DOE/ET/13402 -- T21 (Vol. 44)

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COMMERCIAL PLANT CONCEPTUAL DESIGN AND EVALUATION

VOLUME IV A

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM - PHASE I

PREPARED FOR THE
UNITED STATES
DEPARTMENT OF ENERGY
UNDER CONTRACT NO. ET-77-C-01-2577

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EBASCO

EBASCO SERVICES INCORPORATED
2 RECTOR STREET
NEW YORK, NEW YORK U.S.A.
MARCH 1979

DOE/ET-77-C-01-2577
New No. DE-AC02-77ET13402
Distribution Category UC-89

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PHASE I

COMMERCIAL PLANT CONCEPTUAL DESIGN AND EVALUATION
VOLUME IV A

NOTICE

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Prepared for the

U.S. DEPARTMENT OF ENERGY

Assistant Secretary for Energy Technology
Office of Fossil Fuels

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VOLUME IV

Trade-Off Studies

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY - I
GASIFIER OPERATING PRESSURE

APRIL 1978

PREPARED FOR
EBASCO SERVICES INC.
BY
HUMPHREYS AND GLASGOW LTD.

Synthesis Gas
Demonstration Plant
Program, DOE

Trade-off Study I

Gasifier Operating Pressure

March 1978

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INTRODUCTION

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SECTION 1

INTRODUCTION

1.1 Objective

The objective of Trade-off Study I is to determine the operating pressure for a Commercial Plant to produce 3500 STD ammonia from coal, using the Texaco Coal Gasification Process System (TCGP System), and based on an overall optimisation of operating and capital costs and the examination of the relevant associated technical parameters which change with the selection of pressure.

1.2 Scope

1.2.1 For the study, a range of pressures was selected for the coal gasification stage which adequately extend the range of current knowledge and proven technology into areas which may yield a potentially more economic solution. The four pressures selected for the detailed study were 800, 1200, 1500 and 2500 psig. The pressures refer to the operating condition within the Texaco gasifier. Subsequently, following preliminary results from the study of these four cases, some less detailed engineering definition and estimating work was done for a pressure case at 500 psig to establish cost and performance trends below 800 psig.

1.2.2 Processes which are necessarily associated with the production of ammonia from coal gasification by the Texaco partial oxidation process are also examined for determination of relative sensitivity to operating pressure. This part of the study also includes an examination of associated HP steam generation and the development of a steam system and energy balance.

1.2.3 The study seeks to establish the principal cost step changes and equipment breakpoints which occur as the pressure is increased. Examples examined include the number of process trains, number of heat exchanger shells and unit design, piping and valve specification and rating changes, number of compressor casings and stages required for major duties.

1.2.4 For the purpose of this study only the Rectisol process is considered in detail for the removal of CO_2 and H_2S from the gasifier product stream. Although other washing processes will be considered at a later stage of development of the Phase 1 study it was, for a first approximation, considered that all washing processes should be equally sensitive to pressure on a differential basis, and therefore the use of alternative processes to Rectisol would not influence the overall decision for the selection of the optimum pressure.

1.2.5 No attempt is made to optimise the ammonia synthesis loop section.

1.2.6 Comparative data are considered from competing suppliers to optimise the CO shift system.

1.2.7 Coal preparation plant, offsite power generation and sulphur removal Units are considered as common entities to all pressure levels, differing from case to case only marginally in scale.

1.2.8 Air separation is similarly regarded as common to all the cases considered except for a variation in capacity of about 5.5% over the range of pressure cases. Raising the oxygen to the pressure required, either by compression or by liquid pumping prior to vaporisation of the oxygen, raises questions of the most suitable technology to be applied, and these are considered in Section 3.4 "Commercial Evaluation and Risk Analysis".

SECTION 2

CONCLUSIONS AND RECOMMENDATIONS

CONTENTS

Section 2

CONCLUSIONS AND RECOMMENDATIONS

Comments in this section relate to the following data which is included in Section 4.

Plant operating and ammonia production cost breakdown. (Fig 4-2)
Energy consumption (Fig 4-3)
Estimated installed plant capital cost - effect of pressure. (Fig4-1
Tabulated operating costs. (Table 4-2)
Tabulated capital costs - for four pressure cases. (Table 4-1)

An overall process block diagram - showing the number of trains in each process unit - is included in Section 3.

2.1 Conclusions

2.1.1 Energy Consumption

Coal provides the feedstock to the gasifier, and also the fuel for firing the HP boilers. Electrical power requirements are imported from the supply grid.

Gasifier feed coal requirements increase by about 9% within the gasifier operating pressure range 500 to 2500 psig. However this increase is compensated by a reduction of 51% in coal required for HP steam generation within the same pressure range.

The net effect of these two trends is that the total coal consumption is at a maximum at 500 psig and is almost constant within the pressure range 1200 to 2500 psig, suggesting that no further advantage is gained by gasifying at pressures much above 1200 psig.

Electricity consumption is at a minimum level at 1200 psig and begins to rise steeply above a pressure of about 1500 psig.

2.1.2 Capital Cost

Preliminary plant capital cost evaluations showed that total plant investment costs increase above a gasifier operating pressure of about 1000 psig. This evaluation was based on limited data for the Texaco gasification stage and a European budget estimate for acid gas removal and nitrogen wash units.

To improve the definition in certain critical areas the following additional work was done:

- Calculation and production of a definitive basis for the Texaco coal gasification process (TCGP) system with the number and size of gasifiers and associated equipment items accurately determined for the proposed coal feedstock and design basis.
- Reevaluation of the number of trains and size of columns required for the Rectisol unit operating at feed gas pressures corresponding to the gasifier unit within a range 800 to 1500 psig, and preparation of a new cost estimate in the U.S. relating this design to proven experience under installed U.S. conditions.

Resulting from this work, decisions affecting the numbers of units and trains in the gasifier and Rectisol process sections based on additional work and data have enabled the production of an overall block process schematic defining the number of trains in all principal process areas. This drawing is included in Section 3.

Revised capital cost estimates based on the improved plant definition show the following trends within the increasing pressure range 500 to 2500 psig.

Units with near-constant cost:

- Sulphur recovery
- Ash and carbon handling
- Coal preparation and handling

Units with steadily decreasing cost:

- Boiler and offsites
- Ammonia synthesis section including compression

Units with steadily increasing cost:

- Air separation plant and oxygen compression
- CO shift; vessels increase in cost, but heat exchangers and catalyst decrease; the overall effect is to increase the unit cost with pressure

Units showing a minimum cost at some point within the pressure range considered:

- Gasification

  The combination of number and size of gasifiers and associated equipment per train, and the effect of pressure on the weight of vessels, produces a minimum total cost for the 1200 psig gasifier pressure case.

- Rectisol

  Recent data indicates a minimum cost for the 1200 psig gasifier pressure case. This result is the outcome of the combined effects of pressure on the solution absorption performance and weight of vessels and exchangers.

- Nitrogen Wash

  The estimates show constant cost from 800 to 1200 psig with increasing cost above this pressure. This trend is explained by:

  a) Low pressure plate fin exchangers can be used for the 800 psig case.

  b) High pressure plate fin exchangers are required for the 1200 psig case.

  c) Coiled exchangers are required above the 1200 psig case.

  d) The cost of the exchangers increases in proceeding from Condition a to Condition c above.

  e) Increasing the pressure in the range 800 to 1200 psig favours the absorption reaction, hence lowering vessel costs; but further increasing the pressure introduces possible problems from interaction of high-density phases.

Overall Capital Cost for 3500 STD Ammonia Plant

Combining the costs estimated for the individual plant sections yields a minimum cost of US$267m for the 1200 psig case. The estimated cost at 800 psig represents an increase of about 5% over this minimum, whilst there is an increase of about 20% for the 2500 psig case.

2.1.3 Ammonia Production Cost

The operating costs for five pressure cases are summarised in Table 4-2 included in Section 4. The total ammonia cost is obtained by adding together the appropriate costs for:

- Raw materials and utilities, principally coal and electric power;
- Operations, including labour, supervision and maintenance;
- variable overhead;
- Credit for sulphur byproduct;
- Capital charges including depreciation and finance charges.

The figures for the last of the above categories, which includes an appropriate allowance for rate of return on investment, are directly proportional to the capital costs given in the bottom line of Table 4-1, and as such are subject to the exclusions in Note 4 to that table; these exclusions are however common to all fives cases and therefore do not invalidate the comparison. The ammonia production costs given in the bottom line of Table 4-2 also make no provision for product distribution and marketing which again does not affect the comparison.

On this basis, the minimum ammonia production cost of US$127 per short ton corresponds to the 1200 psig gasification case. This ammonia cost is about 9% lower than the cost for production at the 500 psig case, and 12% lower than the theoretical cost (ignoring lower potential plant availability and higher risk factors) for the 2500 psig case.

2.1.4 Gasification Pressures Above 1200 psig

Texaco partial oxidation of oil and associated downstream gas washing systems have been demonstrated at about 2400 psig on a pilot scale. Coal gasification by the Texaco process has recently been demonstrated on a semi-commercial plant in West Germany at a design pressure reported to be between 500 and 600 psig. The next stage of pilot plant experimentation currently planned by Texaco within the foreseeable future will only permit demonstration up to pressures not exceeding 1200 psig. The demonstration of gasifier product gas composition, especially the detailed analysis of trace components present, will be of importance for fixing the detailed design of the downstream gas washing and treatment plant.

Demonstration of the Texaco coal gasification process for the production of synthesis gas in a commercial-size plant operating at a pressure of 1200 psig also has the benefit of meeting the requirements to generate gas which, after methanation stages, could be used directly as pipeline gas at 1000 psig.

Unquantifiable risks would be involved in constructing a production unit at an operating pressure not satisfactorily demonstrated by pilot plant operation. The principal objective of the current pilot plant programme is to confirm the predictions of performance at gasification pressures up to 1200 psig used in the preparation of this report, after which the economics and risks of operation at a higher pressure can be more realistically evaluated.

2.1.5 Acid Gas Removal

The correlation of cost versus pressure so far obtained for Rectisol indicates that high pressure results in a considerable increase in equipment costs for this process unit. The same conclusions for Rectisol unit performance versus pressure may not apply with some alternative gas washing processes. The results of Trade off Study II (TOS II), "Gas Purification Alternatives", could indicate that lower overall capital cost and improved operating economics and performance are obtained from one of these alternative processes.

2.2 Recommendations

Proceed with the engineering specification and cost estimation of a Commercial Plant to produce 3500 STD of ammonia from coal using the following design and operating conditions and process selection:

1. Synthesis gas preparation plant section to be based on an operating pressure of 1200 psig in the Texaco gasifier.
2. The plant to consist of four trains of Texaco gasifiers and associated equipment. The plant shall be capable of maintaining the full ammonia production capacity with three gasifier trains on line.

3. Optimisation of the process variables within the TCGP shall be done at a gasifier operating pressure of 1200 psig, and design conditions shall be selected to achieve an overall plant optimisation at this pressure level.

4. The final selection of the acid gas removal process to be used shall depend on the results obtained from TOS II. The feed gas operating pressure shall correspond to a gasification stage reference pressure of 1200 psig.

5. The HP steam system shall be specified at 1500 psig, 940°F, referenced at the steam turbine inlet stop valve.

SECTION 3

TECHNICAL EVALUATION

CONTENTS

SECTION 3

TECHNICAL EVALUATION

3.1 Design Basis

3.1.1 Introduction

The data given below have been used as the design basis for Trade-off Study I. They are based on data provided by Ebasco together with, where necessary, assumptions which are considered to form a reasonable basis for comparative study. Some of these data may be amended as better information becomes available in subsequent activities of Phase I.

Paragraph 3.1.2 summarises the cases taken for comparison. The data in all subsequent paragraphs in this sub-section applies to all four cases.

3.1.2 Pressure Range Studied

Four cases have been evaluated, corresponding to four gasification pressures. For each case, outline data have been prepared corresponding to a nett production of 3500 STD ammonia, the synthesis loop and ammonia recovery section being identical for the four cases.

The cases are summarised as follows:

Case No.	Gasification Pressure, psig	Syngas Compressor & Suction Pressure, psig
A1	800	619
B	1200	952
C	1500	1212
D	2500	2074

Some design parameters, such as wall thickness of vessels and piping, vary more or less continuously with operating pressure (other things remaining constant), and the effects of varying pressure on capital cost can be represented by a continuous curve so far as these aspects are concerned. Other features, however, are subject to step changes at certain pressures corresponding, for example, to a change in design code or standard frame size of equipment. In order to allow for these variations, additional data have been developed in pressure-sensitive areas in order to identify discontinuities in the cost curves.

3.1.3 Ammonia Product

Rated Capacity	3,500 STD
Annual Production	1,115, 000 ST (330 days/year)
Specification:	
State	Liquid anhydrous
Pressure	In atmos pressure tank
Temp.	At equilibrium in tank
Purity	99.8% ww min.
Water	0.2% ww max.
Oil	15 ppm ww. max.

3.1.4 Coal

The same coal will be used for gasification and for firing the auxiliary high pressure boiler. The following properties have been assumed for the present study:

Ultimate analysis, moisture free:		
carbon	72.76	wt %
hydrogen	5.24	
nitrogen	1.63	
sulphur	3.35	
oxygen	7.65	
ash	9.37	
Higher heating value moisture free basis	13,368	BTU/lb.

Since preparing the heat and mass balances based on the above coal properties, more performance data for the gasification section has been developed by Texaco Development Corporation, based on a coal believed to be typical of Western Kentucky coalfields.

This data has been reviewed and allowed for in presenting comparative data in this report. Data previously developed which does not affect the comparison has not been amended.

The price of coal has been taken as $0.748/MM Btu delivered to the site for the purposes of this study.

3.1.5 Sulphur By-product

State	Elementary, molten
Purity	99.8% by weight

3.1.6 Utilities

a) Cooling Water

Supply temperature	89°F
Return temperature (max)	109°F

b) Electricity

Purchased from supply grid

Normal supply characteristics

Cost	1.45 cent/Kw-hr

c) Raw Water

Cost	$0.10/MUSG

3.1.7 Environmental Release Limits

SO_2	1.2 lb/MM Btu fuel input
Fly Ash	0.1 lb/MM Btu fuel input

3.2 Process Description

3.2.1 Introduction

As stated above, the objective of the present study is to evaluate the effects of varying the gasification pressure on the capital and operating costs of the Commercial Plant.

a) With regard to capital costs, the advantages to be expected from increased pressure stem from the smaller gas volumes to be handled and consequently smaller physical dimensions, for a given plant capacity, of certain items of equipment. Thus, if it is accepted that equipment sizes will be limited by considerations of fabrication and shipping, the number of trains required for a given plant capacity will tend to reduce with increase in operating pressure. On the other hand, the following factors will tend to limit or offset the benefits due to this effect:

(i) The wall thickness of a vessel or exchanger increases with increase in pressure for a given diameter, temperature and materials of construction. This not only adds to the weight of the

(i) Cont'd.

vessel but reduces the number of possible vendors where very high thicknesses result, with a possible increase in fabrication costs.

(ii) The discharge pressure of compressors and pumps feeding into the main process stream is greater the higher the operating pressure, resulting in more expensive equipment and, in some cases, involving a change in type of technology required, e.g. in the case of oxygen compression.

(iii) In some processes, the size of equipment is not determined by actual gas volume; in such cases there is no reduction in physical dimensions with increase in pressure and the cost of equipment will actually increase on account of increased wall thickness.

These factors are discussed in more detail below. The nett impact of these effects on capital cost is discussed in Section 2 of this report.

(b) Savings in operating costs with increase in gasification pressure can be expected to result from reduced overall power requirements. There is a significant reduction in the power required for syngas compression in going from Case Al to Case D. Although the power required for oxygen compression, coal slurry pumping and nitrogen compression increases, there is indeed a nett saving in total power for the process plant plus support facilities of about 5%.

The saving in operating costs is not, however, so large, the principal reason being the increased requirement, per ton of ammonia, of coal for gasification at high pressure. Though

(b) Cont'd.

this coal is not exactly wasted, being eventually converted into nitrogen wash tail gas, which is fired in the auxiliary boiler and superheater, this does represent a degrading of energy which reduces the overall efficiency of converting coal to ammonia. This and other aspects of the effect of varying the gasification pressure on process consumptions are discussed in further detail below, and the effects on the economics of ammonia production are discussed in Section 2 of this report.

3.2.2 Process Configuration

In order to prepare comparative data for the four cases specified in paragraph 3.1.2 above, a number of simplifying assumptions have been made. The salient features of the process route may be summarised as follows:

(a) Coal preparation by dry crushing and grinding;

(b) Texaco coal gasification, including carbon and ash recovery;

(c) CO conversion using 'dirty shift' catalyst;

(d) Acid gas removal by the Rectisol process, followed by final clean-up by nitrogen wash;

(e) Synthesis gas compression by turbine-driven centrifugal compressor;

(f) Ammonia synthesis and recovery using two synthesis loops in parallel operating at 3400 psig.

(g) Air separation followed by either oxygen compression or liquid oxygen pumping and vaporisation.

(h) Steam for large power requirements generated in an auxiliary boiler at 1800 psig and 986°F. (Note that these conditions for steam generation have been revised since calculating the mass balances used for this study, but this change does not affect the comparison of gasifier pressures).

i) Sulphur recovery from Rectisol off-gas by the Claus Kiln process.
j) Claus tail gas and boiler stack to be cleaned up to meet environmental requirements.

The main departures from the scheme outlined in the above proposal are as follows:

a) Two stages of CO conversion have been adopted, instead of three as previously proposed.
b) Two levels of medium pressure steam, raised from the make gas train, have been chosen, in place of the one level taken previously, in order to improve heat recovery.
c) The gasifier feed preheaters and condensate stripper have been eliminated from the gasification section.

These changes are referred to further in the appropriate paragraphs below.

A number of the above features may be varied in subsequent work on Phase I of the project as the result of further development and trade-off studies; it is not considered that such changes will invalidate the conclusions of the comparison of gasification pressure.

3.2.3 Overall Flowscheme

The overall flowscheme is shown on the block diagram, Drg. No. 1821-X52-5, at the *end of this section*, which shows the main process units and the number of trains proposed for each unit. The only raw materials delivered to the site are coal and raw water. Large machinery is driven by steam generated on site, and smaller drives use purchased electricity. The requirements of these raw materials and utilities are summarised in sub-section 3.3 below.

Dry grinding of process feedstock has been assumed for this study; subsequent adoption of wet grinding would have a negligible effect on the conclusions of the pressure study.

3.2.4 Coal Preparation

The coal delivered to site is stored in the site stockpiles. From the stockpile, it is crushed to a size suitable for feeding to the pulverisers, and is then divided into process and boiler feeds so that pulverisation for these duties can be carried out separately. Buffer storage is provided both for pulveriser feed and for pulverised process feedstock.

3.2.5 Coal Gasification

The Texaco Coal Gasification process is used; basically this is the reaction between coal and oxygen at high temperature:

$$2C + O_2 = 2CO \text{ and } C + O_2 = CO_2$$

Since water is also present, as slurrying agent for the feed, the water gas reaction also takes place:

$$C + H_2O = CO + H_2$$

The final gas leaving the gasification zone is essentially a mixture of C, CO, CO_2, H_2 and H_2O, with the actual composition determined by the kinetics of the partial oxidation reactions.

There is also an equilibrium between methane and carbon oxides and hydrogen:

$$CO + 3H_2 = CH_4 + H_2O$$

$$CO_2 + 4H_2 = CH_4 + 2H_2O$$

One of the advantages of the high temperature which is a feature of the Texaco process is that these reactions go far to the left resulting in a very low methane content of the product gas. Higher pressure however results in a higher methane yield and this means that from a given amount of coal, less H_2 + CO is produced. Therefore one effect of higher pressure is to increase the amount of coal and oxygen feed to give the required amount of hydrogen in the synthesis loop feed gas.

The high temperature also ensures that the ash contained in the coal is completely melted - an essential feature

3.2.5 Cont'd.

of the Texaco process.

The flow scheme is as follows:

Pulverised coal is slurried with water in the slurry tank equipped with an agitator and slurry circulation pump. Gasifier feed is drawn from the circulation line by the slurry charge pump which injects a controlled flow of slurry to the gasifier. Oxygen is supplied at the necessary pressure to the gasifier and mixes with the slurry in the burner located in the head of the gasifier. The gasifier is a refractory lined pressure vessel in which the reactions referred to above take place. The raw gas produced is quenched with water in the base of the gasifier and the solidified ash is withdrawn as a slag by a lock hopper arrangement. Unconverted carbon, known as soot, is removed from the quenched gas by scrubbing with water, which is fed to the Carbon/Ash Recovery Unit where the carbon is recovered and recycled to the slurry tank.

The capital cost of the Gasification Unit is heavily dependent upon the number of gasifiers required for the duty. As the capacity of a single gasifier is determined by actual gas volumes, the number of gasifiers required, together with associated equipment such as slag hoppers and scrubber separators, reduces as the gasification pressure increases. The number of gasifiers required for the Commercial Plant with different operating pressures has been determined by Humphreys and Glasgow working in collaboration with Texaco Development Corporation and using assumptions on coal properties believed to be typical of Western Kentucky Coalfields. In addition to the four cases defined in paragraph 3.1.2, a fifth case of 500 psig has been calculated to verify the trend established for the range under study. The results, including one stand-by gasifier in each case, are as follows:

Case	Gasification Pressure	No. of Gasifiers
A2	500 psig	8
A1	800 psig	5
B	1200 psig	4
C	1500 psig	4
D	2500 psig	3

The impact of this on capital cost is shown in Section 4 of this report.

3.2.6 Carbon/Ash Recovery

The soot water produced in the Gasification Unit is cooled against make-up quench water and let down via a flash drum into the settler, which is a large tank in which the solids thicken to a concentration suitable for recycling to the slurry tank.

The size, and hence the cost of this Unit is relatively insensitive to the gasification pressure, and in any case accounts for only a small percentage of the total capital cost of the Commercial Plant; detailed analysis is therefore not required for this Trade-off Study.

3.2.7 CO Conversion

Gas from the Texaco gasifier contains over 40% carbon monoxide, which is converted to carbon dioxide by the shift reaction:

$$CO + H_2O = CO_2 + H_2$$

Steam required for the reaction results from the quenching of hot gases where the heat given up to the quench water vaporises enough to enable the above reaction to proceed.

Any remaining CO is removed in the Nitrogen Wash Unit, so there are obvious advantages in converting as much as as possible into $H_2 + CO_2$. This reduces the load on the Nitrogen Wash Unit, and also, because more hydrogen is produced, decreases the load on the gasifier and hence the capacity of the Air Separation Unit, etc. Against this the lower the CO level required, the more expensive the shift system will be. For the purposes of this study, the same condition as in the original proposal has been assumed, ie 3% carbon monoxide (dry basis) in the product gas.

3.2.7 Cont'd

The reaction is carried out over a cobalt/molybdenum oxide catalyst, which is sulphided by the H_2S present in the gas. This catalyst is reasonably active down to about 600°F; however as the reaction is exothermic, one single stage of shift conversion is self-limiting when the gas approaches the equilibrium temperature. One stage of conversion could reduce the CO to around 8% V/V and to obtain a lower concentration, the gas should be cooled and passed over a second stage of catalyst. This process can be repeated if required.

Information from shift catalyst vendors indicates that for a concentration of 3% CO, two stages are adequate; the use of three stages gives no significant saving in catalyst quantities. For instance one vendor shows 5720 cu.ft of catalyst in a two stage system giving 3% CO, and 6880 cu.ft in a three stage system giving 2.5% CO. The latter system also requires an extra vessel and exchanger with associated equipment. Therefore a two stage system is used for this pressure study. In the event of a lower CO level being selected at a later point in Phase I, then a three stage system could become preferable. The effect of such a change on the pressure comparison would however be marginal.

The information received from a vendor with experience of pressures up to about 1125 psig has been used as the basis of design of the shift system. In the range of operating experience, it is observed that higher pressure operation requires a smaller catalyst volume for a given duty. At higher pressures one vendor has been conservative, and not allowed for any further reduction in catalyst requirements. On the other hand another vendor has extrapolated from his experience; from this information, case D requires 60% of the catalyst required by Case A1.

The mechanical design of the shift converters is not straightforward, because to give a reasonable pressure drop (10 psi) through a single bed of catalyst would require a vessel diameter of 18 ft. to 21 ft depending on the case. Vessels of a more convenient diameter can be achieved by dividing the required catalyst volume into a number of beds in

3.2.7 Cont'd

parallel, which may be arranged in separate vessels, or with up to two beds within a single vessel, thus economising on vessel costs. A number of arrangements are conceivable, and some design work has been carried out to determine the arrangement to be used for this study. The arrangement now proposed for the Commercial Plant, which appears to satisfy the requirements of minimum capital cost consistent with operability and manageable vessel dimensions, is two parallel trains, each comprising a single vessel and associated heat exchange equipment. Each vessel is divided into two refractory lined compartments, the upper one holding the first stage catalyst, and the lower one the second stage catalyst bed. Thus gas entering the CO Conversion Unit is divided into two streams, then each stream is heated against converted gas from the second stage, then against gas from the first stage, before being fed to the first stage catalyst bed at about 640°F. The gas temperature rises to 730°F - 790°F (depending on the case), the CO content being reduced to 11% V/V. The gas leaving the first stage is cooled against unconverted gas to about 710°F before being fed to the second stage. The converted gas, containing 3% CO V/V, is cooled, first against feed gas, then in two waste heat boilers in series, operating at two pressures designated medium pressure (MP) and lower medium pressure (LMP). Further heat is recovered in a boiler feedwater heater and a low pressure (LP) boiler before the gas is finally cooled with deaerator feedwater and cooling water. The process condensate, formed during the cooling of the gas, is removed before the two streams pass to the Acid Gas Removal Unit.

The effects of increasing pressure on CO shift vessel design are:

a) Less catalyst is required resulting in a smaller total vessel volume.

b) Because of increased gas density, the velocity through the catalyst is reduced - this means that vessel diameters can be reduced.

c) Wall thicknesses have to be increased.

3.2.7 Cont'd.

So far as the heat exchange equipment is concerned a balance must be struck between maximising heat recovery and reducing capital cost by adoption of large temperature differentials. To achieve this a different LMP Steam level has been chosen for each of the four cases considered. The nett impact of variation of gasifier pressure on the cost of the CO Conversion Unit is discussed in Section 4 of this Report.

3.2.8 Gas Purification

The selection of the most appropriate technology for gas purification is the subject for study in Trade-off Study II As stated in paragraph 3.2.2 above, the present study is based on one of the possible process routes, ie Rectisol Acid Gas Removal, Adsorption and Nitrogen Wash. The feed gas enters the Rectisol Unit and, after methanol injection, is cooled with part of the nitrogen wash product gas and the tail gas. The methanol/condensate, which is removed from the feed gas passes to a water/methanol still. Here it is first heated and flashed in order to remove any absorbed H_2S and then fed into a column where the methanol is stripped using steam reboil and passed into the warm regenerator. On entering the wash column the H_2S/COS are scrubbed out in the bottom section of the column, the bulk of the CO_2 removed in the middle section while the final clean up occurs in the upper section where it is scrubbed with cold lean methanol. The gas then passes to a Nitrogen Wash Unit via a molecular sieve. The rich methanol leaves the wash column in two streams, the CO_2 rich stream from the middle section and the CO_2/H_2S rich stream from the base. These streams are cooled and flashed in order to remove absorbed hydrogen; the flash gas is then compressed and recycled back to the feed. In the cold regenerator the rich methanol is stripped of CO_2 using purge nitrogen from the molecular sieves, and the H_2S rich methanol is pumped from the top section of the column to the warm regenerator. The feeds to the cold regenerator are arranged such that the tail gas has an acceptably low H_2S concentration. This H_2S rich methanol is heated against regenerated methanol before being fed into the warm regenerator. Here it is stripped of H_2S using steam reboil before being fed back to the absorber. The overheads are chilled,

3.2.8 Cont'd

in order to remove any methanol before the H_2S stream is passed to the Sulphur plant.

The Adsorber Unit consists of two sets of two parallel beds of molecular sieves, one set adsorbing while the other is being regenerated. The unit is designed to remove the last traces of CO_2 and H_2S in order to prevent freezing up in the Nitrogen Wash, as well as traces of methanol carried over from the Rectisol Unit. Make gas is fed from the Rectisol Unit and passed through two beds of the unit and then on to the Nitrogen Wash. The regeneration is carried out by passing preheated nitrogen from the Air Separation Unit, counter current through the beds. The gas is then cooled before being used as stripping gas in the cold regenerator on the Rectisol Unit.

The Nitrogen Wash Unit employs a mixture of condensation and washing with liquid nitrogen to remove carbon monoxide and methane from the incoming gas. The unit consists essentially of a multistream heat exchanger and a column. The make gas and nitrogen are cooled against the column overheads before entering the column. The carbon monoxide and methane are condensed and removed from the base of the column and are then revaporised against the feed nitrogen before being sent off as fuel gas. A sidestream is removed from the column overheads before the final heat exchanger section and is used to chill the Rectisol Unit feed. The two streams then recombine before passing to the syngas compressors.

As Rectisol is a physical absorption process, increasing pressure might be expected to improve the economics by reduction of the methanol circulation required. In fact, the problem is not so simple: the particular nature of the gas to be treated in this project gives rise to several effects, some favourable and some unfavourable to the economics. These effects give a different balance at different pressures, the nett effect on power consumption is summarised in sub-section 3.3, below, whilst capital costs are discussed in Section 4.

3.2.8 Cont'd

With regard to the Adsorber, the volume of molecular sieve material required is unaffected by pressure, as this is determined by the quantity of acid gas to be removed. However, as pressure drop is reduced by increase in total pressure (for a given mass flow and vessel diameter), it is possible to increase the L/D ratio of the vessel to achieve a more economical design.

The main effect of pressure on the cost of the Nitrogen Wash Unit, apart from a progressive increase in wall thickness in going from Case A1 to Case D, is the change from plate exchangers to the more expensive tubular exchangers at a certain pressure. The maximum feasible pressure for the former is put at about 680 psig by one vendor and about 1000 psig by another. A further effect is that the physical properties of liquid nitrogen are less favourable to the absorption at very high pressures, so that either the tower height or the liquid nitrogen flow to the column must be increased. The result of all these effects is that the cost of this unit increases progressively with pressure, as discussed in Section 4.

3.2.9 Syngas Compression and Ammonia Synthesis

As stated in paragraph 3.2.2 above, this study is based on the adoption of two quench converter synthesis loops, each with its own synthesis gas compressor/circulator and refrigeration system. The gas leaving the nitrogen wash unit is very pure: this means that it is not necessary to take a gas purge from the synthesis loop, because any traces of argon and methane entering with the syngas will be dissolved out in the product ammonia.

It is not necessary to discuss this section in detail because with the exception of syngas compression, the gasification pressure has virtually no effect on ammonia synthesis. It is worth mentioning the heat of reaction, however, because it is possible to select the level at which this waste heat is recovered. There are many variations of loop design, and by selecting different quench temperatures, the temperature at which waste heat becomes available for recovery can be changed. In this case, the waste heat is used to raise 600 psig saturated steam, since this

3.2.9 Cont'd.

gives useful heat recovery with reasonable temperature approaches. Higher pressure steam could be produced, but this would increase the cost of most of the equipment in the loop.

Although the syngas compression duty reduces continuously with increase in pressure, the cost of the machines will be subject to step changes as the minimum number of stages required comes down. These possible step changes have been investigated, and the impact is discussed in Section 4.

3.2.10 Air Separation and Oxygen Compression/Pumping

Air separation is effected by proven technology for which there are a number of possible vendors available. Whilst there are differences between the anticipated power consumption and estimated capital costs provided by different vendors for the requirements of this project, such differences do not significantly affect the comparison of the different gasification pressures in this study.

Although a number of vendors were invited to put forward proposals based on liquid oxygen pumping, should this appear to have an advantage over oxygen gas compression, the two air separation plant vendors making concrete proposals on this question selected compression. In addition, a number of compressor manufacturers were approached directly. The power requirements are summarised in sub-section 3.3 below, and the capital costs are discussed in Section 4.

3.2.11 Heat Recovery and Steam System

The consumption and distribution of energy is a very important consideration in this pressure study, with high pressure gasification showing a distinct saving in total shaft power as indicated in subsection 3.3 below. Electric motors have been used for small drives only (generally less than 1MW). This accounts for about 16 MW of the total 190-200 MW shaft power.

3.2.11 Cont'd

Some steam is raised by recovering waste heat from downstream of the CO conversion section, and from ammonia synthesis. However, this is insufficient for the 190-200 MW mentioned above, so the remainder of the steam is raised in two independent coal fired boilers. This auxiliary steam is distributed at *1800 psig, 986°F, as noted in paragraph 3.2.2 above. Waste heat from the CO conversion section and synthesis loop is recovered and distributed as 580 psig steam, designated medium pressure (MP), with a small amount as 50 psig saturated steam for heating duties. These levels were selected as being compatible with the availability of heat at the CO shift outlet. However, on investigation, restricting the heat recovery from the make gas to a single steam pressure of 580 psig would impose an unnecessary penalty on the thermal efficiency of the plant as a whole. Accordingly some steam is raised at a lower medium pressure (LMP), the precise value of this pressure being different for each case, A1 to D. This lower pressure has been selected so that the temperature at which heat has to be rejected to cooling water (after preheating deaerator feed and boiler feed waters, and raising a smaller amount of 50 psig steam) is below about 230°F. Production of waste heat steam for the different cases is summarised in sub-section 3.3.

Both MP and LMP steam are superheated in a separate fired heater, using, normally, nitrogen wash tail gas as fuel, the temperature of superheat being selected to ensure that condensing turbines do not exhaust at more than 10% wetness. It is possible to raise these temperatures to improve the efficiency of this heat recovery. However, the HP boiler has a saturation temperature of over 620°F, and will have a considerably greater overall thermal efficiency than the fired heater. This means that generating extra HP steam is certainly more efficient than superheating LMP steam, and probably greater than superheating MP steam. If surplus clean fuel gas has a greater value than coal per B.T.U., this would reinforce the argument for not firing more than necessary in the superheater.

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*Further work on the optimisation of the steam system for the 1200 psig case has resulted in the conditions of the HP steam being changed to 1500 psig, 940°F.

3.2.11 Cont'd

The detailed development of the steam system will be carried out and optimised during the conceptual design of the Commercial Plant for the selected gasification pressure. For this study, an outline scheme has been drawn up which is adequate for comparing gasification pressures. In summary, the effects of increasing gasification pressure are:

a) More waste heat is available to be recovered.
b) This waste heat becomes available at a higher temperature, ie is more efficiently recovered.
c) Exchanger surface area is reduced by about 50% for the principal heat exchangers in going from Case A1 to Case D.
d) Wall thicknesses increase.

3.2.12 Boiler Plant

As stated in paragraph 3.2.2 above, the study has been based on high pressure (HP) steam being generated in an auxiliary boiler to deliver steam at 1800 psig and 986 °F. The boiler is fired with coal, supplemented with fuel gas surplus to the requirements of the MP and LMP steam superheater. The overall thermal efficiency of the boiler has been taken as 83% (HHV basis). The boiler duty for the different cases is given in sub-section 3.3 below.

MP and LP steam are superheated with fuel gas taken from the Nitrogen Wash tail gas, with an assumed efficiency of 68% (HHV basis).

The adoption of a lower pressure for HP steam will result in a different balance between consumptions at different steam levels, but the overall energy consumptions will not be significantly changed.

3.2.13 Sulphur Plant and Stack Gas Clean-up Unit

The Sulphur Plant consumptions and costs have been estimated for the four cases, using in-house procedures. The technology to be used for clean-up of the Sulphur Plant tail gas and Boiler Plant stack gas for the Commercial Plant is the subject of a Trade-off Study to be carried out by Ebasco during Phase I of the project. For the present

3.2.13 Cont'd

study the system adopted for the original proposal has been taken as a basis for scaling for all four cases. Overall, increasing the gasification pressure produces a minor reduction in utility consumptions of these units, but has a negligible effect on equipment sizes.

3.3 Anticipated Performance

3.3.1 Introduction

The following performance figures were calculated according to the design basis given in sub-section 3.1 above. Although there have been some changes in this design basis and other possible changes are still under review, the figures are considered to give a fair basis for comparison of plant performance with different operating pressures.

3.3.2 Overall Consumption

TABLE 3-1
RAW MATERIAL AND POWER CONSUMPTIONS

CASE	UNITS	A1	B	C	D
Coal to Gasifier	MM BTU/h	5033	5079	5169	5441
Coal to Boiler	MM BTU/h	1146	1055	999	652
Electric Power	MW	17.52	17.10	17.49	18.91
Raw Water	USGPM	6133	6009	5989	5885

TABLE 3-2
POWER REQUIREMENTS OF LARGE DRIVES
(All in MW)

CASE	A1	B	C	D
Carbon/Ash Handling	0.5	0.7	0.9	1.4
Rectisol	10.9	10.2	11.0	8.5*
Nitrogen Wash	17.7	21.8	23.4	26.8
ASU and O_2 Compression	79.5	80.5	83.6	90.5
Syngas Compression, Ammonia Synthesis and Refrigeration	56.9	47.0	41.2	28.4
Utilities	23.3	22.6	22.2	21.4
TOTAL	188.8	183.1	183.3	177.0

* Nett figure after allowing for power recovery from methanol let-down (Case D only).

3.3.2 Cont'd

TABLE 3-3
ELECTRIC POWER CONSUMPTIONS
(All in MW)

CASE	A1	B	C	D
Gasification	2.08	2.31	2.55	2.33
Carbon/Ash Handling	0.17	0.17	0.18	0.18
CO Shift	0.60	0.64	0.69	0.91
Rectisol	2.05	2.23	2.38	3.22
Nitrogen Wash	0.82	0.23	0.25	0.29
ASU/O_2 Compression	*(1.20)	(1.20)	(1.20)	(1.19)
Ammonia Synthesis	0.56	0.51	0.46	0.39
Sulphur Recovery	0.12	0.12	0.13	0.13
Coal Handling	5.23	5.20	5.20	5.16
Stack Gas Clean-Up	2.43	2.30	2.26	1.93
Utilities	3.16	3.09	3.09	3.06
Lighting, Instruments, etc	1.50	1.50	1.50	1.50
TOTAL	17.52	17.10	17.49	18.91

* There is a net surplus of electric power from the ASU. arising from the use of power recovery gas expanders.

3.3.3 Air Separation and Oxygen Compression Power Requirements

Power consumptions in paragraph 3.3.2 above are based on data from one vendor for air separation and oxygen compression (designated Vendor 'A'). A second air separation vendor has also provided data (Vendor 'B') and data has also been received from two compressor vendors (Vendors 'C' and 'D') and the results are summarised in Table 3-4.

TABLE 3-4
AIR SEPARATION POWER REQUIREMENTS
(All in MW)

VENDOR	CASE	A1	B	C	D
A	Gross Consumption, ASU	55.0	55.4	56.4	58.0
	Power Recovery, ASU	1.9	1.9	2.0	2.0
	Nett Consumption, ASU	53.1	53.5	54.4	56.0
	Consumption, Oxygen Compression	26.6	30.1	32.3	-
B	Consumption, ASU	54.5	55.1	55.9	57.6
	Consumption, Oxygen Compression	29.8	32.7	34.5	39.1
C	Consumption, Oxygen Compression	24.5	25.4	27.2	32.5
D	Consumption, Oxygen Compression	23.5	25.7	27.3	-

3.3.4 Waste Heat Steam Generation

TABLE 3-5
WASTE HEAT STEAM GENERATION

CASE	A1	B	C	D
M.P. Distribution pressure psig	580	580	580	580
temperature °F	654	654	654	654
* Production ST/hr	241	255	260	395
L.M.P. Distribution pressure psig	180	235	275	395
temperature °F	470	505	530	590
Production ST/hr	316	303	308	197

* Includes 160 ST/hr from Ammonia Synthesis Loop in each case.

3.4 Commercial Evaluation and Risk Analysis

3.4.1 Introduction

Although varying the gasification pressure affects the design of almost all the process units and support facilities, it only raises questions of commercial experience and viability in the case of a limited number of these units.

Coal gasification is a special case. The whole object of the Demonstration Plant Programme is to bring new coal gasification processes, in this case the Texaco Coal Gasification Process, to commercialisation. Consequently there is no need to review the commercial status of this process at this stage, although the risk, as a function of pressure, must be considered.

The objective of this sub-section is to assess the commercial status and to identify the commercial risks involved in implementing a project for a 3500 STD ammonia plant based on the process technology outlined in sub-section 3.2. These risks will clearly tend to reduce with time as more development work is undertaken to extend the limits of technology; the following paragraphs summarise the current (March 1978) situation.

3.4.2 Coal Gasification

The Texaco process for partial oxidation of oil and associated downstream gas washing systems have been demonstrated at 2300 psig on a pilot scale on Texaco's Montebello pilot plant. Coas gasification has been

3.4.2 Cont'd.

demonstrated at Montebello at 400 psig, and a semi-commercial coal-based plant, also designed for operation at 400 psig, has been demonstrated at Oberhausen, W. Germany. The next stage of pilot plant experimentation currently planned by Texaco will permit demonstration up to pressures not exceeding 1200 psig.

The above facts, taken together with Texaco's proven ability to design and successfully operate commercial oil gasification plants based on scale-up of pilot plant data, indicates that there are sound reasons for anticipating minimum risk in a commercial plant operating at pressures up to 1200 psig, provided the results of the 1200 psig pilot plant are available before proceeding to the definitive design.

For pressures greater than 1200 psig there will be unquantifiable risks, increasing progressively as the difference between design pressure and pilot plant operating pressure increases. It will be possible to evaluate these risks more realistically after completion of the proposed pilot plant programme.

3.4.3 Acid Gas Removal

As stated previously the Rectisol process has been taken as the method of acid gas removal for the purposes of this study.

The highest pressure so far adopted for a commercial unit is about 1100 psig used by Linde for a 1600 STD ammonia plant currently under construction in India. The highest pressure used on any scale of operation is 2260 psig on a plant operated in conjunction with Texaco's pilot oil gasifier at Montebello, Calif. In both these cases the Rectisol unit was associated with nitrogen wash, though the Montebello (nitrogen wash) unit never operated for any extended period at pressures greater than 1250 psig.

A significant factor is that Rectisol has not been operated in conjunction with the Texaco coal gasification process,

3.4.3 Cont'd

though Rectisol has been used in conjunction with the Koppers Totzek process, which produces a gas somewhat similar in composition, at Modderfontein, South Africa. The Linde view is that there is sufficient commercial basis for designing a Rectisol and nitrogen wash unit for all cases provided the concentration of trace components in the new gas can be accurately specified. For Cases Al and B there is already adequate data available; for Cases C and D some laboratory tests would be required.

From this it follows that there is a limited risk involved in proceeding with a Rectisol/nitrogen wash unit to operate at pressures up to 2300 psig, provided the composition of the feed gas can be specified precisely. However as the content of trace constituents can only be predicted with certainty from pilot plant operation under comparable conditions, the same considerations apply as for gasification, i.e. there will be a progressively increasing risk as the gasification pressure is raised above 1200 psig.

3.4.4 Air separation and Oxygen Compression

Air separation units have been erected and are being constructed with capacities of up to 2300-2400 STD (by L'Air Liquide in Europe and South Africa). The 3500 STD Ammonia Plant will require a capacity of 4800-5000 STD, so two streams of air separation involve only a slight increase of capacity compared with current experience. It follows that the availability of a three-stream unit is not in doubt, and two streams would seem to be completely feasible. Gasification pressure has no affect on this conclusion as only a slight change in capacity is involved.

Oxygen compression by centrifugal compressors is currently being carried out up to 800 psig, and in one case (at VEBA, Gelsenkirchen) to 960 psig. The casing sizes required for single train compression up to about 1100 psig are already available; however the higher pressure casings suggested by G.H.H. and Sulzer are not in use on oxygen service.

3.4.4 Cont'd

Air Products have suggested the use of centrifugal compression up to about 1000 psig, with reciprocating compressors following for cases B, C and D. This means that higher pressures come into the range of commercial experience - Sulzer have a large labyrinth-type reciprocating compressor operating at around 1550 psig, and there would seem to be little difficulty in operating at 1700 psig - i.e. suitable for Case C. The use of reciprocating compressors would certainly increase the capital cost of the installation; six would probably be required in addition to the low pressure centrifugal compression. Experience of compression up to 2800 psig is limited to small machines of about 300 scfm capacity, so there would need to be substantial development before suitably large machines become available.

