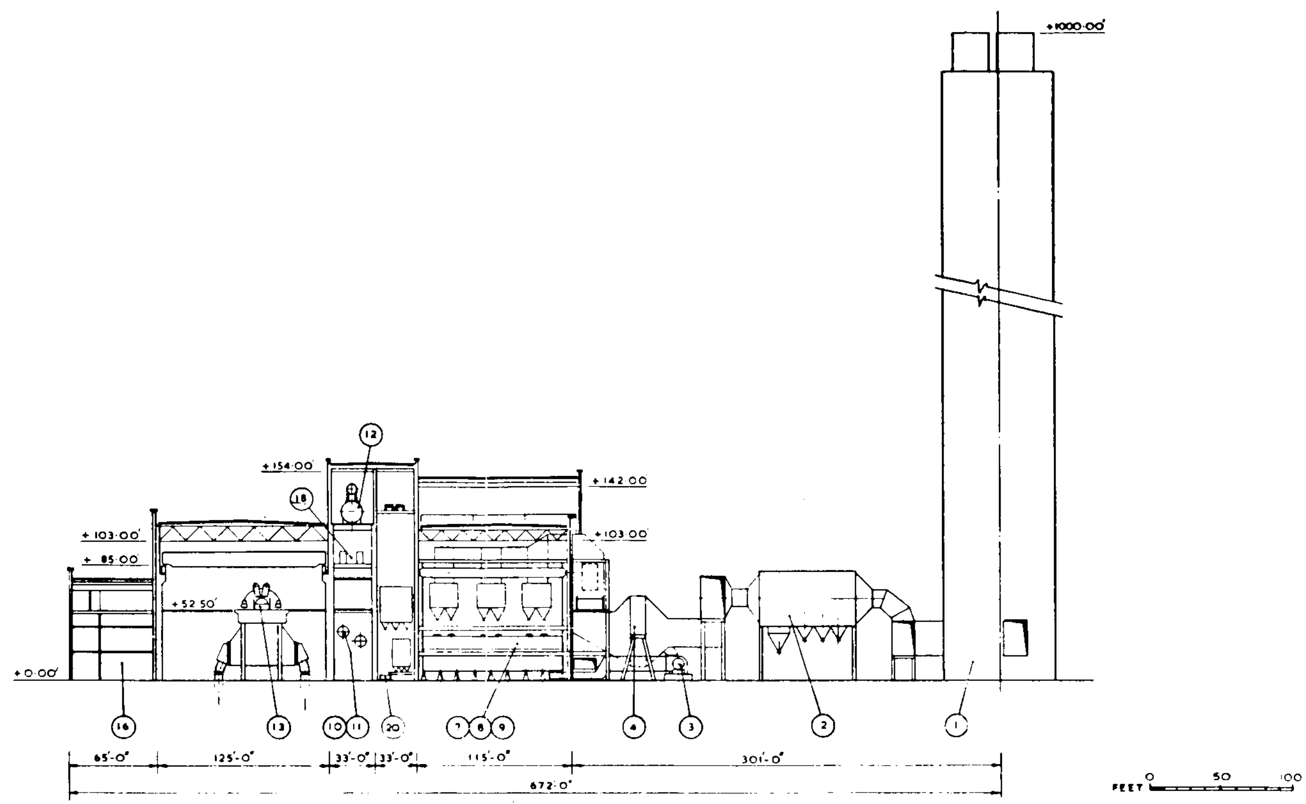
15283-101-005



BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

FLUIDISED BED COMBUSTION STUDY
POWER STATION PLANT ARRGT-SECTIONAL ELEV
ATMOSPHERIC UNITS

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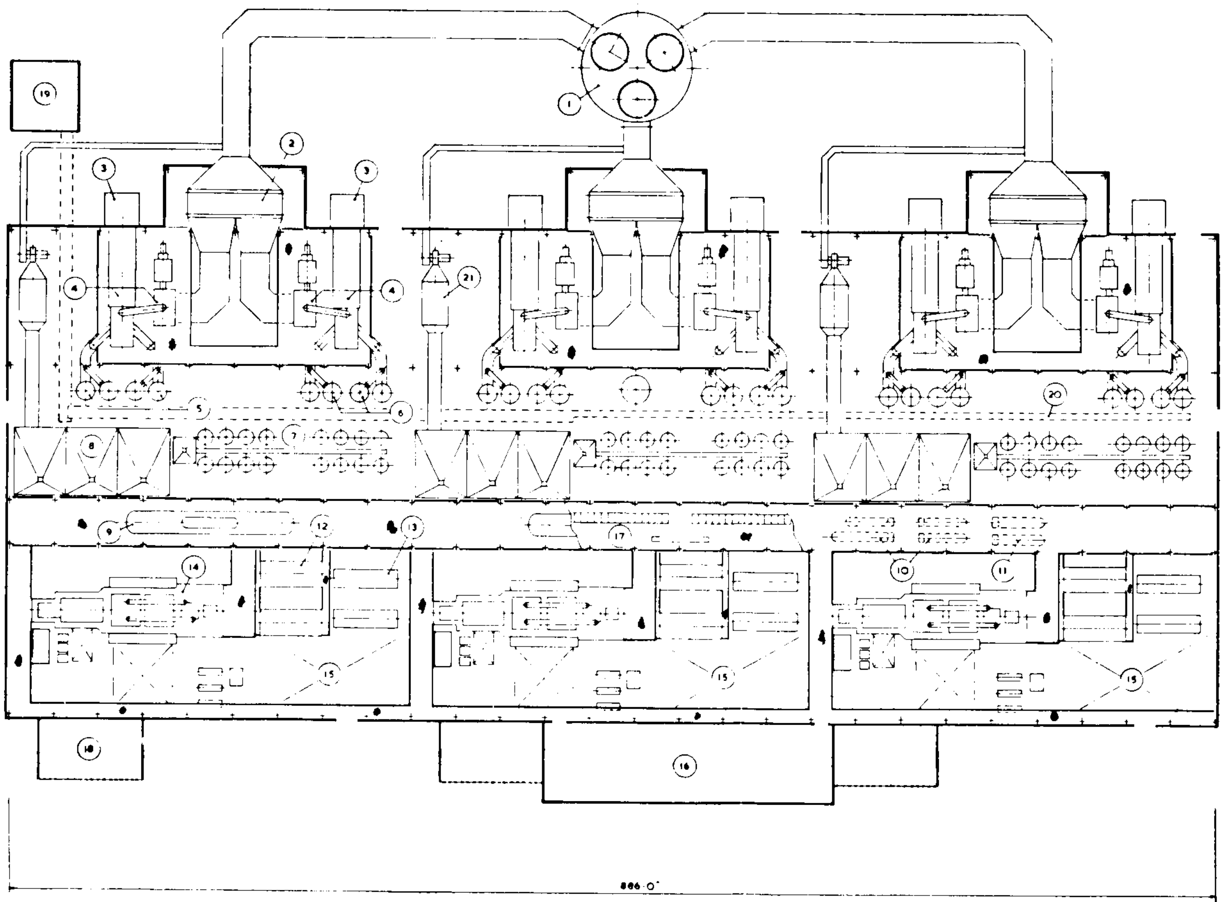
Drawing No. 15283-101-006



KEY:-

1. STACK
2. ECONOMISER
3. AIR INTAKE
4. GAS TURBO-GENERATORS
5. BOILER MODULES
6. GAS CLEANING MODULES
7. COAL FEEDER UNITS
8. COAL STORAGE BUNKERS
9. DE-AERATORS
10. L.P. HEATERS
11. H.P. HEATERS
12. BOILER FEED PUMPS - TURBO DRIVEN
13. BOILER FEED PUMPS - ELECTRIC DRIVEN
14. TURBO-GENERATOR
15. LOADING BAY & MAINTENANCE LAY-DOWN AREA
16. CONTROL ROOM ANNEXE
17. SWITCHGEAR
18. GENERATOR TRANSFORMERS
19. ASH PLANT
20. ASH CONVEYORS
21. COAL DUST PRECIPITATORS

FEET 0 10 20 30 40 50 100 150

BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

FLUIDISED BED COMBUSTION STUDY
POWER STATION PLANT ARRANGEMENT - PLAN
PRESSURISED UNITS

epd Engineering & Power Development Consultants Limited
in association with Combustion Systems Limited

Rev	Date	Location/Description	By	Chkd
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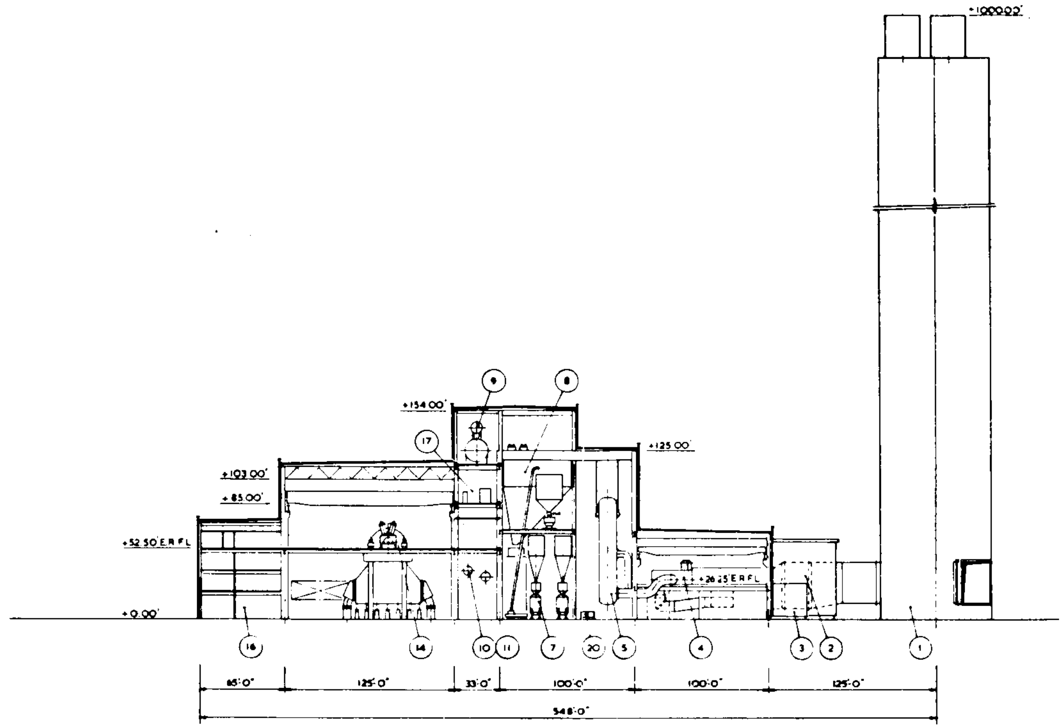

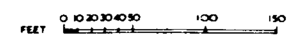

BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

FLUIDISED BED COMBUSTION STUDY
POWER STATION PLANT ARRGT·SECTIONAL ELEV
PRESSUR BED UNITS

Engineering & Power Development Consultants Limited
in association with Combustion Systems Limited

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	19.9.75	By	Chkd	Approved

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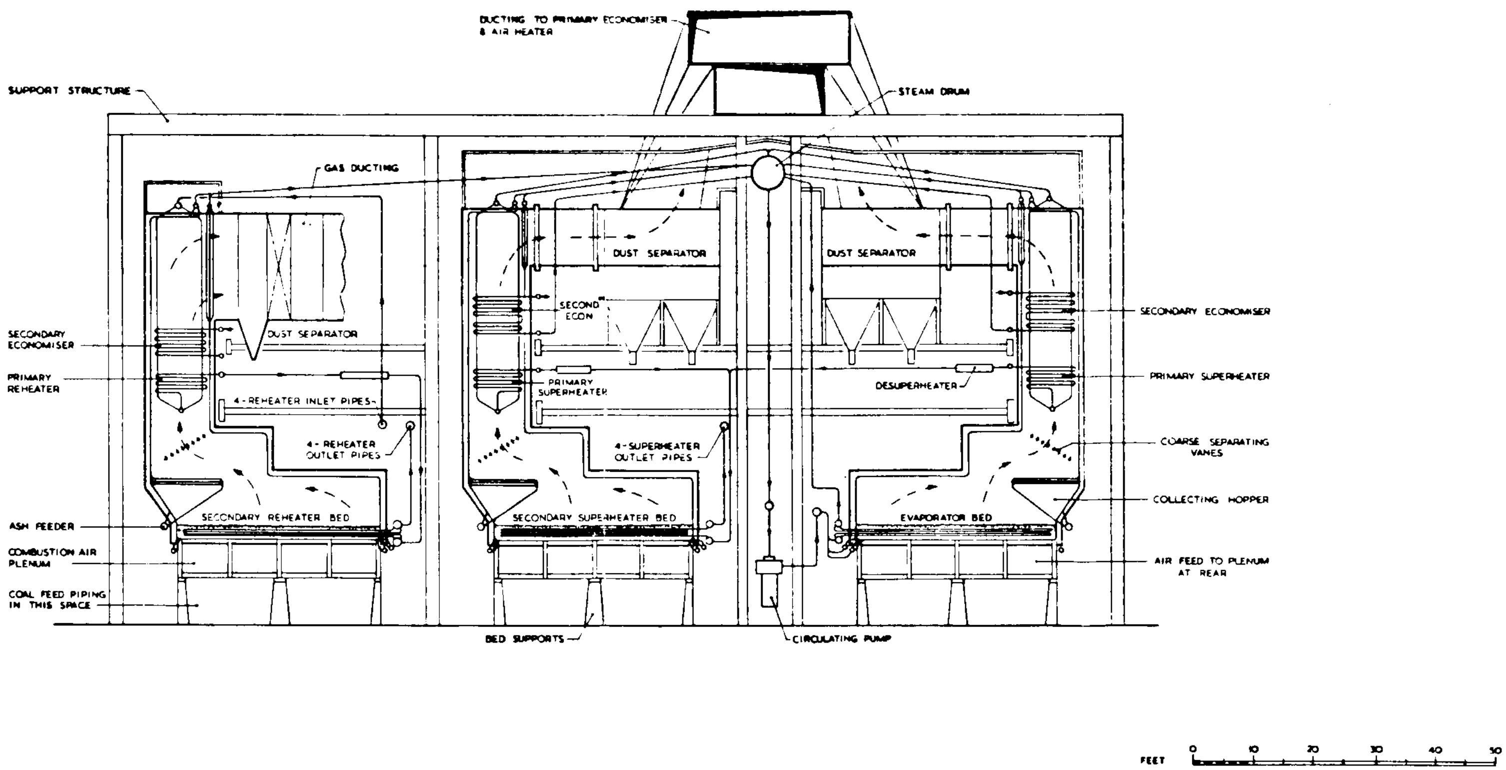


BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

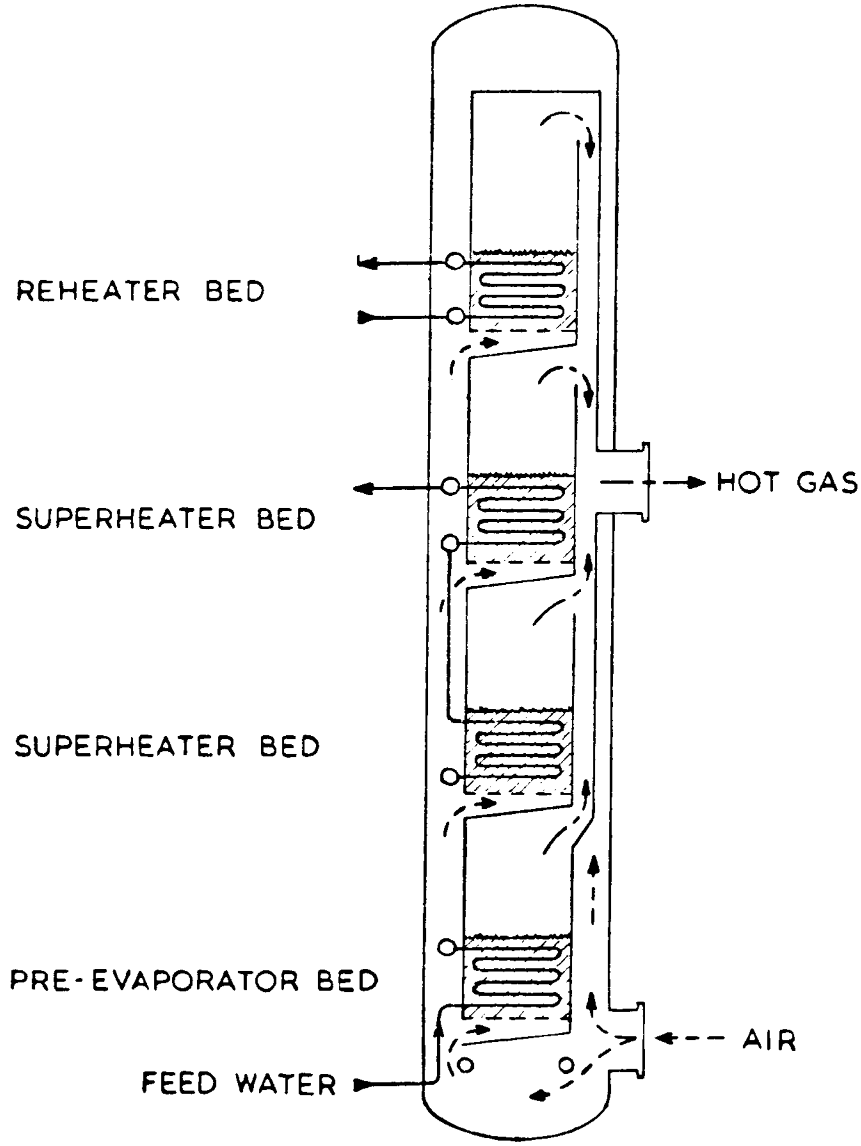
ATMOSPHERIC BOILER UNIT

epd Engineering & Power Development Consultants Limited in association with Combustion Systems Limited

AS INDICATED

15283-101-009

19 9 75 J.H.



B.C. HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

DIAGRAM OF BOILER MODULE

Engineering & Power Development Consultants
in assoc. with Combustion Systems Limited

Rev.	Date	Location/Description	By	Chkd.
	19.9.75	By J.H. Chkd. FP	Approved	

epd

Scale: N.T.S

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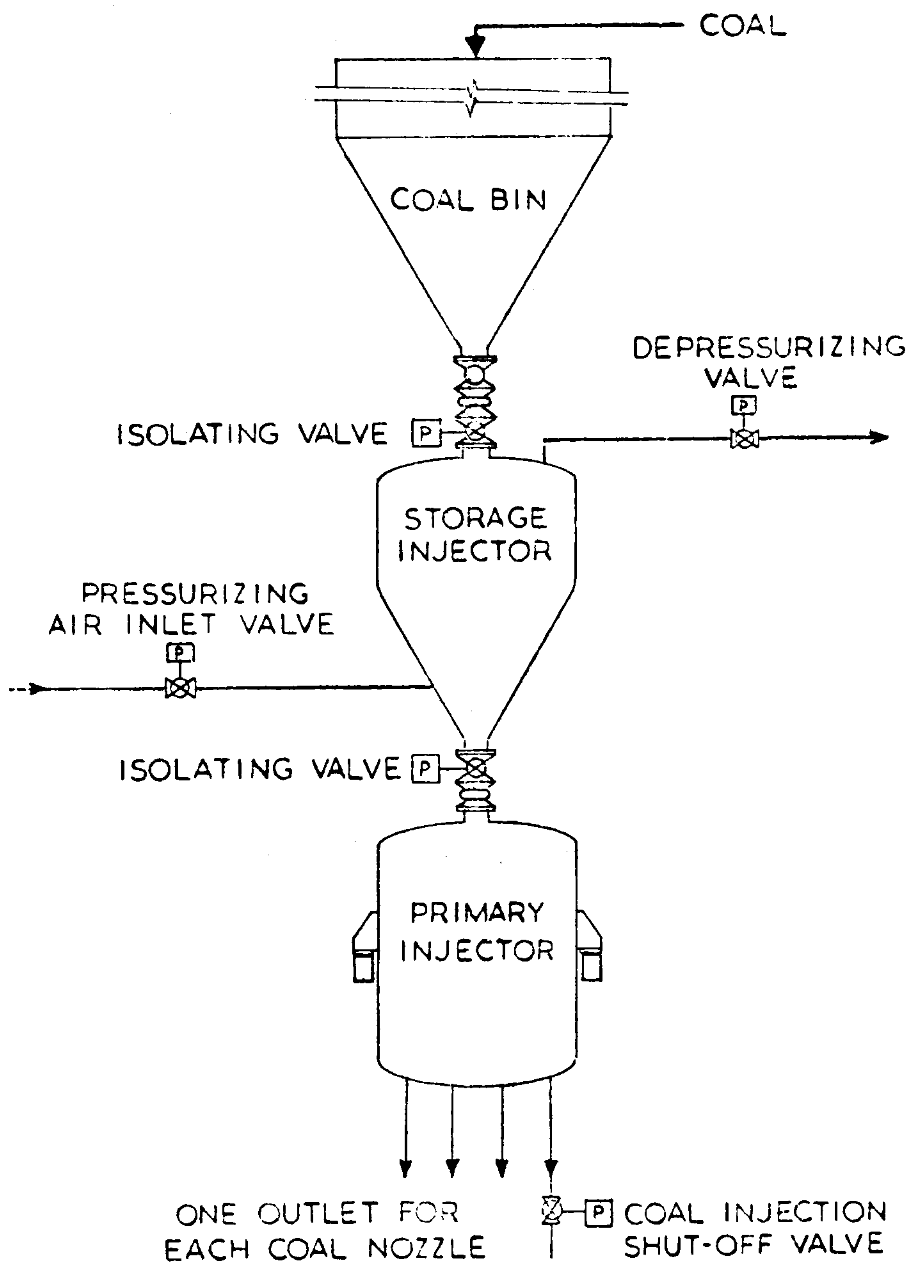

B.C. HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT.

PRESSURIZED
COAL FEEDING SYSTEM.

epd

Engineering & Power Development Consultants
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Rev	Date	Location/Description	By	Chkd
	19.9.75	By J.H. Chkd.	Approved	

Scale N.T.S.

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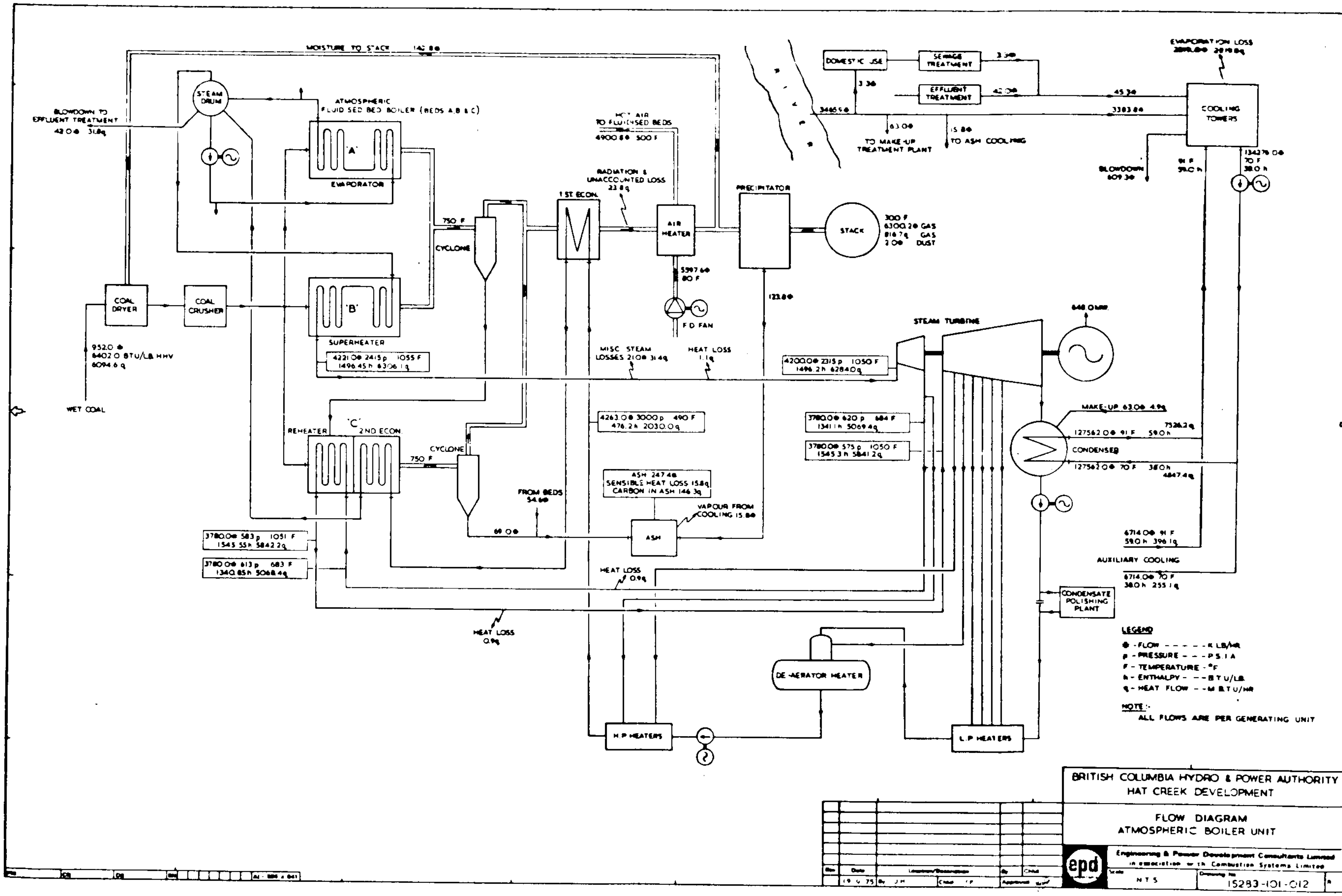
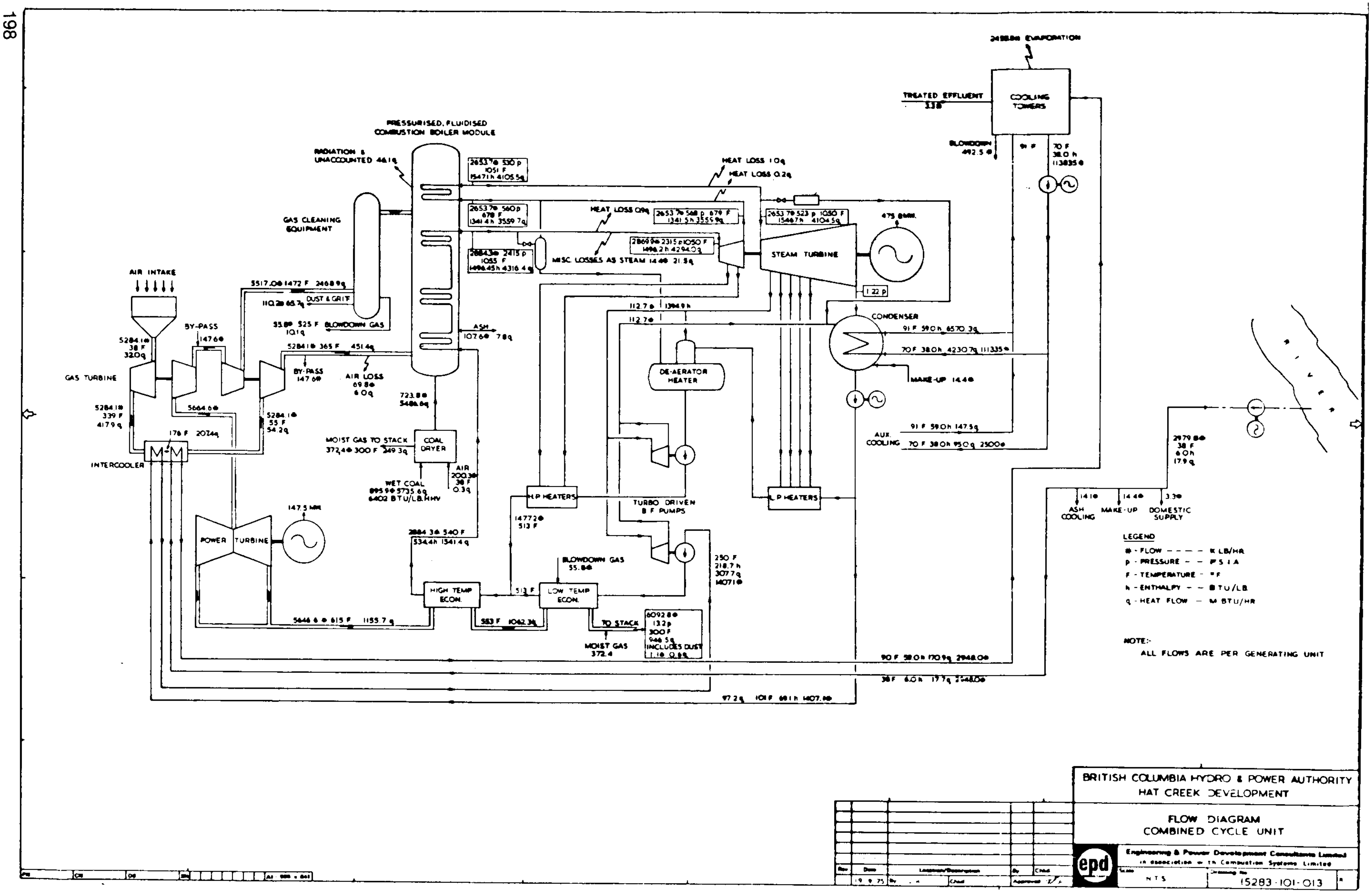
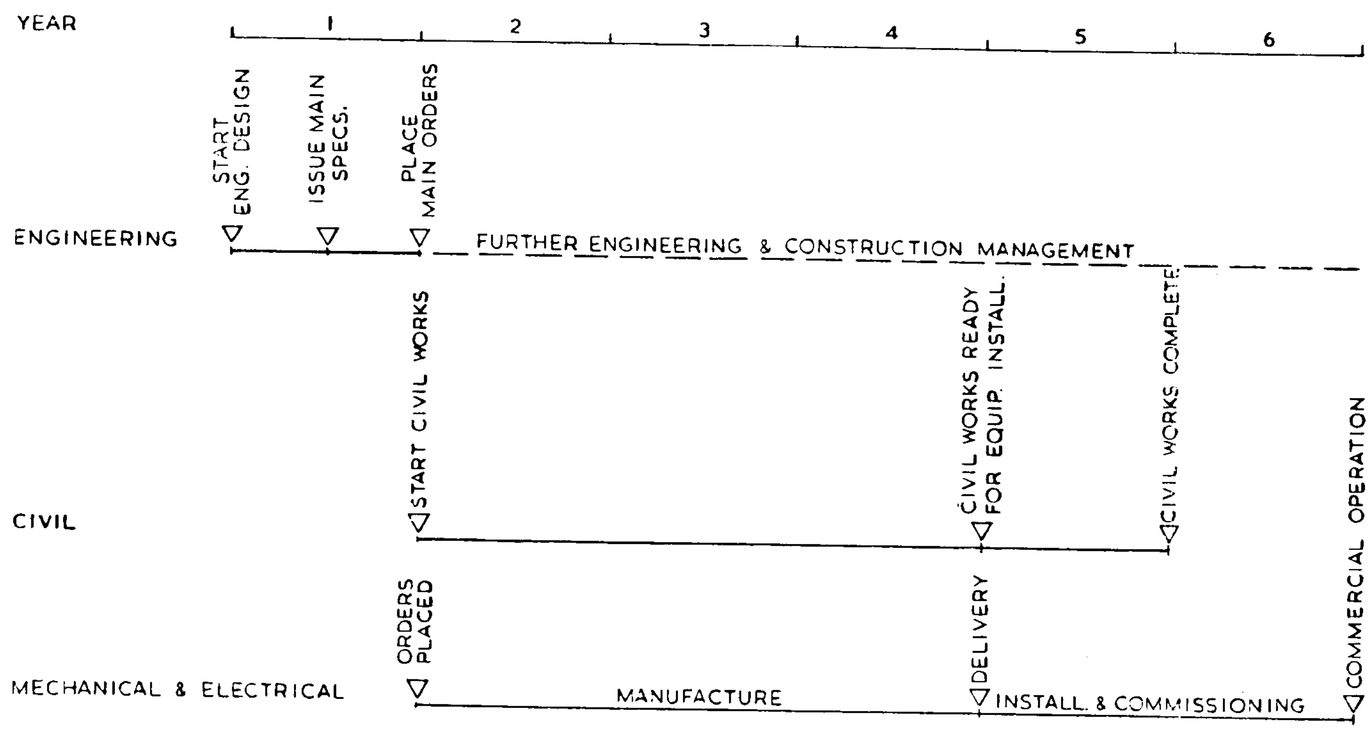


MOISTURE TO STACK 142.8⊕
BLOWDOWN TO EFFLUENT TREATMENT 42.0⊕ 31.8q
STEAM DRUM
ATMOSPHERIC FLUID BED BOILER (BEDS A,B & C)
'A'
EVAPORATOR
'B'
SUPERHEATER
'C'
REHEATER
2ND ECON
COAL DRYER
COAL CRUSHER
WET COAL
9520 ⊕
6402.0 BTU/LB HHV
6094.6 q
4221.0⊕ 2415p 1055 F
1496.45h 6306.1q
750 F
CYCLONE
HOT AIR TO FLUIDISED BEDS
4900.8⊕ 500 F
RADIATION & UNACCOUNTED LOSS 23.8q
1 ST ECON
AIR HEATER
5397.6⊕
80 F
F.D FAN
PRECIPITATOR
STACK
300 F
6300.2⊕ GAS
816.7q GAS
2.0⊕ DUST
123.8⊕
MISC STEAM LOSSES 21.0⊕ 31.4q
HEAT LOSS 1.1q
4200.0⊕ 2315p 1050 F
1496.2h 6284.0q
STEAM TURBINE
648 OMW
4263.0⊕ 3000p 490 F
476.2h 2030.0q
ASH 247.4⊕
SENSIBLE HEAT LOSS 15.8q
CARBON IN ASH 146.3q
VAPOUR FROM COOLING 15.8⊕
FROM BEDS 54.6⊕
69.0⊕
ASH
HEAT LOSS 0.9q
HEAT LOSS 0.9q
3780.0⊕ 583p 1051 F
1545.55h 5842.2q
3780.0⊕ 613p 683 F
1340.85h 5068.4q
3780.0⊕ 620p 684 F
1341.1h 5069.4q
3780.0⊕ 575p 1050 F
1545.3h 5841.2q
DE-AERATOR HEATER
H.P HEATERS
L.P HEATERS
RIVER
DOMESTIC USE
SEWAGE TREATMENT
EFFLUENT TREATMENT
3.3⊕
42.0⊕
45.3⊕
3465.5⊕
3383.8⊕
63.0⊕
TO MAKE-UP TREATMENT PLANT
15.8⊕
TO ASH COOLING
EVAPORATION LOSS
COOLING TOWERS
BLOWDOWN 609.3⊕
91 F
59.0h
134276.0⊕
70 F
38.0h
MAKE-UP 63.0⊕ 4.9q
7526.2q
127562.0⊕ 91 F 59.0h
CONDENSER
127562.0⊕ 70 F 38.0h
4847.4q
6714.0⊕ 91 F
59.0h 396.1q
AUXILIARY COOLING
6714.0⊕ 70 F
38.0h 255.1q
CONDENSATE POLISHING PLANT
LEGEND
⊕ - FLOW - - - - - - K LB/HR
p - PRESSURE - - - P.S.I.A
F - TEMPERATURE - °F
h - ENTHALPY - - - B.T.U/LB
q - HEAT FLOW - - M B.T.U/HR
NOTE:-
ALL FLOWS ARE PER GENERATING UNIT
BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT
FLOW DIAGRAM
ATMOSPHERIC BOILER UNIT
epd
Engineering & Power Development Consultants Limited
in association with Combustion Systems Limited
N.T.S.
15283-101-012



EVAPORATION
TREATED EFFLUENT
COOLING TOWERS
BLOWDOWN
PRESSURISED, FLUIDISED COMBUSTION BOILER MODULE
RADIATION & UNACCOUNTED 46.1q
GAS CLEANING EQUIPMENT
AIR INTAKE
BY-PASS
GAS TURBINE
DUST & GRIT
BLOWDOWN GAS
ASH
AIR LOSS
HEAT LOSS 1.0q
HEAT LOSS 0.2q
MISC. LOSSES AS STEAM 14.4# 21.5q
STEAM TURBINE
475 MW
CONDENSER
MAKE-UP 14.4#
DE-AERATOR HEATER
AUX. COOLING
INTERCOOLER
MOIST GAS TO STACK
COAL DRYER
WET COAL
AIR
POWER TURBINE
147.5 MW
H.P. HEATERS
L.P. HEATERS
TURBO DRIVEN B F PUMPS
HIGH TEMP ECON.
LOW TEMP ECON.
BLOWDOWN GAS
TO STACK
MOIST GAS
INCLUDES DUST
ASH COOLING
MAKE-UP
DOMESTIC SUPPLY
RIVER
LEGEND
- FLOW - - - - K LB/HR
p - PRESSURE - - P S I A
F - TEMPERATURE - °F
h - ENTHALPY - - B T U/LB
q - HEAT FLOW - M BTU/HR
NOTE:- ALL FLOWS ARE PER GENERATING UNIT
BRITISH COLUMBIA HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT
FLOW DIAGRAM
COMBINED CYCLE UNIT
epd
Engineering & Power Development Consultants Limited
in association with Combustion Systems Limited
N T S
15283-101-013



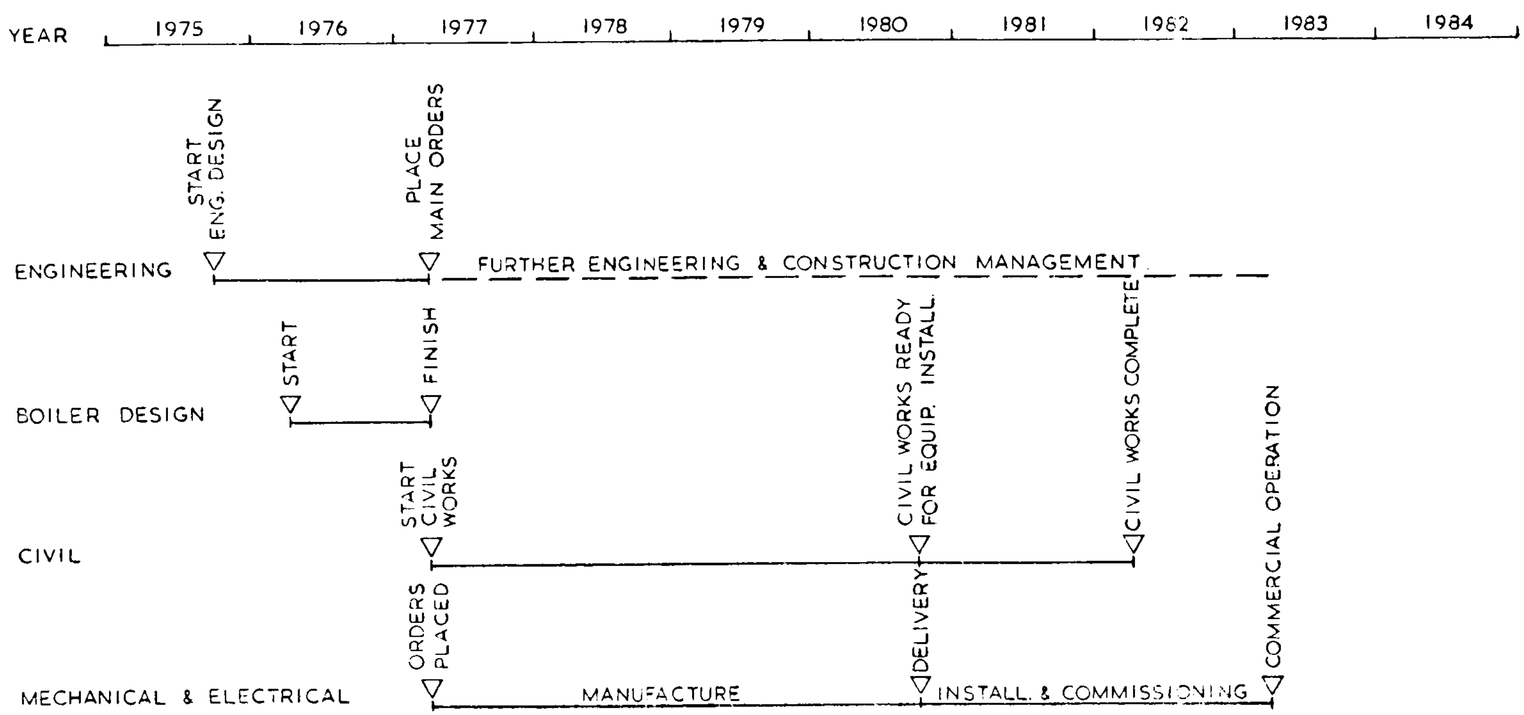
ATMOSPHERIC PRESSURE UNITS

SCHEDULE FOR PLANT WITH DEVELOPED DESIGN AVAILABLE

B.C. HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

SCHEDULE

13283-101-014



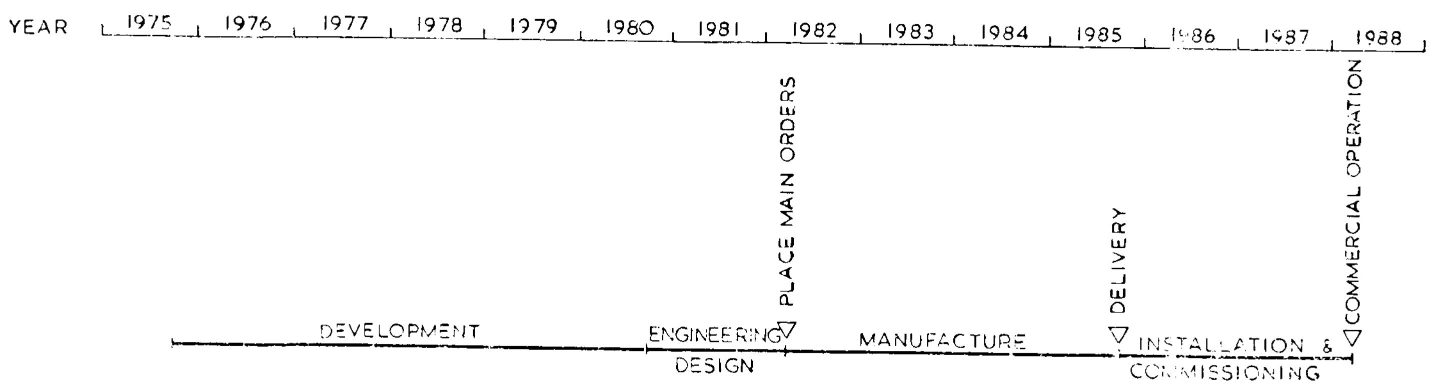
ATMOSPHERIC PRESSURE UNITS.

SCHEDULE FOR PLANT COMMENCED IMMEDIATELY.

B.C. HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

SCHEDULE

5203-101-015



PRESSURISED UNIT

SCHEDULE FOR EARLIEST COMMERCIAL OPERATION.

B.C. HYDRO & POWER AUTHORITY
HAT CREEK DEVELOPMENT

SCHEDULE

15283-101-016

STUDY B — COMBINED CYCLE GASIFICATION

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LIST OF FIGURES AND TABLES

1. SUMMARY

This report forms part of a comprehensive study commissioned by British Columbia Hydro and Power Authority for the comparative evaluation of traditional and new technologies aimed at exploiting the Hat Creek coal deposit for energy conversion.

This study — designated Study B — deals with the status and feasibility of coal gasification combined cycle technology for power generation purposes. It contains estimates and comparison of alternative methods for the generation of electricity in a combined cycle plant of 2000 MW nominal capacity using low-Btu gas derived from the gasification of Hat Creek coal.

This new technology requires an intermediate step in the conversion of the chemical energy of coal, namely the process of gasification. Through this step, however, coal is converted to a clean burning gas, which is suitable for use in high efficiency combined cycles, whereas coal itself is not. Increased performance and greatly reduced pollution are the benefits when compared to conventional, pulverized coal fired steam power plants.

Four systems are reported on. Three are being developed in the United States by General Electric, Westinghouse and United Technologies respectively. Work is in the pilot plant stage. The fourth system, developed in West Germany by STEAG, had reached commercialization after three and half years of demonstration at the Kellermann Power Station of STEAG in Lunen. STEAG's experience had demonstrated, what is also recognized by the U.S. developers, that the difficulties and risks with this new power generation technology are mainly associated with the coal gasification process itself. Both STEAG and General Electric are using the commercially mature Lurgi pressure gasification process with minor modifications to suit their special requirements. Westinghouse to a lesser extent, and United Technologies to a greater extent, are experimenting with new gasification technologies.

In accordance with the Terms of Reference, the emphasis has been placed on the STEAG-Lurgi system. As the suitability of the coal is of vital importance to the Lurgi process, for the purpose of this study analytical tests were performed on a small sample of Hat Creek coal by the Lurgi lab in Frankfurt, West Germany. In addition, the probable performance and cost of a Lurgi gasification system required for the 2000 MW plant, were evaluated separately. The results of these investigations on the coal are included in Appendix.

While the development of an independent Canadian coal gasification combined cycle technology is considered unnecessary and beyond the means of Canadian research and development capability, it appears that the introduction into Canada of a mature, indigenous coal conversion technology is both desirable and feasible.

In Section 10 a pilot project is outlined — modelled after the successful STEAG demonstration plant at Lunen — which could serve the dual purpose of providing the basis for a Canadian research and development facility as well as being a commercially useful power generating plant at the same time.

2. CONCLUSIONS

In the following, our findings are described on the suitability of Hat Creek coal, the current status of the various systems, their potential and development time scale, their estimated cost, performance, environmental effects and water requirements.

2.1 SUITABILITY OF HAT CREEK COAL FOR LURGI GASIFICATION

The coal, as represented by the sample analyzed in the Lurgi lab, was found to be of the lignitic type. It was deemed to be eminently suitable for Lurgi pressure gasification. The ash melting behaviour was found to be very favourable, resulting in low steam consumption. Disintegration and dust formation during carbonization was minimal. Reactivity was somewhat below the average for this kind of coal, requiring slightly higher than average oxygen (air) consumption. The relatively high ash content — 29.3% in the lab sample — is acceptable for the process, but it causes increased handling and processing costs.

Analytical lab test results do not yield accurate enough parameters for the design and optimization of the gasification process. Only actual full-scale gasifier tests can provide those data.

2.2 CURRENT STATUS OF THE SYSTEMS

At present, four coal gasification combined cycle systems are known to exist at various stages of development.

The STEAG combined cycle, integrated with a Lurgi gasification plant of 77 tons per hour capacity, has been adequately demonstrated on commercial scale by a 170-MW prototype unit installed at the Kellermann Generating Station of STEAG, in Lunen, West Germany. The individual components of the unit are large enough to have true validity in development and proving of the overall technology concerned, in order to apply this technology with confidence to units of larger size. Between its commissioning in February 1972 and October 9th, 1975, the demonstration unit had produced 590 million kilowatthours and had accumulated 6400 operating hours with the power plant and 4800 operating hours with the gasification plant. The unit normally is on peaking duty, requiring 40 minutes to reach full load after an 8 to 12 hour shutdown. Cold start requires two hours. The unit is equipped with auxiliary oil firing, enabling the power plant to operate independently from the gasification plant.

The 500-MW and 1000-MW units reported on herein are the results of STEAG's development work to date. The 500-MW unit is being currently designed. The components of this unit are either improved replicas or close extrapolations of the equipment used in the demonstration plant. The 500-MW unit has been optimized for STEAG's conditions and for their coal, which is almost twelve times as expensive as Hat Creek coal.

Work on the three U.S. systems is in the conceptual design and component development stage. General Electric appears to be the most advanced in the gasification plant, through their adoption of a modified Lurgi process. United Technologies and

Westinghouse are aiming at the building of demonstration units — of the size comparable to STEAG's — in the early eighties.

2.3 DEVELOPMENT OBJECTIVE, TIME SCALE, AVAILABILITY

Coal gasification technology has been in existence for many years for the production of town gas and synthesis gas, and commercially proven processes are available for those purposes. Most of the existing process technology is, of the atmospheric pressure type and of small capacity for recent North American requirements. Also, the production of low-Btu gas had not been attempted in the past for the purpose of power generation.

Similarly, combined cycle technique has been known and employed before, but with classical fuels such as oil and natural gas. Coal remained an untouchable fuel for this technique. Although coke oven gas and blast furnace gas as well as refinery gas have been used before in special boilers and gas turbines designed for these fuels, the use of low-Btu gas as a fuel for large capacity, high efficiency combined cycle units is a novel and commercially untried concept as yet.

The development of modern coal gasification combined cycle technology, therefore, requires development in many directions and as the survey shows, can follow different paths.

Development along these lines may proceed independently and concurrently in a co-ordinated fashion and at one time or another all building blocks may be assumed ready for integration. The short history of coal gasification combined cycle technology shows that this step also requires development. Furthermore, the scale-up from pilot plant size to demonstration plant size and to the eventual commercial plant size requires further adaptation and experimentation too.

It was mentioned before that the major difficulties and risks involved in the implementation of coal gasification combined cycle technology are likely to be associated with the gasification process itself. This seems to be the case with the STEAG system, which employs a standard steam cycle and an unfired gas turbine with very conservative gas inlet temperature. The gas is burned in the pressurized boiler, which is the special feature and undoubtedly the most successful component of the STEAG cycle.

The General Electric system, which is also based on Lurgi gasification technology, might encounter one or two additional difficulties, such as the development of the gas turbine combuster for low-Btu gas and the raising of the gas turbine inlet temperature. Westinghouse and United Technologies in addition will have to assume the difficulties, risks and the time lag involved in developing new gasification technologies as well.

It is often suggested or claimed, particularly by developers of new technology, that a complete new system can be conceived and implemented successfully on a commercial scale, if the technology of all the building blocks comprising the new system is already known and proven in isolated application. In STEAG's experience this is an illusion.

In the case of the STEAG-Lurgi technology, pressurized coal gasifiers existed and were in successful large scale commercial use. The same was true of gas and steam turbines and of gas cleaning technology in even more severe applications. Only the pressurized boiler might be said to have been a development or extension of an existing technology, and this item has posed only very minor problems in the integrated system. However, much time, money and intensive development work have proved necessary to marry these existing and proven diverse components to produce a successful, complete new power generation technology.

After many years of development and demonstration, STEAG's target of being able to commission their first 500-MW commercial unit in 1982, appears realistic in view of the results achieved to date.

U.S. developers cannot, in our opinion, offer commercial units of the 500-MW to 800-MW size before the late eighties, assuming that sufficient maturity — based on adequate and successful demonstration and testing — is a requirement for commercialization. General Electric's progress hinges on advanced gas turbine technology. G.E. expect to reach 2400°F by the mid-eighties and 3000°F by 1990. With this temperature, the efficiency of the G.E. system is expected to reach 41.8%, through the use of a 2400 psig/1000°F/1000°F steam cycle. United Technologies are also tied to this time frame, assuming that their gasification technology matures concurrently. Westinghouse's development schedule calls for the construction of a gasification pilot plant by 1980 for processing 60 tons of coal per hour, arranged to supply fuel gas to a separate combined cycle plant.

2.4 COST OF VARIOUS SYTEMS

The costing of the STEAG combined cycle units is based on their 500-MW design optimized for expensive German coal and includes 100% auxiliary oil firing equipment for the pressurized boiler. The costing of the Lurgi gasification plant comes from our independent study, based on processing Hat Creek coal and on complete desulphurization of all fuel gas produced. It is believed that the cost of a 2000-MW plant, optimized for cheap Hat Creek coal, without auxiliary oil firing and with partial treatment only of the fuel gas sufficient to satisfy environmental regulations, would be significantly less. The establishing of this cost is, however, beyond the scope of this study.

The estimate of the General Electric — Lurgi system, for an 800-MW unit, optimized for medium load range, moderate efficiency, low cost and for processing Montana sub-bituminous coal, is based on the company's publications. We have reassessed the cost of the gasification plant to suit Hat Creek coal.

United Technologies are currently working on the integration of their COGAS cycle with the modified KELLOGG molten salt gasification process, which is in its early pilot stage and is expected to be competitive with other gasification processes previously considered by the company. The costing of the system is based on fragmentary information obtained from the company, intended for publication in one of their recent, classified reports.

The cost estimate for the Westinghouse system is taken from a recent paper, co-authored by the company's engineers and presented at the University of Pittsburgh Second Annual Coal Gasification Symposium, in August 1975.

The costs of the four systems appear to be close to one another. Table No. 2.1 provides a comparison. As the basis of the individual estimates varies from detailed estimates (STEAG) to conceptional estimates (G.E.) and to allowances, especially for the gasification plant (U.T. and Westinghouse), the confidence in the figures must be related to the degree of maturity of the respective system. In spite of the obvious discrepancy which exists in this respect, a uniform contingency has been used for all systems, in accordance with our instructions.

2.5 PERFORMANCE

The efficiency and heat rate of the various sytems is shown in Table No. 2.1

Inherent in the efforts of coal gasification combined cycle development is the strive for high overall efficiency. This is done not only to offset the losses suffered in the gasification step, but also to counter rising fuel costs and to effect conservation of fuel resources.

As can be seen from the table, there is a wide spread in the heat rates, which the respective cycles can achieve, especially if the gas turbine inlet temperature is limited to 1950°F. The STEAG cycle can attain 40% overall efficiency with only 1560°F gas turbine inlet temperature.

By contrast, the Westinghouse cycle requires 2200°F temperature in order to attain 42% overall efficiency. Both STEAG and Westinghouse are taking advantage of an efficient steam reheat cycle within the combined cycle. General Electric and United Technologies seem to have optimized their system around a simple, almost rudimentary steam cycle, most likely to keep the costs down. It should be borne in mind that optimization is as much an economic exercise as a thermodynamic one and economics can greatly govern.

It is obvious that the STEAG system, which has been optimized for very expensive German coal, cannot present the most economic choice for a very cheap Canadian coal.

There are, however, two aspects which should merit special consideration in favour of a high efficiency cycle. Firstly, the 20% improvement between the efficiency of the G.E. cycle (33%) and the STEAG cycle (40%) can ensure a 20% longer life of the Hat Creek coal field to its exhaustion. Secondly, the amounts of gaseous emissions and other effluents are also reduced by an equal percentage throughout the life of the plant. Water consumption is also reduced. These are important and beneficial results in the domain of energy conservation and environmental protection, which however, cannot be quantified by using the economic criteria issued for the study.

In this context, of further interest could be the potential of the various systems in achieving higher efficiency. In this respect, STEAG's approach differs markedly from the others. The difference stems partly from thermodynamic design and partly from corporate objectives. STEAG, being a utility company, is interested only in finding, testing and using viable new methods of coal utilization and power generation. These new methods are to replace traditional technology, which is becoming uneconomical and burdensome in meeting environmental requirements, which the company has to face in Germany. The use of and reliance upon standard components are the cornerstone of STEAG's philosophy. Hence the use of the commercially available Lurgi process, a standard reheat steam cycle and the unfired gas turbine with moderate gas inlet temperature.

By contrast, U.S. development is being pioneered by manufacturers of power plant equipment, especially rotating machinery, who are vitally interested in the development of advanced components, such as high efficiency gas turbines. The key to high efficiency is high gas turbine inlet temperature. The cycles of both Westinghouse and United Technologies are currently based on 2200°F inlet temperature and the aim is to develop technology for 3000°F temperature. The integration of the diverse technologies, such as gasification and power generation, appears to be of secondary importance to these developers; they depend on other developers to deliver the technology ready to be used for integration. In brief, STEAG's approach is that of a user, whereas the approach of G.E., U.T., and Westinghouse is closer to that of a seller.

The thermodynamic difference boils down to circumventing the limitation posed by the permissible gas turbine inlet temperature. STEAG do circumvent it by the use of a heat exchanger — the pressurized boiler — ahead of their unfired gas turbine. In this manner, the fuel gas can be burned nearly stoichiometrically. It is not possible to do this with either of the U.S. schemes as yet, but the development tends towards this goal. High gas turbine inlet temperatures are therefore a must for the U.S. developers to close the efficiency gap. The gap will, however, not disappear, as the STEAG cycle also can derive benefits from increased gas turbine temperatures. In our opinion, the use of the pressurized boiler, which is available now, is an excellent starting point for building a high efficiency combined cycle and will remain so for sometime to come. Besides, the pressurized boiler itself can be considered as a valuable alternative to conventional boiler technology in other than combined cycle applications.

2.6 ENVIRONMENTAL EFFECTS

The use of gasified Hat Creek coal in combined cycle systems using present day technology would satisfy the most stringent emission regulations existing in Canada.

The sulphur that is present in the coal is one of the primary reasons that gasification processes are being developed. During gasification the sulphur is converted to hydrogen sulphide H_2S, which is subsequently removed from the gas during purification. The combustion of the purified gas is virtually free of sulphur oxides; the concentration would be about one hundredth of the acceptable level.

The nitrogen in the coal tends to gasify simultaneously with the carbon to form ammonia in the raw gas, which is largely removed during the water scrubbing process. Nitrogen compounds (NO_x) are formed during the combustion of the gas. The control of these compounds can be effected by suitable control of the combustion conditions. Because of the lower combustion temperature and the shorter residence time, lower NO_x formation is expected, than would result from direct combustion of coal in conventional equipment. The reduction is estimated to be one third to one half, by various authorities.

The emission of particulates is minimal, as the use of gas turbines requires a very high degree of particulate removal, which is achieved during the gas cleanup process.

The coal contains small quantities of chlorine. The majority of this chlorine should appear in the raw gas as HCl. This and other chlorine compounds are expected to be removed from the gas with the wash water.

A wide range of trace metals also occurs in the coal. The results of an EPA-IGT study show that a large portion of the trace metals should appear in the gasifier ash. The study also indicates that those elements which would appear in the gasifier effluent, would not survive the gas purification step.

Liquid and solid waste effluents from coal gasification combined cycle plants may be less of a problem than that resulting from the alternative of tail gas cleaning of flue gases following conventional coal combustion.

2.7 WATER REQUIREMENTS

The make-up water received for the 2000-MW STEAG type plant is 18, 500 US gpm., assuming six-fold concentraion in the cooling tower blowdown.

The gasification plant receives approximately 5300 US gpm. make-up water for steam raising, quenching and process cooling.

The total water requirement for the coal gasification combined cycle power plant of the STEAG-Lurgi type is, therefore, 23,800 US gpm.

The American systems would require less water for the power generating plant, because of their higher ratio of gas turbine capacity versus steam turbine capacity. The General Electrical system would require the least amount of water, as it employs the highest ratio.

TABLE 2.1
SUMMARY OF CHARACTERISTIC DATA
COAL GASIFICATION COMBINED CYCLE UNITS STUDIED

ITEM	DESCRIPTION		STEAG		G.E.	UNITED TECH.	WESTINGHOUSE
1	Unit Size	MW	500	1000	800	800	500
2	Steam Turbine Output	MW	369.5	739.0	199	223.8	224.7
3	Gas Turbine Output	MW	127.0	254.0	707.2	595.4	260.5
4	Auxiliary Power Req'd	MW	10.0	20.0	21.1	10.6	16.1
5	Power for Gasification Req'd	MW	3.0	6.0	incl.	71.6	incl.
6	Net Unit Output	MW	438.5	967.0	885.1	737.0	469.1
7	Steam Conditions psig/°F		2813/986/986		1250/900	1.250/816	1800/970/970
8	Gas Turbine Inlet Temp. °F		1562		1950	2200	2200
9	No. of Gas Turbines per Unit		1	2	8	8	8
10	Overall Net Heat Rate $\frac{BTU}{KWH}$		8465		10.300	10.869	8100
11	Overall Net Efficiency %		40.32		33.14	31.4	42.0
12	Steam/Gas Turbine Power Ratio		2.91		0.28	0.376	0.863
13	Fuel Cost Mills/KWH		1.983		2.413	2.546	1.898
	Specific Cost (Sept. 75) $/KW						
14	Contingencies included-"-		415.5		422.2	405.4	414.3
15	Engineering included -"-		448.8		456.0	437.7	447.4
16	Corp. Overhead -"- -"-		471.2		478.8	459.6	469.8

3. SUITABILITY OF HAT CREEK COAL FOR LURGI GASIFICATION

The results of the analytical laboratory tests, carried out by the Fuel R & D Laboratory of Lurgi Mineraloeltechnik Gmbh., in Frankfurt, are presented in Appendix.

The coal appears to be very suitable for the Lurgi process insofar as the ash characteristics are concerned. This quality and the caking behaviour are the most important factors in processing the coal through the Lurgi gasifier. The ash of some coals exhibits a very narrow temperature range between the ash softening point and the ash flow point. Such an ash may fuse or agglomerate rather rapidly under slightly changing temperature conditions, and may eventually block the grate. The addition of extra steam can reduce the temperature so that the ash cannot fuse or agglomerate, but then the ash turns into a powdery form resembling fine sand and causes difficulties in its removal from the gasifier. If the temperature is slightly increased by careful reduction in the steam flow, then the ash begins to fuse or sinter into larger pieces, which can be extracted by the rotating grate. Further reduction in steam, however, increases the temperature and may cause fusing of the whole mass; this should, of course, be avoided. Hence, coals with ash of a rather narrow temperature range are problematic ones for the Lurgi gasifier. The behaviour of the ash from the sample coal was found to be excellent by the tests.

The moisture content appears on the sheets in many forms. The equilibrium moisture, which is what the grains can absorb without moistening the surface, is an important feature. This is established in the lab over 36 hours of conditioning of the sample.

Lurgi had developed over the years special laboratory test facilities and methods. The results of such special tests have been compared to actual performances obtained in large scale equipment; i.e., Sasol, Westfield, and correlations were established which are of great value for the purpose of design. Lurgi can simulate, for instance, the behaviour of the coal as it is being heated rapidly from ambient temperature to 800 - 1000°C. Some coals disintegrate. This is the condition, however, which the coal is passing through when introduced into the gasifier.

The lab advises engineering on their findings and points out items of particular concern, which should be specifically considered in an eventual engineering study or design assignment, which may follow the lab tests.

At our special request, Lurgi Canada had obtained some preliminary estimates from Lurgi Frankfurt and telexed the following information on October 2nd, 1975:

"Coal tests for B.C. Hydro, Hat Creek

(a) Heating value of gas:
Approximately 176.49 Btu/SCF dry gas which equals approximately 143.53 Btu/SCF wet gas.

(b) Steam requirements per ton coal:
Approximately 2170 lbs. per metric ton of dry, ash free.

(c) Air requirements per ton coal:
Approximately 64950 SCF per metric ton of daf coal.

(d) Number of gasifiers for a 2000 MW plant; a combined cycle of said capacity (net output) would require 5 modules with 8 gasifiers each, which gives a total of 40 gasifiers.

All figures are a rough estimate only".

4. COAL GASIFICATION COMBINED CYCLES

The initial steps involved in converting coal to a low-Btu gas suitable for thermal utility fuel are similar to those required for producing a high-Btu gas but, instead of oxygen, air can be used. Thus, the gas produced is diluted with nitrogen, resulting in a lower heating value. The process itself is, however, less complicated and thus less expensive. In addition the gas cleanup is simpler.

The primary incentive behind low-Btu gasification is the conservation of more noble forms of energy such as oil and natural gas. However, in the domain of coal utilization for power generation, gasification also holds the promise of becoming a competitive, if not a superior alternative, to pulverized coal combustion which is burdened by the particulate and noxious gas emission problem. By gasification, coal can be converted into a clean fuel gas. So coal can be made to be an acceptable fuel for the gas turbine, a machine which requires a fuel of utmost cleanliness.

