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TRADE-OFF STUDIES REPORT IV
AIR QUALITY CONTROL ALTERNATIVES

SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

PHASE I

SEPTEMBER 1978

PREPARED UNDER CONTRACT NO. ET-77-C-01-2577

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NOTICE

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Prepared for the

U.S. DEPARTMENT OF ENERGY

Assistant Secretary for Energy Technology
Office of Fossil Fuels

Under CONTRACT ET-77-C-01-2577
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SYNTHESIS GAS DEMONSTRATION PLANT PROGRAM

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PREPARED FOR THE
DEPARTMENT OF ENERGY
DIVISION OF ENERGY TECHNOLOGY
UNDER CONTRACT ET-77-C-01-2577

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 rule-making 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 shown in 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 <u>multicyclonic mechanical dust collector</u> (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.

<u>Electrostatic precipitators</u> (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:

<u>Process Development Status</u>

a) Successful operation of 100 MW equivalent size class

b) Existence of qualified supplier

<u>Process Capabilities and Requirements</u>

a) Emission level performance capability

b) Environmental acceptability of process waste products

c) Acceptable interface with the balance of the plant

<u>Technical and Economic Feasibility</u>

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 successful 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 by-product.

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 ef-efficiency 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

* 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. <u>Energy Requirements</u>

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

* 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 byproduct. 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