The ASU vendors have preferred to have one compressor train per ASU, rather than one large common machine. This, while raising the total installed cost, will result in improved reliability and greater efficiency on turndown, particularly where there is more than one normally operating gasifier.

Clearly before proceeding to the design of a plant corresponding to Case D, and probably for Cases B and C also, it will be necessary to investigate liquid oxygen pumping and compare it with compression. We have not so far received any information from vendors of such pumps. As it was considered that this would form an integral part of the ASU three potential oxygen plant sub-contractors were invited to make recommendations for all four cases, selecting compression or liquid pumping as they felt would be most appropriate. The replies tend to favour compression for pressures within the range of cases A1, B and C, though discussions with vendors will be necessary before finalising the conceptual design of the Commercial Plant. For Case D, should this merit investigation because of advantages in other areas, a more detailed evaluation of compression versus pumping would be necessary before any recommendation could be made.

Low-pressure cycle air separation units with reversing exchangers and capacities up to 2200 STD have proved reliable, and with proper operation the use of two or

3.4.4 Cont'd.

three such units will entail low risks. The main potential hazard is the accumulation of hydrocarbons (particularly acetylene) in the liquid oxygen at the base of the L. P. column. The general solution to this is to adsorb these hydrocarbons from the liquid and also to ensure that there is a small purge of liquid from the column base. Obviously liquid oxygen pumping removes the need for this purge.

Recent bad experiences with oxygen compression to high pressures have led to a reluctance to use centrifugal compressors to more than about 800 psig. However this does not mean that compression should be ruled out.

There is a machine in Germany operating at 960 psig and European manufacturers are confidently offering machines up to 1100 psig. Above this pressure less is known about the burning properties of metals, although it is known that safety margins reduce above this pressure. For these duties materials such as silver, bronze and stainless steels are recommended, and with careful design it should be possible to demonstrate that centrifugal compressors are safe up to 1400 psig (for Case B) and possibly 1700 psig (Case C). However at the present time the use of centrifugal compression for 2700 psig discharge (Case D) cannot be recommended.

Because of lower speeds, and hence lower potential metal temperatures in case of contact, reciprocating compressors have an intrinsic advantage and if large capacities are required at pressures around 2700 psig it is probable that this type of machine would offer safer designs in the nearer future than centrifugal compressors.

3.4.5 CO Conversion

The proposed system converts the carbon monoxide in undesulphurised gas by the use of cobalt molybdenum sulphide. This type of catalyst is well proved by BASF (Type K8-11) both in experimental work up to 1400 psig and in commercial operation to over 1100 psig.

3.4.5 Cont'd

Other catalyst vendors are offering similar catalysts but with less experience. Girdler, for instance, offer their type G-93 which has been tested up to 1000 psig.

Therefore cases A1 and B should present no problem and Case C involves little extension of commercial experience, and already there should be enough experimental data to allow this duty to be selected with confidence. Case D, on the other hand, does require an extension of current experience, and will certainly require some development work to determine

a) whether the catalyst can withstand the higher pressure and partial pressure of steam,

b) how much catalyst will be required,

c) whether any undesirable side reactions become significant.

The opinion of catalyst vendors is that no problems are expected for a) and c). They differ however on b) and it is this aspect which will probably require most investigation.

From the foregoing it can be concluded that cases A1 and B can be specified with certainty, case C would probably be low risk if a conservative margin on catalyst quantity were allowed. Case D would involve moderate risk unless further development work were undertaken.

3.4.6 Conclusion.

On the assumption that the pilot plant programme involving gasification at 1200 psig is carried out according to the schedule currently envisaged, it appears that there is an acceptable risk in proceeding with a commercial plant operating at a pressure up to 1200 psig. Further, by the time this pilot plant programme has been completed, the range of acceptable risk might be extended to about 1500 psig. It seems unlikely that the risk would be acceptable at pressures much greater than 1500 psig, but this conclusion must be subjected to continuous review as Phase I of the project proceeds.

SECTION 4

ECONOMICS

CONTENTS

SECTION 4 - ECONOMICS

4.1 Capital Cost Analysis

4.1.1 Basis of Estimate

The capital cost estimates developed for this study have been carried out with a view to establishing valid comparative costs using engineering data from a variety of sources.

This is augmented with budget quotations from vendors for major items of equipment, and in some cases, complete equipment packages, such as the Air Separation Unit and the Rectisol and Nitrogen Wash Unit. Finally, use has been made of Ebasco's estimates for their Proposal to ERDA of May 1976 for certain areas of the plant.

The above sources are on a number of different bases, e.g. erected Gulf Coast U.S.A. (Air Products) erected in France (Air Liquide) and f.o.b. W. German Port (Linde). As the composite estimate has been compiled from process data generated within Humphreys and Glasgow and using H & G in-house estimating methods, it was decided that the consistent basis for comparison would be that of erection in the United Kingdom, and all externally supplied information was corrected to this basis. The figures are then expressed in U.S. Dollars using a currency conversion of $1.90 = £1.00.

The estimating method used reflects the need for comparative data rather than absolute costs. The engineering schedules and costing have been done using in-house data and methods based on similar projects carried out by H & G.

4.1.2 Summary of Capital Costs

The breakdown of capital costs into the various areas of the ammonia plant, plus support facilities, is given in Table 4-1 on the following page, and presented graphically in Figure 4-1.

4.1.3 Analysis of Estimate

The trends shown by the capital cost curves are discussed in Section 2, Conclusions. Individual process sections behave differently with the effect of changing pressure. In general solidshandling sections at the front end of the plant are not sensitive to pressure as the amount of coal feedstock does not change markedly. Pressure sensitive areas are those primarily where the process fluid is in the gaseous phase,

(Cont'd on page 4-4)

TABLE 4-1

BUDGET CAPITAL COST ESTIMATE FOR 3,500 STD AMMONIA

Estimate Base Date : January, 1978

All Figures Quoted are Thousands of U.S. Dollars

CASE	A1	B	C	D
Coal Preparation	12,670	12,510	12,500	12,390
Coal Gasification	20,330	18,900	19,860	20,430
Ash & Carbon System	5,510	5,970	6,380	7,700
CO Shift	18,500	19,760	21,090	25,330
Rectisol	51,810	43,470	47,580	79,900
Ammonia Synthesis	41,380	38,590	37,630	36,190
Nitrogen Wash	10,110	10,180	13,070	21,450
Air Separation (Vendor A)	47,500	48,500	49,400	50,400
Boiler	35,670	33,935	33,290	30,860
Utilities	33,000	32,700	32,490	32,260
Sulphur Recovery	2,660	2,660	2,660	2,660
TOTALS	279,140	267,175	275,950	319,570

Notes to Table

1. Costs for Air Separation quoted by other vendors are as follows:

CASE	A1	B	C	D
Vendor B	63,270	66,880	68,020	70,300
Vendor E	72,200	76,950	79,800	83,600

2. All costs are in \$U.S. based on \$1.90 = £1 Sterling
Other currency coversions used:

9.00 French Francs = 4.0 German DM = £1 Sterling

Notes to Table 4-1 (Cont'd)

3. All prices are at January 1978, turnkey U.K., and are Budget only.

4. The following items are excluded:

   forward escalation;
   contingencies;
   contractor's fee;
   insurance costs;
   Texaco licence fee;
   customs fees, taxes, and import duties;
   land costs, site preparation and infrastructure

e.g. gasification, CO shift, acid gas removal, oxygen compression and synthesis gas compression.

The accuracy of the overall cost estimate is higher for the lower pressure cases as the figures relate to knowledge of existing plant having similar capacity and performance characteristics. This is particularly true for the Rectisol and Nitrogen Wash Units where large extrapolations have been made to generate capital costs for the 2500 psig pressure case. The argument does not, however, apply to the gasification and CO shift systems where an accurate specification of the equipment can be made leading to cost data of good accuracy. The ammonia synthesis loop cost is almost independent of gasification pressure, although the cost of the synthesis gas compressor and driver reduce with higher pressure, and also become potentially more reliable due to simpler installations.

The conclusions from analysis of the capital cost estimate is that the comparative figures show a minimum cost at about 1200 psig, which is well within the 500 to 1500 psig pressure range where the absolute cost data is of good accuracy.

4.2 Operating Cost

Table 4-2 overleaf gives the operating costs for five pressure cases, together with the capital charges calculated on the basis given below. Also included in the table is the sum of the operating cost and capital charges, which is the ammonia production cost.

The basis of the calculation is as follows, the notes below corresponding with the references in Column 2 of the table.

(a) Ammonia production is taken as 330 days at 3500 STD, i.e. 1,155,000 ST per annum.

(b) The price of coal for gasification and boiler fuel is taken as $20 per ST. The coal is assumed to contain 10 per cent moisture as recieved and to have a HHV of 13,368 Btu/lb on a dry basis; this is equivalent to a price of $0.8312 per million Btu on an HHV basis.

(c) Electric power is purchased at an average price of $0.0185 per kwh.

(d) Raw water is priced at $0.10 per thousand U.S. gallons.

(e) Catalysts and chemicals costs are calculated from manufacturers recommended life and consumptions.

4.2 (f) Labour costs are based on 75 operators each costing $20,000 per year.

(g) Supervision costs are based on 20 men each costing $22,500 per year.

(h) Maintenance costs are taken as 4% per annum of the total capital cost given in the bottom line of Table 4-1.

(i) Variable overhead is taken as:

- direct: 30% of labour plus supervision
- indirect: 65% of labour plus supervision plus maintenance.

(j) Byproduct sulphur is given a credit of $50/ST.

(k) Depreciation of both onsite and offsite facilities is taken as 15 year straight line, i.e. 6.67% per annum of the total capital costs quoted in Table 4-1.

(l) "Finance Charges" is a single figure of 25% of installed capital cost per annum to cover all charges related to capital cost not included under any of the previous headings, e.g. interest on loans during construction and operation, and return on equity capital invested.

The trends apparent from the figures of Table 4-2 are presented graphically in Figure 4-2, whilst the breakdown of energy consumption, which is responsible for the major portion of operating costs, can be seen by reference to Figures 4-3 and 4-4. Inspection of the figures shows that total operating cost is at a minimum at 1200 psig gasification pressure, with the values at 500 and 2500 psig being each about 5 to 6% higher than this minimum.

As the individual curves for total operating costs and for capital charges each show the same shape, the combination of these two parameters further emphasises the trend for minimum ammonia production cost to correspond with the 1200 psig gasification pressure case. Production cost is about 9.5% higher than the minimum at 500 psig and about 13.6% higher at 2500 psig.

TABLE 4-2

PRODUCTION COSTS IN $U.S. PER SHORT TON AMMONIA

(For basis of calculations please refer to Paragraph 4.2**)

CASE	**Note Ref	A2	A1	B	C	D
Raw Materials & Utilities						
Gasification Coal	(b)	28.40	28.69	28.95	29.46	31.01
Boiler Coal	(b)	7.61	6.53	6.01	5.69	3.72
Electric Power	(c)	2.17	2.22	2.17	2.22	2.40
Raw Water	(d)	0.26	0.25	0.25	0.25	0.24
Catalysts & Chemicals	(e)	0.59	0.55	0.51	0.49	0.45
Operations						
Labour	(f)	1.30	1.30	1.30	1.30	1.30
Supervision	(g)	0.39	0.39	0.39	0.39	0.39
Maintenance	(h)	10.39*	9.67	9.25	9.56	11.07
Variable Overhead						
Direct	(i)	0.51	0.51	0.51	0.51	0.51
Indirect	(i)	7.85*	7.38	7.11	7.31	8.29
Byproduct Credit						
Sulphur	(j)	(2.63)	(2.63)	(2.62)	(2.61)	(2.60)
Total Operating Cost		56.84*	54.86	53.83	54.57	56.78
Capital Charges						
Depreciation	(a) (k)	17.32*	16.11	15.42	15.93	18.45
Finance Charges	(a) (l)	64.94*	60.42	57.83	59.73	69.17
Total Capital Charges		82.26*	76.53	73.25	75.66	87.62
Ammonia Production Cost		139.10*	131.39	127.08	130.23	144.40

* Based on an extrapolated installed capital cost of $300 million.

4.3 Economics of Operation

The cost and performance curves indicate a theoretical situation. This must be accompanied in practice by high plant on-stream availability and on-stream flowsheet production of ammonia on an annual basis if the overall actual marketing forecast requirement is to be achieved, yielding profitable production.

The chance of obtaining high availability is greater for plant operation at lower pressure because the process units have been demonstrated on a commercial scale, but with poorer economics than can be obtained by raising the pressure. (Refer to the discussion in Section 3.4 for details of unit scale which have been proven).

The factor for risk in relation to reliability can be set at 1.0 for gasifier operating pressure up to 1200 psig, and trouble-free performance at this pressure can be demonstrated later on the pilot plant scale. The risk factor for plant availability becomes proportionately higher as the pressure increases, and each process unit moves further into the unproven area. From the data presented there is no justification for taking these greater risks because performance and cost do not show improvement. The question for examination is only whether the best pressure is exactly 1200 psig, or at some relatively small variation either way from this figure.

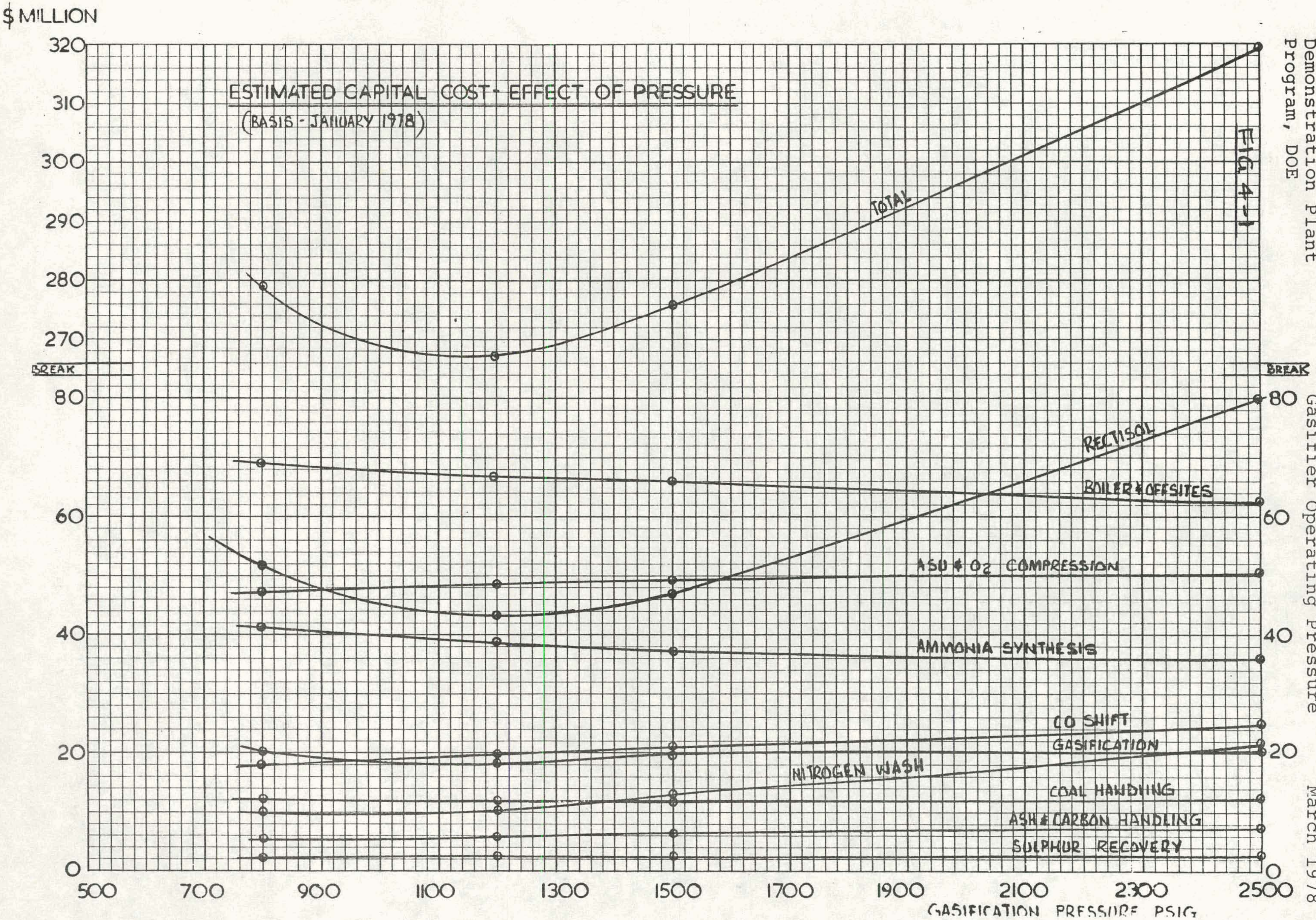
Breakpoints in technology have been sought for process units and equipment hardware. The following are relevant:

- The number of Texaco gasifiers required for economic throughput increases by one when the pressure is reduced below 1100 psig, indicating a benefit from a higher pressure. Note that each gasifier is intimately associated with its own slag lock hopper and soot scrubber separator, hence effecting a major reduction in equipment items at pressures above 1100 psig, which also raises the reliability factor.

- The CO shift system exhibits a breakpoint above about 1100 psig due to a reduction in the catalyst vessel diameters which should permit the CO conversion to take place in only two vessels arranged in parallel, each containing one first stage and one second stage bed. Lower pressures would require more vessels and beds because of overall shell size limitations. Also, the shift catalyst has been proved to 1100 psig and experimentally to 1400 psig; further increases in pressure involve extrapolation of performance.

- The Acid Gas Removal Unit shows a breakpoint above the 1200 psig case, when the number of wash columns required increases by a factor of two.
- The Nitrogen Wash Unit has a sharply increasing cost breakpoint at a feed gas inlet pressure corresponding to gasification at 1200 psig, due to different heat exchanger configuration.
- The Air Separation Unit shows a breakpoint somewhere above the 1200 psig case, where an additional stage of centrifugal compression for oxygen is required.
- The Ammonia Synthesis Unit is benefited by a gasification pressure which reduces the number of casings for the synthesis gas compressor. Three casings are required at 1000 psig, and only two for the 1200 psig case.
- Piping and valve pressure/temperature ratings and associated costs at different plant presssure alternatives are being investigated. Further study is required to establish the overall economic effect of passing from one ANSI rating to the next, and this work is being continued.

[redacted] [redacted]



ESTIMATED CAPITAL COST- EFFECT OF PRESSURE
(BASIS - JANUARY 1978)

FIG 4-1

FIG. 4-2

OPERATING & PRODUCTION COST BREAKDOWN

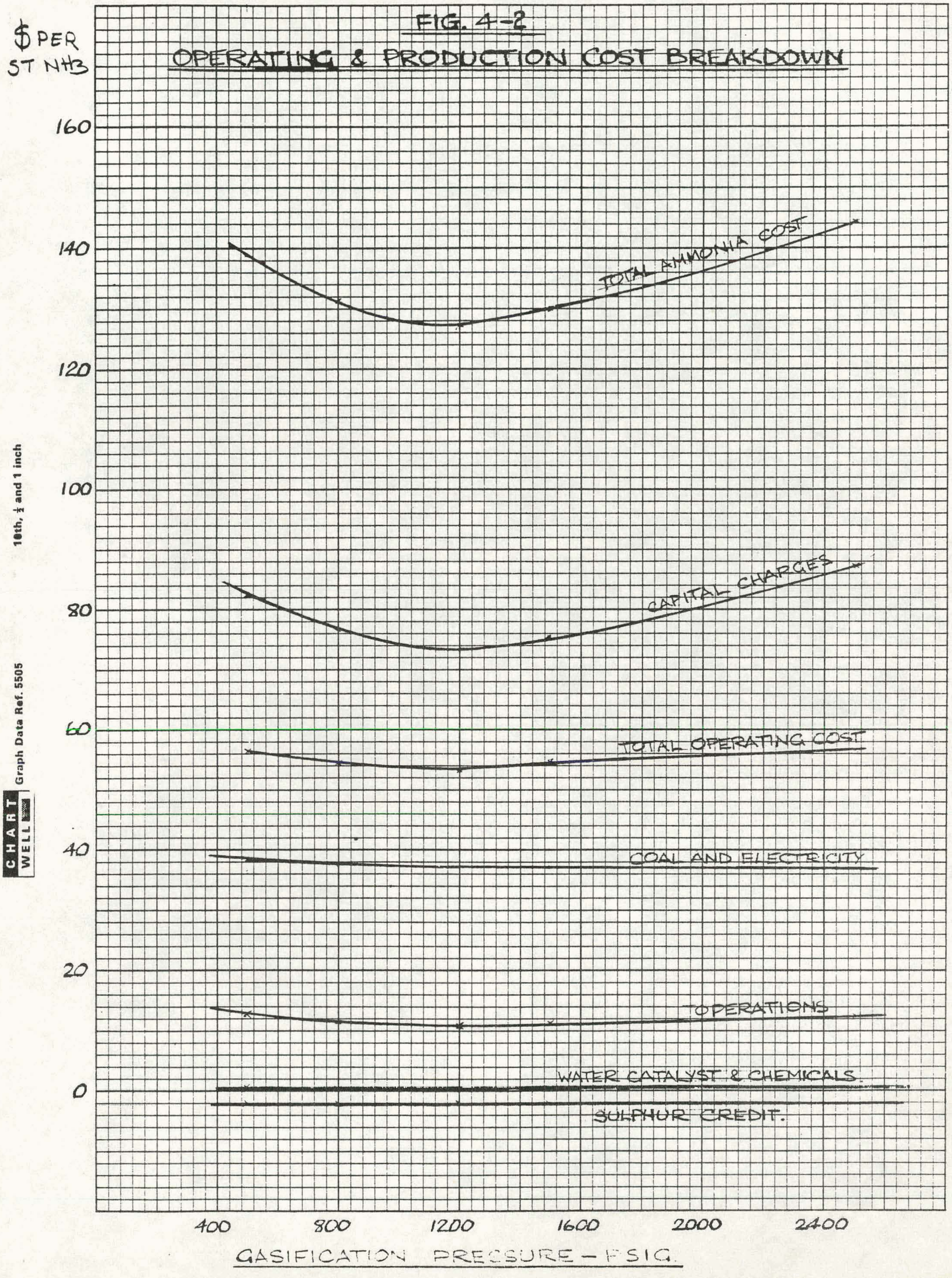


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FIG. 4-3
ENERGY CONSUMPTION
INCLUDING ESTIMATE OF 500 PSI CASE

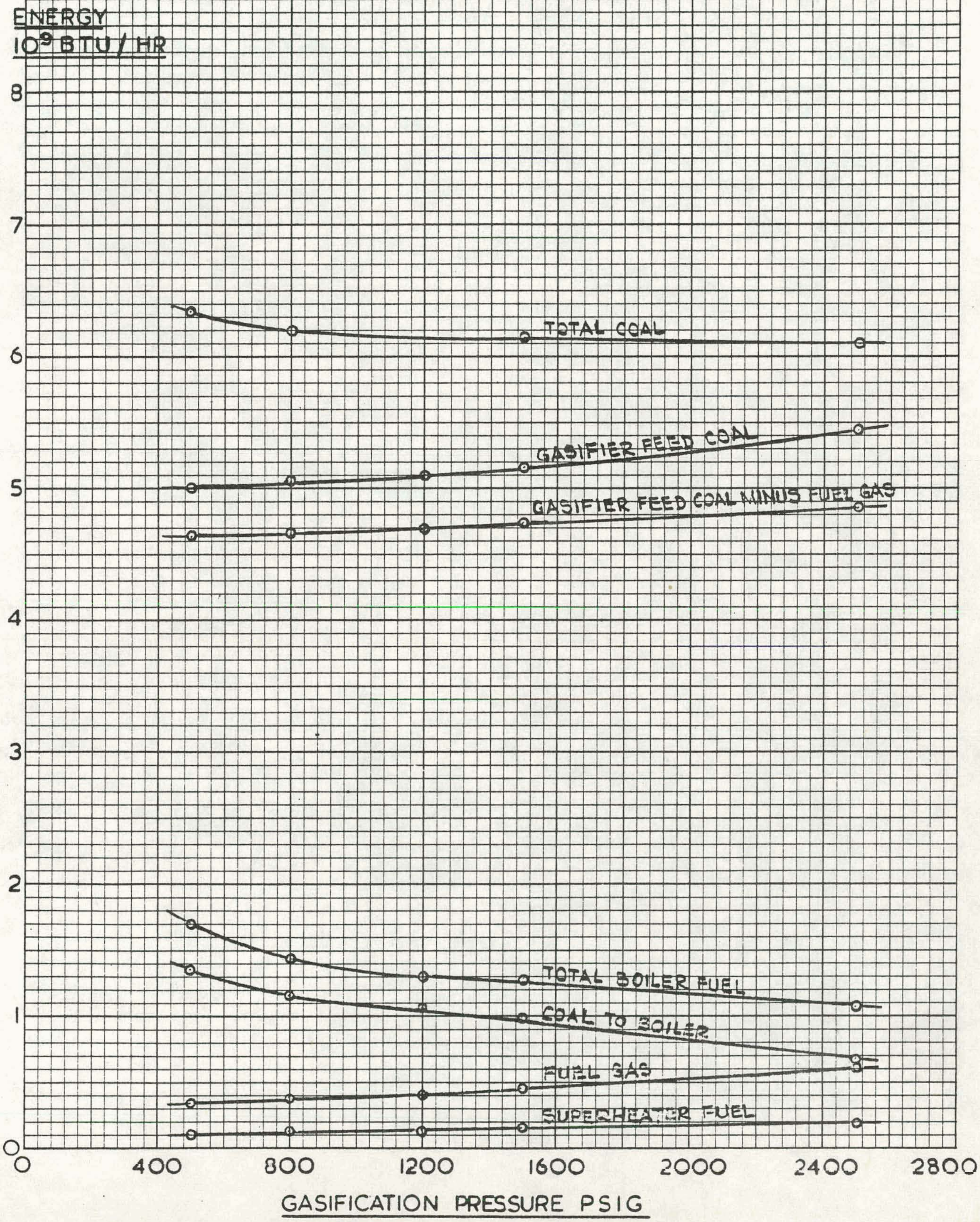
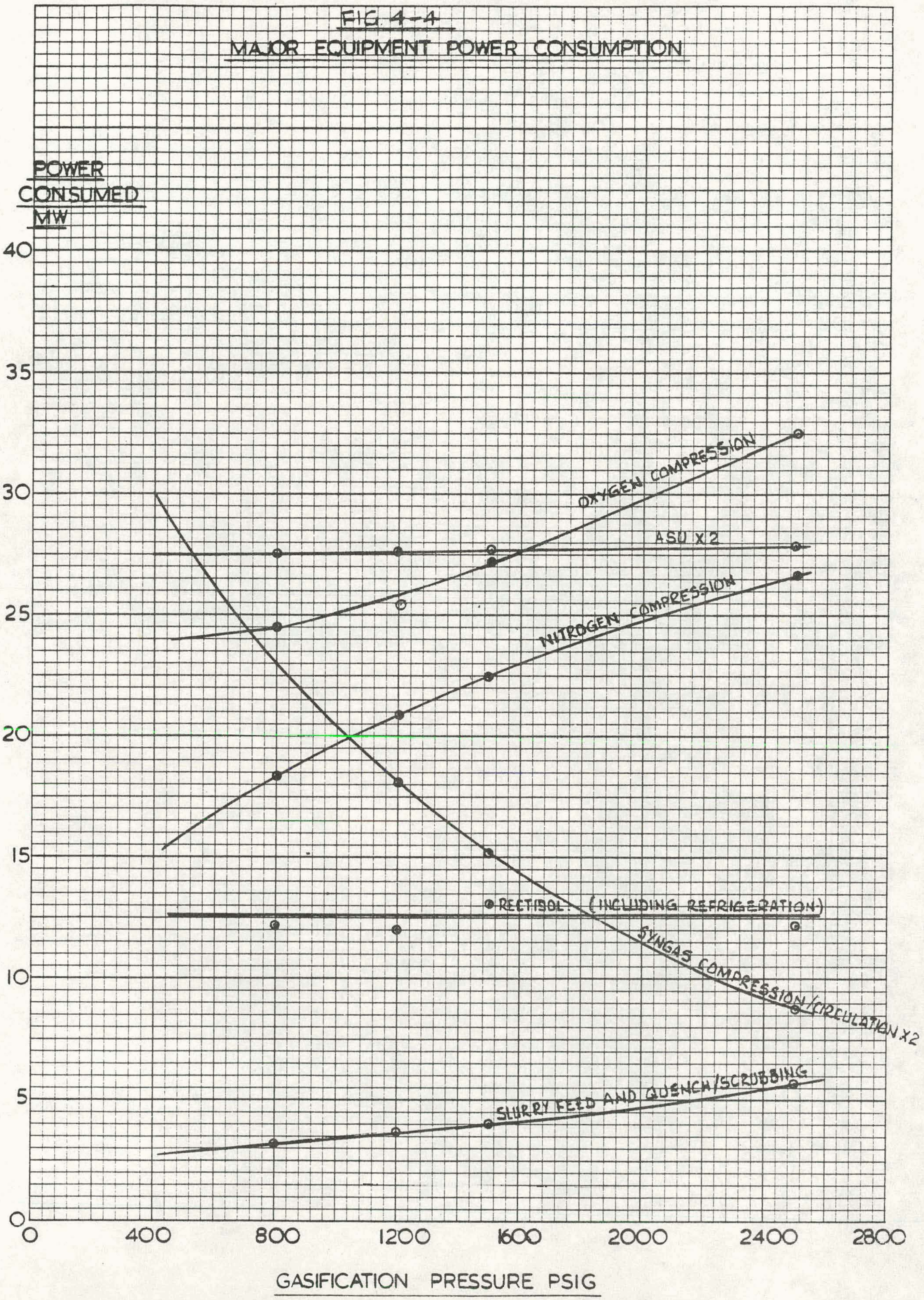


FIG. 4-4

MAJOR EQUIPMENT POWER CONSUMPTION



(H&G. DRG. No. 1821-X52-5)

* EQUIPMENT INCLUDED IN THESE SECTIONS:-

COAL GASIFICATION.
- GASIFIER/QUENCH VESSEL.
- LOCK HOPPER.
- SCRUBBER SEPARATOR.
- SCRUBBER CIRCULATION PUMP.
- SLURRY CHARGE PUMP.

SLAG SCREENING.
- SLAG SCREEN.
- SLAG SUMP.
- SLAG SUMP PUMP.

SOOT WATER HANDLING.
- CARBON WATER EXCHANGER.
- CARBON WATER COOLER.
- SOUR GAS COOLER.
- FLASH DRUM.

CARBON RECOVERY
- SETTLER.
- SOUR CONDENSATE K.O. DRUM.
- SURGE DRUM.
- SOUR CONDENSATE PUMP.
- CARBON SLURRY PUMP.
- QUENCH WATER PUMP.
- WASTE WATER PUMP.

U.S. DEPARTMENT OF ENERGY

W. R. GRACE & CO.
MEMPHIS, TENNESSEE

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
INDUSTRIAL PROJECT "B" - PHASE I
CONTRACT NO. ET-77-C-01-2577

PROCESS BLOCK DIAGRAM

EBASCO SERVICES INCORPORATED

HUMPHREYS & GLASGOW, LTD.

SCALE NONE
APPROVED
DATE 17th March 1978
SECT. CODE NO.
DIV
DR. S. TALLENT
CH.
EBASCO WRG
ISSUE 1

SYNTHESIS GAS DEMONSTRATION
PLANT PROGRAM

TRADE-OFF STUDY - II
GAS PURIFICATION ALTERNATIVES

SEPTEMBER 1978

PREPARED FOR
EBASCO SERVICES INC.
BY
HUMPHREYS AND GLASGOW LTD.

TRADE-OFF STUDY II - GAS PURIFICATION ALTERNATIVES

FINAL REPORT

INTRODUCTION

Trade-Off Study II incorporates proprietary information. The presentation of data has been divided into two volumes.

1) A Final Report attached, which contains the conclusions and recommendations together with a non-proprietary version of the data and process information.

2) The Proprietary Data Volume, which contains the raw data, cost estimates and records of discussions involving the proprietary processes evaluated.

TABLE OF CONTENTS

SECTION 1

INTRODUCTION

CONTENTS

SECTION 1

INTRODUCTION

1.1 OBJECTIVE

The objective of Trade-Off Study II is to define the gas purification system for a Commercial Plant to produce 3500 STD ammonia from coal, using the Texaco Coal Gasification Process operating at about 1200 psig. The system chosen shall also be applicable to a Demonstration Plant designed to produce a synthesis gas for subsequent conversion to ammonia at a production rate equivalent to 1200 STD ammonia, also operating at a gasifier pressure of about 1200 psig. The results are based on an overall optimization of capital and operating costs, supported by the applicability and proved reliability of the process concerned.

1.2 SCOPE

1.2.1 The feed gas to the purification system contains H_2, N_2, CO, Ar, CO_2, H_2S, COS and trace constituents; there is also available a source of high pressure N_2. The purpose of the unit is to separate the gas components so that the following streams result:

a) Ammonia synthesis gas: $H_2 + N_2$
b) Vented gas: CO_2
c) Sour gas: H_2S, plus any undecomposed COS
d) Fuel gas

The following design parameters apply:

i. For technical reasons the synthesis gas a) must not contain CO, CO_2, H_2S, COS or H_2O.
ii. For environmental reasons the vented gas b) should not contain H_2S or COS.
iii. Similarly, the fuel gas d) should not contain H_2S or COS.

iv. It is wasteful if Ar or CH_4 go to a), or if CO or CH_4 go to b) or c).

v. Costs are increased if the H_2S in c) is less than about 25%.

vi. Hydrogen leaving in the vented gas and sour gas streams loses all potential as feedstock or fuel; hydrogen recovered in the fuel gas stream is degraded from its higher potential as feed stock for synthesis gas but retains its heating value.

vii. Processes were evaluated to ensure that any conclusion reached would not be invalidated by emission control requirements for CO and CH_4 which might be applicable later in the program.

1.2.2 It is not feasible to produce the purified synthesis gas and to control the other gas streams with a single process. Two processes are required: an Acid Gas Removal process to remove the sulfur gases and CO_2 from the gas, followed by a final clean-up to produce the synthesis gas.

1.2.3 The Acid Gas Removal system must absorb CO_2, H_2S and COS from the feed gas, and allow separation of the CO_2 from the other two gases. This can be done by selective absorption, selective regeneration, or both. The primary object of producing a purified ammonia synthesis gas can be achieved by many processes; the need for separating CO_2 from H_2S and CO_2 from COS eliminates the majority of these, unless the uneconomic course is adopted of using two different processes in series. The high partial pressure of acid gases and the presence of COS both preclude use of chemisorption (ethanolamine solutions for example), and thus a selective, physical absorption process is indicated.Gases produced from coal are always liable to contain trace impurities which can lead to

1.2.3 (cont'd)

operating problems, such as foaming or sludge formation.

It is therefore valuable to know if a particular acid gas removal process has been successfully applied for several years to the treatment of "dirty" gas.

The process chosen must also integrate well with the final clean-up unit.

To obtain an objective comparison it is necessary to use data on competing processes obtained from competing chemical plant contractors.

1.2.4 The processes available for the final clean-up duty are adsorption, copper liquor, methanation and nitrogen wash. The fact that the gas from the first process will be cold (and free of condensable components) and the necessity of adding nitrogen at this point reduces the economic applicability of the first three named processes. The remaining process, nitrogen wash, integrates well with the other process systems in the overall plant and is specified for its moderate investment cost and good efficiency of operation.

1.2.5 The Rectisol and Selexol processes were chosen for Acid Gas Removal, as fulfilling all requirements.

The designs for this study were provided by Lotepro Corp./ Linde AG (Rectisol and Nitrogen Wash), and by Air Products and Chemicals Inc./Allied Chemical (Selexol and Nitrogen Wash). Only these organisations were able to carry out the work within the time and budget desired. It had been hoped that a competing design could be obtained from Air Liquide, but this did not prove possible.

SECTION 2

CONCLUSIONS AND RECOMMENDATIONS

CONTENTS

SECTION 2

CONCLUSIONS AND RECOMMENDATIONS

2.1 CONCLUSIONS

2.1.1 Gas purification must be carried out in two stages for economic and technical reasons.

2.1.2 The requirements of the first stage can be met by a Rectisol plant by Lotepro/Linde AG, or by a Selexol unit by Air Products/Allied Chemical.

2.1.3 The second stage in either case should be a Nitrogen Wash unit.

2.1.4 The Rectisol solvent (methanol) is freely available worldwide. Selexol solvent is obtainable only from single sources in the U.S.A. and in Germany, but serious supply problems are not anticipated.

2.1.5 Selexol solvent is relatively non-toxic, non-volatile and non-flammable. Although methanol is classified as toxic and flammable, these potential hazards are controlled to acceptable levels of safety by application of standard methods for plant design and operational handling.

2.1.6 The reduction of impurities (other than sulfur) in the CO_2 stream to vent does not form part of this trade-off study. However, a qualitative study of feasibility has been made, with the following findings:

i. Solvent vapor in this stream can be reduced to an extremely low level by a water scrub. The polluted water can be used in the gasifiers, so that no liquid effluent is produced.

2.1 CONCLUSIONS (Cont'd)

2.1.6 (cont'd)

ii. A large percentage reduction in the CO and CH_4 going to vent is achievable at the cost of extra equipment and increased operating cost.

iii. It is not economically feasible to treat the CO_2 vent stream to remove CO and CH_4 to a very low level. The design offered removes CO and CH_4 to currently accepted emission standards. Both investment and operating costs increase at a disproportionate rate if CO and CH_4 vented volumes are limited further. Also, no significant improvement in environmental performance would result from changing the design basis.

2.1.7 As will be seen from Section 4, the alternative gas purification systems have almost identical installed plant cost.

2.1.8 As will also be seen from Section 4, the operating cost of the Selexol system is substantially higher than that for the Rectisol version. This difference is large compared to any advantages of the Selexol process enumerated above.

2.2 RECOMMENDATIONS

2.2.1 The Rectisol plant offered by Lotepro/Linde AG, plus a Nitrogen Wash unit, should be specified for the Commercial Plant.

2.2.2 This recommendation applies also to the Demonstration Plant.

2.2.3 Before any such plant is ordered the process requirements (particularly concerning effluents) should be defined as exactly as possible.

SECTION 3

TECHNICAL EVALUATION

CONTENTS

SECTION 3

TECHNICAL EVALUATION

3.1 DESIGN BASIS

3.1.1 Introduction

The data given below have been used as the design basis for Trade-Off Study II; they are taken from the appropriate "Statement of Work". Some of this data may be amended as more information becomes available from subsequent studies; the modifications are not expected to be so large as to invalidate the conclusions.

3.1.2 Feed Gas

Composition, mol%:

CO	3.0
H_2	49.19
N_2	0.41
Ar	0.12
CO_2	45.45
CH_4	0.75
H_2S	1.08
	100.0

The feed gas also contains trace constituents:

COS	60	ppm v/v
NH_3	0.3	" "
C_2 + hydrocarbons	2	" "

3.1 DESIGN BASIS (Cont'd)

3.1.2 Feed Gas (cont'd)

The gas is available at 1024 psia and 104°F, saturated with water vapor, and in quantity sufficient to produce 3500 short tons per day of ammonia.

3.1.3 Product Gas

The product gas is to be of suitable quality for ammonia synthesis, with total oxygen compounds not over 5 ppm v/v; the pressure not less than 974 psia. The quantity of H_2 and N_2 is to be 34,400 lb. mol/hr, in the ratio of 3.00:1.

3.1.4 Byproduct Gas Streams

a) CO_2 to be vented:
H_2S 5 ppm v/v max.
H_2S + COS 10 ppm v/v max.

b) Sour gas to sulfur recovery:

H_2S	25% v/v min.
Pressure	25 psia min.

c) Nitrogen wash tail gas;
Composition, etc., to be stated by the vendor of the gas purification plant.

3.1.5 Utilities

The following are available, in addition to nitrogen:

Steam at 50 psig saturated
" " 230 " and 540°F
" " 240 " saturated
" " 580 " and 730°F
" " 600 " saturated
" " 1500 " and 940°F

Cooling water at 89°F max., temp. rise 20°F;B.L. pressure 55 psig, prssure drop 20 psi max. Small changes in the

3.1 DESIGN BASIS (Cont'd)

3.1.5 Utilities (cont'd)

parameters would not alter the findings of this study.

Electricity:	below 1	h.p. motors	110 v.	1 - phase	60 Hz
	1 to 250	" "	440 v.	3 "	"
	over 250	" "	4,160 v.	3 "	"

Process water: raw, clarified and deaerated, or demineralized.

Instrument air.

(Refrigeration may be regarded as a utility for convenience).

3.1.6 Effluents

TOS II is concerned solely with the process requirements listed above. The plant as-built will be required to conform with the regulations then in force governing effluents, etc., and where appropriate, consideration has been given to the practicality of incorporating equipment to further limit gaseous and liquid emissions.

3.2 PROCESS DESCRIPTIONS, MASS BALANCES, FLOWSHEETS:

3.2.1 Rectisol System

a) Process Description

i. Rectisol Unit

(Refer to Process Schematic on page 3-12). Owing to the large feed gas flow at the inlet to the Rectisol plant section the gas is treated in two streams, in order to limit the necessary dimensions of the washing towers T410 and the stripping columns T411.

The two stream design basis is used also for the heat exchangers and separators which are associated with these columns. This avoids problems resulting from maldistribution. The feed gas is supplied to the battery limits of the Rectisol plant at 1024 psia.

After injection of methanol to prevent ice and hydrate formation the gas is cooled down in the heat exchangers E400 against tailgas and in E402 against cold syngas.

After separation of the condensed methanol-water-mixture in D400, the feed gas is fed to the wash tower T410, to be washed by methanol. In the bottom section H_2S and COS are absorbed down to 1 ppm. In order to minimize the heat of solution in this section, a split stream of CO_2-loaded methanol from the CO_2-wash section of the absorber is used for sulfur removal. In the upper part of the absorber CO_2 is removed down to 20 ppm. As the CO_2 solubility in methanol is less than the solubility of H_2S, the methanol circulation in the CO_2 section is

3.2.1 Rectisol System (cont'd)

is greater than in the H_2S section. The methanol surplus from the CO_2 section of the absorber is branched-off from the middle of the column. The heat of solution in the CO_2 absorption section is compensated by cooling the methanol in the coolers E410 and E411 against cold methanol from T411 and against refrigerant at -31°F.

The purified gas leaving the top of the absorber is sent to the adsorber station of the N_2-wash unit.

A certain part of synthesis gas coming from the Nitrogen Wash Unit is routed back to the Rectisol wash, where it is warmed up in the heat exchangers E412 and E402.

The H_2S loaded methanol from the bottom of the wash tower T410 is flashed in D410. Most of the H_2 and CO dissolved together with the H_2S and CO_2 is flashed off, and the remaining methanol stream is flashed into the middle of the column T411.

The CO_2 loaded methanol stream from the CO_2 washing part of the wash columns T410 is flashed into D411 and then finally flashed into the top of the column T411. Due to CO_2 flashing the temperature of the methanol decreases.

The flash gas leaving the vessels D410 and D411, mainly H_2 and CO_2, is recycled via recycle compressor C400 and water cooler E401, to the feed gas before heat exchanger E400.

To get the required concentration of H_2S in the H_2S fraction, CO_2 is stripped in the lower section of the columns T411 by nitrogen, and to prevent the stripping of H_2S in this column, the H_2S is absorbed by sulphur-free CO_2-loaded methanol from D411 in the upper section.

In order to get a better CO_2 stripping, the H_2S-loaded methanol from the H_2S absorption part of the columns T411 is warmed up in the heat exchangers E413 and E411, and thereby a rather large amount of the abosrbed CO_2 is flashed out.

The methanol from the bottom of the columns T411 enriched with H_2S in solution, is pumped via heat exchangers E432, E433 where it is warmed up against warm regerated methanol, into the regeneration column T430, where the complete stripping of H_2S and CO_2 is effected by methanol vapors.

The regenerated methanol leaving the bottom of the column T430 is cooled down in the heat exchangers E433, E432, E431, E430 and E413 against loaded methanol refrigerant at -31°F. and tail gas. The methanol drum D430 serves as storage for reduced holdup of the columns during partial load and as storage for feed to pump P430.