The advances made and still anticipated in coal gasification have revitalized the interest in the combined cycle technology. Options of more efficient power generation cycles have emerged, employing the gas turbine and the steam turbine in novel combinations and in larger power blocks, thus reducing the cost of the power generated.

4.1 COMBINED CYCLES

The preferred power cycle to be considered in connection with coal gasification is the combined cycle. The term relates to the thermodynamic integration of the conventional steam (Rankin) cycle with the combustion gas turbine (Brayton or Joule) cycle. Such an integrated cycle can harness a greater temperature spread than the conventional steam cycle, hence it can offer higher energy conversion efficiencies. The integration can, however, take so many forms that the term "combined cycle" used alone is an inadequate description.

Combined cycles may be divided into two generic categories, depending on how the steam generator is operated with respect to the gas turbine.

In the first category, the steam cycle is powered exclusively by the exhaust heat of the gas turbine. Steam conditions are inherently low and unit ratings moderate, requiring a large proportion of gas turbines in a large capacity scheme. Higher unit ratings and more advanced steam conditions including steam reheat can be achieved by the firing of supplementary fuel into the gas utilizing the residual oxygen in the gas turbine exhaust. As the additional heat so imparted is effective only through the steam cycle, the efficiency is not necessarily improved.

Most American manufacturers employ cycles of this category and use acronyms for cycle identification such as: STAG-General Electric; PACE-Westinghouse; COGAS-United Aircraft; etc. In all these cycles, the gas turbine cycle is thermodynamically superimposed upon the steam cycle.

In the second category, the gas turbine is powered by the combustion products of the steam generator which is pressurized for that purpose. The characteristic of this cycle is the large ratio of steam turbine power to gas turbine power. The steam cycle can also be supported by waste heat feedheaters utilizing the gas turbine exhaust. Here, therefore, the gas turbine cycle is thermodynamically interlocked within the steam cycle. The Steag cycle falls into this category.

In general, the overall efficiency of a given combined cycle depends on the efficiency of the contributing gas- and steam-turbine cycles. An approximate relationship for the first category, (simple waste-heat recovery type cycle) can be written as:

$$Yc = Yg + (1 - Yg)\,Ys\,\frac{Tx - Tc}{Tx - Ta}$$

where Yc = combined cycle efficiency
Yg = gas turbine efficiency
Ys = steam cycle efficiency
Tx = gas turbine exhaust temperature
Tc = stack gas (chimney) temperature
Ta = ambient temperature

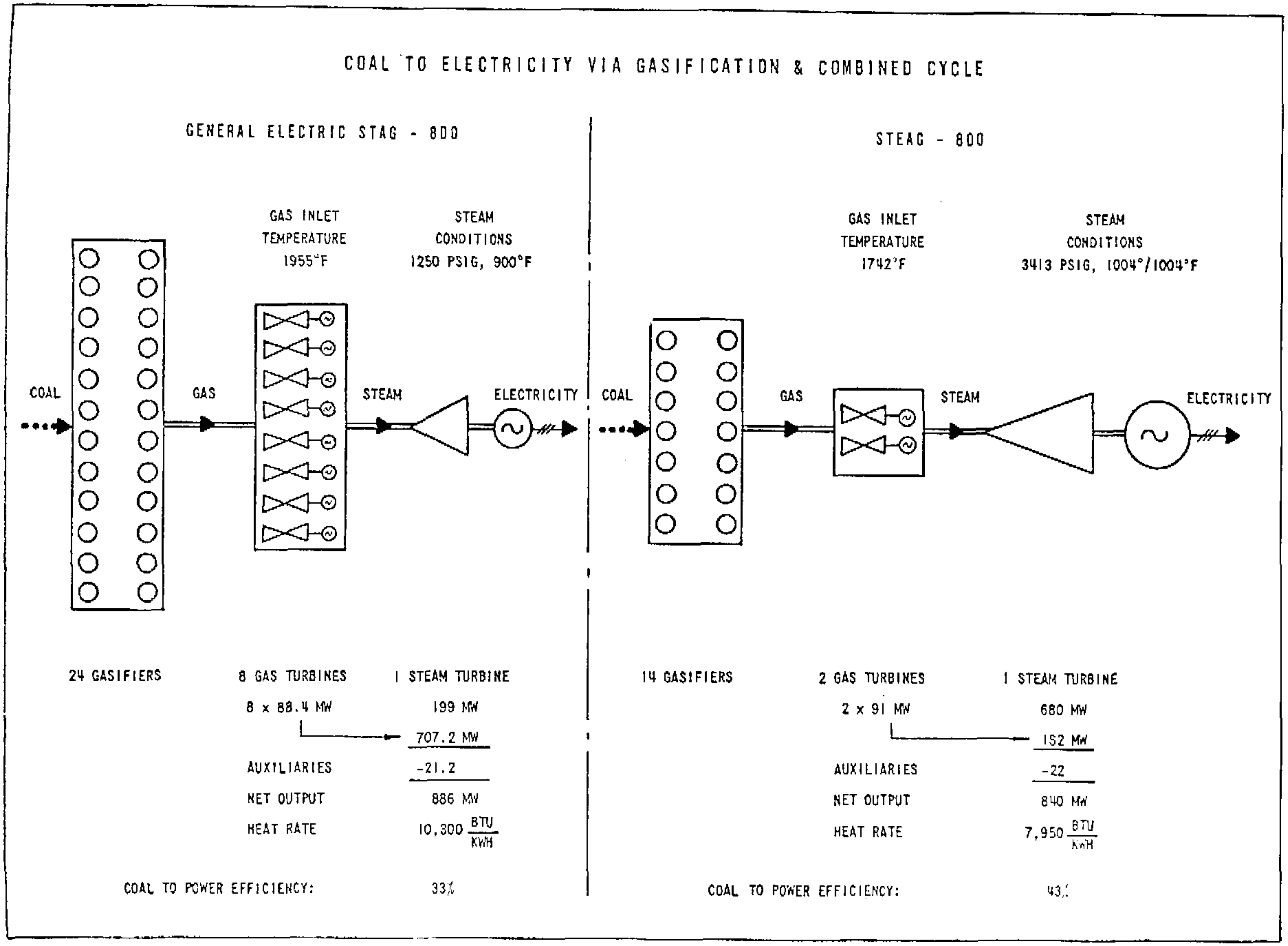
The predominant factor is the gas turbine efficiency. This depends on the gas turbine inlet temperature, which in turn, is limited to what the turbine blades can reliably endure. Stoichiometric combustion of the fuel would result in far too high a gas inlet temperature. In practice, therefore, both excess air and blade cooling are employed to moderate the temperature conditions. This technique requires additional air to be compressed, which results in a performance penalty. Accordingly, the gas turbine efficiency does not increase in proportion with increased turbine inlet temperature. However, the specific power — net power per unit air flow — does increase significantly. Specific power is a measure of the amount of power which a given gas turbine can produce. High specific power therefore translates into low specific costs: $/kW. This, of course, is a desirable result.

The efficiency of the steam cycle — in the waste heat recovery type combined cycle — is also a function of the gas turbine cycle parameters, primarily that of the gas turbine exhaust temperature. Because of the modest throttle steam conditions which can be obtained, the steam cycle configuration is usually quite simple and rather inefficient. Higher gas turbine inlet temperture, resulting in higher exhaust gas temperatures, can produce improvements.

The key to high overall efficiencies in this category of combined cycle plants is therefore the raising of the gas turbine inlet temperature. The realization of expected higher inlet temperatures will be the result of improvements in materials and cooling technique. This is the philosophy and approach of most of the American manufacturers who are active in developing combined cycles for coal gasification.

With combined cycles of the second category, acceptable gas turbine inlet temperatures can be readily achieved with only a minimum of excess air sufficient for stable combustion. Here, the pressurized boiler moderates the high gas temperature down to a level acceptable to the gas turbine. The temperature drop is converted to high pressure steam, with high superheat and reheat, thereby providing the basis for a highly efficient steam cycle. As regards the gas turbine, its power output increases dramatically, owing to the drastic reduction in compressor work or due to the increase in mass flow of combustion products in proportion to compressor air flow. This is the approach of STEAG in building their combined cycle, in which the gas turbine power output at 1.15 excess air increases to 160 percent of the normal rated capacity of the gas turbine employed in the cycle.

It is interesting to compare the two philosophies of combined cycle development at work as illustrated on the following diagram. Both cycles are using the LURGI coal gasification system for the production of fuel gas from coal. The information for the preparation of this diagram comes directly from publications by General Electric and STEAG respectively. No effort has been made to account for obvious differences in the coal, site conditions, assumptions, etc. — no effort has been made to reduce the two schemes to a common denominator. The comparison is therefore only superficial, but the difference in efficiency illustrates the superiority of the STEAG combined cycle approach. It should be noted that both the steam cycle and the gas turbine cycle operate at high temperatures than in the systems described later on. This is the reason for the higher efficiency — 43 percent versus 40.3 percent.


COAL TO ELECTRICITY VIA GASIFICATION & COMBINED CYCLE
GENERAL ELECTRIC STAG - 800
STEAG - 800
GAS INLET TEMPERATURE 1955°F
STEAM CONDITIONS 1250 PSIG, 900°F
GAS INLET TEMPERATURE 1742°F
STEAM CONDITIONS 3413 PSIG, 1004°/1004°F
COAL
GAS
STEAM
ELECTRICITY
COAL
GAS
STEAM
ELECTRICITY
24 GASIFIERS
8 GAS TURBINES
8 x 88.4 MW
707.2 MW
1 STEAM TURBINE
199 MW
AUXILIARIES -21.2
NET OUTPUT 886 MW
HEAT RATE 10,300 BTU/KWH
COAL TO POWER EFFICIENCY: 33%
14 GASIFIERS
2 GAS TURBINES
2 x 91 MW
152 MW
1 STEAM TURBINE
680 MW
AUXILIARIES -22
NET OUTPUT 840 MW
HEAT RATE 7,950 BTU/KWH
COAL TO POWER EFFICIENCY: 43%

4.2 THE LURGI COAL GASIFICATION PROCESS

As mentioned previously, two of the combined cycles dealt with in this report employ the Lurgi gasification process. It is therefore deemed appropriate to briefly describe the process in the following.

During the thirties it was discovered that coal gasification conducted under elevated pressure produced a methane-rich gas. This led to the recognition that gasification and partial catalytic enrichment could be combined into one operation under pressure. Practical tests also showed that the elevated pressure yields a high ration of hydrogen to carbon monoxide. Gasifier output was also found to increase with the square root of the gasification pressure. The advantages of pressure make themselves further felt in the subsequent purification and cooling of the gas. The hot potassium carbonate process for H_2S removal works best at around 300 psig pressure. Finally, the high pressure of the gas can be dropped to the eventual utilization pressure in an expansion turbine with the benefit of power production.

The Lurgi coal pressure gasification process is an example of a technology designed to take advantage of all these opportunities. At the time of its appearance this technology was viewed as being the link between the old water gas reaction and the modern conversion of coal to gases and oils.

The Lurgi reactor has evolved over the past forty years from its original version, designed for lignite and with a capacity of 280 MSCFH crude dry gas, to its present-day version, designed for nearly all kinds of coal for a capacity of 1300 to 1800 MSCFH dry crude gas.

The chemistry of the Lurgi process is rather complex. The reactions are interrelated as a result of high pressure gasification in a fixed bed reactor under countercurrent flow conditions. The path of the coal from top to bottom is as follows: prepared, sized coal is charged to the gasifier via automatically operated coal locks. A rotating distributor at the top spreads the coal over the cross section of the reactor. When gasifying caking coal, attachments to the distributor prevent the coal particles from fusing together as the coal reaches the plastic state during its temperature rise. The distributor can also serve for injecting the tar, where tar recycling is employed. The coal stays in the gasifier for about one hour. During this time it gradually descends to the ash grate while being constantly purged with the rising gasification agents and gas products. Meeting higher and higher ambient temperature on its descent, the coal is first dried, then devolatized, then gasified and finally the remaining carbon is burned to provide the reaction heat, to cover the heat loss of the gasifier and to heat the gas.

The ash is removed by a rotating grate. The amount of the ash removal can be controlled through the speed of the grate.

The gasification agents — steam and air (or oxygen) — are introduced through the rotating grate into the ash bed. The amount of the air is only sufficient to burn the last few inches at the bottom of the coal bed, which is the burning zone, where normal combustion occurs. Above this zone a reducing atmosphere prevails and further burning is precluded. The injection of steam has a tempering effect on the temperature: thermochemically through the endothermic reactions that occur between the reactants, and thermodynamically by absorbing heat and thereby becoming superheated. Not all the steam is decomposed or chemically converted; a portion passes through the bed and forms part of the raw off-gas. Adding to this vapour is the original moisture contained in the coal, which is driven off during the drying and devolitilization, and which can never reach deep enough zones to chemically react with its own coal. Sulphur is converted to hydrogen sulphide; H_2S. The raw off-gas contains H_2, CO, CH_4, which are the combustibles; CO_2, N_2, which constitute the inert ballast, and some tar, oil vapours, phenols, fatty acids, ammonia, H_2S, and traces of coal dust, which represent the impurities in the gas.

The subsequent purification treatment consists of quenching and washing the gas with a hot tar-water solution. This removes the solids, tar, alkali and chlorine, and increases the steam content of the gas to 50 percent H_2O. The carbonization products from the coal, such as tar oil, naptha, phenols, ammonia, etc., still remain in the gas as combustibles.

Sulphur (H_2S) removal follows, which involves heat consuming reactions. The heat is provided by the gas itself, as it is first cooled in a heat exchanger by cooling water, then the heated water returns its heat to the clean gas some process steps later.

In between, the gas is washed by counter-current contact with a potassium salt solution. This solution selectively removes H_2S, but leaves the other gaseous components and the hydrocarbons essentially unaffected. As a result, no dilution of the heating value accompanies the purification step. The gas leaves the system essentially with the same composition as it entered, but is now substantially free of H_2S. It is now ready for further processing as syngas, or for use as a fuel in a boiler or in a gas turbine. When so used, the gas, which is still at high pressure, can first be expanded in an expansion gas turbine, thus generating power to drive the air compressor which supplied the air to the gasifier.

The mass flow ration of gas to air is such that the expansion turbine can deliver surplus energy, the amount of which depends on the amount of the gasification reactants and the respective gas expansion — air compression ratios. This is a special advantage of the Lurgi pressurized coal gasification process for power generation and especially for combined cycle applications.

Besides its many advantages, there are some limitations associated with the Lurgi process. As the gasification depends on a permeable and uniformly resistant coal bed, the coal must be sized between 1/8" and 1¼", must not contain more than 7 percent fines, and must have a low swelling index. The ash melting point is another important factor.

Lurgi have successfully gasified some 70 kinds of coal, ranging from anthracite to lignite. High ash and water content does not present a technological problem but carries economic consequences. While the suitability of a given coal sample for gasification can be predicted by laboratory tests, it cannot be extrapolated to and guaranteed for bulk quantities. To positively prove suitability, it is prudent and recommended that full scale, actual gasification tests be carried out on representative and substantial amounts of the coal in question. Suitable test facilities are presently available only in Europe and South Africa.

5. PROBABLE RESULTS WITH LURGI GASIFICATION

In this section the report of our coal gasification consultant on the gasification of Hat Creek coal is summarized.

Hat Creek coal is a low grade subbituminous coal with high ash and moisture content. For the purpose of this analysis the average ash content of Hat Creek coal was assumed to be 25% and for the worst conditions an ash content of the coal of 31% was assumed. The coal analysis, taken as basis for estimates in the report, is given below:

TABLE 5.1
HAT CREEK COAL ANALYSIS

	AVERAGE	WORST
Ash	25.00%	31.00%
Moisture	20.41%	20.41%
Volatiles	27.30%	24.30%
Fixed Carbon	27.29%	24.29%
Total Organic	54.59%	48.59%
C	37.83%	33.67%
H	2.94%	2.62%
S	0.34%	0.24%
N	0.94%	0.84%
O	12.59%	11.21%
NHV	3457.23 kcal/kg	3086.12 kcal/kg

Ash fusability

softening	1362°C
melting	1482°C
fluid	1510°C

The coal gasification plant should be designed so that 2000 MW of power be generated in the STEAG combined cycle generation plant. The base assumptions are summarized below:

Plant capacity:	2000 MW STEAG power plant
Overall efficiency:	40%
Number of gasifiers:	20 gasifiers of 5.18 m I.D.
	35 gasifiers of 3.9 m I.D.
Gasifier throughput:	11,225.2956 Mcal/hr m^2 grate area
	3.236 ton/hr m^2 grate area

TABLE 5.2
HAT CREEK PROJECT, QUANTITIES PER TON FEED COAL AND PER HOUR FOR A 2000 MW OF POWER GENERATION

	AVERAGE CASE		WORST CASE	
	QUANTITY PER TON OF COAL	QUANTITY PER HOUR	QUANTITY PER TON OF COAL	QUANTITY PER HOUR
Mined coal ton	1.081	1344.514	1.081	1453.161
Feed coal ton	1	1243.769	1	1344.275
Ash to be disposed ton	0.250	310.942	0.310	416.725
Air compressed Nm^3	455.630	566,689.470	392.629	527,801.349
Steam total kg	570.734	709,861.256	548.181	736,906.014
Steam from jacket kg	189.775	236,036.262	169.375	227,686.578
Steam from turbine kg	380.959	473,824.995	373.588	502,205.009
Raw gas from Producer Nm^3	1340.448	1,667,207.669	1,218.333	1,637,774.594
Gas liquor total kg	416.129	517,568.350	412.156	554,051.007
Gas liquor to treatment kg	138.710	172,523.198	137.385	184,683.221
Gas liquor recycled kg	277.419	345,045.152	274.771	369,367.786
Gas from gas liquor Nm^3	6.784	8,437.729	7.496	10,076.685
Lock Hopper gas Nm^3	41.337	51,413.679	41.337	55,568.296
Air to gas purification Nm^3	8.874	11,037.206	8.524	11,458.600
Water from purification kg	3.238	4,027.324	4.686	6,299.273
Elemental Sulphur kg	3.06	3,805.933	2.75	3,696.273
Fine coal to boiler ton	0.081	100.745	0.081	105.886
Ammonia (as gas) kg	3.632	4,517.369	5.488	7,377.381

TABLE 5.3
COMPARISON STEAG COAL WITH HAT CREEK COAL

	STEAG COAL	HAT CREEK COAL
Moisture	12%	20.41%
Ash	20%	25.0%
NHV	5660 kcal/kg	3457.23 kcal/kg
NHV maf	8323.5 kcal/kg	6352.8 kcal/kg
Heat to steam in jacket	1.4% = 79.24 kcal/kg	189.775 kg = 110.069 kcal = 3.17%
CO_2 VOL% in gas	9.0	15.25
CO	10.2	15.89
CH_4	3.2	4.66
C_nH_m	0.3	0.56
H_2	16.0	21.51
N_2	25.6	21.56
H_2O	34.3	20.06
H_2S	0.2	0.13
Kcal/Nm^3 NHV	1200	1,817.7
Gas exit temp°C	620	265
Thermal efficiency of gasifier %	93.7%	93.4
Nm^3 gas/ton coal from and NHV in gas	4,716.7	1,628.675
Ton steam/ton coal required	0.671	0.5724
Kg steam/kcal NHV of coal	0.1185	0.1652
Kg air/Mcal	0.3225	0.1654

Quantities per ton feed coal and per hour time for a 2000 MW of power generation for average and worst case coal are summarized in the Table 5.2.

The thermal efficiency of the gasification system without the recovery of heat losses was estimated at 68.4% for the average conditions (25% ash coal) and 63.9% for the worst condition (31% ash). A large percentage of heat losses can be recovered; tar can be reinjected into the gasifier, lock hopper and other gas losses can be used as a fuel gas, part of the sensible heat from gases and ash can be recovered. With heat recovery the thermal efficiency of the gasification was estimated at 93.4 and 92.0% for the average and worst case respectively. STEAG's estimate was 94%. It was agreed to retain STEAG's figure, in order to preserve the validity of their charts and other supporting calculations. As the price of Hat Creek coal at $3 per ton results in a very low, almost insignificant component in the cost of power, the small difference in the efficiency cannot materially influence the final results. Table 5.3 compares the results of gasification of STEAG coal with the results for the Hat Creek coal.

The total capital for the gasification part of a combined gas-steam cycle power station, operating with the STEAG principle for a capacity of 2000 MW was estimated at $320.48 million. The breakdown of costs is given in Table 5.4.

TABLE 5.4
CAPITAL COSTS FOR THE LURGI GASIFICATION PLANT

Coal preparation and storage	$19.58 million
Gasification	$130.28 million
Utilities	$16.89 million
Purification	46.01 million
Gas Liquor Treatment	14.50 million
Process piping	20.23 million
Electrical	12.02 million
Instruments and Authorization	8.55 million
Insulation	5.78 million
Services	19.64 million
Rails and yard improvements	6.47 million
Civil works	16.64 million
Miscellaneous	3.89 million
Total	$320.48 million

Total manufacturing costs without credits for by products were estimated at $111.5 million or $6.941 per MWhr. With credits for sulphur, tar, tar oils, BTX and ammonium fertilizers, the total costs were estimated at $77.79 millions or $4.861 per MWhr.

At a meeting held in our offices on September 16, 1975, in the presence of the author and two engineers of STEAG, the results of the consultant's report were reviewed, and compared to those used by STEAG in their study. Good agreement was found as regards costs and overall gasification efficiency, STEAG's estimate being $340 million for the gasification plant using 4 meter dia. gasifiers. The estimate of the consultant was $320 million using 5 meter dia. gasifiers which are being developed. The difference in cost between the two versions was assessed as being $22 million in favour of the larger units. It was agreed to use the consultant's estimate as it provides for a breakdown and is cross-referenced with his report.

6. THE STEAG-LURGI SYSTEM

6.1 THE STEAG ROUTE FOR POWER GENERATION VIA COAL GASIFICATION

About ten years ago, in the era of cheap oil, it became apparent to STEAG's owners, the coal producers of the Ruhr, that coal would not be competitive with oil as a fuel to generate electricity unless the investment cost of coal-fired stations could be reduced or their efficiency increased, or both, STEAG, therefore, instituted a research program with these two aims in mind. The conclusion reached was that neither aim could be achieved with developments of the conventional technology of burning coal to generate electricity. Conventional technology has been pushed to its limits and, moreover, is today hampered by increasingly stringent requirements to minimize pollution of the environment.

STEAG then investigated the new technology of combined cycles, i.e., a combination of combustion (gas) turbines with steam turbines. This technology, however, requires that the fuel burned be sufficiently clean burning, a requirement that is met only by natural gas, light oils and certain heavy oils. STEAG's solution to this problem was to consider gasification of coal with subsequent clean-up of the gas produced.

STEAG investigated many available combined cycles, and selected a new combination as offering the greatest advantages. Virtually all combined cycles operating to date, except STEAG's, have the gas turbine exhausting into the boiler, whereas in the STEAG cycle the boiler exhausts into the gas turbine thus avoiding the combustion chambers for the gas turbines. A disadvantage of the STEAG cycle compared to some of the orthodox combined cycles is that in the latter the gas and steam turbines can be operated independently of one another, whereas in the STEAG cycle, this is not so. Some U.S. developers are moving in the same direction; G.E. for example. However, STEAG considered this disadvantage to be minor in the light of the ever increasing experience with, and reliability of, large industrial type gas turbines. As it is, the STEAG gas turbine operates at very modest gas inlet temperature, when compared to its U.S. counterparts.

However, the STEAG cycle requires that the steam boiler be operated with a combustion chamber pressure about ten times as high as that of a conventional boiler, the combustion chamber of which operates at virtually atmospheric pressure. This pressurized boiler is the only newly developed component in the STEAG system. Although it was expected that developmental troubles would be encountered with it, in the event the troubles actually experienced were very minor, and this item is now considered fully proven.

An important advantage of the pressurized boiler is its small size and low cost compared with a conventional boiler. Another advantage is that, even for the largest plants contemplated, it can be divided into units entirely shop fabricated. This avoids most of the problems, and the long erection periods associated with conventional boilers, which, in large sizes, have to be largely constructed at site.

STEAG investigated all methods and types of gasifier that were available and selected the Lurgi type of gasifier. This selection was based upon its relatively high efficiency, the large amount of experience available with Lurgi gasifiers compared to other types, and its being a pressurized unit. STEAG are continuing their study of all methods of gasifying coal and the selection made remains valid today and, it is considered, will remain so for the next decade at least.

Having reached decisions upon the type of combined cycle and type of coal gasifier, STEAG decided to build a commercial size engineering prototype plant. Engineering was started 1969 and the plant was built at STEAG's Kellerman Generating Station at Lunen in West Germany.

It is incidental to the original aims of STEAG in developing their system (i.e., low cost and high efficiency) that their system also offers minimal pollution of the environment as compared with a conventional plant. At the same time, the recent energy crisis has raised the price of fuels alternate to coal — such as oil and gas — and placed their future availability in doubt. In some circumstances, in Germany, a STEAG system plant already offers the only prospect of generating electricity from coal at acceptable cost while being able at the same time to meet current regulations in respect of pollution of the environment.

6.2 THE STEAG — LURGI PROTOTYPE PLANT

A system employing a STEAG combined cycle power plant integrated with a Lurgi coal gasification plant, is shown schematically on Diagram 72/6650. This diagram refers to the demonstration plant in existence since 1971 at the Kellerman station at Lunen, West Germany. The technical features of the plant are further described in detail in the pamphlets attached under appendix.

Its major parameters are:

Gas turbine capacity	74	MW
Gas turbine inlet temperature	1508	°F
Gas turbine outlet temperature	756	°F
Steam turbine capacity	96	MW
Steam generator output	749, 360	lb/hr
Steam conditions	1885	psia, 997°F
Feedwater temperature	628	°F
Number of gasifiers	4 + 1	standby
Coal input	77	ton/hr.
Combined cycle output	170	MW
Overall efficiency	36.9	%

This prototype plant was designed for maximum simplicity and to use standard components as far as practicable. Therefore, its design is not an optimum. However, in spite of this, its overall efficiency (coal to net power output) is quite comparable with that of an optimized conventional plant. An optimum design, now practicable in the light of experience gained at Lunen, would have an efficiency higher than that of a conventional plant.

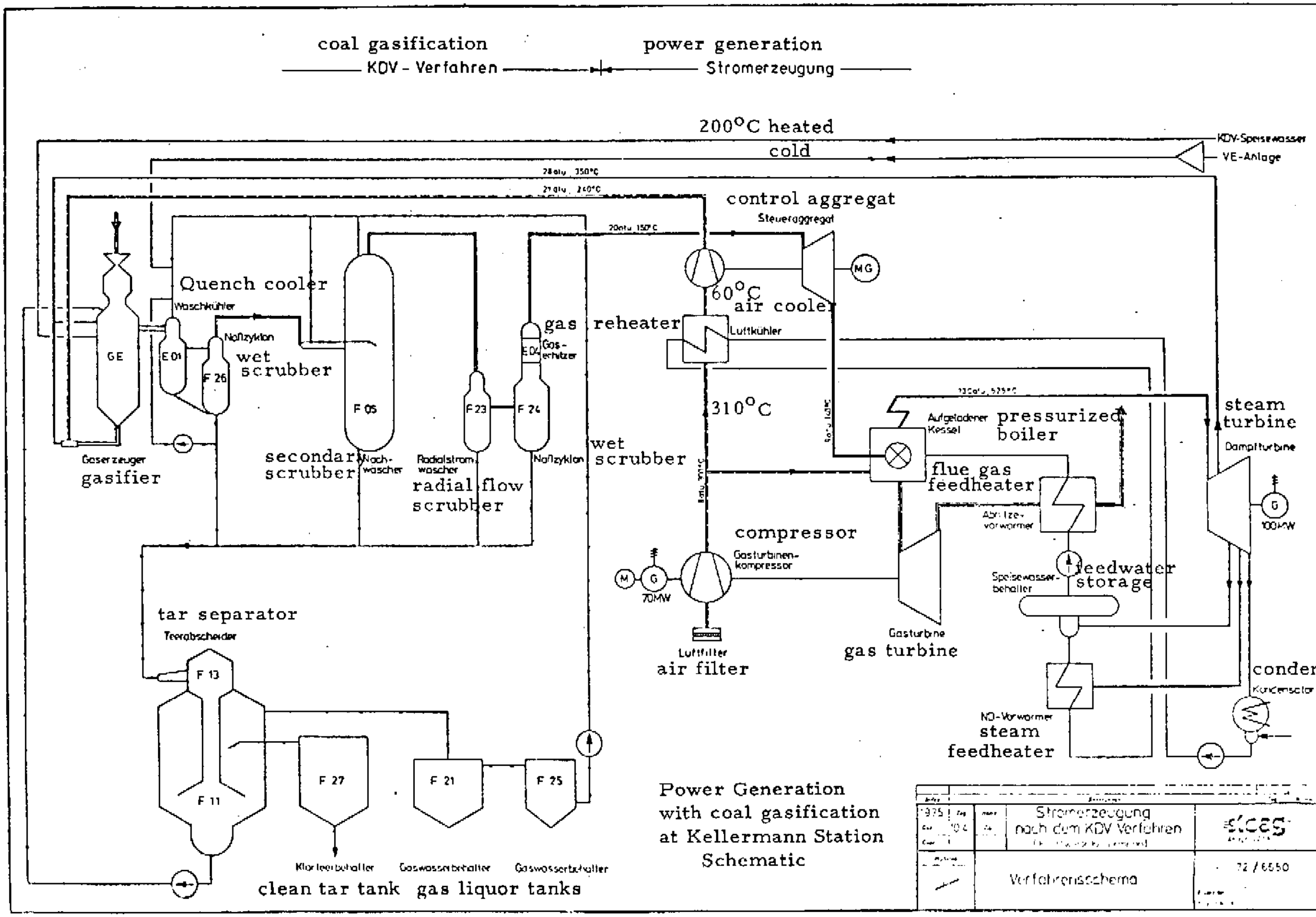
The steam turbine is a single housing, very robust unit, with fast starting capability. The turbogenerators are hydrogen cooled.

The design of the supporting gasification system deviates somewhat from the standard practice of Lurgi. There are five gasifiers, one of which acts as a standby. The raw gas produced at 1110°F temperature is first quenched and cooled to about 350°F and becomes saturated with water vapour. Droplets of water are removed in a wet scrubber. The major portion of the effluent from the quench cooler and the scrubber, which contains tar and coal dust, is recycled to the wet scrubber. The balance is sent to the tar separation system. The gas leaving the wet scrubber is further washed in two subsequent stages and water droplets carried over are removed in a second wet scrubber. If and when necessary, the saturated clean gas can be dried or superheated prior to admission to the gas burner. The effluent from the washing process is also sent to the tar separation system, where by sedimentation the tar is separated from the gas liquor. The tar removed from the separator is recycled to the gasifier via an intermediate storage tank. The gas liquor is also stored intermittently and then returned to the scrubbers. The water absorbed by and saturating the gas is made up from the demineralized water storage system.

The plant is equipped with an experimental sulphur removal system, which presently handles only a portion of the total gas flow. The plant, however, complies with the regulations in force as regards air pollution by particulates, oxides of nitrogen, as this is inherent in the plant design. Sulphur emission is also within the limits permitted. In fact, the purity of the gas stipulated by the gas turbine manufacturer far exceeds the environmental requirements; the sulphur in the gas however is of little concern to the turbine manufacturer.

STEAG found the capital cost of this type of plant to be lower than the capital cost of a conventional plant, if both plants are required to comply with today's regulations as regards air pollution and particularly sulphur emission. The major reasons for the lower cost of the STEAG plant are the advantages inherent in the pressurized boiler, the exploitation of the gas turbine capability, the greater degree of shop fabrication of plant components with consequently shorter erection time and the application of gas clean-up to the fuel before burning it rather than to the flue gases after combustion.

The individual components of the Lunen plant are large enough to have true validity in development and for proving of the overall technology concerned in order to apply this technology with confidence to units of larger capacity. The Lunen plant has undergone intensive testing during the past four years and solutions to all problems have been found. STEAG have the backing of the German Federal and State Governments to develop and build 500 MW and 1000 MW plants for operation in the eighties. The 500 MW plant has been optimized and engineering is progressing, so that the first unit can commence operation by 1981. The 1000 MW plant, using multiples of the components required for the 500 MW plant will have to await the proven reliability of the 700 MW steam turbine, which is expected in the early eighties.


coal gasification
KDV - Verfahren
power generation
Stromerzeugung
200°C heated
cold
KDV-Speisewasser
VE-Anlage
28 atu 350°C
21 atu 240°C
20 atu 150°C
control aggregat
Steueraggregat
MG
Quench cooler
Waschkühler
Naßzyklon
wet scrubber
GE
E 01
F 26
F 05
F 23
F 24
E 04
gas reheater
Gas-erhitzer
60°C
air cooler
Luftkühler
310°C
Gaserzeuger
gasifier
secondary scrubber
Nach-wäscher
Radialstrom-wäscher
radial flow scrubber
Naßzyklon
wet scrubber
13 atu 525°C
Aufgeladener Kessel
pressurized boiler
steam turbine
Dampfturbine
flue gas feedheater
Abhitze-vorwärmer
G
100MW
compressor
Gasturbinen-kompressor
M
G
70MW
feedwater storage
Speisewasser-behälter
tar separator
Teerabscheider
F 13
F 11
F 27
F 21
F 25
Luftfilter
air filter
Gasturbine
gas turbine
condenser
Kondensator
ND-Vorwärmer
steam feedheater
Klarteerbehälter
Gaswasserbehälter
Gaswasserbehälter
clean tar tank
gas liquor tanks
Power Generation
with coal gasification
at Kellermann Station
Schematic
Stromerzeugung
nach dem KDV Verfahren
Verfahrensschema
72 / 6550

6.3 THE 500 MW STEAG KOMBI-BLOCK

STEAG refer to their combined cycle unit as KOMBI-BLOCK and use the abbreviation: KDV for pressurized coal gasification, namely the Lurgi gasification system.

Accordingly "KDV 400 MW KOMBI-BLOCK" stands for the designation of a combined cycle unit integrated with a coal gasification plant, with a nominal power output of 400 MW. The actual capacity of the unit is approximately 500 MW. The drawings and diagrams prepared for this study and included under appendix bear this designation.

The basis of the KOMBI-BLOCK is the availability of a proven and standard steam turbine and a gas turbine. These form the backbone of the combined cycle unit. Both turbines are standardized to a greater or lesser degree by the practice of the power industry. The use and reliance of standard components is one of the cornerstones of the STEAG combined cycle philosophy. The capability of these two major components — once selected — therefore define the power output of the combined cycle unit and determine the requirements for supporting facilities, such as the steam generation plant and the gasification plant. The pressurized boilers and the gasifiers can then be provided to suit these requirements in the form of multiple units. The waste heat recovery unit — being a conventional component — can also be built in the required size.

The major task henceforth entails the integration of the two turbo-units in such a manner, that both turbines operate at their maximum capacity. This determines the major parameters for the design of the pressurized boiler, as the available energy from the fuel gas is essentially split into enthalpy for the steam cycle and enthalpy for the gas turbine cycle in the pressurized boiler. The pressurized boiler converts the chemical (and some sensible) energy of the fuel gas into enthalpy of the flue gas, a part of which is imparted directly to the steam cycle and the balance be applied directly to the gas turbine cycle and indirectly back to the steam cycle in the required proportion.

Also, heat and material export and import exist between the gasification plant and the power plant in order to minimize all possible overall losses and to reduce irreversibilities within each system as much as possible.

The 400 MW KOMBI-BLOCK has been optimized by Kraftwerk Union, Erlangen, under contract to STEAG. The basis of the optimization is the performance of steam and gas turbine, respectively, which are also built by KWU, and many of which are in operation at various power stations in Germany. The other essential components of the 400 MW commercial unit are careful and close extrapolations of the equipment employed in the 170 MW demonstration plant. So, for instance, the pressurized boiler will have a diameter of 16.4' as against 10' at Lunen and its height will increase by 21'. The capacity of the new boiler will be three times that of the Lunen prototype, which was found to have been sized very conservatively originally. The gasifiers will have a diameter of 13.2' as against 11.6' at Lunen and the height of the reaction zone will be increased by 6.6'. STEAG have already ordered one new gasifier for the purpose of experimentation prior to implementation.

The design of the boiler plant is being presently done by DURR' Ratingen in collaboration with the Benson Division of KWU in Erlangen. The boilers will be built in two 50 percent modules. The design of the boiler plant is expected to advance to such a stage in October, 1975, that firm price quotations can be obtained.

The following are the main data of the unit at full capacity:

Steam generators:	steam flow	2, 292, 784	lb/hr.
	steam pressure	2, 827.5	psia
	steam temperature	986	°F
	reheat pressure	609/536	psia
	reheat temperature	986	°F
Gas turbine:	power output	127	MW
	flue gas quantity	1303	lb/sec.
	flue gas inlet temperature	1562	°F
Steam turbine:	power output	369.5	MW
	condenser pressure	2"	Hg approx.

The design is based on the use of a steam turbine driven feedpump, which receives steam at 103 psia from the main turbine. The output of this steam turbine is 12 MW. The performance data of the units are shown in the following tabulation.

PERFORMANCE DATA WITHOUT AND WITH GASIFICATION

BOILER LOAD	%	100	70	40
Steam Turbine Output	MW	369.5	263	153
Gas Turbine Output	MW	127	102	67
Gross Output*	MW	496.5	365	220
	(WITHOUT GASIFICATION)			
Auxiliary Power Required	MW	10	8.7	7.2
Net Output	MW	486.5	356.3	212.8
Net Efficiency (Fuel Gas to Power)	%	43.16	41.68	37.92
Net Heat Rate	Btu/KWH	7908	8188	9000
	(WITH GASIFICATION)			
Power Required for Gasification	MW	3	3	3
Net Output	MW	483.5	353.3	209.8
Gasification Efficiency (Coal to Fuel Gas)	%	94	94	94
Overall Net Efficiency	%	40.32	38.85	35.14
Overall Net Heat Rate	$\frac{\text{Btu}}{\text{KWH}}$	8465	8785	9713
Heat Input	10^6 Btu/HR	4102.3	3111.3	2042.6

*The gross output and efficiency figures are inclusive of the steam turbine driven feedpump power requirement.

6.4 THE 1000 MW STEAG KOMBI-BLOCK

The next larger unit size which STEAG will be developing is the 1000 MW KOMBI-BLOCK. It employs a 700 MW conventional steam turbine and two gas turbines identical to that used in the 500 MW unit for a total output of 1016 MW gross and 966 MW net respectively after auxiliaries and the power requirements of the supporting coal gasification plant are accounted for.

The components of this unit are duplicates of the 500 MW plant, but the number of the gasifiers and the pressurized boilers is double. The detail development and optimization of this unit size hinges on the choice of a 700 MW steam turbine, different makes and/or types of which STEAG are presently projecting for conventional power plants.

It seemed appropriate for this study to include the 1000 MW KOMBI-BLOCK as an alternative to the 500 MW units. For the moment, the performance of the 1000 MW unit is assumed to be equal to that of the 500 MW unit and the specific cost is estimated to be only marginally less.

The cost estimate and the financial evaluation have therefore been based on two 500 MW units and the one 1000 MW unit composing the 2000 MW plant. The plant, however, can be built with four 500 MW units without invalidating the financial results presented in the report.

6.5 DESIGN, PART LOAD PERFORMANCE AND LAYOUT OF THE KOMBI-BLOCKS

6.5.1 DESIGN FEATURES

The following is a brief description of the major components of the combined cycle unit.

6.5.1a PRESSURIZED BOILER

Two required, each 50 percent capacity, for the 500 MW unit and four for the 1000 MW unit arranged on alternate sides of the gas turbine, connected by co-axial ducts; the inner duct carrying the gas the outer duct carrying the air. Firing capacity of two units together: 377 x 10^9 Btu/hr. Gas burners operated with 1.15 excess air, inclusive of air leakage loss. Each boiler approximately 72' high, 16.4' ϕ. Manufacturer: Balcke-Durr, Ratingen in collaboration with the Benson Division of Kraftwerk Union, Erlangen, West Germany. Furnace pressure 145 psia; draft loss 118" W.C. Live steam conditions: 2828 psia, 986°F; steam flow rate: 2, 292, 784 lb/hr. Feedwater inlet pressure 3580 psia, 608°F. The boilers are fitted with auxiliary oil firing equipment for 100% steam capacity. The cost of this, complete with piping and controls, is included in the cost estimate.

6.5.1b GAS TURBINE

One required for the 500 MW unit and two for the 1000 MW unit, arranged between the pressurized boilers. Manufactured by Kraftwerk Union, Erlangen, West Germany, type V93, for the following design conditions.

Maximum power output	127	mw
Air Flow	1058.2	lb/sec.
Exhaust Gas Flow	1222.6	lb/sec.
Pressure Ratio	10	
Gas Inlet Temperature	1562	°F
Exhaust Gas Temperature	842	°F

6.5.1c STEAM TURBINE

Standard design by KWU, three cylinder arrangement for the 500 MW unit with the following design parameters:

Steam Pressure	2683	psia
Steam Temperature	977	°F
Generator Output	369.5	MW
Condenser Pressure	2"	Hg (approx.)
Cooling Water Temperature	71.6	°F
Reheat Pressure	609/539.4	psia
Reheat Temperature	591/977	°F

The type and make of the 700 MW turbine required for the 1000 MW KOMBI-BLOCK is subject to further investigation and operating experience.

6.5.1d FEEDHEATING SYSTEM

In accordance with Drawing No. 30 17 0995/8.020 for the 500 MW unit.

Parallel flow through steam feedheaters and flue gas feedheaters controlled by three-way valves in response to flue gas temperature.

6.5.1e CONTROL AGGREGATE

In the integrated coal gasification — power generation process steam and air are provided from the power generation system to the gasification system.

For technological reasons the two systems operate at different pressures: the gasification system at 300 psia and the combined cycle at 150 psia. The air for gasification is provided by the main compressor at 150 psia, is intercooled and then compressed to 300 psia by a booster compressor. This compressor is driven by an expansion turbine.

The gas produced by the gasifiers, neglecting the pressure drop, is also available at 300 psia. The gas pressure is reduced via the expansion turbine to 150 psia, which is the operating pressure of the pressurized boiler.

Both machines, the booster compressor and the expansion turbine have a common shaft and form the so-called control aggregate. Under normal conditions, the power input to the compressor and the output of the expansion turbine balance each other.

The primary purpose of the aggregate is to save compressor work. However, the functional role of the aggregate is to control the performance of both processes: the gasification and the power generation. The expansion turbine is equipped with throttling valves by which the required amount of fuel gas, and hence the power output of the combined cycle, can be controlled.

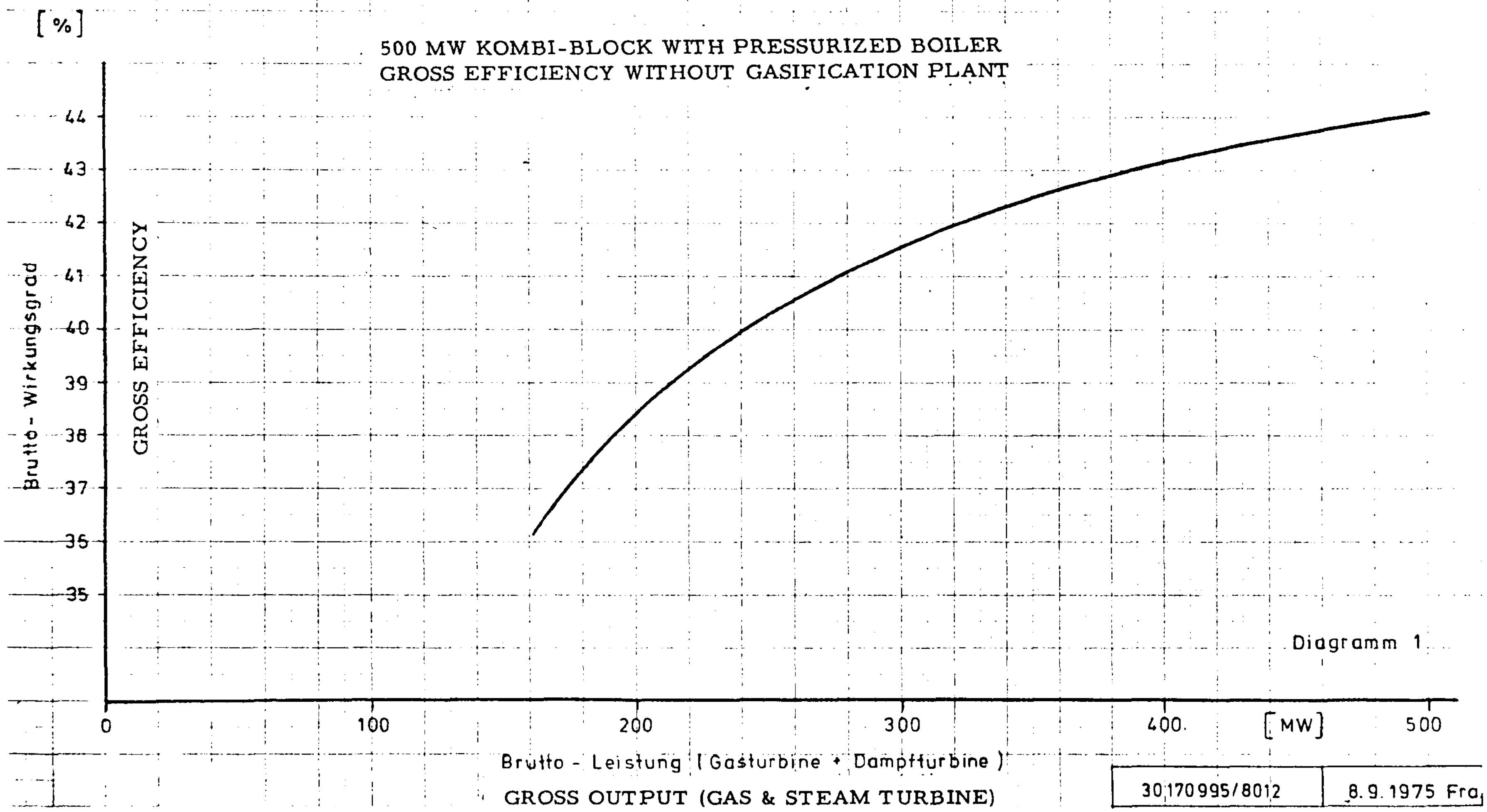
6.5.2 PART-LOAD PERFORMANCE

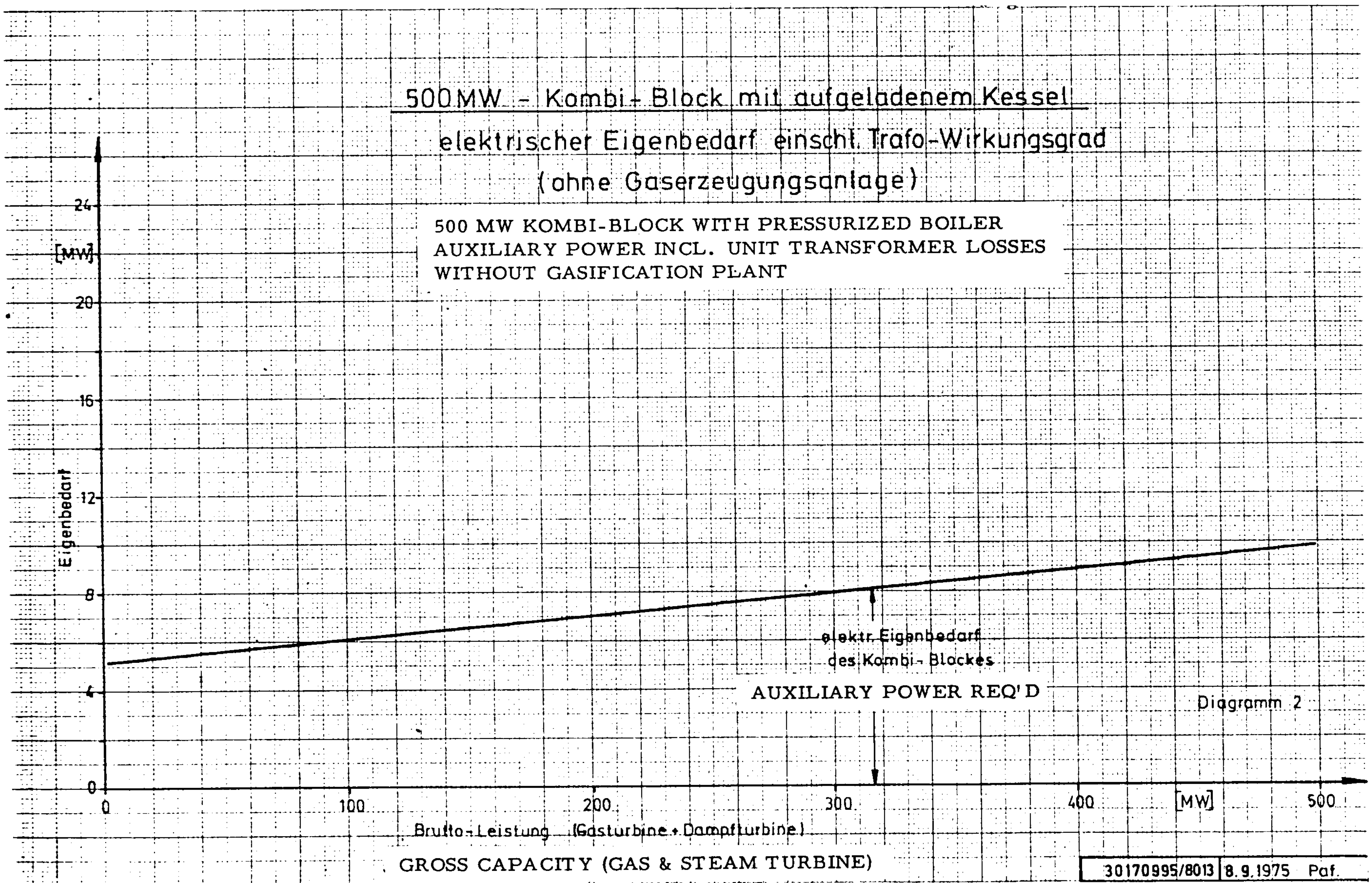
The performance of the 500 MW KOMBI-BLOCK is illustrated on the diagrams listed below, which follow overleaf:

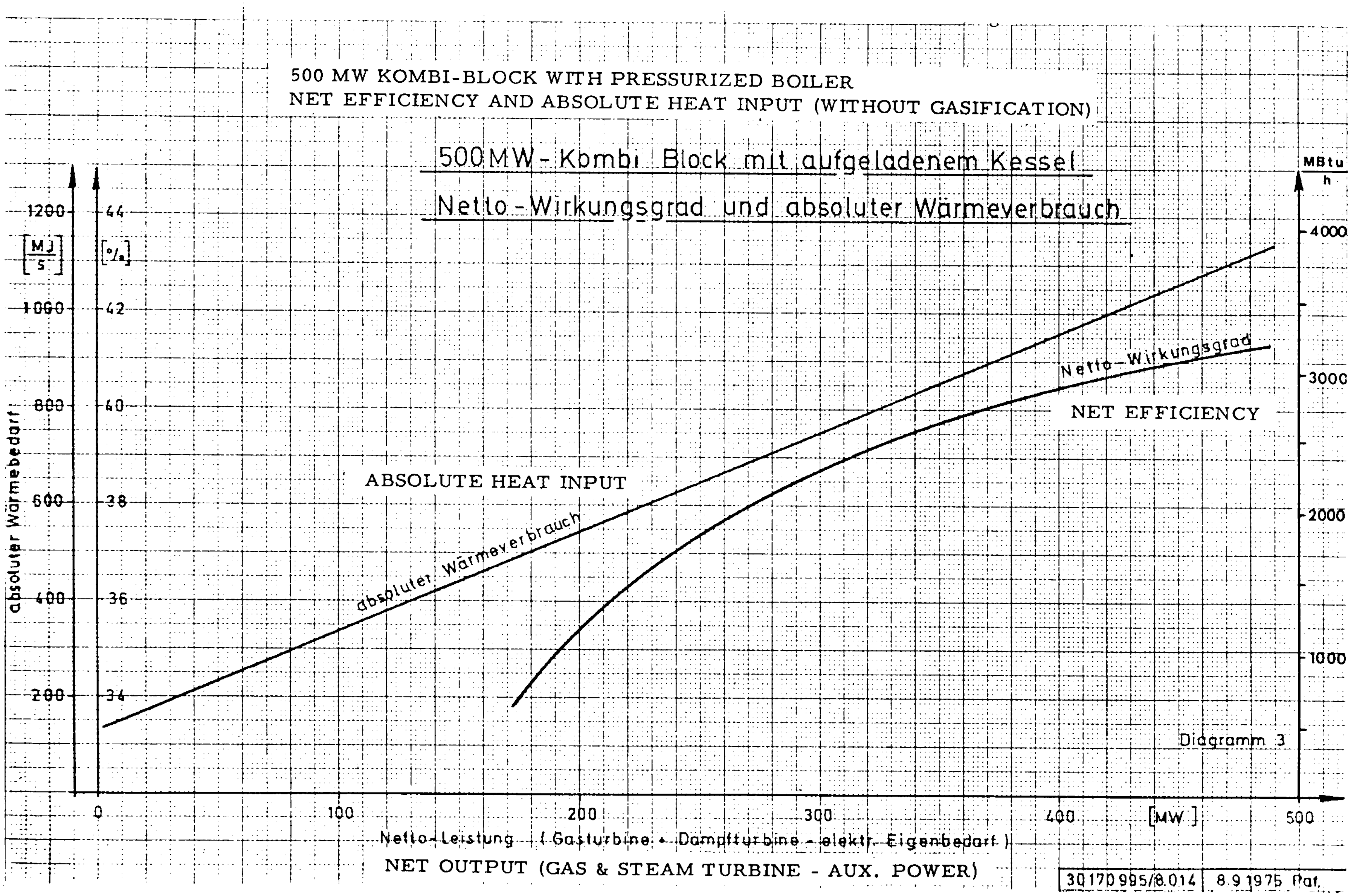
No.	Title
Diagram 1	500 MW unit — Gross efficiency without gasification
Diagram 2	500 MW unit — Auxiliary power required without gasification
Diagram 3	500 MW unit — Net efficiency and heat rate without gasification
Diagram 4	500 MW unit — Power required for plant and gasification
Diagram 5	500 MW unit — Net efficiency and heat rate with gasification

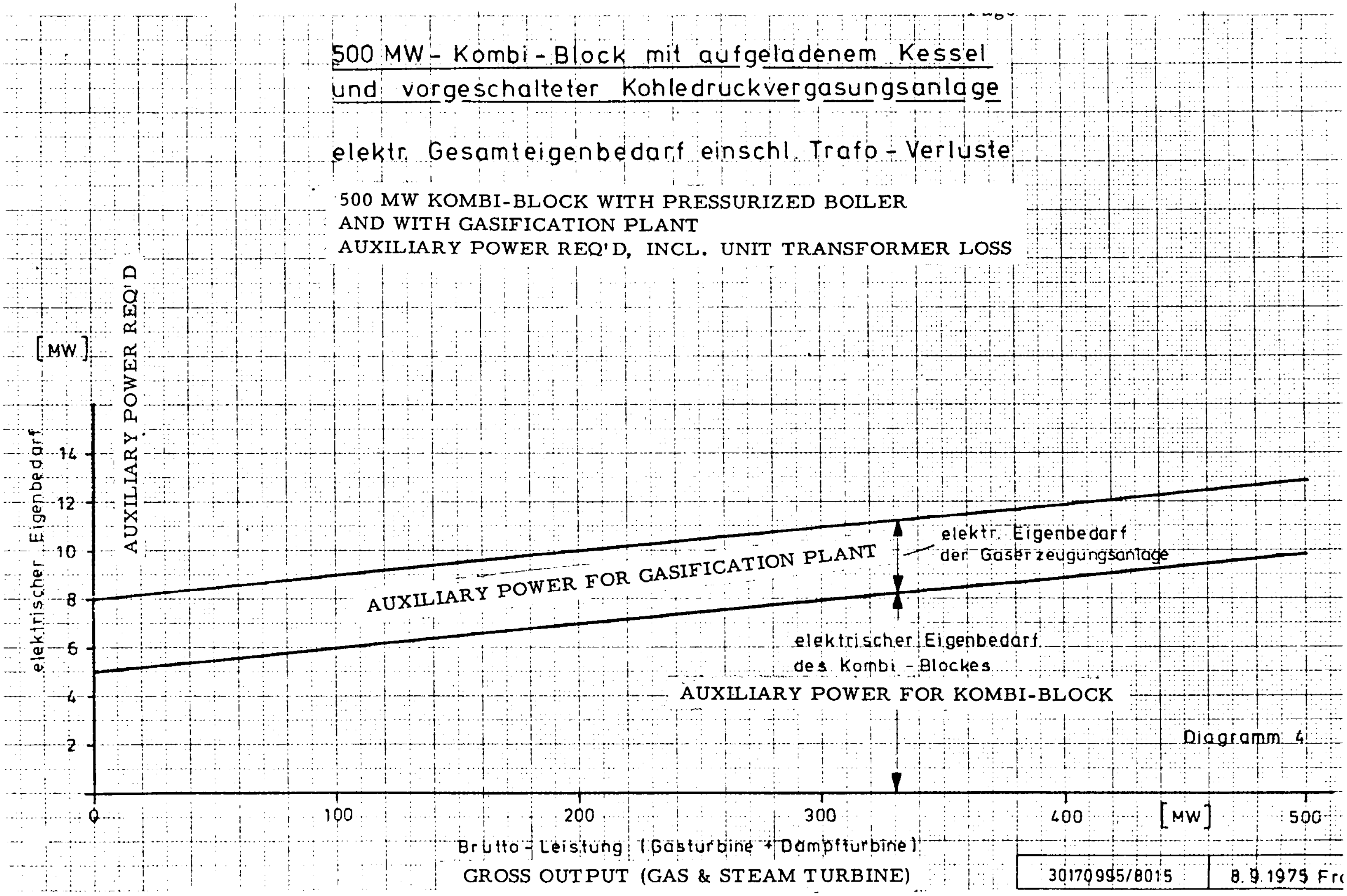
500 MW - Kombi - Block mit aufgeladenem Kessel.

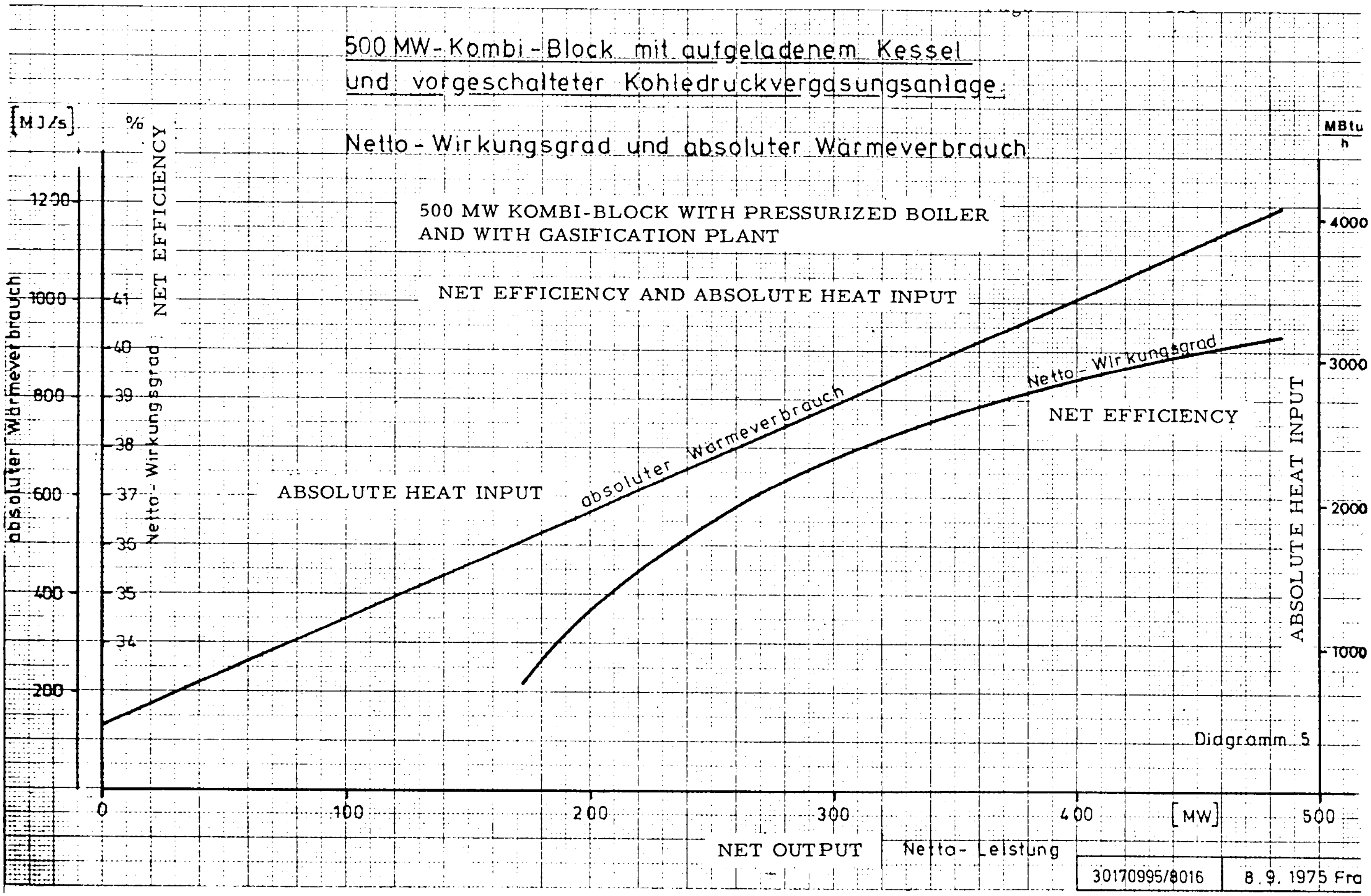
Brutto - Wirkungsgrad
(ohne Gaserzeugungsanlage)




500 MW - Kombi-Block mit aufgeladenem Kessel
elektrischer Eigenbedarf einschl. Trafo-Wirkungsgrad
(ohne Gaserzeugungsanlage)
500 MW KOMBI-BLOCK WITH PRESSURIZED BOILER
AUXILIARY POWER INCL. UNIT TRANSFORMER LOSSES
WITHOUT GASIFICATION PLANT
[MW]
Eigenbedarf
0
4
8
12
16
20
24
elektr. Eigenbedarf
des Kombi-Blockes
AUXILIARY POWER REQ'D
Diagramm 2
0
100
200
300
400
500
[MW]
Brutto-Leistung (Gasturbine + Dampfturbine)
GROSS CAPACITY (GAS & STEAM TURBINE)
30170995/8013
8.9.1975
Paf.


500 MW KOMBI-BLOCK WITH PRESSURIZED BOILER
NET EFFICIENCY AND ABSOLUTE HEAT INPUT (WITHOUT GASIFICATION)
500 MW-Kombi Block mit aufgeladenem Kessel
Netto-Wirkungsgrad und absoluter Wärmeverbrauch
absoluter Wärmebedarf
[MJ/s]
[%]
1200
1000
800
600
400
200
44
42
40
38
36
34
MBtu/h
4000
3000
2000
1000
Netto-Wirkungsgrad
NET EFFICIENCY
ABSOLUTE HEAT INPUT
absoluter Wärmeverbrauch
Diagramm 3
0
100
200
300
400
500
[MW]
Netto-Leistung (Gasturbine + Dampfturbine - elektr. Eigenbedarf)
NET OUTPUT (GAS & STEAM TURBINE - AUX. POWER)
30170995/8.014
8.9.1975 Raf.