* 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

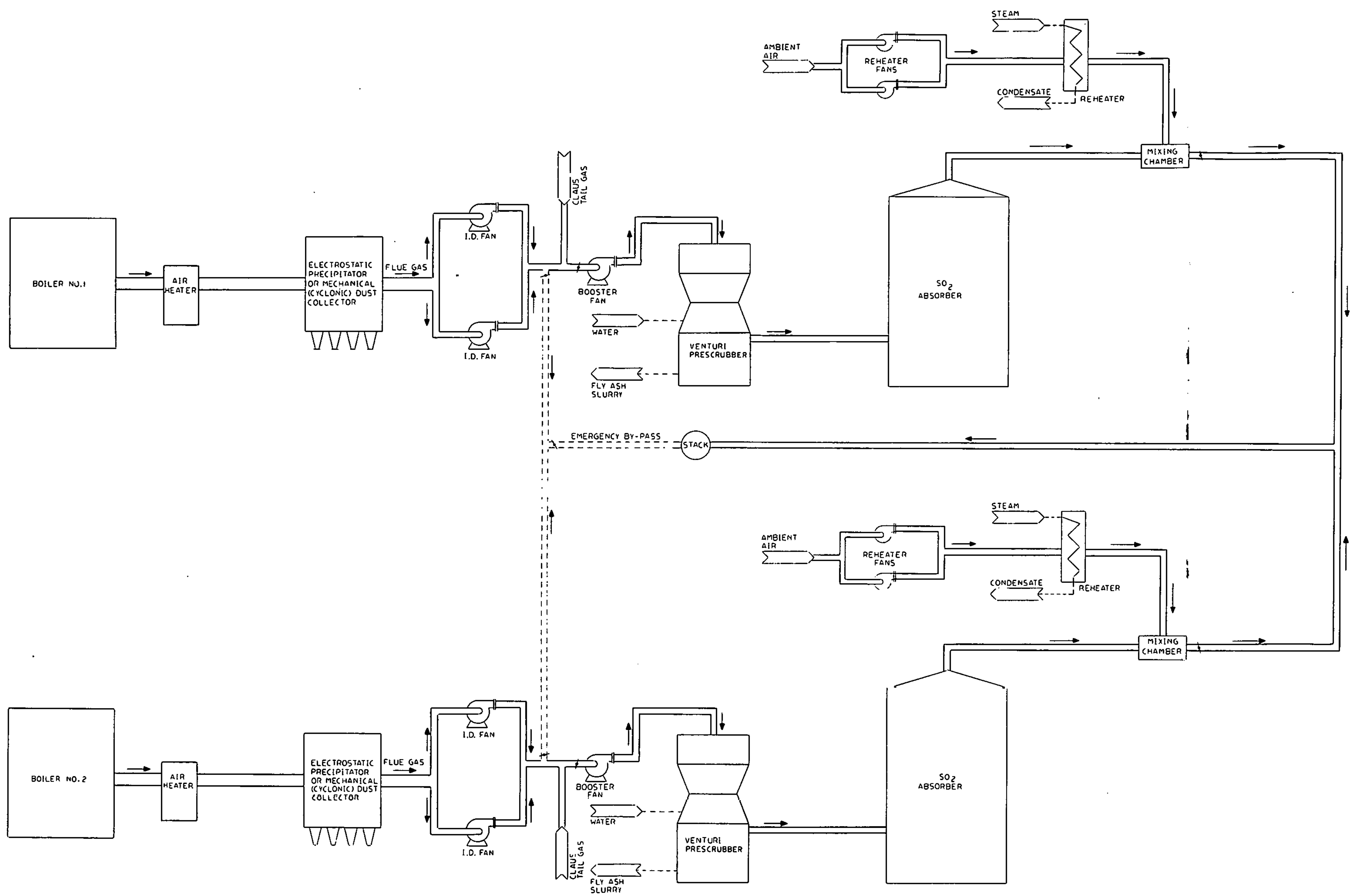

EXHIBIT 5

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.		SECT. CODE NO.
DR. J. KARLICK		SK-8376 M-4
CH.		ISSUE P0

SK-8376 M-5

NO.	DATE	REVISION	BY	CH.	APPROVED	REVIEW

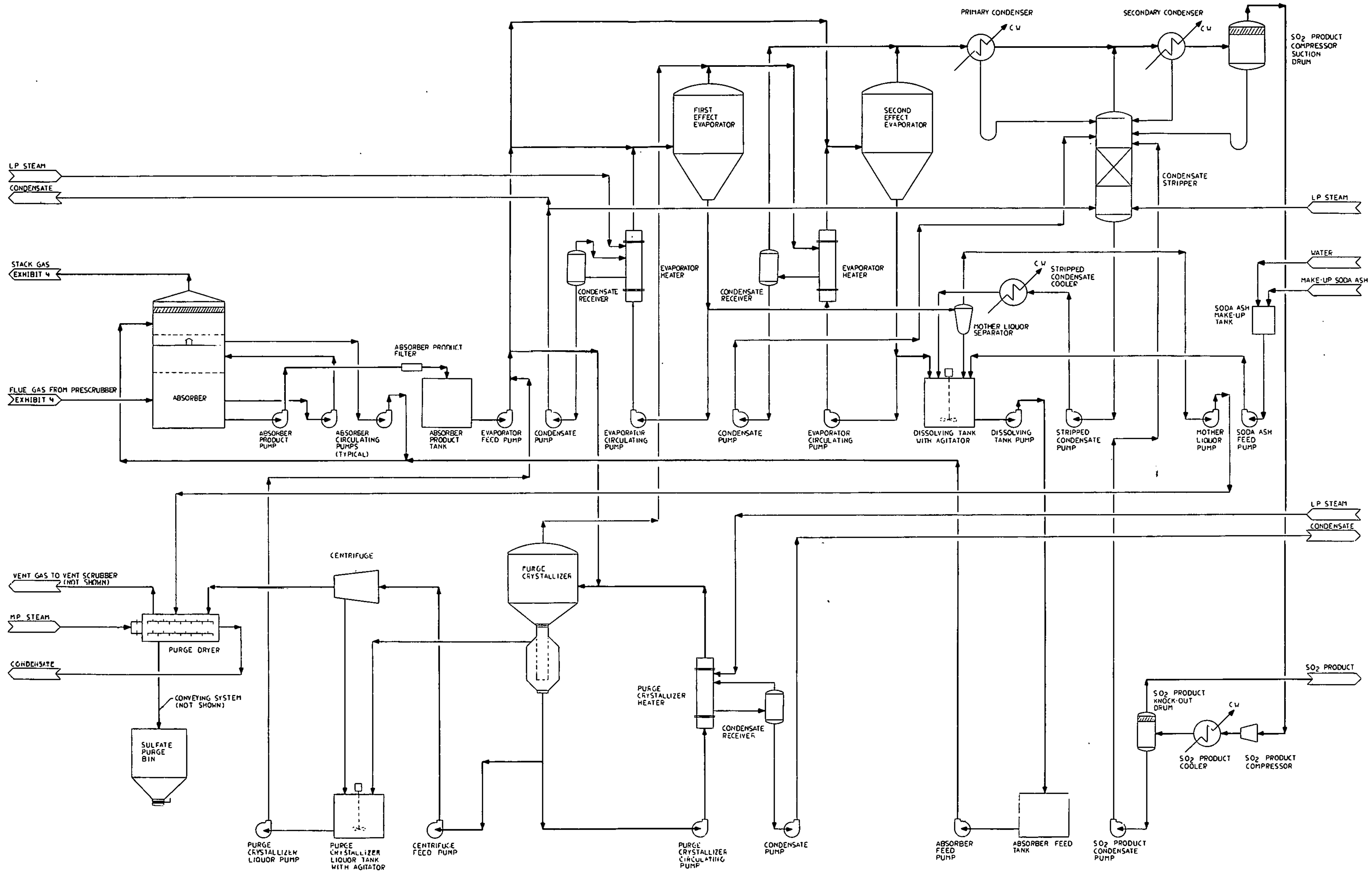

EXHIBIT 6

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		SECT. CODE NO.
DR. OLIVETO		SK-8376 M-5 / ISSUE P0
CH. -		