The H_2S-rich stream leaving the top of the regeneration column T430 is cooled with cooling water and condensate is removed in separator D431. The gas is further cooled in heat exchangers E436 and

and E437 to a temperature of -22°F against cold and refrigerant. Most of the methanol is condensed from the gas and returned to the top of the regeneration column T430.

The tail gas from the top of the stripping columns T411 is saturated with methanol. In order to satisfy the air pollution requirements concerning methanol content of vented gases, the methanol-containing tailgas is washed with water.

The water/methanol mixture from the bottom of columns T431 is routed in the water stream to the gasifier.

The methanol/water mixture withdrawn from the separator D400 is warmed up in heat exchanger E439 and is sent to the methanol/water separation section.

The flashed gas from separator D432 is mixed with the H_2S-rich gas leaving the top of the regeneration column T430 for further treatment, whereas the liquid from the separator D432 is fed to the column T432, in which the mixture is separated into methanol and water. Methanol vapor leaving the top of T432 is condensed in heat exchanger E439 and routed to column T430. Methanol from the bottom of the regeneration column T430 serves as reflux to column T432.

ii. Nitrogen Wash

(Refer to Process Schematic on page 3-13)
Methanol and CO_2 are removed from the feed gas in inter-changeable adsorbers, to prevent solidification of these components in the cryogenic section.

A heat exchanger is used to cool gaseous nitrogen to the same temperature as the feed gas, by reheating the tail gas and a split stream of the purified synthesis gas.

Feed gas and nitrogen are cooled to the nitrogen wash temperature in a second heat exchanger where the heat recovered is used to vaporize tail gas and reheat synthesis gas.

Liquified nitrogen in the wash column dissolves CO, Ar and CH_4 from the feed gas.

Liquified nitrogen is blended into the synthesis gas to establish the stoichiometric ratio of 3 H_2 for each N_2.

Refrigeration losses due to non-ideal heat exchange and insulation are compensated for by expanding the nitrogen to its partial pressure in the synthesis gas and in the tail gas.

This unit can be designed to operate without precooling the feed gas in a Rectisol unit.

3.2.1. Rectisol System (cont'd)

b) Mass Balance

Stream numbers refer to schematics on pages 3-12 and 3-13

	Stream 1		Stream 2		Stream 3		Stream 4	
	Feed Gas To Rectisol		Stripping Nitrogen		Sour Gas To Claus Kiln		CO_2 To Vent	
	lb mol/hr	mol %	lb mol/hr	mol %	lb mol/hr	mol %	lb mol/hr	mol %
COS	3,3	(60 ppm)			3.2	0.22	0.1	(5 ppm)
CO_2	24,134.2	45.45			832.3	57.97	23,301.3	95.79
H_2S	573.5	1.08			573.4	39.94	0.1	(5 ppm)
N_2	217.3	0.41	786.7	100.00	25.4	1.77	766.2	3.15
Ar	66.4	0.12					5.4	0.01
H_2	26,122.3	49.19			1.1	0.08	155.7	0.64
CO	1,593.1	3.00			0.1	0.01	58.4	0.24
CH_4	398.5	0.75			0.1	0.01	41.4	0.17
	53,108.6	100.00	786.7	100.00	1435.6	100.00	24,328.6	100.00

3.2.1. Rectisol System (cont'd)

b) Mass Balance (cont'd)

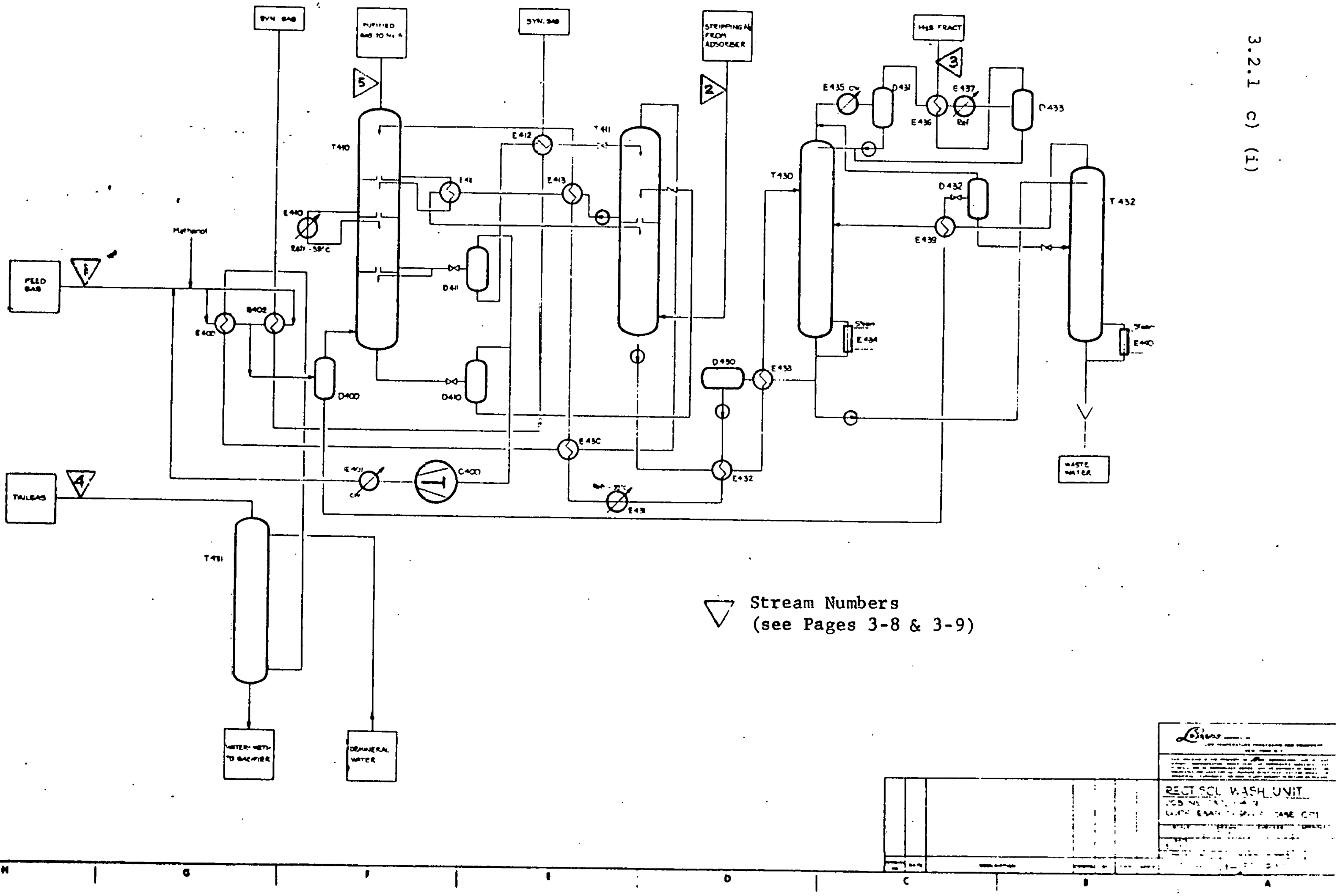
	Stream 5		Stream 6		Stream 7	
	Make Gas To Ammonia Synthesis		N_2 To Nitrogen Wash		Tail Gas From Nitrogen Wash	
	lb mol/hr	mol %	lb mol/hr	mol %	lb mol/hr	mol %
COS						
CO_2						
H_2S						
N_2	8,600.0	25.00	9,782.9	99.98	1.395.3	39.71
Ar	1.5	(43 ppm)	2.0	0.02	61.5	1.75
H_2	25,800.0	75.00			165.5	4.71
CO	0.2	(5 ppm)			1,534.4	43.67
CH_4					357.0	10.16
	34,401.7	100.00	9,784.9	100.00	3,513.7	100.00

3.2.1 Rectisol System (cont'd)

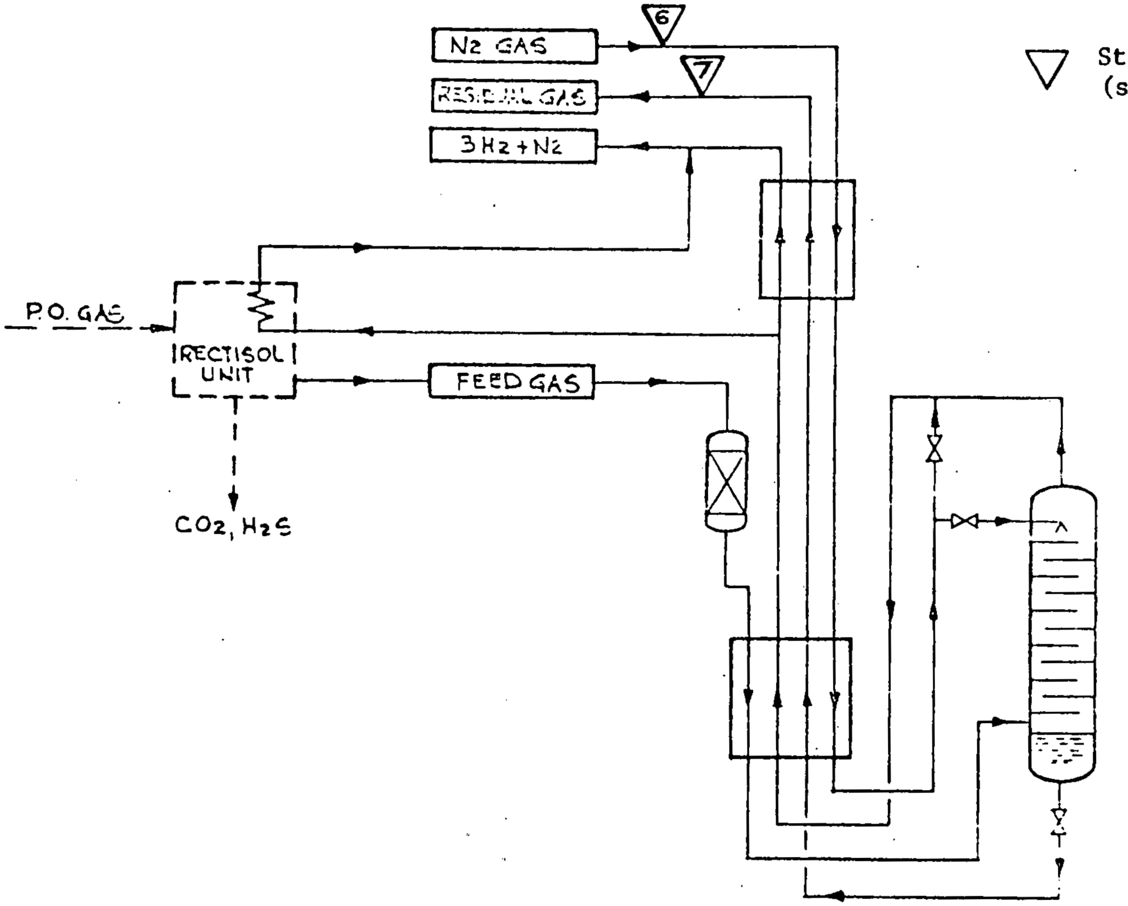
c) Schematics

(i) Rectisol Unit (1-FS-637)

(ii) Nitrogen Wash Unit (LO-1422)



▽ Stream Numbers
(see Pages 3-8 & 3-9)



▽ Stream Numbers
(see P 3-9)

Lotepro CORPORATION
LOW TEMPERATURE PROCESSING AND EQUIPMENT
NEW YORK N.Y.

TITLE

N2 WASH UNIT
JOB No 23200429

CHECKED	APPROVED	DATE
		9/14/78

PROJECT No	DRAWING No	REV.
LOT 749	LO-1422	

3.2.2 Selexol System

a) Process Descriptiom

i. H_2S Removal Unit

(Refer to Process Schematic on page 3-24)
Feed gas at 1,024 psia and 104°F is combined with recycle flash gas and then passes through the H_2S Absorber (T-101) where H_2S is removed by countercurrent contact with a solvent lean in H_2S and rich in CO_2. Using a solvent preloaded with CO_2 prevents a large temperature rise in the H_2S Absorber and for a given solvent temperature at the top of the H_2S Absorber results in a lower solvent rate to the H_2S Absorber. This absorber uses many stages to achieve a low solvent rate and thereby minimize the amount of co-absorbed CO_2 to produce a richer Claus gas. The rich solvent leaving the bottom of the absorber is let down in pressure through a Hydraulic Power Recovery Turbine (HT-101). The flash gas evolved is separated from the solvent in the High Pressure H_2S Absorber Recycle Flash Drum (D-101). The solvent leaving the bottom of this flash drum is further let down in pressure through a second Hydraulic Power Recovery Turbine (HT-102). The flash gas evolved is separated from the solvent in the Low Pressure H_2S Absorber Recycle Flash Drum (D-102). Finally, the flash gas passes through two stages of compression and cooling and is combined with the gas from the High Pressure H_2S Absorber Recycle Flash Drum. The combined stream goes through another stage of compression followed by cooling and combines with the feed gas to the H_2S Absorber. The purpose of this recycle flash is to return most of the co-absorbed CO_2 to the absorber, and thereby maintain a low level of CO_2 in the Claus gas.

3.2.2 Selexol System (cont'd)

The solvent leaving the Low Pressure H_2S Absorber Recycle Flash Drum enters the H_2S Stripper Feed Pump (P-101) which serves to provide enough head to get the solvent through the Rich Solvent-Lean Solvent Exchanger (E-107) and to the top of the H_2S Stripper (T-102). The capacity of this pump has been set to provide for recycling some of the normal flow through a Side Stream Filter (F-101). The rate is equivalent to 10% of the combined solvent flow through both the H_2S and CO_2 removal sections of the plant. The rich solvent is heated against hot stripped solvent. Make-up water is added to the rich solvent which then passes to the top of the H_2S Stripper where a large portion of the remaining dissolved gases flash off.

The flash liquid then enters the packed section of the stripper where the remaining solutes are stripped by countercurrent contact with steam generated by boiling water out of the solvent in the H_2S Stripper Reboiler (E-108). The flashed and stripped gases plus steam leaving the top of the stripper are cooled in the H_2S Stripper Condenser (E-106). Most of the steam is condensed. The condensate is returned to the H_2S Stripper through the H_2S Stripper Condensate Pump (P-102). The gas leaving the condenser contains the H_2S, COS and CO_2 which is the feed to the Claus Plant. On a dry basis the composition of this gas is 25 vol. % H_2S.

Stripped solvent enters the CO_2 Absorber Lean Solvent Booster Pump (P-103) and flows through the Rich

Solvent-Lean Solvent Exchanger where it is cooled. It then enters two 50% CO_2 Absorber Lean Solvent Pumps (P-104 A&B) and is further cooled in the Refrigerated Solvent Cooler (E-109) before entering the top of the CO_2 Absorber (T-201).

ii. CO_2 Removal Unit

(Refer to Process Schematic on page 3-24)

Gas leaving the H_2S Absorber is combined with recycle flash gas and passes through the CO_2 Absorber (T-201) where CO_2 is removed by countercurrent contacting with a cold solvent lean in CO_2. This absorber has a lean solvent entering at the top and a semi-lean solvent entering in the middle. This dictates a column with a large diameter at the bottom and a smaller diamter at the top to achieve a capital cost savings. Rich solvent from the absorber is split into two streams. Some of it passes through the H_2S Absorber Rich Solvent Pump (P-106) before entering the H_2S Absorber. The balance is let down in pressure through a Hydraulic Power Recovery Turbine (HT-103). The flash gas evolved is separated from the solvent in the CO_2 Absorber Recycle Flash Drum (D-103), and compressed, cooled and recycled to the CO_2 Absorber. The purpose of this flash is to return most of the co-absorbed H_2 to the absorber and thereby maintain a very high level of H_2 recovery. Solvent leaving the bottom of the CO_2 Absorber Recycle Flash Drum is further depressurized in another Hydraulic Power Recovery Turbine (HT-104). Solvent leaving the turbine passes into the top of the CO_2 Stripper (T-202) where a large portion of the dissolved gases flash off. The remaining CO_2 is removed with nitrogen stripping gas. The solvent leaving the stripper is pumped through two 50% capacity CO_2 Absorber Semi-Lean Solvent Pumps (P-105 A&B) and then enters the middle of the CO_2 Absorber.

3.2.2 Selexol System (cont'd)

iii. COS Removal Unit

(Refer to Process Schematic on page 3-25)
The design requirements call for 10 ppm maximum total sulfur in the vent stream. To meet this requirement, a vapor stream is taken up from an intermediate section in the H_2S Absorber and the COS in this stream is catalytically hydrolized to H_2S, and the converted stream returned to the H_2S Absorber.

A packed section is included in the H_2S Absorber above the point where the hydrolized stream is returned to provide for the additional H_2S loading resulting from the COS hydrolysis. A booster pump (P-107) delivers liquid to the bottom section from the top section of the H_2S Absorber to account for the pressure drop across the COS hydrolysis section.

Vapor from an intermediate section of the H_2S Absorber is preheated against COS hydrolysis effluent in the COS Hydrolysis Preheater (E-110). This temperature is controlled so that after the high pressure superheated steam (1500 psig, 940°F) is added to achieve a steam to gas ratio of 0.1, the resultant vapor temperature is 350°F. The stream then splits and enters two parallel beds (V-106 A&B) containing United Catalysts (formerly Catalysts and Chemicals, Inc.) C-53 COS Hydrolysis Catalyst. Two parallel beds are used so that each bed can have a reasonable length/diameter ratio. In addition, two parallel beds can permit operation at half-rate and thus allow for change out of the catalyst without shutting down completely. At 350°F and a steam to gas ratio of 0.1, thel COS is hydrolyzed over the catalyst to less than 2 ppm(v).

3.2.2 Selexol System (cont'd)

The COS hydrolysis effluent is cooled in the E-110 exchanger. It is further cooled against cooling water in the E-111 COS Hydrolysis Effluent Cooler, and finally, high level refrigeration is used in E-112 to cool it to the temperature at which it left the H_2S Absorber. The two-phase stream then enters the COS Hydrolysis High Pressure Separator (D-106). The vapor stream from D-106 is returned to the H_2S Absorber.

To keep hydrogen losses to a minimum, the liquid from D-106 is flashed in the COS Hydrolysis Medium Pressure Separator (D-107). The vapor off D-107 is recycled to the H_2S Absorber through the suction of the 3rd stage of the H_2S Absorber Recycle Compressor (C-103). The liquid from D-107 is flashed in the COS Hydrolysis Low Pressure Separator (D-108). The vapor off D-108 is recycled to the H_2S Absorber through the suction of the 1st stage of the H_2S Absorber Recycle Compressor (C-101).

iv. N_2 Wash Unit

(Refer to Process Schematic on page 3-27)

Following the Selexol Unit, the feed gas passes through the D-201, Drier Feed Condensate Trap in which any entrained liquids are removed from the gas. The gas then enters one of the T-301 A or B switching molecular sieve Drier-Adsorbers. The remaining water vapor, carbon dioxide, and other impurities are removed by adsorption. The adsorption process is designed to reduce the water, carbon dioxide, and impurities in the feed gas to levels sufficient to prevent freezeout in the cold box.

Prior to entering the cold box, the feed gas is filtered to remove any dust particles in the F-201 A or B Drier-

Adsorber Afterfilter. Feed gas and nitrogen enter the cold box and are cooled by cold product gas streams in the plate-fin heat exchanger. The nitrogen is cooled below its critical temperature and enters the top of the T-401 Nitrogen Wash Column as reflux. As the reflux flows down the column it is contacted on each tray by feed vapor rising from the column sump. By this process methane, carbon monoxide, and argon are washed from the feed. Purified gas passes overhead from the column, and is sent to the core exchangers. Bottoms product from the T-401 Nitrogen Wash Column is flashed. This reject stream is sent to the core exchangers. The hydrogen/nitrogen product and reject streams both provide refrigeration for the incoming feed and nitrogen gas streams. The hydrogen/nitrogen product from the cold box contains a maximum of 5 ppm of oxygen compounds. Nitrogen is blended into this hydrogen/nitrogen product stream to hold a constant 3:1 H_2:N_2 molar ratio feeding the ammonia synthesis loop.

The T-301 A and B adsorbers are operated on an 8-hour reactivation cycle. The off-stream adsorbent bed is reactivated using a portion of the reject stream from the cold box. The reactivation gas is heated in the E-201 Steam Reactivation Heater and the E-202 electric Reactivation Booster Heater and passed through the off-stream bed. When removal of water, carbon dioxide and any other impurities from the adsorbent is complete, the adsorber is cooled to operating temperature. The cooled, reactivated adsorber is then placed on-stream, and the saturated adsorber is removed from service for subsequent reactivation. The reject, after being used for adsorber reactivation is cooled in exchanger E-203, mixed with the remaining portion of the reject and delivered to the battery limits at a minimum pressure of 15 psia and a maximum of 80°F.

When defrost of the cold box is required, feed gas is heated in the E-204 Defrost Heater. The hot gas is passed through the cold box piping and equipment until a sufficient temperature has been attained to insure complete vaporization of any solidified impurities. The defrost time is reduced by draining the liquid inventory from the cold box equipment. These liquids are vaporized for disposal to flare in the E-205 vaporizer.

3.2.2. Selexol System (cont'd)

b) Mass Balance

Stream numbers refer to schematics on pages 3-24 to 3-27

	Stream 1		Stream 2		Stream 3		Stream 4	
	Feed Gas To Selexol		Stripping Nitrogen		Sour Gas To Claus Kiln		CO_2 To Vent	
	lb mol/hr	mol %	lb mol/hr	mol %	lb mol/hr	mol %	lb mol/hr	mol %
COS	3.3	(60 ppm)			0.7	0.03	0.1	(5 ppm)
CO_2	24313.4	45.48			1734.5	74.92	22565.0	81.26
H_2S	577.7	1.08			579.9	25.05	0.1	(5 ppm)
N_2	219.3	0.41	5107.2	99.98			5107.1	18.39
Ar	64.2	0.12	1.0	0.02			1.0	(37 ppm)
H_2	26276.7	49.15					49.5	0.18
CO	1604.9	3.00					29.9	0.11
CH_4	401.2	0.75					14.7	0.05
	53460.7	100.00	5108.2	100.00	2315.1	100.00	27767.4	100.00

3.2.2. Selexol System (cont'd)

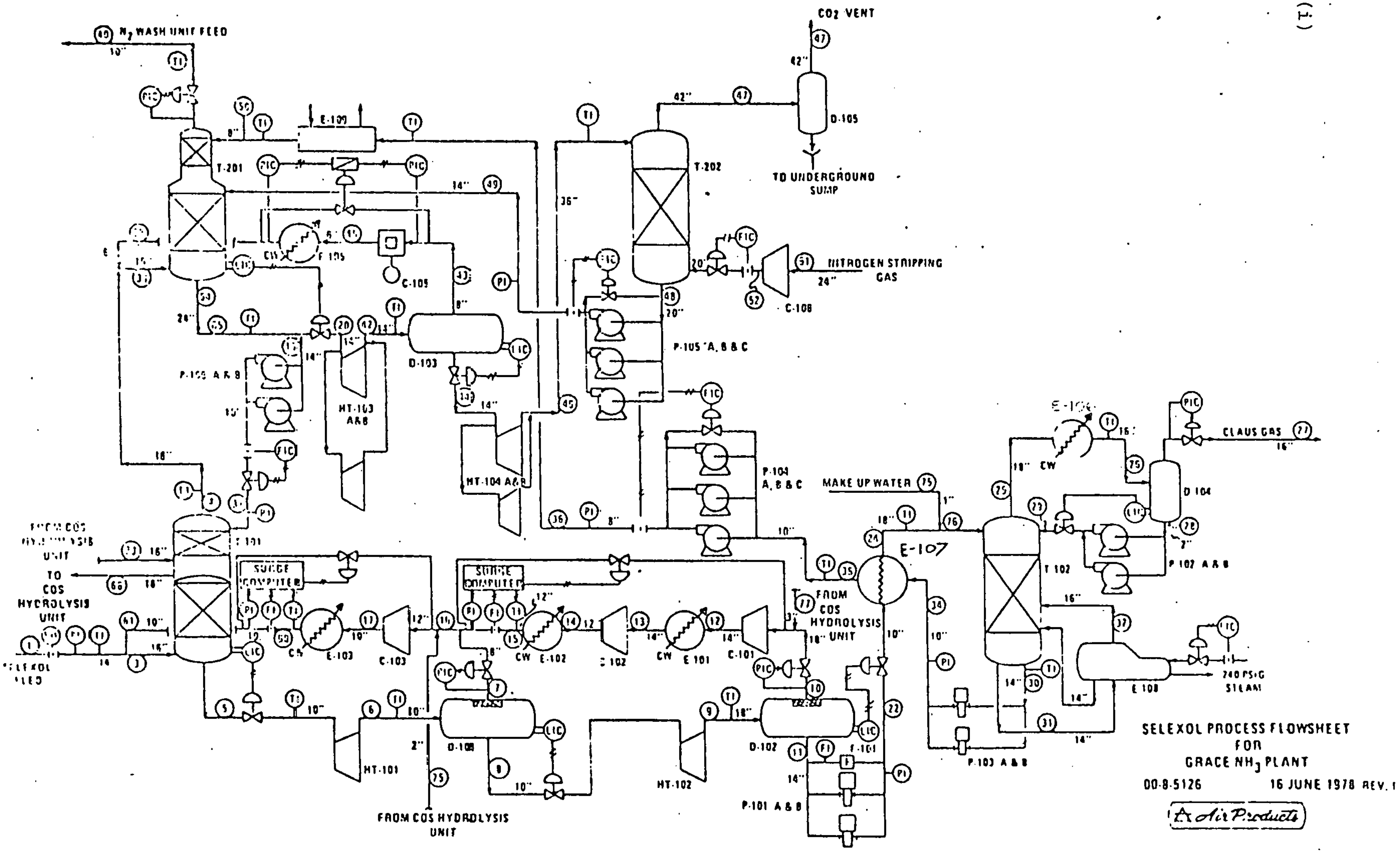
b) Mass Balance (cont'd)

	Stream 5		Stream 6		Stream 7	
	Make Gas To Ammonia Synthesis		N_2 To Nitrogen Wash		Tail Gas From Nitrogen Wash	
	lb mol/hr	mol %	lb mol/hr	mol %	lb mol/hr	mol %
COS						
CO_2					1.4	0.03
H_2S						
N_2	8600.0	25.00	10026.8	99.98	1646.2	40.15
Ar	1.5	(44 ppm)	2.0	0.02	64.7	1.58
H_2	25800.0	75.00			426.7	10.41
CO					1574.9	38.41
CH_4					386.5	9.42
	34401.5	100.00	10028.8	100.00	4100.4	100.00

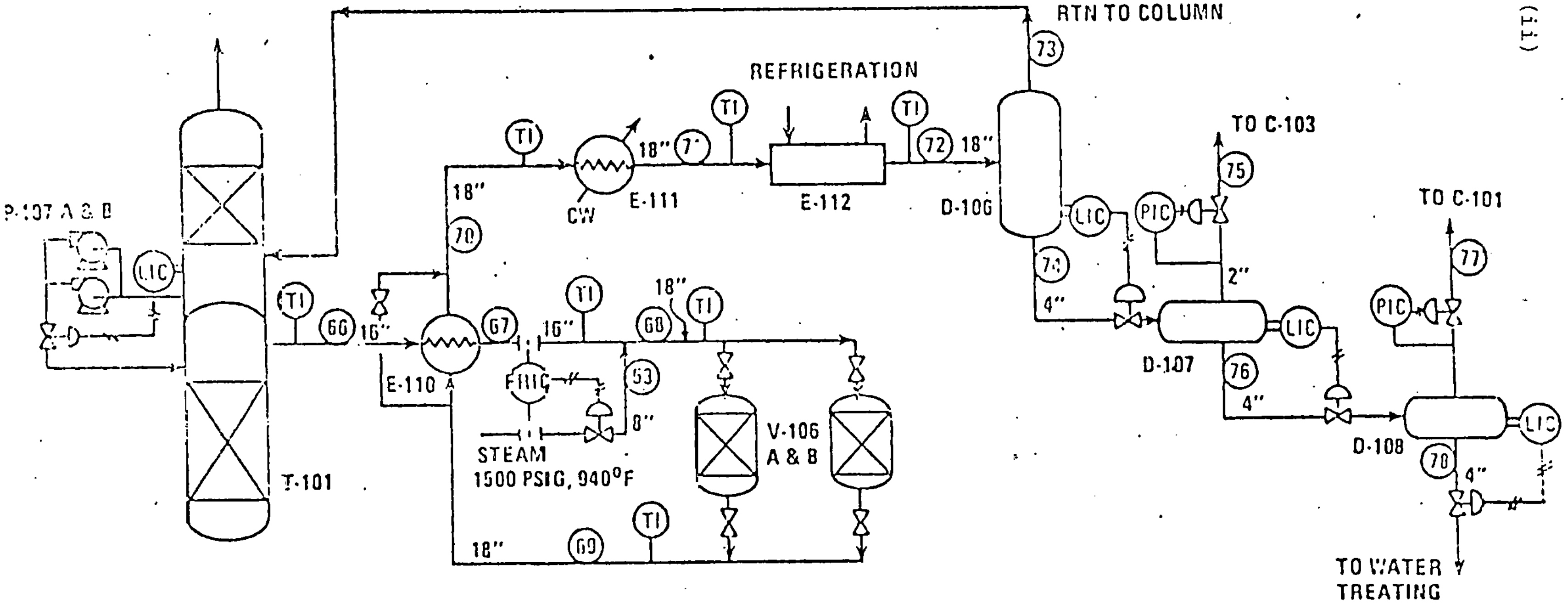
3.2.2. Selexol System (cont'd)

c) Schematics

- (i) Selexol Process Flowsheet
- (ii) COS Hydrolysis Unit
- (iii) Overall Process Flowsheet
- (iv) Cold Box Process Flowsheet N_2 Wash Unit



SELEXOL PROCESS FLOWSHEET
FOR
GRACE NH_3 PLANT
00-8-5126 16 JUNE 1978 REV. 1
Air Products



COS HYDROLYSIS UNIT
FOR
W. R. GRACE NH_3 PLANT

00-8-5126 12 JULY 1978 REV. 1

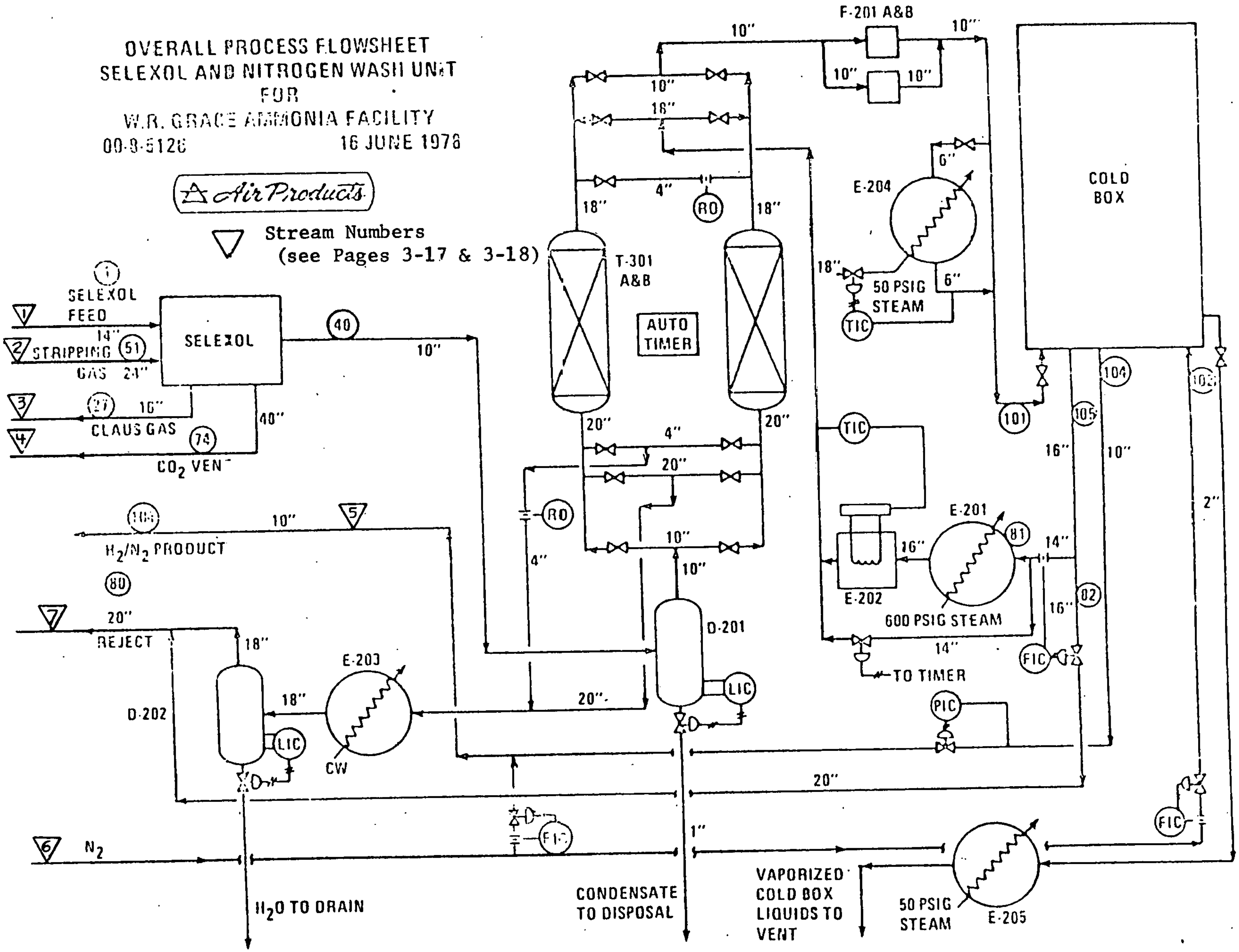
Air Products

OVERALL PROCESS FLOWSHEET
SELEXOL AND NITROGEN WASH UNIT
FOR
W.R. GRACE AMMONIA FACILITY

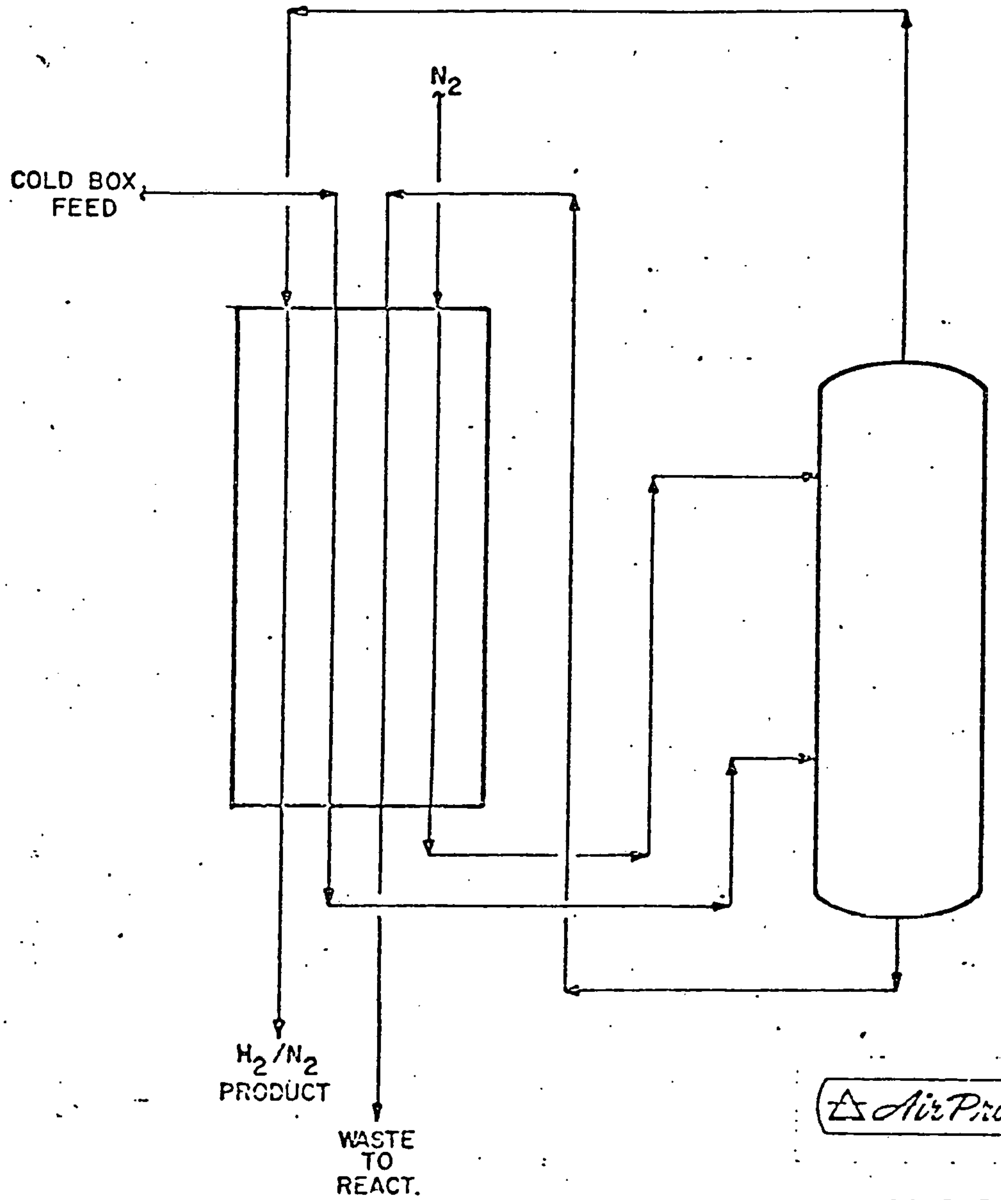
00-8-5126 16 JUNE 1978

Air Products

▽ Stream Numbers (see Pages 3-17 & 3-18)



3.2.2 c) (iv)



Air Products

COLD BOX
PROCESS FLOWSHEET
N_2 WASH UNIT
for
W.R. GRACE NH_3 PLANT
00-3-1125 11 SEPT. 1978

3.3 ANTICIPATED PERFORMANCE

3.3.1 Rectisol

a) For running the Rectisol and Nitrogen Wash plants, between one and two men are needed on each shift. A direct operating labor force totalling seven men is required.

b) Utility consumption is as follows, given on the basis of one short ton of ammonia produced:

Steam at 50 psig saturated:	657 lb
" " 240 " "	66 lb
" " 1500 " 940°F	21 lb
Cooling water	2830 US gal.
Deaerated water	80.5 US gal.
Electricity	31.9 kwhr.
Refrigeration at -31°F	20.6 ton hr
" " 23°F	12.7 " "

(1 ton hour equals 12,000 Btu)

c) Make-up of methanol from all causes totals about 2.3 lb per ton of ammonia produced.

3.3.2 Selexol

a) For running the Selexol and Nitrogen Wash plants, between one and two men are needed on each shift. A direct operating labor force totalling seven men is required.

b) Utility consumption is as follows, given on the basis of one short ton of ammonia produced:

Steam at		240 psig	saturated:	587	lb
"	"	580 "	$730^{\circ}F$	1828	lb
"	"	600 "	saturated	42.5	lb
"	"	1500 "	$940^{\circ}F$	818	lb

There is an export of 700 lb of saturated steam at 50 psig.

Cooling water	26,000 US gal
Electricity	9.0 kwhr
Refrigeration at $0^{\circ}F$	42.0 ton hr
" " $60^{\circ}F$	11.8 " "

(1 ton hour equals 12,000 Btu)

c) Make-up of Selexol solvent from all causes totals about 0.15 lb per ton ammonia produced.

d) The Licensor has projected that the utilities for the plant could be reduced by optimization of the design with some consequent increase in the installed cost. This projection has been evaluated and shown not to affect the overall conclusions.

3.4 COMMERCIAL EVALUATION AND RISK ANALYSIS

3.4.1 General

The object of this sub-section is to examine the risks involved in applying current technology to the gas purification requirements for this plant.

No gas absorption plants are carrying out the exact duty required here, and it is important to consider the effects of change in various parameters.

3.4.2 Size

The large capacity of the plant produces correspondingly large sizes for the equipment. The high operating pressure and low absorption temperatures employed for both acid gas removal processes, would permit the design of either plant as a single stream. However, there are practical benefits in limiting the maximum size of equipment; these relate to aspects of fabrication, transportation and erection, and also ease of operation. For these reasons, two streams are specified in some areas.

The nitrogen wash operates with small gas volumes (owing to the low temperatures) and low liquid rates. There is no difficulty in designing for either one or two streams; the costs appear similar.

The effects of size in this case present no problem.

3.4.3 Pressure

The feed gas pressure (and particularly the partial pressure of carbon dioxide) is considerably higher than in traditional plants. There are three main effects:

(i) There is greater danger of reactions between gas components and the absorbent liquids;

(ii) The system will be further from ideality, and design methods will be less accurate;

3.4.3 Pressure (cont'd)

(iii) Heating effects of absorption will be more pronounced, as for example when a given amount of CO_2 gas is dissolved into a smaller quantity of solvent.

The first point will be dealt with in the next paragraph. The non-ideality of the system requires that design data be measured at (and preferably well beyond) the area of operation; no theoretical calculation is reliable under these conditions. Both licensors state that this information is on file and that it was used for this design.
The unusually sharp temperature rise is easy to understand; less solvent flow involves a smaller thermal capacity.
The important effect here is that higher temperature may reduce the solubility of a gas component in the solvent, setting a practical limit to the partial pressure that can usefully be employed by a particular process or by a particular plant configuration using it.

3.4.4 Gas Impurities

Coal may be regarded as a mixture of rather complex chemicals, and gas made from it is likely to contain trace impurities that may react chemically with the solvent. It is therefore necessary to consider the characteristics of the solvent employed, and the operating experience with similar problems.

Both methanol and Selexol solvent are chemically robust, and at the low temperatures employed appear unlikely to react with the gas or to attack the materials of construction of the plant. Operating experience bears this out, in that solvent degradation and equipment corrosion appear rare for either process.

The most reliable information is provided by plant operation. Both processes have been employed over many years in a number of plants, including cases where the gas is "dirty", for example gas from partial oxidation of heavy fuel oil.

Lists of plants are given in the next sub-section, and support the claims of the process licensors.

The Rectisol process has an incidental advantage where contamination of the solvent may occur. To maintain an operating water balance, a side-stream of solvent is continuously distilled; any dissolved dirt is likely to be removed with the discarded water.

3.4.5 Effluents

As noted above, the feasibility of limiting these merits examination.

In plants of this type in which CO_2, say, is absorbed in a liquid and subsequently regenerated and vented to atmosphere, certain effluents are produced by the nature of the plant operation:

a) The vent gas will contain vapor and perhaps droplets the circulating liquid;

b) The plant contains circulating liquid, which may leak owing to a minor mishap (for example, a pump seal failure);

c) The vented CO_2 is likely to contain traces of all components present in the feed gas.

Droplets of liquid will be removed from the vent gas by an efficient demister mounted at the outlet from the regeneration tower; this is standard practice. Vapors and perhaps fog will pass this, but can be removed by scrubbing with water in tray towers; the liquid effluent can be pumped to the gasifiers for easy disposal.

3.4.5 Effluents (cont'd)

The area drains to sumps which are automatically pumped out to the disposal facilities. In the case of a methanaol spill, the area would need hosing down to dilute the methanol and suppress its evaporation.

When CO_2 is absorbed into the solvent some hydrogen and carbon monoxide will also be dissolved; the quantity of CO is 3 to 6 thousands of tons per year. The only ways in which the CO_2 purity can be increased (and the CO emission reduced) are to flash the CO_2 - rich solvent, and either compress the flash gas back into the gas feed line (thus conserving hydrogen, etc.,), or pass it to flare. The first method is limited by the rapid increase in compression cost, the second by the fact that excessive flashing produces a gas that is non-flammable because of its high CO_2 content. A reasonable design using both measures would reduce CO emission to about 600 tons per year.

Other apparently polluting, streams such as the nitrogen wash tail gas would be burned or flared, completely destroying objectionable components.

3.5 Lists of Plants

3.5.1 Linde Rectisol Units

Note: Three units were built by Messer Griesheim GmbH, whose know-how was acquired and transferred to Linde AG in 1971.