500 MW- Kombi-Block mit aufgeladenem Kessel
und vorgeschalteter Kohledruckvergasungsanlage
elektr. Gesamteigenbedarf einschl. Trafo-Verluste
500 MW KOMBI-BLOCK WITH PRESSURIZED BOILER
AND WITH GASIFICATION PLANT
AUXILIARY POWER REQ'D, INCL. UNIT TRANSFORMER LOSS
[MW]
elektrischer Eigenbedarf
AUXILIARY POWER REQ'D
14
12
10
8
6
4
2
0
100
200
300
400
500
[MW]
elektr. Eigenbedarf
der Gaserzeugungsanlage
AUXILIARY POWER FOR GASIFICATION PLANT
elektrischer Eigenbedarf
des Kombi-Blockes
AUXILIARY POWER FOR KOMBI-BLOCK
Diagramm 4
Brutto-Leistung (Gasturbine + Dampfturbine)
GROSS OUTPUT (GAS & STEAM TURBINE)
30170995/8015
8.9.1975 Fr


500 MW-Kombi-Block mit aufgeladenem Kessel
und vorgeschalteter Kohledruckvergasungsanlage.
Netto-Wirkungsgrad und absoluter Wärmeverbrauch
500 MW KOMBI-BLOCK WITH PRESSURIZED BOILER
AND WITH GASIFICATION PLANT
NET EFFICIENCY AND ABSOLUTE HEAT INPUT
[MJ/s]
%
NET EFFICIENCY
absoluter Wärmeverbrauch
Netto-Wirkungsgrad
1200
1000
800
600
400
200
41
40
39
38
37
36
35
34
ABSOLUTE HEAT INPUT
absoluter Wärmeverbrauch
Netto-Wirkungsgrad
NET EFFICIENCY
MBtu/h
4000
3000
2000
1000
ABSOLUTE HEAT INPUT
Diagramm 5
0
100
200
300
400
[MW]
500
NET OUTPUT
Netto-Leistung
30170995/8016
8.9.1975 Fra

6.5.3 FLOW DIAGRAM AND LAYOUT OF KOMBI-BLOCKS

Representative arrangements, layouts and sections, as well as the basic flow diagram depicting the KOMBI-BLOCKS, are listed below. Prints are included under Appendix 3.

Drawing No.	Title	
Series 30 17 0995/		
8.020	400 MW KDV KOMBI-BLOCK	— Flow Diagram
8.004	2 x 500 MW KOMBI-BLOCK	— Layout
8.005	500 MW KOMBI-BLOCK	— Layout
8.006	500 MW KOMBI-BLOCK	— Section for Layout
8.007	1000 MW KOMBI-BLOCK	— Longitudinal Section
8.008	1000 MW KOMBI-BLOCK	— General Arrangement
8.009	1000 MW KOMBI-BLOCK	— Section Through Coal Conversion Plant

6.6 DEDUCTION OF THE GASIFICATION EFFICIENCY IN THE INTEGRATED COMBINED CYCLE SYSTEM

The designing of a well integrated and optimized combined cycle gasification plant is a tedious exercise. Material and energy flow from the fuel preparation plant (gasification) to the fuel energy conversion plant (combined cycle) and vice-versa provides for a tightly interwoven system. In addition, the objective in a power plant should be the minimum production of by-products and the maximum conversion of coal to kilowatthours. It is therefore not surprising that once integrated, the separation of the gasification efficiency from the overall power conversion efficiency is an awkward and unwarranted exercise. This is the opinion of all developers we dealt with during the study. One is however tempted to insist on some rational account on the losses incurred during the fuel preparation process. In the following we present STEAG's method of accounting based on their coal, and right after our consultant's evaluation based on Hat Creek coal.

It should be noted that the power requirements for the gasification process are acocunted for in the net output (and net efficiency) of the combined cycle.

By interpretation of the consultant's figures, the thermal efficiency of the gasification process with 26 percent ash content in the coal is approximately 93.2 percent.

6.6.1 STEAG METHOD OF ACCOUNTING FOR GASIFICATION EFFICIENCY

Definition: efficiency is equivalent to useable heat in the fuel gas divided by useable heat plus losses, plus evaporation of jacket water, minus additional heat imported to the gasification system from the combined cycle.

Useable heat in gas	100.00%
Heat losses:—	
unburnt and ash	1.40%
radiation	.28%
cooling during gas clean-up	6.67%
	8.35%
Evaporation of jacket water	1.23%
Additional heat from combined cycle:—	
gasification air	1.64%
gasification steam	1.41%
make-up water for jacket	.23%
	3.28%

$$\text{Gasification efficiency} = \frac{100}{100 + 8.35 + 1.23 - 3.28} = .9407 \text{ or } 94\%$$

6.6.2 CONSULTANT'S EVALUATION

Ash Content of Coal	25%	31%
"Cold Gas" efficiency	68.387%	63.876%
Latent heat in tar, etc., recycled	17.700%	18.473%
Latent heat in lock hopper gas	1.841%	2.264%
Latent heat in dissolved gas	.432%	.060%
Sensible heat in raw gas & liquids	5.144%	7.005%
Overall thermal yield	93.504%	91.678%

6.7 THE START-UP OF A KOMBI-BLOCK WITH COAL GASIFICATION

In general, the system can be started-up in two basic methods:

1. The gasification plant is started first and the initial gas is flared until the quality of the gas is adequate for the combined cycle.
2. The combined cycle is started first with oil firing and the gasification plant is started thereafter.

6.7.1 FIRST METHOD

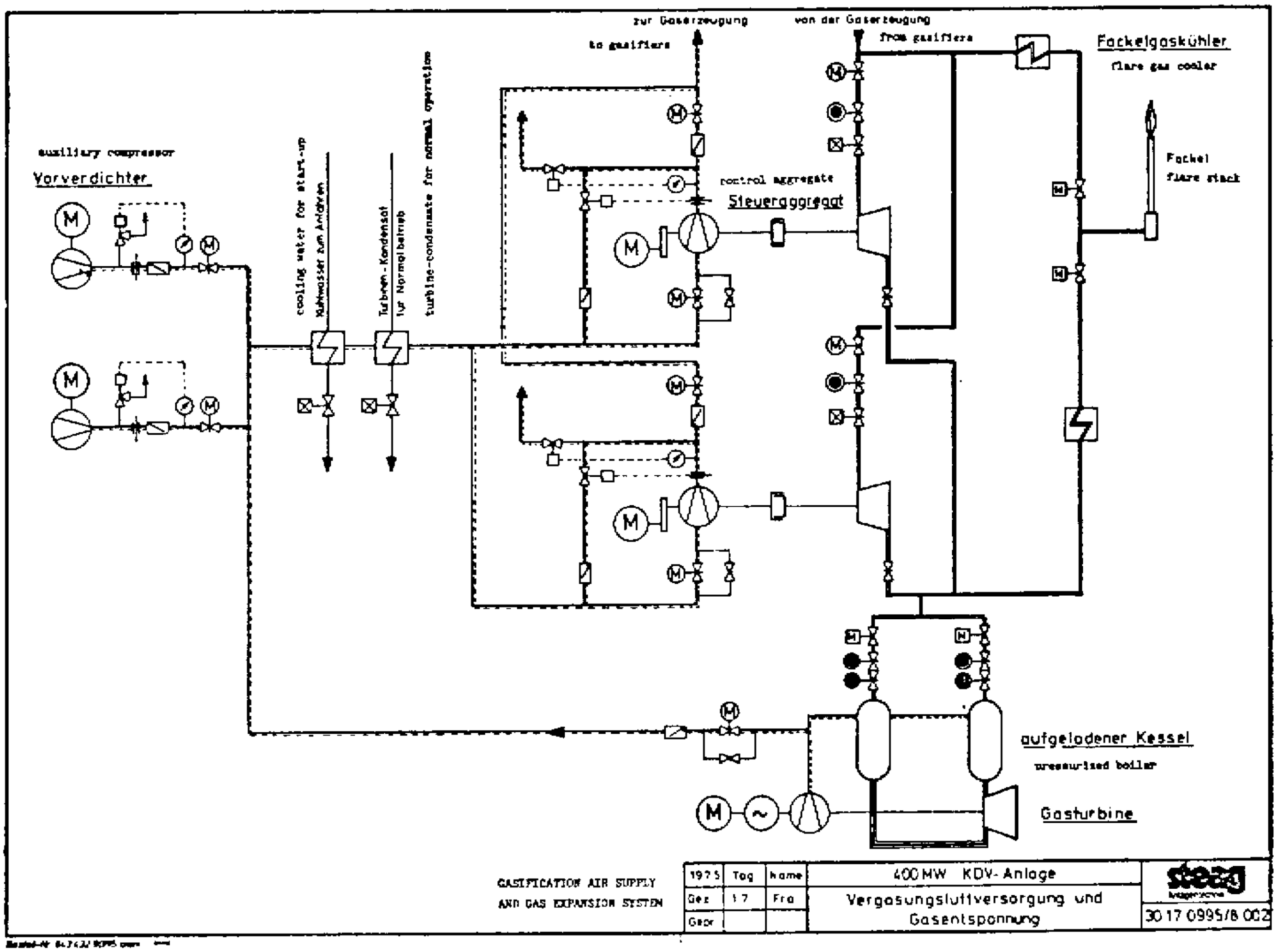
The equipment required for this alternative consists of a start-up boiler, which can be fired either with fuel oil or with the light fractions of the tar produced by the gasification plant. Also, a compressed air supply and storage system is required.

The start-up boiler provides the steam necessary for:

— the preheating of the feedwater for the combined cycle;
— the steam tracing for the tar handling system;
— the supply of initial gasification steam.

The start-up air system supplies the air for initial gasification.

The start-up procedure can be followed from diagram No. 30170995/8.002.



START-UP SYSTEM FOR KOMBI-BLOCKS

When all necessary equipment receives steam and is sufficiently pre-heated, the auxiliary air compressors are started. Depending upon the condition of the gasifier, the booster compressor of the control aggregate is also started with the expansion turbine disconnected.

The compressed air between the auxiliary compressor and the booster compressor is inter-cooled. The air is admitted to the gasifier, together with the start-up steam. The gas produced is either directed straight to the flare stack via a gas cooler, or is admitted first to the expansion turbine, then cooled and flared. The expansion turbine is then coupled with the booster compressor and the drive motor unloaded. By increasing the steam/air flow, the gasifier performance and the quality of the gas will be brought to a level that the combined cycle can be started. To this effect, feedwater circulation will be established and the boiler will be fed with 40% flow rate at approximately 400°F.

For feed heating purposes, the function of the start-up air inter-cooler is replaced by the turbine-condenser heat exchanger. This reduces the load on the start-up boiler. When stable conditions are reached, the gas turbine is activated and the pressurized boiler lit-up. The boiler promptly produces steam which is directed to the main condenser at first, then admitted to the steam turbine. At one point, the main steam system takes over the function of the start-up boiler. As the gas turbine output increases, the gasification air would be provided by the main compressor and the auxiliary air compressors stopped.

The start-up period is completed when the combined cycle reaches the stage to take over the air and steam supply for the gasification plant.

6.7.2 SECOND METHOD

With this method, after the steam is preheated with steam produced by the start-up boiler, the gas turbine is started. The fuel used is the same as for the start-up boiler. The steam generation in the pressurized boiler promptly begins and the steam is either blown-off to the main condenser, or used for pre-heating the steam turbine. When the gas turbine reaches sufficient load, a portion of the air flow is admitted to the booster compressor and delivered to the gasifier. (The expansion turbine would be disconnected.) The gas initially formed is flared. By increasing the fuel oil input, the capacity of the combined cycle will increase and the production of fuel gas follows: At one point, the start-up boiler can be de-activated.

As the gasifier pressure increases, the expansion turbine would be re-connected and the drive motor unloaded. Upon reaching adequate gas quality, the gas is admitted to the burner of the pressurized boiler and the oil firing accordingly reduced. The start-up process is completed when the combined cycle is capable of replacing the function of the auxiliary steam — and compressed air systems.

The unit is capable of taking full load in 40 minutes, when brought up from a warm, dormant stage after, say an 8 to 12 hour shutdown. When started from cold, two hours are required to reach full load. Five percent load change per minute is attainable.

6.8 CAPITAL COST ESTIMATE FOR A 2000 MW PLANT

6.8.1 CAPITAL COST ESTIMATE

The capital cost estimate for a nominal 2000 MW power plant consisting of three STEAG — combined cycle units and the supporting Lurgi gasification system is given under 6.8.3.

The power generating units employed are:

1st Unit:	500 MW KOMBI-BLOCK	486.5 MW net
2nd Unit:	500 MW KOMBI-BLOCK	496.5 MW net
3rd Unit:	1000 MW KOMBI-BLOCK	973.0 MW net
		1946.0 MW net

The power requirements of the gasification plant reduces the station net output to 1934 MW.

The estimate is based on the following conditions:

— The cost of the power plant has been established by STEAG, based on firm prices and quotations obtained in August, 1973. Prices are brought to September 1975 level.
The conversion rate used in \$1.00 = 2.50 DM.

— The station is built with the units listed above; one unit following the other at six to twelve month intervals. The first unit bears the site development costs.

— Spare parts are not included.

— All components, with the exception of the pressurized boilers and the large diameter Lurgi gasifiers are commercially available. These two items are prototypes, being currently developed and tested respectively for commercialization.

— The KOMBI-BLOCKS are equipped with 100 percent fuel gas and 100 percent heavy oil firing equipment. The cost of an oil storage and handling system adequate for 20 full load days has been established, but is not included in the estimate. The cost of the oil firing equipment and associated piping, controls, etc., could not be separated from the boiler price and therefore it is included.

— Special site conditions are not considered.

— The costing of common items such as site preparation, rail spur, 1000 ft. stack, water supply to station, coal and ash handling plant, are based on INTEG's estimate. The cost of the cooling towers and associated C.W. system is based on STEAG's estimate, as this power plant uses less cooling water than a conventional steam turbine plant.
(\$18,240,000 versus \$25,400,000 as established by INTEG.)

6.8.2 DERIVATION OF GASIFICTION PLANT COST CARRIED IN ESTIMATE UNDER PARAGRAPH 6.8.3.

Total as estimated in consultant's report.		\$320,480,000
Less credit for items included in above sum, but accounted for in Paragraph 6.8.3 using INTEG's estimates;		
Coal handling system	\$19,580,000	
Ash handling system	\$ 2,600,000	
Water filtration and intake	\$ 6,030,000	
Rails and yard improvements	\$ 6,470,000	
		\$ 34,680,000
Net total carried in cost estimate under 6.8.3.		\$285,800,000
Distribution of total cost per combined cycle units.		
No. 1 Unit — 30%		\$ 85,740,000
No. 2 Unit — 25%		\$ 71,450,000
No. 3 Unit — 45%		\$128,610,000
		\$285,800,000

6.8.3 COST BREAKDOWN FOR A 2000 MW PLANT

TABLE 6.1

	UNIT NET OUTPUT MW	NO.1 483.5	NO.2 483.5	NO.3 967.0	TOTAL 1934.0
ITEM	DESCRIPTION	SEPTEMBER 1975 PRICES IN $1000			
1	Site Preparation	6,250			6,250
2	Rail Spur	8,125			8,125
3	Water Supply to Station	25,800			25,800
4	Stack	8,850			8,850
5	Coal and Ash Handling Plant	32,700			32,700
6	Buildings	11,040	9,680	13,080	33,800
7	CIVIL SUB-TOTAL:	92,765	9,680	13,080	115,525
8	Boilers with Dual Burners	18,400	18,400	36,800	73,600
9	Boiler Auxiliaries	1,840	1,840	2,080	5,760
10	BOILER PLANT SUB-TOTAL:	20,240	20,240	138,880	79,360
11	Steam Turbo-Generators	16,000	16,000	27,200	59,200
12	Gas Turbines	10,400	10,400	20,800	41,600
13	Boiler Feed Pump Turbine	1,600	1,600	2,400	5,600
14	C.W. System	5,080	5,080	8,080	18,240
15	Feedheaters, Pumps, Tanks	2,960	2,960	4,520	10,440
16	Piping, Fittings, Insulation	11,080	9,880	15,040	36,000
17	Auxiliary Equipment	1,680	1,680	1,800	5,160
18	TURBINE PLANT SUB-TOTAL:	48,800	47,600	79,840	176,240
19	Unit Transformers	2,640	2,640	4,480	9,760
20	Station Services	2,000	2,000	2,480	6,480
21	Motors and Cabling	6,320	6,320	8,520	21,160
22	Controls and Switchgear	1,840	1,840	2,360	6,040
23	ELECTRICAL SUB-TOTAL:	12,800	12,800	17,840	43,440
24	Automation, Instrumentation	5,120	5,120	7,000	17,240
25	**Total items 7 + 10 + 18 + 23 + 24:**	179,725	95,440	156,640	431,805
26	Contingency 10%	17,973	9,544	15,664	43,181
27	Gasification Plant	85,740	71,450	128,610	285,800
28	Contingency 15%	12,861	10,718	19,291	42,870
29	**Total Items 25 + 26 + 27 + 28:**	296,299	187,152	320,205	803,656
30	Engineering and Supervision 8%	23,704	14,972	25,616	64,292
31	**Total Items 29 + 30:**	320,003	202,124	345,821	867,948
32	Corporate Overhead 5%	16,000	10,106	17,291	43,397
33	**Grand Total Items 31 + 32:**	336,003	212,230	363,112	911,345
	Specific Costs $/KW				
34	Based on Item 25	371.7	197.4	162.0	223.3
35	Based on Item 29	612.8	387.1	331.1	415.5
36	Based on Item 31	661.8	418.0	357.6	448.8
37	Based on Item 33	694.9	438.9	375.5	471.2

7. GENERAL ELECTRIC-LURGI SYSTEM

7.1 INTRODUCTION

General Electric are investigating modifications to the Lurgi-type fixed bed gasifier, which would make it suitable for moderately caking coals. A small experimental system using air to produce low-Btu gas has been in operation for some years. A larger gasifier is being built, designed to operate at 20 atm., and to gasify 12 ton/day of coal. Studies were made to evaluate the cost and the performance of a STAG-type combined cycle power plant with the modified Lurgi gasification plant.

Visits were made to G.E. offices at Schenectady to verify that this presentation was representative of the plant which G.E. could provide for the Hat Creek application. In general it was suggested that the plant described in a paper presented to the American Power Conference April 1975 should be used. (Ref. G.E.i).

The choice of gasifier is explained in another paper (Ref. G.E.ii) and a summary is given here.

The cycle designed is based on equipment available today. The reasons indicated by G.E. for not including the pressurized boiler in this application were that large research funds have been made available for development of gas turbines with very high turbine inlet temperatures. G.E. opinion is that, when these machines become available the pressurized boiler will be redundant. Therefore they feel that experience should be gained now in the type of cycle which eventually will be the most beneficial.

7.2 CHOICE OF GASIFIER

A paper published by G.E. (Ref. G.E. ii) shows the reasoning behind choosing the fixed bed gasifier for power generation. They have been conducting research into using coal as a fuel for gas turbines since 1945. The gasification approach avoids the problems of particulate impingement on the blades and investigations have been conducted into the fluidized bed, entrained bed and fixed bed gasifiers. The results of these studies indicate that further development of the already well established Lurgi gasifier offers the most promising hope of success.

The areas of development which are presently being pursued in an experimental rig are as follows:

1. The lock hopper coal feed system is a potential maintenance problem and incurs losses of product gas. G.E. are developing a device which by mixing fine coal with a binder, such as tar extracted in the gas clean-up process, can extrude the coal into a convenient shape and consistency for injecting directly into the gasifier. This would mean that all the coal from the mine including the fines which are presently limited to 7 percent for Lurgi gasification could be used.

2. Heavily caking coals present problems for Lurgi gasification and G.E. are developing a suitable stirring device which would permit use of these coal types.

3. G.E. consider the Lurgi grate as having only limited clinker breaking capability and are just beginning experiments with a strong clinker breaking grate modelled after the eccentric grate used in the Wellman-Galusha gasifier. With the new grate G.E. expect reduction in the gasification steam flow and thereby an increase in the thermal efficiency of the system.

4. Gas clean-up processes presently available for extracting H_2S also extract CO_2. This represents a significant loss of mass flow to the gas turbine. G.E. is developing a liquid membrane gas clean-up system which is very selective in absorbing only H_2S.

7.3 GASIFICATION/COMBINED CYCLE PLANT DESCRIPTION

The description is taken from a paper by G.E. (Ref. G.E. i) which shows a design of plant suitable for using a sub-bituminous coal. Table 7.1 shows a comparison of this coal with Hat Creek average coal.

TABLE 7.1

	COLSTRIP MONTANA SUB-BITUMINOUS	HAT CREEK LIGNITE
PROXIMATE ANALYSIS	%	%
Moisture	28.0	20
Ash	9.0	25
Volatile matter	27.8	25
Fixed Carbon	35.2	30
ULTIMATE ANALYSIS		
C	48.4	37.7
H	3.2	2.9
O	10.2	12.9
N	0.5	0.9
S	0.7	0.4
HEATING VALUE (HHV Btu/lb)	8300	6402

The plant consists of a single 875 MW unit, smaller unit size such as 435 MW would not incur a very significant increase in specific cost. The components have been arranged with provision for necessary access and maintenance room with rail crane service to all turbo-machinery.

Some degree of layout optimization has been included to reduce the runs of the large fuel gas feed piping, steam, boiler feedwater and air feed lines. The plant arrangement requires approximately 122 acres including coal storage and handling. The gas turbines and heat recovery steam generators are laid out such that a back to back stack design results giving a more effective plume rise. An overall layout is shown on pages 244 and 247 including wet cooling towers.

Fuel requirements to meet full maximum load at 34°F requires the full output of about 23 gasifiers. 24 gasifiers are provided to cater for a forced gasifier outage when the maximum plant output has to be met during periods approaching 34°F ambient. During periods when the ambient temperature is above 60°F, 21 gasifiers would be sufficient to meet full load requirements. Below 10°F some form of inlet air heating would be required to avoid load limitations at such low temperatures caused by gas turbine compressor surge restrictions.

The gasifiers are supplied with high pressure air extracted from the combined cycle gas turbine compressors and process steam which is extracted from the main steam turbine. Air extraction permits the use of standard gas turbine aerodynamic designs and avoids compressor/turbine flow mismatching when using low Btu fuel. This is a useful feature permitting dual fuel gas turbine operating capability for use during start-up or in the event of an under supply of low Btu fuel.

Each of the eight General Electric heavy duty MS-7001 gas turbines is arranged with an individual dual-pressure heat recovery steam generator (HRSG). The high pressure elements of the HRSG's generate 1250 psig, 900°F steam for use in a single 200 MW, automatic extracting steam turbine. The low pressure elements generate about 43% of the 400 psig, saturated steam used in the gasification system. The balance of the process requirements are furnished by the steam turbine from the automatic extraction point and the waste heat boiler in the incinerator exhaust.

The combination of these relatively low steam conditions, and modularized HRSG's (which can readily accept gas turbine thermal transients), enables the plant to retain the operating flexibility characteristics of combined cycle plants including fast start-up capability and high availability.

Air for the gasification process is extracted from each gas turbine, intercooled and boosted in a steam turbine-driven booster compressor. The fuel gas to steam exchanger provides superheat to the entrained water vapor to prevent condensation in the fuel gas line and valves.

The first stage of feedwater heating is provided by the process air intercooler between the extraction point at the gas turbine compressor and the booster compressor suction. The second and final stage of feedwater heating is provided by extraction from the steam turbine.

The design of plant is a practical compromise to accommodate major available component designs and operating experience to reduce new design risks, costs and lead time required for commercial operation. There is significant opportunity for performance and economic improvement as gasifier and gas turbine technology evolves.

7.4 PLANT PERFORMANCE

In the integrated plant, the efficient combined cycle is significantly depreciated by the losses associated with the fuels plant. These losses include: carbon in the ash, gases lost during lock hopper operation, heat to the gasifier jacket coolant, sensible and latent heat lost in the gas scrubbing, cooling and resaturation, chemical heat lost in H_2S and ammonia removal, sensible heat lost in various waste and gaseous effluents and unaccounted-for losses.

The heat lost in gas scrubbing, cooling and resaturation is a function of temperature of the raw gas to the clean-up system. (Low in the case of gasifying Hat Creek coal).

7.5 PART LOAD OPERATION

Modulation of the gas turbine variable inlet guide vanes is used for load changes at high loads, resulting in a relatively constant heat rate in this range. The most efficient part load operation is achieved by using the minimum number of gas turbines in a highly loaded condition.

The steam turbine will operate at three distinct throttle pressures in order to maintain favorable moisture conditions in the latter stages of the steam turbine. As load is reduced with all gas turbines in operation, steam production decreases and the steam turbine control valves modulate to maintain constant throttle pressure. At a selected throttle temperature, the throttle pressure set point will be modified, allowing suitable moisture conditions to be maintained.

The plant is provided with dual fuel capability to permit gas turbine/HRSG start-up on liquid fuel providing a source of process air and steam for the gasification plant start-up. Automatic fuel changeover under load is then accomplished as low Btu fuel gas becomes available, thus permitting the rapid start-up and load response, characteristics of the combined cycle, to be maintained. Start-up on low Btu gas fuel is possible and would be the normal start-up mode when at least one turbine/HRSG was operating on low Btu fuel, supplying gasifier reactants, and additional gasifiers were in a standby mode.

Gasifiers can be held in a pressurized hot standby condition by pulse firing every two hours for about 15 minutes. From this ready condition, approximately 30 minutes is required to ramp up to full gas generation. Thus, a hot start-up can be simultaneously on the gas turbines and fuels plants; either on low Btu fuel or on oil, transferring from distillate to coal gas rapidly as the low Btu fuel becomes available.

Cold gasifier start-up is initiated by igniting a combustible material such as wood or fuel oil and using about 10% of the fuel process air requirement. Combustion products are vented until the stack gas is at a level suitable for flaring. Gas flow is transferred to the clean-up train and gasifier steam injection initiated to gradually pressurize the system. As the gasifier is pressurized, the air flow is ramped from the 10% initial value to about 25%. The gas flow increases with air flow reaching about 25% by the time the gasifier is at the normal operating pressure. From cold start, about 4 hours will be required to reach this condition, which is then equivalent to the hot standby state. Beyond this point, the response of the gasifier is the same as a hot restart with the capability of going to full output in about 30 minutes.

In the event of a plant trip which requires a fuels plant trip, the gasifiers are bottled with no external steam and air flow required for about two hours. This shutdown is accomplished without flaring gas or blowing relief valves. During this two hours, the gasifiers are available for hot restart. Beyond the two hours, pulse firing will be required to maintain the hot standby condition. This design approach enables the integrated plant to maintain the rapid start-up load response and cycling capabilities important in meeting the electric utility application requirements.

7.6 PLANT COSTS

Costs taken from the G.E. paper have been adjusted to September 1975 and contingency factors have been applied in accordance with instructions given by B.C. Hydro. They are shown in Table 2 together with an estimate of cost of supplying cooling water to such a plant in the Hat Creek application.

TABLE 7.2
CAPITAL COST ESTIMATE FOR A "STAG" 800 POWER PLANT
INTEGRATED WITH A LURGI — G.E. COAL GASIFICATION PLANT

	THOUSANDS OF DOLLARS		
	G.E. ESTIMATE JANUARY 1975	SECo UPDATE SEPTEMBER 1975	NOTE
Fuels Plant			
COAL HANDLILNG AND ASH	18,700	17,465	No.1
Includes all conveying, screening, bricketting, unloading with intermediate storage and related controls.			
GAS PRODUCTION	92,800		
Includes gasifiers, gas scrubbing, cooling, ammonia removal, final gas saturation, H_2S absorbent regeneration, conversions to elemental sulfur and storage in liquid form, and related controls.		151,354	No.2
FUELS PLANT SUPPORT FACILITIES AND SITE COSTS	16,300		
Includes tar handling, controls, incinerator with waste heat boiler, flare stack, control storage and maintenance buildings, fire protection, fuel storage, rail sidings, sulfur loading docks, and site preparation			
Sub-total	127,800	168,819	No.3
Contingency 15%		25,323	No.4
Fuel plants sub-total		194,142	
Power Plant			
Major Combined Cycle Equipment	114,500	123,660	No.3
Contingency 10%		12,366	No.4
Includes gas turbines, heat recovery steam generators, steam turbine, booster compressor sets, power plant control and set-up transformers.			
Balance of Plant Equipment & Installation	35,000	37,800	No.3
Contingency 15%		5,670	No.4
Includes mechanical and electrical support items such as condensers, boiler and circulating pumps, cooling towers and circulating water system.			
Power plant sub-total	149,500	179,496	
Overall contingency used by G.E. 7%	19,300		
TOTAL	296,600	373,638	
Specific Cost $/kW	335.1	422.2	No.5

7.7 EXPLANATORY NOTES TO TABLE 7.2

1. The cost of this item is derived from INTEG's price of \$32,700,000, pro-rated to account for differences in plant output and efficiency:

   $\$32{,}700{,}000 \times \frac{885}{2000} \times \frac{40}{33} = \$17{,}465{,}000$

2. G.E.'s estimate for the gas production plant and supporting facilities is \$109,100,000 based on processing 549 ton/hr Montana coal. The specific cost is therefore \$198,725 per ton/hr coal, based on January 1975 price level, which becomes \$215,623 using September 1975 prices. Our estimate for the STEAG system is \$285,800,000 for processing 1224 ton/hr Hat Creek coal, giving a specific cost of: \$229,743 per ton/hr coal. Because the higher ash content of Hat Creek coal, the difference in specific costs appear justified. The G.E. gasification system, however, would have to process more Hat Creek coal than the STEAG system, because of its lower overall efficiency. The increase would be proportionate with the ratio of the efficiencies, or about 40.3 : 33.1 = 1.21. This would require more gasifiers and associated equipment. Assuming 710 ton/hr coal consumption for 885 MW output, the equivalent cost for using Hat Creek coal could be estimated (using G.E.'s unit cost):

   $710 \times \$229{,}743 = \$151{,}354{,}000$

3. G.E. prices are multiplied by 1.08 to bring them to September 1975 level.

4. Contingencies applied as directed.

5. This specific cost of \$422.2 per kW is indicative of what can be achieved with one 800 MW unit and is therefore somewhat comparable to item 36 of the tabulation shown in Paragraph 6.8.3, which gives \$418.0 and \$357.6 for a 500 MW and a 1000 MW KOMBI-BLOCK respectively.

REFERENCE

G.E. (i) Economics of Power Generation from Coal Gasification for Combined Cycle Plants.

Ahner, Sheldon, Garrity, Kasper
American Power Conference — April 1975
Palmer House — Chicago — Illinois

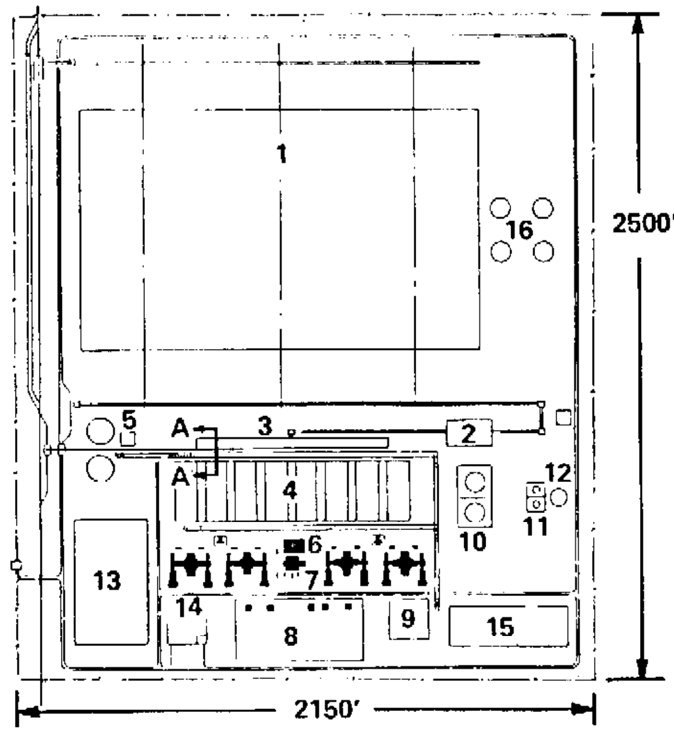
G.E. (ii) Coal Gasification Research at General Electric

Past and Present
Bueche and Kydd
Sixth Synthetic Pipeline Gas Symposium
October 1974 — Chicago — Illinois

G.E. (iii) Integrated Gasification — Gas Turbine Cycle Performance

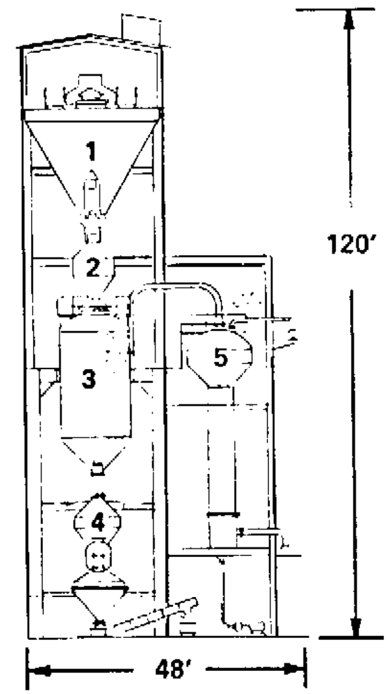
P.H. Kydd
Report No. 75 CRD 021 — March 1975

GASIFICATION/COMBINED CYCLE PLANT LAYOUT (NOMINAL CAPACITY - 875MW)



1 COAL HANDLING & STORAGE
2 BRIQUETTING
3 GASIFIERS
4 GAS TREATMENT & TAR FACILITY
5 ASH HANDLING
6 CONTROL ROOM
7 POWER PLANT
8 SUB STATION
9 BFW TREATMENT
10 LOCK GAS STORAGE
11 INCINERATORS
12 GAS FLARE
13 STRETFORD
14 MAINTENANCE & FIREHOUSE
15 COOLING TOWER
16 FUEL OIL STORAGE TANKS

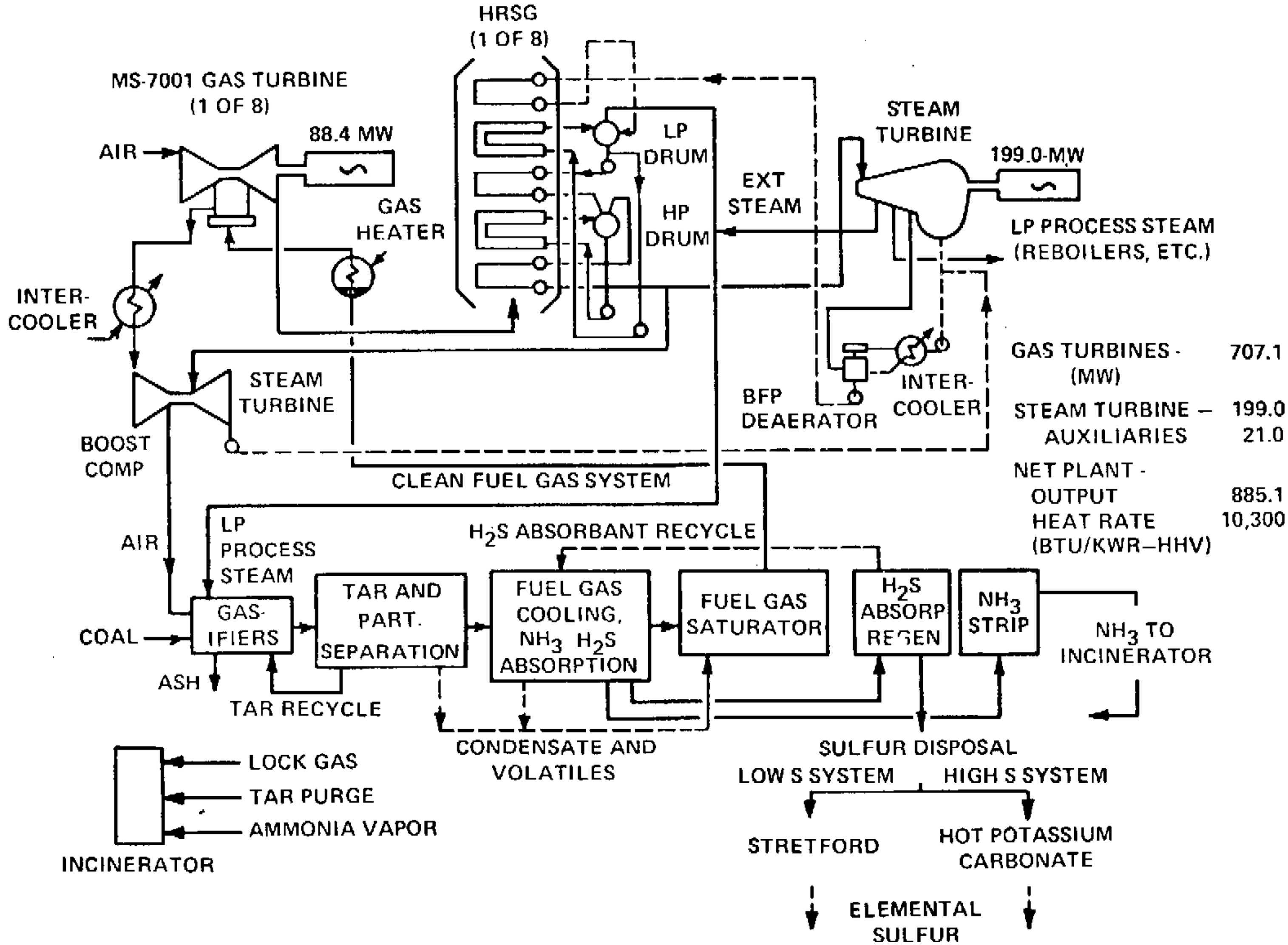
LURGI GASIFIER



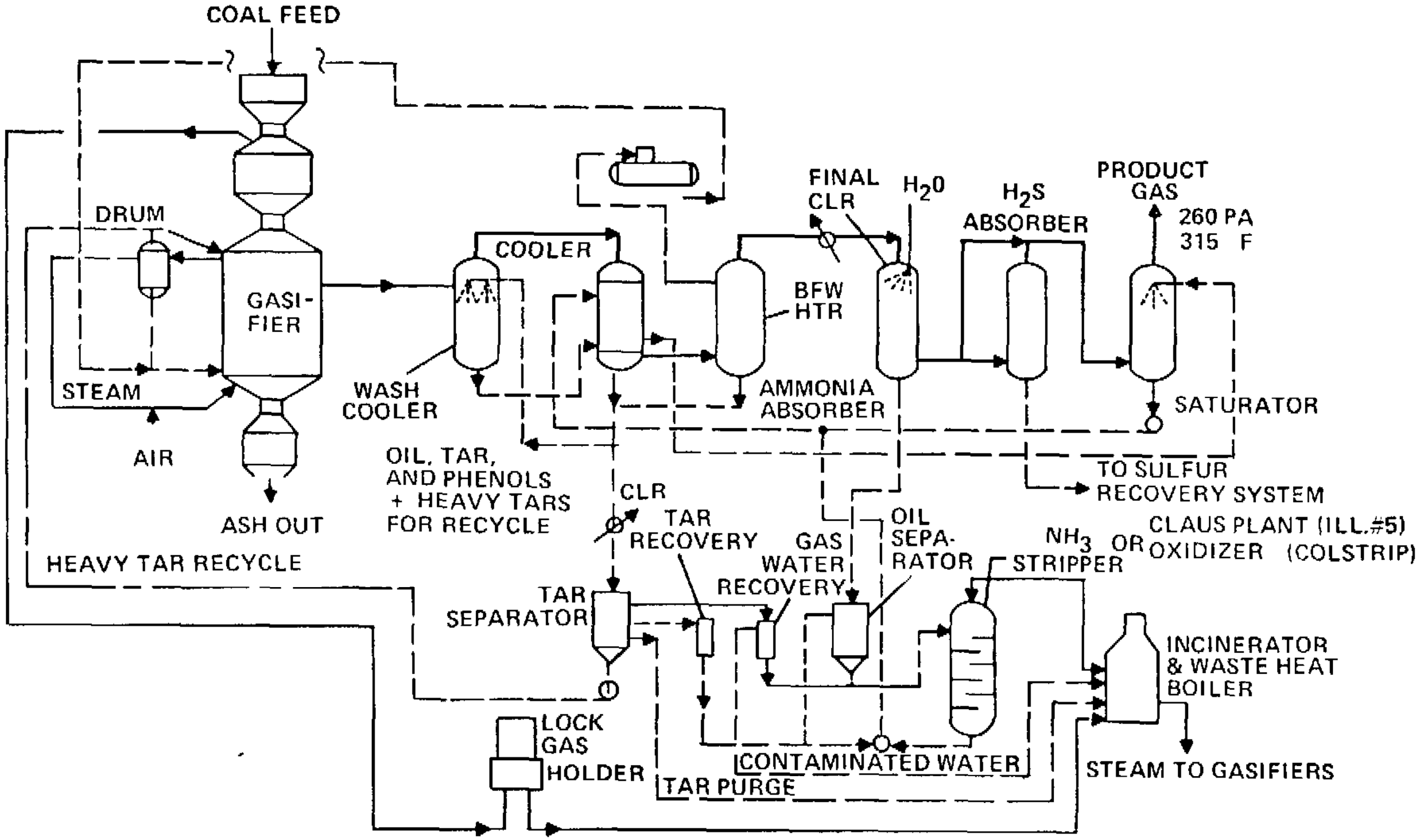
SECTION A-A

1 COAL HOPPER
2 COAL LOCK HOPPER
3 REACTOR VESSEL
4 ASH LOCK HOPPER
5 WASH COOLER

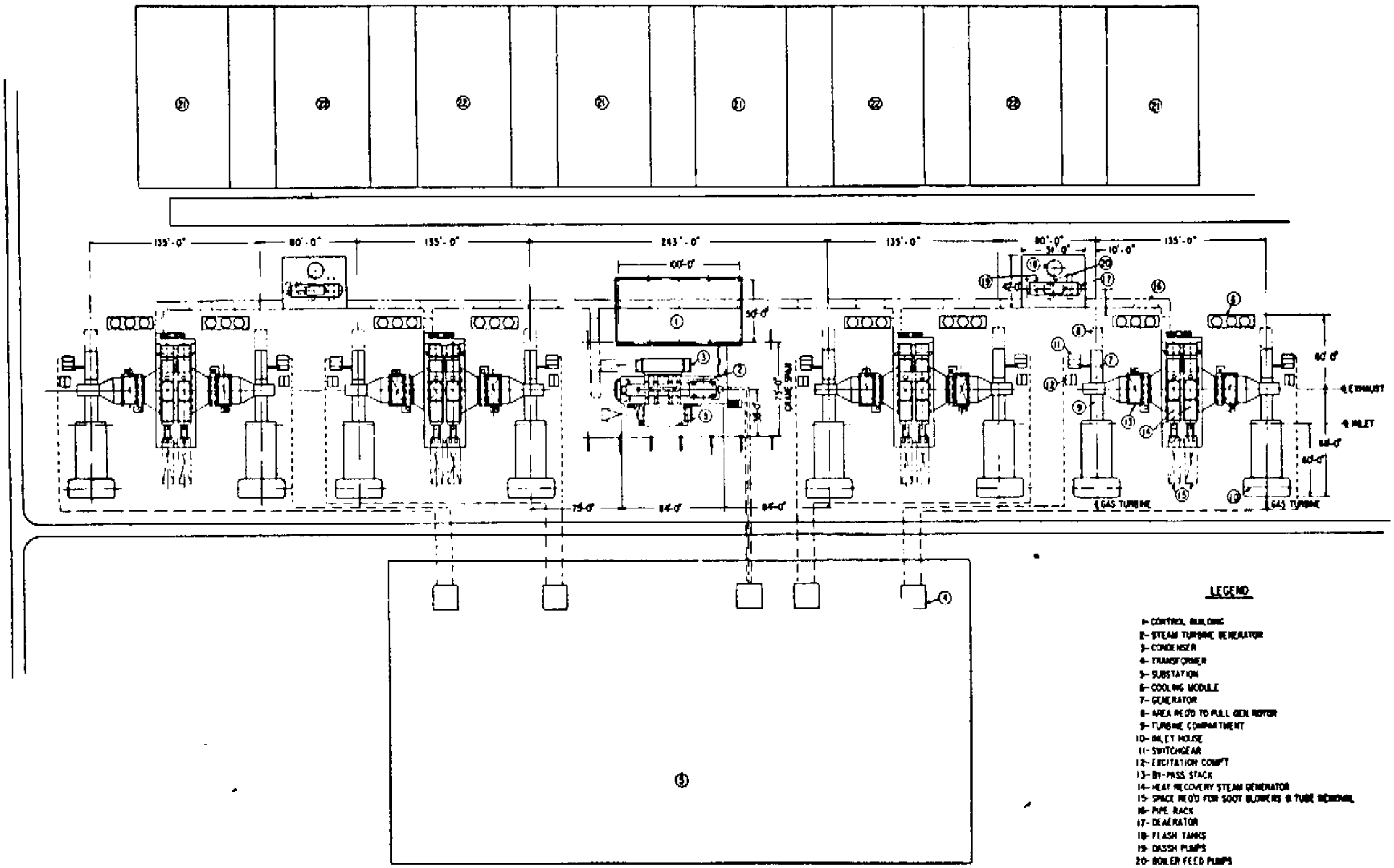
INTEGRATED GASIFICATION COMBINED-CYCLE PLANT



FIXED-BED GASIFICATION SYSTEM



COAL GASIFICATION POWER PLANT



LEGEND

1- CONTROL BUILDING
2- STEAM TURBINE GENERATOR
3- CONDENSER
4- TRANSFORMER
5- SUBSTATION
6- COOLING MODULE
7- GENERATOR
8- AREA REQ'D TO PULL GEN ROTOR
9- TURBINE COMPARTMENT
10- INLET HOUSE
11- SWITCHGEAR
12- EXCITATION COMP'T
13- BY-PASS STACK
14- HEAT RECOVERY STEAM GENERATOR
15- SPACE REQ'D FOR SOOT BLOWERS & TUBE REMOVAL
16- PIPE RACK
17- DEAERATOR
18- FLASH TANKS
19- DASSH PUMPS
20- BOILER FEED PUMPS
21- GAS TREATMENT
22- TAR FACILITY

8. UNITED TECHNOLOGIES — KELLOGG SYSTEM

8.1 INTRODUCTION

United Technologies Research Laboratories have been pursuing federally sponsored development work for some years, involving two-step combustion and large gas turbines. The studies stem from a National Air Pollution Control Administration Contract, completed in December 1970. Present activities are in the field of flue gas scrubbing, fluidized bed technology and combined cycle development in connection with advanced coal gasification systems. Gas turbine development is aimed at building a demonstration unit — consisting of one gas turbine of 150 MW and one steam turbine of 100 MW by the end of 1981 operating at 2600°F inlet temperature. The demonstration plant would run until 1983 and commercialization is expected around 1985. The cost of the program until 1981 is $50 million and the clean-up of the fuel forms part of this program.

UTRL were presenting a report in September this year on their development of gas turbine combustors to burn low-Btu gas.

The aim of the development program is to build 1000 MW coal gasification combined cycle modules. The power plant would employ one steam turbine and multiple gas turbines. The choice of the gasifier between four alternatives — Kellogg molten-salt, Westinghouse — fluidized bed, BOM stirred bed and Foster Wheeler entrained bed — would be decided upon the completion of demonstration results by the end of 1978. UTRL are looking favourably at the Kellogg process, as it removes sulphur directly from the raw gas. This process, in their opinion, is similar to that used in the paper industry and should have a good chance of success. UTRL regard the Lurgi gasifier as the present day most commercially applicable gasifier, however, they consider the Kellogg process as having the most promising development potential.

8.2 THE KELLOGG MOLTEN SALT PROCESS

The M.W. Kellogg Co. have recently decided to develop the process in collaboration with Atomics International Corporation. Federal Research funds are not involved in this development and it is inferred that this is the reason why there is substantially less information publicity available about the development.

The description here is taken from publications by UTRL and Kellogg. The information is somewhat dated but basically shows the technique of gasification. The line of development being pursued by Kellogg and Atomics International was indicated at meetings with UTRL to be away from using steam in the gasifier and to use coal of up to ¾ - 1" mesh. No published data was made available to us on this development but UTRL indicated that they were investigating the requirements for minimum hydrogen for combustibility in a burner.

The molten salt process was designed by Kellogg to produce synthesis gas which could be upgraded to synthetic natural gas. Heat for the endothermic steam-carbon reaction would be supplied by circulating a stream of molten sodium carbonate between the coal gasification vessel and the melt regeneration vessel. In the latter vessel, a portion of the coal would be burned with air to heat the melt and supply heat for the gasifier. Experimental data indicate that the melt, besides acting as a heat-transfer medium, would act as an H_2S absorbent.

The Kellogg molten salt process has been modified to produce a high pressure, low Btu fuel gas for COGAS power systems. A schematic flow sheet of the modified process is shown on page 258. Approximately 11,566 lb/min of coal would be fed from a 90-day storage pile through two stages of hammer mills which would grind the coal to 12 mesh size consistency. After grinding, the coal would be lock hoppered into the gasifiers. Gasification would occur at 1930°F and 230 psia in a suspension of coal and molten sodium carbonate. Based upon experimental data presented by Kellogg, the reaction rate was estimated at 10 lb carbon/hr/ft^3 of melt. Therefore, eleven 16 ft. diameter by 40 ft. high gasification vessels would be required. During the gasification of coal practically all sulphur in the coal would be absorbed by the melt. The purpose would be merely as a sulphur acceptor, not as a heat transfer medium. For the purpose of this evaluation, it was assumed that none of the sulphur in the coal would be transferred to the fuel gas. Experimental data indicate that this would be a valid assumption.

A stream of molten salt would be withdrawn from the gasifiers to maintain a 4 percent carbon and 8 percent ash inventory in the reactor. The presence of this amount of carbon and ash appears to catalyze the gasification reaction to some extent. The spent melt, 9500 lb/min, would then be regenerated. The stream would first be quenched and dissolved in water and the insoluble ash, and unreacted carbon would be filtered from the solution. The solution would then be recarbonated with CO_2.

The relatively insoluble sodium bicarbonate would precipitate and be recovered by filtration. Sodium bicarbonate with a small amount of make-up sodium carbonate would then be recycled to the gasifiers. Calcination of the bicarbonate would occur insitu. The H_2S, which would be evolved during the recarbonation of the spent melt, would serve as a feed to a Claus conversion plant.

The CO_2 requirements for melt regeneration would be met by cooling about 55 percent of the product gas from the gasifiers, and then scrubbing this stream in a hot-carbonate system to recover the CO_2. During the cooling of this stream of product gas, 447,000 lb/hr of steam and 458,000 ft^3/min of air would be preheated to 1000°F. After CO_2 recovery, the cooled and uncooled gas streams would be recombined to yield a stream of 1100°F product gas at 14 atm pressure.

8.3 MATERIAL BALANCE

Raw material requirements and yields for the molten salt gasification process which would supply a clean fuel gas for a second-generation 1000 MW COGAS power system are summarized in Table 8.1 Approximately 2.13 million tons/yr coal would be required. Electrical requirements for the gasification process would be 92 MW to produce a clean gas at a rate of 893,570 ft^3/min at 14 atm and 1100°F. The gas would have a higher heating value of 110.8 Btu/ft^3 and a sensible heat above 60°F of 21 Btu/ft^3 (total heating value 131.8 Btu/ft^3). The total sulphur oxide emissions from the power station would be less than 20 g/million Btu, all of which would be in the Claus plan effluent. The overall efficiency of the process, net power out/total coal in, was estimated to be 40.5 percent. The efficiency of the gasifier is given as 90.9 percent by Kellogg.

TABLE 8.1

Material Balance for Kellogg Molten Salt Gasifier/COGAS 1000 MW combined cycle plant at 70 percent load factory.

IN		OUT	
Coal:	2128 million ton/year 11566 lb/min (as received)	Gas:	893570 scfm 49453 lb/min.
		HHV	110.8 Btu/scf
Electricity:	92 MW	Sensible Heat:	21 Btu/scf
		Analysis	
		Analysis	Vol. %
		H_2O	13.8
		H_2	15.5
		CO	17.2
		CO_2	7.2
		CH_4	0.5
		N_2	45.8
		Sulphur — 59700 Ton/year 20 g/MM Btu	

8.4 COGAS COMBINED CYCLE PLANT

8.4.1 ADVANCED-CYCLE GAS TURBINES

While meaningful improvements in aerodynamic performance are projected for future gas turbines, the most significant future technological advances are expected in the area of turbine inlet temperature. Part of the increase in turbine inlet temperatures will be achieved by the use of improved turbine blade materials. Historically, maximum turbine blade temperatures have advanced approximately 20°F per year because of materials improvements, and this trend is predicted to continue. Significant increases in turbine inlet gas temperature beyond those levels attainable with improved materials are possible by introducing advanced turbine blade cooling techniques developed for aircraft engines. Current industrial gas turbines have not taken full advantage of these cooling techniques and thus are limited to turbine inlet temperatures of approximately 1600 to 1800°F for base-load ratings.

8.4.2 ADVANCED TURBINE MATERIALS

In current aircraft gas turbines extensive use is made of nickel-based alloys in the hot turbine sections. Casting alloys such as B-1900 and IN-100 have superior thermal fatique characteristics when used for turbine blades. By proper heat treatment it appears that formation of the troublesome sigma phase can be avoided so that these alloys should be suitable for long lifetime service that could be expected of base-load machinery, i.e., approximately 30,000 to 100,000 hr.

Turbine blade materials for industrial gas turbine designs anticipated by the early 1980's will include high-temperature nickel alloys, such as modified B-1900 A, and un-directionally sodified eutectic alloys, such as Ni_3 AL-NI_3Cb currently under development for advanced, high-temperature aircraft turbines. Although an accurate prediction of materials which will be available for use during the 1990's is difficult, it is reasonable to assume that chromium — and columbium-type materials currently being investigated will be used.

Coatings for blades and vanes, such as the aluminum-base Type US and JO-coat, while having lifetime for only several thousand hours in aircraft applications could be modified for use in advanced industrial turbines to meet the much longer lifetime requirements. Also, recent progress in the ability to coat columbium-base alloys may allow their usage by the 1990 time period. One of the principle objectives of coatings for industrial gas turbines is to protect blades and vanes against high-temperature sulfidation. In future industrial gas turbine applications operating on gasified and desulfurized fuels, the fuel sulfur content may be low enough to eliminate the sulfidation problem, thus permitting longer blade lifetimes and/or higher gas turbine operating temperatures.

8.4.3 ADVANCED TURBINE COOLING TECHNIQUES

Currently, only the first-stage vanes and disks of industrial gas turbines are cooled. Thus cooling is presently accomplished by means of air extracted from the compressor and injected directly into the hot turbine sections to be cooled. It will be necessary to cool successive stages of blades and vanes if long-life operation at high turbine inlet temperature is to be realized.

The use of advanced impingement-convection cooling techniques should allow base-load turbine operation at turbine inlet temperatures as high as 2400°F. Another cooling technique that could be used is film cooling, in which air from the hollow core of the blade would be injected through slots in the blade wall to form a layer of cool air, which acts like an insulating blanket over the surfaces to be protected. Film cooling schemes are currently under development for advanced aircraft propulsion systems and should allow base-load operation at temperatures approaching 2600°F. Transpiration cooling, another advanced cooling technique in which cooling air passes through porous blade material, could be used to achieve turbine inlet temperatures approaching 3000°F.

8.4.4 WASTE-HEAT RECOVERY STEAM SYSTEMS

Cycle studies have demonstrated that when the boiler inlet gas temperature is below approximately 1200°F, single-pressure steam systems would result in stack temperatures in excess of 300°F. By adding a second low-pressure steam cycle, it is possible to extract additional heat from the stack gases and drop the stack temperature to 300°F. The temperature distribution for representative single- and two-pressure heat recovery systems presented in Figure 8.1 illustrates the difference in stack temperature and the additional heat recovered by a two-pressure steam system.

8.4.5 COST ESTIMATE

The following tabulation originates from fragmentary information made available to us at our September 10, 1975, meeting with four representatives of UTRL, for the preparation of this section. The information relates to cost estimates and technical data included in UTRL's September 1975 report, mentioned before and forms page numbers 240, 263, 269, 270, 273 and 274 of that report. Reproductions of these pages are attached at the end of this section for record purposes. As it can be seen, the Kellogg process is not represented amongst the alternatives listed.

For the purpose of this report, we have used the first-generation BOM gasification system with the SELEXOL purification process to form the basis of our cost estimate. The reason for this selection is the limitation of 1950°F gas turbine inlet temperature prescribed in the Base Engineering Data. While the selected system does not comply with this requirement, it is one of the least exaggerated versions for which supporting cost estimates were available. The basis of the pricing is mid-1974. The tabulation overleaf shows the UTRL figures and our extrapolations in order to bring the prices to September 1975 level as well as to apply the contingencies prescribed. The specific cost is approximately $420 per kW.

From our conversation with Mr. W.A. Blecher, Senior Research Engineer, we learned that he found the specific cost of a nominal 1000 MW COGAS-Kellogg plant to be $415 expressed in early 1975 dollars. The plant would use eight gasifiers, each capable of processing 60 to 80 tons of coal per hour, four gas turbines, four heat recover steam generators and one steam turbine for a net output of 990 MW. Also, we were told, that the cost of the Kellogg gasification system would not differ much from that of the BOM system. In the absence of better information, we therefore consider our estimate as acceptable for the purpose of this report, but perhaps on the low side, subject to modifications in the light of the findings of Study C by Lummus Co. of Canada.

COST ESTIMATE FOR COGAS 800 MW COMBINED CYCLE UNIT
WITH COAL GASIFICATION (NET OUTPUT: 737 MW)

FPC NO.	DESCRIPTION	THOUSANDS OF DOLLARS	
		UTRL MID-1974	SECo SEPTEMBER 1975
341	Structures and Improvements	7,759	8,923
342	Prime Movers (gas turbines)	24,501	28,176
344	Generators for Above	10,538	12,119
312	Boiler Plant	27,350	31,452
314	Steam Turbine-Generator	18,595	21,384
345 & 353	Electrical Equipment	8,769	10,084
346	Misc. Equipment	359	413
	Other Expenses	1,957	24,800*
	Sub-total	99,828	137,351
	Contingency	7,986	20,603
	Engineering & Supervision	14,974	12,638
	Gasification Plant	101,994	117,293
	Contingency	N/A	23,459
	Engineering & Supervision	N/A	11,260
	TOTAL	224,782	322,604
	Corporate Overhead 5%		16,130
	TOTAL		338,734
	Specific Cost $/kW		459.6

*include: coal and ash handling, water supply, higher stack, rails, etc., prorated from INTEG's reference estimates.