CLIENT	FEEDGAS	PRESSURE	COMPONENTS REMOVED (VOL.%)	CONTRACT AWARDED IN
Typpi Oy Oulu, Finland	16 MMSCFD hydrogen rich gas from partial oxidation of heavy fuel oil	356psia	CO_2,33.4% - 50ppm H_2S,0.4% - 1 ppm CO_2-Prod. 2.1 MMSCFD 99%	1960
Dansk-Norsk Kvaelstoff-fabrik, Copenhagen, DK	13.8 MMSCFD hydrogen rich gas from partial oxidation of heavy fuel oil	356 psia	CO_2 33.4% - 50 ppm H_2S 0.4% - 1 ppm CO_2 -Prod. 1.45 MMSCFD 99%.	1961
Texaco Inc. Los Angeles, USA.	79.6 MMSCFD hydrogen rich gas from partial oxidation.	481 psia	CO_2 33.3% - 10 ppm H_2S 0.49% - 1 ppm Tailgas: max. 5 ppm H_2S no CO_2 required. H_2S-Prod. 10% design result 25.6% H_2S	1966
H. Koppers GmbH, Essen, for Kutahya, Turkey.	46.9 MMSCFD converted gas from coal gasification.	351 psia	CO_2: 37.6% - 60 ppm H_2S: Traces to be removed CO_2-Prod. 12.9% MMSCFD 99%	1966

3.5.1 cont'd.

CLIENT	FEEDGAS	PRESSURE	COMPONENTS REMOVED (VOL.%)	CONTRACT AWARDED IN
American Air Liquide Inc. for Monsanto Texas City, USA.	53 MMSCFD hydrogen rich gas from steam reformer	351 psia	CO_2 10.2% - 20 ppm	1969
Long Island Lighting Comp. New York, USA.	4.3 MMSCFD natural gas.	602 psia	Odorants, CO_2, C_2+ CO_2 0.92 - 5 ppm	1969
Pilot - and Demonstration Plant for Texaco Inc. Monte Bello CAL.	1 MMSCFD hydrogen rich gas from partial oxidation of heavy hydrocarbons.	1140-2560 psia	CO_2 33% - 10 ppm H_2S 0.7% - 1 ppm H_2S in tailgas max. 5 ppm	1966
Borden Chemical Comp. New York, USA.	16.8 MMSCFD	313 psia	CO_2	1966
Rohm and Haas Comp. Philadelphia, USA.	12.6 MMSCFD	356 psia	CO_2, C_2H_2	1966
Brooklyn Union Gas Brooklyn New York, USA	11.9 MMSCFD	356 psia	CO_2	1966
Rohm GmbH, Darmstadt, West Germany-	7.9 MMSCFD natural gas.	498 psia	S-Components, C_2+ CO_2 5% - 50 ppm H_2S o.01% - 15 ppm	1970

3.5.1 Cont'd

CLIENT	FEEDGAS	PRESSURE	COMPONENTS REMOVED (VOL.%)	CONTRACT AWARDED IN
H.Koppers GmbH, Essen, for Modderfontein, South Africa.	146 MMSCFD hydrogen rich gas from coal gasification.	427/711 psia	CO_2 42.6% - 20 ppm H_2S 0.95% - 1 ppm H_2S-Prod. 80% H_2S/COS CO_2-tailgas: max. 10 ppm H_2S CO_2-Prod. 10.8 MMSCFD 99.9%	1972
Krupp Koppers GmbH, Essen, for Kafue, Zambia.	32.9 MMSCFD hydrogen rich gas from coal gasification.	398/341 psia	CO_2 42.3% - 20 ppm H_2S 1.0% - 20 ppm H_2S-Prod. 97% H_2S/COS CO_2-tailgas: max. 150 ppm H_2S	1975
Celanese Chemical Co. Houston, Texas, USA	Over 35 MMSCFD hydrogen rich gas.	441 psia		1975
Syngas Co. Houston, Texas, USA	from partial oxidation of heavy residual oil before and after CO-shift (2 stages)	853 psia	CO_2, H_2S, COS, HCN - 0.1 ppm H_2S + COS	1976
Gujarat State Fertiliser Co., Baroda, India.	168 MMSCFD Converted gas from partial oxidation of heavy fuel oil (1 stage)	1067 psia	35.9% CO_2 - 20 ppm 0.75% H_2S - 1 ppm CO_2-Prod. 26.8 MMSCFD - 98.5% CO_2	1976
UBE Industries Ltd., Tokyo.	21.1 MMSCFD from partial oxidation of heavy fuel oil.	612 psia	6.7% CO_2 - 20 ppm 1.46% (H_2S + COS) - 0.1 ppm H2S-fraction: 50% (H_2S + COS)	1977

3.5.2. Linde Nitrogen Wash Units

Note: Linde A.G. has built more than 60 Liquid nitrogen was units since 1950. The list below is an excerpt only.

CLIENT	FEEDGAS	PRESSURE	CONTRACT AWARDED
Anic S.p.A. Rom. Italy	21.0 MMSCFD hydrogen rich gas	185 psia	1955 (1958)
Hindustan Steel Ltd., New Delhi, India	15.7 MMSCFD Coke oven gas	185 psia	1959
Neyveli Lignite Corp. Ltd., Neyveli, India	17 MMSCFD x 2 shift-converted gas from coal gasification	228 psia	1959
Typpi Oy	10.7 MMSCFD shift-converted gas from oil gasification	356 psia	1960
Dansk Norsk Kvaelstoff-fabrik Copenhagen, Danmark.	9.1 MMSCFD shift-converted gas from oil gasification	341 psia	1961
H. Koppers GmbH for South Vietnam	10.7 MMSCFD shift-converted gas from coal gasification	327 psia	1963
U.S. Steel Corp. Pittsburg, USA.	131 MMSCFD Purified coke oven gas	185 psia	1964
H. Koppers GmbH for Kütahya, Turkey.	28.9 MMSCFD shift converted gas from coal gasification	341 psia	1966
VEBA-Chemie AG Gelsenkirchen West Germany.	98 MMSCFD shift-converted gas from oil gasification	683 psia	1970
H. Koppers GmbH for Modderfontein South Africa	82.2 MMSCFD shift-converted gas from coal gasification	668 psia	1972
Fertilizer Corp. of India Sindri.	76 MMSCFD shift-converted gas from oil gasification	583 psia	1975

3.5.2. Cont'd.

CLIENT	FEEDGAS	PRESSURE	CONTRACT AWARDED II
VEBA-Chemie AG Brunsbüttel West Germany	133 MMSCFD shift-converted gas from oil gasification	683 psia	1975
Krupp Koppers GmbH for Kafue, Zambia	19.1 MMSCFD shift-converted gas from coal gasification	341 psia	1975
Gujarat State Fertilizer Baroda, India	104 MMSCFD shift-converted gas from oil gasification	1038 psia	1976

3.5.3. Selexol Units

Owner & Location	Contractor Start Up	Description
	Bulk CO_2 Removal - Synthesis Gas	
Allied Chemical Corp Omaha, Nebraska	Allied Chemical 1965	CO_2 from 18% to 0.5% 45 MM SCFD @ 565 psia. Air stripping, no steam required. Carbon steel. Original solvent charge. Reforming of natural gas.
Ammoniaque Synthetique et Derives S.A. (ASED) Willebroek, Belgium	Davy Powergas GmbH Cologne, W.Germany 1975	CO_2 from 33% to $<1\%$ H_2S from 200 ppm to $<$ 4ppm HCN from 100 ppm to $<$ 10 ppm 47 MM SCFD @ 370 psia POX of crude oil. Replaced H_2O wash.
	Bulk CO_2 Removal - Natural Gas	
Coastal States Gas Producing Co. Six Shooter, Texas Grey Ranch Plant	Fish Engineering Houston, Texas 1969	CO_2 from 43% to $<$ 3.5% H_2S from 5 ppm to $<$ 1 ppm H_2O from saturation to $<$ 7#/MM SCF 275 MM SCFD @ 1000 psia No stripping, no refrigeration.
Lone Star Gas Co Ft. Stockton, Texas Pikes Peak Plant	Olsen Engineering Davy Powergas, Inc. Houston, Texas 1974	CO_2 from 43% to $<$ 3.5% H_2S from 30 ppm to 4 ppm Maintains $<$ 7#/MM SCF 50 MM SCFD @ 1000 psia No stripping, no refrigeration.
	1975	Expansion to 100 MM SCFD
Northern Natural Gas Co. Ft. Stockton, Texas Mitchell Plant	Ortloff Corp. Midland, Texas 1977	CO_2 from 28% to $<$ 3.5% H_2S from 16 ppm to $<$ 4 ppm H_2O from saturation to $<$ 7#/MM SCF 180 MM SCFD @ 900 psia No stripping. Replaced DEA.

3.5.3. (cont'd)

Owner & Location	Contractor Start Up	Description
	Selective Sulfur Removal - Natural Gas	
Northern Natural Gas Co Ft. Stockton, Texas Oates plant	Fish Engineering Houston, Texas 1969	H_2S from 100 ppm to < 4 ppm CO_2 from 18% to $<$ 2.5% Maintains H_2O @ < 7 # /MM SCF 130 MM SCFD @ 1000 psia Inert gas stripping -- split flow
Norddeutsche-Erdgas-Aufberaitungs-GmbH Mobil, Esso, Shell Sulingen, West Germany NEAG II plant	Davy Powergas GmbH (formerly Pintsch Bamag) Cologne, W.Germany 1970	H_2S from 4000 ppm to $<$ 2 ppm CO_2 from 7% to $>$ 5% COS from 130 ppm to 70 ppm Maintains H_2O @ < 7 # /MM SCF 62 MM SCFD @ 1000 psia Inert gas stripping Replaced potassium carbonate
	Mobil Oil A.G. in Deutschland 1973	H_2S from 9.2% to $<$ 2 ppm CO_2 from 9.5% to 8 % COS from 130 ppm to 70 ppm RSH from 100 ppm to $<$ 1 ppm 80% H_2S to Claus plant Maintains H_2O @ < 7 # /MM SCF 62 MM SCFD @ 1090 psia Inert gas stripping Replaced Alkazid
	Mobil Oil A.G. in Deutschland 1976	Expansion to 67 MM SCFD
Wintershall AG Barnstorf, W.Germany Dueste I plant	Davy Powergas GmbH (formerly Pintsch Bamag) Cologne, W.Germany 1970	H_2S from 7% to $<$ 1000 ppm CO_2 from 9% to 5% Maintains H_2O @ < 7 # /MM SCF 32 MM SCFD @ 1000 psia Inert gas stripping Replaced propylene carbonate and glycol-amine

3.5.3. (cont'd)

Owner & Location	Contractor Start Up	Description
	Selective Sulfur Removal - Natural Gas	
	Wintershall AG 1973	Expansion to 45 MM SCFD 15 MM SCFD to $<$ 2 ppm H_2S
Wintershall AG Ruetenbrock, W.Germany	Comprimo b.v. Amsterdam 1976	H_2S from 60 ppm to $<$ 2 ppm CO_2 from 5% to 4% Maintains H_2O @ $<$ 7 #/MM SCF 45 MM SCFD @ 1400 psia Inert gas stripping
Wintershall AG Barnstorf, W.Germany Dueste II plant	Comprimo b.v. Amsterdam 1977	H_2S from 7% to $<$ 2 ppm CO_2 from 9% to 5% COS from 118 ppm to 60 ppm RSH from 100 ppm to $<$ 1 ppm 68% H_2S to Claus plant Maintains H_2O @ 7#/MM SCF 45 MM SCFD @ 1000 psia Steam stripping
	Selective Sulfur Removal - Synthesis Gas	
ERDA/AGA Homer City, Pennsylvania Bi-Gas pilot plant	Stearns-Roger Denver, Colorado 1976	H_2S from 7000 ppm to $<$ 4 ppm CO_2 from 31% to $<$ 1% 35% H_2S to Claus plant $<$ 20 ppm H_2S to CO_2 vent 11 MM SCFD @ 1500 psia max. Steam stripping Gasification of coal to SNG

3.5.4. Air Products Nitrogen Wash Units

Location	Year	Syngas Output
Collier Carbon and Chemical Corporation Brea, California, USA	1954	125 tons/day
Co-operative Farm Chemical Association Lawrence, Kansas, USA	1954	125 tons/day
John Deere and Company Pryor, Oklahoma, USA	1955	200 tons/day
Northern Chemical Industries (Girdler), Searsport, Maine, USA	1955	125 tons/day
Sun Oil Company Marcus Hook, Pennsylvania, USA	1956	315 tons/day
Gonzales Chemical Industries Guanica, Puerto Rico	1956	125 tons/day
Phillips Pacific Chemical Company Kennewick, Washington, USA	1957	220 tons/day
W. R. Grace & Co Memphis, Tennessee, USA	1959	260 tons/day
Government of Venezuela, Ministry of Mines Moron, Venezuela	1959	120 tons/day
E.I. DuPont Gibbstown, New Jersey, USA	1959	300 tons/day
Georgia Pacific Plaquemine, Louisiana	1978	575 tons/day

SECTION 4

ECONOMICS

CONTENTS

SECTION 4

ECONOMICS

4.1 CAPITAL COST ANALYSIS

4.1.1 Basis

The total installed cost of the gas purification system, excluding the cost of land and of unusually stringent measures to reduce effluents, was estimated by Air Products for the Selexol version and by Lotepro for the Rectisol system. The basis is for erection in Kentucky, with all costs referred to June 1978; the effects of inflation have not been included.

The above figures were examined, and a number of omissions and anomalies identified and corrected. It became obvious that certain types of equipment had been estimated on different bases for the two plants, and where this was so an independent estimate was made of the relevant items in the two plant versions. In this way there is assurance that the relative plant costs are as accurate as possible, and that reliance can be placed on any difference between them. The figures for installed cost should not be used for other purposes without further study.

For both of the plants studied a royalty is payable; this has been included in the capital cost figure. No royalty is payable for Nitrogen Wash.

4.1.2 Table 4-1

Installed Cost of Gas Purification Systems

	Rectisol	Selexol
Acid Gas Removal Plant, including royalty, $ million	47.5	49.0
Nitrogen Wash $ million	9.0	9.0
Total	56.5	58.0

4.2 OPERATING COSTS

4.2.1 Basis

a) All operating costs are expressed in (June 1978) dollars per ton (2,000 lb) ammonia product.

b) The Commercial Plant concerned produces 3500 tons of ammonia per day, for 330 days per year.

c) Capital charges are represented by 15 year straight-line depreciation.

d) Direct operating labor is taken as 7 men total (about 1½ men per shift) for either plant; this includes the nitrogen wash. The cost is assumed $20,000/man year.

e) The cost of supervision is taken as 30% of the cost of labor.

f) The cost of maintenance is taken as 4% of installed cost per year.

g) Direct overhead is calculated as 30% of d) plus e).

h) Indirect overhead is calculated as 65% of d) plus e) plus f).

i) The cost of steam is:

50	psig	saturated	$1.64/ST(Steam)
230	"	540°F	2.75
240	"	saturated	2.56
580	"	730°F	3.58
600	"	saturated	3.22
1500	"	940°F	4.49

Exhaust steam is credited at the above levels.

j) (Bought-in) electricity cost $0.0185/kwhr.

k) Cooling water cost is $0.03/1000 USg circulated.

l) Raw water is available at$0.10/1000 USg; demineralisation adds $0.50 to this; for the Rectisol plant clarified deaerated water is required, at a cost assumed to total $0.20/1000 USg.

4.2.1 Basis (cont'd)

m) The cost of refrigeration (regarded as a utility) has been calculated as follows; note - a ton-hour is defined as 12,000 Btu:

-31°F	\$0.097/ton-hr
0°F	0.068
23°F	0.049
60°F	0.027

n) Methanol is currently quoted at \$142 per short ton, Selexol solvent at \$1 per lb.

4.2.2.

Table 4-2

Operating Cost of Gas Purification Plant - $ per ST Ammonia

	Rectisol				Selexol			
Capital Charges								
Depreciation				3.261				3.346
Labour Costs								
Direct Labour			0.121				0.121	
Supervision			0.036				0.036	
Maintenance			1.957				2.008	
Direct Overhead			0.047				0.047	
Indirect Overhead			1.374				1.407	
Total				3.535				3.619
Utilities, Chemicals								
Steam: 1500 psig	21 lb	0.047			818 lb	1.836		
600 psig					42.5 lb	0.068		
580 psig					1828 lb	3.272		
240 psig	66 lb	0.085			587 lb	0.751		
50 psig	657 lb	0.538			-700 lb	(0.574 Cr)		
Subtotal			0.670				5.353	
Electricity	31.9 Kwhr		0.590		9.0 Kwhr		0.166	
Cooling Water	2830 gal		0.085		26,000 gal		0.781	
Deaerated Water	80 gal		0.016					
Refrigeration:								
-31°F	20.6 ton.hr	2.001						
0°F					42.0 ton.hr	2.858		
23°F	12.7 ton.hr	0.622						
60°F					11.8 ton.hr	0.320		
Subtotal			2.623				3.178	
Solvent Make-up	2.3 lb		0.162		0.15 lb		0.152	
Total				4.146				9.630
Total Production Cost				10.942				16.595
		Delta:						5.65 S/ST

4.3 ECONOMICS OF OPERATION

For true profitability a gas purification unit must have reliability and ease of operation, or apparent costs mean little. The evidence presented in Section 3 is of sufficient scope and in sufficient detail for confidence in the process - see in particular Section 3.4.

Operating costs of all types are detailed and compared in this Section 4. It is clear that quite large changes in the relationship between different costs (say, capital charges and electricity) would not alter the findings of this study.

Confidence can therefore be placed in the conclusions in Section 2.

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY III

METHOD OF COAL PREPARATION

EBASCO SERVICES INCORPORATED
Two Rector Street
New York, NY 10006

April 1979

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1. INTRODUCTION

1.1 OBJECTIVE

The objective of this study was to evaluate the performance, reliability and the capital and operating costs of wet grinding as compared to dry grinding in the preparation of a coal slurry for use in the Texaco Coal Gasification System of the Commercial Plant. The study also included a discussion of the entire coal handling system from barge and rail unloading points to the grinding equipment.

1.2 SCOPE

1.2.1 Coal Handling

A conceptual design is discussed for the Coal Handling based on Ebasco's experience in the design of fossil fuel power plants. Availability of equipment, and the relationship between Commercial Plant and Demonstration Plant sizes are included in the discussion.

1.2.2 Coal Preparation

1.2.2.1

The study evaluated a dry grinding system based on the use of a cage mill and wet grinding systems based on ball and rod mills.

1.2.2.2

The study did not evaluate the coal grinding system used to prepare the coal for firing in boilers because the equipment used for this system is well established, and has been in successful operation in numerous fossil fuel power plants.

1.2.2.3

The following procedures were used to develop the data which is included in this study:

a) Surveyed vendors with coal grinding or related grinding experience in order to obtain data for equipment specifications, capital costs, and expected grind distribution.

b) Prepared a conceptual design for alternate coal preparation systems using data obtained from vendors. Developed order-of-magnitude capital investment and operating costs.

c) Evaluated the conceptual designs for the alternate coal preparation systems, and selected a system for pilot plant tests.

d) Selected vendors to conduct the pilot plant tests based on the availability of suitable test facilities and experience with coal grinding.

e) Evaluated the pilot plant test results for acceptability of the grind size distribution.

f) Utilized the pilot plant test results to develop recommendations for the Commercial Plant which take into account economics, operability and risk factors.

2. CONCLUSIONS AND RECOMMENDATIONS

2.1 SUMMARY OF CONCLUSIONS

2.1.1 Coal Handling System

The Coal Handling System employs commercially known and accepted technology used extensively in existing utility and industrial complexes.

2.1.2 Coal Preparation System

2.1.2.1

Grinding a coal/water slurry in a ball mill in open circuit operation will not produce a product free of significant quantities of plus 14 U.S. mesh material without overgrinding.

2.1.2.2

Adjustment of ball mill operation in open circuit to produce a slurry with the required top size results in a product high in the minus 325 fraction. This change adversely affects slurry properties and will reduce efficiency in downstream filtration operations.

2.1.2.3

Solids content for grinding of a coal/water slurry in a ball mill appears to be limited to a maximum of 53 percent. However it is noted that this limitation may be related to the specific mechanical configuration of the equipment used in pilot plant tests.

2.1.2.4

On the basis of preliminary tests it is believed that the required top size specification can be obtained by wet grinding of coal in a ball mill operated in a closed circuit.

However present data indicates that a low solids content must be used, resulting in dewatering requirements.

2.1.2.5

Grinding a coal/water slurry in a rod mill in open circuit operation will produce a product with a suitable grind classification except for the presence of 0.2 to 0.5 percent of plus 14 U.S. Mesh.

2.1.2.6

Solids content for grinding of a coal/water slurry in a rod mill can be in excess of sixty percent without blockage in the mill or the discharge system.

2.2 RECOMMENDATIONS

2.2.1 Commercial Plant

Wet grinding of coal in a rod mill operated in open circuit, using a trommel screen for removal of oversize material is recommended for the conceptual design of the Commercial Plant for the following reasons:

a) Wet grinding of coal in a rod mill operated in open circuit has been successfully demonstrated in a pilot plant.

b) The open circuit operation is a simpler system with fewer items of equipment, and avoids some of the potential difficulties associated with the closed circuit such as water balance and chloride buildup.

c) Present indications are that a 14 U.S. mesh trommel screen will perform satisfactorily in this service, however it is recommended that further work be conducted prior to the final design.

d) Wet grinding of coal presents no serious environmental problems and associated economic risks such as are introduced by dry grinding of coal.

2.2.2 Demonstration Plant

2.2.2.1 Coal Handling System

It is recommended that the design of the Coal Handling System for the Demonstration Plant be based on using equipment similar to, but at one-third of the size of equipment specified for the Commercial Plant. The one major exception to this recommendation involves the barge unloader where it is recommended that for the Demonstration Plant a single grab bucket be used rather than the twin bucket ladder specified for the Commercial Plant. Because of the smaller size required in the Demonstration Plant a twin bucket ladder barge unloader would cost approximately one million dollars more than the single grab bucket type. The savings in operating cost for the twin bucket does not warrant the extra capital cost.

2.2.2.2 Coal Preparation System

The impact of oversized product material on the performance of downstream equipment, and in particular on the ability of the slurry pumps to deliver a consistent volume of slurry has been considered for this study, but its effect has not yet been demonstrated in tests. Conflicting opinions have been expressed by slurry pump vendors as to the maximum slurry top size which can be handled without causing flow interruptions.

In consideration of the severe consequence which will occur in downstream processing equipment if the slurry flow is suddenly interrupted it is recommended that the Demonstration Plant design be based on a system which will more positively eliminate the possibility of oversize material reaching the slurry pumps.

The use of an open circuit system involving rod mill grinding, vibrating screening for separation and recycling of 14 plus U.S. mesh material, followed by dewatering facilities should be considered.

3. TECHNICAL APPROACH

3.1 DESIGN BASIS

3.1.1 General

The study was to be based on a high sulfur, agglomerating Eastern coal. Kentucky coal from the No. 9 seam was selected as the basis for study and test work.

3.1.2 Coal Handling System

The Coal Handling System of the Commercial Plant shall be capable of handling all the coal required for steam generation and for slurry preparation. The system includes all the equipment required to unload coal from barge and rail unloading points, to stack coal in live and dead storage piles, and to reclaim coal from the piles for delivery to separate silos for steam generation and slurry preparation.

The unloading and stacking equipment shall handle approximately 60,000 short tons of eight inch top size coal per week, during one shift per day, for five to six days per week. Three inch top size coal will be stored.

The reclaiming system shall continuously deliver one-quarter inch top size coal at a rate of 351 short tons per hour to the silos. Approximately 271 short tons per hour of coal shall be delivered to the slurry preparation silos, and 80 short tons per hour to the steam generation silos.

3.1.3 Coal Preparation System

The Coal Preparation System shall be capable of grinding 271 short tons per hour (ST/hr) (dry basis 238 ST/hr). The grind product shall have a nominal classification of

100% minus 14 U.S. mesh.

The grinding system may be wet or dry, and may be operated in open or closed circuit.

For wet grinding allowance must be made for the addition of a process carbon and ash recycle stream to the 271 ST/hr of coal.

3.2 PROCESS CONSIDERATIONS

3.2.1 <u>System Descriptions</u>

3.2.1.1 Coal Handling System

Coal is transported to the Plant by barges and/or railcars. The barge unloading system is designed for an average rate of 1500 short tons per hour. The coal is transported to the surge bin by a belt conveyor. Magnetic separators remove all ferrous metal items from the coal stream. The rail unloading system is designed for an average rate of 1300 short tons per hour. The system is complete with a thawing shed to handle frozen cars in winter operation. A rotary car dumper is used to unload the rail car into a track hopper. A frozen coal cracker is positioned beneath the track hopper, and is used during winter months when the coal is in large frozen lumps. During the remainder of the year the breaker plates are pushed back to allow coal to pass through the frozen coal cracker. From the coal cracker, the coal is transferred to a conveyor by means of a vibrating feeder. The material is conveyed to the same surge bin used for barge unloading.

From the surge bin, material is transferred to the coal breaker spreader type vibrating feeders. The coal is reduced from 8" x 0" to 3" x 0" and oversize extraneous foreign material is removed in the screening operation. The material is conveyed to a live storage pile with a capacity of four (4) days storage for 100 percent plant load. Excess coal is transported to the dead storage area by means of a bulldozer and a scraper where it is compacted and shaped. Dead storage has a minimum of a 30 day coal supply at 100 percent plant load.

The coal reclaim system is independent of the unloading system. Coal is reclaimed from the live storage area by means of one of the four vibrating type coal pile dischargers. Reclaimed coal feed rate is controlled by the vibrating feeder under each coal pile discharger. The material is transferred to a surge bin by means of a series of conveyors. The conveyor system has a magnetic separator for removal of tramp iron and a belt scale to measure the reclaimed coal quantity. From the surge bin, vibrating feeders are used to transfer the coal to hammer mills where the material is reduced from 3" x 0" to 1/4" x 0". Each mill is designed for 3000 short tons per hour. The coal is then transferred by a series of conveyors to storage silos for the boiler and gasification plant.

3.2.1.2 Coal Preparation Systems

a) Coal Preparation System - Scheme I

Refer to Exhibit #4 for a system schematic arrangement. Three parallel trains each of 50% plant capacity would be installed in the plant for dry grinding of coal in a cage type mill and conveying the grind product to the slurry mix tanks. Two trains would normally be in operation. Coal is transferred to the cage mill by means of a weigh belt feeder. In the mill the material is reduced from 1/4" x 0" to nominally 100% minus 14 U.S. mesh.

The discharge from the mill is conveyed to a bucket elevator and transferred to a storage bin. The material is then transferred to one of four (4) slurry mix tanks by means of a dense phase pneumatic conveying system.

b) Coal Preparation System - Scheme II

Refer to Exhibit #5 for a system schematic arrangement. Three parallel trains each of 50% plant capacity would be installed in the plant for wet grinding of coal in a ball mill operated in open circuit and for collection of product slurry and transfer to the slurry mix tank and dewatering facilities. Two 100% capacity centrifuges or filters would be used in the dewatering operation and product cake would be transferred to the slurry mix tanks by two 50% conveyors.

Coal, process recycle slurry and make-up water are added to the ball feed inlet chute at controlled rates to make a slurry concentration of 45% solids by weight. The grind product from the mill is a nominal 100% minus 14 U.S. mesh, which is discharged into a sump tank. The slurry is transferred by means of a centrifugal pump to the slurry mix tank and dewatering operation. The fraction of slurry dewatered is controlled to obtain, in the final product slurry concentration, the specified percent solids by weight.

c) Coal Preparation System - Scheme III

Refer to Exhibit #6 for a system schematic arrangement. Three parallel trains each of 50% plant capacity would be installed in the plant for wet grinding of coal in a rod mill operation in open circuit and for collection of product slurry and transfer to the slurry mix tank.

Coal, process recycle slurry and make-up water are added to the rod mill feed inlet chute at controlled rates to provide a slurry concentration 60 plus percent solids by weight. The product grind slurry is passed through a trommel screen to remove oversize material. Product slurry is transferred by means of a centrifugal pump to the slurry mix tank for final addition of water to obtain the desired solids content.

3.2.2 System Evaluation

Both dry and wet grinding systems were considered feasible for the coal preparation study. Preliminary analysis of capital investment and operating costs based on vendor data indicated that the dry grind system offered some economic advantage over the wet grinding system. However, the wet grinding system was selected for the Commercial Plant after consideration of the following process parameters.

3.2.2.1

Dry grinding of coal in open circuit cage mill operation has previously been demonstrated in pilot plant operation. However, the product grind was screened to remove oversize material before preparation of the coal/water slurry. If the product grind coal must be screened, a dry screening operation would require approximately thirty-two (32) large screens which would increase the investment and operating costs substantially above a wet grinding system using a wet screening operation. The system would also create potential environmental problems associated with handling and recovery of fine coal dust.

3.2.2.2

Taking into account the product size distribution required, a dry grinding system may be subject to pluggage of product coal during periods of high moisture content in the feedstock.

3.2.2.3

The product coal discharge from the mill must be transported to the gasifier as a coal/water slurry. Using a dry grinding system, the slurry preparation system for the Commercial Plant would be designed in a batch mode to insure thorough mixing of dry coal and water followed by final adjustment of the product slurry for the specified percent solids by weight.

3.2.2.4

The problem of mixing coal and water in a wet grinding system does not exist. The system also offers the possibility of continuous operation for slurry preparation with potential savings in the number and size of product slurry mix tanks.

3.2.2.5

In the Commercial Plant design, a substantial quantity of process material is recycled for use in the preparation of product slurry. In a dry grinding system, this stream would be returned to the slurry mix tank as shown in Exhibit 4. To preclude excessive quantities of oversize material in the product slurry feed to the gasifier, the recycle stream may require a separate wet grinding system.

3.2.2.6

In a wet grinding system, the process recycle stream can be added to the mill inlet, thereby eliminating the oversize problem.

3.3 COMMERCIAL EVALUATION

3.3.1 Survey of Grinding Equipment Manufacturers

A total of thirteen major manufacturers were contacted to provide cost estimates and technical information for various types of mills that might be used to grind coal for the process slurry. The types of mills investigated were rod, ball, bowl, hammer, cage and disc. Both wet and dry grinding methods were considered using open and closed circuits. Each manufacturer was requested to provide cost estimates, sizing and power requirements for their equipment on the basis of grinding 220 tons per hour of the typical coal considered for this program. This is the nominal design flow rate of coal feedstock for a Commercial Plant to produce 3500 STD of ammonia from coal.

Prior coal grinding experience and shop facilities to perform grinding tests later on actual samples of coal were also covered in discussions with the manufacturers.

The manufacturer's data, listed in Exhibit 3, were analyzed and evaluated to select coal slurry preparation systems for further study. Of the two manufacturers contacted on disc mills, one declined to provide the requested data at this time and the second manufacturer provided preliminary data on number of mills required for the specified duty, but without product grind size distribution. In the U.S. there is no service experience with wet grinding of coal with disc mills that characteristically operate at approximately 1200 RPM. Although this type of mill may find application in wet grinding of coal in the future, the application of a disc mill was excluded from further study in this program.

3.3.2 Pilot Plant Tests

3.3.2.1 Dry Grinding

Dry Grinding of coal in a cage mill has been demonstrated in pilot plant tests conducted prior to this trade-off study, as being capable of producing the required size classification. These tests utilized screening to separate oversize material from the mill product. Additional pilot plant tests were not deemed necessary.

3.3.2.2 Wet Grinding

a) Ball Mill Pilot Plant Tests

Vendor "A" was selected for pilot plant investigation of wet grinding of coal based on previous experience in a commercial installation, estimated product size distribution using a ball mill in open circuit and availability of tests facilities. A truckload of coal considered closely representative of that which will be used in the program was shipped to the Vendor's test center.

The initial test program was based on investigating the following areas of interest:

1) Determine the feed rate to the ball mill to obtain a minus 16 U.S. mesh top size product in open circuit operation.

2) Determine the effect of finer grind product top size on product size distribution and ball mill operation.

3) Determine the maximum percent of solids by weight that can be achieved by wet grinding without flow restrictions through the mill.

4) Determine the effectiveness of a disc filter in removing moisture from the product slurry discharged from the mill.

As a result of preliminary test results obtained in the pilot plant operation, the program was modified to investigate wet grinding of coal in a closed circuit ball mill using a spiral classifier for separation of oversize material.

The pilot plant investigation used a 2' x 4' ball mill, spiral classifier and one four-foot diameter disc filter. Size distribution of coal feedstock to the ball mill is tabulated in Exhibit 7. Tests results of open and closed circuit ball mill operation are tabulated in Exhibits 8 and 9 respectively.

The duration of the test program did not permit sufficient time to study the effect of ball loading, extensive variation in percent of critical speed, or modification of the mill discharge piping configuration. Within these limitations, the test results indicated a maximum slurry concentration of 53% dry solids before pluggage occurred in the system; open circuit ball mill could not meet the program requirements on top size; a closed circuit ball mill operation could possibly satisfy the program requirements with additional test work although the classifier product slurry would be only 15% by weight dry solids. This would require that the major portion of the product slurry be dewatered to obtain the optimum slurry concentration for the gasifier operation.

b) Rod Mill Pilot Plant Tests

Vendor "D" had received a contract from the Department of Energy (DOE) to investigate various modes of coal grinding for a coal gasification program sponsored by the government. As a result of the experience gained in the pilot plant investigations, a computer simulation model for coal grinding had been developed. Before actual pilot plant work was commenced, a series of simulation runs were made on wet grinding of coal in open and closed circuit ball mill operation. The simulation results tended to confirm the preliminary results obtained in Company "A's" test center. The simulation run for wet grinding of coal in a rod mill open circuit operation indicated that this may be the promising approach to meet the program required objectives on top size material and maximum product slurry concentration. A closed circuit wet coal grinding system would require addition of water to the classifier regardless of the type used, in order to effect a separation of top size material. The product slurry concentration from the mill would also contain the major fraction of the dilution water resulting in a final product slurry of lower concentration.

Size distribution of coal feedstock used in the tests is shown in Exhibit 10. A three-foot diameter rod mill was used in all the tests. The rod charge and percent of critical speed was not varied during the investigation. The test results indicate a slurry concentration as high as 62% dry solids can be obtained without pluggage of the rod mill or mill discharge slurry transfer system for the coal tested. The maximum slurry concentration was not limited by viscosity properties but by the mill inlet piping configuration used to transport the coal and water to the mill inlet. The fraction plus 16 U.S. mesh varied from 0.2 to 0.5 percent of the product grind.

The test results of wet grinding in a rod mill with open circuit operation are not included in this report because they contain data considered proprietary by Texaco Development Corporation.

3.4 RISK ANALYSIS

3.4.1 General

Successful performance is expected of a Coal Preparation System constructed in accordance with the system recommended for the conceptual design of the Commercial Plant, i.e. wet grinding in a rod mill with a trommel screen for removal of oversize material. However, the translation from a conceptual design to an actual plant involves some elements of risk which are evaluated in the following paragraphs.

3.4.2 Equipment Availability

In the Commercial Plant 220-280 ST/hr of solids must be handled in the coal preparation system. The coal preparation plant for the Black Mesa Pipeline, using rod mills in open circuit operation, processes 660 ST/hr of coal. Therefore the availability and size of equipment required for the Commercial Plant presents no significant risks.

3.4.3 Deviations from Pilot Plant Tests

3.4.3.1

In the pilot plant study, 3/4" x 0 top size was used as feedstock to the rod mill. A 2" x 0 top size feedstock is proposed for the Commercial Plant design. The extrapolation of top size feedstock from pilot plant to the Commercial is considered to be of nominal risk.

3.4.3.2

The coal/water slurry preparation work in the pilot plant used city or town water for adjustment of the percent solids in the slurry concentration. In the commercial plant, the water for preparation of the coal/water slurry will be a process recycle stream containing various chemicals compounds dissolved in the water. The impact if any, on corrosion of liners, rods, etc,has not been demonstrated. It is assumed that the corrosion problems can be controlled by pH adjustments of the coal/water slurry.

3.4.4 Untested Elements of the System

The impact of the carbon/ash recycle streams on the product size distribution and fraction of product oversize has not been demonstrated in the pilot plant investigation because recycle material was not available. It is assumed that the product size distribution will not change significantly when carbon/ash recycle is added to the coal slurry in the rod mill. Provision has been made in the design for the potential of a higher power requirement with recycle. Oversize material as a result of the carbon/ash recycle is not expected to limit the plant capacity. However, since a representative sample of carbon/ash recycle will not be available until the actual plant operation, its effect on the performance of the Coal Preparation System must be listed as an uncertain factor.

3.4.5 Elimination of Oversize Material

The process for elimination of oversize material has not been demonstrated. The risks must therefore be evaluated in terms of plus 14 U.S. mesh material in the product slurry going to the gasifier. There exists the possibility that oversize material can be caught in the suction or discharge valves of the pumps used to feed the slurry to the Gasifier at high pressure. Such an occurrence can cause a sudden reduction in the rate of slurry to the gasifier. Without a corresponding change in oxygen flow, dangerously high temperatures will occur in the Gasifier. Because of this danger, the question of oversize material must be considered a high risk factor. It is essential that further test work be conducted prior to the final design, to evaluate the performance of equipment to remove oversize material, and to determine the effect of oversize material on the slurry pump performance.

4. ECONOMICS

4.1 CAPITAL COST ANALYSIS

Cost estimates were obtained from each of the manufacturers for the type of mills selected for Coal Preparation System schemes I, II and III. Cost estimates were developed for the additional equipment required for each system which are included in Exhibit 11. These are summarized below:

Coal Preparation System	Estimated Capital Investment $1,000,000
Scheme I (Cage Mill, Dry Grinding)	3.6
Scheme II (Ball Mill, Wet Grinding)	5.0
Scheme III (Rod Mill, Wet Grinding)	3.8

4.2 OPERATING COSTS

4.2.1 Basis

a. All operating costs are expressed in 1978 dollars per short ton of ammonia product.

b. The Commercial Plant produces 3500 tons of ammonia per day for 330 days per year.

c. Capital charges are represented by 15 year straight line depreciation.

d. Direct operating labor is assumed as four men total for all three systems. The assumed cost is $20,000/ Man Year.

e. The cost of supervision is taken as 30% of the cost of direct labor.

f. The cost of maintenance as a percent of plant installed plant cost is taken as 5.42 for scheme I, 2.84 for scheme II, and 3.42 for scheme III.

g. Direct overhead is calculated as 30% of d) plus e).

h. Indirect overhead is calculated as 65% of d) plus e) plus f).

i. Purchased electricity cost $0.0185/kw hr.

4.2.2 Total Production Costs ($/ST NH_3)

	SCHEME I DRY GRINDING CAGE MILL		SCHEME II WET GRINDING BALL MILL		SCHEME III WET GRINDING ROD MILL	
LABOR COSTS						
Direct Labor	0.069		0.069		0.069	
Supervision	0.021		0.021		0.021	
Maintenance	0.169		0.123		0.112	
Direct Overhead	0.027		0.027		0.027	
Indirect Overhead	0.168		0.138		0.131	
Total		0.454		0.378		0.360
CAPITAL CHARGES DEPRECIATION		0.208		0.289		0.219
UTILITIES						
ELECTRICITY (Kwhr)	(10.6)*	0.196	(11.5)*	0.213	(16.4)**	0.303
TOTAL PRODUCTION COST ($/ST NH_3)		0.858		0.880		0.882

* Feedstock Topsize: 1/4" x 0

** Feedstock Topsize: 2" x 0

4.2.3 Cost Adjustment

During the development of information for this trade-off study it was determined that the use of a 2" x 0 feedstock topsize to the rod mill would permit better control of fines in the product grind. Accordingly, capital and operating costs for scheme III - wet grinding in a rod mill, were obtained on this basis. The data obtained earilier for dry grinding in a cage mill (scheme I), and wet grinding in a ball mill (scheme II) were for 1/4" x 0 feedstock topsize. Therefore the total production costs listed in 4.2.2 are based on different feedstock topsizes. Implicit in the data is the fact that for schemes I and II there will be equipment, and a corresponding power consumption for obtaining the quarter-inch topsize. For purposes of a better comparison of the three schemes, an adjustment has been made to the total production costs for scheme III to account only for electric power costs. With this adjustment it is estimated that the total production costs for scheme III, assuming 1/4" x 0 feedstock topsize would be 0.794 $/ST of ammonia.

4.3 RISK ANALYSIS

4.3.1 Escalation

The investment estimates prepared for each of the schemes investigated for the Coal Preparation System are order-of-magnitude estimates based on 1978 equipment costs. It is not expected that the price escalations as a result of inflation would change the conclusions and recommendations of this study.

4.3.2 Screening to Remove Oversize Material

4.3.2.1 Dry Grinding - Cage Mill

The capital cost analysis does not include any allowance for screening of oversize material. If dry screening is required, it is estimated that approximately 32 screens each about 4 feet wide x 20 feet long will be required. The equipment cost alone is expected to be in excess of one million dollars. Consideration of the building requirements and the dust collection equipment costs for such a large screening area makes the economic risk intolerable for dry grinding.

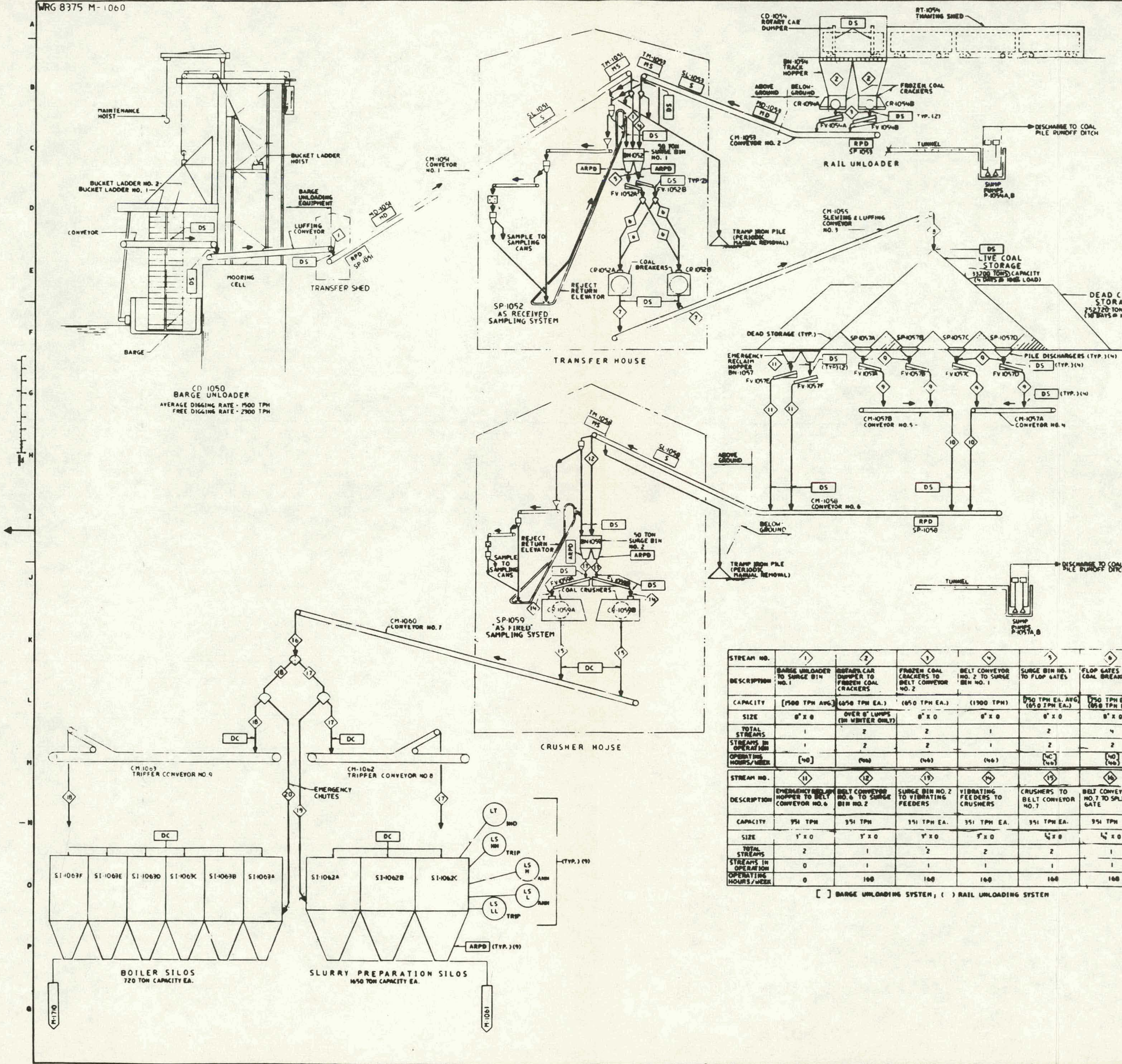
4.3.2.2 Wet Grinding - Ball and Rod Mills

The capital cost analysis does not include any allowance for screening of oversize material other than the use of the trommel screen in the case of the rod mill. If wet screening is required with ball or rod mill operation, it is estimated that approximately 16 screens each about 4 feet by 8 feet will be required. The cost of the screens alone will be approximately $160,000. Additional dewatering equipment required with the wet screens will have order-of-magnitude cost of $320,000.

4.3.3 Carbon/Ash Recycle

The impact of carbon/ash recycle for the wet grinding systems has been discussed in section 3.4.4. In the case of dry grinding with a cage mill, the wet carbon/ash recycle stream can not be introduced into the mill, but must be added at the slurry mix tanks. If the carbon/ash recycle stream contains oversize material, it will be necessary to add an additional grinding system for the recycle stream.

WRG 8375 M-1060



STREAM NO.	◇1	◇2	◇3	◇4	◇5	◇6	◇7	◇8	◇9	◇10
DESCRIPTION	BARGE UNLOADER TO SURGE BIN NO. 1	ROTARY CAR DUMPER TO FROZEN COAL CRACKERS	FROZEN COAL CRACKERS TO BELT CONVEYOR NO. 2	BELT CONVEYOR NO. 2 TO SURGE BIN NO. 1	SURGE BIN NO. 1 TO FLOP GATES	FLOP GATES TO COAL BREAKERS	COAL BREAKERS TO SLEWING & LUFFING CONVEYOR NO. 3	SLEWING & LUFFING CONVEYOR TO LIVE COAL STORAGE	LIVE COAL STORAGE TO BELT CONVEYOR NOS. 4 & 5	BELT CONVEYOR NOS. 4 & 5 TO BELT CONVEYOR NO. 6
CAPACITY	[1500 TPH AVG.]	(650 TPH EA.)	(650 TPH EA.)	(1300 TPH)	[750 TPH EA. AVG] (650 TPH EA.)	[750 TPH EA. AVG] (650 TPH EA.)	[750 TPH EA. AVG] (650 TPH EA.)	[1500 TPH AVG] (1300 TPH)	351 TPH EA.	351 TPH EA.
SIZE	8" X 0	OVER 8" LUMPS (IN WINTER ONLY)	8" X 0	8" X 0	8" X 0	8" X 0	3" X 0	3" X 0	3" X 0	3" X 0
TOTAL STREAMS	1	2	2	1	2	4	2	1	4	2
STREAMS IN OPERATION	1	2	2	1	2	2	2	1	1	1
OPERATING HOURS/WEEK	[40]	(46)	(46)	(46)	[40] (46)	[40] (46)	[40] (46)	[40] (46)	168	168

STREAM NO.	◇11	◇12	◇13	◇14	◇15	◇16	◇17	◇18	◇19	◇20
DESCRIPTION	EMERGENCY RECLAIM HOPPER TO BELT CONVEYOR NO. 6	BELT CONVEYOR NO. 6 TO SURGE BIN NO. 2	SURGE BIN NO. 2 TO VIBRATING FEEDERS	VIBRATING FEEDERS TO CRUSHERS	CRUSHERS TO BELT CONVEYOR NO. 7	BELT CONVEYOR NO. 7 TO SPLITTER GATE	SPLITTER GATE TO SLURRY PREPARATION SILOS	SPLITTER GATE TO BOILER SILOS	FLOP GATE TO SLURRY PREPARATION SILO (SI-1062A)	FLOP GATE TO BOILER SILO (SI-1063A)
CAPACITY	351 TPH	351 TPH	351 TPH EA.	351 TPH EA.	351 TPH EA.	351 TPH	271 TPH	80 TPH	271 TPH	80 TPH
SIZE	3" X 0	3" X 0	3" X 0	3" X 0	1¼" X 0	1¼" X 0	1¼" X 0	1¼" X 0	1¼" X 0	1¼" X 0
TOTAL STREAMS	2	1	2	2	2	1	1	1	1	1
STREAMS IN OPERATION	0	1	1	1	1	1	1	1	0	0
OPERATING HOURS/WEEK	0	168	168	168	168	168	168	168	0	0

[ ] BARGE UNLOADING SYSTEM; ( ) RAIL UNLOADING SYSTEM

NO	DATE	REVISION	BY	CH	APPROVED	NF VIEW
1	8-3-78	REVISED: (C17, J17) NOTE; (A1, Q23) DWG NO. WAS M-173; (D10, F13, G16, H13, H15, 19, J8, J10, K9, L9) DS; (J20) GEN. NOTES. ADDED: EQUIP. NOS. (123) REF. NO.; (E4) TRANS. SHED; (Q1, Q7) REF. DWG. NO.; (G12, I12, O13) STREAM	J.O.	[illegible]	[illegible]	
2	[illegible]	REVISED: (L4, L9, N1) STREAM NO. 15 THRU 19; (M2, M5) WAS SHUTTLE CONVEYOR; (L11 THRU L21) TABLE; (E16; F17) CAPACITY. ADDED: (N4) STREAM NO. 20. DELETED: (E15) TELESCOPIC CHUTE; (L6) CM-1008; (J21) NOTE 3	JK	JO	[illegible]	

EQUIPMENT NOT SHOWN

WB-1050	WORK BOAT
BS-1050	SHUTTLE BARGE
YL-1054	YARD LOCOMOTIVE
SP-1064	BULLDOZER
SP-1065	SCRAPER

LEGEND

◇	STREAM NO.
◿	VIBRATING FEEDER
▽	FLOP GATES
DS	DUST SUPPRESSION
MD	METAL DETECTOR
MS	MAGNETIC SEPARATOR
RPD	RIP PROTECTION DEVICE
S	SCALE
ARPD	ARCHING & RATHOLING PREVENTIVE DEVICE
DC	DUST COLLECTION
◇	SPLITTER GATE

GENERAL NOTES

1 THIS IS A CONCEPTUAL DESIGN DRAWING
2 COAL CAN BE SUPPLIED EITHER FROM BARGE OR FROM RAILROAD.
SEE STREAMS ◇1 TO ◇8

EXHIBIT I

U.S. DEPARTMENT OF ENERGY
W.R. GRACE & CO.
MEMPHIS, TENNESSEE
SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
INDUSTRIAL PROJECT "B" - PHASE I
CONTRACT NO. ET-77-C-01-2577
COMMERCIAL PLANT
PROCESS SCHEMATIC
COAL HANDLING
UNLOADING & RECLAIMING
EBASCO SERVICES INCORPORATED
SCALE NONE
APPROVED
DATE 4-26-78
DIV. MECHANICAL
SECT. CODE NO. 050
DR. R. POMERENK
CH. S. BALAKES
WRG 8375 M-1060
ISSUE 2

EASTERN COAL - EXPECTED RANGE OF ANALYSES*
EXHIBIT #2

		RAW COAL	
PROXIMATE ANALYSIS - AS RECEIVED		Minimum	Maximum
%	Moisture	4.19	12.00
%	Ash	13.90	22.03
%	Volatile	27.79	36.30
%	Fixed Carbon	40.00	49.23
		Not Additive	
ULTIMATE ANALYSIS - DRY BASIS			
%	Carbon	57.95	68.61
%	Hydrogen	4.0	5.00
%	Nitrogen	0.50	1.57
%	Chlorine	0.03	0.24
%	Sulfur	3.35	6.90
%	Ash	14.88	24.77
%	Oxygen	3.50	6.81
		Not Additive	
Btu - As Received (Min-Max. Range)		9 248	11 948
Btu - Dry Basis (Min-Max. Range)		10 397	12 470
% Sulfur - As Received		3.05	6.30
% Sulfur - Dry Basis		3.35	6.90
Hardgrove Grindability Index		48	66

* Range of properties given to manufacturers of grinding equipment.

LINE-UP OF MANUFACTURER'S COST ESTIMATE FOR GRINDING MILL EQUIPMENT

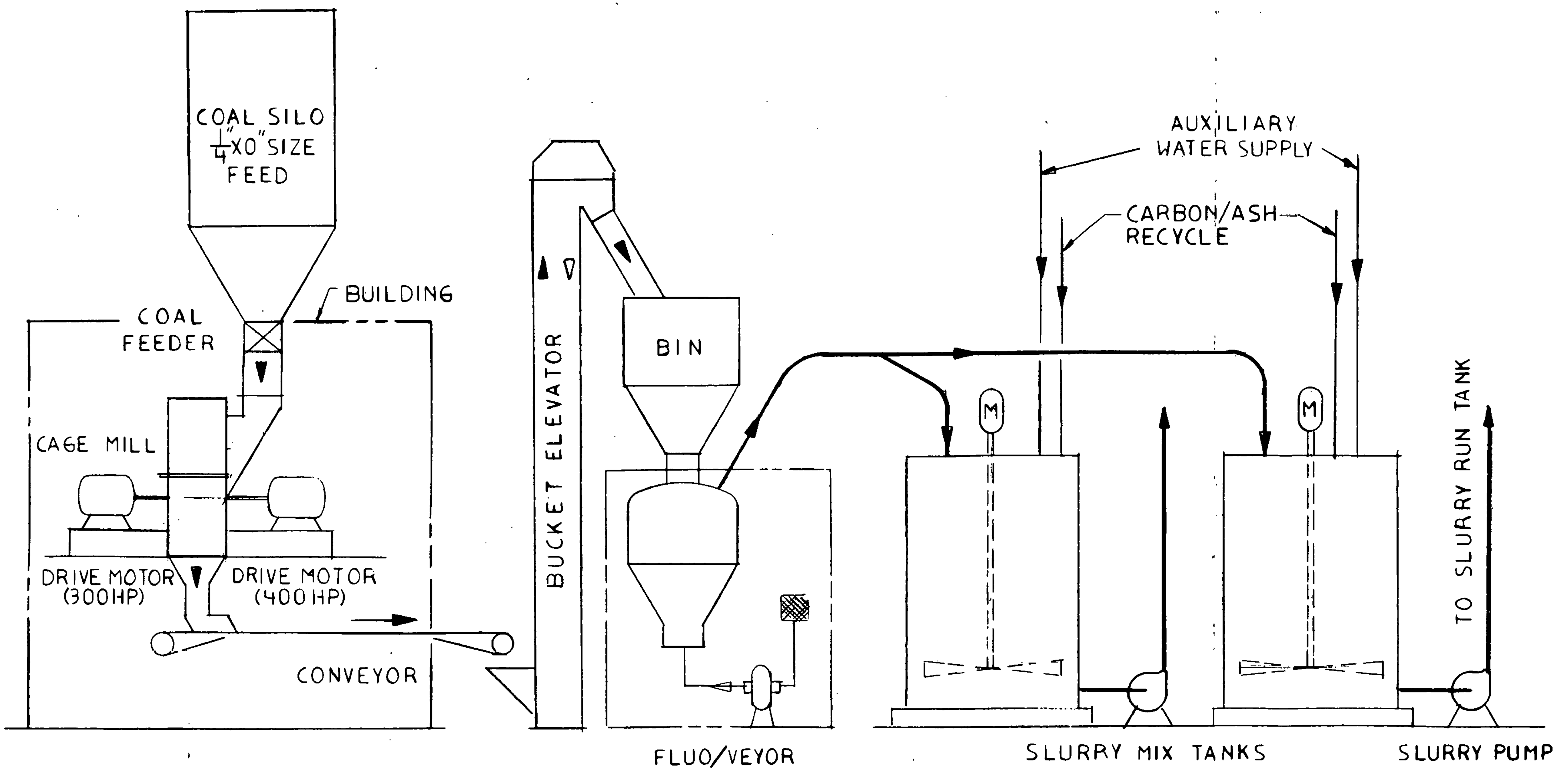
EXHIBIT #3

MANUFACTURER	A	B	C	D	E	F
TYPE OF MILL	Ball	Cage	Hammer	Ball	Cage	Rod
WET OR DRY GRIND	Wet	Dry	Wet	Wet	Dry	Wet
OPEN OR CLOSED CIRCUIT	Open	Open	Closed	Closed	Open	Open
GRIND SIZE-% THRU # 14 MESH	100	100	89	100	100	99.5
GRIND SIZE-% THRU # 40 MESH	100	90	56	-	100	90
GRIND SIZE-% THRU #100 MESH	44.3	50	25	-	80	60
GRIND SIZE-% THRU #200	27.9	30	6	35	-	80
GRIND SIZE-% THRU #325 MESH	27.8	22	2	25	25	-
SLURRY-% SOLIDS (WGT)	50	--	50	50	-	50
MILL UNIT DATA						
COAL FEED - TPH	110	220	110	220	220	110
DRYING AIR FLOW - LB/HR	-	-	-	-	-	-
DRYING AIR TEM - F	-	-	-	-	-	-
P OF AIR FLOW-IN. H_2O	-	-	-	-	-	-
DRIVE MOTOR(S) & HP	900	1 - 1750	1 - 600	3000	1-700 & 1-800	1-1500
MOTOR/MILL - RPM	180/25	1750/900	1200/1200		V-Belt	180/-
MOTOR -KW						
MILL & MOTOR SIZE-FT X FT X FT						
ESTIMATED UNIT COST						
MILL - $1000	330	205	77	1160	76	668
MOTOR(S) - $1000	25	45	Incl	Incl	44	Incl
ROB OR BALL CHARGE - $1000	30	-	-	Incl	-	Incl
DRYING AIR EQPT - $1000	-	-	-	-	-	-
LINERS OR OTHER EQPT $1000		1	-	Incl feeders	11	-
TOTAL	385	251	77	1160	131	668
TOTAL NO. MILLS & CAPACITY -%	2-50%	1-100%	2-50%	1-100%	1-100%	3-50%
TOTAL COST-ALL MILL - $1000	770	251	154	1160	131	2005
PRIOR COAL SLURRY EXPERIENCE:	Black Mesa Pipeline Slurry Prep Use 3-Rod Mills 100%-14 mesh as Primary Mills	Black Mesa Pipeline Secondary Crusher for Mill Feed	2 Secondary Crushers at Cadiz, Ohio Slurry Pipeline & Black Mesa Pipeline has Secondary Impactor	No Coal. But many Limestone wet Grind Mills for AQCS	None, Mfr says this Cage Type Mill cannot be used for wet grinding	None

LINE-UP OF MANUFACTURER'S COST ESTIMATE FOR GRINDING MILL EQUIPMENT

EXHIBIT #3

MANUFACTURER	G	H	I	J	K
TYPE OF MILL	Ball	Bowl	Bowl	Ball	Rod
WET OR DRY GRIND	Dry	Dry	Dry	Wet	Wet
OPEN OR CLOSED CIRCUIT	Closed'	Closed	Closed	Closed	Closed
GRIND SIZE-% THRU # 14 MESH	98	-	100	100	98
GRIND SIZE-% THRU # 40 MESH	72	98	-	75	85
GRIND SIZE-% THRU #100 MESH	53	-	-	45	-
GRIND SIZE-% THRU #200	35	70	-	30	26
GRIND SIZE-% THRU #325 MESH	25	-	25	25	-
SLURRY-% SOLIDS (WGT)	-	-	-	60-70	50-54
MILL UNIT DATA					
COAL FEED - TPH	55	60	73.3	110	75
DRYING AIR FLOW - LB/HR	143,000		218,000	-	
DRYING AIR TEM - F	520		600	-	
P OF AIR FLOW-IN. H_2O	12.5			-	
DRIVE MOTOR (S) & HP	1-900	470	705	1-900	1-700
MOTOR/MILL - RPM	1200/-			700/-	522/-
MOTOR -KW			412		
MILL & MOTOR SIZE-FT X FT X FT					
ESTIMATED UNIT COST					
MILL - $1000	500	405	525	400	245
MOTOR(S) - $1000	Incl			25	56
ROB OR BALL CHARGE - $1000	Incl	-	-	30	
DRYING AIR EQPT - $1000				-	
LINERS OR OTHER EQPT $1000					26
TOTAL				455	
TOTAL NO. MILLS & CAPACITY -%	4-25%	4-25%	3-33%	2-50%	3-33%
TOTAL COST-ALL MILL - $1000	2000	1620	1575	910	981
PRIOR COAL SLURRY EXPERIENCE:	None	None	None	None Dry Grind coke for rotary kilns	None



NOTE:

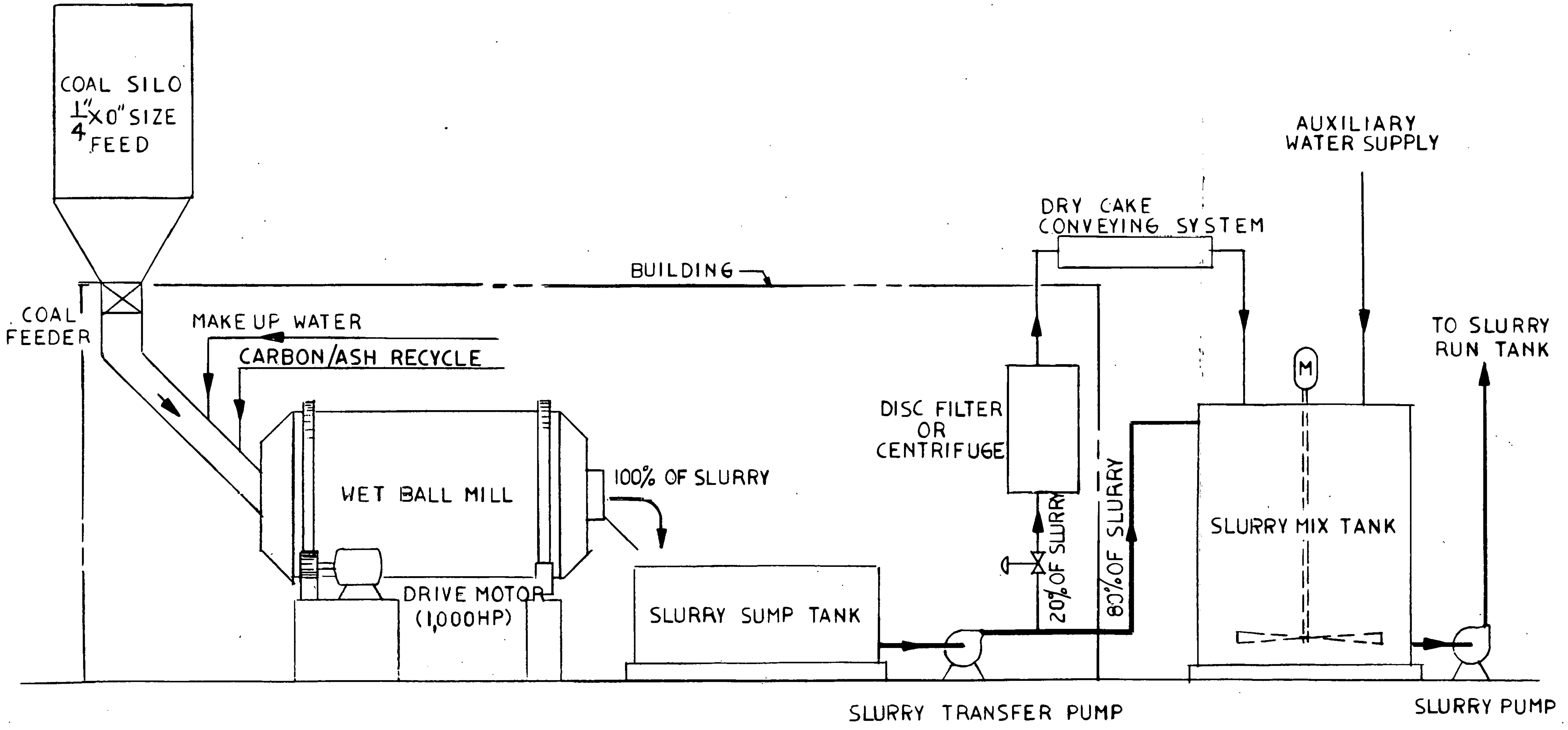
1 THERE ARE THREE 50% CAPACITY CAGE MILLS, CONVEYORS, BUCKET ELEVATORS, BINS AND FLUO/VEYORS.

2 FOUR SLURRY MIX TANKS ARE THE SAME FOR ALL SCHEMES

3 THE SPARE CAGE MILL SHALL BE CAPABLE OF FEEDING EITHER GROUP OF TWO SLURRY MIX TANKS

SYNTHESIS GAS
DEMONSTRATION PLANT PROGRAM

EBASCO SERVICES INCORPORATED		COMMERCIAL PLANT	EXHIBIT
DIV. MECH DR. EGS	APPROVED	TOS III METHOD OF COAL	4
CH.		PREPARATION SCHEME I	
DATE 8/10/78			



NOTE:

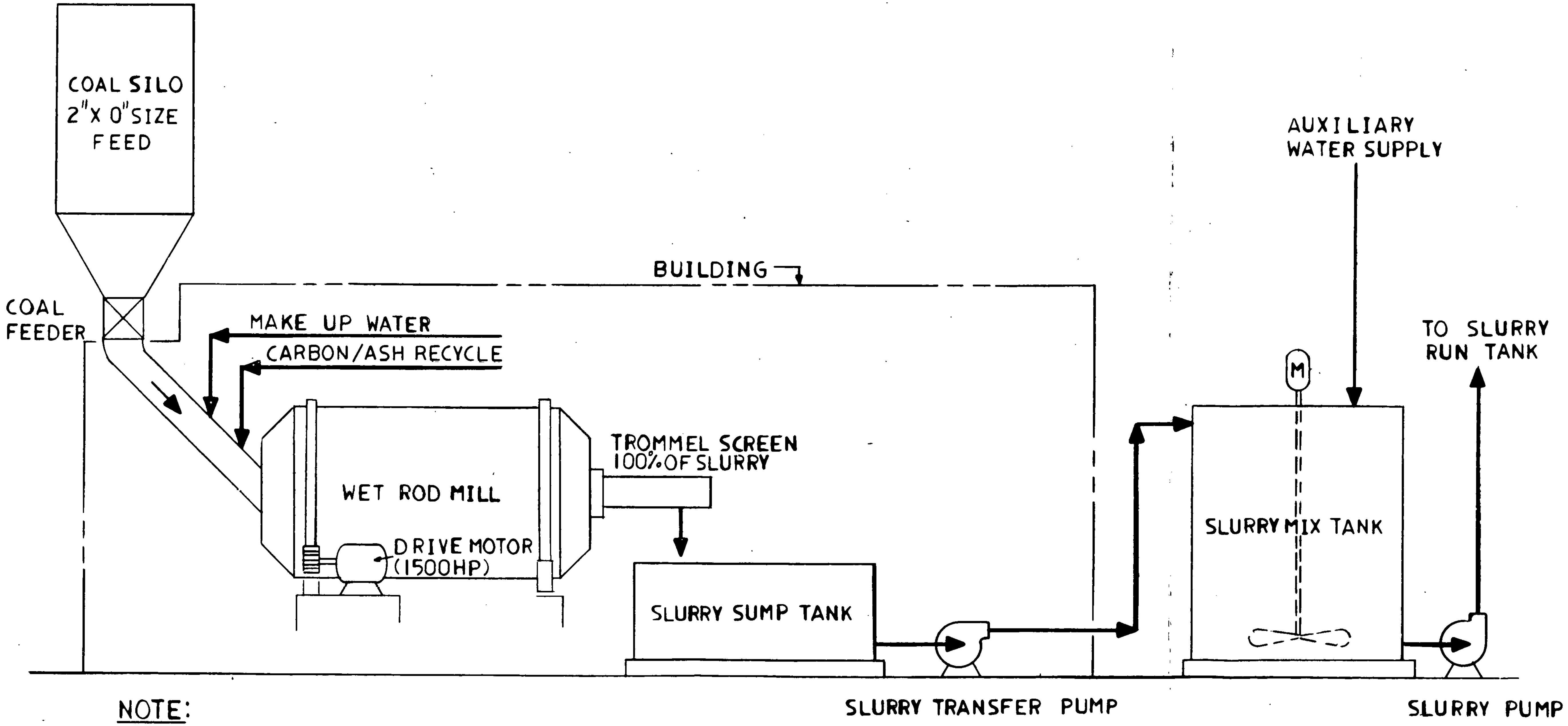
1. THERE ARE THREE 50% CAPACITY BALL MILLS, SLURRY SUMP TANKS & SLURRY TRANSFER PUMPS, TWO 100% CAPACITY DISC FILTERS OR CENTRIFUGES & TWO 50% CAPACITY DRY CAKE CONVEYING SYSTEMS,
2. FOUR SLURRY MIX TANKS ARE SAME FOR ALL SCHEMES

SYNTHESIS GAS
DEMONSTRATION PLANT PROGRAM

EBASCO SERVICES INCORPORATED
DIV. MECH DR. EGS
CH.
DATE 8/10/78
APPROVED

COMMERCIAL PLANT
TOS III . METHOD OF COAL PREPARATION SCHEME II

EXHIBIT
5



NOTE:
THERE ARE THREE 50% CAPACITY ROD MILLS, THREE TROMMEL SCREENS, THREE SLURRY SUMP TANKS, SIX SLURRY TRANSFER PUMPS, ONE SLURRY MIX TANK WITH AGITATOR AND TWO SLURRY PUMPS.

SYNTHESIS GAS
DEMONSTRATION PLANT PROGRAM

8				4			
7				3			
6				2			
5				1			
REV	DATE	BY	APPROVED	REV	DATE	BY	APPROVED

EBASCO SERVICES INCORPORATED
DIV.______ DR.______
CH.______
DATE______
APPROVED

COMMERCIAL PLANT
TOS III METHOD OF COAL
PREPARATION SCHEME III

EXHIBIT
6

EXHIBIT #7 SIZE DISTRIBUTION OF COAL FEEDSTOCK TO BALLMILL

TEST RUNS	1 & 2	3 & 4	5,6 & 8
TOP SIZE	½"	¼"	¼"
FEED SIZE DISTRIBUTION			
U.S. SIEVE % RETAINED			
+ 40	82.5	71.5	77.1
- 40 + 100	10.5	15.8	13.2
-100 + 200	3.5	5.9	4.5
-200	3.5	7.1	5.2
	100.0	100.0	100.0

EXHIBIT #8 TEST RESULTS OF OPEN CIRCUIT BALLMILL OPERATION

TEST NUMBER	1	2	2	3	4	5	5	5	5	5	6
TIME @ SAMPLING MINUTES	30	50	115	60	60	50	65	50	65	-	30
TYPE OF GRINDING CIRCUIT	Open	Open	Open	Open	Open	Open	Open	-	-	-	Open
COAL FEED RATE TO MILL, LB/HR.	350	300	300	300	250	250	250	-	-	-	175
LOCATION (SAMPLE)	Mill Disch.	Mill Disch.	Mill Disch.	Mill Disch.	Mill Disch.	Mill Disch.	Mill Disch.	Filter Disch.	Filter Disch.	Filtrate (H_2O)	Mill Disch.
% SOLIDS	38.0	42.0	53.0	53.0	49.0	34.0	33.0	62.0	63.0	1.0	40.0
% MILL CRITICAL SPEED	77	77	77	77	77	66	66	-	-	-	66
MILL H P/ TON OF FEED	18.9	19.2	19.2	23.1	28.4	261	26.1	-	-	-	35.4
U.S. SIEVE ANALYSIS											
WT% RETAINED											
+ 16	4.8	5.3	4.7	1.2	1.4	6.5	4.4	0.4	0	0.0	5.0
- 16 + 40	4.3	4.5	3.1	3.6	2.3	5.0	4.3	2.0	2.3	0.0	3.0
- 40 +100	21.4	20.8	19.0	17.2	21.2	23.2	20.8	20.5	20.6	11.1	14.0
-100 +200	12.4	20.8	22.2	20.6	19.8	20.8	21.6	21.4	23.0	7.4	23.3
-200 +325	20.1	10.4	10.5	11.5	9.2	9.4	11.5	12.1	12.1	11.1	14.2
-325	37.0	38.2	40.5	45.9	46.1	35.1	37.4	43.6	42.0	70.4	40.0
MEAN PARTICLE DIAMETER (MM)	.317	.334	.253	.133	.152	.331	.257	.119	.115	-	.249

EXHIBIT # 9 TEST RESULTS OF CLOSED CIRCUIT BALLMILL OPERATION

TEST NO	8	8	8	8
TIME @ SAMPLING MINUTES	60	60	60	60
TYPE OF GRINDING	Closed	Closed	Closed	
COAL FEED RATE TO MILL #/hr	400	-	-	
LOCATION (SAMPLE)	Mill Discharge	Classifier Overflow	Classifier Underflow	Filter Cake
% SOLIDS	43	13	68	75
% MILL CRITICAL SPEED	77	-	-	-
MILL H P/TON OF FEED	16.0	-	-	-
US SIEVE ANALYSIS				
WT % RETAINED				
+ 16	11.0	0	33.8	2.7
- 16 + 40	13.5	5.4	29.0	10.1
- 40 +100	30.3	33.3	24.8	36.6
-100 +200	15.8	20.7	6.2	17.5
-200 +325	7.2	9.1	1.4	8.3
-325	22.2	31.5	4.8	24.8
MEAN PARTICLE DIAMETER (MM)	-	0.168	-	0.272

COAL FEEDRATE - Does not include solids recycled to mill from classifier sand discharge

EXHIBIT #10 SIZE DISTRIBUTION OF COAL FEEDSTOCK TO RODMILL

SIZE	CUM. % PASSING
3/4"	100
5/8"	84.9
1/2"	56.5
3/8"	49.3
1/4"	42.8
6 Mesh	30.7
8	25.3
16	15.5
40	7.6
100	4.2
200	2.8
325	2.3

EXHIBIT #11

ESTIMATED INVESTMENT FOR COAL PREPARATION SYSTEM SCHEME I, II AND III

Scheme I (220 TPH Capacity Dry Cage Mill System)

Scheme I consists of the following items:

Three (3) dry cage mills, each 50% capacity.
Six (6) mill motors.
Three (3) conveyors each 50% capacity.
Three (3) bucket elevators, each 50% capacity.
Structural steel, foundations, electrical piping,, instrumentation, insulation and painting for above.

Pricing is based on (a) underflow from the Carbon Ash System being recirculated directly to the slurry mix tanks, and (b) no screening required for oversize from the cage mills.

Total installed cost = $3,600,000.

Scheme II (290 TPH Capacity Wet Ball Mill System)

Scheme II consists of the following items:

Three (3) wet ball mills, each 50% capacity.
Three (3) mill motors.
Three (3) slurry sump tanks, each 50% capacity.
Three (3) slurry sump pumps, each 50% capacity.
Two (2) centrifuge each 100% capacity.
Two (2) centrifuge main drive motors.
Two (2) screw conveyors, each 100% capacity.
Two (2) reversible belt conveyors, each 50% capacity.
Two (2) bucket elevators, each 50% capacity.
Two (2) belt conveyors with tripper, each 50% capacity.
Structural steel, foundations, electrical, piping instrumentation, insulation and painting for above.

Pricing is based on underflow from the Carbon Ash System being recirculated directly to the ball mills.

Total installed cost = $5,000,000.

Scheme III (290 TPH capacity Wet Rod Mill System)

Scheme III (290 TPH capacity Wet Rod Mill System

Three (3) rod mills, each 50% capacity
Three (3) mill motors and gear units.
Three (3) trommel screens.
Three (3) slurry pump tanks each 50% capacity
Six (6) slurry sump pump each 50% capacity

One (1) slurry storage tank complete with agitator.

Structural steel, foundations, electrical, piping, instrumentation, insulation and pointing for above.

Building as per sketch SK 8375-B-33 dated 2-26.79

Total installed cost - $3,800,000.

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY IV

AIR QUALITY CONTROL ALTERNATIVES

EBASCO SERVICES INCORPORATED
TWO RECTOR STREET
NEW YORK, N. Y. 10006
SEPTEMBER 1978

TABLE OF CONTENTS

I - INTRODUCTION

II - CONCLUSION AND RECOMMENDATION

III - TECHNICAL APPROACH

IV - ECONOMIC EVALUATION

V - EXHIBITS

APPENDIX A

I - INTRODUCTION

A. Objective

The objective of this study is to define and evaluate alternate Air Quality Control (AQC) Systems to determine an optimum process with respect to capital/operating costs, reliability of operation, technical viability, feasibility of change to accommodate size changes which may occur during overall plant design evolution, and general feasibility of integration with the overall plant. This evaluation includes a review of AQC System decisions incorporated into the original design as presented in the proposal. The AQC System must also be capable of achieving applicable emission standards with respect to particulate and sulfur dioxide (SO_2) pollutants discharged.

B. Scope

The scope of the study includes:

1. Review of available AQC system technologies. (For SO_2 removal the review is confined to recovery type Flue Gas Desulfurization (FGD) processes.)
2. Application of screening criteria to identify suitable processes/systems.
3. Development of a conceptual design based on the selected AQC systems including description of major subsystems and interfaces with the balance of the plant.
4. Identification of potential problem areas that may have an adverse impact on operational and performance reliability.
5. Comparative economic analysis of the alternatives with respect to investment and operating cost.

The conceptual design is sized for the Commercial Unit (3500 TPD ammonia). Applicability of the selected AQC system alternative for the Demonstration Unit (1200 TPD ammonia) is discussed but not evaluated.

The study does not address itself specifically to the conversion of the recovered SO_2 into saleable by-products. (To be included in Trade-off Study VI By-Product Sales Analysis.) It has been assumed for the purpose of material balances that the final Commercial Plant by-product will be elemental sulfur. In the event recovery of the sulfur is in another form appropriate changes in the material balance for final design will be required. The production of a by-product other than elemental sulfur is not expected to alter the results of this study.

The study also does not address nitrogen oxides (NO_x) control systems since NO_x control technology is related to design/operating parameters for the steam generator and emission limits with respect to NO_x will be met as part of the steam generator design. Selection of an AQC system is essentially unaffected by NO_x control technology.

II - CONCLUSION AND RECOMMENDATION

A. Recommendation

For particulate removal systems four alternate methods were considered:

- Wet Venturi Scrubbers
- Combination of Multicyclonic Mechanical Dust Collectors (MDC) and Wet Venturi Scrubbers
- Electrostatic Precipitators (ESP)
- Baghouses

For sulfur dioxide (SO_2)removal eleven alternate systems were considered:

- Magnesia Slurry Scrubbing
- Sodium Sulfite Scrubbing (Wellman-Lord)
- Ammonia (clear liquor)
- Citrate
- Phosphate (Aqua-Claus)
- Steam Stripping
- Aqueous Carbonate
- Ammonia (semi-dry)
- Carbon Sorption
- Copper Oxide
- Catalytic Oxidation

On the basis of this investigation it is recommended that a 99.65 percent efficiency ESP/Wellman-Lord System be utilized for this project.

This recommendation is made on the following bases:

1 - Elimination of high fly-ash concentrations at the venturi scrubber and at the fans which results in greater operating reliability for this equipment.
2 - Ability to bypass the FGD system.
3 - Overall system reliability with respect to performance and operability.

It is recognized that in spite of the recommended system not being the most economical it is nevertheless the preferred system. In the final design selection a lower efficiency ESP should be considered to reduce costs (making the ESP and MDC systems economically comparable) recognizing that by-passing of the FGD system would no longer be feasible.

A short description of the technical and economic basis for this recommendation follows.

B. Technical

1 - Particulate Removal

The proposed system is designed to meet an emission level of .05 lb/million Btu with a 99.65 percent removal efficiency.

ESP'S are the first alternative considered. They are the most widely accepted collection devices and if properly designed and maintained, are capable of 99+ percent removal efficiency and have demonstrated a high degree of availability. The coal characteristics point to the selection of a cold-side ESP (located downstream of the air heater).

For the next alternative, approximately 60 percent of the inlet particulates (primarily coarse) are removed in a mechanical dust collector (MDC) and the remainder in a wet scrubber.

For either of these two alternatives, a venturi scrubber is required to humidify the flue gas, to remove chlorides and also to remove any particulates remaining in the flue gas. The venturi particulate removal is negligible in the ESP alternative as the ESP is designed for 99.65 percent removal efficiency to meet the New Source Performance Standards (NSPS) emission levels. On the other hand, in the MDC alternative, a high efficiency (99+%) venturi scrubber is needed to meet NSPS.

Baghouses are generally not acceptable in high sulfur coal applications because potential sulfuric acid attack results in reduced bag life. Likewise an all "wet" system (venturi scrubbers alone) is considered undesirable for it requires that the fans be located downstream of the AQC System where they are susceptible to fouling and corrosion.

The design includes a by-pass of the Flue Gas Desulfurization (FGD) System. There are indications that EPA may permit its use during emergency situations when the FGD System is completely inoperable as long as the particulate emission levels are satisfied. Thus, the use of by-pass may be feasible with the design which includes the 99.65 percent efficiency ESP; the MDC alternative or a lower efficiency ESP would preclude its use, since particulate emission limitations would not be met.

The manner of particulate removal effects the selection of induced draft (ID) and booster fans. The ESP arrangement permits installation of high efficiency fans of standard construction. The MDC arrangement necessitates lower efficiency and special construction fans because they must handle relatively "dirty" flue gas. Their reliability can be expected to be lower when compared to the fans following an ESP.

2 - Sulfur Dioxide Removal

The Wellman-Lord Process is based on the relatively simple technology of sodium sulfite scrubbing. All its components have been well defined and optimized to a point at which a high level of reliability with respect to performance and operability can be expected as has been demonstrated on installations in the US and in Japan.

The Wellman-Lord Process, developed by Davy Powergas Inc., is therefore the alternative that meets the established selection criteria with respect to technical feasibility and applicability on commercial sized units.

All other recovery Flue Gas Desulfurization (FGD) systems, with the exception of Magnesia Scrubbing, have not reached the stage of development that permits a scale-up to a commercial size unit with a high degree of confidence. The Magnesia Slurry Scrubbing Process, while demonstrated commercially, is characterized by several yet to be resolved technical problems. In addition, its long-term reliability has not been satisfactorily proven.

For these reasons, the risk factors with respect to Magnesia Slurry Scrubbing are deemed significant at the current stage of development.

With certain modifications in the Wellman-Lord System, the overall conceptual design is acceptable for the Demonstration Unit. Single-effect in place of double-efffect evaporators may be the economic choice because of less total SO_2 removed. A one module system may also be feasible provided that only one boiler is used.

Coal properties are based on design conditions developed for the Commercial Plant coal gasification facilities. The range of coal characterics of the ultimate coal sources will influence the final design of the AQC system for the Demonstration Plant. However, this should not alter the conclusions regarding the type of control equipment selected in this report.

C. Economic

1 - The investment and annual operating costs for the alternative systems for particulate removal and sulfur dioxide removal are summarized in the following table. Details of this analysis may be found in Section IV of this study.

	$1000 (1978)	
	ESP/Wellman-Lord	MDC/Wellman-Lord
Investment		
Particulate Removal	5,340	2,155
Sulfur Dioxide Removal	19,900	19,900
Total	25,240	22,055
Differential	3,185	Base
Annual Operating Cost		
Particulate Removal	754	483
Sulfur Dioxide Removal	5,191	5,191
Total	5,945	5,674
Differential	271	Base

The results indicate that for particulate removal, the economics favor the MDC approach over the high efficiency (99.65%) ESP arrangement. As mentioned previously, the ESP system offers other advantages which increase the system reliability and operating performance to offset the apparent economic disadvantages.

2. Lowering the ESP collection efficiency from 99.65 percent to 90 percent results in drastic reductions in the cost of the ESP. If costs are expressed on the same basis as in (1) above, the comparative costs are as follows:

	ESP Efficiency	
	99.65%	90.0%
Investment, $1000	5,340	2,480
Annual Operating Cost, $1000	754	463

Thus, at the ESP 90 percent efficiency level, the economics of the ESP and MDC arrangements are comparable. It must be emphasized that this comparative analysis is predicated on the assumption that a reduction in ESP efficiency does not necessitate a higher pressure drop across the Wellman-Lord venturi scrubber and concomitant increase in energy cost.

4. Escalation of costs (by 20.36 percent for all investment and operating costs except for purchased power cost which has been escalated by 27.0 percent) to year 1981 has no appreciable effect on the relative economics of the evaluated alternatives, as shown below:

	$1000 (1981 Basis)	
	ESP & Wellman-Lord	MDC & Wellman-Lord
Annual Operating Cost	7 188	6 859
Differential	+329	Base

III-TECHNICAL APPROACH

A. GENERAL

The AQC System for the Commercial Unit will be designed to remove particulates and sulfur dioxide from flue gas discharged from two steam generating units, each rated at 403 000 lb/hr steam, 1500 psig and 940 F. The high pressure steam is used for driving various compressor turbines.

In addition to treating the boiler flue gases, the FGD system will be also capable of removing sulfur dioxide from the tail gases generated by the Claus Unit. The Claus Unit flow represents approximately 4.6% of total flow to the FGD system.

The boilers are equipped with an economizer section and a Ljungstrom type air preheater. Steam soot blowers are provided for the coal-fired boilers. Each furnace is designed for balanced draft firing and is served by a full capacity forced draft fan and two 50 percent capacity induced draft fans. Three 50 percent capacity ball type pulverizers are provided for each boiler, which are rated at 16.5 tons per hour. Each mill feeds three burners in the boiler. Primary air fans are used to sweep the mills for the boiler. The boiler efficiency is estimated to be 80 percent.

The steam generating units will be fired with the same coal as is used in the gasification process. A typical analysis is shown in Exhibit 1. At design conditions the total boiler coal firing rate will be approximately 65 tons per hour.

The purpose of this study is to define the alternative control technology and to select a system that controls SO_2 and particulates to meet the required emission levels. Detailed examination of the technical features of the selected control system is presented with respect to system operability, reliability and interface constraints. The economic evaluation represents a development of the capital and operating costs of selected systems. The energy requirements are developed in conjunction with the technical and economical evaluations.

B. REGULATORY REQUIREMENT

1. General

On August 7, 1977, the President signed into law the "Clean Air Act Amendments of 1977" (CAAA). These amendments significantly strengthened the "Clean Air Act" and have had a distinct and measurable impact on the planning, scheduling and economics associated with new facilities subject to the provisions of the Act. Of special concern is the fact that the amended law and the regulations which the Environmental Protection Agency (EPA) has proposed to comply with the law include regulatory constraints such as Prevention of Significant Deterioration (PSD) and Emission Offset. EPA now intends that these two constraints will be applied to most sources with potential emissions (of any pollutant regulated under the Clean Air Act) in excess of 100 tons per year. The CAAA and the proposed regulations of EPA also include provisions concerning emissions limitations, ground level concentrations, preconstruction monitoring, and stack heights.

This study considers only the emission constraints with respect to SO_2 and particulates as imposed by the New Source Performance Standards (NSPS) limitations which are essentially the minimum Best Available Control Technology (BACT) requirements. Other constraints, which are necessarily site specific, will have to be addressed at a later date when site metereology and topography are established and a complete environmental evaluation is feasible. Such evaluation will not only consider the environmental impact of pollutants for which NSPS limitations are not eatablished (e.g. carbon monoxide) but may also dictate controls for SO_2 and particulates that are more restrictive than NSPS in order that the emissions are in compliance with all applicable regulatory requirements.

2. New Source Standards of Performance

The CAAA requires EPA to promulgate revised New Source Performance Standards (NSPS) for fossil fuel fired stationary sources. The revised standards of performance are to include the imposition of two specific requirements: (1) The establishment of allowable emission rate limitations; and

(2) A requirement that the source achieve a specific percentage reduction in emissions.

The draft of revised NSPS for fossil fuel fired utility boilers were circulated for public comment in November, 1977; they are presented in Exhibit 2. Since the issue date, the details respecting the NSPS have been a constant source of controversy and as such have been in a state of flux. Since it is likely that the standards may be modified during the rulemaking proceedings, the revised NSPS shown in Exhibit 2 can be used as a guide only. Promulgation of the final standards is now expected in September 1978. EPA has not proposed revised NSPS for fossil fuel fired industrial boilers but has indicated that the industrial boiler NSPS will generally not be more stringent than those NSPS being considered for utility boilers.