TABLE 45

INTEGRATED SYSTEMS PERFORMANCE SUMMARY

	FIRST GENERATION		SECOND GENERATION		SECOND GENERATION GASIFICATION FIRST GENERATION POWER SYSTEM	
	BOM/ SELEXOL	BOM/ IRON OXIDE	BCR/ SELEXOL	BCR/ CONSOL	BCR/ SELEXOL	BCR/ CONSOL
GAS TURBINE						
Turbine Inlet Temperature — F	2,200	2,200	2,600	2,600	2,200	2,200
Compressor Pressure Ratio	16	16	24	24	16	16
Exhaust Temperature — F	916	913	1,107	1,115	913	920
Output Power — Mw	595.4	626.2	726.6	857.6	642.3	757.6
STEAM CYCLE						
Steam Temperature — F	816	813	1,000	1,000	813	820
Steam Pressure — psia	1,250	1,250	1,250	1,250	1,250	1,250
Condenser Pressure In. Hg. Abs.	4.0	4.0	4.0	4.0	4.0	4.0
Net Steam Cycle Output — Mw	223.8	208.1	293.3	296.6	273.5	271.4
Net Steam Cycle Efficiency	.280	.292	.307	.307	.282	.279
GASIFIER AND CLEANUP SYSTEM						
Coal Feed Rate — lb/hr	700,000	700,000	700,000	700,000	700,000	700,000
Air — Coal Ratio	3.013	2.688	3.088	3.088	3.088	3.088
Steam — Coal Ratio	.405	.349	.567	.567	.567	.567
Air Preheat Temperature — F	800	800	800	800	800	800
Steam Temperature — F	584	584	1,000	1,000	913	920
Steam Pressure — psia	1,250	1,250	1,250	1,250	1,250	1,250
Gasifier Exit Temperature — F	1,000	1,000	1,800	1,800	1,800	1,800
Cleanup System Exit Temperature — F	265	1,070	1,000	1,700	1,000	1,700
Fuel Gas Higher Heating Value Btu/SCF	160.3	165.9	159.3	135.8	159.3	135.8
INTEGRATED STATION						
Gross Power — Mw	819.2	834.3	1,019.9	1,154.2	915.8	1,029
Boost Compressor Power — Mw	43.4	36.1	40.1	40.2	40.1	40.2
Gasifier & Cleanup Aux. Power — Mw	28.2	36.5	58.7	27.6	58.7	27.6
Plant Auxiliaries — Mw	10.6	10.2	13.6	14.5	12.5	13.1
New Plant Output — Mw	737.0	751.5	907.5	1071.9	804.5	948.1
Net Plant Efficiency (HHV-Coal)	.314	.320	.360	.425	.319	.376

TABLE 48
INTEGRATED SYSTEM COST SUMMARY

	FIRST GENERATION		SECOND GENERATION	
	BUMINES SELEXOL	BUMINES IRON OXIDE	BCR SELEXOL	BCR CONSOL
CAPITAL COSTS — $/kw				
Power System Cost	232	219	208	190
Gasification System Cost	111	107	117	99
Cleanup System Cost	88	48	89	35
Total Plant Cost	431	374	414	324
OWNING-PLUS-OPERATING COSTS-mils/kwhr				
Owning Costs (17% of Capital)	11.94	10.36	11.47	8.97
Operation & Maintenance				
Power System	1.32	1.25	1.19	1.08
Gasification & Cleanup	2.75	2.14	2.84	1.85
Fuel Cost at 60c/MM Btu	6.52	6.40	5.69	4.82
Total Cost of Power	22.53	20.15	21.19	16.72

TABLE 51
POWER SYSTEM COST DETAILS — BUMINES/SELEXOL/COGAS

ACCOUNT 341			
341-17	Site Preparation		$848,250
341-18	Administration Building		563,830
341-19	Turbogeneration Building		3,948,000
341-20	Tank Farm		1,236,100
341-23	Condensate Polishing System		800,640
341-24	Stack		362,000
		Total 341:	$7,758,820
ACCOUNT 343			
343-01	Gas Turbine (8) Includes Installation, Labor		$17,643,500
343-02	Starter Motor (8)		85,000
343-03	Torque Converter		804,000
343-04	Lube Oil Purifier & Storage (Pumps, Filters, etc.)		224,000
343-05	Lube Oil Fire Protection		160,000
343-07	Air Compressor Services, Instrumentation		140,000
343-08	Breeching		2,140,600
343-09	Expansion Joints; Not Applicable in COGAS Plant		0
343-10	Inlet Air Filters		555,760
343-12	Energy Cooling Tank Pump & Piping		11,200
343-13	Fuel Oil Heaters & Pumps		123,200
343-14	Miscellaneous Pumps & Tanks		56,000
343-15	Control Panels		560,000
343-16	Computer Controls		560,000
343-17	Fuel Piping		1,249,400
343-18	Fuel Pipe Insulation		187,500
		Total 343:	$24,500,760

ACCOUNT 344			
344-01	Generator For Gas Turbine		$10,537,630

ACCOUNT 312			
312-01	Waste Heat Boiler		$22,432,400
312-02	Boiler Feed Pump		339,390
312-03	Boiler Feed Tank Deaerator		133,070
312-04	Water Treatment (Demineralization)		691,960
312-05	Condensate Storage Tank		29,940
312-08	Miscellaneous Pumps		70,525
312-09	Piping		3,080,240
312-10	Insulation for Piping		264,420
312-11	Controls } Steam Turbine Only		308,025
312-12	Computer		
		Total 312:	$27,349,970

ACCOUNT 314			
314-01	Steam Turbine and Generator (Output per Unit — 105,800 kw)		$9,833,270
314-03	Condenser & Tubes		1,212,180
314-04	Condensate Vacuum Pump & Motor		134,795
314-05	Condensate Pump & Motre		274,780
314-06	Cooling Tower		5,997,080
314-08	Circulation Water Valves & Expansion Joints		
314-09	Circulation Water Pumps		1,146,280
314-10	Make-Up Structure: Screens & Pumps		
		Total 314:	$18,595,385

ACCOUNTS 345 & 353		
	Accessory Electrical Equipment	$8,768,750
ACCOUNT 346		
	Miscellaneous Power Plant Equipment	359,040
	Other Expenses	$1,957,410
	Total Direct Construction Costs	$99,827,765
	Contingency (8%)	$7,986,220
	Engineering & Supervision (15%)	$14,974,165
	Total Unescalated Cost	$122,788,150

TABLE 53

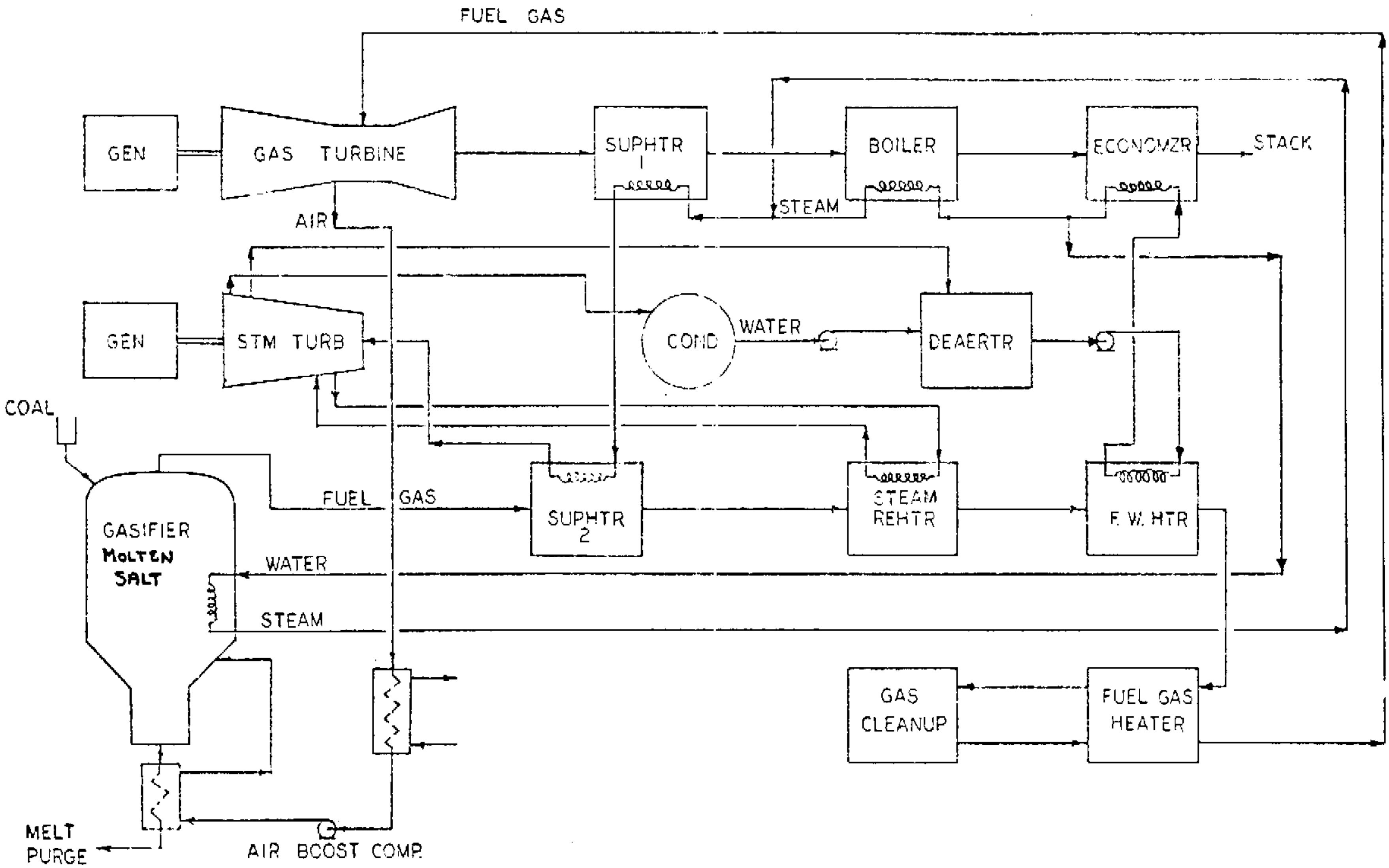
COGAS POWER SYSTEM COST SUMMARY

FPC ACCOUNT NUMBER		COSTS — $1,000 BUMINES/ SELEXOL	BUMINES/ IRON OXIDE	BCR/ SELEXOL	BCR/ CONSOL
341	Structures and Improvements	7,494	7,494	11,011	11,422
343	Prime Movers (Gas Turbine)	23,665	24,637	20,725	23,556
344	Electric Generators (Gas Turbine)	10,178	10,315	9,514	10,077
312	Boiler Plant Equipment	26,417	26,204	29,445	32,360
314	Steam Turbine Generator Units	17,961	18,148	23,230	22,822
345 & 353	Accessory Electric Equipment	8,469	8,679	10,350	12,009
346	Miscellaneous Power Plant Equipment	347	351	385	416
	Other Expenses	1,891	1,917	2,093	2,253
	Direct Construction Costs	96,421	97,749	106,754	114,914
	Contingency, Engineering & Supervision	22,177	22,482	24,553	26,430
	Total Construction Costs	118,597	120,231	131,308	141,344
	Interest & Escalation	52,030	52,746	57,606	62,009
	Total Capital Cost (Power System Only)	170,627	172,978	188,913	203,353

TABLE 54
GASIFIER & CLEANUP SYSTEM CAPITAL COST BREAKDOWN

	BUMINES/ SELEXOL	BUMINES/ IRON OXIDE	BCR/ SELEXOL	BCR/ CONSOL
Gasification	62.42	62.43	83.23	83.23
Gas Cooling	14.86	—	23.78	—
Desulfurization	23.78	20.81	29.72	20.81
Sour Water Stripping	5.94	—	5.94	1.49
Ammonia Recovery	10.40	—	8.92	—
Sulfur Recovery	2.97	10.40	2.97	8.92
Waste Water Treatment	4.82	3.75	6.18	4.58
Boost Compressor & Boiler	10.80	10.72	11.00	11.73
Feedwater Treatment	6.73	5.80	9.42	9.42
Cooling Tower	1.06	—	1.75	.42
Condensate Polishing	.06	.22	.25	.03
Other Expenses	2.88	2.28	3.66	2.81
Total Capital Cost (Includes Escalation & Interest)	146.74	116.41	186.83	143.44

SECo NOTE
From the figures on page 273, interest and escalation was found to represent 43.87 percent increase. The total capital cost of the gasification plant, before escalation and interest is therefore believed to be $101,994,000.



COGAS-KELLOGG COMBINED CYCLE UNIT

FIGURE 919175 (United Technologies)

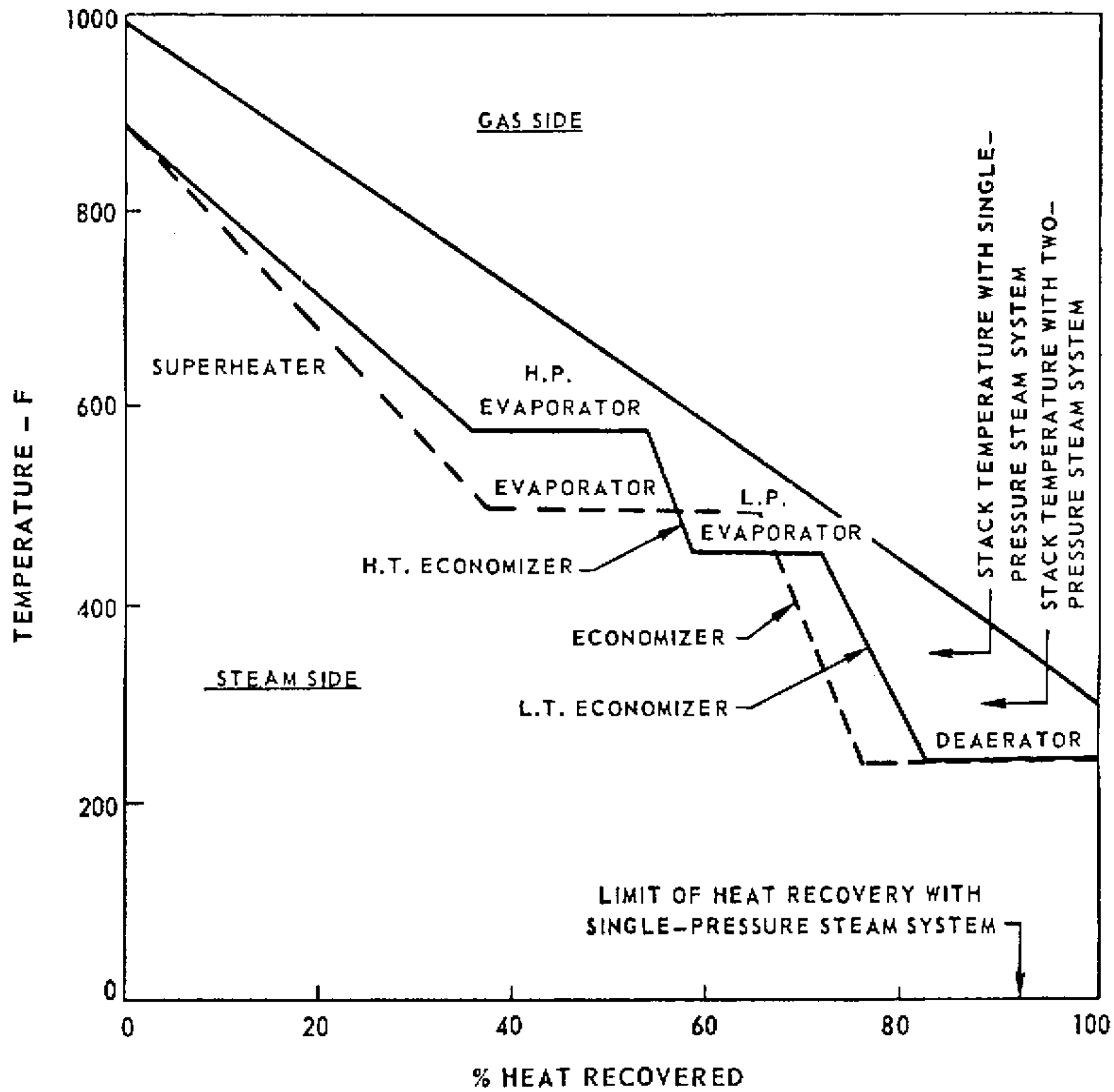
TEMPERATURE DISTRIBUTION IN WASTE HEAT RECOVERY BOILER FOR OIL-FIRED COGAS STATION

MID – 1970'S TECHNOLOGY

TURBINE INLET TEMPERATURE = 2200F

COMPRESSOR PRESSURE RATIO = 16

FIGURE 8.1



9 WESTINGHOUSE SYSTEM

9.1 INTRODUCTION

Extensive telephone conversations with Mr. A. Finizio of Westinghouse resulted in the suggestion that the paper prepared for presentation to the "Second Annual Symposium on Coal Gasification" at the University of Pittsburgh (Ref. W.1.) should be used as a basis for the Westinghouse proposal.

The combined cycle plant is similar to that offered by General Electric and the paper presents a graph showing how the advantage of using a pressurized boiler is nullified at gas inlet temperatures above 2000°F. (see Fig. 9.3) Mr. Finizio indicated that the inlet temperature proposed in the paper of 2200 °F was quite feasible for base load plant in the near future.

The greater part of the paper compares the advantages and disadvantages of four types of gasifier — fixed bed, entrained flow, fluidized bed and molten salt, all of which could be considered for power generation. Westinghouse concludes that fluidized bed gasification has advantages over the others.

Table 9.1 shows a schedule for the proposed development of this gasification combined cycle system which is taken from Ref. W.2.

TABLE 9.1
SCHEDULE OF DEVELOPMENT FOR THE FLUIDIZED BED/COMBINED CYCLE PLANT

	COMPLETION DATE
Develop and Operate Multiple Fluidized-Bed	March 1975
Select Gasifier Concept for Further Development	April 1975
Scale Up Concept and, if Necessary, Build and Operate 5 Ton/hr Gasifier Pilot Plant	September 1977
Complete Design of Generating Pilot Plant for the Dresser Station — Terre Haute, Indiana	September 1978
Complete Construction of Generating Pilot Plant	August 1979
Operate Combined Cycle Plant with Coal Gasifier	May 1981

9.2 FLUIDIZED-BED GASIFICATION

In a fluidized-bed process — shown schematically in Figure 9.1 — the solid phase (coal-char-ash) is supported by a pressure difference created by the flow of gases through the bed. In this fluidized state the solid particles are in random motion within the fluidizing medium, and take on liquid-like characteristics. The main characteristic of a fluidized-bed process is the virtual elimination of temperature zones corresponding to predominantly exothermic and endothermic reactions. The net effect is essentially a mixed temperature dictated by the relative rates of combustion and gasification reactions.

This temperature is generally controlled to just below ash softening temperature to avoid ash agglomeration in the bed. Unless suitably designed, ash agglomeration may cause a loss of fluidization.

TABLE 9.2
FLUIDIZED-BED GASIFIER CHARACTERISTICS

Solid Phase of coal-char-ash supported by gases
Solids and gases fully mixed

ADVANTAGES	DISADVANTAGES
1. Provides superior solids-gas contact.	1. Moderately high loss of sensible heat in product gases.
2. Can tolerate wide variety of fuel quality and particle size.	2. High carry-over loss in char entrained in product gases.
3. High capacity per unit ground area.	3. Loss due to char in ash residue removed from bed.
4. Can be operated over a wide range of output, restricted only by the fluidization characteristics of the solids mixture.	4. Fluidization phenomenon sensitive to fuel characteristics. Strongly caking coals require pretreatment.
5. High degree of process reliability, stability, and safety due to high fuel inventory.	
6. High degree of process uniformity.	
7. Product gases are free of tars.	

The analysis published in the paper identifies the advantages and disadvantages of each gasifier system and concludes that for combined cycle application the requirement is for a gasifier which includes wet scrubbing or particulate and sulphur removal, and operates on air at a pressure close to that of the combined cycle combuster. A further evaluation is made using an appraisal of the following factors:

Design Technology
Operation on various coals
Coal Utilization
Undesirable Carryover
Mechanical Complexity
Tar Production

The paper describes the characteristics of each process under the six headings and rates them in an approximate order of merit, high, average, low. From this assessment it appears that the choice lies between fluidized bed and entrained flow gasification. Comparisons are made between these two on the basis of control and response, turn down ratio, thermodynamic advantages. In all these respects the fluidized bed is indicated to be superior.

9.3 FLUID-BED GASIFICATION/POWER CYCLE DESIGN

Five designs are shown in the paper (Ref. W.1.) from which one — design D — has been selected as appropriate to the Hat Creek application.

The design includes a pressurized air blown fluidized bed gasifier with external cooling and wet scrubbing of the fuel gas and sulphur removal. A gas turbine with a turbine inlet temperature of 2200 °F is included with a heat recovery steam generator. Steam conditions are 1800 psig/970 °F with reheat to 970 °F. Steam for gasification is obtained by flashing surplus feedwater or by extraction from the steam turbine. Air for gasification is taken from the gas turbine compressor discharge at about 16 atmospheres and boosted to over 20 atmospheres by a motor driven compressor. This boost compressor would require approximately 1.5% of gross plant output.

The gasification process is designed to minimize carbon loss in the ash by use of ash agglomeration in the gasifier. The approach temperatures in the HRSG are close: 50°F on the superheater or reheater, 26 °F on the evaporator and 30 °F on the economizer. Table 9.3 summarizes other plant performance data. A schematic diagram of the system is shown in Figure 9.2

TABLE 9.3

Gasifier Pressure	20	ata
Number of gas turbines	2	
Gas turbine inlet temperature	2200	°F
Sulphur removal temperature	200	°F
Gas turbine power	260.5	MW
Steam turbine power	224.7	MW
Auxiliary power	16.1	MW
Net plant output	469.1	MW
Plant heat rate HHV	8100	Btu/kWh
Efficiency	42	%

In order to make maximum use of the heat from the gas turbine exhaust a somewhat complicated feedwater/steam system is employed. A simple reheat system would be unable to absorb the low temperature heat available so the balance is recovered by additional feedwater which is overflowed from the steam drum into flash drums connected to the inlet to the steam reheater and the cross-over pipe between the I.P. and L.P. turbines. The balance of feedwater flows into the deaerator at 20 psia and is recirculated to the economizer.

9.4 COST EVALUATION

The estimate is based on costs derived from the referenced reports updated to September, 1975 levels, as shown in the following tabulation. The specific cost is: $469.8 per kW.

TABLE 9.4

COAL GASIFICATION COMBINED CYCLE 500 MW UNIT
(NET OUTPUT 469.1 MW)

ITEM	DESCRIPTION	COST IN $/KW WESTINGHOUSE MID — 1972	COST IN $/KW SECo SEPTEMBER 1975
1	Land	.5	.7
2	Structures and site Facilities	14.0	19.6
3	Boiler Plant Equipment	13.0	18.2
4	Turbine Plant Equipment	52.0	72.8
5	Electric Plant Equipment	15.5	21.7
6	Misc. Plant Equipment	3.0	4.2
7	Cooling Towers (allowance)	10.0	5.7*
8	Sub-total	108.0	142.9
9	Contingency 15%	16.2	21.4
10	Sub-total for Power Plant	124.2	164.3
11	Engineering and Supervision	N/A	13.1
12	Allowance for Gasification Plant (contingency assumed included)	250.0	250.0
13	Engineering re No. 12	N/A	20.0
14	Corporate overhead 5%	N/A	22.4
15	Anticipated Total $/kW	N/A	469.8

(*computed from INTEG figure: $\$11.87 \text{ per kW} \times \frac{224.7}{469.1} = \$5.68/\text{kW}$).

REFERENCES

W1 "Electric Power from Low-Btu Gas in Combined Cycle Plant"
— R.W. Foster — Pegg, M.L. Jaeger, D.L. Leight — 2nd Annual Symposium Coal Gasification, Lignification and Utilization — Best Prospects for Commercialization, University of Pittsburgh August 1975

W2 Westinghouse ENGINEER July 1975

W3 Anon. "Investment Cost Study" WASH 1230 prepared for the Atomic Energy Commission by the United Engineers and Constructors, Inc. 1972.

FIGURE 9.1 The proposed multistage fluidized bed gasification process combines the sulphur removal task with the coal gasification process to provide efficient and economic generation of clean fuel gas. (Westinghouse)

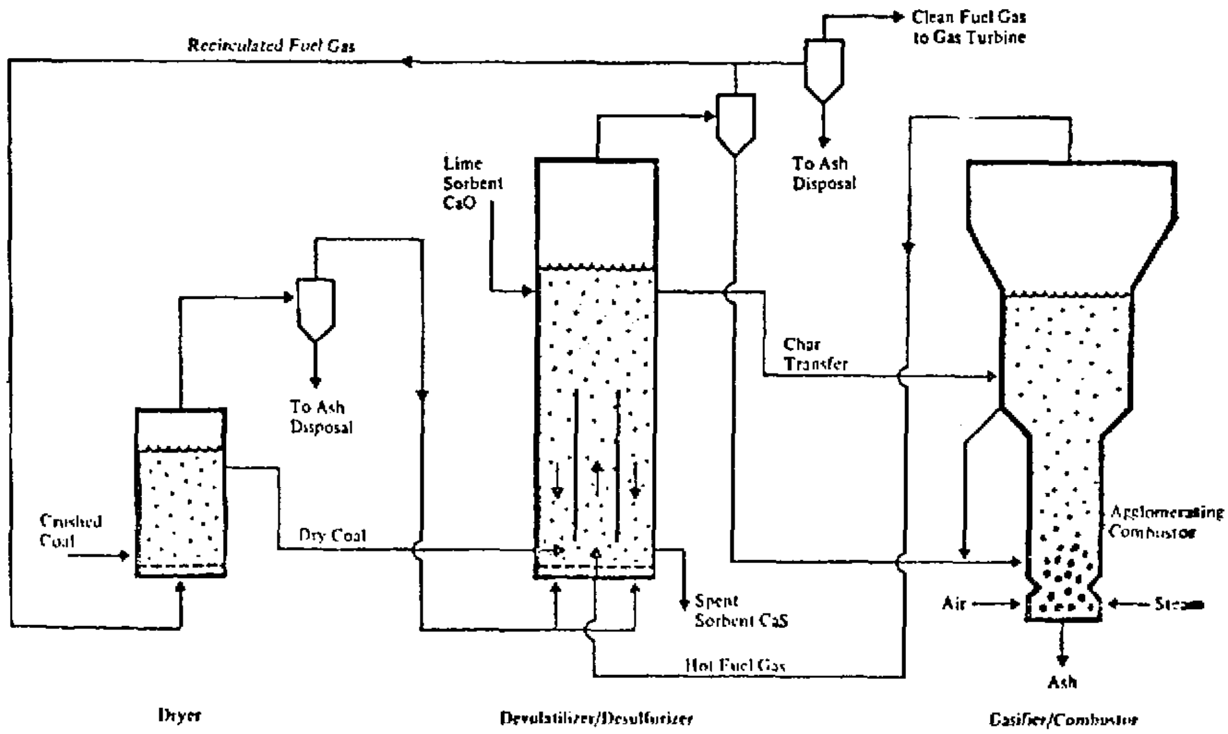


FIGURE 9.2 Schematic Diagram of the Fluidized Bed/Combined Cycle Plant for 469 MW Net Output (Westinghouse)

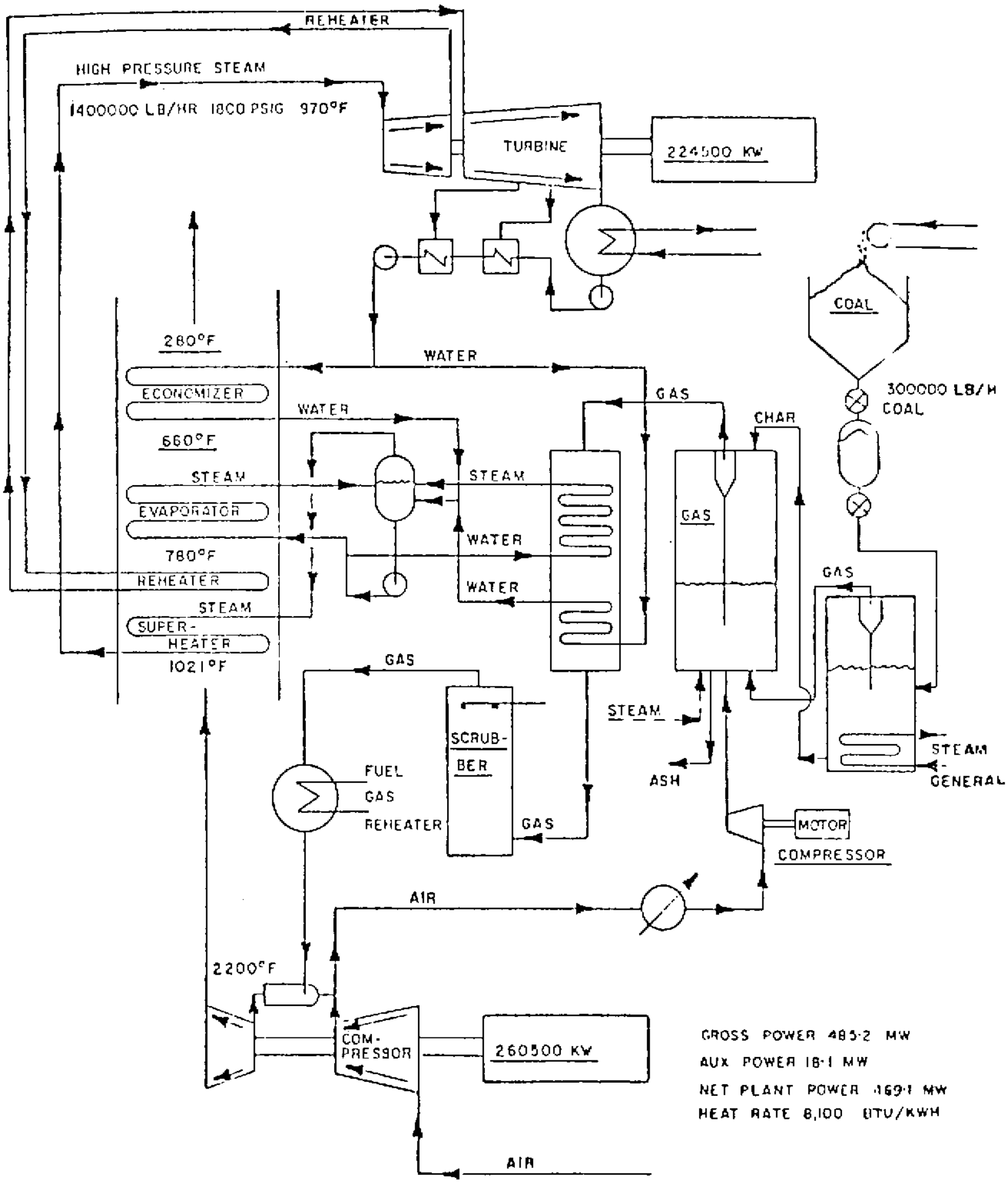
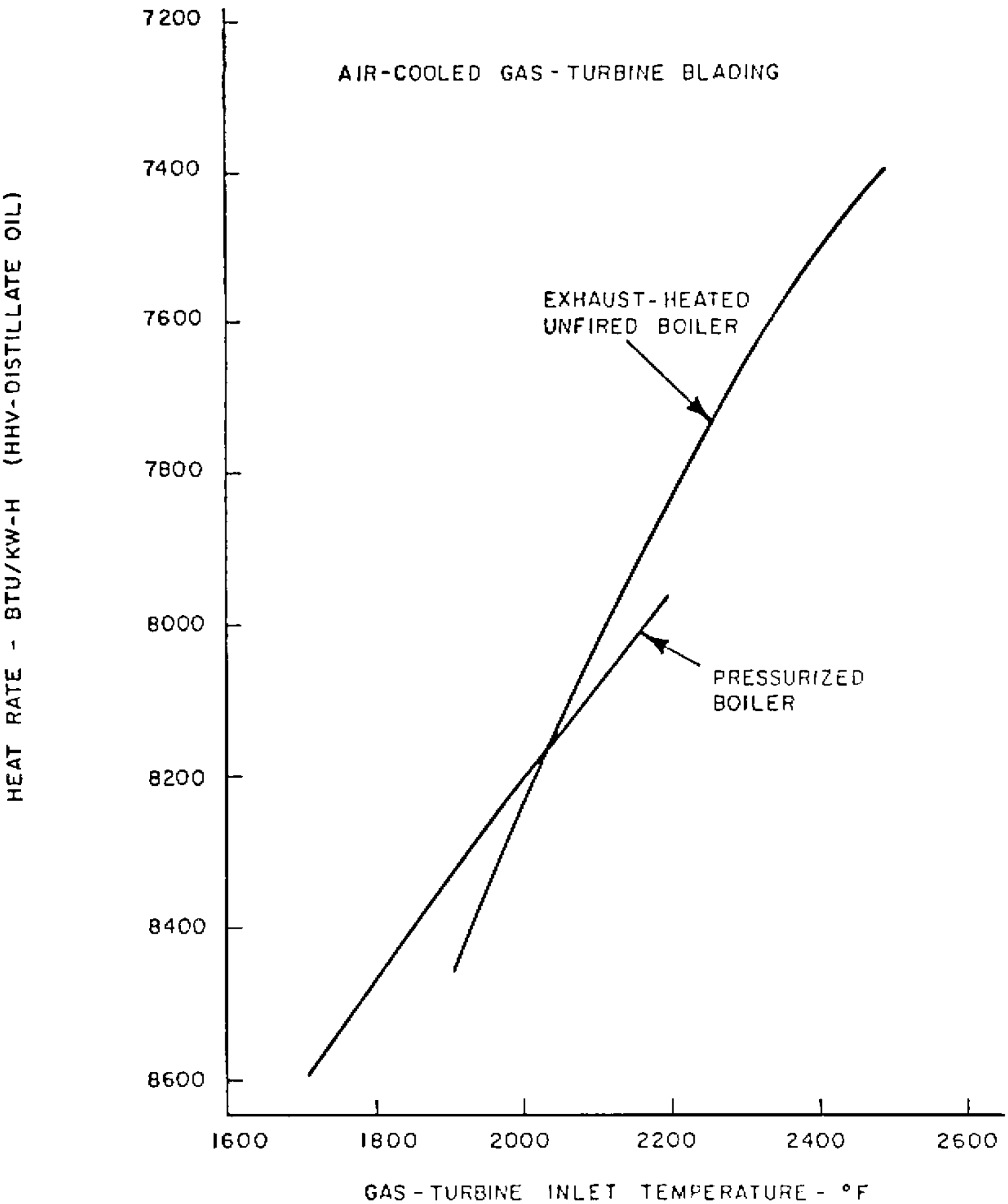


FIGURE 9.3 Performance comparison between exhaust-heated unfired-boiler and pressurized-boiler combined cycles. (Westinghouse)



10. PILOT PROJECT

It appears that the development of an independent Canadian coal gasification technology is considered beyond the means of Canadian research and development. Fortunately, the Canadian energy dilemma is not nearly as acute as that of the United States. Also, the majority of Canadian coals capable of being strip-mined are non-agglomerating and are consequently suitable for commercially available process.

However, the systematic development of an existing, indigenous coal conversion technology appears to be a must for Canada. In this connection it is conceivable, that current and future research efforts and development goals could be effectively served by a scheme, such as proposed herein, at reduced overall costs.

It is recognized that the applicable technology is relatively novel and therefore the introduction of even the most advanced process technology would require demonstration for and adaptation to Canadian conditions.

Gasification of coal is the first step in converting coal to a clean fuel gas, synthesis gas and eventually to synthetic products such as hydrogen, ammonia, methanol, substitute natural gas, etc. The front-end of the technology to reproduce these products is essentially the same. In producing fuel-gas, air can be used instead of oxygen. Thus the process is both cheaper and simpler. Also, the production of fuel-gas shows promise of being commercially viable on a scale magnitude less than the scale of operations necessary for substitute natural gas. Furthermore, power generation from gasified coal can be competitive with conventional coal fired plants and with important incidental advantages over a conventional plant.

Therefore, in considering the introduction of coal gasification technology into Canada, it seems logical to start with the generation of electricity and proceed from there to the next and subsequent processing steps required for other synthetic products, after experience had been gained with the technology concerned.

The scheme envisages a commercial size — not pilot plant size — gasification plant preferably installed at an existing conventional coal-fired generating plant. This would easily and economically assure the plant of operating staff, support services, a fuel supply and a market for the electricity produced. Such a plant could also be used as a test facility and provide a base for research and development required for the future expansion of coal gasification technology. At the same time the plant should be largely self-supporting from the sale of electricity produced.

If the above reasoning is accepted, then it is suggested that a dual purpose facility be built adjacent to an existing, coal fired power station. The facility should consist initially of a Lurgi-type coal gasification plant and a STEAG-type combined cycle plant for power generation built with adequate provisions and features to also serve as a testing facility.

The first Table, which follows, is a graphical presentation of the particular objectives, steps and effects of this proposal in four major categories of endeavour listed below.

RESEARCH, DEVELOPMENT, TESTING

The scheme proposed would introduce into Canada the technology of coal gasification on a demonstration scale and pave the way toward future research, development and testing, with a built-in opportunity to rapidly gain commercially valuable expertise and experience in this field.

POWER GENERATION VIA COAL GASIFICATION

The proposed plant would provide clean, efficient power from coal and through the use of combined cycle technique would open the field toward high efficiency, low-cost, water conserving and non-polluting future power plants.

SYNTHESIS GAS FROM COAL

The plant would provide the basic, initial facilities essential for the utilization of coal gasification products in the manufacture of ammonia and other synthetic products, in order to augment the manufacture of same now obtained from natural gas and from petro-chemical feedstock.

SNG FROM COAL

Through gradual development of the technology and by addition of appropriate process steps, the plant could be extended to produce substitute natural gas and/or serve as a model for large scale SNG facilities built elsewhere. Included in this category are the full scale tests of any type of coal to determine its suitability for gasification, shift conversion and methanation.

The second Table shows the processing steps required to obtain these products from coal.

In the interests of minimum capital cost and minimum time to bring the plant into operation, we suggest that the plant should be based upon the components of the existing operating Lunen plant but with fewer units and, therefore, be smaller in size. Specifically, we suggest it should utilize the same supercharged boiler as is utilized at Lunen, but only one of these boilers instead of two as at Lunen, and use the same gasifier units as are used at Lunen, but only three such units instead of five as at Lunen. The combustion turbo generator would be the nearest standard available unit of about 30/40 MW in rating, and the steam turbo generator would be the nearest standard available non-reheat unit about 60 MW in rating.

The result would be a plant with the following characteristics:

(1) It would have an electrical output of approximately 100 MW.

(2) It would meet the most stringent requirements as regards pollution of the environment.

(3) Its cost and overall efficiency should be comparable with a conventional plant of the same capacity, if no reheating is used in both cases.

(4) There should be the minimum of teething troubles provided the principle was strictly observed of profiting to the full from Lunen experience.

(5) Any two of the three gasifiers would be adequate for full load, with the third available for maintenance, as standby, or as a test facility for different coals. Any or all of the three gasifiers could be arranged for blowing with oxygen as well as air in order to extend their versatility for test purposes.

(6) Excess fines in the coal supply to the plant could be disposed of by using these as fuel for the conventional plant at the same site.

(7) The time required for completion of the plant should not be any greater, and might well be less, than for a conventional plant.

(8) In the event of temporary complete shutdown of the gasifier section of the plant, the plant would be operable at full load on either natural gas or a suitable oil as fuel. The existence of such a plant would offer the following possibilities:

a) The ability to carry out full scale tests on any type of coal to determine its suitability for gasifying, whether such coal was intended for gasifying to produce power, SNG, or a chemical feedstock. There does not appear to be any reason why such a plant should not be designed to incorporate facilities for full scale testing of gasifying tar sands coke. Equally, a test facility for the shift conversion and methanation step to upgrade the gas to SNG could also be added.

b) Different systems could be incorporated as desired in the gas clean-up part of the plant (allowance having been made for this in the initial design) to enable the gas produced to be treated in different ways when testing different types of fuel for gasification, e.g., a sulphur removal system if and when testing the gasifiability of high sulphur content coke.

c) The proving out of the STEAG system of producing electricity from coal, since this system does appear to offer substantial advantages over the conventional system, and British Columbia does appear to have a substantial interest for some time to come in the production of electricity from coal.

d) The introduction into Canada of the technology of coal gasification and the opportunity to build up local expertise and experience in this field.

Such a plant is envisaged as serving the dual functions of being at one and the same time both a commercial plant and a major Canadian test and research facility. To the extent that it was designed to fill the second function as well as the first, the cost may be increased. However, this incremental cost would certainly be substantially less than the cost of any plant designed solely as a test and research facility.

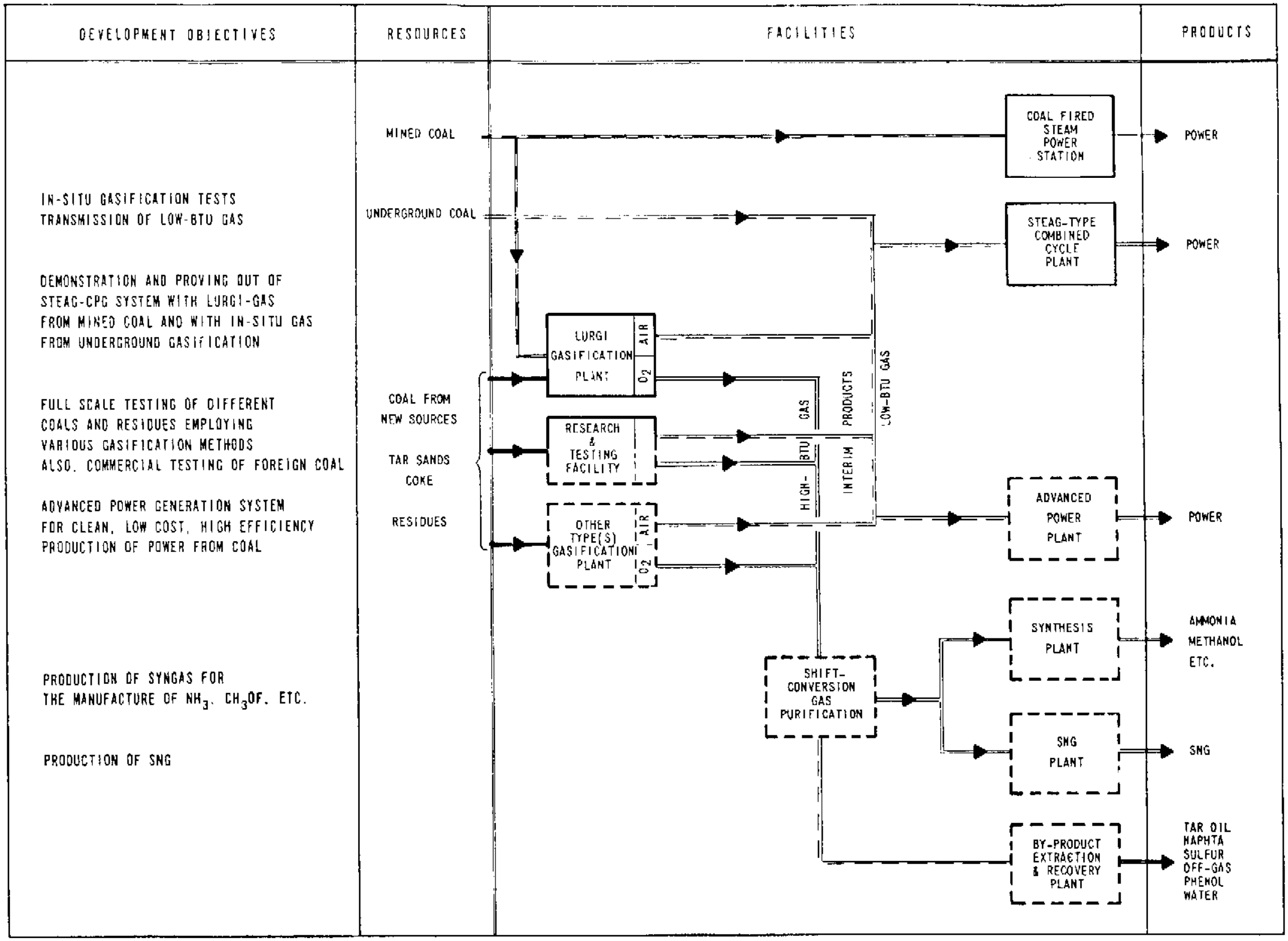
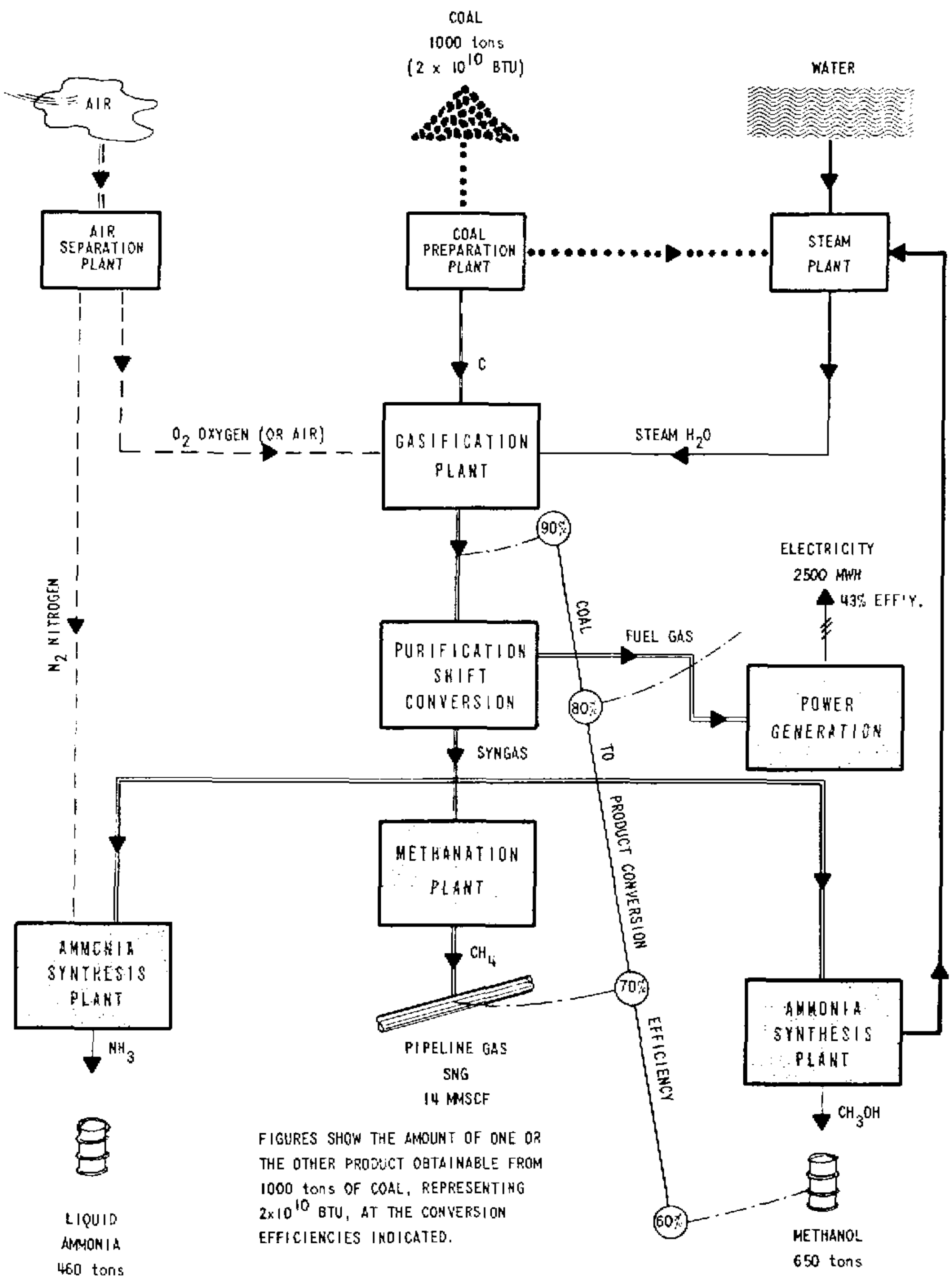


FIGURE 10.1
DIAGRAM OF GASIFICATION/COMBINED CYCLE PILOT PLANT

FIGURE 10.2
FROM COAL TO SYNTHETIC PRODUCT — PROCESSING STEPS



APPENDICES

1. LURGI LAB TEST REPORT

2. STEAG ANLAGENTECHNIK PUBLICATIONS

Combined gas/steam turbine generating plants with bituminous coal high-pressure gasification.
Steam generators of special design.

3. STEAG DRAWINGS

Series 30 18 0995/

8.020	400 MW KDV KOMBI-BLOCK	— Flow Diagram
8.004	2 x 500 MW KOMBI-BLOCK	— Layout
8.005	500 MW KOMBI-BLOCK	— Layout
8.006	500 MW KOMBI-BLOCK	— Section for Layout
8.007	1000 MW KOMBI-BLOCK	— Longitudinal Section
8.008	1000 MW KOMBI-BLOCK	— General Arrangement
8.009	1000 MW KOMBI-BLOCK	— Section through Coal Conversion Plant
8.010	SCHEDULE FOR KDV KOMBI-BLOCKS	

APPENDIX 1

LURGI MINERALÖLTECHNIK GMBH

FRANKFURT (MAIN)

DOLMAGE CAMPBELL & ASSOCIATES LTD.

Via Lurgi Canada ltd.
100 Adelaide St. W
TORONTO, Canada M5H 1S3

Re.: Examination of Hat Creek Coal,
drill holes No. 74-38, 916 - 1036 ft.
Lurgi Code No. 43/75

Gentlemen,

On July 7th, 1975, we received the above mentioned sample. The attached laboratory report No. 112/75 summarizes the result of analyses and tests carried out on the sample. On the basis of these findings, we would like to comments as follows:

1. Based on the values for equilibrium moisture and oxygen content the sample represents a lignite type of coal.
2. The ash content is relatively high. While this is no drawback on the gasification process per se, it does, of course, mean that accordingly more dead solids have to be handled.
3. The ash melting behavior is excellent, with the ash melting, under both reducing and oxidizing conditions, at temperatures in excess of 1,500°C (2,700°F). This can be explained with the rather high silica and alumina contents, accompanied by low concentrations of alkaline oxides. For the actual process, the high ash melting temperatures result in a favorably low steam to oxygen ratio, i.e. high steam decomposition.
4. Although the equilibrium moisture was determined at 20.5%, the sample dried off readily, during lab storage, to around 10%. While the process can easily handle the 20%, the 10% would permit a higher thruput and is therefore preferable. Thus, further investigations of the moisture content on an as-mined basis, as well as of drying during stock-piling, are warranted.
5. Low temperature carbonization and screen analysis of feed and char are tests designed to give information about the degree of disintegration during carbonization. As can be seen from the attached diagrams, very little disintegration does take place and the amount of 1/8" material formed in the process is very small. This result is very advantageous for the gasification process.
6. The reactivity of the sample is somewhat below what is typical for that kind of coal and consequently, the specific oxygen consumption will be slightly higher (although it is still well below of that of, say, caking coals). However, this is expected to be more than offset by the lower steam requirements.
7. The Pressure Reick Degassing test provides data needed to predict the crude gas analysis.

8. The sample's chlorine content is absolutely normal and will pose no problems.

CONCLUSIONS

The coal as represented by the sample submitted to us makes an excellent feedstock for Lurgi gasification. Ash melting characteristics are very favorable so that a low steam to oxygen ratio can be expected. The specific oxygen consumption, though expected to be slightly higher than typical for this kind of coal, is still well below of that of, e.g., caking coals. Also, the low steam requirements are likely to offset this penalty. The fact that very little dust is being formed during carbonization is very advantageous and will help ensure a smooth operation. The somewhat above normal ash content does not affect the process per se, it just means accordingly more solids handling.

We trust the above information will help the development of your project. Should you have any further questions, please feel free to contact us.

Very truly yours,

LURGI MINERALÖLTECHNIK GMBH

ANALYTICAL TEST REPORT NO. 112/75
BGD-50-3910

SAMPLE IDENTIFICATION: **Dolmage Campbell & Associates Ltd., Canada (for British Columbia Hydro and Power Authority):** Composite sample of **Hat Creek Coal** deposit, drill holes No. 74-38, 916-1036 ft, received on 7th July, 1975, Lurgi Code No. 43/75

MOISTURE AS RECEIVED: n.d., surface-dry; carbonate contents (as CO_2) = 1.3 wt%

PROXIMATE ANALYSIS
(from carbonate-free sample, recalculated to original)

		m	maf
Moisture	wt%	9.7	-
Ash (815 °C)	"	29.3	-
Volatile (900 °C)	"	26.8	43.9
Fixed Carbon	"	34.2	56.1
Equilibrium Moisture	wt%		20.5

CARBONIZATION ASSAY:
(Fischer, 520 °C)

		m	maf
Moisture	wt%	9.7	-
Gas Liquor	"	5.1	5.6
Tar	"	6.1	6.8
Coke	"	70.2	77.7
Gas and Loss	"	8.9	9.9

CALORIFIC VALUES:

	m	maf
HCV MJ/kg (kcal/kg)	16.9 (4046)	27.8 (6633)
LCV MJ/kg (kcal/kg)	16.0 (3810)	26.2 (6246)

FUSION PROPERTIES OF ASH: (prepared at 815 °C)
(LEITZ HEATING MICROSCOPE)

Atmosphere:		Reducing	Oxidizing
Softening Point	°C		
Melting Point	°C	1500	1500
Flow Point	°C		
Appendix No.		1	1a

SPECIFIC ASHING TEST
Input Coal (5 - 30 mm)
Moisture wt% n.d. Sulphur wt% n.d.
Ash wt/ n.d. Chlorine wt% n.d.

Ash Analysis	**% Volatilization**
Sulphur wt% n.d.	n.d.
Chlorine wt% n.d.	n.d.

Ash Characteristics
Fine material of grey appearance; most of the ash consists of whitish-brown, dense, but fissured, lightly sintered pieces but also some fused pieces of brown to mauve colour. (ash strongly resembles that of Sigma Mine Coal, SASOL)

ULTIMATE ANALYSIS:
(from carbonate-free sample, recalculated to original)

		m	maf
Moisture	wt%	9.7	-
Ash	"	29.3	-
Carbon	"	41.5	68.03
Hydrogen	"	3.4	5.57
x) Nitrogen	"	0.80	1.31
x) Combust. Sulphur	"	0.26	0.43
x) Chlorine	"	0.03	0.05
Oxygen (diff.)	"	15.01	24.61

x) from original

SULPHUR:		m	maf
Total Sulphur	wt%	0.35	-
Pyritic Sulphur	"	0.09	-
Organic Sulphur	"	0.25	-
Sulphate Sulphur	"	0.01	-

MINERAL ANALYSIS:	**wt%**
Silica, SiO_2	54.3
Alumina, $A1_2O_3$	34.0
Ferric Oxide, Fe_2O_3	4.5
Magnesia, M_3O	1.0
Lime, CaO	1.6
Sodium Oxide, Na_2O	1.0
Potassium Oxide, K_2O	0.3
Sulphur trioxide, SO_3	0.4
Barium Oxide, BaO	0.2
Titania, TiO_2	1.2
Phos. pentoxide, P_2O_5	0.12
Undetermined	1.38

Sample:

*** LTC-DISINTEGRATION TEST/SCREEN ANALYSIS:**

Screen Fraction	feed	char
40 mm - 50 mm wt%		
30 mm - 40 mm wt%		
25 mm - 30 mm wt%	47.0	22.8
20 mm - 25 mm wt%	25.0	31.4
15 mm - 20 mm wt%	14.7	21.8
10 mm - 15 mm wt%	9.0	13.2
5 mm - 10 mm wt%	4.3	7.0
3 mm - 50 mm wt%		2.0
2 mm - 3 mm wt%		0.4
1 mm - 2 mm wt%		0.5
0.8 mm - 1 mm wt%		0.9
0.5 mm - 0.8 mm wt%		
0.315 mm - 0.5 mm wt%		
0.2 mm - 0.315 mm wt%		
0.1 mm - 0.2 mm wt%		
0.063 mm - 0.1 mm wt%		
- 0.063 mm wt%		
Appendix No.		2
Bulk Density g/cm^3	0.678	0.558
Mean Grain Size mm	22.570	19.427
in % of feed		86.1
Yield of Char wt%		65.2

Description of Char wt%

Essentially non-fissured pieces of dense and hard structure, specifically quite heavy.
No dust formation.

*LTC - Low Temperature Carbonization

PRESSURE REICK DEGASSING;

Total gas N l/100 g (690°C) 11.93

Gas Composition (N_2-free)		C_2-C_4-Hydrocarbons
CO_2	Vol% 33.7	C_2H_6 Vol% 2.7
	Vol% -	C_2H_4 Vol% 0.1
CO	Vol% 5.5	C_3H_8 Vol% 0.7
H_2	Vol% 17.1	C_3H_6 Vol% 0.2
CH_4	Vol% 39.5	C_4H_{10} Vol% 0.2
		C_4H_8 Vol% 0.3

Density kg/Nm^3 1.102
HCV kcal/Nm^3 5271
LCV kcal/Nm^3 4740
Residue, wt % 64.6 Moisture of Input, wt% 10.0

PRESSURE REACTIVITY:

CO_2-Conversion Vol%: 6.67 CO, Vol: % 12.5
per gram/input: 0.37 Input coke, g: 18.05
Reactivity Nml CO/g/sec: 0.044 Coke Temp. °C: 800
Temperature °C 800

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TABLE 16
COMPARISON STEAG COAL WITH HAT CREEK COAL

	STEAG COAL	HAT CREEK COAL
Moisture	12%	20.41%
Ash	20%	25.0%
NHV	5660 kcal/kg	3457.23 kcal/kg
NHV per ton maf	8323.5 kcal/kkg	6352.8 kcal/kg
Heat to steam in jacket	1.4% = 79.24 kcal/kg	189.775 kg = 110.069 kcal = 3.174%
CO_2 vol % in gas	9.0	15.25
CO	10.2	15.89
CH_n	3.2	4.66
CnHm	0.3	0.56
H_2	16.0	21.51
N_2	25.6	21.56
H_2O	34.3	20.06
H_2S	0.2	0.13
Kcal/Nm³ NHV	1200	1,817.7
Gas exit temp °C	620	265
Thermal efficiency of gasifier %	93.7% (?)	70.1 (93.4)%
Nm³ gas/ton coal from η and NHV in gas	4,716.7	1,628.675
Ton steam/ton coal required	0.671	0.5724
Kg steam/kcal NHV of coal	0.1185	0.1652
Kg air/Mcal	0.3225	0.1654

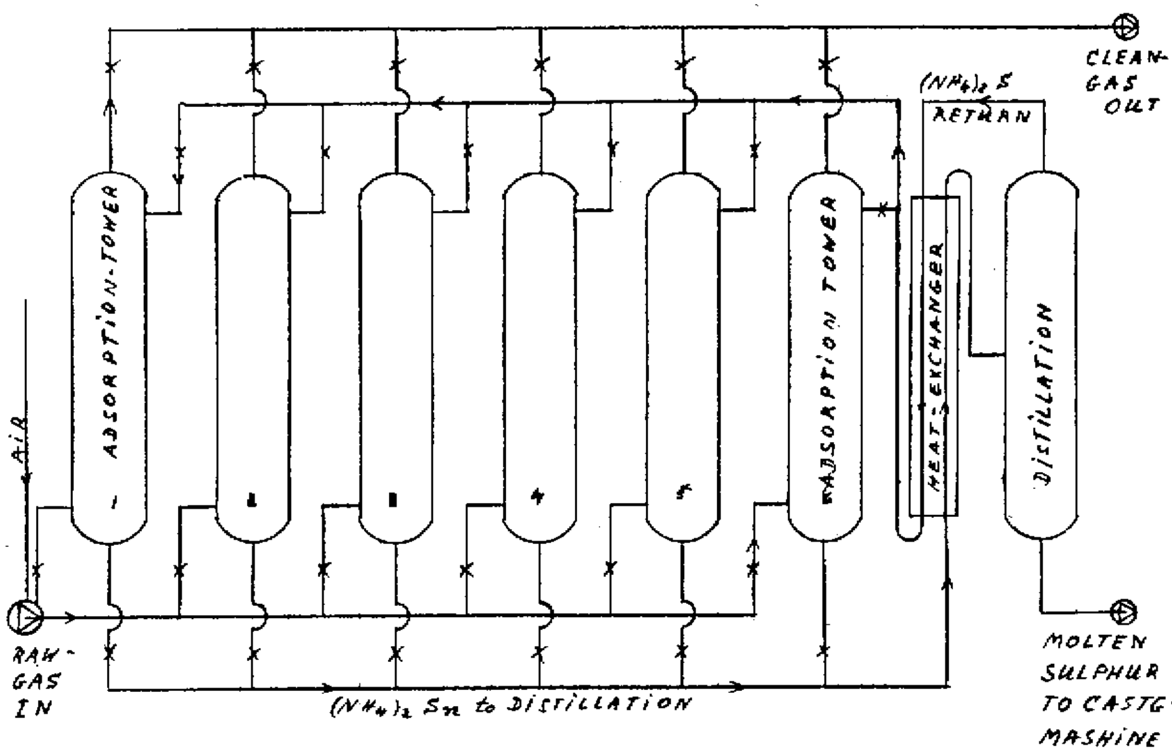

Fig 4 Active Carbon Gas-Desulphurisation

APPENDIX 3

COMBINED GAS/STEAM TURBINE GENERATING PLANT WITH BUTIMINOUS COAL HIGH-PRESSURE GASIFICATION PLANT AT THE KELLERMANN POWER STATION, LÜNEN

Dr.-Ing. Dr. rer. pol. K. Bund
Dr.-Ing. K.-A. Henney
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The demand for electricity throughout the world is still being met predominantly by fossil fuels, and these sources of energy will continue to provide a considerable proportion of the electricity generated for some time to come. Conversion of these fuels into electric power has long been accomplished almost exclusively in plants employing the conventional steam cycle process. Gas turbines have been used only in very few cases so far for generating electrical energy.

The conventional steam cycle process has almost reached the end of its development capabilities. Efficiency can only be further improved by using higher pressures and temperatures and not by enlarging the unit size of the equipment. Plants with a high steam temperature, which operate in the supercritical range and thus necessitate the use of austenitic steels in the high-pressure section, have been built only in special cases because of the high cost of materials and consequent increase in the overall cost of the plant (1;2).

By introducing the gas turbine into power station engineering, for example in the form of the combined gas/steam turbine cycle process described below, it is possible to achieve higher efficiencies and a simultaneous reduction in capital cost as compared with the more conventional plants. This applies particularly where the inlet temperature of the gas turbine can be stepped up even further and the steam generator is a boiler of the pressurized type (3).

So far, only a small number of high-capacity combined gas/steam turbine generating plants have been built (4; 5), the main reason being that in the past the permissible inlet temperatures for gas turbines were limited to figures below the now usual 800° to 900°C; however, plants with inlet temperatures of up to 1100°C have already been tested, and may shortly be expected to have reached the stage where they can be put into commercial service.

Gases and light fuel oils may be employed as fuel for such combined cycle processes. With the present state of the art, coal can only be used if it is first gasified. Such a process, however, has the advantage of permitting the waste gases to be discharged directly into the atmosphere free from dust and, by incorporating an H_2S scrubber behind the high-pressure gasifier, virtually free from SO_2 as well.

The development prospects offered by this type of plant, and the resultant advantages in efficiency and capital cost, induced the Steinkohlen-Eletrizitat AG (Steag) in the spring of 1969 to build a prototype coal gasification plant. At present, this plant is being erected at the Kellermann Power Station of Steag at Lunen, near Dortmund.

THE COMBINED GAS/STEAM CYCLE PROCESS

DESCRIPTION OF THE PROCESS

The possibility of combining coal gasification units with gas turbine systems has frequently been discussed in literature without, however, leading to any decision to build such plants (6;7). After thoroughly investigating all the requirements and present technical circumstances, Steag decided on the gas/steam cycle process, as shown in simplified form in Fig. 1.

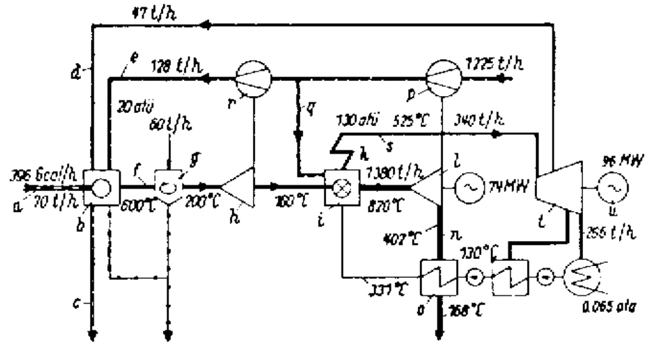


Fig 1 Combined Gas/Steam Turbine Cycle

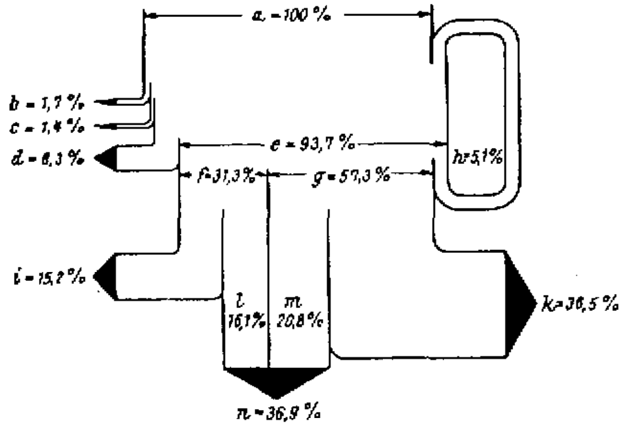
a	coal	l	main gas turbine
b	gasifier	m	generator
c	ash	n	waste gases
d	steam	o	preheater
e	air	p	air compressor
f	fuel gas	q	combustion air
g	scrubber-cooler	r	gasifying air compressor
h	expansion turbine	s	live steam
i	combustion chamber	t	steam turbine
k	combustion gas	u	generator

The coal (a) is gasified in the gasifier (b) and the ash (c) is removed from the process. The gasifying media are steam (d) and air (e). The fuel gas (f) so produced leaves the gasifier at a temperature of 600°C, passes through the scrubber-cooler (g) and enters the expansion turbine (h) at a pressure of approx. 20 atms. gauge and a temperature of 200°C. There, the pressure of the fuel gas is reduced to about 10 atms. gauge. The fuel gas then passes to the combustion chamber (i) in the pressurized boiler. The combustion gas (k) from the boiler enters the main gas turbine (1) at a temperature of 820°C. The waste gases are used for heating the feed water in the feedwater preheater (o) and in the process are cooled to 168°C. The main gas turbine is coupled with and directly drives the air compressor (p). The compressed air is used as combustion air (q) for the combustion chamber and as gasifying air for the gasifier. The gasifying air is compressed in the compressor (r), which is driven by the expansion turbine, from approx. 10 atms. gauge — the pressure of the gas turbine cycle — to 20 atms. gauge — the pressure of the gasifying process.

Live steam (s) is generated in the pressurized boiler at 130 atms. abs. and 525°C to drive the conventional steam turbine set (t). The gasifying steam is bled from an extraction stage of the steam turbine (t).

HEAT FLOW

The heat flow in this circuit, Fig. 2, is governed by the gas/steam turbine cycle process. If the heat chemically combined with the coal is taken as 100%, 36.9% of this heat is converted into electrical energy — 16.1% by the gas turbine and 20.8% by the steam turbine. In addition to the heat in the coal, a further 5.1% of this thermal energy is fed to the gasifier along with the heated gasifying air and gasifying steam as heat circulating in the process.



a heat input from coal
b losses due to ash removal and radiation
c heat of evaporation of the jacket steam
d heat of evaporation of the scrubbing water
e heat input, combustion chamber
f heat in the gas turbine
g heat in the steam turbine
h heat in circulation
i waste gas losses, gas turbine
k cooling-water losses, steam turbine
l electrical energy, gas turbine
m electrical energy, steam turbine
n total electrical energy

Heat losses in the gasification process are accounted for by the ash removed from the gasifier, unburnt fuel in the ash, radiation losses of the water-cooled gasifier, and the heat of evaporation of the jacket steam. These losses amount to 3.1% of the total heat input in the gasifier. A further 8.3% of the heat is lost to the gasification process by evaporation of the scrubbing water, since this lowers the temperature of the fuel gas from 600°C to 160°C. However, in the combined gas/steam turbine cycle this does not constitute a loss, since the spray water increases the flow rate through the gas turbine, and in effect we obtain the conversion of sensible heat into mass flow.