The State of Kentucky Air Pollution Control Regulations include provisions which limit emissions of particulates and sulfur dioxide from fossil fuel combustion units. For boilers with a heat input of 250 million Btu per hour or greater, the limits are 0.10 pounds of particulates per million Btu input and 1.2 pounds of sulfur dioxide per million Btu input. The Federal NSPS for boiler emissions are more restrictive, and compliance with NSPS for boiler emission will insure compliance with the Kentucky boiler limitations.

With respect to the opacity standard, although the 10% requirement is included in the revised standards, EPA may be flexible in its enforcement. Current indications are that EPA may consider the particulate standard to be the controlling factor. For sources which meet the particulate emission level but exceed the opacity requirement, EPA may establish a higher opacity standard which corresponds to the compliance particulate emission level.

3. Prevention of Significant Deterioration (PSD)

EPA's PSD Regulations have been adopted for the purpose of preserving the air quality in areas in which the existing air quality is better than that established by the National Ambient Air Quality Standards (NAAQS). The regulations require that before construction can commence on a major facility, a PSD construction permit must be obtained. The application for this permit must be supported by an analysis which demonstrates that the emissions from the facility will not cause air pollution levels in excess of any NAAQS and will not result in increases in air pollution levels beyond certain increments specified in the Amended Clean Air Act.

The PSD application must also include a demonstration that the air pollution control systems proposed for the facility will use BACT. BACT is considered to be the maximum degree of emission reduction possible with considerations given to energy, environmental, and economic impacts. It should be noted that BACT can never be less stringent than any applicable NSPS.

Any PSD permit application submitted after August 7, 1978, must be supported by continuous air quality monitoring data collected for the purpose of determining whether the emissions from the proposed facility will cause pollutant concentrations in excess of the allowable PSD increments or the NAAQS. The continuous air quality data are to be gathered for a period of one year preceding the date of the application. The CAAA also state that monitoring periods of less than one year may be allowed if the reviewing agency, in accordance with regulations proposed by EPA, determines that a complete and adequate analysis can be conducted with less than a full year of data.

Projected emissions for the proposed source as shownin Exhibit 3 indicate that the plant will be a major source (potential emissions greater than 100 tons/year) with respect to sulfur dioxide, particulates and NO_x. Therefore the plant is subject to the PSD regulations and the associated BACT analysis for these pollutants.

4. Emission Offset Policy

Areas which are not meeting the NAAQS have been designated as non-attainment areas. Sources exceeding EPA's minimum size criteria and having the potential to cause a significant impact upon a non-attainment area, will be subject to the Offset Policy for those pollutants for which the area is designated non-attainment. The Offset Policy requires that the new emission from the proposed source be "traded off" against emissions from an existing source at a greater than a one to one ratio, with trade-off also resulting in a net air quality benefit for the region. An additional requirement of the Offset Policy is that the proposed source must employ controls which will provide for the Lowest Achievable Emission Rate (LAER) of the non-attainment pollutants. LAER is essentially the most stringent emission rate being required or achieved in the United States and may well be stricter than the emission rates associated with BACT or NSPS.

5. Stack Height Limitation

Provisions of the CAAA and EPA proposed regulations restrict the stack height that can be used for demonstrating compliance with NAAQS and PSD requirements. The stack height used in atmospheric dispersion modeling studies cannot exceed a "Good Engineering Practice" (GEP) stack height. necessary to avoid excessive pollutant concentrations in the vicinity of the source due to atmospheric downwash created by nearby structures or terrain features. The CAAA indicates that the stack used in modeling analyses for demonstrating compliance with the CAAA and EPA regulations may not exceed 2-1/2 times the height of nearby structures.

C. AQC SYSTEM ALTERNATIVES

Currently the most commonly used AQC System involves the electrostatic precipitator for the removal of particulates and the limestone/lime throwaway FGD System for the removal of sulfur oxide. The electrostatic precipitator is considered a viable alternative for the Synthesis Gas Demonstration

Plant Program. However, the throwaway FGD Systems have not been included in this study because the overall concept of the Project is predicated on the conversion of coal to useful products. Throwaway processes generate mixed sulfite/sulfate salts of calcium or sodium which are of little commercial value. Exclusion of throwaway processes limits the FGD selection to processes which recover SO_2 in useful forms such as sulfuric acid or elemental sulfur and regenerate the absorbent used for the removal of SO_2 from the flue gases. The on-site availability of a reductant (required in a number of recovery processes for conversion of SO_2 to elemental sulfur) is an important consideration that favors the recovery process option. Since a reductant, in the form of H_2S, will be available on site, installation of a recovery FGD system is a logical approach.

1. Particulate Removal

The alternate methods of flyash removal from flue gas are as follows:

- wet venturi scrubbers (all "wet" system)
- combination of mechanical dust collectors (MDC) and venturi scrubbers
- electrostatic precipitators (ESP)
- baghouses

Venturi scrubbers are not commonly used for primary control of particulates unless it is in conjunction with wet SO_2 removal systems. Advantages of venturi scrubbers are their relative insensitivity to coal chemical composition and to variations in flue gas temperatures. On the other hand, the fractional collection efficiencies decrease rapidly with decreasing particle size in the sub-micron range. Since no current theory allows particle size distribution to be predicted for a new installation on a wide range of coal sources, confidence in the performance capability in the absence of pilot testing is not as high as for the alternative particulate control methods.

Major drawbacks inherent to "wet" scrubbing are as follows:

- The fans can no longer be operated dry, creating potential for corrosion and imbalance. Even if located downstream of the AQC System, fans are susceptible to fouling due to mist eliminator carryover.
- The FGD System cannot be by-passed.
- The scrubber must be protected against erosion and abrasion due to high flyash content in the flue gas.
- The ability of scrubbers, at a reasonable pressure drop, to meet emission levels of less than 0.05 lb/million Btu has not been fully demonstrated.

The operating reliability of most existing "wet" particulate removal systems has been adversely affected due to corrosion, abrasion and plugging problems. An acceptable level of reliability can be achieved, at an economic penalty, with the prudent selection of materials of construction and vigorous maintenance efforts.

An alternate scheme is the combination of dry and wet particulate removal. Approximately 60 percent of the flyash is collected dry in a multicyclonic mechanical dust collector (MDC) and the remainder in the wet scrubber. It lessens but not eliminates the drawbacks of an all "wet" system listed above. The major function of the dust collector, aside from the removal of coarse particles, is to permit locating the fans upstream of the scrubber and the FGD System, allowing the higher efficiency fans and avoiding the corrosion potential associated with downstream location.

Electrostatic precipitators (ESP) are the most commonly used devices for high efficiency removal of particulates from the combustion gases of coal-fired steam generators. The size of the ESP, and hence the cost required to meet a given level of emission control, varies with the characteristics of the coal ash. Ash resistivity is a major factor affecting the ESP size.

Larger ESP systems are needed as the resistivity of flyash increases and the levels of emission control are reduced. One of the key variables affecting the resistivity of flyash is the sulfur content in the flue gas. Flue gas with low sulfur oxide concentrations has a high flyash resistivity in the 250-350 F temperature range (typical temperatures of the gas exiting the air heater). However, the same flyash when subjected to an electrostatic precipitation field in the 600-750 F range undergoes a major decrease in resistivity. Therefore, high resistivity flyash (low sulfur coal) is normally easier to precipitate in a hot-side ESP, located on the hot side of the air heater, while low resistivity flyash (high sulfur coal) favors the installation of the ESP downstream of the air heater (cold-side ESP). Wherever the precipitation characteristics do not require a hot side ESP, a cold-side ESP is generally an economic choice.

Baghouses have been applied extensively to various industrial processes, but until recently represent the least applied particulate removal device for coal-fired boilers. The renewed interest in baghouses has been brought about by the more stringent emission regulations requiring 99.9 + percent removal efficiency in some applications. Since a baghouse is capable of such high removal efficiencies at a minimal increase in capital investment (unlike ESP where costs increase substantially with efficiency), baghouses have been penetrating the precipitator market in recent years. This penetration has been almost exclusively in high resistivity ash (low sulfur) applications where ESP is no longer competitive due to very large Specific Collection Area (SCA) requirements.

On the other hand, the need for baghouses in much less pronounced in high sulfur (above 3%) coal applications because the collection of low-resistivity flyash from these coals does not require SCA's in excess of 550-600-the range below which the economics generally favor a cold-side ESP over a baghouse. Nearly all currently operating baghouses have been designed for coal sulfur levels of 1 percent or less with only pilot plant data available on the impact of operation at higher sulfur levels. The major concern has been the durability of fabric filters. Under some operating conditions, particularly at low loads, the flue gas temperature can easily approach the acid dewpoint at which the fabric filters are exposed to the corrosive attack of sulfuric acid. The necessity to remove and replace the bags on a periodic basis, and its concomitant negative economic impact, has been one of the major drawbacks of baghouses. A two year bag life appears to be a reasonable assumption considering the current state-of-the-art of the fabric technology. Fiberglass bags have been used on most coal-fired boilers.

In the installation under study, it is projected that coal sulfur level will not be less than 2.5 percent and will be in excess of 3 percent at design conditions. In the absence of operating experience of baghouses on coal-fired boilers at such high sulfur levels, it is not prudent to consider baghouses as a viable particulate control technology for this application. The exclusion of baghouses can be further justified on economic grounds as discussed in Section IV where it is shown that the economics in terms of operating costs favor a cold-side ESP over a baghouse. It must be emphasized that these conclusions are predicated on current technologies and the relative merits of baghouses and precipitators may be subject to future reevaluation based on developments in fabric technology pointing to a longer bag life.

Of the options available, the cold-side ESP and the combination of mechanical dust collector/venturi scrubber are considered viable alternatives and are examined in more detail in Section III-E.

2. SO_2 Removal - FGD Systems

As previously indicated, the throwaway FGD processes are not being considered as alternatives for SO_2 control in this project because of the expressed intent to recover sulfur in the form of saleable by-products. Exclusion of throwaway processes, which are the more widely accepted and technically developed SO_2 control systems, narrows the list of alternatives to recovery process. A large number of FGD recovery processes are currently at various stages of development ranging from laboratory to full commercial sized facilities.

Several of the more important recovery processes are listed in Exhibit 4. These processes are broken down into a number of logical categories. The first level of categorization is whether the process operates wet or dry or, recently, semi-dry. This distinction provides some indication of the characteristics of the technology employed. Wet technology usually implies that the dirty flue gas is contacted with a large and generally recirculated flow of absorbent which absorbs the SO_2 and cools the gas by evaporation of water to a temperature slightly above the water dew point. It is the usual practice to reheat the saturated flue gas prior to its discharge into the stack. Wet technology implies materials handling by pumping, low temperature operation and corrosion/materials of construction as major problem areas.

Dry technology usually implies high temperature operation, materials handling by mechanical or pneumatic conveying and abrasion/erosion rather than corrosion as the major areas of concern. Dry processes have an advantage in not requiring stack gas reheat because the flue gas is not contacted by water. The recently introduced semi-dry technology involves contacting the flue gas by small quantities of aqueous absorbent in a spray dryer followed by dry collection in a baghouse or ESP of both the SO_2 reaction products and the particulates. In this case, the absorbent is dried, the flue gas is only partially cooled, and reheat can generally be avoided.

The second distinction, which applies only to wet technology, is whether the absorbent liquid contacting the flue gas is a slurry or a clear liquor. The use of slurries generally implies abrasion, deposition and scaling as additional operating problems.

Finally, the third classification refers to the suitability of the process to produce either sulfur or sulfuric acid or both.

With this philosophy of classification, eleven recovery FGD processes have been listed in Exhibit 4. Although the status is not indicated, essentially all have progressed to the 1 MW equivalent pilot plant size and have been reported in the open literature.

In order to reasonably address the large number of possible technology alternatives, it is necessary to establish a logical set of selection criteria which can be applied as a screening procedure to identify that techology considered suitable for this application. The selection criteria and sequence of application are listed below:

Process Development Status

a) Successful operation of 100 MW equivalent size class
b) Existence of qualified supplier

Process Capabilities and Requirements

a) Emission level performance capability
b) Environmental acceptability of process waste products
c) Acceptable interface with the balance of the plant

Technical and Economic Feasibility

a) Technical feasibility
b) Economics of investment and operation
c) Energy requirements

The application of the first selection criteria (Process Development Status) implies demonstrated capability to engineer and design equipment in a modular size range typical of commercial sized equipment without undue scale-up. Only two processes, Wellman-Lord and Magnesia Slurry can be categorized as having been applied commercially.

The Wellman-Lord process is a first-generation recovery FGD process which has been applied in this country and in Japan for SO_2 removal from Claus and sulfuric acid plant tail gases and from oil-fired power plant flue gases since the early seventies. The most recent installation, and the most significant for this study, is the 115 MW unit at Northern Indiana Public Service Company's (NIPSCO) Dean H Mitchell Station. It represents the first coal-fired application; after having successfully completed a short-term performance test, it is currently undergoing a comprehensive one year demonstration program. In addition, several Wellman-Lord systems are now under construction or in design stages: three at Public Service Company of New Mexico's San Juan Station (1715 MW total); one system at the 55 MW unit at Getty Refining Co's Delaware City coal-fired boiler; and one system at a unit treating 250 000 ACFM of flue gas from ARCO Polymer Co's coal-fired industrial boiler.

The Magnesia (MgO) Slurry Scrubbing Process is also a first-generation recovery FGD process. SO_2 removal is achieved by scrubbing with an aqueous solution of MgO to produce a by-product slurry of magnesium sulfite which is concentrated, dried and shipped to a reprocessing plant for regeneration. Magnesium sulfite, along with coke for reduction of any magnesium sulfate, is calcined in a rotary kiln to produce SO_2 gas as feedstock to a sulfuric acid plant with the regenerated MgO returned to the absorption system for reuse.

The Chemico version of the MgO Process operated intermittently as a 150 MW prototype on Boston Edison's Mystic No. 6 Unit from 1972 to 1974 for about 3000 operating hours. Primary problems were of a material handling nature. Operating time was judged insufficient to develop reliable data on regeneration and transportation losses of reagent. The project was terminated in 1974. The process has also been tested at Potomac Electric Power Company's 95 MW Dickerson No. 3 Unit. Operating problems were comparable to those experienced by Boston Edison.

The United Engineers' version of the MgO process has been intermittently tested at Philadelphia Electric 120 MW Eddystone Station from 1974 to present. The test runs have experienced a multitude of problems of mechanical and chemical nature. Pending results of further operating experience, Philadelphia Electric intends to install an additional 500 MW capacity at the Eddystone and Cromby Stations.

There are three MgO process installations in Japan, none of which operate solely on boiler flue gas, and for which specific operating data have not been published.

Two processes, Carbon Sorption and Copper Oxide, have been operated on a prototype size scale (20-40 MW) for limited time periods. Their technical viability and scale-up capability have not been fully demonstrated.

The Citrate and Aqueous Carbonate Processes have undergone extensive pilot plant development but the integration of the various subprocesses has not been demonstrated at this time. A complete integration of the Citrate Process is going to be performed for the first time at St Joe Minerals Corp's 60 MW G F Weaton coal-fired electric generation station now under construction. Also, a program is now underway to integrate the Aqueous Carbonate Process on a 100 MW unit at the Niagara Mohawk Power Corp's Huntley Station with operating and testing due to begin in 1980.

All other FGD recovery processes are essentially confined to pilot plant stage development (up to 5 MW capacity). Design and operating data for these processes are not judged adequate for scale-up to a commercially-sized unit. Operation of prototype units is imperative in order to verify their technical viability.

It is apparent from the preceding analysis that the application of the process developmental status selection criteria reduces the list of FGD recovery process alternatives to two: the Wellman-Lord and the Magnesia Slurry Scrubbing Processes. A more rigorous application of this criterion, namely the <u>successful</u> operation of a 100 MW equivalent size unit, raises serious doubts as to the viability of the magnesia process at its current level of technical development. According to the study prepared by Radian Corporation for the Electric Power Research Institute (Evaluation of Regenerable Flue Gas Desulfurization Processes, January 1977), the magnesia slurry scrubbing process "still faces many problems both of a chemical and equipment nature. Thus far process operations have been aimed more at getting the process to run after it has been built rather than developing the basic chemical kinetic data which might help understanding the process". The Radian study further suggests that more investigation be carried out in the recovery section of the process with respect to equipment design, precipitation characteristics of $MgSO_3$ hydrates and dissolution rates of recovered MgO. The low reliability of the U S installations suggests that the magnesia scrubbing process needs improvements before it can be applied on new units requiring a high degree of reliability.

On this basis, it is felt that the Magnesia Slurry Scrubbing be excluded from further consideration, leaving the Wellman-Lord Process as the only viable alternative.

D. WELLMAN-LORD PROCESS

In the preceding section it was established that the Wellman-Lord technology is the only recovery FGD process that has been adequately demonstrated on a commercial scale. This section presents the salient features of the Wellman-Lord Process, and its acceptability in terms of the established technical selection criteria and environmental constraints.

The information presented herein is based on data furnished by the developer of the Wellman-Lord Process, Davy Powergas Inc, and on data available from open literature.

1. Process Description

The Wellman-Lord Process is based on the aqueous absorption of SO_2 by sodium sulfite to form sodium bisulfite. The scrubbing liquor is thermally regenerated to produce an SO_2 rich stream which can be converted into sulfuric acid or elemental sulfur. The regenerated absorbent is returned to the absorber. Sodium sulfate produced by oxidation in the absorption process must be purged from the system. A solution of soda ash must be added into the system to replenish sodium losses resulting from the purge of sodium sulfate.

The process consists of four basic functional subsystems: gas pretreatment, SO_2 removal, purge treatment, and absorbent/SO_2 recovery.

Gas Pretreatment

This subsystem serves two basic functions: to humidify the inlet flue gas and to effect particulate and chloride removal. The level of particulate removal depends on the type of particulate removal equipment that precedes the FGD system. A venturi-type prescrubber effects both the humidification and solid removal functions. Continuous purge from the prescrubber recirculating loop is required to maintain desired suspended and dissolved solids levels. This bleedstream is then routed to the waste disposal pond.

SO_2 Removal

Humidified gas (at approximately 130 F) enters the absorption tower where it is contacted with the recirculating sodium sulfite-bisulfite solution to effect the required SO_2 removal.

The principal reaction in the absorber is between SO_2 in the flue gas and sodium sulfite in the absorbing solution:

$$SO_2 + Na_2SO_3 + H_2O \longrightarrow 2NaHSO_3$$

The bisulfite anion HSO_3 is found only in solution. When water is removed from the sodium bisulfite solution, a solid sodium pyrosulfite ($Na_2S_2O_5$) is formed:

$$2\ NaHSO_3 \longrightarrow Na_2S_2O_5 + H_2O$$

Some oxidation of the sodium sulfite occurs by oxygen in the flue gas and by absorption of SO_3 from the flue gas:

$$Na_2SO_3 + 1/2\ O_2 \longrightarrow Na_2SO_4$$

$$2\ Na_2SO_3 + SO_3 + H_2O \longrightarrow 2\ NaHSO_3 + Na_2SO_4$$

The sodium sulfate (Na_2SO_4) must be removed from the absorbing solution in the purge treatment area.

The cleaned gas passes through a mist eliminator and is reheated prior to being discharged to the atmosphere.

Purge Treatment

The spent absorbing solution leaving the absorber is split into two streams: the main stream is pumped to the evaporators for SO_2 recovery while a slip stream is directed to the high temperature purge crystallizer for removal of sodium sulfate by-product. In the purge crystallizer, the solution is heated in a shell and tube exchanger by condensing low-pressure steam. An internal liquid solid separation chamber is incorporated in the design in order to maintain a relatively high solids content in the slurry product. The slurry is fed into a centrifuge where most of the remaining liquor is removed and the resultant cake is dried by steam in a rotary type dryer. The crystalline product, a mixture of sodium sulfite, sodium sulfate and small amounts of sodium thiosulfate and sodium pyrosulfate, is pneumatically conveyed to the sulfate purge bin for storage. The mother liquor overflow from the purge crystallizer and the centrifuge liquor is recycled to the absorber product liquor stream entering the evaporator loop. Vent gases from the dryer are passed through an eductor-type vent gas scrubber to remove any remaining SO_2 and dust particulate before discharging to atmosphere or to the inlet flue gas stream.

Absorbent SO_2 Recovery

Regeneration and SO_2 recovery involves a simple reversal of the absorption reaction by addition of heat:

$$2\ NaHSO_3 \longrightarrow Na_2SO_3 + SO_2 + H_2O$$

However, higher temperatures also increase the formation of sodium thiosulfate by a disproportionation reaction:

$$6\ NaHSO_3 \longrightarrow 2\ Na_2SO_4 + Na_2S_2O_3 + 2\ SO_2 + 3\ H_2O$$

These regeneration reactions occur in the double effect evaporator. In the first effect, the rich absorbing solution is heated in a shell and tube exchanger by condensing low-pressure steam. In the second effect, the solution is heated by condensing overhead vapors from the first effect evaporator and from the purge crystallizer. In the evaporators, which operate under a vacuum, SO_2 and H_2O vapors are released while Na_2SO_3 crystals precipitate from the solution. The slurry product from each evaporator is discharged by gravity to the dissolving tank. Overhead SO_2 and H_2O vapors from the evaporators are subjected to partial condensation to remove most of the water and thus concentrate the SO_2. The condensate flows to the stripper for removal of dissolved SO_2. The stripped condensate is returned to the dissolving tank to redissolve the Na_2SO_3 crystals and dissolve the make-up sodium carbonate. Soda ash in the dissolving tank reacts with sodium bisulfite to form additional sodium sulfite:

$$Na_2CO_3 + 2\ NaHSO_3 \longrightarrow 2\ Na_2SO_3 + H_2O + CO_2$$

The combined regenerated solution from the dissolving tank provides absorber feed.

The SO_2 exiting from the condensers is compressed and cooled by a rotating liquid ring compressor. The resultant gas-liquid mixture flows to the knock-out drum where the two phases separate. The SO_2 rich (96.5% SO_2 and 3.5% H_2O) gas is then available for conversion to the desired byproduct.

2. Process Capabilities

In general, the Wellman-Lord process is quite simple and consists of unit operations which are understood. It has been operated successfully on different SO_2 sources and modifications are still being made (primarily in the purge treatment area of the process).

Sulfur dioxide removal efficiencies in exceess of 90 percent have been achieved with all the operating Wellman-Lord FGD Systems. Removal efficiencies in excess of 98 percent have been reported at units installed in Japan; they have exceeded 97 percent at sulfuric acid and Claus units in the U S and 90 percent at the acceptance trials completed at NIPSCO's D H Mitchell Station.

The process is capable of achieving high SO_2 removal levels due to the relatively high (as compared to calcium scrubbing e.g.,) affinity of sodium to absorb SO_2 and by varying the number of absorption stages used in the absorber. Like other clear liquor scrubbing processes, the absorption efefficiency is not limited by the slow dissolution of the absorbent and does not require high L/G's (Liquid to Gas ratios)that are characteristic of lime/limestone slurry scrubbing.

As the absorber does not recirculate slurry and prescrubber recirculating slurry operates as a separate loop, scaling in the SO_2 absorber has not been reported as a problem in any Wellman-Lord installation. Scale-free operation enhances the reliability of the absorption system.

An important consideration in adapting the Wellman-Lord process to coal-fired plants is the removal of particulate matter and chlorides ahead of the absorber. Primary removal of flyash is generally provided by an ESP or by a venturi or tray type prescrubber. Regardless of what type of primary removal equipment is used, a prescrubber is always required to humidify and cool the flue gas and also to remove chlorides that cannot be tolerated in downstream processing. In a well designed prescrubber, 99^+percent of the chlorides are removed. As a result of low pH (1-2) conditions, the prescrubber must be constructed of acid-resistant material.

The design area of major concern is the oxidation of the sodium sulfite to the unreactive sodium sulfate. Its formation requires a purge from the absorber to maintain the level of reactive sodium sulfite and to reduce the possibility of fouling of evaporator surfaces. The sulfate purge can amount to 5 to 10 percent of total sulfur removed which means higher make-up costs and need to dispose of relatively large quantities of sodium sulfate by-product. Attempts to reduce oxidation by means of organic antioxidants have been abandoned because of high cost. Selective removal of sodium sulfate by chilled-wall crystallizers has been more successful, resulting in a five to six-fold decrease in the purge stream. This purge treatment has been employed in all recent Wellman-Lord installations including NIPSCO. The latest development (proposed for this project) involves the use of high-temperature purge crystallizers with the purpose of reducing energy requirements. This represents the only subsystem that has not been proved on commercial scale at this time.

Despite design improvements, the recovery area of the Wellman-Lord process remains a major consumer of energy as significant amounts of steam are required in the evaporators and SO_2 strippers. The evaporators may be either a single or double effect type, the amount of SO_2 removed being the governing factor affecting the choice.

Water make-up is required to replenish water losses due to evaporation in the prescrubber, loss in the product SO_2, drying of purged solids and prescrubber blowdown. The only solid waste effluents are prescrubber blowdown (primarily flyash) and the sodium sulfate purge. The latter has only limited commercial value although the sodium sulfate purge solids have been reported to be acceptable for paper industry consumption. Some adverse environmental impacts could be caused by dust emisisons from the dryer in the purge treatment area. However, a properly designed vent scrubber on the dryer should reduce this emission source to an insignificant level or, it can be eliminated entirely if the vent scrubber gases are recycled to the inlet flue gas stream.

Turndown (capability of operating at lower than design load) of the Wellman-Lord absorbers is estimated at 50 percent which is relatively high. Lower turndowns (to 30 percent) are feasible but only at the expense of pressure drop across the system.

Space requirements for the equipment are comparable to those for other regenerable FGD Systems. However, only the SO_2 removal equipment (prescrubbers and absorbers) need be near the boiler area while the remainder of the system can be situated at some other location without any major capital or operating cost increase.

The Wellman-Lord process can be easily adapted to treat tail gases from the Claus Unit together with the main boiler flue gas. The feasibility has been demonstrated at NIPSCO where the Claus tail gases from the Allied Chemical sulfur recovery plant are fed into the boiler flue gas ahead of the prescrubber for SO_2 removal.

In the proposed installation, waste steam from the various sources of the gasification process can be interfaced advantageously with the steam requirements of the Wellman-Lord process.

In summary, the Wellman-Lord process is a viable FGD alternative in terms of SO_2 removal capabilities and overall technical feasibility.

E. AQC SYSTEM CONCEPTUAL DESIGN

The AQC System design presented in this Section includes a cold side ESP for particulate removal and the Wellman-Lord FGD System for SO_2 removal. Alternate particulate removal by means of a mechanical dust collector is also described. The design is conceptual, and although representing a workable system, it by no means reflects the optimal system with respect to equipment configuration/sizing, process material flow and energy utilization. Design information on the Wellman-Lord process was obtained from its developer, Davy Powergas Inc, and was suppplemented, as required, by data available from literature sources. It must be recognized that much of the detailed design information including internal material flows and equipment sizing optimization, is proprietary to Davy Powergas and will not be generally available until the project advances beyond the conceptual stage and more comprehensive specifications are issued.

With regard to environmental requirements, the conceptual design is based on the best available information. However, as indicated in Section III.- B further analyses will be required which may cause design modifications including changes in removal efficiency.

1. System Description

Flue Gas Circuit

The AQC System for each steam generator consists of one (1) electrostatic precipitator, one (1) booster ID fan, and one (1) FGD train. One regeneration facility common to the two FGD trains is provided. The rationale for providing one train for each steam generator is predicated on reliability considerations: if any AQC System component of the train becomes inoperative, it will still be possible to treat at least 50 percent of the total flue gas flow.

As shown in Exhibit 5 the modular concept is applied also to the entire flue gas circuit as each steam generator is drafted by means of a separate set of ID fans. This arrangement is not only a logical consequence of furnishing two 50 percent steam generators but also consistent with the established design principle strongly favoring an independent drafting mechanism per steam generator.

The flue gas from each air heater is directed to a cold side ESP where 99.65 percent of the flyash is removed. After leaving the ESP, the flue gas flows through two parallel ID fans which draft the steam generator, the air heater and the ESP. A booster ID fan then delivers the flue gas into the FGD System and thence to the stack. A crossover plenum between the two trains is provided at the booster fan inlet permitting treatment of flue gas from either steam generator in one FGD train in the event the other train is inoperative. The tail gas from the Claus Unit is introduced into the flue gas stream at the booster fan inlet.

Upon leaving the Wellman-Lord absorber and prior to discharge into the atmosphere the saturated flue gas stream is reheated to 170 F in a mixing chamber by introducing heated air into the wet gas.

The ductwork is arranged to permit any or all of the flue gas to by-pass the FGD System and flow directly to the stack. The by-pass is intended for emergency purposes only. Its inclusion in the design is contingent upon future governmental regulations that may prohibit the usage of by-pass.

For the alternate particulate removal using a mechanical dust collector, the system configuration is identical, except that the ESP is replaced with a mechanical dust collector in each flue gas train.

Reagent/SO_2 Recovery and Purge Treatment

This portion of the system has already been described in Section III-D and is shown schematically in Exhibit 6 .

Conceptual layout of the AQC System equipment is shown on the plot plan in Exhibit 7.

2. Design Assumptions

The equipment is designed to treat the flue gases discharged from the two steam generators operating at 115 percent of MCR (MCR is defined as maximum continuous rating conditions of the steam generators) and fired with coal having the properties as shown in Exhibit 1. In calculating the composition of the flue gas entering the AQC System (Exhibit 1), 20 percent excess air and 10 percent air heater in-leakage were assumed.

The composition and flow rates of the incinerated Claus tail gas are shown in Exhibit 8.

Inlet SO_2 concentration was calculated on the assumption that 100 percent of the coal sulfur (Design Coal) is converted to SO_2. In calculating the inlet particulate loading it was assumed that 85 percent of the coal ash (Design Coal) is emitted as flyash while 15 percent is collected as bottom ash.

The inlet design parameters can be summarized as follows:

		Per Train	Total (2 boilers)
1.	Flue Gas to AQC System* (at air heater exit)		
	pounds/hour	833 500	1 667 000
	SCFM**	177 700	355 400
	ACFM @300F and -13 in WG	268 300	536 600
	SO_2, pounds/hr	4 415	8 830
	ppm by vol dry	3 070	3 070
	Particulate, pounds/hr	9 495	18 990
	grains/SCF dry	7.55	7.55
2.	Claus Tail Gas (incinerated)		
	pounds/hr	34 800	69 600
	SCFM**	6 900	13 810
	SO_2, pounds/hr	948	1 895
	ppm by vol dry	8 625	17 250

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* Includes 15 percent margin over MCR conditions

** SCFM = Standard Cubic Foot per Minute at 60 F and 408 inches WG.

3. Particulate Removal

The design inlet particulate loading is 7.55 grains/Standard Cubic Foot (gns/SCF) dry equivalent to 14.45 lbs/million Btu. As described in Section III-B the emission levels that this installation will have to meet are not fully defined since applicable New Source Performance Standards (NSPS) have not been promulgated by EPA. Of the two particulate standards that EPA is currently considering, 0.03 and 0.05 lb/million Btu, the latter has been selected for the purpose of this study. In the event the lower standard becomes applicable, either of the two particulate removal systems discussed in this section could be used - however, their investment and operating costs would be higher. To achieve 0.05 lb/million Btu emission level, 99.65 percent overall removal emission is required assuming the above inlet particulate loading.

Two alternate removal systems will be considered: cold-side ESP and MDC/Venturi Scrubber combination.

Two ESP's are included, one per flue gas train. Each precipitator has a Specific Collection Area (SCA) of 499 Ft^2/1000 ACFM, surface collection area of 133,900 Ft^2 and is capable of achieving the desired removal efficiency of 99.65 percent with 10 percent of the electrical bus sections out of service. For the purpose of this study, the discharge electrodes are based on the weighted wire design; the rigid frame design is equally suitable and may be considered in the final design.

It should be noted that the ESP is designed to meet the expected NSPS with no additional particulate removal required in the downstream prescrubber. Justification for this conservative approach is that the particulate-free flue gas may be vented through the emergency by-pass in the event the FGD system is not operational (provided that a variance can be obtained for short time periods). Generally, a very high efficiency ESP permits a low pressure drop downstream prescrubber with the resulting lower energy requirement to drive the booster ID fans. In this application however, the reduction in energy is not indicated for reasons discussed later.

If the requirement to meet the NSPS emission level by the ESP alone is waived, the ESP can be undersized. For example, lowering the removal efficiency to 99 percent, the size of the ESP is reduced by 17 percent (416 SCA); for 90 percent efficiency, the size can be reduced by 64 percent (180 SCA). Impact on investment of such reductions is discussed in Section IV.

Mechanical dust collectors (MDC) are not capable of particulate removal efficiency to meet NSPS because they are extremely ineffective in removing particulate less than 10 microns. In the 10+ micron size, removal efficiencies of 95 percent are feasible with a pressure drop of 3 inches W G. Typically, flyash from a pulverized coal-fired boiler averages 44 percent less than 10 microns. Thus, in the proposed alternative the MDC is expected to remove the bulk of the coarse particles, with the remainder being collected in the prescrubber. The mechanical dust collector removal efficiency of 60 percent was assumed (3 inches W G pressure drop) giving an inlet loading of 3 gns/SCF dry to the prescrubber. Verification of these assumptions will be required once a definitive flyash particle size distribution becomes available.

Using the MDC a 99.23 percent removal efficiency in the prescrubber is needed to meet the required overall system emission level of 0.023 gns/ /SCF dry (0.05 lb/million Btu). Davy Powergas have indicated that a 12 inch W G pressure drop across the prescrubber is needed to remove chlorides from the flue gas and that no additional pressure drop is required to reduce the particulates to the 0.023 gns/SCF dry level. This seems to be a somewhat optimistic assumption considering the predominance of small, difficult to remove particles in the flyash entering the prescrubber. Based on Ebasco's experience, a pressure drop in the range of 15-18 in. W G would be a more realistic estimate. However for the purpose of this conceptual design and in the absence of better information as to the characteristics of the flyash, the lower pressure drop, as proposed by Davy Powergas, will be used.

The use of MDC requires that ID fans and booster fans be of special construction to permit processing flue gas having relatively high particulate concentration. Compared to fans handling very clean flue gas (if ESP is used), the fans following the MDC's will be less efficient and will require frequent replacement of liners due to the errosive effect of fly-ash. In general, their overall operating reliability can be expected to be poorer.

Major design parameters of the two alternate particulate removal systems are presented in Exhibit 9.

4. SO_2 Removal

The required SO_2 removal applicable to this installation is likewise contingent on the final promulgation of NSPS by EPA and the results of a site specific environmental evaluation. SO_2 removal efficiency of 90 percent is assumed in this study.

The total SO_2 content in th flue gas entering the FGD system is as follows:

In Boiler Flue Gas:	8 830 lbs/hr (6.72 lbs/million Btu)
In Claus Tail Gas:	1 895 lbs/hr
Total	10 725 lbs/hr

Considering the emissions in the boiler flue gas only, ninety percent removal efficiency will result in SO_2 emission level of 0.67 lbs/million Btu - approximately 44 percent below the current maximum allowable NSPS limit of 1.2 lbs/million Btu.

To achieve the required SO_2 reduction Davy Powergas proposes a 3 - stage absorption system. The flue gas from the prescrubber flows upward and is contacted with the recirculating absorbing solution at each stage which in the proposed design is a valve tray. Each tray is individually recirculated to maintain adequate flow for good hydraulic characteristics and sufficiently large L/G ratio (est 1.7 gpm/1000 ACFM) for efficent mass transfer. The solution on the bottom tray overflows into the bottom section of the absorber and thence pumped to the absorber product surge tank. The top stage (tray) is fed with the regenerated solution which is pumped from the absorber feed surge tank. The absorber is a concrete tiled tower (20 ft by 20 ft and 60 ft high) with the trays constructed of 316 stainless steel.

As previously indicated, the Wellman-Lord process is capable of SO_2 removal efficiency in excess of 90 percent with only moderate impact on capital cost.

The pressure drop across each absorption stage is 3 inches W G and across the entire absorption system including the mist eliminator and ductwork is estimated at 15 inches W G.

Major equipment proposed for the Wellman-Lord Process is listed in Exhibit 10.

5. Reheat of Scrubber Flue Gases

When a wet scrubber is inserted between the air heater and stack, the flue gas exiting the scrubber is humidified and cooled to its saturation temperature. Discharge of the wet gas to the stack produces water condensation and corrosion in downstream equipment and impaired stack plume rise due to lower gas buoyancy. To correct these undesirable aspects of wet scrubbing, the treated flue gas is normally reheated to a temperature above its dewpoint. However, there is significant economic penalty associated with gas reheat, particularly in terms of high energy requirements. For this reason and also since the function and level of reheat are not clearly defined alternate approaches other than reheating are being currently considered. Most prominent of these alternatives is no reheat at all, that is, operating under wet stack conditions which necessitate specific consideration as to the stack design (low velocity stack) and the selection of corrosion resistant duct lining materials. A non-reheat alternative could be considered for the proposed design as additional experience is gained on FGD installations based on this design. However, for the purpose of this conceptual design, flue gas reheat is being proposed with the understanding that it may be modified in the final design.

Flue gas can be reheated in several ways. Reheat methods currently in use include:

1. Direct in-line reheat - using steam or hot water heat exchangers.

2. Direct combustion reheat - using gas or oil in either in-line burners or external combustion chambers.

3. Indirect hot air reheat - using steam to heat air to mix with the wet gas.

4. Bypass reheat - bypassing a portion of the untreated hot flue gas to mix with the treated gas. (This method is not applicable on high sulfur coals when 90 percent removal is required because bypass is not feasible).

The indirect hot air reheat method has been selected for this design, despite its recognized higher energy requirements as compared to the other reheat methods, because of its demonstrated reliability. In this method, the heat transfer surfaces are not exposed to the saturated flue gas stream which has been a major source of operating problems (tube corrosion and solids build-up) on installations using direct in-line reheaters. The major problems with direct combustion reheat have been attributed to failures due to vibration fatigue and flame instability. Another disadvantage is the need for auxiliary fuels which could be in short supply.

In the selected reheat method, ambient air is reheated to 300 F through condensing steam in a heat exchanger which is then injected into the flue gas in a mixing chamber raising the temperature to 170 F. (approximately 40 F above the flue gas dew point which is consistent with the level of reheat currently advocated by EPA and characteristic of reheat systems now in operation). Because the heat exchanger is outside the wet flue gas duct, corrosion and fouling problems are virtually eliminated. However, as a result of external reheat, the energy requirement is more than doubled as compared to the direct in-line reheat method, and the diameter of the stack is increased to accommodate the additional hot air flow. The hot air injection reheat system is shown schematically in Exhibit 5.

6. Stack Design

The reheated flue gases from each FGD train recombine and enter the stack through a common breeching. This conceptual design assumes a high-velocity stack (exit gas velocity of 90 Ft/sec) and one flue liner (11.4 Ft in diameter) designed for a gas flow of 548,000 ACFM @ 170 F and 1 in. W G. Additional entry is provided for the emergency bypass. The flue liner is of steel construction and coated with a suitable corrosion resistant material to provide protection against potential acid attack. In view of the recently reported failures of acid-resistant coatings, particularly at temperatures above 200 F, it is suggested that alternate materials, such as acid brick lining, be considered in the final stack design. Also, a quench system may be necessary to cool the flue gas during high temperature excursions (above 200 F). For the purpose of this conceptual design, a stack height of 300 Ft. was assumed.

7. Overall Material Balance

Major flows entering and leaving the AQC System are shown schematically in Exhibit 11 and further described in Exhibit 12*. Material balances were performed using the design parameters defined earlier in this Section and on the data furnished by Davy Powergas. Inlet flue gas flow rates were established on the assumption that the steam generators are operating at MCR conditions (no design margins applied).

* For sake of clarity, only the ESP alternative flows are shown.

The AQC System yields five major streams leaving the system: SO_2 product, sodium sulfate purge, prescrubber flyash/chloride purge, flyash collected in ESP/MDC and treated flue gas.

On the basis of 90 percent SO_2 removal in the Wellman-Lord Process, approximately 86 percent of the inlet sulfur is converted to the useful SO_2 product and about 4 percent is lost as a result of the sodium sulfate purge. In order to replenish this loss, 660 lbs/hr of Na_2CO_3 must be added into the system as make-up.

Another important make-up stream is water which must be added to replenish evaporative loss in the prescrubber, the prescrubber flyash/chloride purge, and the water leaving the system with the SO_2 product stream. The following fresh water makeup is indicated:

	Fresh Water Make-up, gpm
Evaporative Loss	142
Prescrubber Purge	63**
SO_2 Product	1
Make-up	206

Additional water may be required to provide flush water for pump seals in the Wellman-Lord Process. On the other hand, the total fresh water make-up may be reduced in the final design, if the water in the prescrubber purge is recycled from the waste disposal pond back to the system. The feasibility of the recycle will be contingent on the overall water management (now being evaluated by Ebasco) of the waste disposal pond that would yield an acceptable water quality in the recycle stream with respect to dissolved solids, particularly the chlorides.

** Prescrubber purge would be increased to 286 gpm for the MDC/prescrubber alternative due to higher amounts of fly ash removed.

8. Energy Requirements

The following energy requirements have been considered:

- Energy needed to drive the ID fans and the booster fans.
- Energy associated with the operation of the particulate removal equipment, and the Wellman-Lord Process.
- Energy required to reheat the treated flue gas.

The ID fans draft the boilers, air heaters and the particulate removal equipment. For the ESP alternative which has an overall pressure drop of 14 in. WG, the electric power requirements for the ID fans are estimated to be 880 kW. In the MDC design, the fan power requirements are increased to 1350 kW because the overall system pressure drop is higher (17 in. WG) and the fan efficiency is lower due to greater particulate loading in the gas.

The booster fans deliver the flue gases through the FGD System into the stack. The required pressure drop (based on information furnished by Davy Powergas) is 27 in. WG regardless what equipment is used for primary particulate removal because the pressure drop across the prescrubber is controlled by chloride removal. However, since the booster fans following the MDC's have a lower efficiency as compared to those following the ESP (due to higher particulate loading in the flue gas), the energy required to drive them is somewhat higher. The following power requirements are indicated: 1650 kW (ESP used) and 1770 kW (MDC used).

Total operating power consumption attributable to ESP is estimated at 800 kW. This includes transformer/rectifier sets, hopper heaters, rappers and dampers. No electrical energy is consumed by the MDC.

The total power requirements for the Wellman-Lord Process (excluding the gas reheat) have been estimated at 1530 kW. Breakdown for individual equipment is not available but it is assumed that the evaporator recirculating pumps are the major users of electrical power in the Wellman-Lord Process.

Steam represents the major energy requiremet in the Wellman-Lord Process. By far the largest users of steam are the double-effect evaporators; other smaller users are the SO_2 strippers, purge crystallizers and dryers. All but the dryers require low pressure (25 psig max.) steam. Since the lowest pressure steam available from the gasification process is at 50 psig, it has been proposed by Davy Powergas to use superheated steam from the available 230 psig header. This steam would be used to drive the booster fans with the turbine discharging steam at 25 psig. The energy remaining in the turbine exhaust steam would then be utilized in the evaporators, SO_2 strippers and purge crystallizers.

For the purpose of this study, it is assumed that the turbine will be of such design that sufficient steam will be exhausted to meet the Wellman-Lord Process low-pressure steam requirements which Davy Powergas estimate to be 81,000 lbs/hr. Assuming that steam at 230 psig and 540 F is used to drive the booster ID fans and saturated steam at 25 psig is exhausted by the fan turbines, the following energy requirements are indicated:

	Energy Requirement MM Btu/Hr
To drive the booster ID Fans	9.4
Wellman-Lord Process	81.2
Total	90.6

For stack gas reheat, approximately 36 MM Btu/Hr (39,500 lb/hr steam at 230 psig, 540 F) will be needed to raise the stack gas exit temperature to 170 F. An additional 140 kW will be required to drive the gas reheat air fans.

Operating energy requirements for the two proposed AQC system alternatives are summarized as follows:

		ESP/FGD	MDC/FGD
ESP	kW	800	---
ID Fans	kW	880	1350
Wellman-Lord	kW	1530	1530
Gas Reheat Air Fans	kW	140	140
Total	kW	3350	3020
Steam @230 psig, 540 F			
Booster ID Fans	MM Btu/Hr	9.4	9.4
Wellman-Lord	MM Btu/Hr	81.2	81.2
Gas Reheat	MM Btu/Hr	36.0	36.0
Total	MM Btu/Hr	126.6	126.6
	Lbs/Hr	121,300*	121,300*

It is imperative that a more comprehensive energy optimization study be performed when a more detailed process proposal is received from Davy Powergas.

9. Claus Unit Interface

Assuming that the SO_2 product from the Wellman-Lord Process is fed to the Claus Unit, the operation of the FGD system must be integrated with that of the gasification process. Of primary concern is the potential mismatch in the operating time of the steam generator and the Claus Unit. For example, during cold start-up at least one of the steam generators will have to be on line to provide steam for driving the process compressor turbines while the Claus Unit will not be started as yet. Obviously, the chemical recovery portion of the Wellman-Lord system cannot operate but the absorption system must treat the flue gases. Consequently, the Absorber Product and

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* It is estimated that approximately 118,350 Lbs/Hr of condensate (210 F) will be available for return back into the thermal cycle. The condensate used in dissolving the sodium carbonate makeup is not included in this amount.

Absorber Feed Tanks, located upstream and downstream of the chemical recovery plant, need to be of sufficient capacity to sustain the operating requirements of the absorption system. In the conceptual design both tanks are sized to provide a surge capacity of 461,000 gallons (equivalent to design flow rate for 48 hours, one boiler operating).

At design condition, the SO_2 product mass flow rate is 9,210 lb/hr (100% SO_2). In order to maintain the proper H_2S to SO_2 feed stoichiometry, some modification in the Claus Unit design may be required to accommodate the increased SO_2 input.

As previously indicated, the Wellman-Lord FGD System is designed to treat Claus Unit Tail Gases (CUTG). For the purpose of this study it has been assumed that CUTG's are introduced at the AQC System battery limits after they have been incinerated and cooled from the design temperature of 1400 F to 300 F in a waste heat boiler. This simplifying assumption may prove unworkable because at 300 F the CUTG may be below the acid dewpoint and thus provide a corrosive atmosphere in the ductwork. Humphreys & Glasgow have been requested to determine the CUTG dewpoint and the allowable temperature may have to be modified based on their findings. Higher CUTG temperature would cause a slight increase in the inlet temperature to the FGD system and a corresponding increase in the evaporative loss in the prescrubber.

Some consideration has been given to utilizing the thermal energy in CUTG for reheating the flue gases exiting the Wellman-Lord absorber. Direct gas to gas heat exchanger is deemed impractical because a very large heat exchange surface area would be needed. Also potential corrosion problems may arise. The feasibility of using the steam generated in the waste heat boiler as a supplementary source for the Wellman-Lord Process steam requirements warrants further consideration and this alternative should be evaluated in the framework of future heat optimization studies.

Another alternative that has been considered but not evaluated fully entails incineration of the CUTG directly in the steam generators. Preliminary discussions with one steam generator supplier indicate that direct incineration may yield NO_x emissions in the flue gases above acceptable levels. However, with some modification in the furnace design, this apparent problem should be overcome and it is suggested that direct incineration be evaluated in more detail because it offers simplicity and possible economic advantage over the external incineration method.

10. SO_2 Product Conversion

The SO_2 product recovery is 9210 lb/hr (100% basis) at design conditions. The product is suitable for conversion to either sulfur or sulfuric acid. Relatively minor processing steps are needed for conversion to either by-product. However, as previously indicated, modification in the design of the Claus Unit may be required in the event conversion to sulfur is contemplated.

Any steam produced in the SO_2 conversion facilities has not been considered in determining the energy requirements for the AQC System.

11. Scale-down to Demonstration Unit Size

Essentially the configuration of the Commercial Unit AQC System is adaptable to the Demonstration Unit size. Scale-down of the Wellman-Lord subsystems should be straightforward since the equipment used is not unusual and the scaledown factors are well defined. Since the total amount of SO_2 removed will be less, single-effect rather than double-effect evaporators may be the economic choice in terms of trade-off between capital investment and steam cost.

While the lower gas flow would indicate a single module absorption system, turn-down and reliability considerations would favor maintaining a two-module arrangement.

F. Risk Factors

Each subsystem of the proposed AQC System has been applied on commercial scale units. However, there have been and still are problems associated with AQC Systems. With increased experience, significant progress has been made in identifying the problem areas and in reducing the severity of their impact on the system reliability. The reliability of a specific AQC System will depend to a large extent on the soundness of its overall design, degree of redundancy, selection of materials of construction and, most importantly, on how well the system is maintained. The Owner must be prepared to devote substantial effort, continuously and on a skilled level, to the operation and maintenance of the system.

With regard to particulate removal, electrostatic precipitators are the most commonly used equipment. A recent survey conducted by Ebasco, covering 250 cold side ESP's, indicates 98 + percent (weighted average) availability* for a period of 10 years. Common causes of failures were due to flyash reentrainment, breakage of discharge electrodes, thermal expansion and problems associated with flyash handling. Such failures can be minimized by proper design and maintenance.

Failure to meet performance can be likewise minimized by proper selection of key design parameters, including SCA, aspect ratio, rapping intensity and electrical sectionalization. Many of the reported ESP failures with regard to performance have been due to undersizing, and to the problems related to handling high-resistivity flyash in equipment not designed specifically for such operating conditions.

MDC's are inherently simple to operate and maintain. However, as with any other equipment handling erosive materials, periodic replacement of certain components can be expected. A major risk factor associated with mechanical dust collector arrangement is the anticipated reduced reliability of ID and booster fans which, necessarily, handle flue gases with relatively

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* Availability is defined as the ratio of hours the ESP is available for operation (whether operated or not) to hours in the period.

high ash loading. While proper fan design can minimize outage time, providing replacement liners which protect both blades and housing against ash erosion and abrasion is deemed essential. In contrast, fans following the ESP are more reliable and require lower maintenance. Likewise the reliability of the prescrubber may be adversely affected as a result of higher ash loading. As previously indicated, wet particulate removal implies potential for solid deposit formation, which is aggravated at increased ash levels, creating maintenance problems and unscheduled shutdowns.

In terms of reliability, several Wellman-Lord systems installed in the U.S. and in Japan are noteworthy for their successful operating histories by having demonstrated on-stream factors of 97 to 98 percent.

Scaling and plugging due to slurry scrubbing has been the major source of maintenance and shut-downs in lime/limestone FGD System. Since Wellman-Lord System is based on clear liquor scrubbing, downtime due to scaling or plugging of absorbers has not been reported on Wellman-Lord units.

However, as with any complex chemical plant, the Wellman-Lord System must be properly maintained to insure sustained reliability. The areas that may have an adverse impact on reliability are the prescrubbers and the evaporators. The very low pH conditions in the prescrubber necessitate the use of high alloy materials to minimize corrosion. Thermal deposits of sulfites and sulfates on the evaporator heat exchanger surfaces require periodic (approximately every 6 months) shutdowns so that the surfaces can be washed.

A spare absorption module is not included in the conceptual design. As the absorption system has been sized to include a 15 percent margin on flow, the two installed modules are capable of handling 115 percent of MCR (maximum continuous rating of the boilers) gas flow. In the event one of the modules is taken out of service for maintenance, the remaining module could still treat 57.5 percent of total gas flow and sustain the operation of one of the boilers. Further "overloading" of the absorber may be feasible at a somewhat reduced SO_2 removal efficiency and increase in pressure drop. In addition, the ductwork arrangement permits any or all of the flue gas to by-pass the FGD System and flow directly to the stack. (If EPA regulations allow a variance to maintain operations in the event of an absorber loss.)

It is estimated that the addition of a spare module would increase the cost of the FGD System by 35-40 percent. On the basis of the performance of the Wellman-Lord units now operating and the margins built into the design, it is felt that the additional investment for a spare module is not warranted. It should be noted that at NIPSO no spare is provided; on the other hand at the San Juan Station of New Mexico Public Service, each unit will have one spare module.

G. EFFECTIVE INTERFACES

Major interfaces of the AQC System with the balance of the plant (Claus Unit, waste disposal, water and steam utilization) have been discussed earlier in this section in light of the various design parameters considered. As the overall process design evolves, the conceptual AQC System design may have to be modified to accommodate any changes in the interface areas.

Areas that need to be re-examined or considered in more detail are as follows:

- The feasibility of closing the loop with respect to liquid effluents
- Steam and condensate usage optimization
- Raw material handling and storage
- Integration of the AQC System with the SO_2 product conversion facilities
- Modification of AQC System to accommodate any changes in coal source. (The current design is based on a coal analysis developed for Texaco's design conditions. Further examination of expected maxima with regard to sulfur, ash and other coal constituents is mandatory as soon as this information becomes available.)

IV. ECONOMIC EVALUATION

A. General

The economic evaluation factors used in the study are tabulated below:

Average Annual Capacity Factor	%	90
Depreciation Charge Rate	%	6.67
Electrical Energy Charge	$/kwhr	0.0185
Steam (230 psig, 540 F)	$/ton	4.73
Sodium Carbonate Delivered Cost	$/ton	90
Cooling Water Cost	$/1000 gal	0.03
Maintenance Material & Labor		
Electrostatic Precipitator	% of Investment	1.0
Mechanical Dust Collector	% of Investment	2.0
Wellman-Lord FGD System	% of Investment	3.5
Operating Labor	$/man-year	25 000
Supervision	$/man-year	40 000

All costs are in 1978 dollars.

B. Investment Estimate

Comparative order-of-magnitude estimates have been made of the investment associated with the AQC Systems described in Section III.

The scope of each investment estimate includes both the equipment supplied by the Vendor and that which the Owner will have to provide (such as foundations, flyash handling equipment to battery limits, HV transformers, switchgears, motors above 250 HP and wiring). The scope excludes the waste disposal pond because its investment cost for the two particulate removal alternatives is estimated to be equivalent.

Equipment and erection costs for the Vendor supplied equipment are taken from budgetary proposals. The costs of the Owner supplied equipment have been developed by Ebasco based on available information and are subject to change upon receipt of more detailed design data. The conceptual estimate received from Davy Powergas for the Wellman-Lord system included both Vendor and Owner supplied equipment and therefore, no major adjustments by Ebasco were required.

On the gas side, the limits of the estimate may be identified as the air heater outlet to and including the stack.

All estimates are at 1978 pricing levels and include installed direct costs only.

C. Comparative Annual Operating Cost

The following costs are included in the annual operating cost analysis:

- Depreciation charges on direct plant construction cost
- Electrical Energy Charge
- Sodium Carbonate Make-up
- Steam Consumption
- Cooling Water Requirements
- Operating Labor & Supervision
- Maintenance Material & Labor

Not included are the costs attributable to process water consumption because the source of make-up water has not been established as yet. Its impact on the total operating cost will be insignificant as the consumption is relatively small.

All annual costs are based on an annual average capacity factor of 0.9 and on 1978 pricing level. The cost items which are affected by the coal sulfur content, namely steam and sodium carbonate consumption, are predicated on the assumption that design sulfur coal is burned.

D. Results

Economic evaluation was performed on the two design alternatives described in Section III.

> Electrostatic precipitator (ESP) at a 99.65 percent collection efficiency, followed by the Wellman-Lord FGD System. Particulate loading in the flue gas exiting the ESP is at 0.05 lbs/million Btu assumed to be the NSPS emission level.

> Mechanical dust collector (MDC) at a 60 percent collection efficiency, followed by the Wellman-Lord FGD System. Particulates not collected in the MDC are removed in the prescrubber to meet the NSPS emission level.

Detailed tabulation of comparative investment and owning and operating cost estimates is presented in Exhibits 13, 14, 15 and 16 and summarized as follows:

	ESP/Wellman-Lord	MDC/Wellman-Lord
	$1000 (1978)	
Total Direct Investment	25 240	22 055
Differential	+3 185	Base
Annual Operating Cost	5 945	5 674
Differential	+ 271	Base

Since the design and the costs associated with the Wellman-Lord FGD System are assumed to be the same for both alternatives, the cost differential between them is attributable solely to the respective particulate removal systems selected.

It is therefore noteworthy to examine the cost impact of reducing the collection efficiency of the electrostatic precipitator and effecting increased particulate removal in the prescrubber. As discussed in Section III, the controlling factor in establishing the pressure drop across the prescrubber appears to be the removal of chlorides from the flue gas. At the estimated pressure drop of 12 in. WG, required for chloride removal, acceptable particulate removal can be expected even if the ESP collection efficiency is reduced (higher particulate loading entering the prescrubber).

Three lower ESP collection efficiencies were considered: 99 percent, 98 and 90 percent. As would be expected, there is a significant reduction in the investment as follows:

ESP Efficiency %	SCA Ft^2/1000 ACFM	Estimated Direct Cost $1000
99.65	499	4 540
99.0	416	3 760
98.0	312	2 910
90.0	180	1 680

If the design were based on a 90 percent in lieu of a 99.65 percent efficient ESP, the annual operating cost would be reduced from $754 000 to $463 000 - primarily due to lower capital charges as shown in Exhibit 17. Sizing the ESP for 90 percent efficiency reduces the differential between the ESP and MDC designs to only $20 000 per year.

It must be noted that this analysis is predicated on the assumption that a reduction in ESP collection efficiency does not necessitate a higher pressure drop across the prescrubber. If this assumption were not made, cost reductions due to smaller ESP sizes would be at least partially offset by the higher energy requirements associated with the operation of the prescrubber.

E. Escalated Costs

It is the intent of this study to develop the comparative annual operating cost for the year 1981. It has been estimated that all costs except purchased power increase by 20.36 percent during the 1978 to 1981 period; in the same time period the cost of purchased power is expected to increase by 27 percent. It is apparent that escalation of the costs presented in Exhibit 16 will have a minimal impact on the relative economics of the alternatives, because the cost of electrical energy (purchased power) represents only about 8 percent of the total cost.

Comparative costs for the total AQC System expressed at 1981 cost levels are shown below:

	ESP/FGD	MDC/FGD
Annual Operating Cost, $1000	7 188	6 859
Differential	+329	Base

F. Baghouses

As stated in Section III baghouses are not considered a viable particulate control technology in the application under study because of lack of demonstrated performance of fabric filters in high sulfur coal service. Further justification can be made on economic grounds by comparing the operating costs of a baghouse with that of a cold-side ESP. This comparison, as presented in Exhibit 18, indicates that the annual operating cost of a cold-side ESP is estimated to be $265 000/yr less than that of a baghouse. The differential is due to a higher pressure drop (increased power to drive the ID fans), increased fuel consumption to maintain air heater exit temperature above the acid dewpoint and the bag replacement cost based on a two year bag life. These charges offset the lower capital investment associated with the baghouse system which is estimated to be $4 632 000 as compared to $5 340 000 for the cold-side ESP. Both the baghouse and the cold ESP are capable of achieving particulate emission level of 0.05 lb/million Btu.

EXHIBIT 1

DESIGN COAL CHARACTERISTICS *

(As Received Basis)

Ultimate Analyses		Weight Percent
Carbon		57.82
Hydrogen		3.65
Nitrogen		1.13
Sulfur **		3.40
Oxygen		4.80
Ash		17.20
Moisture		12.00
		100.00
Gross Heating Value	Btu/Lb	10 116

Flue Gas Composition	Weight Percent
CO_2	18.98
H_2O	5.19
SO_2	0.61
O_2	4.88
N_2	70.34
	100.00

* Coal characteristics based on design conditions developed for the coal gasification facilities (Commercial Unit).

** Includes 0.24% Chlorine.

EXHIBIT 2

REVISED NEW SOURCE PERFORMANCE STANDARDS UNDER CONSIDERATION FOR FOSSIL FUEL FIRED UTILITY BOILERS

(as of November 1977)

Sulfur Dioxide	
Emission Limitation	1.2 lbs/10^6 BTU's
Percent Reduction	90%(1)
Floor Value	0.2 lbs/10^6 BTU's
Particulates	
Emission Limitation	0.03 lbs/10^6 BTU's(2)
Percent Reduction	99%
Opacity Limitation	10%
Nitrogen Oxides	
Emission Limitation	
Subbituminous Coal	0.5 lbs/10^6 BTU's
Bituminous Coal and Certain Lignites	0.6 lbs/10^6 BTU's
North Dakota, South Dakota and Montana Lignites(3)	0.8 lbs/10^6 BTU's
Percent Reduction	65%

---

(1) 85 percent possible.

(2) 0.05 lbs/10^6 BTU's possible.

(3) Utilizing a cyclone boiler.

EXHIBIT 3

COMMERCIAL PLANT BOILER EMISSIONS(1)

Emissions	Sulfur Dioxide(2)	Particulates	Nitrogen Oxides(3)
Potential (uncontrolled)			
Pounds per Hour	10,724	18,987	788
Tons per Year(4)	42,274	74,846	3,106

---

(1) Emissions based on preliminary design data.

(2) Approximately 20 percent of these emissions are from the Claus Unit tail gases.

(3) Nitrogen oxides emissions assumes compliance with proposed standard of 0.6 lbs/MM Btu

(4) Based on 90 percent usage rate.

EXHIBIT 4

FLUE GAS DESULFURIZATION RECOVERY PROCESSES

CLASSIFICATION	TYPE	SUPPLIER
I Wet Processes		
1. Slurry	Magnesia (A)	Envirotech/Chemico United Engineers
2. Clear Liquor	Sodium Sulfite (Wellman-Lord) (A)	Davy Powergas
	Ammonia (B)	Catalytic
	Citrate (B)	Bureau of Mines Peabody Morrison Knudsen
	Phosphate (Aqua-Claus) (B)	Envirotech/Chemico
C	Steam Stripping	--
II Semi-Dry		
(Spray Dryer)	Aqueous Carbonate (B)	Atomics International
	Ammonia	Carborundum
III Dry	Carbon Sorption (A)	Foster Wheeler/ Bergbau
	Copper Oxide (A)	Shell/UOP
	Catalytic Oxidation (C)	Monsanto

(A) Process suitable for both sulfur and sulfuric acid production.
(B) Process suitable for sulfur production only.
(C) Process suitable for sulfuric acid production only.

SK-837 M-4

NO.	DATE	REVISION	BY	CH.	APPROVED	REVIEW

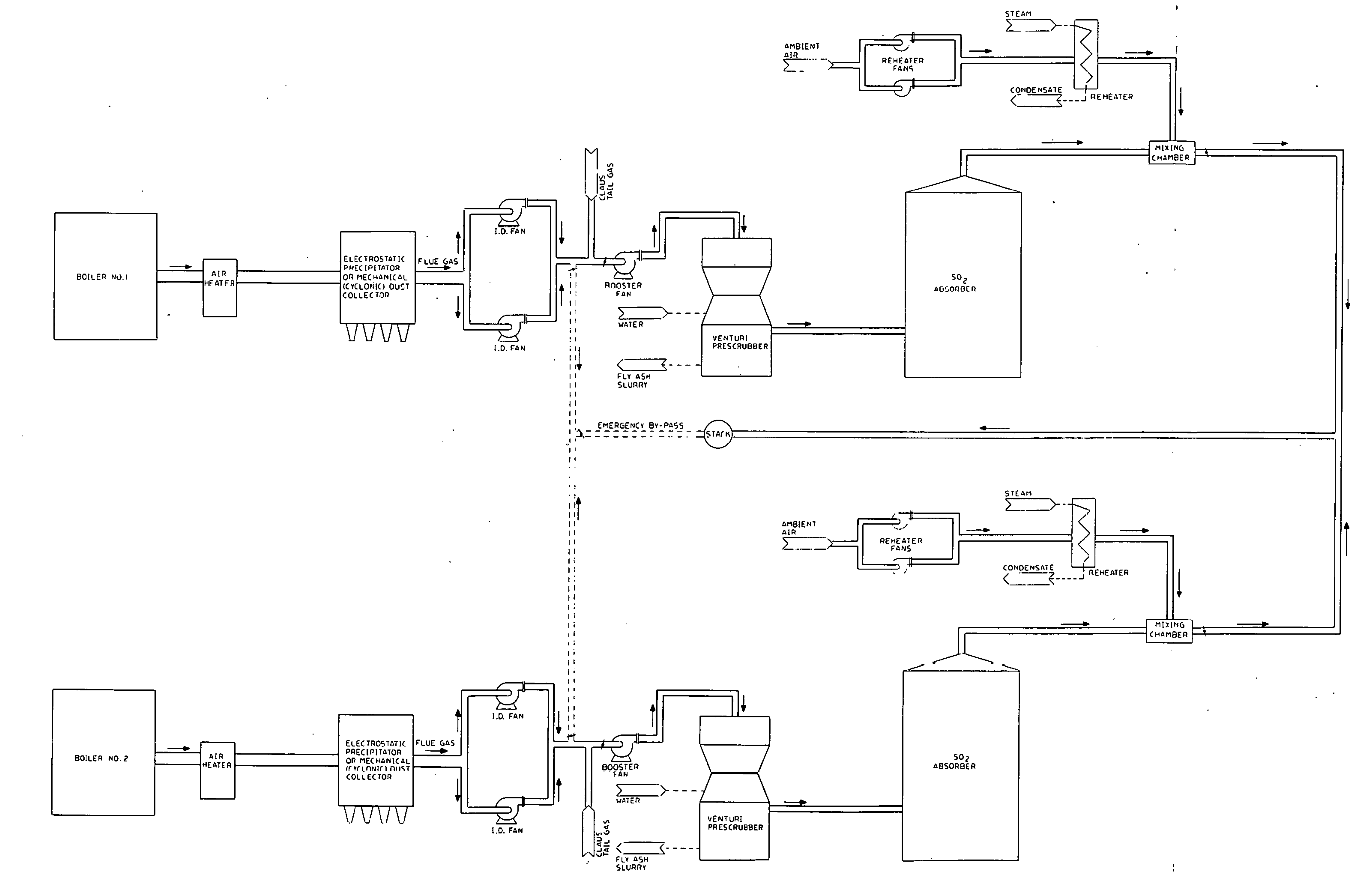


EXHIBIT 5

U.S. DEPARTMENT OF ENERGY
W. R. GRACE & CO.
MEMPHIS, TENNESSEE
SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
INDUSTRIAL PROJECT "B" - PHASE I
CONTRACT NO. ET-77-C-01-2577
COMMERCIAL PLANT
PROCESS SCHEMATIC
AIR QUALITY CONTROL SYSTEM-FLUE GAS CIRCUIT
EBASCO SERVICES INCORPORATED

SCALE NONE
APPROVED
DATE 6-21-78
DIV. MECH. DES.
DR. J. KARLICK
CH.
SECT. CODE NO.
SK-8376 M-4
ISSUE P₀

SK-8376 M-5

NO.	DATE	REVISION	BY	CH.	APPROVED	REVIEW

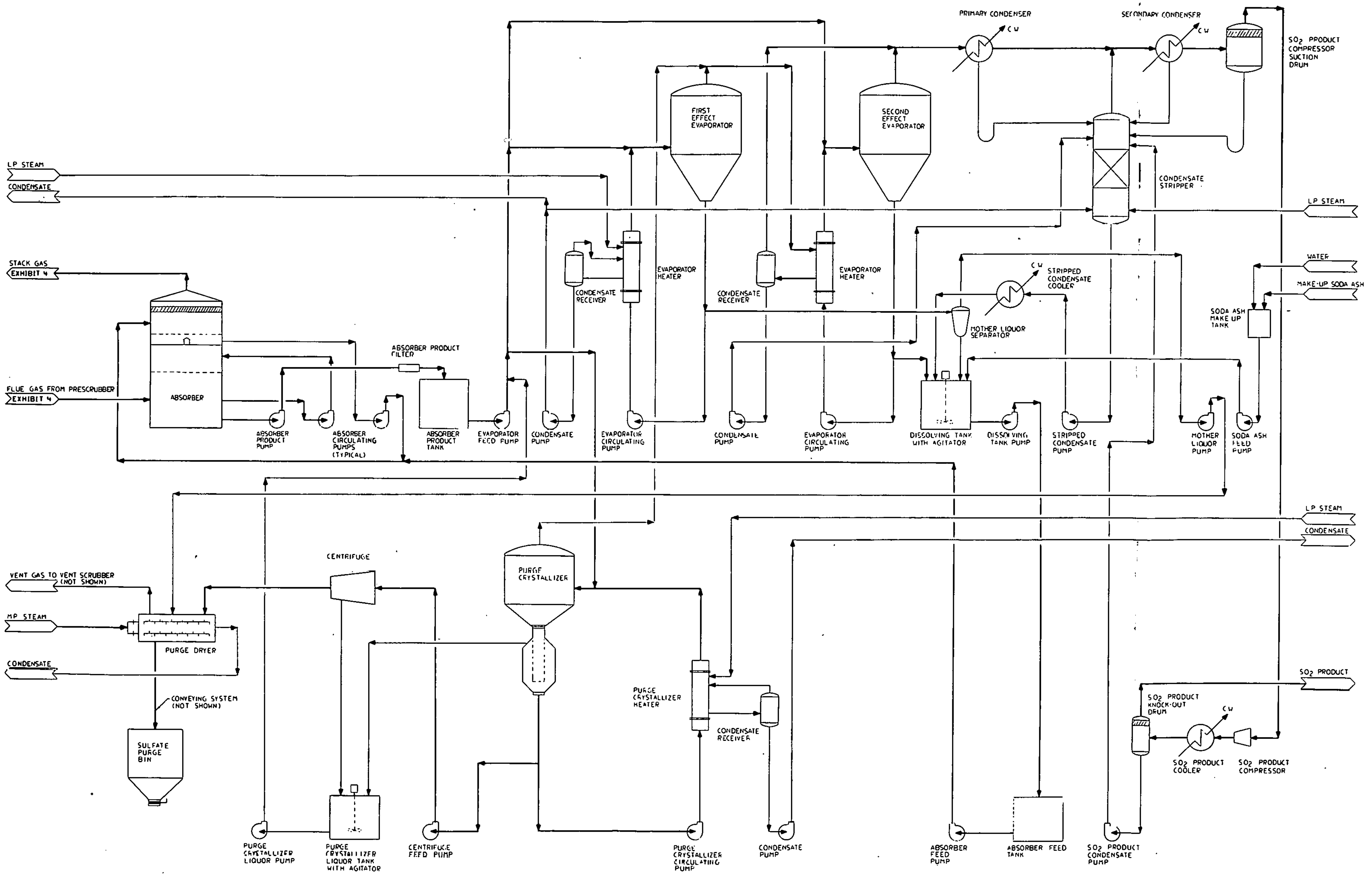


EXHIBIT 6

U.S. DEPARTMENT OF ENERGY
W. R. GRACE & CO.
MEMPHIS, TENNESSEE
SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
INDUSTRIAL PROJECT "B" - PHASE I
CONTRACT NO. ET-77-C-01-2577
COMMERCIAL PLANT
FLOW DIAGRAM
AIR QUALITY CONTROL SYSTEM
FLUE GAS DESULFURIZATION SYSTEM-WELLMAN LORD PROCESS
EBASCO SERVICES INCORPORATED
SCALE NONE
APPROVED
DATE 6-21-78
DIV. MECH DES
DR. OLIVETO
CH.
SECT. CODE NO.
SK-8376 M-5
ISSUE P0

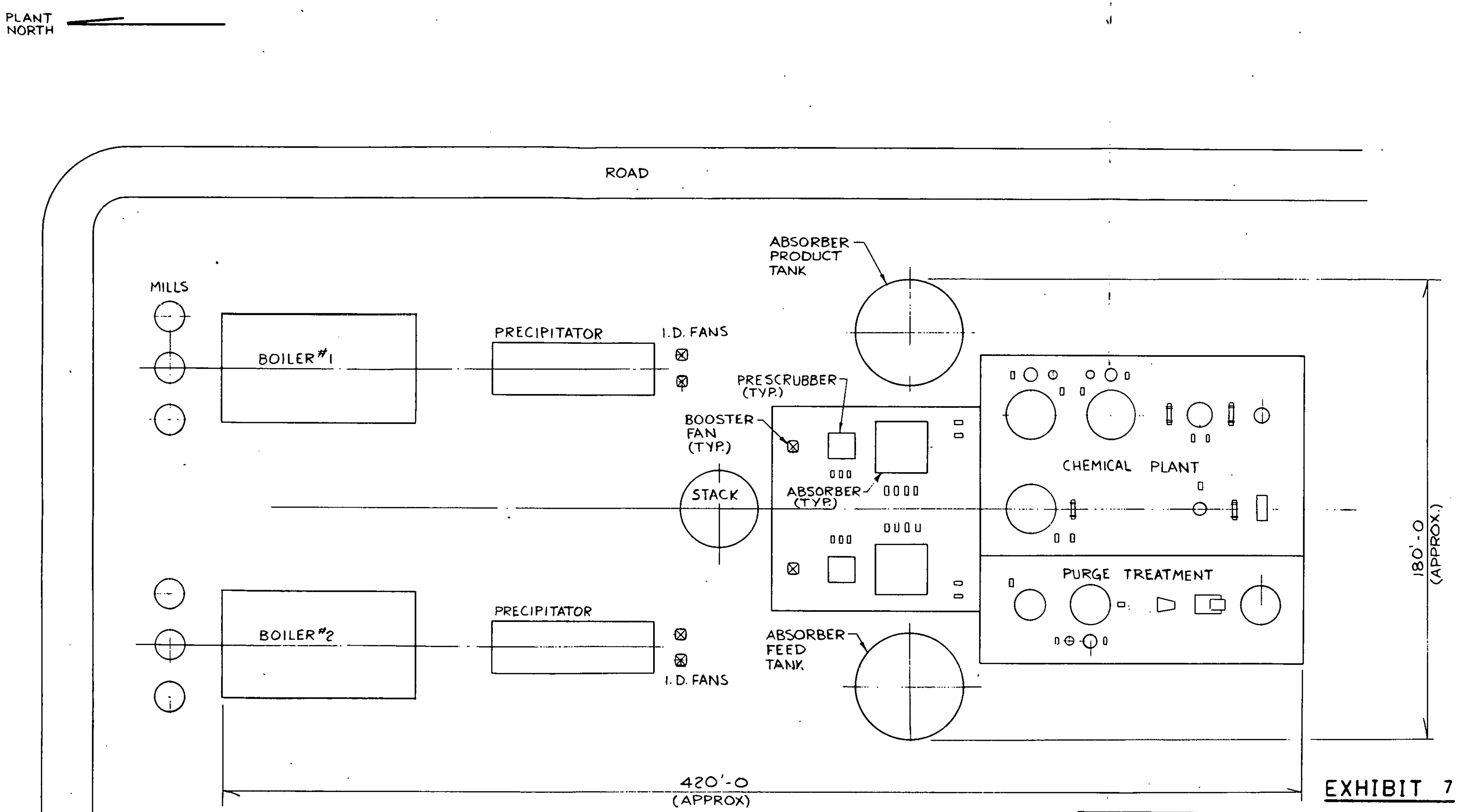


EXHIBIT 7

U.S. DEPARTMENT OF ENERGY

W. R. GRACE & CO.
MEMPHIS, TENNESSEE

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
INDUSTRIAL PROJECT "B" - PHASE I
CONTRACT NO. ET-77-C-01-2577

PLOT PLAN
AIR QUALITY CONTROL SYSTEM

EBASCO SERVICES INCORPORATED

5			
4			
3			
2			
1			
REV.	DATE	BY	APPROVED

EBASCO SERVICES INCORPORATED
DIV. ______ DR. ______
CH. ______
DATE ______
APPROVED

SCALE NONE
DIV. MECH. DES.
DR. OLIVETO
CH. ______
APPROVED
DATE 6-21-78
SECT. CODE NO.
SK-8376
M- 6
ISSUE P0

EXHIBIT 8

INCINERATED CLAUS TAIL GAS

DESIGN CONDITIONS

Composition	Weight Percent
H_2O	12.44
N_2 + Argon	37.16
CO_2	47.68
SO_2	2.72
	100.00

Design Mass Flow Rate,	Lbs/Hr	69,577

EXHIBIT 9
Sheet 1 of 3

PARTICULATE REMOVAL EQUIPMENT
TECHNICAL SUMMARY

A. ELECTROSTATIC PRECIPITATOR (ESP)

Quantity		Two (One per boiler)
Specific Collection Area (SCA)	Ft^2/1000 ACFM	499
Maximum Flue Gas Velocity	Ft/sec	4.1
Type Discharge Electrode		Weighted Wire
Aspect Ratio (Depth to Height)		1.5
Plate Spacing	inches	9
Rapper Cleaning Method		Impact Type
Pressure Drop Across ESP		
(including gas distribution)	in. W G	0.5
Overall (including ductwork)	in. W G	3.0
Installed Power	Kw	800

Per Each ESP:

Collecting Surface Area	Ft^2	133 882
No. of Electrical Fields		10
Field Depth	Ft	4.5
No. of Gas Passages		48
Plate Height	Ft	30
No. of Hoppers		6
No. of Transformer/Rectifiers		10
No. of Bus Sections		20
Overall Dimensions	Ft	
Height		62
Depth		52.5
Width		37.5

EXHIBIT 9
Sheet 2 of 3

A. ELECTROSTATIC PRECIPITATOR (ESP) (Cont'd)

ID Fans

Quantity		Four (Two per boiler)
Type		Radial
Blades		Airfoil
Design Conditions*		
ACFM/Fan		140,000
Static Pressure	in. W G	20
Efficiency	percent	87.5
BHP/Fan		510

B. MECHANICAL DUST COLLECTOR

Quantity		Two (One per boiler)
Type		Multicyclone

Per Collector

Number of Cyclones		30
Number of Banks		2
Overall Dimensions	Ft	
Height		32
Depth		100
Width		12
Pressure Drop Across Collector	in. W G	3
Overall (including ductwork)	in. W G	6

EXHIBIT 9
Sheet 3 of 3

B. MECHANICAL DUST COLLECTOR (Cont'd)

ID Fans

Quantity		Four (Two per boiler)
Type		Modified Radial
Blades		3/8" replaceable hardened steel liners
Design Conditions*		
ACFM/Fan		140,000
Static Pressure	in. W G	24.5
Efficiency	percent	69
BHP/Fan		780

* 20% margin on flow and 44% margin on Static Pressure

EXHIBIT 10

Sheet 1 of 3

WELLMAN-LORD PROCESS

EQUIPMENT LIST

EQUIPMENT	QUANTITY
Prescrubber Circulating Pump	4 + 2
Absorber Circulating Pump	6
Absorber Product Pump	2 + 2
Evaporator Feed Pump	1 + 1
Fly Ash Sump Pump	1 + 1
First Effect Condensate Pump	1
First Effect Evaporator Circulating Pump	1
Second Effect Condensate Pump	1
Second Effect Evaporator Circulating Pump	1
Mother Liquor Pump	1
Dissolving Tank Pump	1 + 1
Absorber Feed Pump	1 + 1
Stripped Condensate Pump	1 + 1
Seal Water Pump	1 + 1
Crystallizer Condensate Pump	1
Crystallizer Circulating Pump	1
Centrifuge Feed Pump	1
Crystallizer Liquor Pump	1
Chemical Plant Sump Pump	1 + 1
Condensate Pump	1 + 1
Soda Ash Feed Pump	1 + 1
Soda Ash Unloading Pump	1
Vent Gas Scrubber Circulating Pump	1 + 1
Fly Ash Sump Agitator	1
Dissolving Tank Agitator	1
Crystallizer Liquor Tank Agitator	1

NOTE: + 1 denotes 1 spare

EXHIBIT 10
Sheet 2 of 3

WELLMAN-LORD PROCESS
EQUIPMENT LIST (Cont'd)

EQUIPMENT	QUANTITY
Flue Gas Booster ID Fan including steam driven turbines	2
SO_2 Product Compressor	2 + 1
Fly Ash Filter	1
Centrifuge	1
First Effect Evaporator Heater	1
Second Effect Evaporator Heater	1
Primary Condenser	1
Secondary Condenser	1
Stripped Condensate Cooler	1
Sulfate Purge Dryer	1
Seal Water Cooler	1
Prescrubber	2
Absorber Inlet Gas Mist Eliminator	2
Absorber	2
Condensate Stripper	1
Vent Gas Scrubber	1
Sulfate Purge Bin Activator	1
Sulfate Purge Bin Slide Gate	1
First Effect Evaporator	1
Second Effect Evaporator	1
First Effect Condensate Receiver	1
Second Effect Condensate Receiver	1

EXHIBIT 10
Sheet 3 of 3

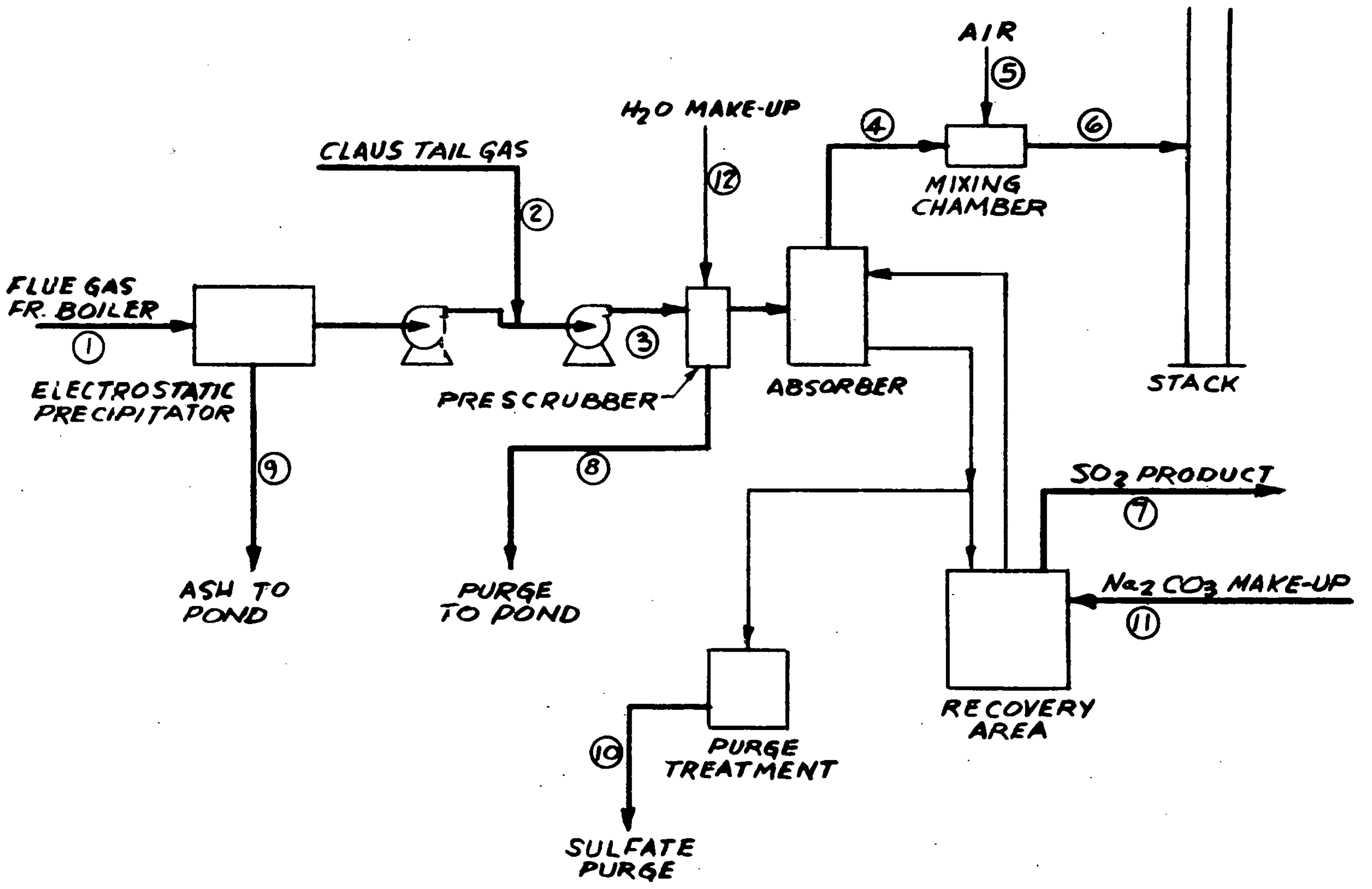
WELLMAN-LORD PROCESS
EQUIPMENT LIST (Cont'd)

EQUIPMENT	QUANTITY
Mother Liquor Separator	1
Purge Crystallizer	1
Purge Crystallizer Condensate Receiver	1
Steam Condensate Surge Drum	1
Absorber Product Tank	1
Dissolving Tank	1
Absorber Feed Tank	1
Evaporator Dump Tank	1
Evaporator Wash Water Tank	1
Crystallizer Liquor Tank	1
Sulfate Purge Bin	1
Soda Ash Storage Tank	1
Sulfate Purge Pneumatic Conveying System including one each of the following: Air Blower Air Filter Surge Hopper Surge Hopper Rotary Feeder Dust Collector Dust Collector Rotary Feeder	1
Gas Reheaters	2
Gas Reheat Fans	2 + 2

EXHIBIT 11

SYNTHESIS GAS DEMONSTRATION PLANT

SCHEMATIC FLOW DIAGRAM
AIR QUALITY CONTROL SYSTEM



EBASCO SERVICES INCORPORATED

DIV.____ DR.____
DATE____ CH.____
SCALE ____

APPROVED

EXHIBIT 12

OVERALL MATERIAL BALANCE
AIR QUALITY CONTROL SYSTEM
ELECTROSTATIC PRECIPITATOR AND WELLMAN-LORD SYSTEM
(2 STEAM GENERATORS)

Stream		(1) Flue Gas to AQCS	(2) Claus Tail Gas	(3) Flue Gas to FGDS	(4) Flue Gas FGD Outlet	(5) Reheat Air	(6) Flue Gas to Stack	(7) SO_2 Product
Dry Gas	lbs/hr	1 365 669	59 030	1 424 699	1 424 699	510 719	1 935 418	
SO_2	lbs/hr	8 830	1 894	10 724	1 072		1 072	9 210
H_2O	lbs/hr	75 182	8 653	83 835	155 026		155 026	490
Total	lbs/hr	1 449 681	69 577	1 519 258	1 580 797	510 719	2 091 516	9 700
Flyash	lbs/hr	18 987		66	66		66	
Chlorides	lbs/hr	312		312	Negligible			
Volume Flow	ACFM	466 540	21 621	441 600	390 600	161 790	553 960	
Temperature	F	300	300	300	128	300	170	120
Pressure	in. W G (psia)	-13	(15)	+28	+2	+2	+1	(15)

Liquid & Solid Flows		(8) Flyash/Chloride Purge	(9) Flyash to Pond	(10) Sulfate Purge	(11) Na_2CO_3 Make-up	(12) Water Make-up
H_2O	lbs/hr	31 442				103 023
Solids	lbs/hr	See Note	18 920		660	
Total	lbs/hr	See Note	18 920	840	660	103 023
Volume	gpm	63				206
Chlorides	lbs/hr	314				2

Notes: Flow at MCR (no margin) conditions
Design sulfur, flyash and chlorides
All required flyash removed in ESP. Under normal operating conditions some flyash will be removed in prescrubber and purged in Stream (8)
Stream numbers refer to flows in Exhibit 10

EXHIBIT 13

AIR QUALITY CONTROL SYSTEM
INVESTMENT ESTIMATES

$1000 (Present Day)

	Electrostatic Precipitator & FGD System	Mechanical Dust Collector & FGD System
A. Particulate Removal Equipment incl Ductwork		
1. Vendor Supplied		
Materials	1 700	641
Erection	1 500	214
Total Vendor Supplied	3 200	855
2. Owner Supplied Equipment & Erection	1 340	200
3. ID Fans Equipment & Erection	800	1 100
4. Sub-total Particulate Removal	5 340	2 155
Differential	+3 185	Base
B. Wellman-Lord FGD System Equipment & Erection	18 800	18 800
C. Stack	1 100	1 100
Total Direct Investment	25 240	22 055
Differential	+3 185	Base

EXHIBIT 14

PARTICULATE REMOVAL EQUIPMENT & ID FANS
ANNUAL OPERATING COST

1978 Cost Basis

			Electrostatic Precipitator (ESP)		Mechanical Dust Collector (MDC)	
	Item	Unit Cost	Quantity	$1000	Quantity	$1000
1.	Depreciation	6.67% of Inv	$5 340 000	356	$2 155 000	144
2.	Electrical Energy					
	ID Fans	$0.0185/kwhr	6.9 MM kwhr/yr	128	10.6 MM kwhr/yr	196
	ESP	$0.0185/kwhr	6.3 MM kwhr/yr	117	-	-
3.	Operating Labor	$25 000/man/yr	4 Men	100	4 Men	100
4.	Maintenance Labor	1.0% of Inv	$5 340 000	53		
	& Materials	2.0% of Inv			$2 155 000	43
5.	Total Annual					
	Operating Cost			754		483
	Differential			+ 271		Base

Basis: Steam generator at 100% MCR Conditions
Capacity Factor at 0.9
Costs associated with prescrubber excluded (charged to FGDS)

EXHIBIT 15

WELLMAN-LORD PROCESS - FGD SYSTEM
ANNUAL OPERATING COST

1978 Cost Basis

Item	Unit Cost	Quantity	$1000
1. Depreciation	6.67% of Inv	$19 900 000(1)	1 327
2. Sodium Carbonate	$90/ton	2600 tons/yr	234
3. Electrical Energy	$0.0185/kwhr	13.2 MM kwhr/yr	244
4. Steam @230 psig, 540F	$4.73/ton	478 165 tons/yr	2 262
5. Cooling Water	$0.03/1000 gal	2885 MM gal/yr	87
6. Operating Labor	$25 000/man/yr	12 men	300
Operating Supervision	$40 000/man/yr	1 man	40
7. Maintenance Labor & Materials	3 1/2% of Inv	$19 900 000(1)	697
8. Total Annual Operating Cost			5 191

Basis: Steam Generators at 100% MCR Conditions
Capacity Factor @ 0.9

(1) Includes Stack

EXHIBIT 16

AIR QUALITY CONTROL SYSTEM
ANNUAL OPERATING COST
$1000 (1978)

Item	Electrostatic Precipitator & FGD System	Mechanical Dust Collector & FGD System
1. Depreciation	1 683	1 471
2. Electrical Energy	489	440
3. Raw Materials	234	234
4. Steam	2 262	2 262
5. Cooling Water	87	87
6. Operating Labor & Supervision	440	440
7. Maintenance Labor & Materials	750	740
8. Total Annual Operating Cost	5 945	5 674
Differential	+ 271	Base

EXHIBIT 17

PARTICULATE REMOVAL EQUIPMENT & ID FANS
COMPARATIVE ANNUAL OPERATING COST
AT VARIOUS COLLECTION EFFICIENCIES

$1000 (1978)

		ESP Efficiency, %				Mechanical Dust Collector
		99.65	99	98	90	
1.	Depreciation	356	304	247	165	144
2.	Electrical Energy					
	ID Fans	128	128	128	128	196
	ESP	117	98	73	45	-
3.	Operating Labor	100	100	100	100	100
4.	Maintenance	53	50	40	25	43
5.	Total Operating Cost	754	680	588	463	483
*	Differential	+291	+217	+125	Base	+20

EXHIBIT 18

COMPARATIVE ANNUAL OPERATING COST
BAG HOUSE AND COLD-SIDE ESP
(1978 COST BASIS)

	Bag House			Cold-Side ESP
	Unit Cost	Quantity	$1000	$1000 (1)
1. Depreciation	6.67% of Inv	$4 632 000	309	356
2. Electrical Energy				
ID Fans (2)	0.0185/kwhr	11.8 MM kwhr/yr	218	128
Bag House or ESP	$0.0185/kwhr	1.0 MM kwhr/yr	19	117
3. Fuel Charge (3)	$0.8312/MMBtu	194 242 MMBtu/yr	161	-
4. Operating Labor	$25 000 /man/yr	4 men	100	100
5. Maintenance				
Labor & Material	1% of Inv	$4 632 000	46	53
Bag Replacement (4)	$60/bag	2 760 bags/yr	166	-
6. Total Annual				
Operating Cost			1 019	754
Differential			+ 265	Base

(1) Costs as presented in Exhibit 14

(2) Pressure drop across Bag House system estimated at 8 in. W G.