Accordingly, with the gas turbine unit selected for Lunen, the flow rate through the gas turbine is 43 kg/s more than the flow rate through the compressor. On the other hand, if natural gas were used as fuel, the flow rate would be only 4 kg/s more. This large increase in the flow rate at Lunen, which is due to the fuel used and which is attributable half to the extra volume of steam required for gasification and half to the evaporation of spray water in the scrubber-cooler, leads to an increase of 19 MW in the power output of the gas turbine as against the power output of the same turbine using natural gas. Thus, the apparent loss of sensible heat in the scrubbing process is more than made up by this effect.

DESCRIPTION OF THE LÜNEN PLANT

PRINCIPLE OF THE GASIFIER

Various methods have been used over the years to gasify bituminous coal and lignite. Up to now, the most successful has been the Lurgi high-pressure gasification process, and based on this process a total of 58 plants have already been built both in Germany and abroad (8; 9; 10).

The Lünen plant features 5 high-pressure gasifiers of this type. They can be charged with non-caking or slightly caking lump coal in sizes between 3 and 30 mm, and with a permissible undersize fraction of up to 7%. Ash contents up to 30% and water contents up to 15% are permissible; the total content of incombustible matter must not exceed 35%.

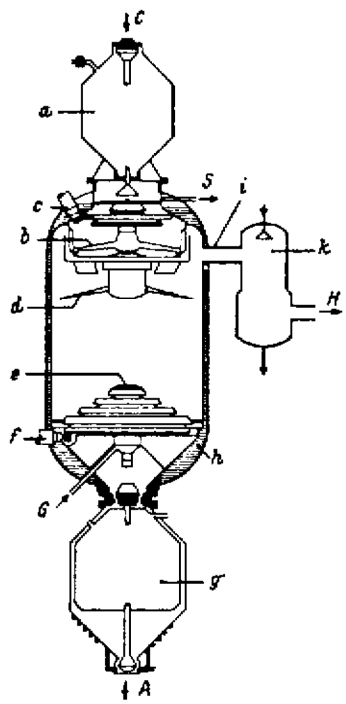


Fig. 3 Gasifier Setup

A ash
C coal
G gasifying media
H fuel gas
S jacket steam

a coal lock
b distributor
c drive of distributor
d agitator blades
e rotating grate
f drive of rotating grate
g ash lock
h water jacket
i gas outlet
k scrubber-cooler

Fig. 3 illustrates the setup of this type of high-pressure gasifier. Coal (C) is fed into the gasifier from above through a fully automatic coal lock (a). A rotating distributor (b) ensures uniform charging, the drive (c) of the distributor being located outside the gasifier. The agitator blades (d) below the distributor prevent agglomeration of the fuel bed when caking coal is used.

As the coal descends in the gasifier it is dried by the counterflowing gas, which also drives off the volatiles and gasifies the coal. The coal that remains ungasified is burnt in a thin combustion layer on a rotating grate (e). This generates the heat required for the process. The grate drive (f) is of the external type. The grate carries the ash downwards into the automatic ash lock (g), from where the ash (A) is transferred off.

The gasifying media (G) — air and steam — are introduced into the gasifier from below through the grate. The gasifier operates under a pressure of approx. 20 atms. guage and is cooled by a water jacket (h). The steam (S) thus generated is also used in the gasification process. The temperature in the combustion zone is about 1200°C, falling to 600°C at the gas outlet (i) of the gasifier. The fuel gas contains tar and small amounts of dust, which are removed in the scrubber-cooler (k) using hot tar-containing water in closed cycle, part of which is elutriated in a secondary circuit. The cleaned steam-saturated fuel gas (H) leaves the scrubber at a temperature of 160°C. The scrubbing water in the closed cycle passes through a separator to precipitate the tar-dust mixture. This mixture is then pumped back into the gasifier, where it is cracked and gasified.

After scrubbing, the fuel gas has the following analysis:

H_2	=	16.0%	H_2O	=	34.3%
CO	=	10.2%	N_2	=	25.6%
CH_4	=	3.2%	CO_2	=	9.0%
C_nH_m	=	1.2%			
C_nH_m	=	1.2%	H_2S	=	0.2%
NH_3	=	0.3%			

Min. calorific value: 1 200 kcal/m_n^3

Fuel gas readings taken on high-pressure gasification plants in operation show that the solids content after scrubbing is less than 1.5 mg/m_n^3 The solids content is thus below the figure of about 2 mg/m_n^3 stipulated by the gas turbine suppliers. Any compounds of chlorine, sodium or potassium in the gas are washed out in the scrubber-cooler.

The gasifiers at the Lünen plant have an external diameter of 3.5 m and an overall height, including coal and ash locks, of approx. 20 m. They are completely manufactured and assembled in the workshops, so that work at site is limited to erection only. Each gasifier has a coal throughput of 10—15 t/h.

Plants in operation have shown that maintenance of such gasifiers is limited to wearing parts of the seals of the coal and ash locks and to the rotating parts of the grate and coal distributor. The plant is so designed that 4 gasifiers can produce full output while maintenance on the 5th gasifier is in progress.

GAS TURBINE AND PRESSURIZED BOILER

A VF 93 type gas turbine from the "Kraftwerk-Union" was selected for this project, a large number having already been supplied and in satisfactory operation (11). The 17-stage compressor and 4-stage gas turbine feature a common shaft carried in two bearings. The shaft comprises axially tensioned rotor discs featuring Hirth-type serrations.

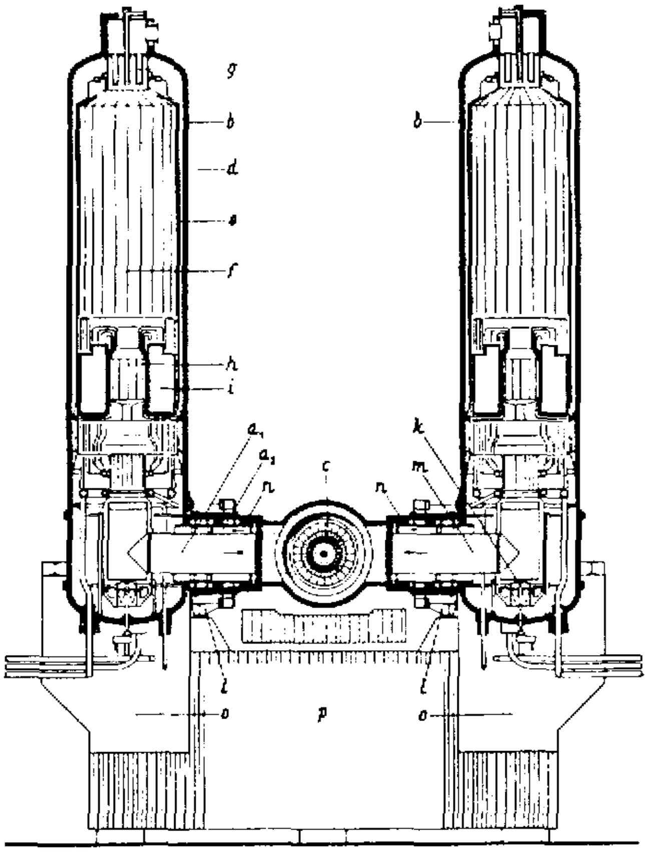


Fig. 4 Arrangement of Pressurized Boilers

a_1	gas inlet	h	displacement body
a_2	air outlet	i	superheater surfaces
b	boiler	k	annular slide valve
c	gas turbine	l	fixing point
d	outer jacket	m	flue gas pipes
e	tube cage	n	corrugated expansion joint
f	evaporating heat surface	o	concrete extensions
g	gas burner	p	gas turbine foundation

As the type of gas turbine employed features two gas inlets (a_1) and two air outlets (a_2) arranged concentrically on both sides of the turbine, the boilers (b) are arranged in two units, one on each side of the gas turbine (c). Each unit features a cylindrical pressure-resistant outer jacket (d). The actual steam generator inside the jacket consists of a cylindrical spirally-wound welded tube cage (e), which forms the evaporating heat surface (f). The combustion air supplied by the compressor at a pressure of 10 atms. gauge flows through the gap between outer jacket and tube cage to the gas burners (g) located at the top of the boiler.

The superheater heating surfaces (i) are of the contact type and arranged around a displacement body (h) in the bottom section. The flue gases leave the boilers at the bottom and enter the gas turbine through pipes (m), the temperature of these gases being regulated by adding air through a sliding cylindrical valve (k).

A new feature of the installation is the direct connection between the pressurized boiler and the gas turbine. Whereas in previous designs the combustion chambers of the gas turbine were flexibly suspended to allow for thermal expansion of the gas turbine, such an arrangement was impracticable for pressurized boilers because of the weight involved — about 100 t per boiler — and because of the bending moments of the incoming pipes. The boilers are therefore rigidly supported at point (l) underneath the flue gas pipes (m) leading to the gas turbine. These flue gas pipes are connected to the gas turbine by two corrugated expansion joints (h), one of which is braced to the boiler by means of a steel framework. The setting on which the boiler is anchored is connected directly to the gas turbine foundation (p) by two concrete extension pieces (o).

The construction of this type of pressurized boiler was first made possible after the development of modern techniques of welded tube walls, and led to unit sizes of 3.6 m external diameter by 18 m high. The boiler units are manufactured and assembled in the workshops, so that erection time at site is reduced to a minimum.

High-velocity pressurized boilers have been built in large numbers as Velox boilers (12). At Lunen, however, the boilers are not operated in the high-velocity range. The flow rates on the flue gas side at 9 m/s and are within normal limits. The draught loss is 2000 mm WG. The firebox has a volume load of 0.42 Gcal/m^3 h atm., a cross-sectional load of 2.65 Gcal/m^2 atm., a surface load of 0.28 Gcal/m$_2$ atm., and a heat-flux density of 0.23 Gcal/m^2 h. With such a high thermal load, particular attention must be paid to uniform distribution of steam in the tubes, and for that reason a comparatively high water flow rate of 3.5 m/s was selected. The pressure drop on the water side of the boiler is 54 atms.

TECHNICAL DATA

The Lunen plant has a total gross electrical output of 170 MW, made up of 96 MW from the steam turbine and 74 MW from the gas turbine. Although the total gas turbine output is 180 MW, 106 MW are required to drive the air compressor, so that only 74 MW are left over for actual output. The only other auxiliary system loads are the drives for the boiler feed pump and the cooling-water pumps, which require 5 MW, i.e. 3%. This figure is lower than in conventional plants as other auxiliary system loads for the boiler and coal crushing plant are not required. The main technical data for the Lunen plant are shown in Fig. 1.

Fig. 5 shows the partial-load performance of the plant in comparison with a conventional 150 MW steam turbine block. At the design point the plant efficiency is 36.9%, corresponding to 2,330 kcal/kWh, as related to gross output. Allowing for auxiliary system loads, the heat rate of the plant is 2,400 kcal/kWh. When assessing these figures it must be remembered that the plant is a prototype and has a relatively low power output in comparison with the unit ratings now usual in conventional power stations.

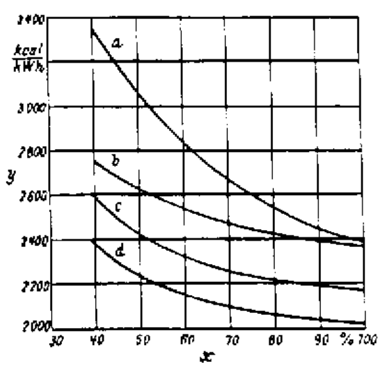


Fig. 5 Partial-Load Performance

- x net output
- y heat rate
- a 165 MW, 820°C combined cycle
- b 150 MW, conventional set
- c 330 MW, 810°C combined cycle
- d 330 MW, 1050°C combined cycle

At half-load operation, efficiency of the plant under discussion is 28%, or 3,050 kcal/kWh. 1,300°C. 130°C, and so the outlet temperature was raised to 168°C. turbine, and is attributable to the fact that in the prototype plant the gas turbine has to be run with a comparatively large air surplus of 2.5 so as not to exceed the gas inlet temperature of 820°C.

CAPITAL COST

Capital expenditure for a conventional coal-fired 150 MW generating unit with reheat equipment is DM 460/kW_{net} on the present price basis and supervising construction, commissioning, and building loan interest. In comparison, the capital expenditure for the combined gas/steam turbine generating plant of 165 MW without reheater is DM 390/kW_{net}. This plant is thus 15% cheaper than a conventional coal-fired plant with reheater.

Arrangements for Future Operation

Before the final decision was made to build the Lunen plant, the possible risks were investigated and assessed. Except for the pressurized boiler, which was a new design, and the connection between boiler and gas turbine, only well proven components were included in the cycle process.

To prevent corrosion when using coal containing chlorine, the gasifiers are lined with plates of Remanite, a steel containing 29% chromium and 9% nickel. The chlorine enters the scrubber-cooler as ammonium chloride and is washed out and then neutralized with soda lye.

From the scrubber, 4 to 6 m^3 of waste water must be elutriated per hour. This waste water contains about 7 g/ltr. phenol, 3 g/ltr. free ammonia, 5 g/ltr. H_2S and 2 g/ltr. fatty acids, as well as 50—70 g/ltr. common salt, depending on the chlorine content in the coal. The waste water can be disposed of by evaporation in an oil-fired kiln, the phenol, ammonia and fatty acids being burnt off in the process. In the Lunen plant an experiment is being made to spray the waste water directly into the existing boiler plant.

The fuel gas is not at the moment desulphurized in the Lunen plant. Sulphur is present in the form of H_2S. To prevent corrosion of the first-stage turbine blades, the solids content — in particular alkalis and alkaline earth — must be less than 1.5 Mg/m_n^3 . Measurements of the dust content have shown that the high-pressure gasification plant can meet this requirement.

Tube bursts in the pressurized boiler, coupled with heavy outflows of water, might possibly cause water droplets to enter the gas turbine. For that reason, the water and steam throughput of the pressurized boiler are measured on a continuous basis. On tube bursts occurring, the plant shuts down automatically. Minor leaks in the boiler tubes cannot damage the gas turbine, since any water leaking out would evaporate immediately at the flue gas temperature of 1300°C.

To protect the expansion turbine from possible precipitation of tar and water when the fuel gas is expanded, the gas is heated from 160° to 210°C before it enters the turbine. Efficiency of the expansion turbine is such that the gas outlet temperature is always above 160°C.

The waste-heat boiler (or economizer), which is located behind the main gas turbine, might be damaged if the temperature of the flue gas drops below the dew point. Although the original intention was to operate with an economizer outlet temperature of

120°C, this proved impractical because the dew point in the flue gas is around 130°C, and so the outlet temperature was raised to 168°C.

In addition, a water/water heat exchanger was installed upstream of the economizer to ensure that the feedwater inlet temperature would not drop below 130°C. In addition, a start-up preheater to preheat the feedwater to 165°C is installed to prevent the gas temperature from dropping below the dew point during start-up operations.

PLANT START-UP AND SHUTDOWN

As the combined gas/steam turbine power plant is to operate in the medium-load range, an important criterion is the behaviour of the plant on a hot start, i.e. after a shutdown period of not more than 8 hours.

On the plant being shut down, it is possible to keep the high-pressure gasifiers under pressure since the operating pressure will drop only slightly due to heat radiation. The gas pipes between the gasifier and the expansion turbine are also kept under pressure to avoid having to blow them out. The steam turbine and part of the associated live-steam pipes are also kept at operating temperature level.

The plant can be started up using the fuel gas from the high-pressure gasification plant; neither gas from an external source not light fuel oil is required.

The hot restart begins with starting up the gasifying plant using steam and air from outside sources. The boiler feed pump is then started and feedwater pumped through the pressurized boiler. It takes approximately 6 minutes to run up the turbine using a starting motor, synchronize it and then load it. Between the 6th and 13th minute the boiler is brought to operating temperature, the water plug being ejected through the starting relief value. After the 13th minute, the steam turbine is run up and synchronized. In 20 minutes at the most after a hot restart, the plant may be run up to full load. The plant is designed for a mean temperature change rate of 30°C/min; however, the permissible rates of change in the lower temperature range are higher and in the upper temperature range lower than the mean value quoted above.

OUTPUT CONTROL

Because of the combined form of construction, it was necessary to coordinate the output control systems of the gas and steam turbines with the output control system of the gasifiers. Separate control systems for the various plant components were not provided. The power output of the entire plant is controlled by varying the flow of fuel gas to the pressurized boiler, the nozzle group valves of the expansion turbine being utilized for this purpose.

On the flow of fuel gas being varied by such action, this will simultaneously alter the pressure in the gasification plant, which can be controlled by adjusting the vanes of the booster fan driven by the expansion turbine. The gasifying air that the air/steam ratio necessary for reliable gasification is maintained for any load condition.

On the combustion gas throughput being varied, the heat input in the pressurized boiler will vary in direct proportion, as will the steam output of the steam turbine. The steam turbine itself is operated using initial-pressure control equipment, the temperature of the live steam being regulated by varying the feedwater flow. The fuel gas flow is taken as the command variable for this control procedure by comparing it with the feedwater flow at any given time.

The gas inlet temperature to the gas turbine is not automatically controlled. When the gas plant is put into service, the air bypass valve behind the pressurized boilers is so adjusted that the maximum permissible inlet temperature is not exceeded. A thermostat is provided which, on the temperature being exceeded, will actuate the nozzle group valves of the expansion turbine to restrict the flow of fuel gas.

FURTHER DEVELOPMENT PROSPECTS

There are various promising lines of development for the combined gas/steam turbine cycle process. While retaining the present type of gas turbine, the unit rating may be increased by reducing the air ratio to 1:1 at the inlet to the gas turbine, and operating the steam turbine with reheated steam. The total net output would thus be increased to 330 MW, the gross output of the gas turbine then being 110 MW and that of the steam turbine 230 MW. At the same time, the net heat rate would drop from 2,400 kcal/kWh to 2,170 kcal/kWh. Reducing the air ratio would also improve the partial-load performance of the entire plant, as may be seen from curve (c) in Fig. 5, and the specific capital expenditure would drop from DM 390/kW net to DM 320/kW net.

Further possibilities of development are offered by selecting larger gas turbine units or, with the same size power plant, by raising the gas inlet temperature. Thus, the net heat rate would be 2,030 kcal/kWh on raising the gas inlet temperature from 820°C to 1,050°C, and with a net output of 330 MW the gas turbine output woud be 145 MW. Improving efficiency will reduce the specific capital expenditure for the plant as a whole to the same extent; the percentage of both coal-dependent and steam-dependent plant parts will decrease as efficiency improves, thus leading to a reduction in specific capital expenditure. Unlike a conventional steam power station, the cost per kW in a combined gas/steam turbine power plant will drop simultaneously with improvement in efficiency.

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APPENDIX 4

STEAM GENERATORS OF SPECIAL DESIGN

K. H. Schmahl, Dipl.-Ing.

3.1. COMBINED GAS/STEAM TURBINE PROCESSES

The linking of gas and steam turbines permits considerable savings in capital expenditure for fossil fuel power stations, and simultaneously improves thermal efficiency. This is particularly true where the conventional steam turbine process features not just a superposed gas turbine, the waste gases from a normal open-cycle process being used as combustion air to fire a conventional steam boiler, but where the gas turbine is integrated into the process in such a way that steam from the steam turbine is generated in the gas turbine combustion chamber; in other words, the combustion chamber is a "supercharged" steam generator, the furnace being operated under positive pressure. The gas turbine waste gases are then utilized to preheat the feedwater.

Figure 21 shows the simplified arrangement of a prototype combined gas/steam turbine cycle process featuring, in this case, a fuel gasifier for solid fuels. This is not required when using liquid or gaseous fuels, of course.

3.2. BITUMINOUS COAL GASIFICATION

The coal is gasified in pressure vessels (Figure 22) using air and steam as gasifying media. The fuel gas so produced is available as pressure gas, the entrained dust being removed in a scrubber. An expansion turbine/gasifying air compressor assembly, which has no effective output of its own, is used for adjusting the pressure of the gas to suit the various pressure stages at which the coal gasifier and gas turbine operate at optimum.

3.3. THE STEAM GENERATOR

The forerunner of this supercharged steam generator is the high-velocity boiler (Velox boiler), of which numerous have been built and which feature small surface in keeping with high flue-gas flow rates.

Figure 23 shows a cross-section through the gas turbine and supercharged boiler. Owing to the design of the gas turbine, the boiler had to be split into two sections; these are arranged on either side of the turbine and connected in parallel.

Each unit generates 170 t/h steam at 130 atms.g. and 535°C. As the furnace pressure is 10 atms., the pressure vessel has an outside diameter of only 3.6 m and an overall height of 18 m. This permits complete workshop assembly. Also, the burner dimensions are within present-day customary limits.

3.3.1 AIR AND FLUE-GAS FLOW PATTERN

The air-flue-gas path in the boiler is also governed by the gas turbine design. The

Translation of an abridged extract from "Jahrbuch der Dempferzeugungstechnik", 2nd Edition, Vulkan-Verlag Dr. W. Classen, Essen

turbine emits compressed air through a pipe arranged coaxially around the gas intake sockets, and so helps to cool the pressurized outer wall of the pipe.

The combustion air in the boiler is guided along the vessel wall to the burner in the same manner, while the flue gas is kept away from the vessel wall by welded tangent-tube walls, leaving the boiler coaxially to the incoming air.

Since steam generation and thus fuel input are limited in this particular case, the air not used for the combustion process is previously mixed with the flue gas via an air vale. This permits some degree of regulation of the excess air at the burner.

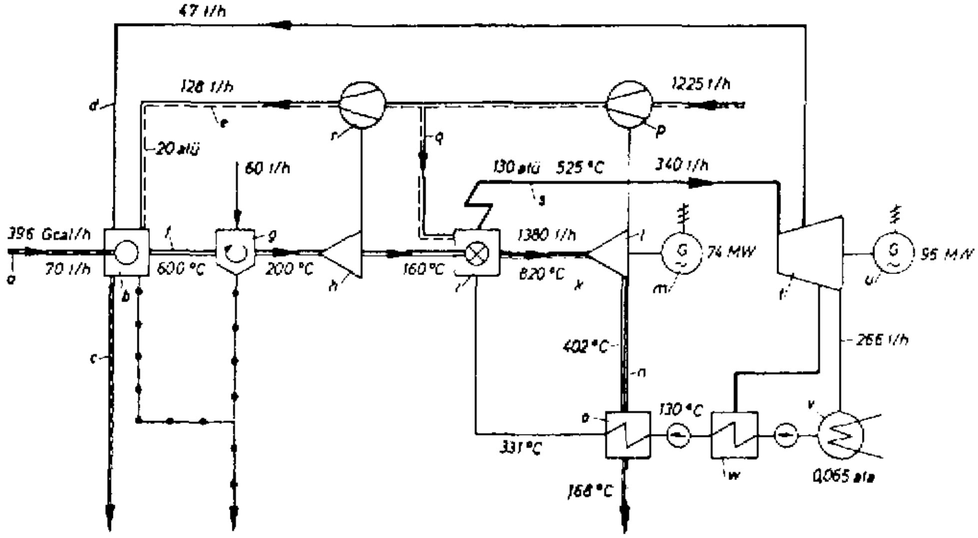


Fig. 21: Combined Gas/Steam Turbine Cycle Process (STEAG System)

a	coal	h	expansion turbine	p	air compressor
b	gasifier	i	combustion chamber	q	combustion air
c	ash	k	combustion gas	r	gasifying air compressor
d	steam	l	gas turbine	s	live steam
e	air	m	generator, gas turbine	t	steam turbine
f	fuel gas	n	waste gases	u	generator, steam turbine
g	scrubber cooler	o	preheater	v	condenser
				w	preheater

A further facility of reducing the flue gas temperature is provided by a flue-gas bypass. This is formed by the tubeless core of the tube bank, and permits an adjustable partial flow of flue gas to bypass the heating surface.

3.3.2 THE PRESSURE HEATING SURFACES

The steam generator system is in fact a Benson system. The demineralized feedwater is first intensively preheated in the economizer using waste gases from the gas turbine. The feedwater enters an annular manifold through three pipes at the bottom of the boiler and, while starting to evaporate, flows through the cylindrical tangent-tube wall; this comprises 48 parallel tubes, which were welded together and spirally wound. The wall separates the air from the flue gas, and terminates in an annular manifold at the top. The steam then flows through 96, down pipes, which subsequently form the supporting-tube bottom, and an inside, welded tube jacket — the flue-gas bypass — up this bypass and into the tube bank heating surface. This consists of concentric, spirally-wound but non-welded tube cylinders. The steam passes through a bottom flow-reversing annual manifold and, after again flowing up and down through the heating surface, leaves the boiler at the bottom via the outlet annular manifold and three connecting pipes.

The steam flows in parallel through both boiler sections; however, the sole means of regulating the temperature of the superheated steam is adjusting each feedwater flow rate.

The design of the superheater tube bank heating surface (of concentric, cylindrical spiral windings) and dimensioning of the combustion chamber are such that, in the event of a tube fracture, the cylinder containing the damaged tube can be disconnected and drawn into the combustion chamber for repair there.

The space underneath the superheater is accessible through a manhole in the vessel wall. The combustion chamber can be entered after the burner has been dismantled.

The heating surfaces are made of ferritic material only. Particular attention has been paid to appropriate dimensioning of tube and manifold wall thicknesses to cater for thermal stresses caused by the fluctuating operating conditions of the quick starter and also by the daily shutdowns. The permissible temperature gradient on starting, for example, averages 70°C/min.

3.3.3 THE GAS FIRING EQUIPMENT

Based on results of small-scale tests, multi-lance type circular burners were provided for the firing system. The 6 lances inject the gas into the air current through radial slots. These lances can be moved during operation and permit adjustment of the flame starting point and, to a certain extent, also the shape of the flame. This can be observed through peepholes in the vessel jacket and tube wall.

The burners are designed for a capacity of 190 Gcal/h each.

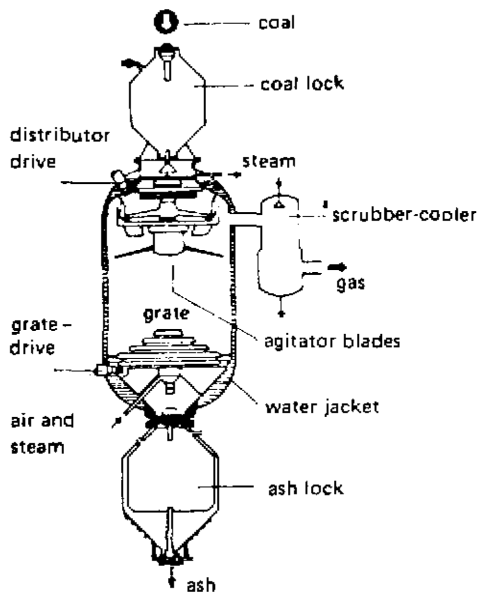


FIGURE 22
PRESSURE GASIFIER (LURGI DESIGN)

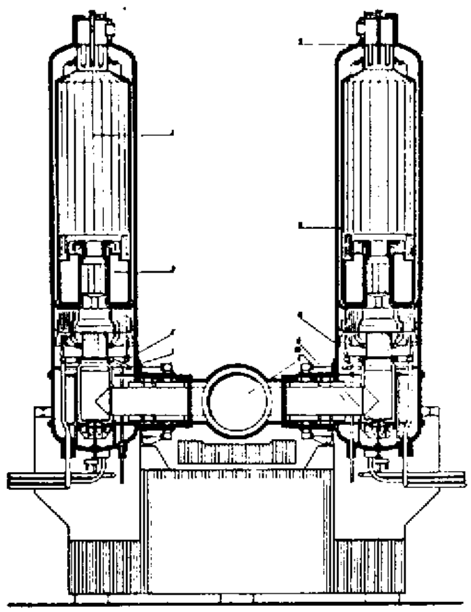


Fig. 23: Pressurized Steam Generator (Durr design)

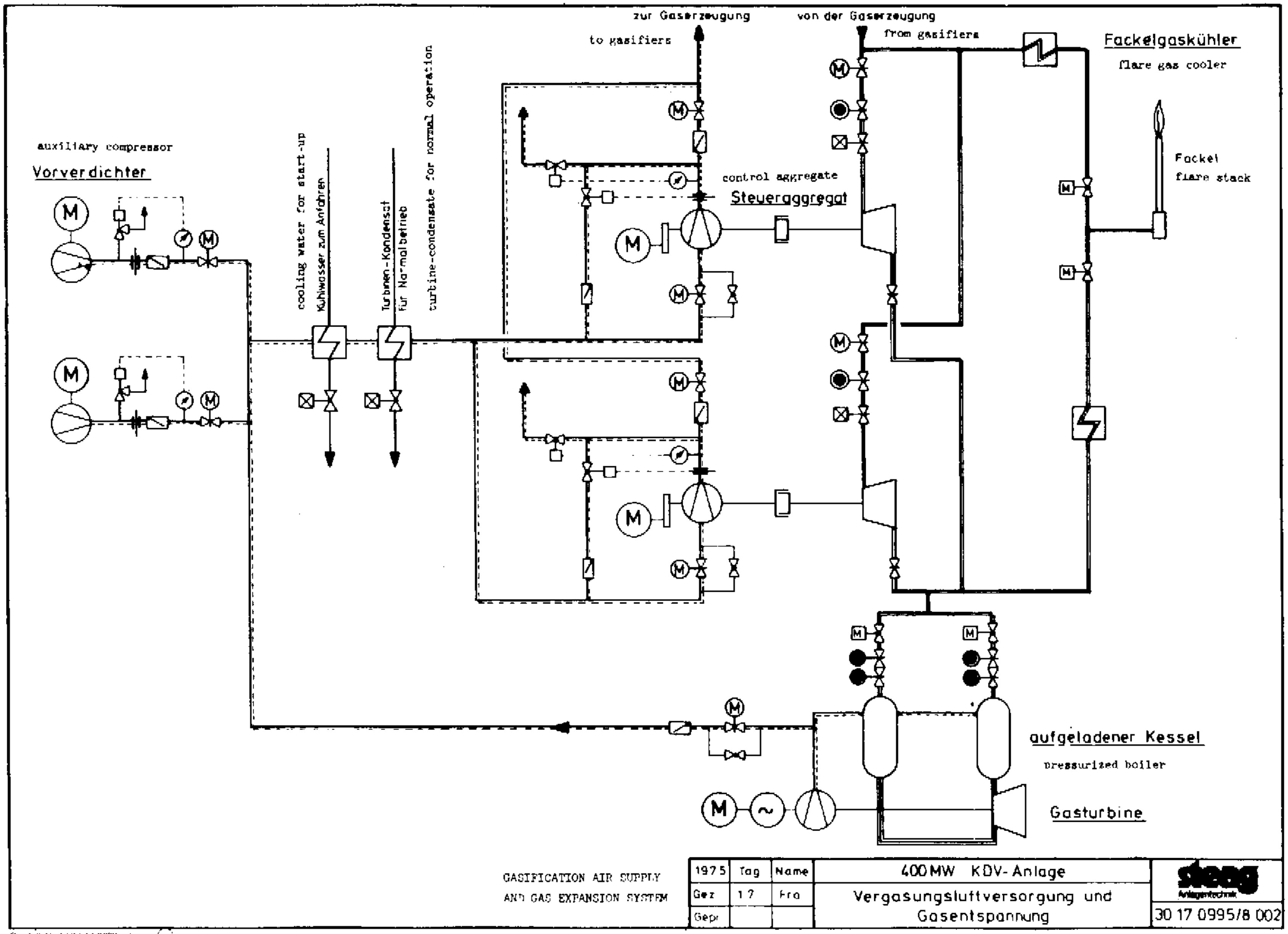
1	feedwater inlet	6	air jacket
2	evaporator	7	gas burner
3	superheater	8	air mix
4	superheat steam outlet	9	air control valve
5	gas turbine	10	gas outlet

3.4 PLANT OPERATION

The supercharged steam generator operates with conventional flue gas flow rates. The reduction in heating surface is effected only by the intake pressure, which — as a factor of the medium density — takes effect on the flue gas heat transfer coefficient with the same exponent as the velocity. At the same time, heat transfer through radiation is also greater as a result of increase in the partial pressure of the radiant gas constituents.

That is the reason for the small dimensions of the steam generator, which permit complete workshop assembly in pressure vessels that still permit normal handling and transportation.

With this prototype plant, it was decided not to utilize fully the steam generating capacity of about 800 t/h possible with the air volume supplied by the gas turbine. Appropriate studies have shown that this is fundamentally possible. To exploit the advantages of workshop assembly, a total of 6 pressure vessels are built for this output, 4 equipped with combustion chambers, 2 containing only contact heating surfaces. In this case, it is easily possible to include reheater heating surfaces, which have not as yet been fitted in the prototype.


zur Gaserzeugung
to gasifiers
von der Gaserzeugung
from gasifiers
Fackelgaskühler
flare gas cooler
Fackel
flare stack
auxiliary compressor
Vorverdichter
cooling water for start-up
Kühlwasser zum Anfahren
Turbinen-Kondensat
für Normalbetrieb
turbine-condensate for normal operation
control aggregate
Steueraggregat
M
aufgeladener Kessel
pressurized boiler
Gasturbine
GASIFICATION AIR SUPPLY
AND GAS EXPANSION SYSTEM
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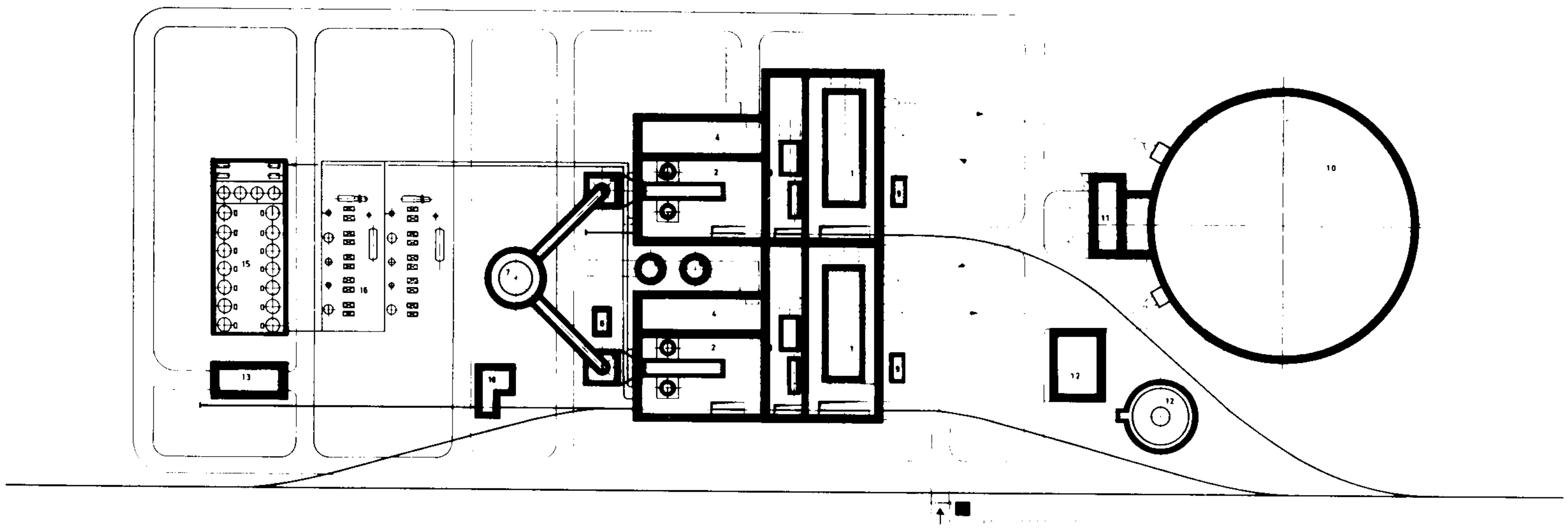

Legende:

STEAM TURBINE HALL	1	Maschinenhaus (Dampfturbine)
GAS TURBINE HALL	2	Maschinenhaus (Gasturbine)
PUMPHOUSE, HOT WATER STORAGE	3	Pumpenhaus, Warmspeicher
SWITCHGEAR, CONTROL ROOM	4	Schaltanlage, Warte, Verwaltung
WASTE HEAT RECOVERY UNIT	5	Abhitzekessel
START UP BOILER	6	Anfahrkessel
STACK	7	Schornstein
COLD WATER STORAGE	8	Kaltspeicher
UNIT TRANSFORMER	9	Maschinentrafo
NATURAL DRAFT COOLING TOWER	10	Naturzugkuhlturm
C.W. PUMPHOUSE	11	Hauptkuhlwasserpumpenhaus
C.W. MAKE UP SYSTEM	12	Kuhlturmzusatzwasseraufbereitungsanlage
ASH SILO	13	Aschebunker
DESULPHURIZATION PLANT	14	Entschwefelung
GASIFICATION PLANT	15	KDV - Anlage
WORKSHOP, STORAGE	16	Werkstatt, Lager
GATE HOUSE	17	Pfortner

LAYOUT

0 10 20 30 40 50 m

1975 25.6.

2 x 500 MW Kombi - Block
B.C. Hydro - Study

1 : 500

Lageplan

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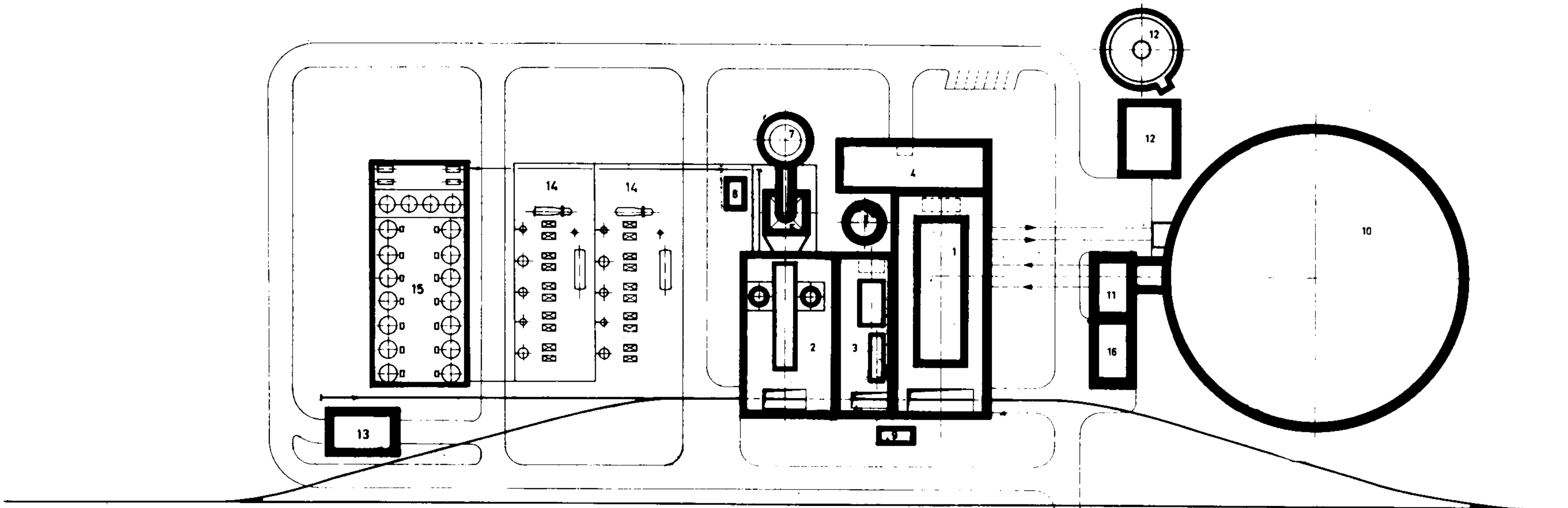

Legende:

STEAM TURBINE HALL	1	Maschinenhaus (Dampfturbine)
GAS TURBINE HALL	2	Maschinenhaus (Gasturbine)
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WORKSHOP, STORAGE	16	Werkstatt, Lager
GATE HOUSE	17	Pförtner

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LAYOUT

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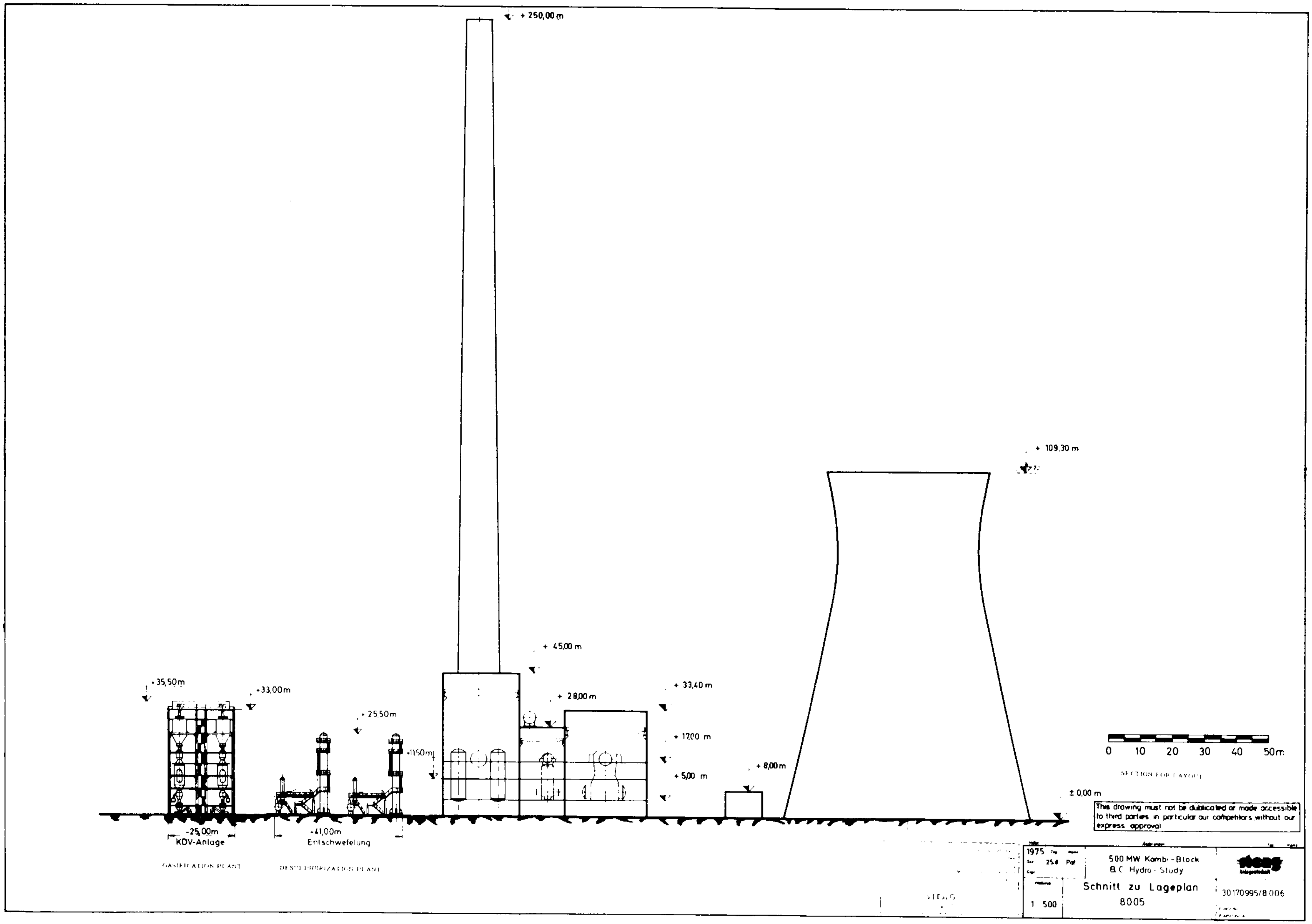
1:500	Lageplan	30170995/B.005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± 0,00 m
0 10 20 30 40 50m
SECTION FOR LAYOUT
-25,00m
KDV-Anlage
GASIFICATION PLANT
-41,00m
Entschwefelung
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500 MW Kombi-Block
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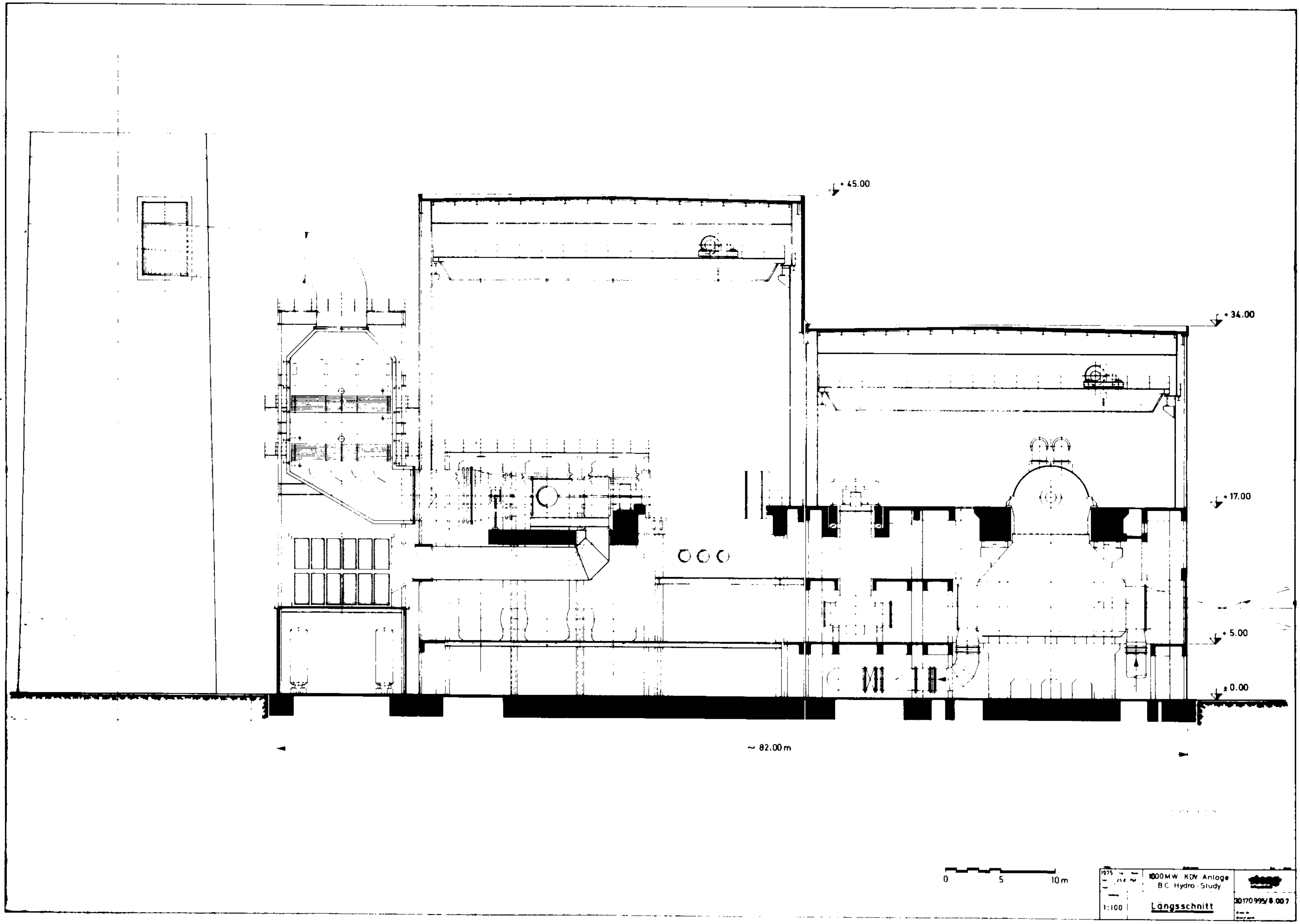
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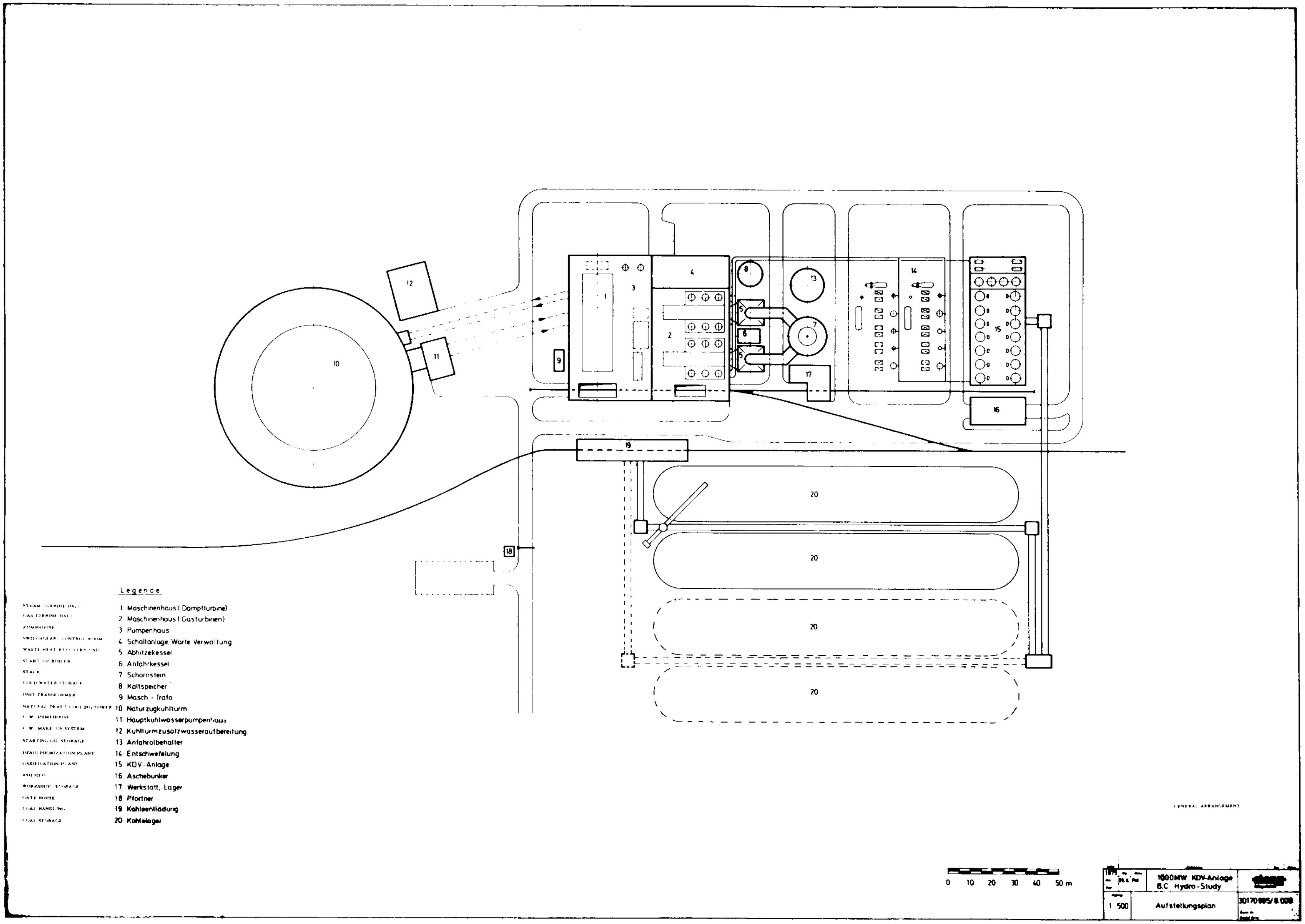
Legende
STEAM TURBINE HALL 1 Maschinenhaus (Dampfturbine)
GAS TURBINE HALL 2 Maschinenhaus (Gasturbinen)
PUMPHOUSE 3 Pumpenhaus
SWITCHGEAR, CONTROL ROOM 4 Schaltanlage, Warte Verwaltung
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STACK 7 Schornstein
COLD WATER STORAGE 8 Kaltspeicher
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STARTING OIL STORAGE 13 Anfahrolbehalter
DESULPHORIZATION PLANT 14 Entschwefelung
GASIFICATION PLANT 15 KDV-Anlage
ASH SILO 16 Aschebunker
WORKSHOP STORAGE 17 Werkstatt, Lager
GATE HOUSE 18 Pfortner
COAL HANDLING 19 Kohleentladung
COAL STORAGE 20 Kohlelager
GENERAL ARRANGEMENT
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1000MW KDV-Anlage
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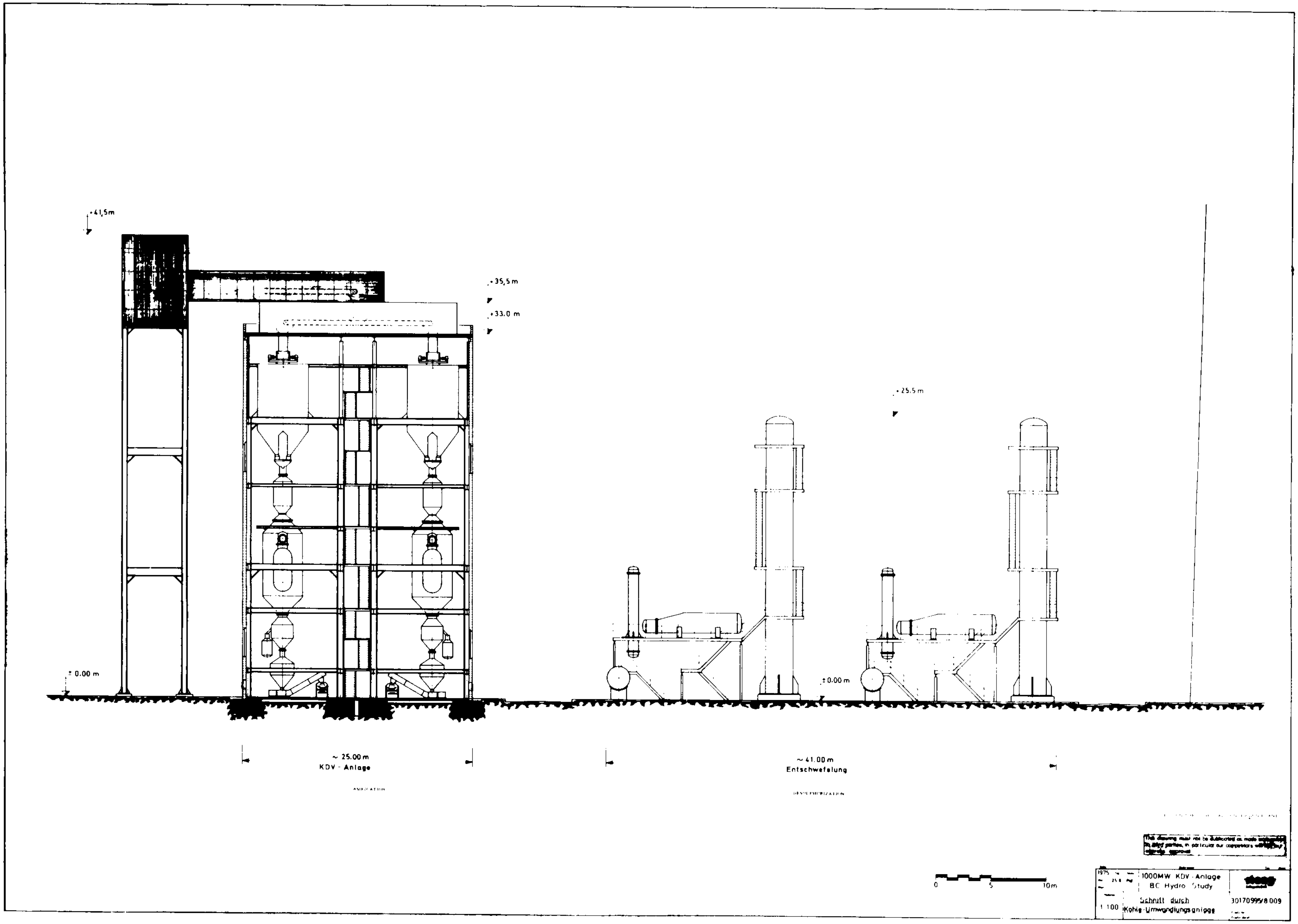
Aufstellungsplan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+35,5 m
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KDV - Anlage
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Schnitt durch
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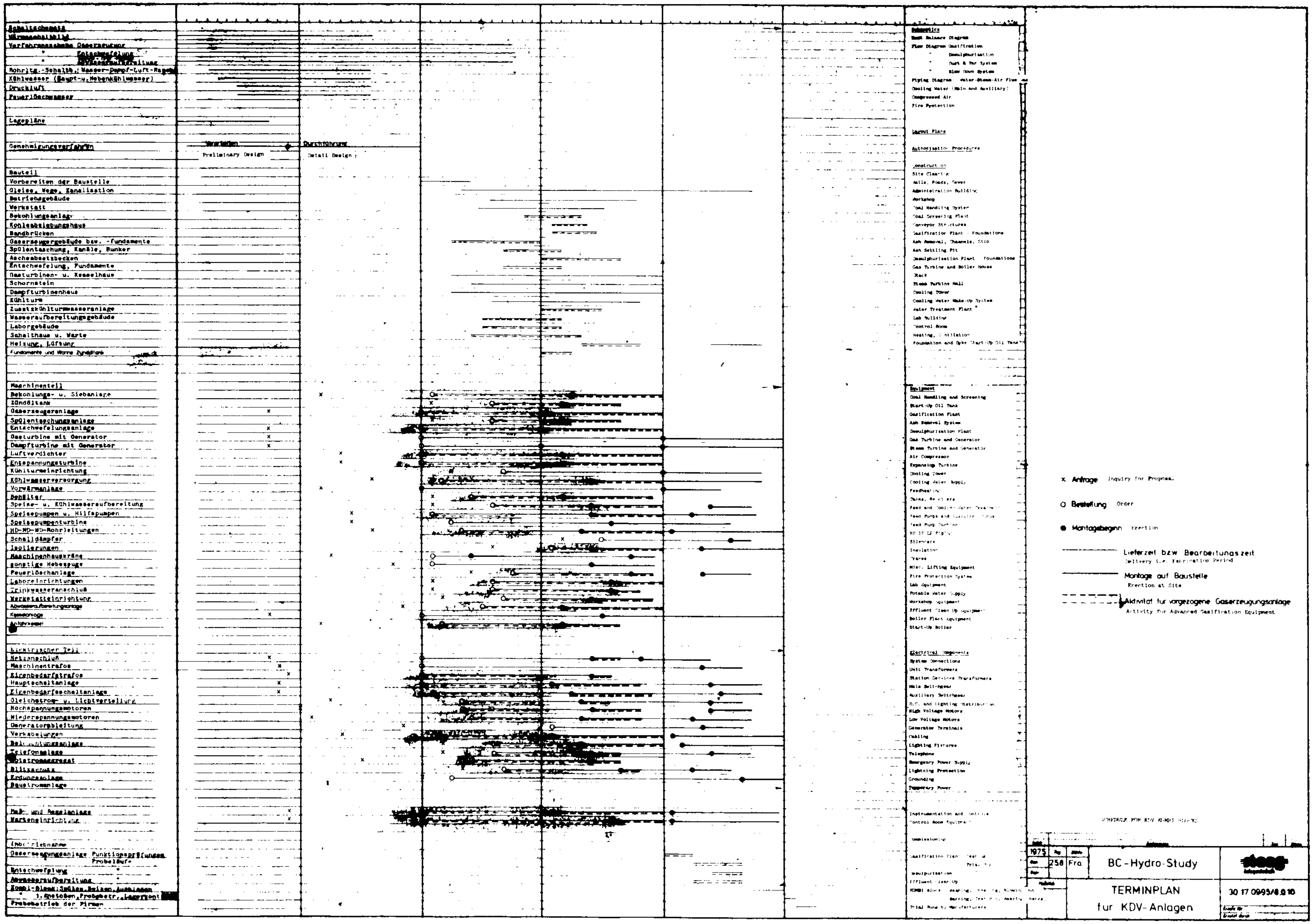

Schaltschemata
Wärmeschaltbild
Verfahrensschema Gaserzeugung
Rohrltg.-Schaltb., Wasser-Dampf-Luft-Rauchgas
Kühlwasser (Haupt-u. Nebenkühlwasser)
Druckluft
Feuerlöschwasser
Lagepläne
Genehmigungsverfahren
Vorplanen
Preliminary Design
Durchführung
Detail Design
Bauteil
Vorbereiten der Baustelle
Gleise, Wege, Kanalisation
Betriebsgebäude
Werkstatt
Bekohlungsanlage
Kohlesiebbrechhaus
Bandbrücken
Gaserzeugergebäude bzw. -fundamente
Spülentaschung, Kanäle, Bunker
Ascheabsetzbecken
Entschwefelung, Fundamente
Gasturbinen- u. Kesselhaus
Schornstein
Dampfturbinenhaus
Kühlturm
Zusatzkühlturmwasseranlage
Wasseraufbereitungsgebäude
Laborgebäude
Schalthaus u. Warte
Heizung, Lüftung
Maschinenteil
Bekohlungs- u. Siebanlage
Zündöltank
Gaserzeugeranlage
Spülentaschungsanlage
Entschwefelungsanlage
Gasturbine mit Generator
Dampfturbine mit Generator
Luftverdichter
Entspannungsturbine
Kühlturmeinrichtung
Kühlwasserversorgung
Vorwärmanlage
Speise- u. Kühlwasseraufbereitung
Speisepumpen u. Hilfspumpen
Speisepumpenturbine
HD-MD-ND-Rohrleitungen
Schalldämpfer
Isolierungen
Maschinenhauskräne
sonstige Hebezeuge
Feuerlöschanlage
Laboreinrichtungen
Trinkwasseranschluß
Werkstatteinrichtung
Elektrischer Teil
Netzanschluß
Maschinentrafos
Eigenbedarfstrafos
Hauptschaltanlage
Eigenbedarfsschaltanlage
Gleichstrom- u. Lichtverteilung
Hochspannungsmotoren
Niederspannungsmotoren
Generatorableitung
Verkabelungen
Beleuchtungsanlage
Fernsprechanlage
Notstromaggregat
Blitzschutz
Erdungsanlage
Baustromanlage
Meß- und Regelanlage
Warteneinrichtung
Inbetriebnahme
Entschwefelung
Probebetrieb der Firmen
Schematics
Heat Balance Diagram
Flow Diagram Gasification
Piping Diagram
Cooling Water (Main and Auxiliary)
Compressed Air
Fire Protection
Layout Plans
Authorisation Procedures
Construction
Equipment
Electrical Components
x Anfrage Inquiry for Proposal
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Lieferzeit bzw. Bearbeitungszeit
Delivery i.e. Fabrication Period
Montage auf Baustelle
Erection at Site
Aktivität für vorgezogene Gaserzeugungsanlage
Activity for Advanced Gasification Equipment
1975
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TERMINPLAN
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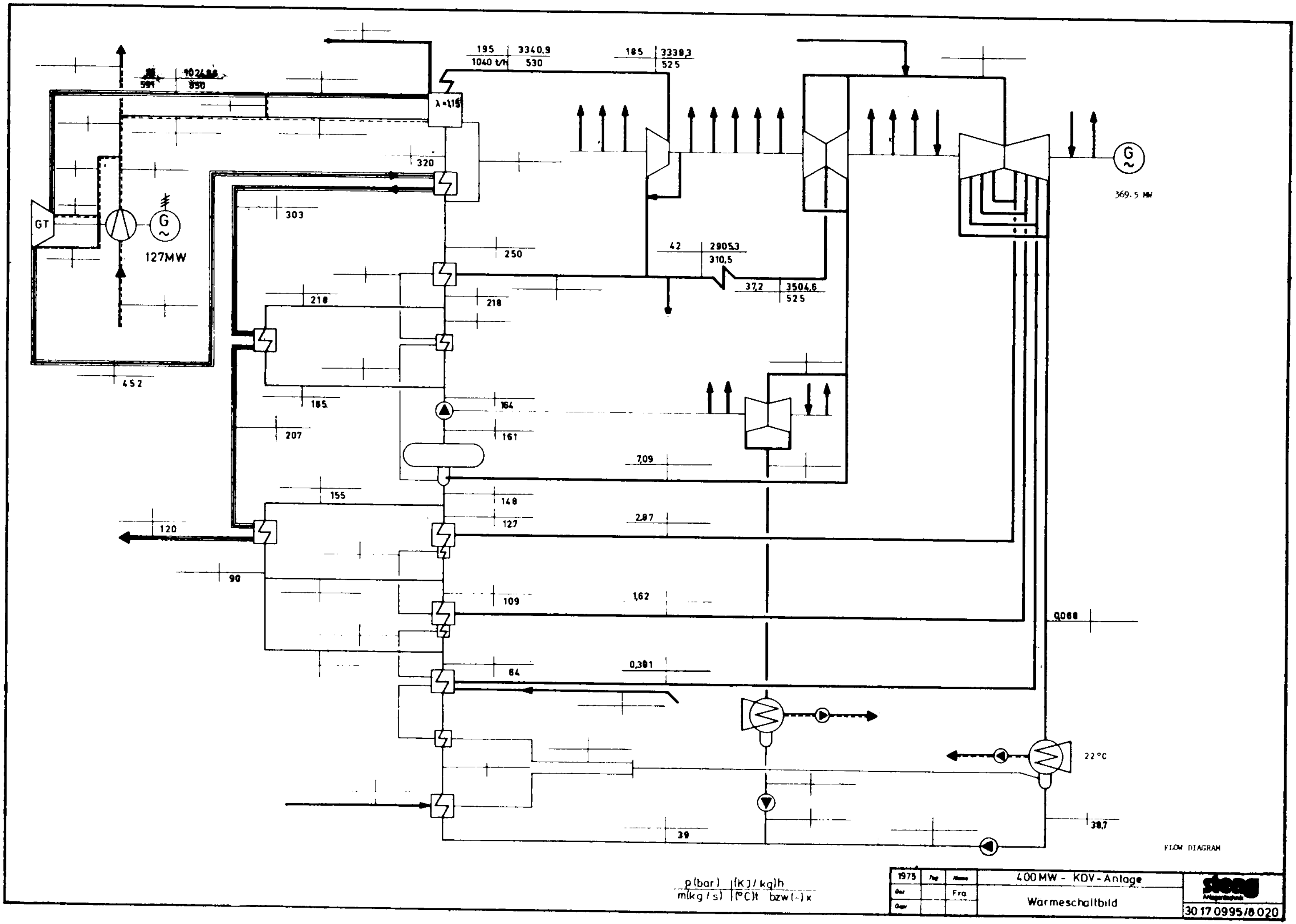

GT
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127MW
λ = 1,15
1024,86
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195 | 3340,9
1040 t/h | 530
185 | 3338,3
525
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42 | 2905,3
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FLOW DIAGRAM
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400 MW - KDV-Anlage
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STUDY C — SNG, MEDIUM AND LOW Btu GASIFICATION

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SECTION **CONTENTS**

1.0 INTRODUCTION

This report covers the technical and economic analysis of coal gasification processes to manufacture low to medium Btu fuel gas for electric power generation, medium Btu gas for distribution as town gas, high Btu gas as pipeline-quality gas, and the liquefaction of coal for the production of fuel oil.