(3) Due to increase Air Heater Temperature to 375 F which is equivalent to 1.875 percent increase in fuel consumption.

(4) Two year bag life.

APPENDIX A

LIST OF ABBREVIATIONS

ACFM	Actual Cubic Foot per Minute
AQC	Air Quality Control
BHP	Brake Horsepower
BACT	Best Available Control Technology
Btu	British thermal unit
CAAA	Clean Air Act Amendments of 1977
CUTG	Claus Unit Tail Gas
EPA	Environmental Protection Agency
ESP	Electrostatic Precipitator
F	Degree Fahrenheit
FGD	FLue Gas Desulfurization
GEP	Good Engineering Practice
gpm	gallons per minute
gns/SCF	grains per Standard Cubic Foot
ID Fan	Induced Draft Fan
in. W G	Inches Water Gauge
kwhr/yr	Kilowatt hours per year
L/G	Liquid to Gas Ratio
LAER	Lowest Achievable Emmission RAte
lbs/hr	pound per hour
MCR	Maximum Continuous Rating
MDC	Multicyclonic Mechanical Dust Collector
MM	Million
MW	Megawatts
NIPSCO	Northern Indiana Public Service Company
NAAQS	National Ambient Air Quality Standard
NSPS	New Source Performance Standards
PSD	Prevention of Significant Deterioration
ppm	parts per million
SCA	Specific Collection Area
SCFM	Standard Cubic Foot per Minute
TPD	Short Tons per Day

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY V

COOLING TOWER OPTIMIZATION

EBASCO SERVICES INCORPORATED
Two Rector Street
New York, NY 10006

November 1978

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY V

COOLING TOWER SYSTEMS

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I - INTRODUCTION

A - OBJECTIVE

The objective of this study is to determine the optimum design for a cooling tower system for use in the Synthesis Gas Demonstration Plant Program. The system selected is to be suitable for both the Demonstration and Commercial Plant.

B - SCOPE

Data developed herein are based on the Commercial Plant. The applicability to the Demonstration Plant is discussed.

The cooling tower system is to operate in conjunction with the synthesis gas plant, receiving hot water discharge from a variety of heat rejection equipment including direct equipment coolers, heat exchangers and mechanical drive turbine condensers. The hot water mixture will be cooled and returned to the plant at a temperature consistent with the economics of the cooling system and its effect on plant performance. The report identifies the optimum cooled water temperature developed from design data currently available.

The investigation will be confined to towers of the evaporative type including natural and mechanical draft designs. There appears, at this time, no justification for the substantial added investment required for special types designed specifically for water conservation or plume abatement.

The study utilizes the Ebasco Computer Program for the approximate sizing and pricing of cooling towers. A series of cold water temperatures and cooling ranges are used as input to the program which in turn calculates cooling tower design parameters, performance, and price for various combinations. Data thus generated can be compared with data solicited from suppliers for validity.

The Ebasco computer program for the economic selection of steam condensing systems was also used with a simulated condenser to establish a base for estimating system costs other than the cooling tower. The study plan enumerates items of investment and operating costs which can be examined in arriving at finite cost estimates. This study is intended only to identify cooling systems in order of ranking. The following listed items do not have an impact on this result since they are either constant adders or multipliers. They therefore have been omitted.

Investment Costs

Cost of Land
Escalation
Interest During Construction

Operating Costs

Makeup Water Supply and Treatment

The study examines cooled water temperatures from 83°F to 94°F and cooling ranges from 17°F to 30°F. Supplementary data explores reducing the design inlet temperature to 81°F and increasing the cooling range to 35°F.

II - CONCLUSIONS AND RECOMMENDATIONS

The basic results of the economic investigation of rectangular mechanical draft cooling towers for the commercial plant application are tabulated in Exhibit I (Sheets 1 to 3) and plotted in Exhibit II. The results show that the annual costs, adjusted for the influence of cooled water temperature and cooling range on plant equipment costs and performance, follow a well defined pattern. Specifically, we find that for a given design cooled water temperature, costs are reduced when the cooling range (or temperature rise through the plant) is increased. The rate of cost reduction decreases however, indicating an eventual reversal of the trend caused by a rapid increase in the cost of plant equipment designed for high range and high temperatures.

It is also noted that costs decrease as the design cooled water temperature approaches the design wet bulb temperature. This trend is due to the influence of plant costs since the trend in cooling tower costs is opposite. An optimum occurs at 83°F cooled water temperature due to rapidly increasing cooling tower costs.

The plotted results show that for all cooling ranges up to 35°F, a cooled water temperature of 83°F is the economic choice. This is also the lowest practical choice for cooling tower design. Suppliers have been known to guarantee a 5°F approach (81°F) but specifying this as a performance requirement is certain to restrict participation in the bidding.

The investigation has not been carried beyond a cooling range of 35°F although shape of the curves indicates that the trend reversal and most economical point appears to occur at a higher range.

In combination with an 83°F cooled water temperature, a 35°F range will produce a hot water temperature of 118°F entering the tower. This approaches the 120°F limit which cooling tower suppliers are reluctant to exceed because of adverse effects on materials. To allow for the specfication of plant equipment to accept a cooled water temperature in excess of 83°F, and to allow for some deterioration in tower performance without exceeding 120°F hot water temperature, it is recommended that a range of 30°F be used, considering the nominal increase in cost associated with this change. Furthermore, in the interest of conservatism, to mitigate any effects of biological fouling or mechanical deterioration, we recommend that equipment using cooling water be designed for a cooled water temperature higher than 83°F.

Summarizing, it is recommended that the cooling tower design be based upon a cooled water temperature of 83°F and a cooling range not to exceed 30°F. It is further recommended that the cooling tower be of the conventional rectangular mechanical design.

A detailed tabulation has not been prepared for the Demonstration Plant. The current state of design indicates that an extrapolation of data presented in this report will be valid and that the relationships established will hold. However, further investigation of the original criteria for selection of the maximum wet bulb temperature causes us to recommend that the Demonstration Plant criteria be based on that temperature which will be exceeded for 5 percent of the summer months. This results in a design wet bulb temperature of 77°F. An approach of 7°F (84°F cooled water temperature) and a range of 30°F, used as a basis for the Demonstration Plant, results in a 114°F hot water temperature and no change in basic cooling tower design. System equipment for the Demonstration Plant is recommended to be designed for a cooled water temperature higher than 84°F consistent with the explanation above.

III - TECHNICAL APPROACH

A - DESIGN BASIS

This section presents a discussion of the basic design criteria and operating parameters which must be satisfied by the cooling system under consideration. The study is based on the following design and economic factors developed for the Commercial Plant.

Design

1.	Heat Load - 10^6 Btu/hr	2720
2.	Operating Hours/year	7920
3.	Circulating Water Pumps	
	Number	2
	Type	Vertical
	Element	Removable
	Material	Steel
4.	Circulating Water Conduit	
	Length - ft	1200
	Base Friction Loss - ft (for conduit sizing)	25
5.	Design Water Level (below intake deck) - ft	3
6.	Design Ambient Temperatures	
	Wet Bulb	76°F
	Dry Bulb	85°F

Economics

1.	Unit Fuel Cost - $\$/10^6$Btu	0.8312
2.	Depreciation - %	6.67
3.	Incremental Capability Charge $/kW	334
4.	Cost of Incremental Steam	
	1500 psig 940°F $/ST	7.66
	580 psig 730°F $/ST	6.11
	230 psig 540°F $/ST	4.73

B - PROCESS CONSIDERATION - PLANT OPERATION

Operation of the plant is considered to be at full capacity, 330 days per year for the purpose of this study. No part load operation has been considered. Accordingly, when operating the cooling water system is assumed to be at full capacity regardless of the season of the year. The cooling system is to be designed for optimum cooled water temperature when ambient conditions are at design levels.

For this study, condensing turbines have been assumed to be designed for 4.5" Hg back pressure at the design cooled water temperature and temperature rise specified. This provides for operation in the 3"-3.5" Hg range under average seasonal conditions. Improvement in operating performance based on a weighted annual cooling water inlet temperature has been included in the Total Adjusted Annual Cost.

C - COMMERCIAL EVALUATION

1.1 SYSTEM

The cooling water system associated with each cooling tower selection consists of an intake structure, circulating water pumps and reinforced concrete conduit to conduct water to the plant and return.

Pumps are standard-vertical mixed flow, removable impeller type for fresh water service. Motors are of weatherproof construction, as it is assumed they will be out-of-doors. For a given flow, the same basic pump selection will be suitable for any tower selection, the variation in pumping head being small as compared to the total required head.

Steel piping connections are used to connect pumps and equipment to the concrete conduit and large valves are of the butterfly type.

The system as considered in this study makes no provision for extra or standby capability. Addition of spare pumping capacity would not affect the results of this study.

1.2 DESCRIPTION OF COOLING TOWERS

Evaporative cooling towers of two basic designs are considered, both of fire resistant construction. These include the natural draft type and two variations of the mechanical draft type. The design wet bulb temperature is derived from a review of meteorological data for the City of Evansville, Indiana over the period 1958-1964. A statistical analysis of these data indicates that a wet bulb temperature of 76°F will be exceeded, on an average 18 days(5 percent) annually. This occurs most frequently in connection with a dry bulb temperature of 85°F (approximately 67 percent relative humidity). By comparison a wet bulb temperature of 77°F is expected to be exceeded 2.5 percent of the time. Reduction in design point to 75°F would result in increasing expected excursions over design conditions to 12 percent which is considered unacceptable.

a) Natural Draft Towers

Two basic designs exist for natural draft cooling towers:

1. The crossflow type distributes the water peripherally around the base of the shell so that the air path is horizontal and perpendicular to the water falling through a fill section which is entirely external to the shell.

2. The counter flow type encloses the fill section within the shell so that air flow through the fill is upward, counter to the falling water. For the same performance, this type will have a higher shell and higher discharge elevation than the cross flow design.

The chimney effect of the high shell induces the flow of air through the fill area where approximately 80 percent of the cooling is due to evaporation of a portion of the water and the remainder is by transfer of sensible heat. Loss of water by windage is appreciably reduced by the long passage up to the discharge elevation.

Air flow through the natural draft tower is not controlled as it is in a mechanical draft design. Cross flow design natural draft towers subjected to low ambient temperatures have experienced severe icing damage and are generally unsatisfactory under these conditions. Counterflow towers, however, have a relatively good record under severe conditions. For this reason, counterflow design only is considered.

b) Mechanical Draft Cooling Towers

Two mechanical draft designs are considered. The first is the conventional rectangular type having the required number of cells arranged linearly and oriented to have the axis parallel to the prevailing winds. The Commercial Plant will require one tower of this design type, having ten (10) cells. For the Demonstration Plant, a single tower having four (4) slightly smaller cells will meet requirements. The number, size of cells and tower height for each application varies slightly among suppliers and may be cause for additional environmental evaluation in the course of final selection.

A round mechanical design is available for the Commercial Plant in either counterflow or crossflow designs, competitively priced with each other. There is a low duty limitation due principally to physical considerations and the Demonstration Plant requirements have been found to be too low for this design. Improved plume buoyancy, because of the clustered arrangement of the fans, is an attractive characteristic of this design. It also exhibits reduced recirculation as compared with the rectangular design.

Compared with natural draft designs, drift loss will be somewhat higher due to proximity of the discharge to the drift eliminators. Droplets which are not collected, or happen to be reentrained are almost immediately discharged to atmosphere.

D - RISK ANALYSIS

All components of the systems proposed are of time proven design. The study suggests that reliability is such that standby capability is not required for any of the equipment. Cost estimates do not therefore include redundant items.

E - EFFECTIVE INTERFACES

The study is based on information developed by Humphreys and Glasgow and Ebasco with regard to cooling requirements for the process plant and the variation of equipment costs with cooling water temperature level. These data have permitted identification of a system and the design operating parameters.

IV - ECONOMICS

A - CAPITAL COSTS (INVESTMENT)

Tabulated in Exhibits I, III, and IV are installed costs for the system considered. Five cooling ranges are reviewed. These ranges are 17°F, 20°F, 25°F and 30°F. For the rectangular mechanical design, 35°F is shown as supplementary information.

Meetings were held with various cooling tower suppliers to obtain up-to-date costs for some of the tower selections involved. Information developed through these contacts is used to update and validate cooling tower costs determined by computer. The Marley Cooling Tower Company, Zurn Industries and the Ecodyne Corporation were contacted for this purpose. All attended discussion meetings but information requested was received from Marley and Zurn only.

Costs for items other than the cooling tower and basin were estimated from available in-house data developed over a period of time from typical designs. Material and labor costs were applied as follows.

	Material	Labor
Circulating Water Conduit		
60" Dia. - $/ft	120	123
72" Dia.	159	153
96" Dia.	232	201
120" Dia	364	258
144" Dia.	594	317
Intake Structure $/cu. ft.	1.90	4.25
Grading Cooling Tower Area $/cu. yd.		3.00
Piling for CT Foundation $/sq. ft.	4.18	2.13
High Voltage Cable $/mva/ft.	8.30	11.00

	Material	Labor
Low Voltage Cable $/mva/ft.	8.00	13.00
Control Wiring - Circ Water Pump - $/ft.	18.10	35.00
Control Wiring - C T Fans - $/ft.	7.86	16.33
C.W Pumps per information supplied by Byron Jackson		
C.W Pump Motors - Westinghouse Price Book		

Investment costs have been adjusted to reflect differential cost variation of process plant equipment and condensing equipment with changes in cooled water temperature and cooling range. In general these show that process plant equipment increases in cost with rising cooled water temperature and also with increasing range. The cost of condensing equipment tends to increase with increasing range and for any given range increases with increasing cooled water temperature.

As a result of the review of investment costs it was concluded that the system costs for the natural draft type and the round design variation of the mechanical type were out of range for these applications. The round mechanical design is, considerably more costly ($300,000 annually at 30°F cooling range). The natural draft tower is slightly more attractive with its energy advantage but cannot be expected to achieve an approach as low as 7°F.

The conclusions are therefore based upon rectangular mechanical draft cooling towers.

B - OPERATING COSTS

Annual costs of owning and operating the system consist of the fixed charges on investment plus the cost of energy used to operate fans and pumps, the cost of makeup water and the cost of maintenance. Makeup costs (including treatment) have been omitted as they vary only slightly among the tower selections reviewed. This is because evaporation and blowdown are functions of heat load and solids concentration which are constant for all situations.

Steam driven mechanical equipment exhausting to condensers will show a variation in performance due to the condenser pressure which is determined by the cooled water temperature and the cooling range of the system. The adjustment for this variation reflects the cost of providing sufficient steam to maintain required power output based upon average seasonal operating temperatures.

Variation in cooled water temperature has negligible effect on process efficiency. This assumes equipment properly sized for the design temperature level to be experienced.

C - RISK ANALYSIS

A high level of confidence in the system and equipment renders analysis of financial risk unnecessary.

The study does not recognize possible unfavorable environmental impact of the system selected. The atmospheric effects are discussed in an appendix to this report entitled Atmospheric Effects of Alternate Cooling Tower Operation at the Syngas Demonstration and Commercial Plants. This appendix bases the acceptability of the mechanical draft on a location and orientation which will mitigate tower induced fogging on nearby Tscharner Road. It is expected that this can be accomplished.

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EXHIBIT 1
Sheet 1 of 3

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION

RECTANGULAR MECHANICAL TOWER

(For Commercial Plant)

COOLING RANGE - F	17			20			25				30				35	
COOLED WATER TEMP. - F	83	89	94	83	89	94	81	83	89	94	81	83	89	94	81	83
A. DESIGN CONDITIONS																
1. Heat Load - 10^6 Btu/hr	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720
2. Circ. Water - 10^3 GPM	320	320	320	272	272	272	217.6	217.6	217.6	217.6	181.3	181.3	181.3	181.3	155.4	155.4
B. TOWER DESCRIPTION																
1. No. of Towers	2	2	1	2	2	1	2	2	2	1	2	2	2	1	2	1
2. Basin Length (ft)	400	352	400	360	288	400	352	280	280	360	280	240	200	320	240	400
3. Basin Width (ft)	52	52	52	52	52	52	52	52	52	52	52	52	52	52	52	52
4. Height - Overall (ft)	57	57	57	57	57	57	57	57	57	57	57	57	57	57	57	57
5. Pumping Head (ft)	43	43	43	43	43	43	43	43	43	43	43	43	43	43	43	43
6. No. Fans/Tower	10	8	10	9	8	10	8	7	7	9	7	6	5	8	6	10
C. INVESTMENT ($1000)																
1. Site Preparation (Grading & Excavation)	42	31	22	35	25	21	39	31	24	20	33	27	20	18	23	20
2. Piling	313	231	165	265	189	157	289	234	184	151	245	201	151	134	168	144

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EXHIBIT 1
Sheet 2 of 3

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION

RECTANGULAR MECHANICAL TOWER

(For Commercial Plant)

COOLING RANGE - F	17			20			25				30				35	
COOLED WATER TEMP. - F	83	89	94	83	89	94	81	83	89	94	81	83	89	94	81	83
3. Cooling Tower Basin	180	133	95	153	109	91	167	135	106	87	142	116	87	77	97	83
4. Intake Structure	473	473	473	402	402	402	322	322	322	322	268	268	268	268	230	230
5. Circ. Water Conduit	280	280	270	260	260	254	248	248	248	243	207	207	207	204	177	177
6. Cooling Tower	3544	2409	1678	3428	2227	1802	3639	2971	2160	1700	3135	2569	1788	1521	2504	2035
7. Circ. Water Pumps & Motors	1360	1360	1360	1129	1129	1129	860	860	860	860	717	717	717	717	615	615
8. Switchgear & Wiring	1077	966	886	1072	963	898	861	767	698	653	781	668	601	575	668	486
9. Instruments & Controls	74	64	63	74	64	63	75	70	60	58	72	68	58	56	68	60
TOTAL DIRECT COST	7343	5947	5012	6818	5368	4817	6500	5789	4662	4245	5600	4841	3897	3573	4550	3850
ADJUSTMENTS ($1000)																
1. Process Equipment	-956	BASE	+722	-566	+254	+936	-332	-39	+829	+1482	+254	+546	+1404	+2067	+839	+1150
2. Condensing Equipment	BASE	+154	+338	+39	+213	+429	+44	+120	+340	+639	+132	+227	+530	+1021	+273	+368
TOTAL ADJUSTED INVESTMENT	6387	6101	6072	6291	5835	6182	6212	5870	5831	6366	5986	5614	5831	6661	5662	5368

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EXHIBIT 1
Sheet 3 of 3

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION

RECTANGULAR MECHANICAL TOWER

(For Commercial Plant)

COOLING RANGE - F	17			20			25				30				35	
COOLED WATER TEMP. - F	83	89	94	83	89	94	81	83	89	94	81	83	89	94	81	83
E. ANNUAL COSTS ($1000)																
1. Depreciation on Adjusted Investment	426	407	405	420	389	412	414	392	389	425	399	374	389	444	378	358
2. Cooling System Energy @ $0.0185/kWh	1601	1436	1317	1343	1256	1140	1274	1139	1036	970	1080	992	892	855	918	870
3. Drive Turbine Performance Adjustment (cost of steam)	BASE	+355	+537	+37	+390	+563	+54	+109	+463	+608	+132	+180	+537	+652	+219	+272
4. Estimated Maintenance	113	91	70	104	86	68	97	86	75	60	84	72	59	52	70	59
TOTAL ADJUSTED ANNUAL COST	2140	2289	2329	1904	2121	2183	1839	1726	1963	2063	1695	1618	1877	2003	1580	1559

EXHIBIT II

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY V
COOLING TOWER OPTIMIZATION

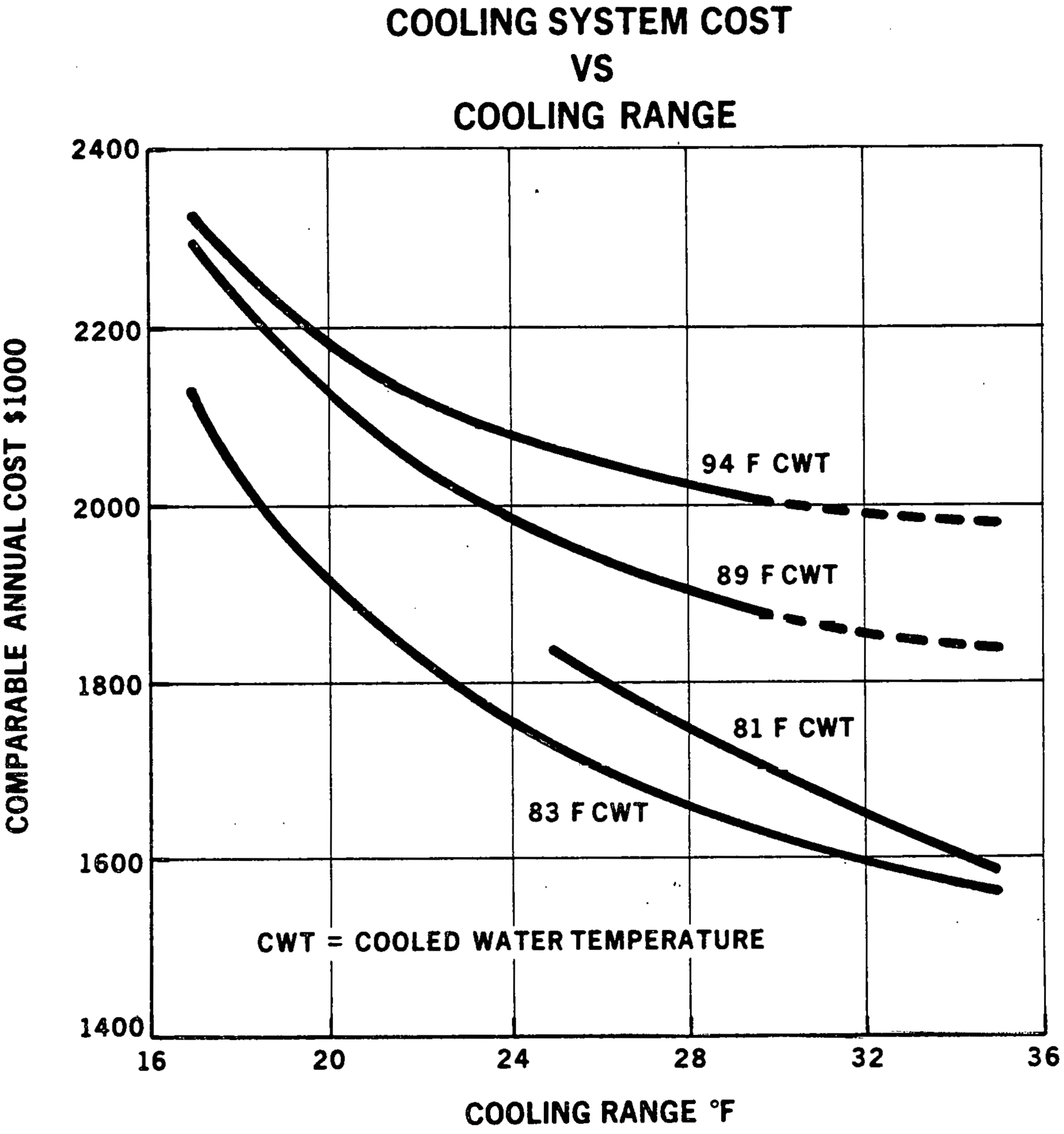

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EXHIBIT III
Sheet 1 of 3

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION

ROUND MECHANICAL TOWER

(For Commercial Plant)

COOLING RANGE - F	17			20			25			30		
COOLED WATER TEMP. - F	83	89	94	83	89	94	83	89	94	83	89	94
A. DESIGN CONDITIONS												
1. Heat Load - 10^6 Btu/hr	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720	2720
2. Circ. Water - 10^3 GPM	320	320	320	272	272	272	217.6	217.6	217.6	181.3	181.3	181.3
B. TOWER DESCRIPTION												
1. No. of Towers	2	1	1	2	1	1	2	1	1	1	1	1
2. Basin Diameter (ft)	275	355	315	260	335	290	250	315	275	355	290	250
3. Height - Overall (ft)	62	62	62	62	62	62	62	62	62	62	62	62
4. Pumping Head (ft)	48	48	48	48	48	48	48	48	48	48	48	48
5. No. Fans/Tower	12	16	13	12	16	12	12	13	12	16	12	12

EBASCO SERVICES INCORPORATED
SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION
ROUND MECHANICAL TOWER
(For Commercial Plant)

EXHIBIT III
Sheet 2 of 3

COOLING RANGE - F	17			20			25			30		
COOLED WATER TEMP. F.	83	89	94	83	89	94	83	89	94	83	89	94
C. INVESTMENT - $1000												
1. Site Preparation (Grading & Excavation)	189	77	58	167	68	51	155	63	47	76	51	38
2. Piling	1532	624	468	1365	556	415	1208	492	369	593	398	296
3. Cooling Tower Basin	737	300	225	655	267	200	580	236	177	284	191	132
4. Intake Structure	473	473	473	402	402	402	322	322	322	268	268	268
5. Circ. Water Conduit	292	280	280	270	260	260	255	248	248	215	206	206
6. Cooling Tower	7492	4772	4119	7043	4486	3872	6424	4092	3532	5692	3626	3130
7. Circ. Water Pumps & Motors	1360	1360	1360	1129	1129	1129	860	860	860	715	715	715
8. Switchgear & Wiring	1115	983	861	960	846	741	792	698	611	739	628	544
9. Instruments & Controls	75	64	55	75	64	50	75	55	50	64	50	50
TOTAL DIRECT COST	13265	8933	7899	12066	8078	7120	10671	7066	6266	8646	6133	5379

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SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION

EXHIBIT III (Cont'd)
Sheet 3 of 3

ROUND MECHANICAL TOWER

(For Commercial Plant)

COOLING RANGE - F	17			20			25			30		
COOLED WATER TEMP. - F	83	89	94	83	89	94	83	89	94	83	89	94
D. **ADJUSTMENTS**												
1. Process Equipment	-956	BASE	+722	-566	+254	+936	-39	+829	+1482	+351	+1404	+2067
2. Condensing Equipment	BASE	+154	+338	+39	+213	+429	+120	+340	+639	+227	+530	+1021
TOTAL ADJUSTED INVESTMENT	12309	9087	8959	11539	8545	8485	10752	8235	8387	9224	8067	8467
E. **ANNUAL COSTS**												
1. Fixed Charges on Adjusted Investment	821	606	598	770	570	566	717	549	559	615	538	565
2. Cooling System Energy @ $0.0185/kWh	1491	1448	1409	1285	1247	1211	1063	1029	996	992	926	887
3. Drive Turbine Performance Adjustment	BASE	+355	+537	+37	+390	+563	+109	+463	+608	+180	+537	+652
4. Estimated Maintenance	181	121	105	169	113	95	155	97	87	118	85	78
TOTAL ADJUSTED ANNUAL COST	2493	2530	2649	2261	2320	2435	2044	2138	2250	1905	2086	2182

EXHIBIT IV
Sheet 1 of 3

EBASCO SERVICES INCORPORATED
SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION

NATURAL DRAFT TOWER

(For Commercial Plant)

	COOLING RANGE - F	17		20		25		30	
	COOLED WATER TEMP. - F	89	94	89	94	89	94	89	94
A.	DESIGN CONDITIONS								
	1. Heat Load - 10^6 Btu/hr	2720	2720	2720	2720	2720	2720	2720	2720
	2. Cir. Water - 10^3 GPM	320	320	272	272	217.6	217.6	181.3	181.3
B.	TOWER DESCRIPTION								
	1. No. of Towers	1	1	1	1	1	1	1	1
	2. Basin Diameter (ft)	340	290	325	280	310	270	280	250
	3. Height - Overall (ft)	450	370	450	370	370	350	370	350
	4. Pumping Head (ft)	44	40	43	39	41	38	39	38

EXHIBIT IV
Sheet 2 of 3

EBASCO SERVICES INCORPORATED
SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION

NATURAL DRAFT TOWER

(For Commercial Plant)

COOLING RANGE - F	17		20		25		30	
COOLED WATER TEMP. - F	89	94	89	94	89	94	89	94
INVESTMENT - $1000								
1. Site Preparation (Grading & Excavation)	74	63	68	47	62	47	56	44
2. Piling	570	485	523	358	476	359	430	338
3. Cooling Tower Basin	273	232	251	172	228	172	206	162
4. Intake Structure	473	473	402	402	322	322	268	268
5. Circ. Water Conduit	280	280	260	260	248	248	207	207
6. Cooling Tower	8972	6949	8202	6498	7615	6225	6381	5700
7. Circ. Water Pumps & Motors	1360	1360	1129	1129	860	860	720	720
8. Switchgear & Wiring	550	550	456	456	347	347	290	290
9. Instruments & Controls	30	30	30	30	30	30	30	30
TOTAL DIRECT COST	12852	10422	11321	9352	10188	8610	8588	7759

EXHIBIT IV (Cont'd)
Sheet 3 of 3

EBASCO SERVICES INCORPORATED
SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
TRADE-OFF STUDY V - COOLING TOWER OPTIMIZATION
NATURAL DRAFT TOWER
(For Commercial Plant)

COOLING RANGE - F	17		20		25		30	
COOLED WATER TEMP. - F	89	94	89	94	89	94	89	94
ADJUSTMENTS								
1. Process Equipment	BASE	+722	+254	+936	+829	+1482	+1404	+2067
2. Condensing Equipment	+154	+338	+213	+429	+340	+639	530	+1021
TOTAL ADJUSTED INVESTMENT	13006	11482	11788	10717	11357	10731	10522	10847
ANNUAL COSTS								
1. Fixed Charges on Adjusted Investment	868	766	786	715	758	716	702	723
2. Cooling System Energy @ $0.0185/kWh	1095	1095	909	909	693	693	577	577
3. Drive Turbine Performance Adjustment	+355	+537	+390	+563	+463	+608	+537	+652
4. Estimated Maintenance	+129	+104	+113	+94	+104	+608	+86	+78
TOTAL ADJUSTED ANNUAL COST	2247	2502	2198	2281	2016	2103	1902	2030

APPENDIX

ATMOSPHERIC EFFECTS OF ALTERNATE COOLING TOWER OPERATION

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM
TRADE-OFF STUDY V
COOLING TOWER SYSTEMS

APPENDIX

ATMOSPHERIC EFFECTS OF ALTERNATE COOLING TOWER OPERATION

I - INTRODUCTION

The purpose of this cooling system study is to qualitatively assess and compare the atmospheric effects of the following cooling tower alternatives:

A. Hyperbolic natural draft tower (NDCT).

B. Rectangular mechanical draft tower (MDCT).

C. Round mechanical draft tower (RMDT).

Cooling tower impacts can only be evaluated within the context of site geography and location with respect to other facilities. The Baskett Site is located about two miles south of the Ohio River at an elevation of about 400 ft msl. The terrain is fairly hilly to the south, east, and west of the site, but elevations are less than 600 ft msl within five miles of the site. A relatively flat flood plain extends north of the site to the Ohio River. The proposed location of the cooling tower (1) is about 800 feet northwest of Tscharner Road, 2,000 feet north of the Louisville and Nashville Railroad, a little less than 1 mile northwest of the town of Baskett and 1.2 miles north of US Route 60 at its closest point.

The major atmospheric effects associated with cooling tower operation are the following:

- Elevated visible plumes
- Ground level fogging and icing
- Salt deposition and drift

Most of these atmospheric effects result from the condensation of water vapor and the entrainment of large water drops from the tower by the exhaust air. Water vapor condensation may lead to fogging, icing or elevated visible plumes and the large drop entrainment may cause deposition of salts from the tower onto surrounding environs. Most studies of cooling tower impact have concentrated on elevated plumes, ground level fogging and icing and salt deposition. Very little research has been done regarding the release of latent heat (and to a much smaller extent, sensible heat) during tower operation and the associated convective instability.

The selected tower type for the Demonstration Plant will have a circulating water rate of 60,400 gpm and a heat rejection rate of 9.07×10^8 Btu/hr. The Commercial Plant tower will have a circulating water rate of 181,000 gpm and a heat rejection rate of 2.72×10^9 Btu/hr. Design dry bulb and wet bulb temperatures for both Plants are 85°F and 76°F respectively. Cooling tower parameters for the Demonstration and Commercial Plants are given in Tables 1 and 2, respectively, and a representative drift mass distribution for the alternate tower types is given in Table 3.

II - EVALUATION OF ATMOSPHERIC EFFECTS

A - ELEVATED VISIBLE PLUMES

Elevated visible plumes are caused by the condensation of water vapor and are therefore, relatively free of impurities. Warm saturated air leaving the tower mixes with cooler, drier ambient air. Because of the non-linear relationship between saturation vapor pressure and air temperature, the mixture is usually supersaturated and the excess moisture condenses in the form of an elevated visible plume. The plumes generally rise to heights of 1,000 to 3,000 feet with no impact at ground level. These elevated plumes are considered the primary atmospheric effect of NDCT and RMDT operation, but are not considered a significant factor in MDCT operation (2) . The effluent from these relatively low towers does not rise to the heights of NDCT and RMDT plumes. Furthermore, these low level MDCT plumes are dispersed relatively close to the tower due to the fan operation and tower downwash.

Based on a ten-year period of record (1951-1960) at Evansville, Indiana, the dominant directions of the elevated plumes would probably be southeast and north-northeast of the tower. A study by Smith (3) indicates that 93 percent of the elevated plumes from selected NDCT in Ohio and West Virginia dissipated within two miles of the tower. Plumes from the Demonstration and Commercial Plant NDCT or RMDT should be shorter than those studied by Smith.

The closest airports to the Baskett Site are Evansville, located 12 miles northwest, Henderson, located 12 miles southwest and Owensboro, located 20 miles southeast of the site. NDCT and RMDT plumes would not extend to these distances and, therefore, are not expected to interfere with airport air traffic.

B - GROUND LEVEL FOGGING AND ICING

Tower induced fogging at ground level occurs when relatively low level plumes from MDCT (and infrequently from RMDT) are brought down to the earth's surface. Such fogging is generally evaluated in terms of frequency exclusive of periods of natural fog. Ground level icing occurs when all criteria are present for tower induced fogging and the temperature of objects impacted by the plume is 32°F or less.

Observations indicate that significant fogging and icing caused by cooling towers do not usually extend beyond 2,000 feet from the tower (4). Since Tscharner Road is 800 feet southeast of the proposed tower, tower induced fogging along this highway is a potentially significant factor associated with MDCT operation. Tower induced fogging along the Louisville and Nashville Railroad is expected to be less significant due to its greater distance from the tower.

Ground level fogging and icing are generally not considered significant effects of natural draft tower operation (5). According to Carson, (2) ground level fogging from ground towers is greater than that from natural draft towers but less than the fogging due to rectangular tower operation. Plumes from both natural draft and round towers usually attain greater heights than those from rectangular towers and there is no elevated terrain of sufficient height in the site vicinity which could lead to plume impingement.

C - SALT DEPOSITION AND DRIFT

A smaller fraction of the circulating water in a cooling tower is entrained in the air rising out of the tower. These water droplets which contain dissolved and suspended chemicals are called drift. Table 3 gives a representative drift mass distribution for the cooling tower types being considered for the Demonstration and Commercial Plants. Most of the solid material carried in these droplets from freshwater cooling towers is calcium sulfate and these "salt" drift droplets may be carried downwind from the tower and subsequently fall to ground level.

Carson has reviewed the literature pertaining to salt deposition from freshwater cooling towers (2) . He concludes that these towers, equipped with state-of-the-art drift eliminators, will have "very small" deposition rates, and that most of the drift will fall to the ground within two thousand feet of the tower.

Roffman indicates that salt deposition rates from NDCT are lower by a factor of five than from MDCT but that the salt drift will travel further downwind from the tower and affect a greater area (6) . Salt deposition rates from RMDT also are less (but affect a larger area) than those from MDCT. These differences are generally a result of the greater heights of NDCT and RMDT compared to MDCT.

III - RESULTS

The results of this study indicate that the major effect of NDCT and RMDT operation at the Demonstration and Commercial Plants would be the generation of visible elevated plumes. Possible impacts of such plumes are aesthetics, ground shading and reduction in visibility to aircraft flying at plume level. No interference with airport air traffic, however, is expected from these plumes.

The major atmospheric effect of MDCT operation is the potential for tower induced fogging on Tscharner Road, located 800 feet from the proposed tower location. Possible mitigative measures include relocating the tower to a greater distance from Tscharner Road and orientation of the tower to minimize tower induced fogging due to downwash.

Salt deposition in the nearby Site region will result from the operation of any one of the three alternate cooling tower types, but salt deposition rates are generally low from freshwater tower operation.

REFERENCES

1. Demonstration Plant Site Plan, SK-8375-M-8, May 1, 1978.

2. Carson, J E, "Atmospheric Impacts of Evaporative Cooling Systems," Argonne National Laboratory, October 1976.

3. Smith, M E et al, "Cooling Towers and the Environment," Am Elec Power Corp, October 1974.

4. "Development of Document of Effluent Limitations Guidelines and New Source Performance Standards for the Steam Electric Power Generating Point Source Category," EPA 440/1-74 029-a, 1974.

5. Hosler, C L, "Wet Cooling Tower Plume Behavior," AICHE Cooling Tower Symposia, Houston, Texas, March 1971.

6. Roffman, A, "Environmental Economic and Social Considerations in Selecting a Cooling System for a Steam Electric Generating Plant," Cooling Tower Environment - 1974, ERDA, 1975.

TABLE 1

COOLING TOWER PARAMETERS* FOR DEMONSTRATION PLANT

BASKETT SITE

PARAMETER(1)	NATURAL DRAFT	RECTANGULAR MECHANICAL DRAFT	ROUND MECHANICAL DRAFT
Design Dry Bulb Temp (F)	85	85	85
Design Wet Bulb Temp (F)	76	76	76
Heat Rejection Rate (Btu/hr)	9.07×10^8	9.07×10^8	9.07×10^8
Circulating Water (GPM)	6.04×10^4	6.04×10^4	6.04×10^4
Exit Velocity (m/sec)	2.1	7.0	6.6
Number of Towers	1	1	1
Salt Concentration (ppm)	2,000	2,000	2,000
Tower Height (ft)	272	42	65
Tower Length (ft)	-	210	-
Tower Width (ft)	-	64	-
Diameter at Top Tower (ft)	131	-	144
Diameter at Base Tower (ft)	218	-	144
Number of Fans	0	3	5
Fan Diameter (ft)	-	42	33
Drift Rate (%)	0.002	0.005	0.005
Air Flow (lb/hr)	2.27×10^7	2.28×10^7	2.28×10^7

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(1) PARAMETERS ARE GIVEN FOR DESIGN CONDITIONS.

DATA SOURCE: EBASCO SERVICES, MAY 1978.

* BASED ON PRELIMINARY SELECTION AND APPROXIMATED FOR ENVIRONMENTAL CONSIDERATION.

TABLE 2

COOLING TOWER PARAMETERS* FOR COMMERCIAL PLANT

BASKETT SITE

PARAMETERS(1)	NATURAL DRAFT	RECTANGULAR MECHANICAL DRAFT	ROUND MECHANICAL DRAFT
Design Dry Bulb Temp (F)	85	85	85
Design Wet Bulb Temp (F)	76	76	76
Heat Rejection Rate (Btu/hr)	2.72×10^9	2.72×10^9	2.72×10^9
Circulating Water (GPM)	1.81×10^5	1.81×10^5	1.81×10^5
Exit Velocity (m/sec)	3.5	6.3	7.3
Number of Towers	1	1	1
Salt Concentration (ppm)	2,000	2,000	2,000
Tower Height (ft)	374	40	65
Tower Length (ft)	-	630	-
Tower Width (ft)	-	64	-
Diameter at Top Tower (ft)	180	-	250
Diameter at Base Tower (ft)	299	-	250
Number of Fans	0	9	14
Fan Diameter (ft)	-	42	33
Drift Rate (%)	0.002	0.005	0.005
Air Flow (lb/hr)	6.81×10^7	6.85×10^7	6.80×10^7

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(1) PARAMETERS ARE GIVEN FOR DESIGN CONDITIONS.

DATA SOURCE: EBASCO SERVICES, MAY 1978

* BASED ON PRELIMINARY SELECTION AND APPROXIMATED FOR ENVIRONMENTAL CONSIDERATION.

TABLE 3

DRIFT MASS DISTRIBUTION FOR ALTERNATE COOLING TOWERS

BASKETT SITE

DROP DIAMETER (MICRONS)	DRIFT MASS (PERCENT)
<50	40
50-100	24
100-200	10
200-350	7
350-500	9
>500	10

ALTERNATE COOLING TOWER TYPES ARE HYPERBOLIC NATURAL DRAFT, RECTANGULAR MECHANICAL DRAFT AND ROUND MECHANICAL DRAFT.

DATA SOURCE: EBASCO SERVICES, MAY 1978.