The study was conducted at the request of the B.C. Hydro and Power Authority as part of an overall program to identify and analyze various schemes for the generation of electric power.

2.0 SUMMARY

The technical and economic components in the production of synthetic gas by coal gasification were developed for various gas products. The LURGI and KOPPERS-TOTZEK coal gasification processes were chosen to analyze the manufacture of low to medium Btu fuel gas for power generation stations of 2000 MW and 900 MW capacity. The LURGI process was selected to study the difference in operating and investment requirements between oxygen and air-blown gasification.

The technical definition and costs of coal gasification plants based on LURGI technology for the generation of 250 MM SCFD of town gas for Vancouver Island and 250 MM SCFD of pipeline-quality gas (SNG) were prepared.

Following the criteria established by the coordinating consultant and B.C. Hydro, the capital investment requirements, with the corresponding gas costs for the power generating stations, are summarized as follows:

Plant Capacity	2000 MW 450 x 10^9 Btu/D		900 MW 230 x 10^9 Btu/D		
Process	LURGI	K-T	LURGI	LURGI	K-T
Gasifying Agent	O_2	O_2	O_2	Air	O_2
Cost of Facilities, $MM	808	1007	433	470	520
Cost of Service, $/MM Btu	1.11	1.50	1.17	1.18	1.50
Btu/Scf	300	292	300	192	292

Pertinent data for the raw materials, utilities, by-products, and capital charges used in the economic analysis are as follows:

Coal	\$3/ST (as received)
Steam (Purchased)	\$1.00/1000 lb
Power (Purchased)	10 mills/Kwh
Liquid Hydrocarbon By-products	\$6/Bbl
Ammonia by-product	\$180/ST
Annual Charges	15.1625 percent of Cost of Facilities
Variable Maintenance	0.3 mills /Kwh

The annual charges include plant operation and maintenance, depreciation, interest expense, other taxes, insurance, and interim replacement.

The study guidelines established by B.C. Hydro and the coordinating consultant limited Lummus to minimal contact with the licensors of the processes and the use of data that are essentially in the public domain. The analysis of the LURGI processes was done on the basis of the document submitted to the Federal Power Commission by American Natural Gas. In order to use the FPC filing document, we assumed that Hat Creek coal would gasify similarly to North Dakota lignite, an assumption that has to be verified by LURGI. The plant area costs listed in that document were adjusted for capacity and escalated to mid 1975. The KOPPERS-TOTZEK process was analyzed on the basis of communications between Lummus and KOPPERS-TOTZEK, covering a heat and material balance for North Dakota lignite and an order of magnitude estimate of the cost of the KOPPERS-TOTZEK sections of the plant.

Caution should be exercised in using the data submitted in this report. If the results of this study lead to a phase where a rigorous analysis of technical and economic requirements are needed, we suggest that the services of the licensors be employed.

The results clearly show that the LURGI process produces synthetic gas at a lower cost than the KOPPERS-TOTZEK process. The major reasons for these results are in the differences in capital investment and thermal efficiencies for these processes. An examination of the cost of facilities and cost of service readily reveal the sensitivity of the gas cost to capital investment and the price of coal. A graph showing the effect of varying coal prices and different capital investment has been included to illustrate this point. (See Figure No. 1.)

The cost data indicates that there is little or no economy of scale between a plant capacity of 230×10^9 Btu/D and a plant capacity of 450×10^9 Btu/D.

The comparison between oxygen and air-blown LURGI coal gasification systems shows relatively little difference in operating or investment costs. It is noted, however, that the air-blown system yields a gas with a heating value (HHV) of 192 Btu/SCF, compared to a gas from an oxygen-blown system with a heating value (HHV) of 300 Btu/SCF. This difference in heating values may have significant effects in the design of boilers that would use this gas and will have to be considered if the manufactured gas is to be transported via pipeline over an extended distance.

The manufacturing cost of the town gas was calculated at \$1.45/MM Btu before enrichment with LPG. The cost of pipeline-quality gas was estimated at \$1.81/MM Btu. Comments similar to the cost of gas for thermal power generating stations apply to town gas and pipeline gas. The cost of facilities, along with the associated capital charge rate and the cost of coal to be used in the calculation, are the significant variables. It should be noted that B.C. Hydro assigned the coal at \$3/T and the capital charge rate at 11.75 percent.

The comparison between the LURGI and KOPPERS-TOTZEK processes in the production of low Btu gas leads to the conclusion that the LURGI process results in lower production costs in the manufacture of town gas or pipeline-quality gas, since the upgrading of the gas obtained from the KOPPERS-TOTZEK gasifier will require substantially greater facilities that those required by the LURGI process.

The use of British Columbia coal in Lummus' "CLEAN FUEL FROM COAL" liquefaction process has been evaluated. Cost of service for this process has been estimated as $1.78/MM Btu of liquid product.

A preliminary review of the COGAS process shows that for lignite-type coal, this process has a lower Thermal Efficiency than the LURGI process. The capital investment for a COGAS plant of 230 MM Btu/D of medium Btu gas is essentially the same as the LURGI plant. If COGAS is evaluated using a bituminous coal, the results show that COGAS is competitive with LURGI in both technical and economic areas. The reason is that the liquid by-product yield from a bituminous coal (Illionis No. 6) is about 4-5 times greater than the liquid yield from a lignite (Glen Harold, North Dakota) coal.

A preliminary review of the SYNTHANE process indicates that with a lignite-type coal, the process has a higher Thermal Efficiency than LURGI, and the cost of service is competitive with LURGI.

Of the second-generation processes examined in this report, only SYNTHANE appears to have advantages warranting further study using lignite coals as a feedstock.

3.0 DISCUSSION OF RESULTS

3.1 FUEL GAS

For the production of low Btu or medium Btu fuel gas, three basic systems were analyzed — namely, LURGI with oxygen, LURGI with air, and KOPPERS-TOTZEK with oxygen. The total cost of facilities determined for each of the systems shows that the LURGI oxygen-blown case is the least expensive one, followed closely by the LURGI air-blown case with sulfur removal via Stretford. A variation of the LURGI air-blown case has been included to show the potential advantage of using a different gas treating system — hot potassium carbonate — instead of Stretford. This system can possibly reduce or eliminate the Phenosolvan unit and the handling of substantial quantities of tars, oils, phenols, and ammonia. It should be attractive from an environmental point of view, as it may reduce or eliminate the disposal of contaminated water streams.

The cost of the KOPPERS-TOTZEK plant is significantly higher for the 900 MW and 2000 MW alternates. The difference in capital investment results from the much higher oxygen requirements for the KOPPERS-TOTZEK process than the LURGI process.

Unlike the LURGI process, the KOPPERS-TOTZEK process does not produce liquid by-products but converts all of the coal to gaseous products such as hydrogen, carbon monoxide, and carbon dioxide. Because LURGI operates at a lower temperature and uses less oxygen, not all the heavy hydrocarbons produced by devolatilization are broken down to hydrogen, carbon monoxide, and carbon dioxide. Significant quantities of by-products such as tars, oils, phenols, and ammonia are formed. An estimated value for these by-products has been included in the cost of service analysis.

There is also a significant quantity of methane and other higher hydrocarbons in the gas stream of the LURGI process. This results in a gross heating value of the product gas greater than that obtained from the KOPPERS-TOTZEK process. For the same quantity of energy then, the volumetric flow of gas from a LURGI unit is smaller than the flow from a KOPPERS-TOTZEK unit.

Another difference between the two processes is that the LURGI process generates the gas at a pressure between 300 and 400 psig, while the KOPPERS-TOTZEK process generates gas at about 1 or 2 psig. This difference in operating pressure results in further capital increases of the KOPPERS-TOTZEK system, in that product compression is needed to supply the gas to the desired system pressure of 30 to 50 psig. In the LURGI case, the available higher pressure of the gas makes it convenient to include a product gas expansion system to supplement the power requirements of the plant. If the product gas is to be transported some distance from the coal gasification plant, the LURGI gas would not be expanded; rather, the available system pressure would be used to transport the gas to its final destination. For example, if the Burrard power station were to be converted to use low or medium Btu gas manufactured at Hat Creek, arrangements for a pipeline would be included. In such a case, the LURGI system would further increase its advantage over the KOPPERS-TOTZEK process.

An article presented by LURGI at the Clean Fuel From Coal symposium of IGT in June, 1975 discussed the feasibility of recycling tars, oils, and other by-products generated in the LURGI system within the gas processing steps instead of separating these compounds. This alternative was used by Lummus to calculate the approximate investment and operating costs for an air-blown LURGI gasification unit. The significant advantage of this process lies in reducing the environmental problems associated with coal gasification, as it promises to significantly reduce the quantity of contaminated water. It also reduces the amount of handling or processing of the by-products, which is an advantage in a remote location. The comparison between the two air-blown cases of the LURGI system results in capital savings of about 12 percent and an improvement in Thermal Efficiency of the process. The final costs of service are in favor of the case where by-products are separated; but the difference is very slight and depends on the sales value of the by-products. This case merits exploration in much greater detail than this study permitted.

The results of the study show that small differences exist in using either air or oxygen in the production of low Btu gas. It should be noted that the elimination of an oxygen plant reduces the potential safety hazard of the plant.

At the present time, the KOPPERS-TOTZEK (with oxygen) and LURGI coal gasification (with oxygen or air) processes are the only ones that have been used in commercial-scale plants and, as such, are the only processes that qualify as proven, reliable technologies.

The sulfur removal process used in the evaluations of the low/medium Btu fuel gas processes was Stretford. The Stretford process has been used at atmospheric pressure, but not yet at elevated pressure. Pilot plant tests done by the British Gas Board on Stretford have indicated that this process is suitable for pressure operation. If a stipulation for the construction of a plant at this time is that all systems be based on fully demonstrated, large-scale installations, an alternative to the use of the Stretford process under pressure is a hot potassium carbonate system, similar to the one employed at the Westfield plant of the Scottish Gas Board. Sulfur conversion would be in a Claus unit or atmospheric Stretford process, depending on the sulfur emission regulations.

The definition of coal gasification units of LURGI and KOPPERS-TOTZEK is based on information used in similar studies. In any forward plan of a coal gasification project, either of the licensors has to be brought in for detailed analysis of the coal and the design of the respective proprietary equipment.

3.2 TOWN GAS

The manufacture of this gas is based on the LURGI coal gasification system and has been patterned to some extent on the Westfield plant of the Scottish Gas Board — a plant that was in operation between 1960 and 1973. The calculated gas cost of \$1.45/MM Btu is higher than the comparable gas cost of the oxygen-blown LURGI system for the production of fuel gas. This extra cost represents the addition of a shift conversion unit to reduce the carbon monoxide concentration in the product gas. The shift reaction decreases the heating value of the product gas, and a larger quantity of gas for the same energy output has to be generated. The study did not include enrichment steps with LPG and did not evaluate the possible advantages that might be obtained if carbon dioxide were removed from the gas stream, as was done in the Westfield plant. The relative advantages of such a step could be the subject of a further evaluation.

3.3 PIPELINE GAS

The production of pipeline gas is based on the oxygen-blown LURGI coal gasification process. The total cost is analogous to that developed for the American Natural Gas project in North Dakota, if one takes into account the differences in coal cost and capital charge rate.

4.0 CONCLUSIONS AND RECOMMENDATIONS

4.1 LOW AND MEDIUM BTU FUEL GAS

The major objective of this study was to determine the most attractive process for the generation of low to medium Btu fuel gas through the gasification of coal. Commercially demonstrated processes were given prime consideration. Only the LURGI (oxygen and air-blown) and KOPPERS-TOTZEK (oxygen-blown) processes come under this category today.

The lowest cost of service for the generation of fuel gas from coal is obtained from the oxygen-blown LURGI coal gasification process. The LURGI oxygen-blown process produces fuel gas at a cost about 20 to 25 percent lower than fuel gas produced by the KOPPERS-TOTZEK process.

For the 900 MW case, an oxygen-blown LURGI process was compared with an air-blown LURGI process. On the basis of using the same gas treating process, the difference in investment is in favor of the oxygen-blown case by about 10 percent, resulting in a difference of about 1¢/MM Btu in the gas cost. This difference is so small that both systems have to be rated equal. This gas obtained from the oxygen-blown case has a considerably higher heating value (300 versus 192 (Btu/SCF), which should result in a less costly boiler design or lower retrofitting costs if an existing gas-fired unit is to be converted from natural gas to low Btu gas. Should pipelining of the fuel gas over a significant distance be required, it would undoubtedly be more economical to generate medium Btu gas using oxygen-blown LURGI instead of low Btu gas via air-blown LURGI.

Further work should be done on the alternate air-blown LURGI process. In this system, tars, oils, and other compounds removed from the gas cooling step are recirculated or processed into the hot gas stream and the LURGI gasifier; and a hot potassium carbonate solution is used for removal of sulfur and some carbon dioxide. This particular case shows the lowest capital investment of the systems, but it is the system which has been studied and analyzed the least. To evaluate this case more accurately requires significant input from LURGI, the licensor of this process.

If the 900 MW power generating station at Burrard is to be converted from natural gas to low Btu gas, the most attractive process appears to be a LURGI oxygen-blown coal gasification plant constructed near the coal mines at Hat Creek. In this case, gas would be made available at the battery limits at about 300-350 psig; and depending on the details of the transmission system, additional product gas compression would, of course, be eliminated. If this case is under serious consideration, Lummus recommends a further detailed study to be done in conjunction with LURGI.

4.2 TOWN GAS

The LURGI coal gasification process has been used to define the investment and operating requirements of a gas plant that will generate 250 MM SCFD of a town gas with a heating value of 280 Btu/SCF. Since town gas, is normally distributed with a heating value in the range of 450-550 Btu-SCF, this gas has to be upgraded by enrichment with either propane, butane, or a mixture of both (LPG). The heating value of the gas leaving the process unit can be increased by removing carbon dioxide to varying degrees in order to reduce the amount of enrichment that may be needed. In view of the results obtained for the low Btu case, it is unlikely that a process other than LURGI would be considered for this type of gas at this time.

4.3 PIPELINE GAS

The production of high Btu pipeline-quality gas was developed using the LURGI coal gasification process. The definition of this type of plant parallels, in many details, the American Natural Gas project planned for North Dakota. Capital investment was corrected for local conditions insofar as is feasible at this time. Adjustments were made for the fact that high-pressure steam and boiler feed water are supplied to this plant instead of the production of these utilities within the coal gasification plant battery limits. The cost of gas is considerably lower than that used by American Natural Gas in its filing document; but as has been stated previously, the cost of coal and the capital charge rate have a significant influence on the cost of service. Should B.C. Hydro and Power Authority advance to another phase in the consideration of a pipeline gas plant, it is recommended that an analysis of such a project be undertaken in much greater detail, together with the process licensor — LURGI — as well as an extensive examination of the selected site conditions, labor conditions, and specific environmental regulations.

4.4 SECOND-GENERATION GASIFICATION PROCESSES

Three coal gasification processes presently in various stages of pilot plant development were analyzed to indicate if these hold any promise for materially affecting the capital or operating requirements in the production of low or medium Btu gas. The processes are those of TEXACO, COGAS, and the U.S. Bureau of Mines SYNTHANE.

The TEXACO coal gasification process was analyzed on the basis of in-house data; but following a meeting with the Texaco Development Corporation, Lummus was advised to retain the data until further information from TEXACO was obtained.

Based on information from the COGAS Development Company, it was determined that this process has a similar thermal efficiency and investment requirements to the LURGI process. The COGAS process appears to be a more interesting process if a higher-ranking coal is available than the British Columbia coal that was used in this evaluation.

The SYNTHANE process of the U.S. Bureau of Mines appears to have a higher thermal efficiency than the LURGI process, with approximately similar investment requirements. The values assigned to the by-products result in the cost of service of gas to be in favor of the LURGI process.

The review of second-generation technology was done within a limited time scale. The capital investment requirements for these processes have been evaluated in terms of an order of magnitude, while the investment data for the LURGI technology is based on extensive work done for other coal gasification projects. With this qualification, the conclusion at this time is that no significant advantages over the LURGI process can be expected from either COGAS or SYNTHANE. After the Bruceton pilot plant of the SYNTHANE process goes into operation, it may be interesting to review that process to determine if more significant data can be applied in this evaluation. Following receipt of data from the Texaco Development Corporation, a brief summary of that process will be submitted to indicate its potential in the generation of low Btu gas.

5.0 DESCRIPTION OF TECHNOLOGY

5.1 LOW BTU FUEL GAS

5.1.1 LURGI 2000 MW and 900 MW

5.1.1a OVERALL PLANT DESCRIPTION

This section of the study covers a grass roots coal gasification plant to be located at Hat Creek, British Columbia. The plant includes all process and utility systems, environmental facilities, tankage, and buildings, excluding auxiliary steam and electric power generation. It is sized to produce a sufficient volume of low Btu fuel gas to support an over-the-fence Thermal Generating Station having a capacity of 2000 MW and, as an alternate, a 900 MW station. The design is based on the well-known and commercially proven LURGI pressure gasification process. Non-proprietary process units, off sites, sulfur recovery and waste effluent control systems were designed by Lummus.

The Process Block Flow Diagram and Material Balance, Sketch No. 1, shows the major processing areas. Sized coal is delivered to the LURGI oxygen-blown gasifiers. The crude product gas is cooled and treated for sulfur removal before delivery to the Thermal Generating Station. Liquids condensed in the cooling process are processed through the gas liquor separation and treating units to obtain the following by-products; tar, tar oil, crude phenols, and anhydrous ammonia. Ash removed from the gasifier is processed and eventually returned to the mine site. The approximate mass balance shown is based on prorating the data disclosed by the ANG Coal Gasification Company in their filing document to the Federal Power Commission of the U.S. Government, and making adjustments dictated by the differences in the coal analysis and the flowsheet.

Approximate energy balances were calculated for the gasification section and the overall plant as follows:

GASIFICATION ENERGY BALANCE

		2000 MW 10^9 Btu/D	900 MW 10^9 Btu/D	HHV %
INPUT				
Coal to Gasifiers (at 20% moisture, 25% Ash)		522.0	266.7	100.0
OUTPUT				
Low Btu Fuel Gas	(300 Btu/lb at 60°F)	450.0	230.0	86.3
Tar	(147,000 Btu/US gal)	14.8	7.6	2.8
Tar Oil	(133,000 Btu/US gal)	23.6	12.1	4.5
Crude Phenols	(122,000 Btu/US gal)	4.9	2.5	0.9
Anhydrous Ammonia	(9800 Btu/lb)	6.2	3.2	1.2
		499.5	255.4	95.7

OVERALL PLANT ENERGY BALANCE

	2000 MW 10^9 Btu/D	900 MW 10^9 Btu/D	HHV %
INPUT			
Net coal to gasifiers	522.0	266.7	81.1
Electric Power	3.1	1.6	0.5
Steam Import	118.7	60.5	18.4
	643.8	328.8	100.0
OUTPUT			
Low Btu Fuel Gas (300 Btu/SCF at 60°F)	450.0	230.0	69.9
Byproducts	49.5	25.4	7.7
Steam Export	3.4	1.7	0.5
	502.9	257.1	78.1

The Overall Plant Energy Balance is based upon coal fed to the gasifiers. The total coal fed to the plant will be about 20 to 25 percent higher because the fines produced in the coal preparation area cannot be fed to the LURGI gasifiers and must be separated. Since B.C. Hydro coal is a lignite type coal, we believe that the quantity of fines produced will be similar to the quantity of fines produced by North Dakota lignite (22%). This, of course, must be confirmed by actual tests. The final disposition of the fines should be considered in any future evaluation.

A preliminary Plot Plan of the operating units is shown on Sketch No. 9 for the 2000 MW equivalent plant. The 900 MW case requires less area but it is recommended that a 100 acre area be purchased in all cases.

The complex is designed with maximum reliability built into the system to minimize the possibility of interruption in gas production. Independent, parallel processing systems are employed to minimize the effect of the loss of a process unit. In addition, spare equipment is provided at critical points. The process train philosophy is illustrated by the Process Train Arrangement Diagram shown on Sketch No. 8.

The environmental control systems are conceived to be integrated with the process units, waste heat recovery, cooling water and ash handling systems and are expected to meet applicable standards and regulations in British Columbia. Where possible, water reuse was incorporated into the design. A description and flow diagram of the liquid waste effluent recovery system is included in the respective sections of the study.

5.1.1b PLANT DEFINITION

The preliminary process design is based on processing lignite type coal in an oxygen-blown LURGI pressure gasification system to produce a sufficient quantity of low Btu fuel gas to feed a 2000 MW Thermal Generating Station and, as an alternate case, a 900 MW station.

The Hat Creek Coal was assumed to have the average properties shown on Table No. 1.

The study was prepared on the assumption that Hat Creek Coal will gasify similarly to the North Dakota Lignite used by ANG Coal Gasification Company in the filing document to the Federal Power Commission of the U.S. Government. This assumption has to be checked by LURGI.

The plant elements forming the basis of this study and estimate are listed below:

AREA	DESCRIPTION
1100	Gasification
1300	Gas Cooling
1600	Phenosolvan
1800	Gas Liquor Separation
2000	Coal Handling and Preparation
3000	Oxygen Plant
4000	Sulfur Recovery
5000	Steam Distribution
5300	Power Distribution
5400	Raw Water Supply and Treating
5500	Cooling Water
5600	Fire Protection System
5700	Miscellaneous Utilities
6000	Offsite Storage and Loading Facilities
7000	Plant Interconnecting Piping
8100	Liquid Waste Effluent System
8200	Ash Disposal
8300	Flare System
8400	Product Gas Expansion

Areas 1100, 1300, 1600 and 1800 are based on published process information on the LURGI technology. The preliminary design of all other areas was done by Lummus. A brief description of each area follows:

AREA 1100 - GASIFICATION

The coal is gasified using a multiple train of LURGI pressure gasifiers. High pressure oxygen and superheated steam are passed in countercurrent flow through a moving bed of coal, resulting in nearly complete conversion of the coal to gaseous compounds some of which are subsequently condensed and processed in the liquid state.

Sized coal is conveyed from battery limits to coal bunkers located above the gasifiers. Coal is charged to gasifiers through automatically operated coal locks which are depressurized before receiving coal from the bunkers. The lock gas from depressurizing is recompressed and reinjected in the gas cooling unit.

After filling with coal, the coal lock is pressurized with crude gas from the gas cooling unit and is equalized with the pressure of the gasifier. The coal is then charged to the top of the coal bed in the gasifier.

The gasification agent consisting of a mixture of oxygen and superheated steam is introduced through a rotating grate below the ash bed at the bottom of the gasifier. Partial combustion of the coal with the oxygen supplies the heat necessary for the gasification reactions.

Besides the crude gas, the process yields tar, oil, naphtha, phenols, ammonia and sulfur.

The ash is removed by the rotating grate and discharged through a semi-automatically operated ash lock into an ash chute. From the ash chute the ash is quenched with water and transferred to ash disposal.

Steam is generated in the jacket of the gasifier from the combustion heat and acts to cool the inner wall of the gasifier. The steam passes through a knockout drum and is returned directly to the gasifier, partially furnishing the steam required for gasification.

The hot crude gas leaving the gasifier is directly quenched in a wash cooler with recycled tarry gas liquor. Tarry gas liquor produced in excess of the recycle is sent to the gas liquor separation unit.

The crude gas freed from dust and heavy tars and saturated with steam is further cooled by generation of 100 psig steam in the waste head exchangers. The crude gas then passes directly to the gas cooling unit.

AREA 1300 - GAS COOLING

The gas cooling unit is designed to cool the raw gas from gasification and to condense the heavier hydrocarbons and unreacted steam. The cooling scheme is arranged to recover and utilize as much of the process heat as is practical. The exit gas stream is conveyed to the sulfur recovery unit.

AREA 1600 - PHENOSOLVAN UNIT

The process water from the Gas Liquor Separation Unit which is contaminated with phenols, ammonia, hydrogen sulfide and carbon dioxide is treated in the Phenosolvan Unit prior to use as makeup water in the process gas liquor cooling tower. Crude phenol and liquid anhydrous ammonia byproducts are produced.

The incoming process water is passed through gravel filters for removal of suspended matter and then through extractors where an organic solvent is used to extract phenols. The organic solvent is distilled and separated from the phenol and recycled to the extractors for reuse.

The crude phenol byproduct is recovered and transferred to storage for subsequent sale.

After removal of all traces of solvent the dephenolized process water is stripped in the deacidifier to remove dissolved carbon dioxide and hydrogen sulfide which is processed in the sulfur recovery unit. The resultant process water is distilled to recover 25% aqua ammonia which is further distilled to produce commercially pure anhydrous ammonia. The remaining process water is then utilized as cooling water makeup to the process water cooling tower.

AREA 1800 - GAS LIQUOR SEPARATION

The gas liquor contains tar, tar oil, naphtha, and dissolved compounds such as phenols, ammonia, carbon dioxide and hydrogen sulfide. Tar is defined as a heavier-than-water organic liquid phase, while tar oil is the lighter-than-water organic liquid phase.

The gas liquor separation is designed to clean up tarry and oily gas liquors by separating the incoming streams into tar, tar oil, recycled gas liquor and clarified aqueous liquor streams. Flash gases released from the gas liquor by pressure reduction are scrubbed to remove ammonia.

The gas liquor streams originating from the gasification and gas cooling units are cooled, combined and reduced in pressure.

The liquor flows to a large separator from which tar is removed. The tar is retained for export as fuel. The net liquor flow is forwarded to a second separator where tar oil is separated and removed.

The gas liquor passes to a final separator where additional tar oil is removed and sent to storage along with tar oil from the second separator.

The aqueous stream from the final separator passes to intermediate tankage before being fed to the Phenosolvan unit.

AREA 2000 - COAL STORAGE AND PREPARATION

Run-of-mine coal is delivered to the plant. After crushing and screening the coal is delivered through a system of conveyors to the live storage pile having a storage capacity of about six days production for the gasification plant. A dead storage pile is also provided with a capacity sufficient to ensure approximately 30 days production requirements. From the live storage pile, the coal is fed continuously through screens to achieve the correct size distribution for gasification. Coal fines not suitable for gasification are routed to storage silos and are available for sale.

AREA 3000 - OXYGEN PLANT

The oxygen facilities are designed to provide gaseous oxygen to the process plant with an oxygen purity of 99.5 percent. 7800 T/D are estimated to be required for the 2000 MW unit and 4000 T/D for the 900 MW unit. Parallel process trains are utilized with each unit consisting of a turbine driven axial/centrifugal air compressor, air separation section (cold box) and a turbine driven centrifugal oxygen compressor. The air compressor turbines are of the extraction/condensing type utilizing 1500 psig, 900°F steam from battery limits. The total makeup steam required for the rest of the complex is extracted at 550 psig with the remainder going to the surface condensers. The oxygen compressor turbines are of the condensing type.

Gasifier startup air and plant general use air is provided from the air compressors. Excess nitrogen from the air separation system is used for the process plant inert gas system. Liquid nitrogen storage is provided to ensure availability of purge gas during a plant outage.

Oxygen plant control will be centralized and combined with the main plant control room.

AREA 4000 — SULFUR RECOVERY

The sulfur in the coal feed to the gasification unit is recovered to a major extent by treating the entire product gas stream in a Stretford unit. Small quantities of sulfur are present in the ash and the liquid by products. Miscellaneous small purge gas streams containing H_2S will be incinerated and disposed of in a 1000 foot stack located at the Thermal Generating Station.

The Stretford process is used to recover elemental sulfur from H_2S present in the product gas stream from the gas cooling unit. This process is licensed by the North Western Gas Board of the British Gas Corporation. The process, using a dilute aqueous solution containing Na_2CO_3, sodium meta-vanadate and anthraquinone di-sulfonic acid (ADA) operates in a continuous regenerative fashion as follows:

H_2S is absorbed from the gas by the alkaline carbonate solution forming HS^- ions. This is accomplished in a counter-current open grid tower. The sulfide is oxidized to free sulfur by vanadate according to the reaction:

$$HS^- + 2V^{5+} = 24^{4+} + S + H^+$$

This reaction proceeds during the absorption step and is completed in a holding vessel. The solution is regenerated by re-oxidation of V^{4+} to V^{5+}. This is accomplished by sparging with air in a separate vessel with ADA acting as a catalyst for the reaction. The sulfur forming as a floating froth is separated from the solution and is processed to produce liquid sulfur.

AREA 5000 — STEAM DISTRIBUTION

High-pressure steam imported from the Thermal Generating Station is used to drive compressors and large pumps and as process steam for coal gasification. Steam generated in the process waste heat boilers is used in turbine drivers and process and heating applications.

1500 psig steam is imported and 550 psig steam is extracted in the air compressor turbines. Low-level steam is generated at 100, 60, and 20 psig. All necessary distribution piping and controls are provided in the design.

AREA 5300 — POWER DISTRIBUTION

A conventional 3-level distribution system for a total requirement of 40,000 KW in the 2000 MW case and 20,000 KW in the 900 MW case are included in the design.

AREA 5400 — RAW WATER SUPPLY & TREATING

Raw water is supplied at the Gasification Plant battery limits. The sequence of water treating steps and flow rates and the interrelation with the waste effluent steps are shown on Sketch No. 7 Water and Waste Effluent Balance.

The system was designed with maximum water reuse. Raw water is required on a continuous basis for cooling tower and potable water makeup.

AREA 5500 — COOLING WATER

Two separate cooling towers are provided to handle the bulk of the plant heat rejection, as follows:

- Cooling tower using treated gas liquor (process water) from the Phenosolvan and Gas Liquor Stripping unit as makeup.
- Separate cooling tower using treated fresh water for the oxygen plants in order to minimize the hazard associated with hydrocarbons entering the system.

Chemical feeding equipment will be provided to permit addition of water treating chemicals to all systems, as required, in order to adjust pH and inhibit corrosion, scale formation, and biological plant growth.

AREA 5600 — FIRE PROTECTION SYSTEM

The fire protection system consists of a fire-water loop, chemical and foam-fire suppression equipment and mobile equipment.

AREA 5700 — MISCELLANEOUS UTILITIES POTABLE WATER

Water for potable and sanitary use is supplied from the raw water treatment plant, where it is filtered chlorinated, and treated prior to distribution.

In addition to supplying water to the major plant buildings, the system will also supply the plant safety showers and eye baths.

INSTRUMENT AIR, PLANT AIR AND INERT GAS

A plant air system is provided to supply compressed air at a nominal 100 psig pressure to shops and service outlets throughout the plant.

Plant air is normally supplied from the oxygen plant main air compressors. During periods of total plant shutdown, motor-driven air compressors supply the plant air requirements. The system is complete with necessary aftercoolers, air receivers, and distribution piping. All plant air will be dried.

An instrument air system is provided to supply clean, dry air for instrument operation.

Instrument air is supplied by oil-free motor-driven compressors, operating at a system pressure of 100 psig. The system is complete with dryers, air receivers, and distribution piping, To ensure maximum reliability during a power outage, these compressors are connected to the emergency power system.

An inert gas system is provided to supply dry nitrogen at 100 psig for purging and blanketing vessels and catalysts in the process areas.

Two half-capacity oil-free motor-driven compressors will deliver nitrogen from the oxygen plant. In addition, a 100 ton liquid nitrogen tank, provided with vaporization facilities, supplies nitrogen to the system during periods of oxygen plant outage.

COMMUNICATIONS

The plant is provided with two communication systems. An in-plant dial telephone system is installed and operated by the telephone company in space provided by the plant. This arrangement avoids capital expenditure and hiring of specialized maintenance skills. The system is automatically monitored against failure to the degree that it is acceptable for fire reporting, and thereby avoids the operation of a separate fire alarm system. The system is arranged to allow outside communication from designated telephones.

Operating communications throughout the plant to roving personnel and vehicles is by radio. Radio paging interconnected with the telephone system is included for contacting non-operating personnel when they are away from their normal stations.

AREA 6000 — OFFSITE STORAGE AND LOADING FACILITIES

Storage facilities are included to store by-products produced in the plant. Storage for a number of raw materials for plant feed is also provided. A minimum of 15-day storage supply is provided for most byproducts and raw materials, except for the anhydrous ammonia storage tank, where 30 days storage is provided.

Liquid elemental sulfur produced in the sulfur recovery unit is stored in a sulfur pit. The sulfur pit, with 15 days storage capacity, is provided with submerged loading pumps and rail loading facilities.

AREA 8100 — LIQUID WASTE EFFLUENT SYSTEM

The liquid waste effluent treatment system is designed to maximize water reuse. The only discharge of water is the water associated with the ash returned to the mine. The effluent treatment system is shown on Sketch No. 7 Water and Waste Effluent Balanc[illegible]

[illegible] k of reusable water is derived from the gas liquor area. After most of the phenols are extracted and H_2S, CO_2 and NH_3 are stripped from this stream, the process condensate is used as make-up water for the process gas-liquor cooling tower. The cooling tower also serves as an oxidation unit for reduction of biological and chemical oxygen demand.

The cooling tower blowdown, together with equipment and area drains and rain water from paved area, is stored in a pond serving as a surge for rain storms and as a safeguard against contaminated cooling water in the event of a heat exchanger leak. The effluent from the pond is treated by gravity oil separation and subsequent flocculation/clarification. Any separated slop oil is stored for disposal by incineration. The underflow from the clarifier, together with sanitary sewage treated in a biological treatment unit, is reused in the Ash Handling system.

AREA 8200 — ASH DISPOSAL

Ash discharge from the gasifier is quenched and sluiced by water to screw classifiers. The classifier discharge drops onto conveyors and is transferred to an ash bin, which is emptied into trucks for disposal in the mine.

AREA 8300 — FLARE SYSTEM

The flare system is capable of flaring the total gas from gasification for short periods. It may be used to flare product during plant startup when gas quality is below the acceptable specifications.

The flare system collects all emergency and operating hydrocarbon vents and burns them at the top of a smokeless flare stack.

The self-supporting flare stack includes ignitor, flame front generators, molecular seal, and continuous burning pilots. Ladder and access platforms will be provided on the flare stack to facilitate maintenance.

AREA 8400 — PRODUCT GAS EXPANDER

A turbine expander and generator were included to recover the energy available when delivering the gas at a pressure of 50 psig.

5.1.2 LURGI AIR BLOWN GASIFICATION

5.1.2a OVERALL PLANT DESCRIPTION

This section of the study covers the design of a grass-roots Coal Gasification plant based upon the LURGI Air Blown Gasification process. The plant includes all process and utility systems, environmental facilities, tankage and buildings but excludes steam and power generation. It is sized to produce a sufficient volume of low Btu fuel gas to support an over the fence Thermal Generating station having a capacity of 900 MW.

The Process Block Flow Diagram and Material Balance Sketch Nos. 3 and 4 show the major processing areas. Sized coal is delivered to the gasifier feed system. The hot gasifier effluent is cleaned of tar and dust by cooling in quench vessels, steam generators, and coolers. The gas stream, which consists of methane, ethane, hydrogen, hydrogen sulfide, carbon monoxide, carbon dioxide and nitrogen, is then purified of hydrogen sulfide, treated, and expanded. Liquids condensed in the cooling process are processed through the gas liquor separation and treating units to obtain the following byproducts: tar, tar oil, crude phenols and anhydrous ammonia. Ash removed from the gasifier is processed and eventually returned to the mine site. The approximate mass balances shown are based on prorating published data and making adjustments due to the differences in the coal analysis and the flowsheet.

Two cases, B and B[1], employing different gas cooling systems, H_2S removal systems and gas liquor treating units were considered. In Case B, the gas is cooled to 250°F prior to entering a hot potassium carbonate wash system. The duty recovered by cooling the gas is used to supply the reboiler heat requirements for the hot potassium carbonate. Additional cooling is accomplished in air-cooled exchangers. Most of the

H_2S and some CO_2 in the process gas are removed by this system. In Case B[1] the process gas is cooled to 90°F and H_2S is removed in a Stretford system. Cooling is accomplished in air-cooled and water-cooled exchangers.

In Case B gas liquor condensed in the waste heat boiler is used to saturate the process gas after it leaves the H_2S removal system. Only the excess gas liquor from the waste heat boiler and gas cooler passes to the gas liquor separation section.

In Case B[1] all of the gas liquor condensed in the waste heat boiler and the gas cooling train is routed to the gas liquor separation section, resulting in a larger gas liquor separation unit than in Case B. In addition, the water leaving the gas liquor separation section in Case B[1] contains a sufficient quantity of phenols to necessitate further treatment in a Phenosolvan unit.

Treated fuel gas leaving the H_2S removal step in both Case B and Case B[1] is preheated and expanded to recover energy. The energy recovered by expanding the gas is used to drive some of the air compressors and to meet most of the plant power requirements. In Case B, an excess of electricity is generated, therefore an export is shown. In case B[1], less energy is recovered, therefore a net import of electricity is required.

High pressure steam imported from battery limits is used to supply the remaining air compression requirements, drive major pumps and as process steam to the gasifiers. Low pressure steam generated in process waste heat boilers in excess of plant requirements is exported. Approximate energy balances were calculated for the gasification section and the overall plant as follows:

GASIFICATION ENERGY BALANCE

	CASE B		CASE B[1]	
	10^9 Btu/D	HHV%	10^9 Btu/D	HHV%
INPUT				
Net Coal to Gasifiers	251.5	100.0	283.1	100.0
OUTPUT				
Fuel(Case B 211 Btu/SCF, Case B[1] 192/SCF at 60°F)	230.0	91.5	230.	81.2
Tar (147,000 Btu/US gal)	11.4	4.5	6.4	2.3
Tar Oil (133,000 Btu/US gal)			12.6	4.5
Crude Phenols (122,000 Btu/US gal)			2.5	.9
Anhydrous Ammonia (9800 Btu/lb)			2.6	.9
	241.4	96.0	254.1	89.8

OVERALL PLANT ENERGY BALANCE

	CASE B		CASE B[1]	
	10^9 Btu/D	HHV%	10^9 Btu/D	HHV%
INPUT				
Net Coal to Gasifiers	251.5	87.0	283.1	87.5
Steam	37.5	13.0	40.2	12.4
Electric Power			.1	.03
	289.0	100.0	323.4	100.0
OUTPUT				
Fuel Gas	230.0	79.6	230.0	71.1
Byproducts	11.4	3.9	24.1	7.5
Electric Power	.5	.2		
	241.9	83.7	254.1	78.6

The same comment concerning Overall Plant Energy Balance which appears in Paragraph 5.1.1a applies here.

The complex is designed with maximum reliability built into the system to minimize the possibility of interruption in gas production. Independent, parallel processing systems are employed to minimize the effect of the loss of a process unit. In addition, spare equipment is provided at critical points.

The environmental control systems are conceived to be integrated with the process units, waste heat recovery, cooling water and ash handling systems and are expected to meet applicable standards and regulations in British Columbia where possible water reuse was incorporated in the design. Descriptions of the liquid waste effluent and sulfur recovery systems are included in the respective sections of the study.

This section of the study is based on the assumption that the gas composition resulting from air gasification of Hat Creek Coal will be similar to the composition from air gasification of a sub-bituminous coal such as New Mexico Coal as described in the filing document submitted to the Federal Power Commission by El Paso Natural Gas Co.

5.1.2b PLANT DEFINITION

The preliminary process design used for this study is based on processing lignite type coal in an air blown LURGI pressure gasification system to produce 230 x 10^9 Btu/D of low Btu gas. The coal was assumed to come from the Hat Creek, B.C. area and to have average properties as shown in Table No. 1.

The Plant Elements forming the basis of this study and estimate are listed below:

AREA	DESCRIPTION
	For both Cases Except Where Indicated
1100	Gasification
1300	Phenosolvan (Case B^1 only)
1600	Gas Cooling
1800	Gas Liquor Separation
2000	Coal Preparation and Handling
4000	Sulfur Removal and Recovery
5000	Steam Distribution
5300	Power Distribution
5400	Raw Water Supply and Treating
5500	Cooling Water
5600	Fire Protection System
5700	Miscellaneous Utilities
6000	Offsite Storage and Loading Facilities
7000	Plant Interconnecting Piping
8100	Liquid Waste Effluent System
8200	Ash Disposal
8300	Flare Sytem
8400	Product Gas Expansion

Areas 1100, 1300, 1600, and 1800 are based on published process information on the LURGI technology. The prelimary design of all other areas was done by Lummus. A brief description of each area follows:

AREA 1100 — GASIFICATION

The description of Area 1100 is essentially the same as that shown in Paragraph 5.1.1b above, with the exception that air and steam instead of oxygen and steam are the gasifying agents.

AREA 1300 — GAS COOLING

The gas cooling is designed to cool the raw gas from gasification and to condense the heavier hydrocarbons and unreacted steam before purification. Gas cooling is accomplished in multiple parallel trains.

In Case B where sulfur is removed in a hot potassium carbonate system the cooling scheme is arranged to recover and utilize some of the process heat to provide the energy required for regeneration of the hot potassium carbonate solution.

In Case B[1] cooling is accomplished in air coolers and final cooling to meet the lower temperature requirements for the Stretford Unit is accomplished by cooling water.

AREA 1600 — PHENOSOLVAN UNIT

The description of Area 1600 for Case B' is identical to the description provided in Paragraph 5.1.1b above.

No Phenosolvan Unit is required for Case B.

AREA 1800 — GAS LIQUOR SEPARATION

The description of Area 1800 is nearly identical to the description provided in Paragraph 5.1.1b above.

The only difference occurs in Case B. No Phenosolvan Unit is required in Case B, the excess water is filtered and reused in the cooling water system.

AREA 2000 — COAL PREPARATION AND HANDLING

The description of Area 2000 is identical to that in Paragraph 5.1.1b above.

AREA 4000 — SULFUR REMOVAL

CASE B — MODIFIED HOT POTASSIUM CARBONATE SYSTEM

Hydrogen sulfide is removed from the process gas by absorption in circulating hot potassium carbonate solution. Since CO_2 is also readily absorbed by potassium carbonate solution, the system is designed to maximize the absorption of H_2S with minimal removal of CO_2. The unit consists of an absorption tower and a regenerator. Process gas contacts the solution in the absorber and H_2S is absorbed relatively quickly. The absorber design uses trays rather than packing, because the residence time between the gas and the solution is minimized, thus minimizing the absorption of CO_2. The spent solution from the absorber flows to a regenerator where the absorbed H_2S and CO_2 are stripped out by heat supplied to the reboiler. The lean solution from the bottom of the regenerator is pumped back to the top of the absorber. The acid gas leaving the regenerator overhead passes to the sulfur recovery unit.

CASE B — CLAUS UNIT

In Case B, the acid gas leaving the H_2S removal section (modified hot potassium carbonate system) passes to a Claus Unit where elemental sulfur is produced. The acid gas is preheated, mixed with air and passed directly to a catalytic reactor containing a bauxite type catalyst where the following reactions occur:

$$H_2S + 3/2\,O_2 = SO_2 + H_2O + \text{Heat}$$
$$2H_2S + SO_2 = 3S + 2H_2O + \text{Heat}$$

The gas is then cooled and sulfur is condensed out and routed to storage.

AREA 4000 — SULFUR REMOVAL

CASE B^1— STRETFORD UNIT

The description of Area 4000 for Case B^1 is identical to that in Paragraph 5.1.1b above.

AREA 5000 — STEAM DISTRIBUTION

Imported steam at 1500 psig and 900°F is let down through extraction condensing turbines driving some of the air compressors. Some of the steam is extracted at 400 psig from the air compressor turbines and is supplied to the gasifiers and to turbine drivers for several pumps. The remaining steam is condensed at 4" Hga.

Steam generated from waste heat in the gasification unit is fed to the 40 psig steam system. This steam is used for heating, tracing, and deaeration. Excess steam from the 40 psig steam system is exported back to battery limits.

Condensate, make-up from battery limits, and 40 psig steam are mixed and deaerated. High pressure boiler feed pumps provide feed water to the gasifier water jackets. Low pressure boiler feed pumps provide feed water to the 40 psig waste heat boilers.

AREA 5300 — POWER DISTRIBUTION

A conventional 3-level distribution system for an estimated total requirement of 1500 KW is included in the design.

AREA 5400 — RAW WATER SUPPLY & TREATING

The description of Area 5400 is identical to that provided in Paragraph 5.1.1b above.

AREA 5500 — COOLING WATER

One cooling tower system is provided to handle the bulk of the plant heat rejection using treated fresh water as well as treated gas liquor (process water) from the Phenosolvan and Gas Liquor Stripping unit as make-up. Chemicl feeding equipment will be provided to permit addition of water-treating chemicals to all systems as required, in order to adjust pH and inhibit corrosion, scale formation and biological plant growth.

The description of the following areas are identical to the description provided in Paragraph 5.1.1b:

Area 5600 — Fire Protection System
Area 5700 — Miscellaneous Utilities
Area 6000 — Offsite Storage and Loading Facilities
Area 8100 — Liquid Waste Effluent System
Area 8200 — Ash Disposal
Area 8300 — Flare System

AREA 8400 — PRODUCT GAS EXPANSION

Following H_2S removal and final treating, the process gas in expanded to 50 psig to recover energy. In Case B the gas is first saturated with gas liquor from the waste heat boilers and preheated in a fired heater prior to entering the expander.

In Case B^1 the gas is just preheated prior to expansion.

5.1.3 KOPPERS-TOTZEK 200 MW AND 900 MW

5.1.3a OVERALL PLANT DESCRIPTION

This section of the study covers a grass-roots coal gasification plant based upon the KOPPERS-TOTZEK atmospheric pressure coal gasification process. It is sized to produce a sufficient volume of low Btu fuel gas to support an over-the-fence Thermal Generating Station having a capacity of 2000 MW and, as an alternate, a 900 MW station. Non-proprietary process unit, offsites, sulfur recovery, and waste effluent control systems were designed by Lummus.

The Process Flow Diagram and Material Balance, Sketch No. 4, shows the major processing areas. Pulverized dried coal is delivered to the KOPPERS-TOTZEK oxygen-blown gasifiers. The crude product gas is cooled and treated for sulfur removal before delivery to the Thermal Generating Station. Liquids condensed in the cooling process are processed through the clarifier, where soot is removed. The water is recycled back to the process. Ash removed from the gasifier is processed and returned to the mine site. The approximate mass balance shown is based on information received from KOPPERS-TOTZEK concerning the gasification of North Dakota lignite.

Approximate energy balances were calculated for the gasification section and the overall plant, as follows:

GASIFICATION ENERGY BALANCE

	2000 MW 10^9 Btu/D	900 MW 10^9 Btu/D	HHV%
INPUT			
Coal to Gasifiers	570.9	293.1	100.0
OUTPUT			
Low Btu Fuel Gas (292 Btu/SCF at 60°F)	450.0	230.0	78.0

OVERALL PLANT ENERGY BALANCE

	2000 MW 10^9 Btu/D	900 MW 10^9 Btu/D	HHV%
INPUT			
Coal to the Plant	570.9	293.1	86.8
Fuel for Drying Coal	23.3	11.9	3.5
Electric Power	2.1	1.0	0.3
Steam Import	61.8	34.3	9.4
	618.1	340.3	100.0
OUTPUT			
Low Btu Fuel Gas (292 Btu/SCF at 60°F)	450.0	230.0	67.6
Low Pressure Steam	6.7	3.4	1.0
	456.7	233.4	68.6

The overall Thermal Efficiency for the KOPPERS-TOTZEK process is significantly lower than that for LURGI because KOPPERS-TOTZEK requires the coal be dried to about 8% moisture, whereas LURGI feeds wet coal (25% moisture) directly into the gasifier. KOPPERS-TOTZEK believes it is less expensive to remove water from the coal before gasification rather than after gasification. There is no net fines production in a KOPPERS-TOTZEK system. The coal is pulverized and all of it is fed to the gasifier.

A preliminary examination of plot requirements indicate that approximately 100 acres is required. The 900 MW case requires less area, but it is recommended that a 100 acre area be purchased for both cases.

The complex is designed with maximum reliability built into the system to minimize the possiblity of interruption in gas production. Independent, parallel processing systems are employed to minimize the effect of the loss of a process unit. In addition, spare equipment is provided at critical points.

The environmental control systems are conceived to be integrated with the process units, waste heat recovery, cooling water, and ash handling systems and are expected to meet applicable standards and regulations in British Columbia. Where possible, water reuse was incorporated in the design. Detailed descriptions of the liquid waste effluent and sulfur recovery systems are included in the respective sections of the study.

5.1.3b PLANT DEFINITION

The preliminary process design is based on processing lignite type coal in an oxygen-blown KOPPERS-TOTZEK atmospheric pressure gasification system to produce a sufficient quantity of low Btu fuel gas to feed a 2000 MW Thermal Generating Station and, as an alternate case, a 900 MW station.

The Hat Creek coal was assumed to have average properties as shown in Table No. 1.

In operation, raw coal is pulverized and dried to about 8%. The pulverized coal is then fed into the KOPPERS-TOTZEK gasifier by means of specially-designed screw feeders. Oxygen and steam are also fed into the reactor. The KOPPERS-TOTZEK gasifier operates at essentially atmospheric pressure. The raw gas and some slag move vertically into the waste heat boiler, and the remainder of the slag flows through the bottom of the gasifier into a quench system. About half the slag goes with the raw gas. Just prior to entering the waste heat boiler, the gas is quenched with a water spray which solidifies any slag so it doesn't adhere to the waste heat boiler tubes.

The raw gas is cooled in the waste heat boiler, generating high-pressure, superheated steam, (1500 psig/900°F).

Because the raw gas contains a high percentage of particulate matter, it must be scrubbed thoroughly. This is accomplished in a two-stage venturi scrubbing system followed by a packed tower. The heat picked up by the water in this scrubbing and cooling sequence is removed from the water by a cooling tower, and the water is recirculated.

The clean gas is then processed in a Stretford Unit, which removes the H_2S. The clean desulfurized gas is then compressed to approximately 30 psig for use in a Thermal Generating Plant.

The slag is quenched and solidified in a holding tank and then is transmitted back to the mine. Most of the quench water is recovered and recirculated through the cooling tower.

The KOPPERS-TOTZEK process is not self-sufficient in high-pressure steam, so steam is imported from battery limits at 1500 psig/900°F. Most of the major equipment drivers are on steam turbines. Condensate and low-pressure steam are exported back to battery limits.

There are two cooling water systems. The process cooling water is used to scrub and cool the raw product gas and quench the slag from the gasifier. The fresh water system is kept separate from the process water and is used for the inter and aftercoolers for the air and oxygen compressors and the surface condensers for the turbine drivers.

The purpose is to prevent possible hydrocarbon leakage into the air separation unit and the oxygen compressor.

Other support facilities such as storage and water treatment are also supplied. The plant elements forming the basis of this study and estimate are listed below:

AREA	DESCRIPTION
1100	Gasification
1300	Gas Cooling and Scrubbing
1900	Product Gas Compression
2000	Coal Preparation and Handling
3000	Oxygen Plant
4000	Sulfur Recovery
5000	Steam Distribution
5300	Power Distribution
5400	Raw Water Supply and Treating
5500	Cooling Water
5600	Firewater
5700	Miscellaneous Utilities
6000	Offsite Storage and Loading Facilities
7000	Plant Interconnecting Piping
8100	Liquid Waste Effluent System
8200	Ash Disposal
8300	Flare System

Areas 1100, and 1300 are based on information received from KOPPERS-TOTZEK on the gasification of North Dakota lignite. All other areas were designed by Lummus. A brief description of each area follows:

AREA 1100 — GASIFICATION

Four-headed gasifiers, capable of gasifying over 800 tons of coal per day each, are used. The oxygen, steam, and coal react in the refractory-lined steel shell gasifier at a slight positive pressure (5-7 psig). Coal, oxygen, and steam are brought together in opposing burner heads spaced 90° apart. These units resemble intersecting ellipsoids having a major axis of approximately 25 feet and a minor axis of 13 feet.

Gasification of the coal is almost complete and instantaneous. Carbon conversion is a function of the reactivity of the coal, approaching 100 percent for lignite type coals.

Exothermic reactions produce a flame temperature of approximately 3500°F. Endothermic reactions, occurring in the gasifier between carbon and steam and heat radiation to the refractory walls, substantially reduce the flame temperature from 3500°F to 2700°F. Low-pressure process steam for the gasifier reaction is produced in the gasifier jacket from the heat passing through the refractory lining.

Ash in the coal feed is liquefied in the high-temperature zone. Approximately 50 - 70 percent of the molten slag drops out of the gasifier into a slag quench tank and is recovered for disposal as a granular solid. The remainder of the slag and any unreacted carbon are entrained in the gas exiting the gasifier. Water sprays quench the gas to drop the temperature below the ash fusion temperature to prevent slag particles from adhering to the tubes of the waste heat boiler mounted atop the gasifier. Ash fusion characteristics can be adjusted if necessary by the addition of flux to the coal feed.

AREA 1300 — GAS COOLING AND SCRUBBING

The raw gas from the gasifier passes into the waste heat boiler, where high-pressure, superheated steam is produced.

After leaving the waste heat boiler, the gas is cleaned and cooled in a high-energy scrubbing system. The system consists of a fixed orifice venturi-type scrubber, for removing the largest particles (95 percent of total), followed by a variable orifice venturi-type scrubber, where more than 99 percent of the remaining particles are removed. The entrained solids in the gas are thus reduced to 0.002 to 0.003 grains per SCF. Following scrubbing, the gas is cooled with water in a packed tower.

Particulate-laden water from the gas cleaning and cooling system is piped to a clarifier. Sludge from the clarifier is pumped to the ash disposal area. Clarified water is recirculated through the venturi scrubbers, and the excess overflows into the cooling tower system at the gas cooler. Evaporation, windage, and blowdown water losses at the cooling tower, plus moisture in the clarifier sludge and slag, necessitate the addition of make-up water to this system.

AREA 1900 — PRODUCT GAS COMPRESSION

The product gas compression unit consists of double flow centrifugal compressor trains, each driven by extraction/condensing steam turbines. The gas is compressed from 2 psig to 30 psig in a single-stage machine, and is delivered to the Thermal Generating Station.

AREA 2000 — COAL PREPARATION

Depending upon rank, the coal is dried to between 2 percent and 8 percent moisture and pulverized to 70 percent through 2000 mesh. The pulverized coal is conveyed with nitrogen from storage to the gasifier service bins. Controls regulate the intermittent feeding of coal from the service bins to the feed bins which are connected to twin variable-speed coal screw feeders. The pulverized coal is continuously discharged from each screw into a mixing nozzle, where it is entrained in oxygen and low-pressure steam. The mixture is then delivered through a transfer pipe to the burner head of the gasifier. Moderate temperature and high velocity in the burner prevent the reaction of the coal and the oxygen until they enter the gasification zone.

AREA 3000 — OXYGEN PLANT

With the exception of capacity, the description of Area 3000 is identical to that given in Paragraph 5.1.1b. The capacities required for KOPPERS-TOTZEK are 20,300 T/D and 10,600 T/D of oxygen for the 2000 MW equivalent and 900 MW equivalent plants respectively.

AREA 4000 — SULFUR RECOVERY

The description of Area 4000 is identical to that in Paragraph 5.1.1b above.

AREA 5000 — STEAM DISTRIBUTION

The KOPPERS-TOTZEK design generates 1500 psig/900°F steam in the waste heat boilers and 15 psig/saturated steam in the reactor jackets. High pressure (1500 psig/900°F) steam is used to drive the large compressors and pumps and is also used in some process applications after extraction at intermediate pressure. As the waste heat boilers do not generate sufficient high-pressure steam, some high-pressure steam is imported from battery limits.

Low-pressure (15 psig) steam is generated in the gasifier jackets. Some of this steam is used in the process, some is used to heat condensate used for boiler feedwater, and the excess is exported along with excess 50 psig steam and steam condensate back to battery limits.

AREA 5300 — POWER DISTRIBUTION

A conventional 3-level distribution system for a total requirement of 26000 KW for the 2000 MW Thermal Generating Station and 13000 KW for the 900 MW Thermal Generating Station are included in the design.

AREA 5400 — RAW WATER SUPPLY AND TREATING

Raw water is supplied at the Gasification Plant Battery Limits. The sequence of water treating steps and the interrelation with the waste effluent steps, are similar to those shown on Sketch No. 7, Water and Waste Effluent Balance.

The system was designed for maximum water reuse. Raw matter is required for cooling tower and potable water make-up.

AREA 5500 — COOLING WATER

Two separate cooling towers are provided to handle the bulk of the plant heat rejection, as follows:

— Cooling tower circulating only process water

— Cooling tower for oxygen plant and compressor intercooler and surface condensers

Chemical feeding equipment will be provided to permit addition of water treating chemicals to all systems, as required, in order to adjust pH and inhibit corrosion, scale formation, and biological plant growth.

OTHER AREAS

The descriptions of the following areas are identical to those in Paragraph 5.1.1b above:

Area 5600 — Fire Protection System
Area 5700 — Miscellaneous Utilities
Area 6000 — Offsite Storage and Loading Facilities
Area 8300 — Flare System

AREA 8100 — LIQUID WASTE EFFLUENT SYSTEM

The liquid waste effluent treatment system is designed to maximize water reuse. The only discharge of water is the water associated with the ash returned to the mine. The effluent treatment scheme is similar to that shown on Sketch No. 7, Typical Water and Waste Effluent Balance.

The fresh water cooling tower blowdown, together with equipment and area drains and rain water from paved areas, is stored in a pond serving as a surge for rain storms and as a safeguard against contaminated cooling water in the event of a heat exchanger leak. The effluent from the pond is treated by gravity oil separation and subsequent flocculation/clarification. Any separated slop oil is stored for disposal by incineration. The underflow from the flocculator/clarifier is sent to ash handling. The overflow from the clarifier, together with sanitary sewage treated in a biological treatment unit is reused in the Ash Handling system. Excess water is returned to the proces water cooling tower as make-up.

AREA 8200 — ASH DISPOSAL

Slag discharged from the gasifiers is quenched and slurried by water to screw classifiers. The classifier discharge drops onto conveyors and is transferred to an ash bin, which is emptied into trucks for disposal in the mine.

5.2 TOWN GAS

5.2.1 OVERALL PLANT DESCRIPTION

This section of the study covers a Town Gas Plant based on LURGI Oxygen-Blown Coal Gasification Technology. The plant is to be located on Vancouver Island, British Columbia. It includes all process and utility systems, environmental facilities, tankage, and buildings, excluding steam and electric power generation. The plant is sized to produce 250 MM SCFD of low Btu town gas for distribution. Non-proprietary process units, off sites, sulfur recovery, and waste effluent control systems were designed by Lummus.

The Process Block Flow Diagram and Material Balance, Sketch No. 5, shows the major processing areas. Sized coal is delivered to the LURGI oxygen-blown gasifiers. The crude gas out of the gasifiers is passed through shift converters to reduce its carbon monoxide content and then cooled and treated for sulfur removal before leaving the plant for distribution. Liquids condensed in the cooling process are processed through the gas liquor separation and treating units to obtain the following by-products: tar, tar oil, crude phenols, and anhydrous ammonia.

Ash removed from the gasifiers is processed and eventually returned to the mine site or other land-fill operation. The approximate mass balance shown is based on prorating the data disclosed by the ANG Coal Gasification Company in their filing document to the Federal Power Commission of the U.S. Government and making adjustments dictated by the differences in the coal analysis and the plant flowsheet. The study assumed that the Hat Creek Coal would gasify similarly to the North Dakota Lignite used by the ANG Coal Gasification Company. This assumption has to be verified by LURGI.

Approximate energy balances were calculated for the gasification section and the overall plant, as follows:

GASIFICATION ENERGY BALANCE

INPUT	10^9 Btu/D	HHV %
Coal to Gasifiers (at 20% moisture & 25% ash)	82.8	100
OUTPUT		
Town Gas (280 Btu/SCF at 60°F)	71.6	80.5
Tar (147,000 Btu/US gal)	2.2	2.7
Tar Oil (133,000 Btu/US gal)	3.5	4.2
Crude Phenols (122,000 Btu/US gal)	0.7	0.8
Anhydrous Ammonia (9800 Btu/lb)	0.9	1.1
	78.9	95.3

OVERALL PLANT ENERGY BALANCE

INPUT	10^9 Btu/D	HHV %
Net Coal to Gasifiers	82.8	77.3
Electric Power	1.0	0.9
Steam Import	23.3	21.8
	107.1	100.0
OUTPUT	10^9 Btu/D	HHV %
Town Gas	71.6	66.9
By-Products	7.3	6.8
Steam Export	0.7	0.6
	79.6	74.3

The statements concerning the Overall Plant Energy Balance which appear in section 5.1.1a also apply here.

The Plot Plan will be similar to the one shown as Sketch No. 9 in the Low Btu section of this report.

The complex is designed with maximum reliability built into the system to minimize the possibility of interruption in gas production. Independent, parallel processing systems are employed to minimize the effect of the loss of a process unit. In addition, spare equipment is provided at critical points.

The environmental control systems are conceived to be integrated with the process units, waste heat recovery, cooling water, and ash handling systems and are expected to meet applicable standards and regulations in British Columbia. Where possible, water reuse was incorporated in the design. Description of the liquid waste effluent and sulfur recovery systems are included in the respective sections of the study.

5.2.2 PLANT DEFINITION

The preliminary process design used for this study is based on processing lignite-type coal in an oxygen-blown LURGI pressure gasification system to produce 250 MM SCFD of low Btu town gas containing approximately 280 Btu/SCF. The Gasification Plant is located on Vancouver Island but uses coal from the Hat Creek, British Columbia area, having average properties as shown in Table No. 1.

The Plant elements forming the basis of this study and estimate are listed below:

AREA	DESCRIPTION
1100	Gasification
1200	Shift Conversion
1300	Gas Cooling
1600	Phenosolvan
1800	Gas Liquor Separation
2000	Coal Handling and Preparation
3000	Oxygen Plant
4000	Sulfur Recovery
5000	Steam Distribution
5300	Power Distribution
5400	Raw Water Supply & Treating
5500	Cooling Water
5600	Fire Water
5700	Miscellaneous Utilities
6000	Off-site Storage & Loading Facilities
7000	Plant Interconnecting Piping
8000	Ash Disposal
8100	Liquid Waste Effluent System
8300	Flare System

Areas 1100, 1200, 1300, 1600, and 1800 are based on published process information on the LURGI technology. The preliminary design of all other areas was done by Lummus. A brief description of each area follows:

The descriptions of the following areas, with exception as noted, are similar to the descriptions in section 5.1.1b.

AREA	DESCRIPTION
1100	Gasification
1300	Gas Cooling
1600	Phenosolvan
1800	Gas Liquor Separation
2000	Coal Handling and Preparation
3000	Oxygen Plant; capacity = 1200 T/D
4000	Sulfur Recovery
5000	Steam Distribution
5300	Power Distribution; capacity = 12400 KW
5400	Raw Water, Supply & Treating
5500	Cooling Water
5600	Fire Protection System
5700	Miscellaneous Utilities
6000	Off-site Storage & Loading Facilities
8100	Liquid Waste Effluent System
8200	Ash Disposal
8300	Flare System

The only area not covered in section 5.1.1b is Area 1200, Shift Conversion. A description of Area 1200 follows:

AREA 1200 — SHIFT CONVERSION

The carbon monoxide content of the crude gas is too high for town gas distribution. It must be reduced by converting a portion of it to carbon dioxide. This is accomplished through the "water gas shift" reaction carried out catalytically in the presence of steam, as follows:

$$CO + H_2O = CO_2 + 16{,}538 \text{ Btu/lb. mole}$$

The CO content is reduced to less than 7.5 percent (mole) in the present study.

The converted gas from this area flows to the gas cooling unit to remove the hydrocarbon by-products and unreacted steam.

5.3 PIPELINE QUALITY GAS

5.3.1 OVER PLANT DESCRIPTION

This section of the study covers a grass roots Coal Gasification Plant to be located at Hat Creek, British Columbia. The plant includes all process and utility systems, environmental facilities, tankage and buildings but excludes steam and power generation. It is sized to produce 250 MM SCFD of pipeline quality gas having a minimum HHV of 950 Btu/SCF measured at 14.72 psia, 60°F and dry basis. The design is based on the well-known and commercially proven, LURGI Pressure Gasification Process. Non-proprietary process units, offsites, sulfur recovery and waste effluent control systems were designed by Lummus.

The Process Flow Diagram and Material Balance Sketch No. 6 shows the major processing areas. Sized coal is delivered to the gasifier feed system. The coal enters each gasifier through a lock-hopper system and passes downward while being gasified. Steam and oxygen are introduced at the bottom of the gasifier to effect the coal gasification reactions. A revolving grate supports the coal bed, cleans out the ash, and distributes the steam-oxygen mixture. The gasifier is designed to remove the ash as a solid particulate through an ash lock hopper. The ash is dumped into a hydraulic sluicing system and is conveyed to the ash handling area where it is concentrated by mechanical means and is subsequently trucked to the mine for disposal.

The hot crude product gas leaving the gasifier reactor is cleaned of tar and dust by cooling in quench vessels, steam generators, and coolers. The tar oil is recovered from the quench water. A portion of the crude gas is passed through shift conversion reactors. The gas leaving the shift converters is combined with the portion of the crude gas that by-passed the shift converters after the streams have been cooled through heat exchangers to recover sensible heat, additional tar, oil and naphtha. The gas stream, which consists of methane, ethane, hydrogen, hydrogen sulfide, carbon monoxide, carbon dioxide and nitrogen, is then purified of hydrogen sulfide and carbon dioxide. This is accomplished by absorption using cold methanol in the Rectisol unit. The purified gas is methanated to remove almost all of the carbon monoxide and a portion of the carbon dioxide by reaction with hydrogen to produce methane. This raises the heating value so that the synthetic gas can be blended with natural gas. The product gas is compressed for delivery to the pipeline.

The oxygen required for the gasifiers is supplied from the self-contained oxygen plant included in the complex.

The sulfur recovery plant consisting of the Stretford, Claus and IFP units receives gas containing hydrogen sulfide from the Rectisol unit and converts it into salable elemental sulfur in liquid form.

The energy requirements of the plant are largely satisfied by recovering waste heat. High pressure steam imported to the plant battery limits is used to drive large compressors, pumps and as feed to the process. Steam generated in process waste heat boilers is used to drive some compressors and pumps and for process and heating service. Electric power is imported to satisfy the remaining energy requirements of the plant.

The approximate mass balance used for this study is based on prorating the data disclosed by the ANG Coal Gasification Company in the filing document to the Federal Power Commission of the U.S. Government and making adjustments dictated by the differences in the coal analysis and the plant flowsheet. The study was prepared on the assumption that Hat Creek Coal would gasify similarly to the North Dakota Lignite used by the ANG Coal Gasification Company. This assumption has to be checked by LURGI.

Approximate energy balances were calculated for the gasification section and the overall plant as follows:

GASIFICATION ENERGY BALANCE

	10^9Btu/D	% HHV
INPUT		
Net Coal To Gasifiers	356.9	100.0
OUTPUT		
Pipeline Quality Gas (970 Btu/SCF at 60°F)	242.5	67.9
Tar (147,000 Btu/US gal)	11.7	3.3
Tar Oil (133,000 Btu/US gal)	17.4	4.9
Naphtha (396,000 Btu/US gal)	17.4	1.5
Crude Phenols (122,000 Btu/US gal)	3.3	0.9
Anhydrous Ammonia (9800 Btu/lb)	4.1	1.1
	284.4	79.7

OVERALL PLANT ENERGY BALANCE

	10^9 Btu/D	% HHV
INPUT		
Net Coal To Gasifiers	356.9	84.8
Steam	59.3	14.1
Electric Power	4.6	1.1
	420.8	100.0
OUTPUT		
Pipeline Quality Gas	242.5	57.6
Liquid Byproducts	41.9	10.0
	284.4	67.6

The statements concerning Thermal Efficiency which appear in Paragraph 5.1.1a also apply here.

The complex is designed with maximum reliability built into the system to minimize the possibility of interruption in gas production. Independent, parallel processing systems are employed to minimize the effect of the loss of a process unit. In addition, spare equipment is provided at critical points. The process train philosophy is similar to that for the Low Btu Fuel Gas Case as shown on the Process Train Arrangement Diagram shown on Sketch No. 8.

The environmental control systems are conceived to be integrated with the process units, waste heat recovery, cooling water and ash handling systems and are expected to meet applicable standards and regulations in British Columbia. Where possible, water reuse was incorporated in the design. Detailed descriptions of the liquid waste effluent and sulfur recovery systems are included in the respective sections of the study.

5.3.2 PLANT DEFINITION

The preliminary process design used for this study is based on processing lignite-type coal in an oxygen-blown LURGI pressure gasification system to produce 250 MM SCFD of pipeline quality gas. The coal was assumed to come from the Hat Creek, B.C. area and to have the average properties shown in Table 1.

The Plant Elements forming the basis of this study and estimate are listed below:

AREA	DESCRIPTION
1100	Gasification
1200	Shift Conversion
1300	Gas Cooling
1400	Rectisol and Refrigeration
1500	Rectisol
1600	Phenosolvan
1700	Methanation
1800	Gas Liquor Separation
1900	Product Gas Compression
2000	Coal Handling and Preparation
3000	Oxygen Plant
4000	Sulfur Recovery
5000	Steam Distribution
5300	Power Distribution
5400	Raw Water Supply and Treating
5500	Cooling Water
5600	Fire Protection System
5700	Miscellaneous Utilities
6000	Offsite Storage and Loading Facilities
7000	Plant Interconnecting Piping
8100	Liquid Waste Effluent System
8200	Ash Disposal
8300	Flare System

Areas 1100, 1200, 1300, 1400, 1500, 1600, 1700, and 1800 are based on published process information on the LURGI technology. The preliminary design of all other areas was done by Lummus. A brief description of each area follows:

AREA 1100 — GASIFICATION

The coal is gasified using a multiple train of LURGI pressure gasifiers. High pressure oxygen and superheated steam are passed in countercurrent flow through a moving bed of coal, resulting in nearly complete conversion of the coal to gaseous compounds some of which are subsequently condensed and processed in the liquid state.

Sized coal is conveyed from battery limits to coal bunkers located above the gasifiers. Coal is charged to gasifiers through automatically operated coal locks which are depressured before receiving coal from the bunkers. The lock gas from depressuring is recompressed and reinjected in the gas cooling unit.

After filling with coal the coal lock is pressurized with crude gas from the gas cooling unit and is equalized with the pressure of the gasifier. The coal is then charged to the top of the coal bed in the gasifier.

The gasification agent consisting of a mixture of oxygen and superheated steam is introduced through a rotating grate below the ash bed at the bottom of the gasifier. Partial combustion of the coal with the oxygen supplies the heat necessary for the gasification reactions.

Besides the crude gas produced, the process yields tar, oil, naphtha, phenols, ammonia and sulfur.

The ash produced is removed by the rotating grate and discharged through a semi-automatically operated ash lock into an ash chute. From the ash chute the ash is quenched with water and transferred to ash disposal.

Steam is generated in the jacket of the gasifier from the combustion heat and acts to cool the inner wall of the gasifier. The steam passes through a knock-out drum and is returned directly to the gasifier, partially furnishing the steam required for gasification.

The hot crude gas leaving the gasifier is directly quenched in a wash cooler with recycled tarry gas liquor. Tarry gas liquor produced in excess of the recycle is sent to the gas liquor separation unit.

The crude gas freed from dust and heavy tars and saturated with steam is further cooled by generation of 100 psig steam in the waste heat exchanger. The crude gas is then divided, with a portion passing directly to the gas cooling unit, whereas the remainder is routed to the shift conversion unit.

AREA 1200 — SHIFT CONVERSION

The amount of methane (the principal component of natural gas) in the crude gas from the gasification unit is quite low and further chemical conversion of the crude gas to increase the methane content is necessary. This conversion is performed in the Crude Gas Shift and Methanation Units. The shift conversion unit is designed to produce hydrogen required to adjust the H_2:CO ratio for proper feed to the methanation unit. This is accomplished through the "water gas shift" reaction carried out catalytically in the presence of steam as follows:

$$CO + H_2O = CO_2 + 16{,}538 \text{ Btu per lb. mole.}$$

Approximately 40 percent of the total crude gas is subjected to shift conversion with the balance bypassed directly to the gas cooling unit. The proportions of the gas streams are adjusted to achieve the desired H_2:CO ratio for methanation.

The converted gas from this area flows to the gas cooling unit to remove the hydrocarbon byproducts and unreacted steam.

AREA 1300 — GAS COOLING

The gas cooling unit is designed to cool the raw gas from gasification and shift conversion to remove the heavier hydrocarbons and unreacted steam before low temperature purification. The cooling scheme is arranged to recover and utilize as much of the process heat as is practical. Further cooling is accomplished in air coolers and final cooling is by cooling water.

The gas cooling is accomplished in two parallel trains, each train being further subdivided into two lines of exchangers. One line is for cooling the crude gas bypassing the shift conversion area and the other for cooling the converted gas. Converted gas is then compressed and combined with the crude gas stream. The mixed gas stream having a pre-determined H_2:CO ratio is conveyed to the gas purification unit.

AREA 1400 & 1500 — RECTISOL UNIT

The gas purification unit utilizes the Rectisol process to remove CO_2, sulfur compounds, and other impurities from the raw gas. Low temperature methanol is utilized to absorb the carbon dioxide and sulfur compounds. Sulfur compounds are removed to a level of less than 0.1 ppm (by volume) so that the gas meets the requirements for methane synthesis.

The crude gas from the gas cooling unit is chilled before entering the prewash tower to recover naphtha and water. The naphtha free gas then enters an absorber where sulfur compounds and the bulk of the CO_2 are removed by a cold methanol wash. The stripped acid gas streams are directed to the sulfur recovery unit for conversion to elemental sulfur. The sulfur-free gas exits the Rectisol Unit and passes to methanation. Following methanation and first stage compression, the gas returns to the carbon dioxide removal section of the Rectisol Unit. After carbon dioxide removal in the Rectisol Unit the SNG is conveyed to the Product Gas Compression Unit. The SNG leaving the carbon dioxide section of the Rectisol Unit will contain approximately 95 percent volume methane.

AREA 1600 — PHENOSOLVAN UNIT

The process water from the Gas Liquor Separation Unit which is contaminated with phenols, ammonia, hydrogen sulfide and carbon dioxide is treated in the Phenosolvan Unit prior to use as make-up water in the process gas liquor cooling tower. Crude phenol and liquid anhydrous ammonia byproducts are produced.

The incoming process water is passed through gravel filters for removal of suspended matter and then through extractors, where an organic solvent is used to extract phenols. The organic solvent is distilled and separated from the phenol and recycled to the extractors for reuse. The crude phenol byproduct is recovered and transferred to storage for subsequent sale.

After removal of all traces of solvent, the dephenolized process water is stripped in the deacidifier to remove dissolved carbon dioxide and hydrogen sulfide, which is processed in the sulfur recovery unit. The resultant process water is distilled to recover 25 percent aqua ammonia, which is further distilled to produce commercially pure anhydrous ammonia. The remaining process water is then utilized as cooling water make-up to the process water cooling tower.

AREA 1700 — METHANATION

The methanation unit converts low Btu synthesis gas to methane-rich high Btu gas by the following exothermic reactions:

$$CO + 3H_2 = CH_4 + H_2O + 94{,}250 \text{ Btu per lb. mole } CH_4$$
$$CO_2 + 4H_2 = CH_4 + 2H_2O + 77{,}700 \text{ Btu per lb. mole } CH_4$$

Other minor reactions which take place are the hydrogenation of ethylene to ethane and hydrocracking of ethane to methane.

Feed gas entering the unit from the gas purification unit is heated and passed through a guard vessel containing zinc oxide for removal of trace sulfur compounds. The flow of the feed gas is proportioned between two catalytic reactor stages along with recycled methanated effluent gas, which serves to limit the temperature rise across the reactors. The reactors are designed as fixed bed downflow units employing a pelleted reduced nickel-type catalyst.

The reaction heat is removed by generation of 600 psig steam in waste heat exchangers at the outlet from each reactor.

Net gas leaving the synthesis loop is passed through the cleanup reactor to accomplish essentially complete conversion of carbon monoxide, and then it is cooled by successive heat exchange with boiler feed water, fresh feed gas, air and cooling water. Water condensed from the gas is separated and forwarded for recovery as boiler feed water. The net product is sent to the gas compression unit.

AREA 1800 — GAS LIQUOR SEPARATION

The gas liquor contains tar, tar oil, naphtha, and dissolved compounds such as phenols, ammonia, carbon dioxide, and hydrogen sulfide. Tar is defined as a heavier-than-water organic liquid phase, while tar oil is the lighter-than-water organic liquid phase.

The gas liquor separation is designed to clean up tarry and oily gas liquors by separating the incoming streams into tar, tar oil, recycled gas liquor, and clarified aqueous liquor streams. Flash gases released from the gas liquor by pressure reduction are scrubbed to remove ammonia.

The gas liquor streams originating from the gasification, shift conversion, and gas cooling units are cooled, combined, and reduced in pressure. The liquor flows to a large separator from which tar is removed. The tar is retained for export as fuel. The net liquor flow is forwarded to a second separator where tar oil is separated and removed. The gas liquor passes to a final separator where additional tar oil is removed and sent to storage along with tar oil from the second separator.

The aqueous stream from the final separator passes to intermediate tankage before being fed to the Phenosolvan unit.

AREA 1900 — PRODUCT GAS COMPRESSION

The product gas compression unit consists of two parallel centrifugal compressors driven by induction condensing steam turbines. Gas leaving the methanation unit is compressed in the first stage of the compressor and then cooled prior to undergoing final carbon dioxide removal in the second stage Rectisol unit.

The product gas is then compressed in the second compression stage to the required pipeline pressure.

AREA 2000 — COAL STORAGE AND PREPARATION

Run-of-mine coal is delivered to the plant. After crushing and screening the coal is delivered through a system of conveyors to the live storage pipe having a storage capacity of about six days production for the gasification plant. A dead storage pile is also provided with a capacity sufficient to ensure approximately 30 days production requirements. From the live storage pile, the coal is fed continuously through screens to achieve the correct size distribution for gasification. Coal fines not suitable for gasification are routed to storage silos and are available for sale.

AREA 3000 — OXYGEN PLANT

The oxygen facilities are designed to provide a nominal 5200 tons/d of gaseous oxygen to the process plant with an oxygen purity of 99.5 percent.

Three parallel process trains are utilized with each unit consisting of a turbine-driven axial/centrifugal air compressor, air separation section (cold box) and a turbine-driven centrifugal compressor. The air compressor turbines are of the extraction/condensing type utilizing 1500 psig, 900°F steam from battery limits. The total make-up steam required for the rest of the complex is extracted at 550 psig with the remainder going to the surface condensing type utilizing 550 psig steam.

Gasifier start-up air and plant general use air is provided from the air compressors. Excess nitrogen from the air separation system is used for the process plant inert gas system. Liquid nitrogen storage is provided to ensure availability of purge gas during a plant outage.

Oxygen plant control will be centralized and combined with the main plant control room.

AREA 4000 — SULFUR RECOVERY

The sulfur in the coal feed to the gasification plant is recovered using technologies based on the characteristics of each sulfur-containing stream. The design of the sulfur recovery system is based on coal containing an average of 0.71 percent sulfur on a DAF basis.

RECTISOL UNIT

The Rectisol unit produces two H_2S containing streams. One, relatively rich in H_2S, is sent to a Claus and IFP sulfur conversion plant. The tail gas from this system is sent to an over-the-fence Thermal Generating Station for complete incineration. The second stream from the Rectisol and Phenosolvan units, low in H_2S, is sent to the Stretford sulfur conversion plant. The Stretford offgas, consisting mainly of CO_2 with small amounts of ethylene, H_2S, COS, and traces of hydrocarbons, is also incinerated at the nearby Thermal Generating Station.

STRETFORD UNIT

The Stretford unit is used to recover elemental sulfur from H_2S present in the lean acid gas stream from the Rectisol unit. This process is licensed by the North West Gas Board of the British Gas Corporation. The process, using a dilute aqueous solution containing Na_2CO_3 sodium metavanadate and anthraquinone disulfonic acid (ADA) operates in a continuous regenerative fashion as follows:

H_2S is absorbed from the gas by the alkaline carbonate solution forming HS^- ions. This is accomplished in a countercurrent open grid tower. The sulfide is oxidized to free sulfur by vanadate according to the reaction:

$$HS^- + 2V^{5+} = 2V^{4+} + S + H^+$$

This reaction proceeds during the absorption step and is completed in a holding vessel. The solution is regenerated by reoxidation of V^{4+} to V^{5+}. This is accomplished by sparging with air in a separate vessel with ADA acting as a catalyst for the reaction. The sulfur forming as a floating froth is separated from the solution and is processed to produce a salable liquid sulfur by-product.

The Stretford process removes only sulfur in the form of H_2S. Other sulfur compounds such as COS, CS_2, and mercaptans are unaffected by the process. The sulfurous compounds are present only in small concentrations and are incinerated to SO_2 at the Thermal Generating Station. A small degree of oxidation of sulfides in solution to thiosulfate and sulfate occurs. These salts are nonregenerable and require a small liquid purge.

The unit is designed in two parallel trains, each with a nominal capacity of 50 percent. The regeneration air blowers are double train with 50 percent spare.

CLAUS UNIT

The Claus unit carries out the stoichiometric reaction of H_2S and SO_2 to produce sulfur according to the reaction:

$$2\,H_2S + SO_2 = 3\,S + 2H_2 + O_2 + \text{Heat}$$

The reaction proceeds catalytically in the vapor phase, using a bauxite-type catalyst. Sulfur is condensed and separated as a liquid from the effluent.

The process is licensed by Amoco.

The unit is designed as two 50 percent trains.

IFP UNIT

This unit continues the Claus reaction in the liquid phase at lower temperatures than the normal Claus unit, permitting a more favorable equilibrium for reaction of H_2S and SO_2. The process is specifically tailored to serve as a Claus unit clean-up.

The reaction is carried out in a circulating high boiling solvent in a packed tower. The solvent contains a catalyst for the reaction and is maintained at a temperature above the melting point of sulfur. The process is a proprietary development of the Institute Francais du Petrole (IFP) and is available through a number of licensees in North America.

Single train design is used, with a bypass provided to permit continuous operation. The tail gas from the IFP unit is incinerated at the nearby Thermal Generating Station.

AREA 5000 — STEAM DISTRIBUTION

Imported steam at 1500 psig and 900°F is let down through extraction condensing turbines driving the air compressors in the oxygen plant. The 550 psig steam for the air compressor turbines is combined with steam generated in the methanation waste heat boilers. This steam is supplied to the gasifiers and to turbine drivers fed from the 550 psig steam system. The oxygen, refrigeration, and lock gas compressor turbines are of the induction type and use 550 psig and 100 psig steam, discharging to condensers at 4" Hga.

Steam generated from waste heat in the gasification unit is fed to the 100 psig steam system. This steam is used for reboilers in the Phenosolvan, Rectisol and Stretford units, steam jet air ejectors, converted gas booster compressor, and methanation recycle compressors. These turbines all exhaust to condensers at 4" Hga. Excess steam from the 100 psig steam system is fed to the induction nozzles of the product gas compressor turbines.

Steam generated from the heat recovered in the shift conversion waste heat boilers and some of the gas cooling waste heat boilers is fed to the 60 psig steam system. This steam is utilized as heat input to reboilers in the Phenosolvan and Rectisol areas and for steam tracing and tank heaters. Pressure is maintained and make-up steam is provided to the header by a let-down station from the 100 psig header.

The remaining gas cooling waste heat boilers generate 20 psig steam to be used as heating steam for reboilers in the Phenosolvan and Rectisol units and for the plant deaerators. Pressure is maintained, and make-up steam is provided to the header by a let-down station from the 60 psig header.

Condensate, make-up water from battery limits, and 20 psig steam are mixed and deaerated in the M.P. deaerator. The M.P. boiler feed pumps provide feedwater to the gasifier water jackets and to the methanation waste heat boilers through process exchangers.

Make-up from battery limits, treated process condensate, blowdown water, and 20 psig steam are mixed and deaerated in the L.P. deaerator. The 100 psig boiler feed pumps take suction from the deaerator and pump the feedwater through process exchangers to the gasification waste heat boilers. The L.P. boiler feed pumps take suction from the deaerator and provide feedwater to the shift conversion and gas cooling waste heat boilers after preheating in process exchangers.

AREA 5300 — POWER DISTRIBUTION

A conventional 3-level distribution system for an estimated total requirement of 53,300 KW is included in the design.

AREA 5400 — RAW WATER SUPPLY & TREATING

Raw and Boiler Feedwater are supplied at the Gasification Plant battery limits. The sequence of water treating steps and the interaction with the waste effluent steps are similar to those shown on Sketch No. 7, Water and Waste Effluent Balance. The system was designed for maximum water reuse. Raw water is required for cooling tower and potable water make-up.

AREA 5500 — COOLING WATER

Three separate cooling towers are provided to handle the bulk of the plant heat rejection as follows:

- Cooling tower using treated fresh water as make-up
- Cooling tower using treated gas liquor (process water) from the Phenosolvan and Gas Liquor Stripping Unit as make-up.
- Separate cooling tower using treated fresh water for the oxygen plants in order to minimize the hazard associated with hydrocarbons entering the system.

Blowdown water from the process gas-liquor cooling tower system is sent to deep-well disposal.

Chemical feeding equipment will be provided to permit addition of water treating chemicals to all systems as required, in order to adjust pH and inhibit corrosion, scale formation and biological plant growth.

AREA 5600 — FIRE PROTECTION SYSTEM

The fire protection system consists of a fire water loop, chemical and foam fire suppression equipment and mobile equipment.

AREA 5700 — MISCELLANEOUS UTILITIES

Water for the potable and sanitary use is supplied from the raw water treatment plant where it is filtered, chlorinated and treated prior to distribution.

In addition to supplying potable water to the major plant buildings, the system will also supply the plant safety showers and eye baths.

INSTRUMENT AIR, PLANT AIR AND INERT GAS

A plant air system is provided to supply compressed air at a nominal 100 psig pressure to shops and service outlets throughout the plant.

Plant air is normally supplied from the oxygen plant main air compressors. During periods of total plant shutdown motor driven air compressors supply the plant air requirements. The system is complete with necessary aftercoolers, air receivers and distribution piping. All plant air will be dried.

An instrument air system is provided to supply clean, dry air for instrument operation.

Instrument air is supplied by oil-free motor driven compressors operating at a system pressure of 100 psig. The system is complete with dryers, air receivers, and distribution piping. To ensure maximum reliability during a power outage, these compressors are connected to the emergency power system.

An inert gas system is provided to supply dry nitrogen from the oxygen plant. In addition, a liquid nitrogen tank provided with vaporization facilities supplies nitrogen to the system during periods of oxygen plant outage.

COMMUNICATIONS

The plant is provided with two communication systems. An in-plant dial telephone system is installed and operated by the telephone company in space provided by the plant. This arrangement avoids capital expenditure and hiring of specialized maintenance skills. The system is automatically monitored against failure to the degree that it is acceptable for fire reporting and thereby avoids the operation of a separate fire alarm system. The system is arranged to allow outside communication from designated telephones.

Operating communications throughout the plant to roving personnel and vehicles is by radio. Radio paging interconnected with the telephone system is included for contacting non-operating personnel when they are away from their normal stations.

AREA 6000 — OFFSITE STORAGE AND LOADING FACILITIES

Storage facilities are included to store by-products produced in the plant. Storage for a number of raw materials for plant feed is also provided. A minimum of 15-days storage supply is provided for most by-products and raw materials except for the anhydrous ammonia storage tank where 30-days storage is provided. The anhydrous ammonia by-product is stored as a liquid at atmospheric pressure in a double wall insulated tank provided with a vapor recovery refrigeration system.

Liquid elemental sulfur produced in the sulfur recovery plant is pumped from a heated sulfur pit included in the sulfur recovery unit to the stockpile area.

AREA 8100 — LIQUID WASTE EFFLUENT SYSTEM

The liquid waste effluent treatment system is designed to maximize water reuse. The only discharge of wastewater from the plant is that going to deep-well disposal and the water associated with ash returned to the mine. The effluent treatment scheme is similar to that shown on the Water and Waste Effluent Balance Sketch, No.7.

The bulk of reusable water is derived from the gas liquor area. Most of the phenols are extracted and H_2S, CO_2 and NH_3 are stripped from this stream, the process condensate is used as make-up water for the process gas-liquor cooling tower. The cooling tower also serves as an oxidation unit for reduction of biological and chemical oxygen demand.

The blowdown from the process gas-liquor cooling tower is sent via filtration to deep-well disposal.

The cooling tower blowdown, together with equipment and area drains and rain water from paved areas, is stored in a pond serving as a surge for rainstorms and as a safeguard for contaminated cooling water in the event of a heat exchanger leak. The effluent from the pond is treated by gravity oil separation and subsequent flocculation/clarification.

Any separated slop oil is stored for disposal by incineration. The underflow from the flocculator/clarifier is sent to ash handling. The overflow from the clarifier, together with sanitary sewage treated in a biological treatment unit, is reused in the ash handling system.

AREA 8200 — ASH DISPOSAL

Ash discharged from the gasifiers is quenched and sluiced by water to screw classifiers. The classifier discharge drops onto conveyors and is transferred to an ash bin which is emptied into trucks for disposal in the mining area.

AREA 8300 — FLARE SYSTEM

The flare system is capable of flaring the total gas from the Methanation Unit in the event of failure of the product gas compressors. The flare system may also be employed to flare product gas during plant startup when gas quality is below the acceptable specifications.

The flare system collects all emergency and operating hydrocarbon vents and burns them at the top of a flare stack.

The self-supporting flare stack includes ignitor, flame front generators, molecular seal, and continuous burning pilots. Ladder and access platforms will be provided on the flare stack to facilitate maintenance.

6.0 BASE DATA AND ASSUMPTIONS

6.1 BASE DATA

The "Base Engineering and Cost Data" Issue No. 3, with Addendum 1 provided by B.C. Hydro, was used in the study.

It must be understood that this study is based on data received from various sources. As such, the accuracy of the various elements in the study varies, and it should not be construed that the numbers presented are definitive estimates of capital costs. There are many areas which require more definition by the Client, the contractor(s), and/or licensor(s). A typical analysis of the sensitivity of the gas cost to capital investment and coal and steam costs is presented in Figure 1 for the LURGI oxygen-blown 900 MW case.

6.2 ASSUMPTIONS

6.2.1

Assumptions for the process design bases and the financial analyses of the various alternatives are defined in the following sections.

6.2.2 LURGI LOW BTU FUEL GAS AND OTHER LURGI FACILITIES

The process concept of this alternative was done on the following basis:

1. Hat Creek Coal would gasify in the same manner as North Dakota Lignite. This assumption allowed Lummus to use information contained in the document filed by American Natural Gas with the Federal Power Commission of the United States Government and prorate the costs of the various units in the LURGI gasification plant directly on a capacity basis. While it is recognized that every coal is different, since Hat Creek coal is essentially a sub-bituminous/lignite-type coal, then the gas analysis and quantity of by-products should be fairly similar to those for American Natural Gas. However, it must be understood that an analysis by LURGI giving the expected composition and perhaps an actual test of the coal at Westfield or Sasol is required to support this assumption.

2. In prorating the cost of the gasification section of the plant, it was assumed that the capacity of the LURGI gasifier would be the same for Hat Creek coal as it is for North Dakota lignite. Because of the differences in ash content and moisture content, this factor must be confirmed by LURGI.

3. Another assumption was that the ash content would not have a significant affect on the operability of the LURGI gasifier. Hat Creek coal will be blended to a maximum 32 percent ash (normal quantity 25-26 percent); whereas, North Dakota lignite has an ash content of approximately 6 percent, and New Mexico sub-bituminous coal (El Paso) has an ash content of 20 percent. This assumption must be confirmed by LURGI.

4. *The design of the LURGI plant was based upon a coal with 25 percent ash.* With an ash content increase to 26 percent, no appreciable differences are expected. If the ash content is as high as 31 percent, the specified ash handling disposal facilities would not be able to handle an increase of about 20 percent in the quantity of ash. Since the installed cost of the ash handling and disposal facilities is a small percentage of the total installed cost (approximately 1.5 percent), a 20 percent increase in its cost would have very little affect on the economics of gasification.

5. Because the HHV of the Hat Creek coal is different from the HHV of North Dakota lignite, an adjustment was made in the coal feed rate to arrive at proportionally the same heat input/output as for North Dakota lignite. This heat balance must be confirmed by LURGI.

6. The financial assumptions are based on information contained in the "Base Engineering and Cost Data." Additional items to be considered are:

a. Price of fuel-type by-products is set at one half its value as feedstock, as stated in the B.C. Hydro Cost Data. If fuel oil is \$12/Bbl, oil by-products will have a value of \$6/Bbl. This is an arbitrary adjustment and should be verified by B.C. Hydro.

b. The contingencies for the various LURGI alternates are different. They are as follows:

1. Low Btu Fuel Gas	15 and 20 percent
2. Town Gas	15 percent
3. Pipeline Gas	10 percent

6.2.3 KOPPERS-TOTZEK

The basis of design of the KOPPERS-TOTZEK unit was provided by the Koppers Company of Pittsburgh, Pennsylvania for North Dakota lignite. Based upon an analysis of the number of gasifiers required, Koppers submitted an order of magnitude cost estimate for their systems, including engineering, equipment, materials, and construction. The Koppers estimate included the coal preparation and handling areas in the total. Lummus estimated the home office cost for the non-KOPPERS' units (8 percent as per *base engineering and cost data) and added these to Koppers' home office costs. For the* KOPPERS-TOTZEK cases, therefore, the engineering portion is about 12 percent of the direct costs, rather than 8 percent, as directed by B.C. Hydro.

7.0 ENVIRONMENTAL CONSIDERATIONS

The pollution control aspects that have been considered are related to essentially three services:

7.1 AIR EMISSION

In the cases where low Btu gas is produced, the cooled and precleaned raw gas is passed through a Stretford desulferization system. The emission of SO_2 resulting from the combustion of the low Btu gas will be compatible with the B.C. regulations.

For the SNG alternate, a Rectisol system will remove H_2S, CO_2, and small amounts of hydrocarbons from the raw gas. The Rectisol system's off-gas is sent to a Stretford system. The cleaned tail gas from the Stretford system will be incinerated.

Vent gases deriving from startups, emergency blowdowns, and temporary local pressure reliefs are sent via headers to an elevated flare system.

7.2 EFFLUENT WATER STREAMS

Intensive reuse of process and utility waste water will be applied within the coal gasification plant. For the LURGI process, the "gas liquor" (process condensate) will be treated to remove tars, oils, phenols, ammonia, and hydrogen sulfide. In the case of the KOPPERS-TOTZEK process, only stripping will be required. This treated "gas liquor" will be used as cooling water makeup. The blowdown from this cooling tower operation can be disposed of either by deep-well injection or reduced in volume by multiple-effect evaporation before disposal together with the ash. Waste heat steam boiler blowdowns, after flash steam recovery, will be reused as cooling tower make-up water. Regeneration and backwash streams from the water treating facilities will be disposed of by either deep-well injection (brines) or with the ash (thickened silt). The blowdown from the cooling tower for the air separation unit will and can be reused for humidification/cooling purposes.

Any of the above-indicated, final disposal methods will have to be ultimately verified. Feasibility studies concerning deep-well disposal will have to be made.

7.3 SOLIDS REMOVAL

Ash and other solid wastes will be sent to the mine. Certain aqueous waste streams, as indicated under 7.2 above, could be disposed of with the ash. Ash leachability studies and methods of preventing contamination of water in nearby wells should be considered.

7.4 SOCIO-ECONOMIC CONSIDERATIONS

During the construction of the plant, a work force of some 2000-3000 people, not including support facilities, will be employed. The socio-economic and environmental impact of these people and the related housing, transportation, sanitary, recreational facilities, etc., must be taken into account.

8.0 OTHER PROCESSES

Technical and economic analyses of various "second-generation" processes are included in this section.

The processes examined are:

8.1 SYNTHANE
8.2 CLEAN FUELS FROM COAL
8.3 COGAS

A summary of the TEXACO process will be submitted following receipt of information from the Texaco Development Corporation.

8.1 SYNTHANE

This section of the report covers the evaluation of the SYNTHANE process to produce a medium Btu fuel gas from coal.

SYNTHANE is a second-generation coal gasification process being developed by the Energy Research and Development Administration of the U.S. Government. There are no commercial units in operation or construction; however a 5 T/hr (ultimate coal feed) pilot plant, designed by Lummus, has been constructed and is expected to begin operations in early 1976. Lummus has the operating contract for this faciltiy.

The SYNTHANE process is a high-pressure fluidized bed process using steam and oxygen as fluidizing/gasification agents. The process was originally considered for generating pipeline quality synthetic natural gas at 1000 psi. The high pressure reduces or eliminates the gas compression requirements, increases the equilibrium concentration of methane in the gasifier effluent, and should result in savings in the gasification and methanation areas. By eliminating shift conversion, CO_2 removal, and methanation, a medium Btu fuel gas (HHV = 375 Btu/SCF) can be manufactured. A simplified schematic block flow diagram is shown in the attached sketch.

The SYNTHANE process has a higher Thermal Efficiency than the LURGI process, as can be seen in the attached table. A specific feature of the SYNTHANE process is the discharge of char containing up to 30 percent of the carbon in the coal fed to the gasifier. In addition, the process also produces by-products such as tars, oils, and phenols. Char is a mixture of ungasified carbon and ash. It may be possible to use these by-products as fuel for steam generation or to sell them.

Since many of the process areas in a SYNTHANE plant are similar to those in a LURGI plant, we have attempted to qualitatively estimate the capital requirements for a SYNTHANE plant producing 230 x 10^9Btu/D of medium Btu fuel gas. A more detailed analysis was not possible for two reasons:

1. The study ground rules did not provide for sufficient time to analyze the SYNTHANE process in depth.

2. No reliable information on the capital requirements of a commercial-scale SYNTHANE plant has been developed to date.

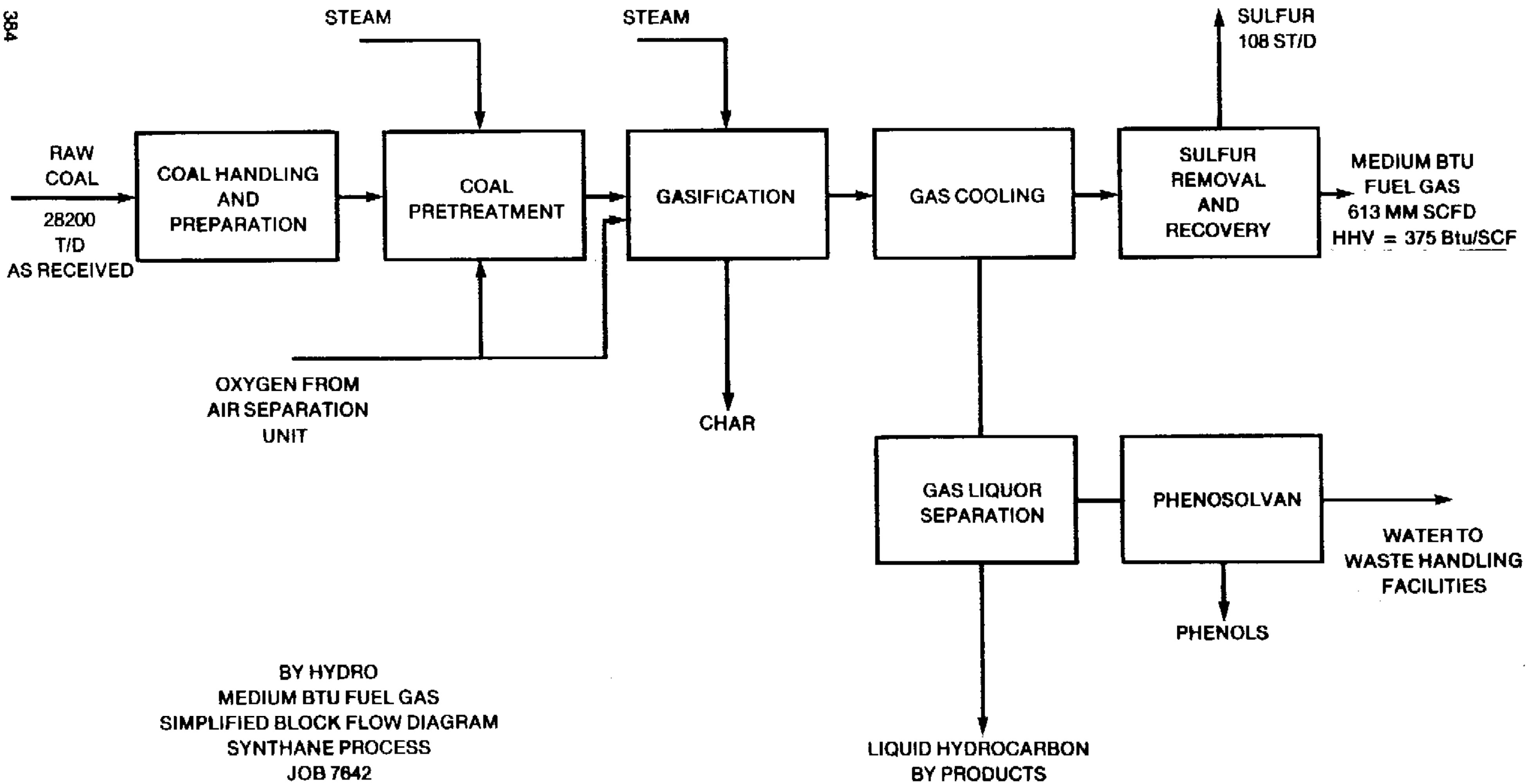
Based upon the assumption of similarity in many of the process areas, we estimated the capital requirements of a SYNTHANE plant, with a capacity as noted above, to be about $450 MM. This, combined with an analysis of the utilities requirements, produces a cost of service in the area of $1.20 to $1.25 per MM Btu. These numbers indicate that SYNTHANE is just competitive with LURGI.

It should be understood that one reason for the slightly higher cost of service for SYNTHANE is the higher quantity of coal required because of the high char quantity produced. If the char is recirculated, a higher percentage of the carbon in the reactor is gasified. This would reduce the raw coal requirements substantially, and the cost of service would also decrease, since it is sensitive to the coal requirements.

The SYNTHANE process appears to be economically and technically competitive with the LURGI process. For the same Btu production, SYNTHANE offers a lower volume of gas at a potentially higher pressure. The SYNTHANE reactors are less complicated mechanically, meaning greater reliability and possible elimination of the requirement for spare reactors. Based on a preliminary analysis of data generated by the U.S. Bureau of Mines, there also may be slightly less liquid products from the SYNTHANE process.

SYNTHANE OVERALL THERMAL EFFICIENCY

	ST/D	10^9 BTU/D	HHV%
INPUT			
Coal (6402 Btu/Lb)	28157	360.52	83.8
Steam	22212	63.48	14.8
Power	47535 KW	3.89	0.9
Fuel for Coal Drying		2.25	0.5
		430.14	100.0
OUTPUT			
Medium Btu Fuel Gas 375 Btu/SCF	613 MM SCFD	230.00	53.5
Liquid Hydrocarbons	546	14.99	3.5
Char (4616 Btu/Lb)	10287	94.97	22.2
L.P. Steam Export	3600	8.48	2.0
		348.44	81.1



BY HYDRO
MEDIUM BTU FUEL GAS
SIMPLIFIED BLOCK FLOW DIAGRAM
SYNTHANE PROCESS
JOB 7642

8.2 CLEAN FUEL FROM COAL

As part of the study investigating the production of Thermal Power Plant Fuels, Lummus has investigated the use of its CLEAN FUEL FROM COAL process to produce a synthetic heavy fuel oil from Hat Creek coal.

PROCESS DESCRIPTION

The key features of this process are:

1. Catalytic hydrodesulfurization of coal to produce a refined liquid coal containing 0.5 wt.% S or less.

2. Special ash separation technique to produce a Clean Fuel product containing less than 0.1 wt.% ash.

A simplified schematic block flow diagram of the C-E Lummus process is shown in the attached sketch.

After crushing and drying, coal is slurried and partially digested in the presence of an aromatic recycled solvent. The resulting coal paste slurry is sent to the hydro-desulfurization section, where reaction with hydrogen in the presence of a commercially available catalyst at elevated pressure and temperature results in liquefaction and desulfurization of the coal. Clean Fuel product sulfur levels of 0.3 wt.% can be readily achieved even with coals having sulfur contents as high as 3-4 wt.%.

Hydrogen sulfide formed in the reaction step is purged and absorbed from the reaction products and converted to elemental sulfur via conventional sulfur recovery techniques.

As a consequence of desulfurization, other constituents of the coal — viz., nitrogen, and oxygen — are also partially removed. However, pilot plant data indicated that the degree of denitrification obtained using commercially available catalyst at practical hydrogen consumptions was not sufficient to result in a fuel that would meet NOX emission standards when burned in a conventional unmodified boiler. Recognizing the need for a low nitrogen product, C-E Lummus developed a new proprietary catalyst that exhibits both high desulfurization and denitrification capabilities. In tests conducted thus far, product nitrogen contents have been reduced to levels of less than 0.5 wt.%, as compared to 1% or more for standard high activity catalysts under similar reaction conditions.

As illustrated in the attached table, the yield of liquids from Hat Creek coal is rather low, primarily because of the high ash and moisture content of the feedstock. An overall thermal efficiency of 55.4 percent was estimated for the liquefaction complex. A factor contributing to the relatively low thermal efficiency is the high hydrogen consumption required for this particular coal. A major factor in hydrogen uptake for younger coals is their oxygen content. Coals with high oxygen content need more hydrogen, since the oxygen is removed primarily as water.

On a very preliminary basis, the total installed cost of the plant to produce 230 x 10^9 Btu/D (HHV) of liquid fuel is about $500 MM for mid 1975. Using the economic ground rules provided by B.C. Hydro, a cost of service calculation indicates that a synthetic heavy fuel oil can be produced from Hat Creek coal for $1.80/MM Btu (HHV).

While it is apparent that for Hat Creek coal liquefaction, a higher unit product energy cost results in comparison to low Btu gasification, it must be pointed out that the liquid product is quite storable and thus uncouples the power plant from the conversion plant. Low Btu gas schemes do not offer this flexibility.

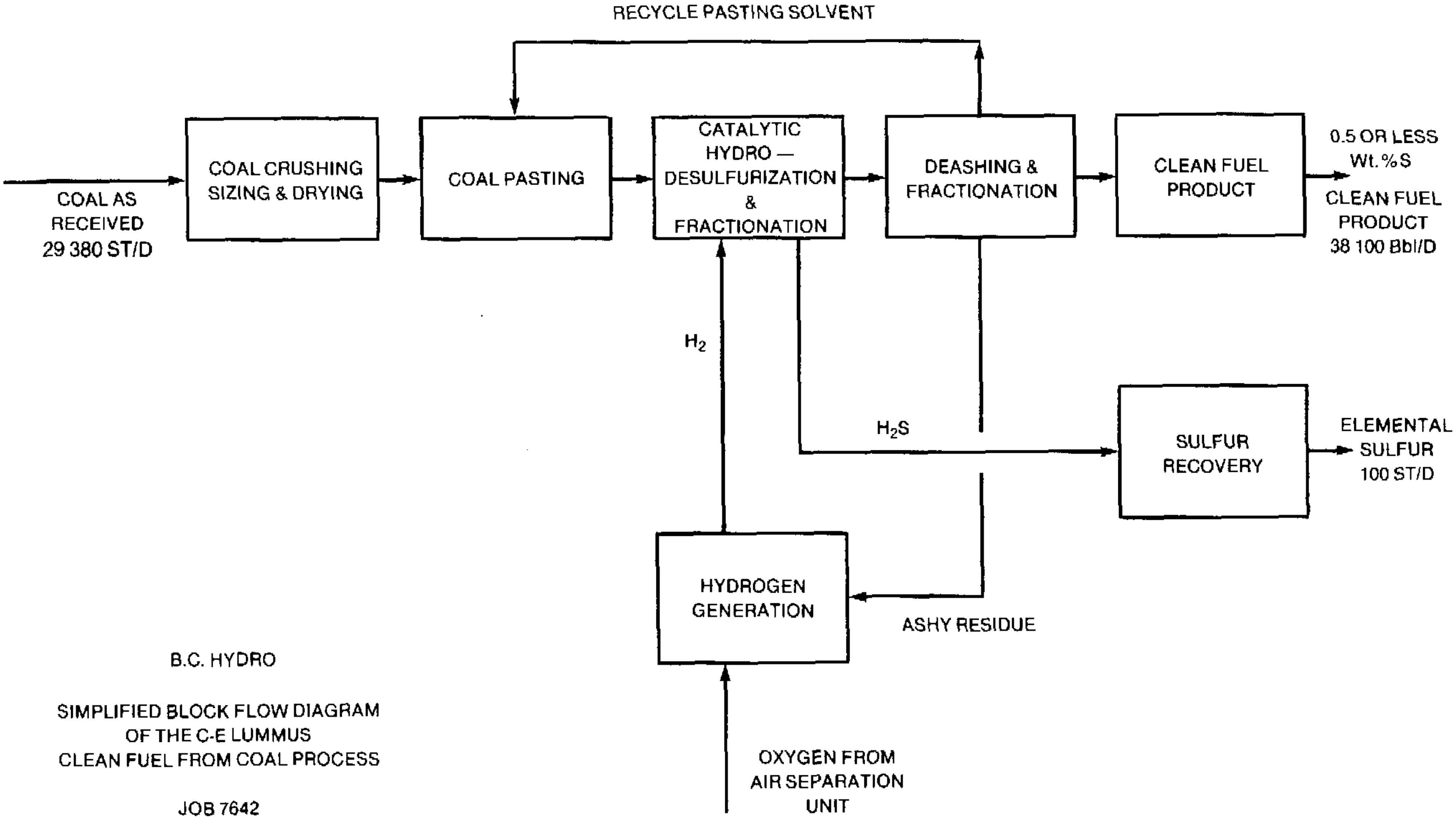
It should be carefully noted that the liquefaction yield estimate for Hat Creek coal is based on adjustments to data derived from Lummus Pilot Plant operations on a lignitic coal or similar ultimate analysis (maf basis). In establishing the yields obtainable from a given coal (in the absence of actual pilot plant testing), a petrographic analysis is needed. The ultimate analysis is insufficient. In the absence of a petrographic analysis of Hat Creek coal, the yields reported should be considered as preliminary.

CLEAN FUELS FROM COAL

OVERALL PRODUCTION AND THERMAL EFFICIENCY

INPUT	ST/D	10^9 Btu/D	HHV %
Coal			
To Liquefaction	20470	262.1	60.3
To Gasification for H_2	8910	114.1	26.3
Steam	19200	54.9	12.6
Power, 43000 KW		3.5	0.8
		434.6	100.0

OUTPUT	ST/D	10^9 Btu/D	HHV %
CFFC Blended Liquids C_{5+} @ 18300 Btu/lb (HHV)	6270	230	52.9
Excess Fuel Gas		11	2.5
		241	55.4



B.C. HYDRO

SIMPLIFIED BLOCK FLOW DIAGRAM
OF THE C-E LUMMUS
CLEAN FUEL FROM COAL PROCESS

JOB 7642

8.3 COGAS

This section of the report covers the evaluation of the COGAS process to produce a medium Btu Fuel Gas from coal.

The COGAS process is a second-generation coal gasification process being offered by the COGAS Development Company. It is, at present, in the pilot plant stage. The technology of the COGAS process is an outgrowth of the FMC COED process, which sought to make liquid hydrocarbon products from coal.

Although no commercial units have been built, the COGAS Development Company has undertaken economic evaluations of commercial-size units to produce both pipeline-quality gas and medium Btu fuel gas.

The economics discussed in this report are based on the COGAS evaluation of its process to manufacture SNG from Illinois No. 6 coal and Glen Harold (North Dakota) lignite coal and medium Btu Fuel Gas from Illinois No. 6 coal. The coals have analyses as follows:

ILLINOIS NO. 6

Proximate Analysis	AS RECEIVED wt.%	DRY wt.%	MAF wt.%
Moisture	10.0	-	-
Volatile Matter	32.85	36.5	41.5
Fixed Carbon	46.35	51.5	58.5
Ash	10.8	12.0	-
	100.0	100.0	100.0

Ultimate Analysis	DRY wt.%	MAF wt.%
C	69.0	78.4
H	5.0	5.7
N	1.3	1.5
S	4.0	4.5
O	8.7	9.9
Ash	12.0	-
	100.0	100.0
HHV Btu/lb	12600	14300

GLEN HAROLD LIGNITE (NORTH DAKOTA)

Proximate Analysis	AS RECEIVED wt.%	DRY wt.%	MAF wt.%
Moisture	28.4	-	-
Volatile Matter	33.4	46.6	50.5
Fixed Carbon	32.7	45.7	49.5
Ash	5.5	7.7	-
	100.0	100.0	100.0

Ultimate Analysis	DRY wt.%	MAF wt.%
C	59.9	64.9
H	4.6	5.0
N	0.86	0.9
S	0.64	0.7
O	26.3	28.5
Ash	7.7	-
	100.0	100.0
HHV Btu/lb	10769	11693

The COGAS Development Company has estimated capital and operating requirements for a medium Btu fuel gas plant with Illinois No. 6 coal as feed. We have made adjustments in these numbers using information supplied by COGAS to relate the costs to a plant assuming lignitic coal and the B.C. Hydro financial conditions. On this basis, the total cost of facilities, including interest during construction, for a plant producing 230 x 10^9 Btu/D of gas (HHV approximately 340 Btu/SCF) would be about $404 MM. This is comparable to the cost of the LURGI process for the same capacity. If we assume that COGAS' evaluation of the utilities requirements for the plant is correct, then the cost of service is $1.17/MM Btu.

The relatively high cost of service is due to two factors: more coal is required than in LURGI, and by-product credit is essentially unchanged. With a bituminous coal, on the other hand, the COGAS cost of service is about $0.90/MM Btu.

PROCESS DESCRIPTION

The process involves pyrolysis of the coal and gasification of the char in fluidized bed reactors, followed by scrubbing, oil recovery, purification, and compression.

Raw coal is pulverized, dried, and fed to the pyrolysis reactors, where volatiles are driven off using recycle gas from the gasification section as both a fluidizing and a heating medium. The pyrolysis gas is scrubbed with water to recover the heavier hydrocarbons (raw oil) and is then compressed, passed through the CO_2/H_2S removal section, and then blended into the make gas from the gasification section.

The char from the pyrolysis section is fed to the gasification section, where it is reacted with steam to form a make gas containing primarily hydrogen and carbon monoxide, with some methane and nitrogen. The heat for the reaction is obtained by burning char fines with air and using hot flue gas to heat and recirculate char. The flue gas is separated from the char prior to recharging the char into the gasifier. The flue gases are sent through a power recovery turbine prior to treating.

The raw make gas is then passed through a CO_2/H_2S removal unit and compressed to the required pressure. The light hydrocarbons remaining in the purified pyrolysis gas are blended into the synthesis gas, producing a fuel gas with a gross heating value of approximately 340 Btu/SCF.

Quantities of a raw oil are also produced as a by-product. This raw oil may be used as fuel in the Thermal Generating Plant or exported.

A simplified block flow diagram is attached.

A thermal efficiency for a COGAS plant producing medium Btu Fuel Gas from lignitic-type coal was estimated to be in the range of 60 to 65 percent. This is much lower than the comparable LURGI process and also much lower than the estimated COGAS efficiency using bituminous coal (Illinois No. 6) as feed. This efficiency is estimated to be about 78 percent.

The major reason for the much lower efficiency with lignitic coal is the smaller quantities of liquid hydrocarbons produced. COGAS has informed us that the raw oil produced from a lignitic coal is about 20 percent of the raw oil from a bituminous coal. The raw oil by-product produced by COGAS with lignitic coal is about the same as is produced by a comparable LURGI process. This affects the overall efficiencies and the economics significantly.

Although COGAS does not appear to be economically attractive for use with a lignitic coal, the process does have some advantages over the LURGI process:

1. No oxygen plant is required.

2. Since COGAS operates at lower pressures, better reliability and operability are expected, leading to increased on-stream time and lower maintenance costs.

3. Estimated thermal efficiencies and carbon yields are better for bituminous coals.

4. The lower severity of process conditions is more favorable to the COGAS process.

5. Coal feed and ash removal are simpler.

6. Construction of COGAS plants should be less affected by bottlenecks in delivery of high-pressure equipment, particularly reactors.

7. Fluidized bed pyrolysis of a wide variety of caking and non-caking coals has already been demonstrated.

8. The LURGI process is limited in the size range of coal it can handle. More coal must be mined for plant feed.

Disadvantages of the COGAS process are as follows:

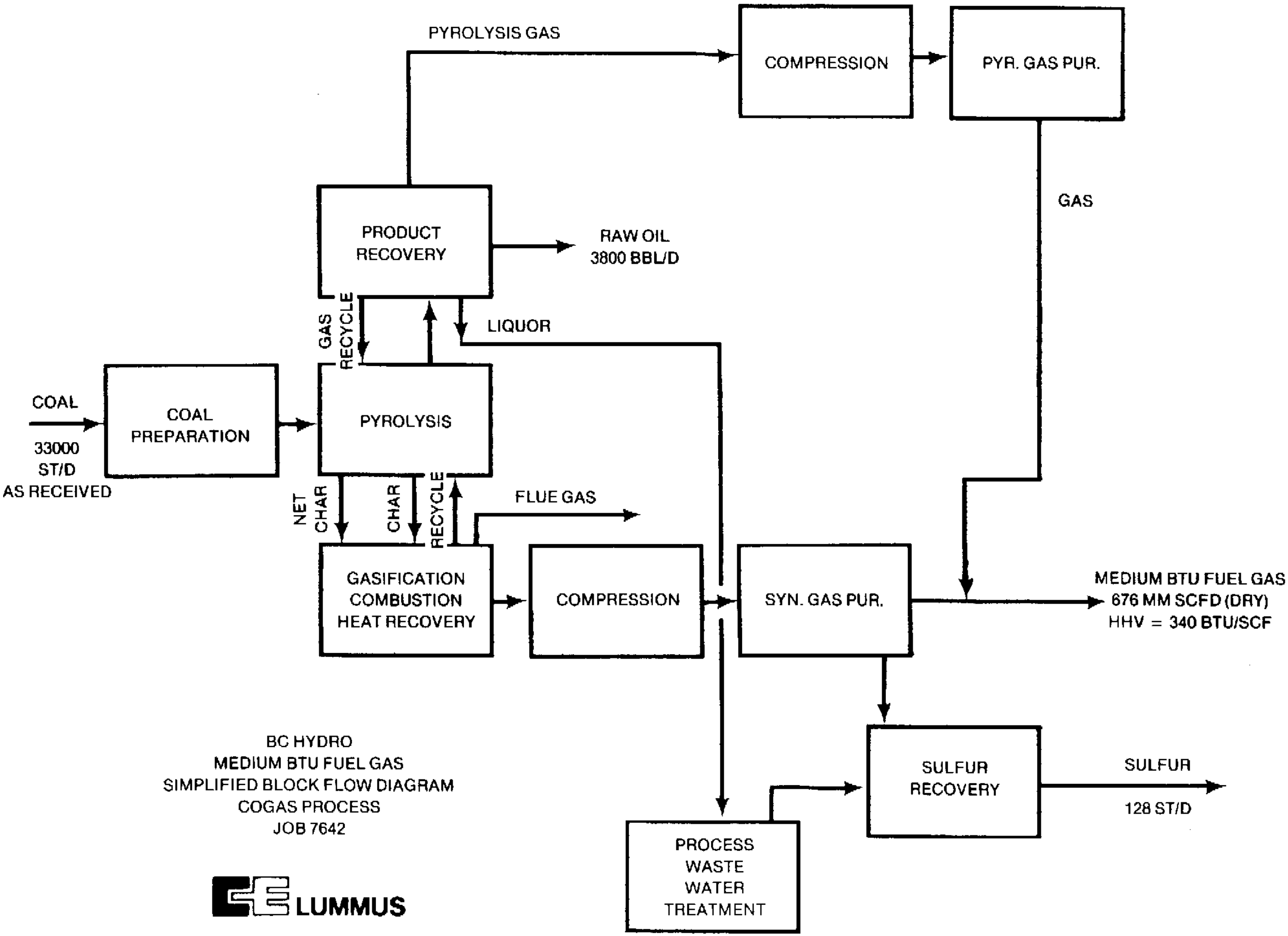
1. The solid carrier recirculation, stripping, and control of COGAS does not exist in the LURGI process.

2. The need for flue gas clean-up of particulates and sulfur dioxide does not exist for LURGI if steam generation is accomplished by firing low Btu desulfurized fuel gas.

3. The low-pressure, raw gas product may require compression, depending on the end use.

4. No commercial experience.

The COGAS process, using lignitic coal as a feedstock, offers no significant advantages over the LURGI process. If a higher rank coal should be considered, the COGAS process does show significant benefits; and in this case, further evaluation would be warranted.



BC HYDRO
MEDIUM BTU FUEL GAS
SIMPLIFIED BLOCK FLOW DIAGRAM
COGAS PROCESS
JOB 7642

LUMMUS

9.0 APPENDIX — INDEX

TABLES

SKETCHES

FIGURES

TABLE NO. 1

AVERAGE COAL PROPERTIES

PROXIMATE ANALYSIS	WEIGHT %
MOISTURE	20
ASH	25
FIXED CARBON	21.1
COMBUSTIBLE VOLATILES	27.2
INCOMBUSTIBLE VOLATILES	6.7
HEATING VALUE	
HHV	6402 Btu/LB

TABLE NO.2

COST OF FACILITIES THOUSANDS OF DOLLARS

SECTION	TITLE	LURGI OXYGEN BLOWN CASE A	LURGI OXYGEN BLOWN CASE B	LURGI AIR BLOWN HOT POT. CARB. CASE B	LURGI AIR BLOWN STRETFORD BASE B[1]	KOPPERS TOTZEK CASE A	KOPPERS TOTZEK CASE B	LURGI PROCESSES TOWN GAS	LURGI PROCESSES PIPELINE GAS
	Case A = 450 x 10^9 Btu/day Case B = 230 x 10^9 Btu/day SM	1498 MM SCFD 300 Btu/SCF	765 MM SCFD 300 Btu/SCF	1090 MM SCFD 211 Btu/SCF	1198 MM SCFD 192 Btu/SCF	1542 MM SCFD 292 Btu/SCF	792 MM SCFD 292 Btu/SCF	250 MM SCFD 280 Btu/SCF	250 MM SCFD 970 Btu/SCF
1100	Gasification	124,663	69,217	83,532	93,443	180,000	98,820	17,281	90,009
1200	Shift Conversion	---	---	---	---	---	---	7,538	12,321
1300	Gas Cooling	10,057	6,216	8,671	9,907	included in Area 1100	included in Area 1100	1,999	11,297
1400-1500	Rectisol	---	---	---	---	---	---	---	63,724
1600	Phenosolvan	46,973	26,693	---	14,507	---	---	8,851	33,609
1700	Methanation	---	---	---	---	---	---	---	33,692
1800	Gas Liquor Separation	14,991	8,519	3,915	8,252	---	---	2,824	10,727
1900	Product Compression	---	---	---	---	27,765	13,594	---	9,709
2000	Coal Preparation & Handling	52,042	32,516	31,987	34,547	included in Area 1100	included in Area 1100	14,289	41,183
3000	Oxygen	79,795	39,898	---	---	223,231	109,189	19,949	59,846
4000	Sulfur Recovery	25,831	13,377	6,898	13,099	20,804	13,038	4,527	13,094
5000	Steam Distribution	12,913	8,110	6,457	6,076	7,469	5,776	2,995	8,348
5200	Stack Gas Scrubbing	---	---	---	---	---	---	---	---
5300	Power Distribution	6,266	3,895	885	983	4,740	2,865	2,808	7,827
5400	Raw Water Supply & Treating	3,706	2,285	6,466	6,955	26,777	17,180	1,568	7,151
5500	Cooling Water	10,864	6,745	5,556	6,002	15,879	10,231	3,956	10,786
5600	Firewater	753	753	541	590	753	394	463	753
5700	Misc. Utilities	946	946	688	738	946	492	583	946
6000	Offsite Storage & Fuel	1,371	808	808	808	590	344	352	1,294
7000	Plant Interconnecting Pipe	7,200	4,100	3,900	4,400	8,500	4,900	1,600	6,900
8100	Liquid Waste Effluent	6,562	4,109	1,770	1,968	8,421	5,614	1,696	4,772
8200	Ash Disposal & Sanitary	7,277	4,551	4,294	4,294	7,438	4,652	2,362	5,853
8300	Flare	7,434	4,639	7,854	7,591	8,751	5,693	917	2,636
8400	Product Gas Expansion	12,676	8,349	21,051	19,884	---	---	---	---
8500	Air Compression	---	---	25,132	26,908	---	---	---	---
	Sub Total (ex engineering)	432,320	245,726	220,405	260,952	542,064	292,782	96,558	437,107
	Engineering Services	37,666	21,409	19,207	22,692	62,432	36,823	8,396	38,010
	Total Direct Cost	469,986	267,135	239,612	283,544	604,496	329,605	104,954	475,117
Corporate Overhead	@5%	23,499	13,357	11,981	14,177	30,225	16,480	5,248	23,756
Location Bonus	@6%	28,199	16,028	14,377	17,013	36,270	19,776	6,297	28,507
Buildings, Land, Equipment	@1.0%	4,700	2,671	2,396	2,835	6,045	3,296	1,050	4,751
Licensor Feeds and Unallocated Costs Costs	@5.2%	24,439	13,891	12,460	14,744	15,717	8,570	5,458	24,706
	Total Capital Cost	550,823	313,082	280,826	332,313	692,753	377,727	123,007	556,837
Startup & Training	@3.8%	20,931	11,897	10,671	12,628	26,325	14,353	4,674	21,159
Contingency	% varies	82,624	46,962	56,165	49,847	69,275	37,772	18,451	55,684
	Total Capital Expenditure	654,378	371,941	347,662	394,788	788,353	430,852	146,132	633,680
	Interest During Construction	154,278	71,296	66,640	75,671	218,799	89,604	20,272	149,449
	Total Cost of Facilities	808,656	443,237	414,302	470,459	1007,152	520,456	166,404	783,129

TABLE NO. 3

COST OF SERVICE (1)

	LURGI OXYGEN BLOWN		LURGI AIR BLOWN		KOPPERS TOTZEK		LURGI PROCESSES	
	CASE A	CASE B	CASE B	CASE B[1]	CASE A	CASE B	TOWN GAS	PIPELINE GAS
Raw Materials and Utilities $M/YR								
Coal @ $3/T	40,504	20,584	19,588	21,912	44,488	22,908	6,308	27,888
Steam @ $1/1000 lb.	26,560	13,944	8,632	9,296	14,276	7,968	5,312	14,608
Raw Water @ $22/1000 Gal.	108	55	214	272	1,651	915	31	366
Power @ 10 mills/Kwh	2,988	1,660	(501)	122	1,992	996	996	4,316
Chemicals	1,470	750	460	900	1,870	960	330	3,330
Subtotal	71,630	36,993	28,393	32,501	64,277	33,747	12,977	50,508
Plant Operations & Maintenance @1.45%	11,726	6,427	6,007	6,822	14,604	7,547	2,413	11,355
Variable Maintenance 0.3 mill/kwh	4,781	2,151	2,151	2,151	4,781	2,151	654	2,668
Administration & General @0.3625%	2,931	1,607	1,502	1,705	3,651	1,887	603	2,839
Insurance @0.25%	2,022	1,108	1,036	1,176	2,518	1,301	416	1,958
Interim Replacement @0.35%	2,830	1,551	1,450	1,647	3,525	1,822	582	2,741
Other Taxes @1.0%	8,087	4,432	4,143	4,705	10.072	5,205	1,664	7,831
Depreciation Expense @1.75%	14,151	7,757	7,250	8,233	17,625	9,108	2,912	13,705
Interest Expense @10%	80,866	44,324	41,430	47,046	100,715	52,046	16,640	78,313
Subtotal	127,394	69,357	64,969	73,485	159,491	81,067	25,884	121,410
Byproduct Credit								
Tar @ $6/Bbl	(4,648)	(2,324)	(3,652)	(2,139)	---	---	(664)	(3,652)
Tar Oil @ $6/Bbl	(8,300)	(4,316)	---	(4,634)	---	---	(1,328)	(6,308)
Naphtha @ $6/Bbl	---	---	---	---	---	---	---	(1,992)
Ammonia @ $180/ST	(18,924)	(9,628)	---	(8,300)	---	---	(2,656)	(12,616)
Phenols @ $6/Bbl	(1,992)	(996)	---	(920)	---	---	(332)	(1,328)
Subtotal	(33,864)	(17,264)	(3,652)	(15,993)	---	---	(4,980)	(25,896)
Total Cost of Service	165,160	89,086	89,710	89,993	223,768	114,814	33,881	146,022
Annual Production 10^{12} Btu/YT	149.40	76.36	76.36	76.36	149.40	76.36	23.24	80.51
Final Cost of Service $/MM/Btu	1.11	1.17	1.18	1.50	1.50	1.17	1.45	1.81

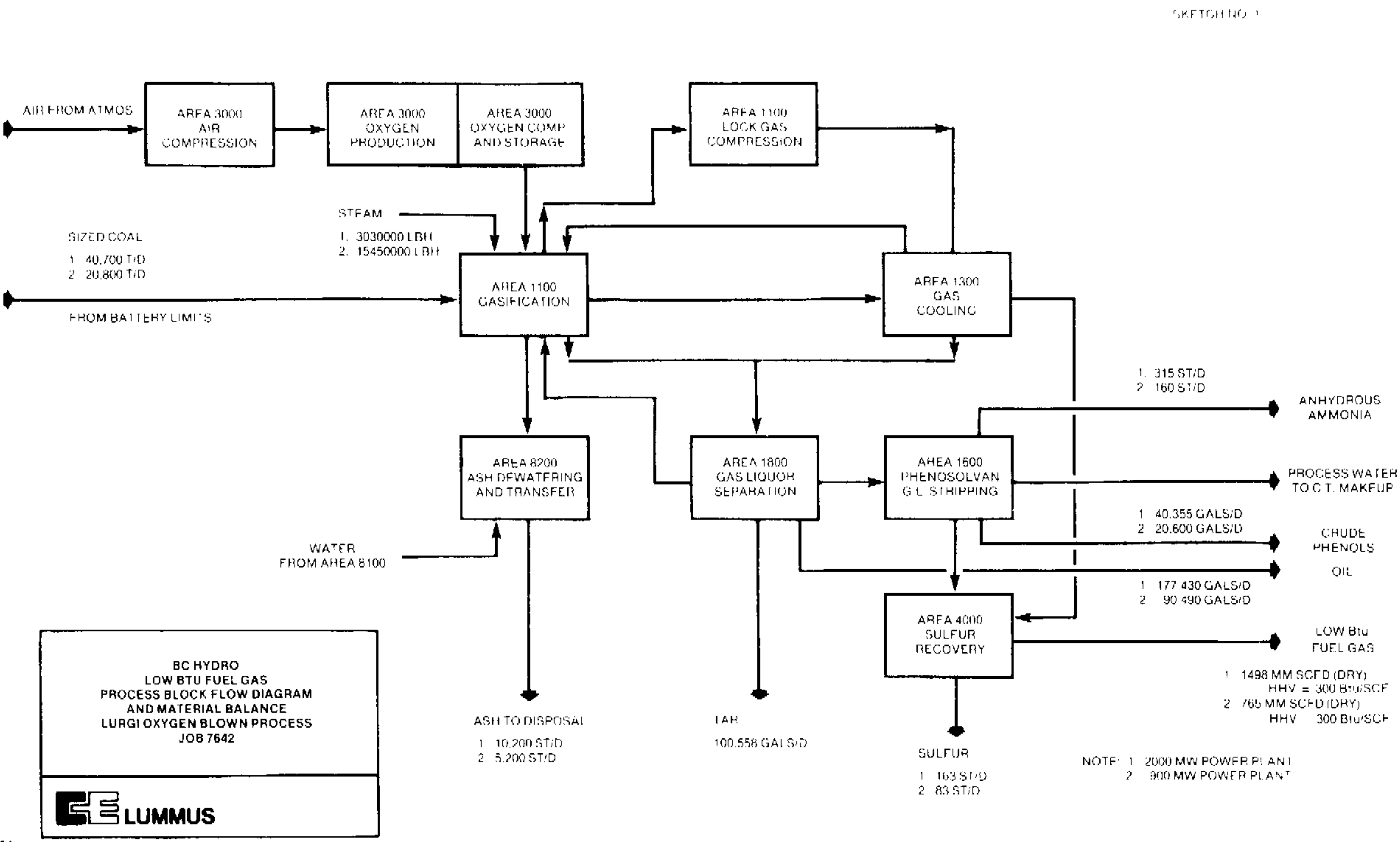
NOTES:
(1) Basis: 332 operating days/year
(2) Parentheses denotes credit

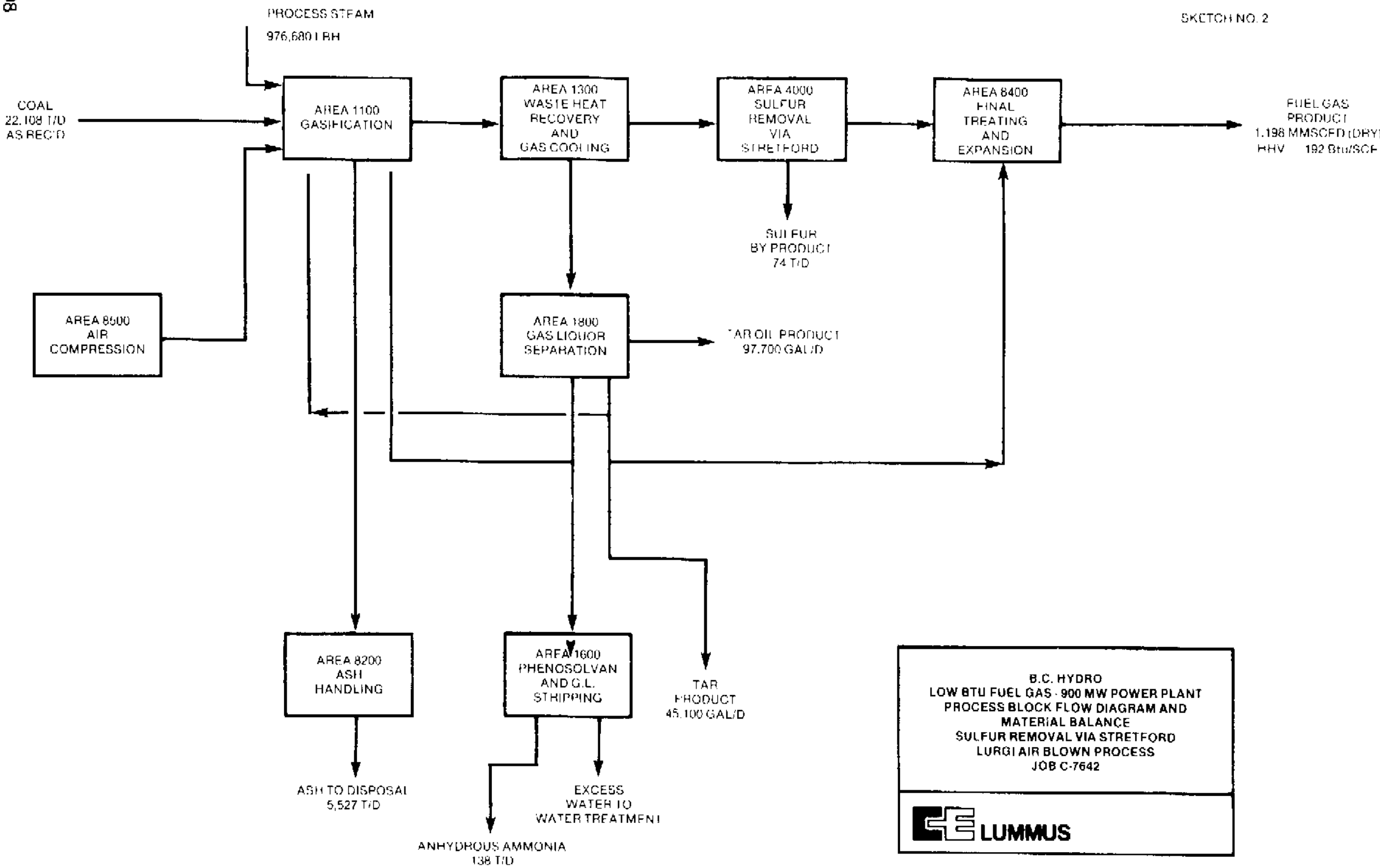
TABLE NO. 4

SUMMARY OF OPERATING REQUIREMENTS

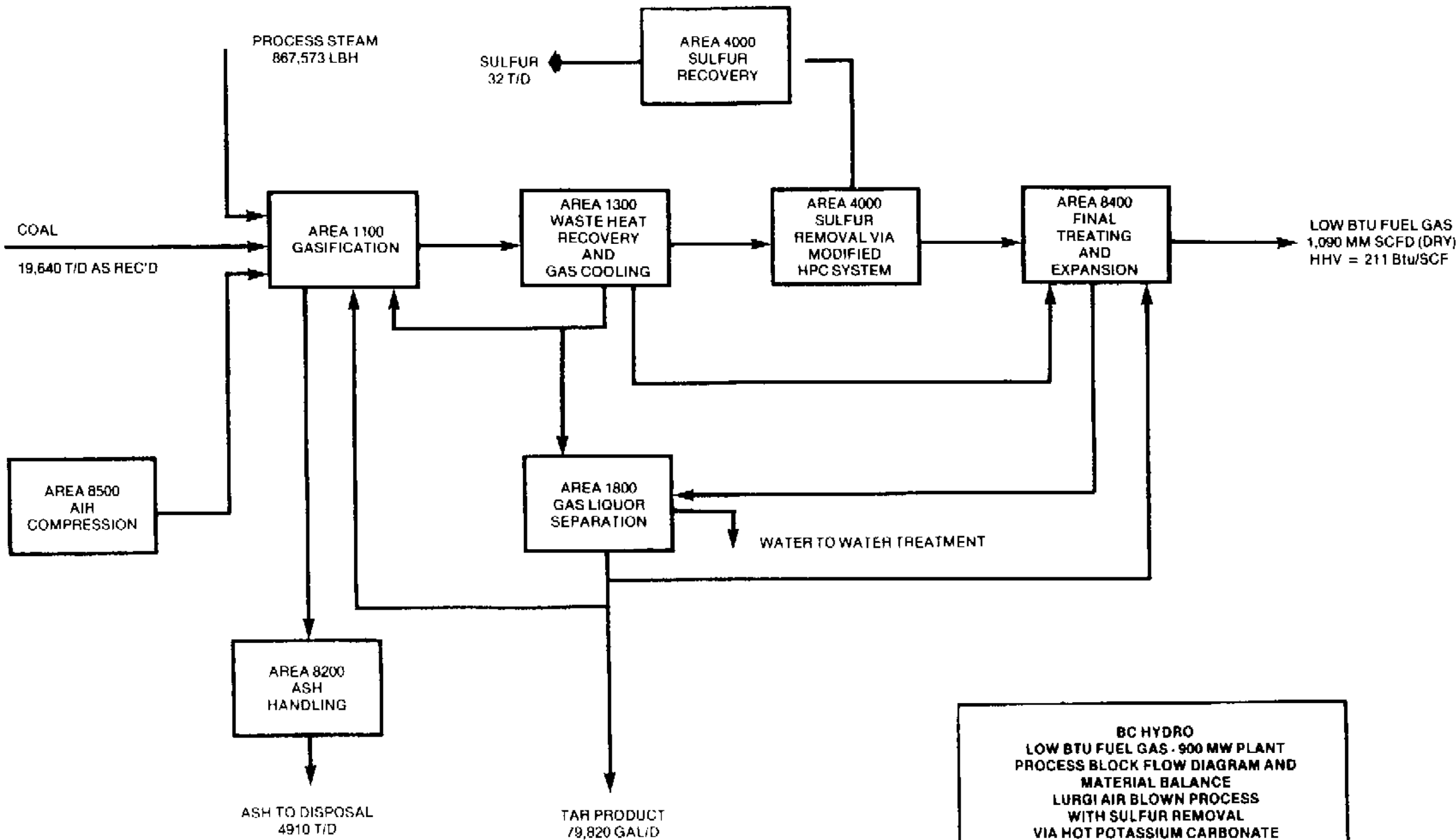
	LURGI OXYGEN BLOWN		LURGI AIR BLOWN		KOPPERS TOTZEK		LURGI PROCESSES	
	CASE A	CASE B	HOT POT. CARB. CASE B	STRETFORD CASE B[1]	CASE A	CASE B	TOWN GAS	PIPELINE GAS
Raw Materials, Utilities								
Coal, ST/D As Rec'd.	40,714	20,805	19,638	22,108	44,590	22,893	6,460	27,875
Steam LB/HR High Press.	3,460,600	1,768,400	1,094,100	1,170,400	1,801,00	1,000,000	679,000	1,821,000
Low Press.	(121,600)	(62,100)	(418,281)	(468,300)	(237,000)	(121,000)	(25,000)	
Condensate, US gpm	-	-	-	-				
Raw Water Makeup US gpm	1,430	735	2,035	2,580	15,700	8,700	291	3,486
Power, KW	39,000	20,000	(6,300)	1,530	26,000	13,000	12,350	53,300
By Products								
Tar, US GAL/D	(100,558)	(51,285)	(79,820)	(45,100)			(15,095)	(79,488)
Tar Oil, US GAL/D	(177,426)	(90,487)		(97,700)			(26,624)	(130,680)
Phenols US GAL/D	(40,355)	(20,581)		(19,400)			(6,059)	(26,706)
Naphtha US GAL/D	-							(43,924)
Ammonia, ST/D	(315)	(161)		(138)			(47)	(209)
Sulfur ST/D	(163)	(83)	(25)	(85)	(164)	(86)	(27)	(89)

Notes (1) (-) Denotes Export


SKETCH NO. 1
AIR FROM ATMOS
AREA 3000 AIR COMPRESSION
AREA 3000 OXYGEN PRODUCTION
AREA 3000 OXYGEN COMP AND STORAGE
AREA 1100 LOCK GAS COMPRESSION
STEAM
1. 3030000 LBH
2. 15450000 LBH
SIZED COAL
1 40,700 T/D
2 20,800 T/D
FROM BATTERY LIMITS
AREA 1100 GASIFICATION
AREA 1300 GAS COOLING
1. 315 ST/D
2 160 ST/D
ANHYDROUS AMMONIA
AREA 8200 ASH DEWATERING AND TRANSFER
AREA 1800 GAS LIQUOR SEPARATION
AREA 1600 PHENOSOLVAN G.L. STRIPPING
PROCESS WATER TO C.T. MAKEUP
1 40,355 GALS/D
2 20,600 GALS/D
CRUDE PHENOLS
WATER FROM AREA 8100
OIL
1 177 430 GALS/D
2 90 490 GALS/D
AREA 4000 SULFUR RECOVERY
LOW Btu FUEL GAS
1 1498 MM SCFD (DRY)
HHV = 300 Btu/SCF
2 765 MM SCFD (DRY)
HHV 300 Btu/SCF
BC HYDRO
LOW BTU FUEL GAS
PROCESS BLOCK FLOW DIAGRAM
AND MATERIAL BALANCE
LURGI OXYGEN BLOWN PROCESS
JOB 7642
LUMMUS
ASH TO DISPOSAL
1 10,200 ST/D
2 5,200 ST/D
TAR
100,558 GALS/D
SULFUR
1 163 ST/D
2 83 ST/D
NOTE: 1 2000 MW POWER PLANT
2 900 MW POWER PLANT


SKETCH NO. 2
PROCESS STEAM
976,680 LBH
COAL
22,108 T/D
AS REC'D
AREA 1100
GASIFICATION
AREA 1300
WASTE HEAT
RECOVERY
AND
GAS COOLING
AREA 4000
SULFUR
REMOVAL
VIA
STRETFORD
AREA 8400
FINAL
TREATING
AND
EXPANSION
FUEL GAS
PRODUCT
1,198 MMSCFD (DRY)
HHV 192 Btu/SCF
SULFUR
BY PRODUCT
74 T/D
AREA 8500
AIR
COMPRESSION
AREA 1800
GAS LIQUOR
SEPARATION
TAR OIL PRODUCT
97,700 GAL/D
AREA 8200
ASH
HANDLING
AREA 1600
PHENOSOLVAN
AND G.L.
STRIPPING
TAR
PRODUCT
45,100 GAL/D
ASH TO DISPOSAL
5,527 T/D
EXCESS
WATER TO
WATER TREATMENT
ANHYDROUS AMMONIA
138 T/D
B.C. HYDRO
LOW BTU FUEL GAS - 900 MW POWER PLANT
PROCESS BLOCK FLOW DIAGRAM AND
MATERIAL BALANCE
SULFUR REMOVAL VIA STRETFORD
LURGI AIR BLOWN PROCESS
JOB C-7642
LUMMUS

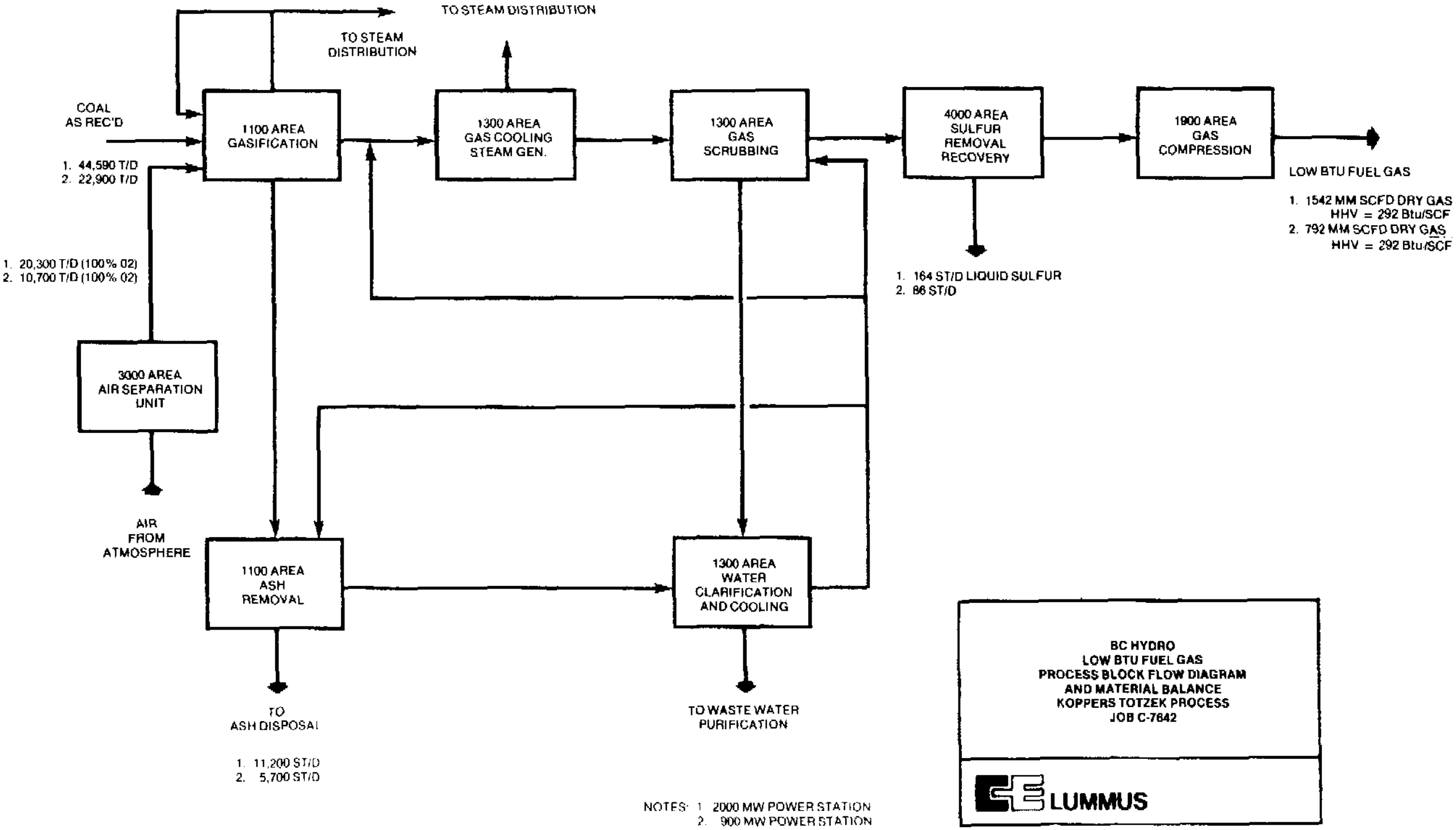
SKETCH NO. 3



BC HYDRO
LOW BTU FUEL GAS - 900 MW PLANT
PROCESS BLOCK FLOW DIAGRAM AND
MATERIAL BALANCE
LURGI AIR BLOWN PROCESS
WITH SULFUR REMOVAL
VIA HOT POTASSIUM CARBONATE
JOB C-7842

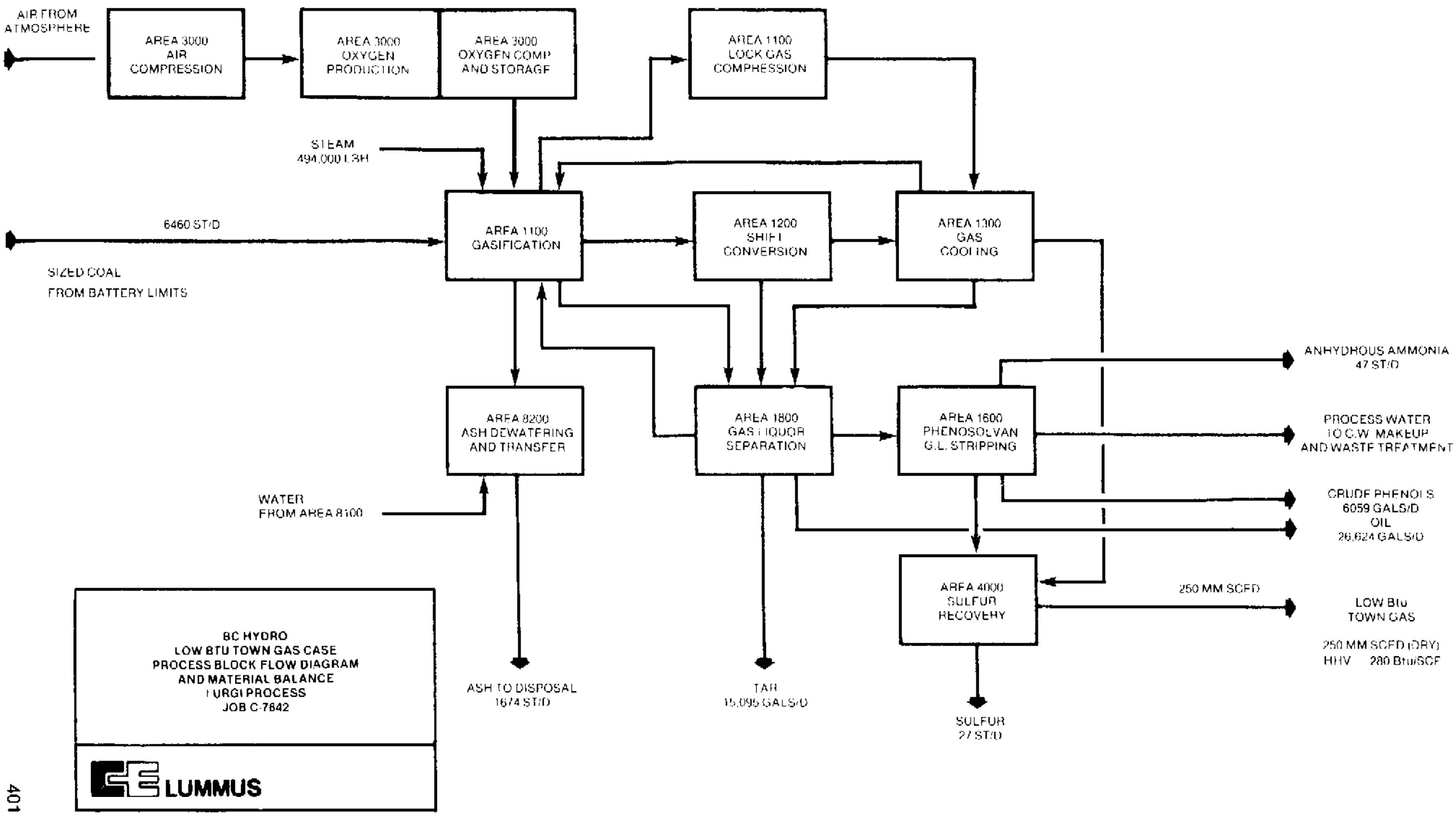
LUMMUS

SKETCH NO. 4

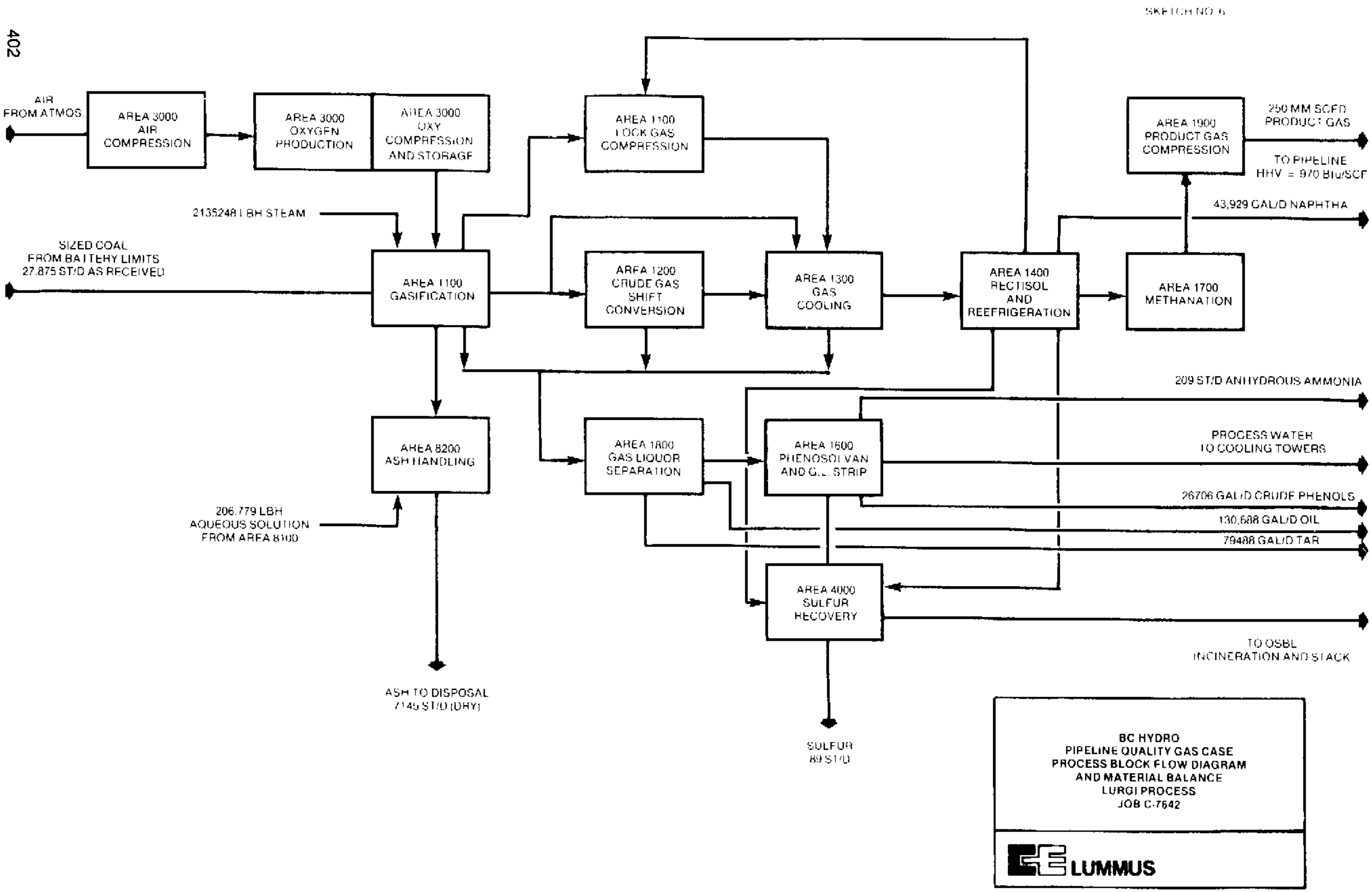


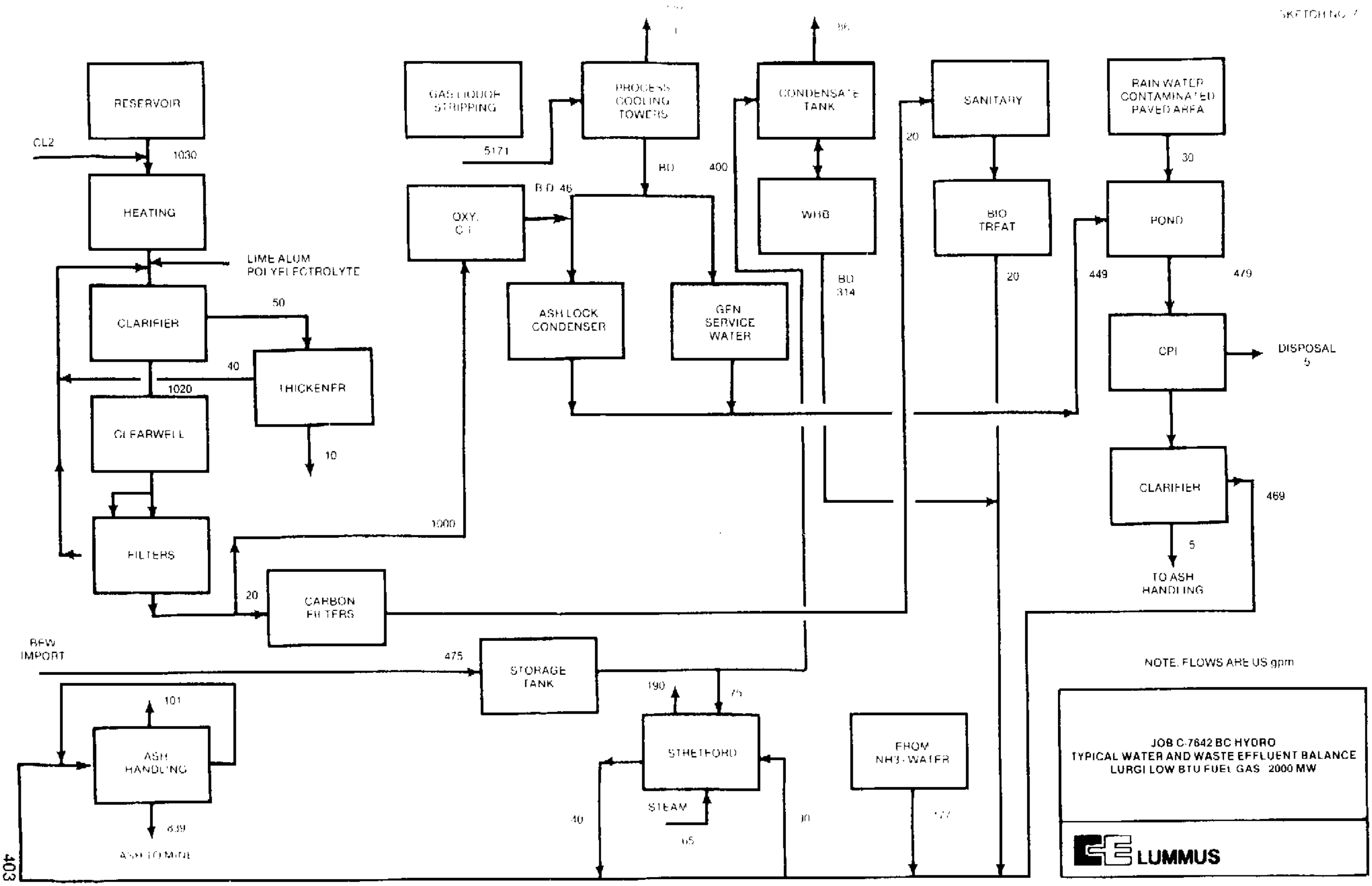
TO STEAM DISTRIBUTION
TO STEAM DISTRIBUTION
COAL AS REC'D
1. 44,590 T/D
2. 22,900 T/D
1100 AREA GASIFICATION
1300 AREA GAS COOLING STEAM GEN.
1300 AREA GAS SCRUBBING
4000 AREA SULFUR REMOVAL RECOVERY
1900 AREA GAS COMPRESSION
LOW BTU FUEL GAS
1. 1542 MM SCFD DRY GAS HHV = 292 Btu/SCF
2. 792 MM SCFD DRY GAS HHV = 292 Btu/SCF
1. 164 ST/D LIQUID SULFUR
2. 86 ST/D
1. 20,300 T/D (100% O2)
2. 10,700 T/D (100% O2)
3000 AREA AIR SEPARATION UNIT
AIR FROM ATMOSPHERE
1100 AREA ASH REMOVAL
1300 AREA WATER CLARIFICATION AND COOLING
TO ASH DISPOSAL
1. 11,200 ST/D
2. 5,700 ST/D
TO WASTE WATER PURIFICATION
NOTES: 1. 2000 MW POWER STATION
2. 900 MW POWER STATION
BC HYDRO
LOW BTU FUEL GAS
PROCESS BLOCK FLOW DIAGRAM
AND MATERIAL BALANCE
KOPPERS TOTZEK PROCESS
JOB C-7642
LUMMUS

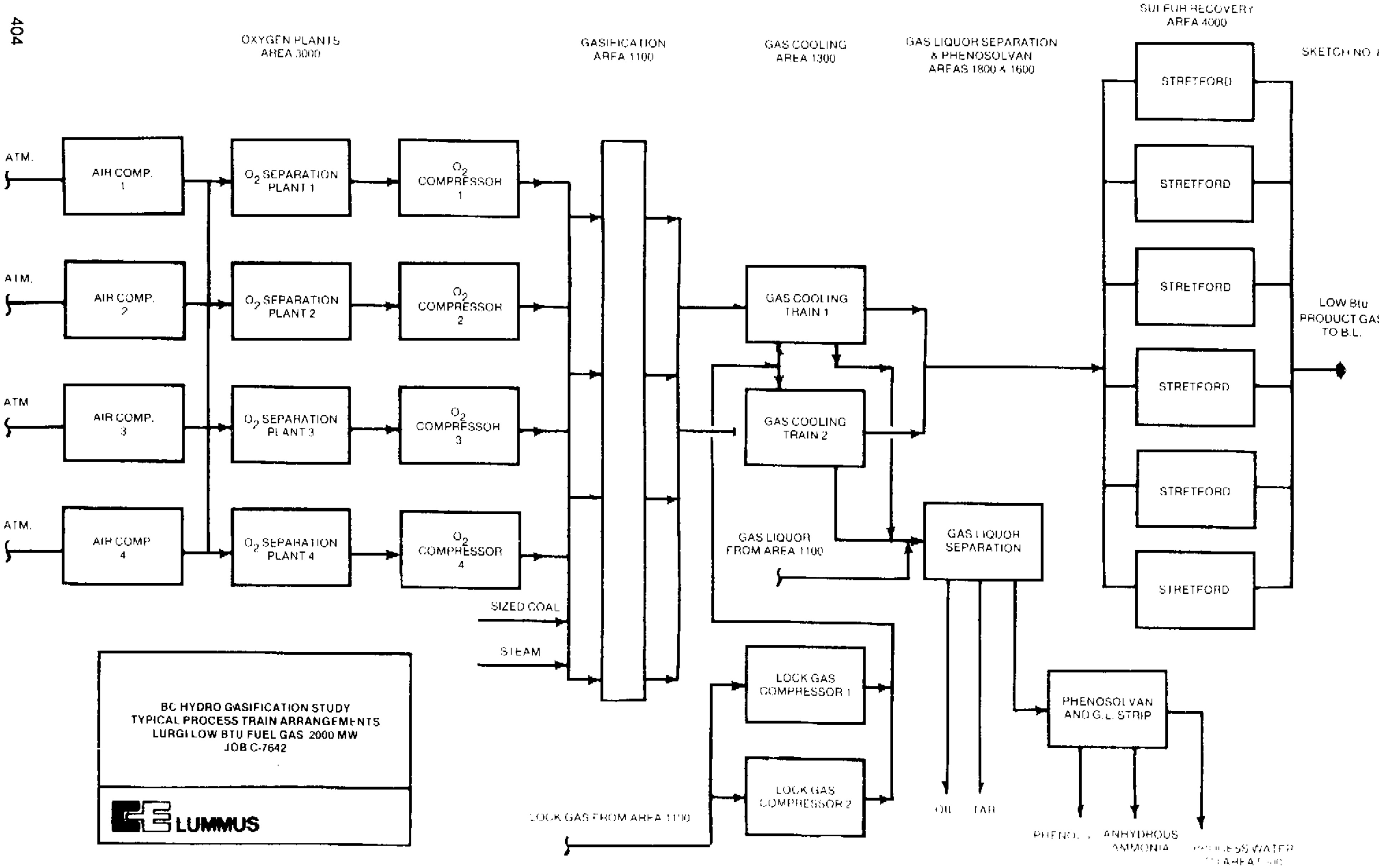
SKETCH NO. 5

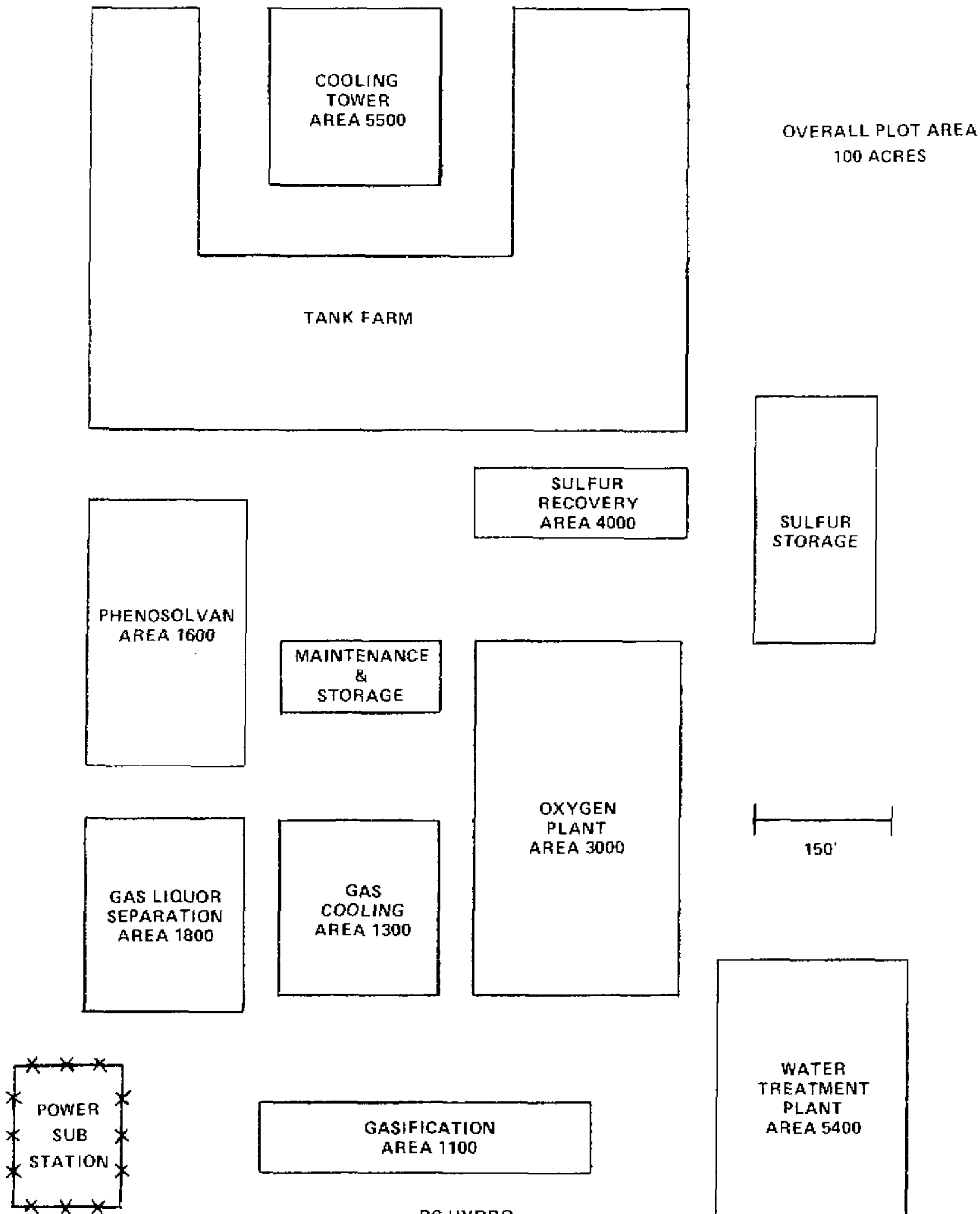


BC HYDRO
LOW BTU TOWN GAS CASE
PROCESS BLOCK FLOW DIAGRAM
AND MATERIAL BALANCE
LURGI PROCESS
JOB C-7642


SKETCH NO. 6
AIR FROM ATMOS.
AREA 3000 AIR COMPRESSION
AREA 3000 OXYGEN PRODUCTION
AREA 3000 OXY COMPRESSION AND STORAGE
AREA 1100 LOCK GAS COMPRESSION
AREA 1900 PRODUCT GAS COMPRESSION
250 MM SCFD PRODUCT GAS
TO PIPELINE HHV = 970 Btu/SCF
43,929 GAL/D NAPHTHA
2135248 LBH STEAM
SIZED COAL FROM BATTERY LIMITS 27,875 ST/D AS RECEIVED
AREA 1100 GASIFICATION
AREA 1200 CRUDE GAS SHIFT CONVERSION
AREA 1300 GAS COOLING
AREA 1400 RECTISOL AND REEFRIGERATION
AREA 1700 METHANATION
209 ST/D ANHYDROUS AMMONIA
AREA 8200 ASH HANDLING
AREA 1800 GAS LIQUOR SEPARATION
AREA 1600 PHENOSOLVAN AND G.L. STRIP
PROCESS WATER TO COOLING TOWERS
26706 GAL/D CRUDE PHENOLS
130,688 GAL/D OIL
79488 GAL/D TAR
206,779 LBH AQUEOUS SOLUTION FROM AREA 8100
AREA 4000 SULFUR RECOVERY
TO OSBL INCINERATION AND STACK
ASH TO DISPOSAL 7145 ST/D (DRY)
SULFUR 89 ST/D
BC HYDRO
PIPELINE QUALITY GAS CASE
PROCESS BLOCK FLOW DIAGRAM
AND MATERIAL BALANCE
LURGI PROCESS
JOB C-7642
LUMMUS


SKETCH NO. 7
RESERVOIR
CL2
1030
HEATING
LIME ALUM
POLYELECTROLYTE
CLARIFIER
50
THICKENER
40
1020
CLEARWELL
10
FILTERS
20
CARBON
FILTERS
1000
GAS LIQUOR
STRIPPING
5171
PROCESS
COOLING
TOWERS
BD
OXY.
B.D. 46
ASH LOCK
CONDENSER
GEN
SERVICE
WATER
400
CONDENSATE
TANK
WHB
BD
314
SANITARY
20
BIO
TREAT
20
RAIN WATER
CONTAMINATED
PAVED AREA
30
POND
449
479
CPI
DISPOSAL
5
CLARIFIER
469
5
TO ASH
HANDLING
BFW
IMPORT
475
STORAGE
TANK
190
75
STRETFORD
STEAM
65
40
FROM
NH3 - WATER
ASH
HANDLING
101
ASH TO MINE
NOTE: FLOWS ARE US gpm
JOB C-7642 BC HYDRO
TYPICAL WATER AND WASTE EFFLUENT BALANCE
LURGI LOW BTU FUEL GAS 2000 MW
LUMMUS


OXYGEN PLANTS
AREA 3000
GASIFICATION
AREA 1100
GAS COOLING
AREA 1300
GAS LIQUOR SEPARATION
& PHENOSOLVAN
AREAS 1800 & 1600
SULFUR RECOVERY
AREA 4000
SKETCH NO. 8
ATM.
AIR COMP. 1
AIR COMP. 2
AIR COMP. 3
AIR COMP. 4
O_2 SEPARATION PLANT 1
O_2 SEPARATION PLANT 2
O_2 SEPARATION PLANT 3
O_2 SEPARATION PLANT 4
O_2 COMPRESSOR 1
O_2 COMPRESSOR 2
O_2 COMPRESSOR 3
O_2 COMPRESSOR 4
SIZED COAL
STEAM
GAS COOLING TRAIN 1
GAS COOLING TRAIN 2
GAS LIQUOR FROM AREA 1100
GAS LIQUOR SEPARATION
LOCK GAS COMPRESSOR 1
LOCK GAS COMPRESSOR 2
LOCK GAS FROM AREA 1100
OIL
TAR
PHENOSOLVAN AND G.L. STRIP
ANHYDROUS AMMONIA
STRETFORD
LOW Btu PRODUCT GAS TO B.L.
BC HYDRO GASIFICATION STUDY
TYPICAL PROCESS TRAIN ARRANGEMENTS
LURGI LOW BTU FUEL GAS 2000 MW
JOB C-7642
LUMMUS

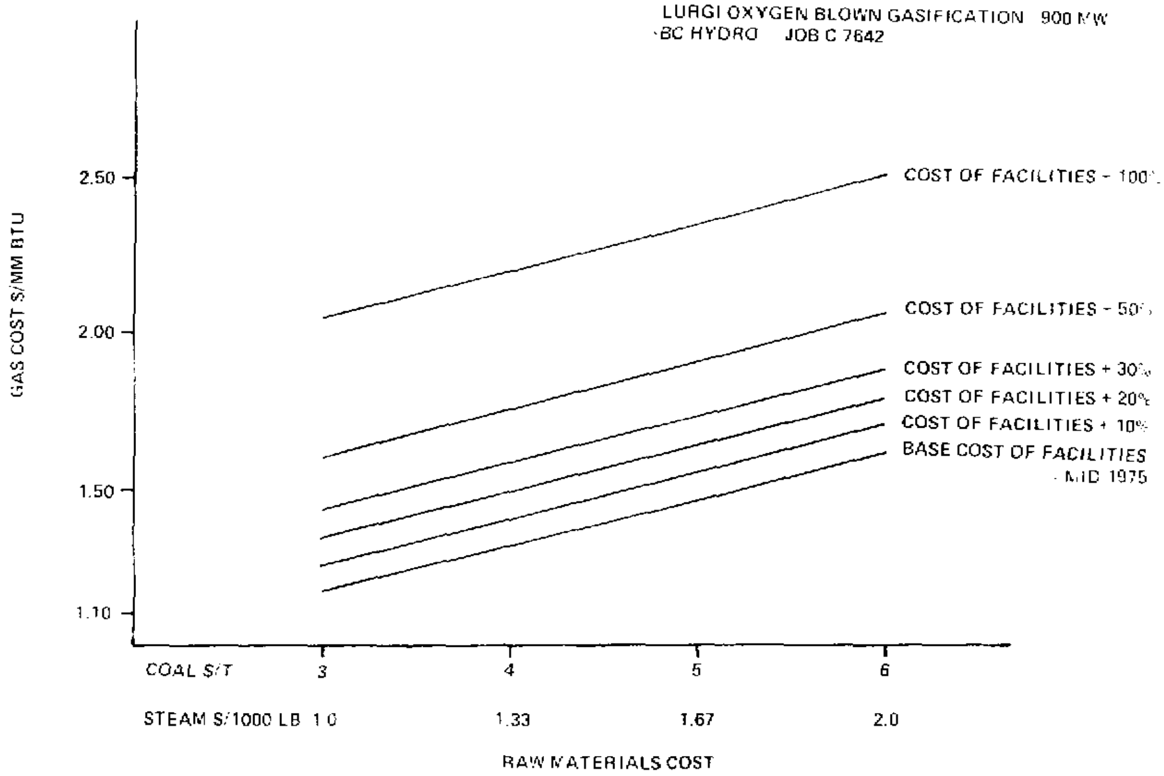


BC HYDRO
TYPICAL PLOT PLAN ARRANGEMENT
PROCESSING AREAS
LURGI LOW BTU FUEL GAS
2000 MW CASE

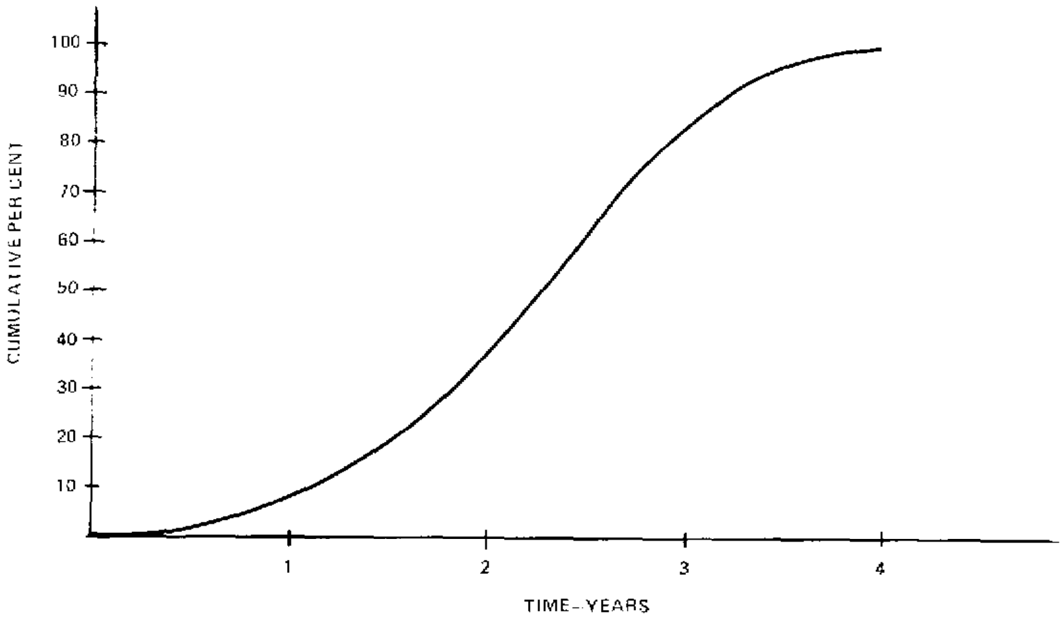
SKETCH NO. 9

FIGURE NO. 1

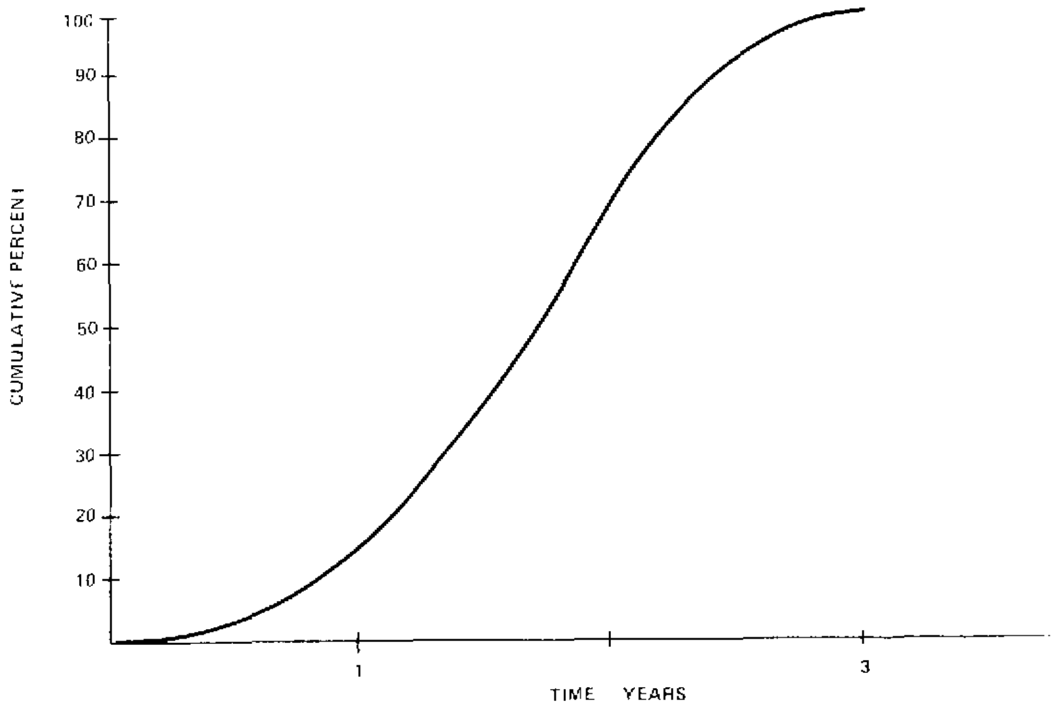
SENSITIVITY OF GAS COST TO CAPITAL AND OPERATING COSTS
LURGI OXYGEN BLOWN GASIFICATION 900 MW
BC HYDRO JOB C 7642



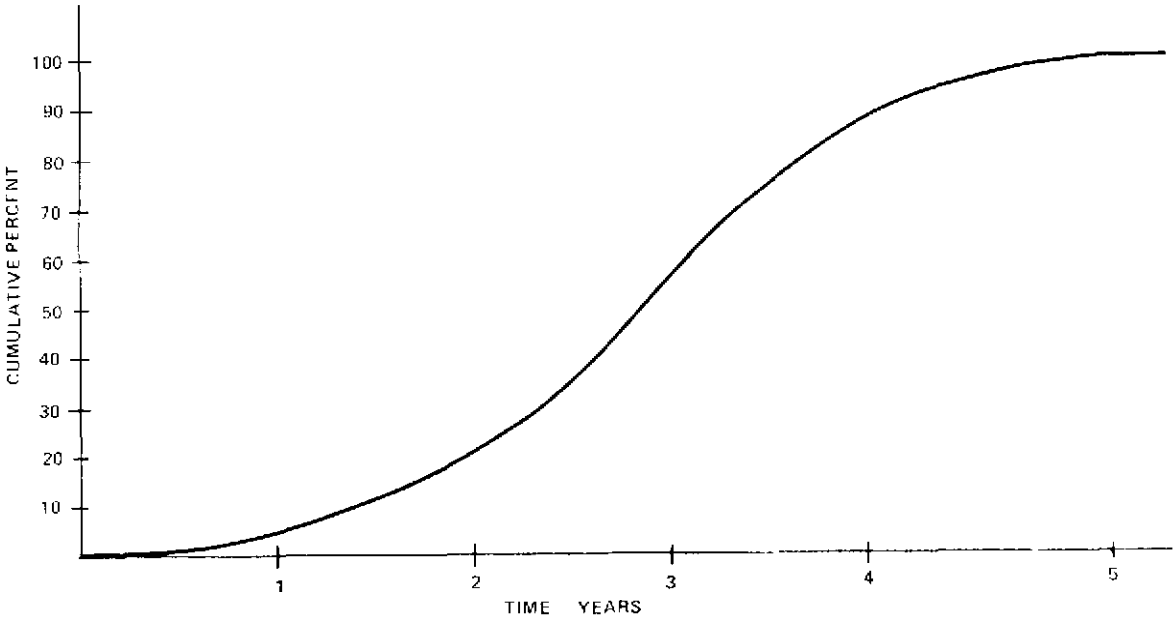
PROJECT SCHEDULE – CASH FLOW CURVE
LURGI OXYGEN BLOWN 900 MW
LURGI AIR BLOWN - BOTH CASES
48 MONTHS
BC HYDRO JOB C–7642
FIGURE NO. 2



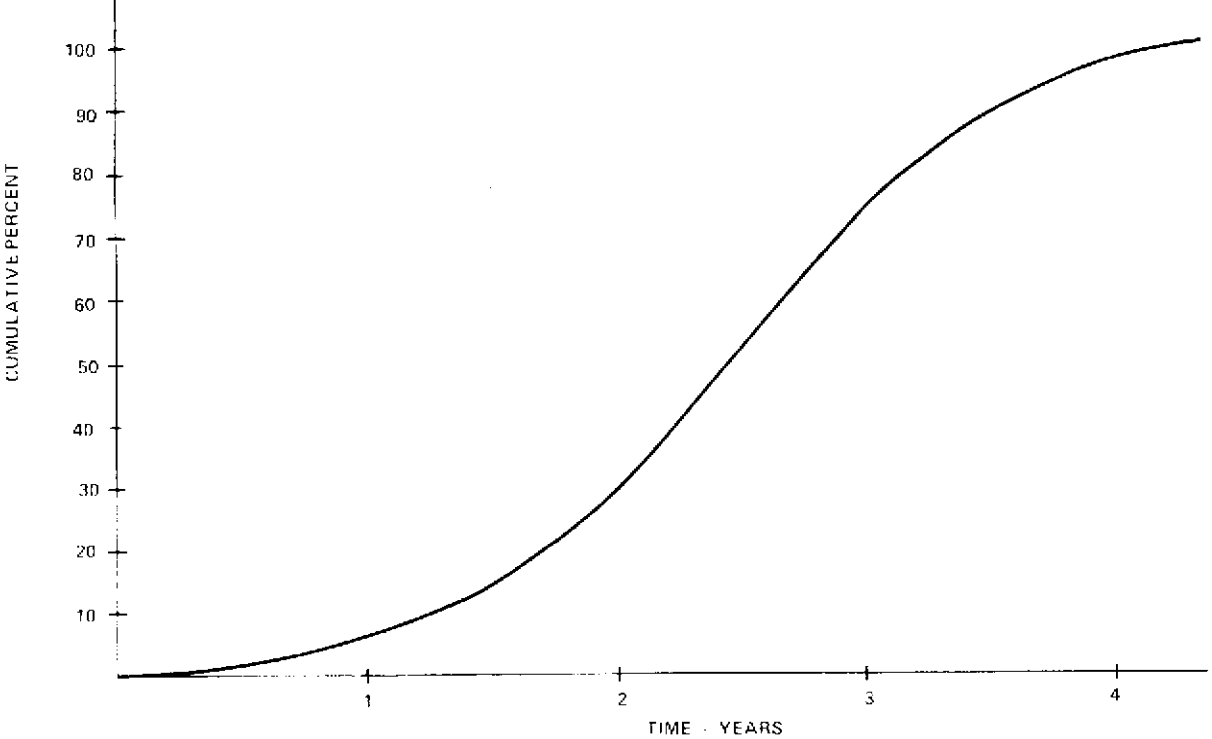
PROJECT SCHEDULE - CASH FLOW CURVE
TOWN GAS - 36 MOS.
B.C. HYDRO JOB C-7642
FIGURE NO. 3



PROJECT SCHEDULE CASH FLOW CURVE
KOPPERS TOTZEK 2000 MW
BC HYDRO JOB C 7642
64 MOS.
FIGURE NO. 4



PROJECT SCHEDULE   CASH FLOW CURVE
KOPPERS TOTZEK 900 MW
BC HYDRO   JOB C-7642
52 MOS.
FIGURE NO 5



PROJECT SCHEDULE - CASH FLOW CURVE
LURGI OXYGEN BLOWN 2000 MW
LURGI PIPELINE GAS
56 MOS.
JOB C-7642 B.C. HYDRO
FIGURE NO. 6