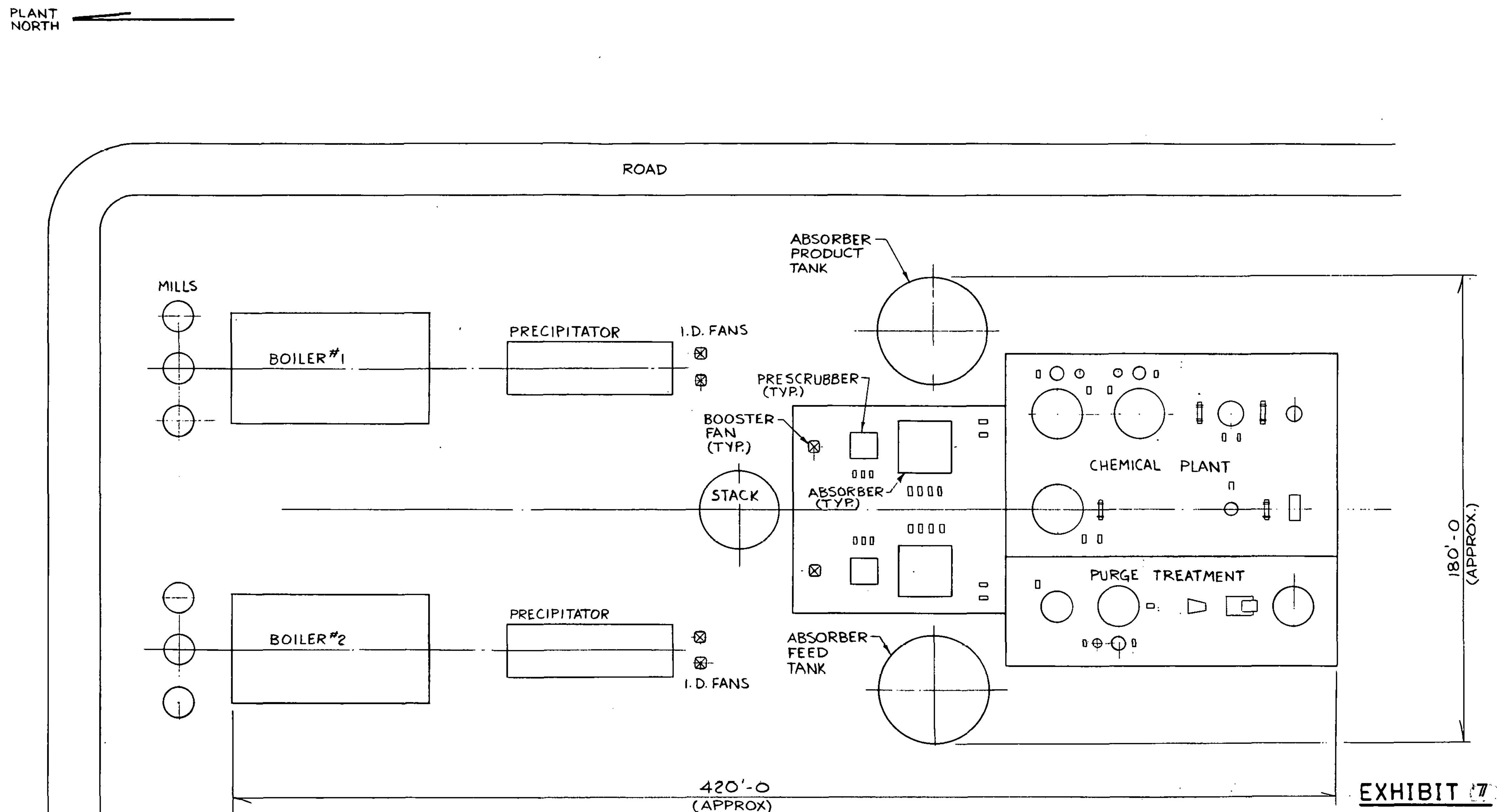

EXHIBIT 7

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

REV.	DATE	BY	APPROVED
5			
4			
3			
2			
1			

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 P.

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

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

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

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

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

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

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:	1
Air Blower	
Air Filter	
Surge Hopper	
Surge Hopper Rotary Feeder	
Dust Collector	
Dust Collector Rotary Feeder	
Gas Reheaters	2
Gas Reheat Fans	2 + 2

SYNTHESIS GAS DEMONSTRATION PLANT

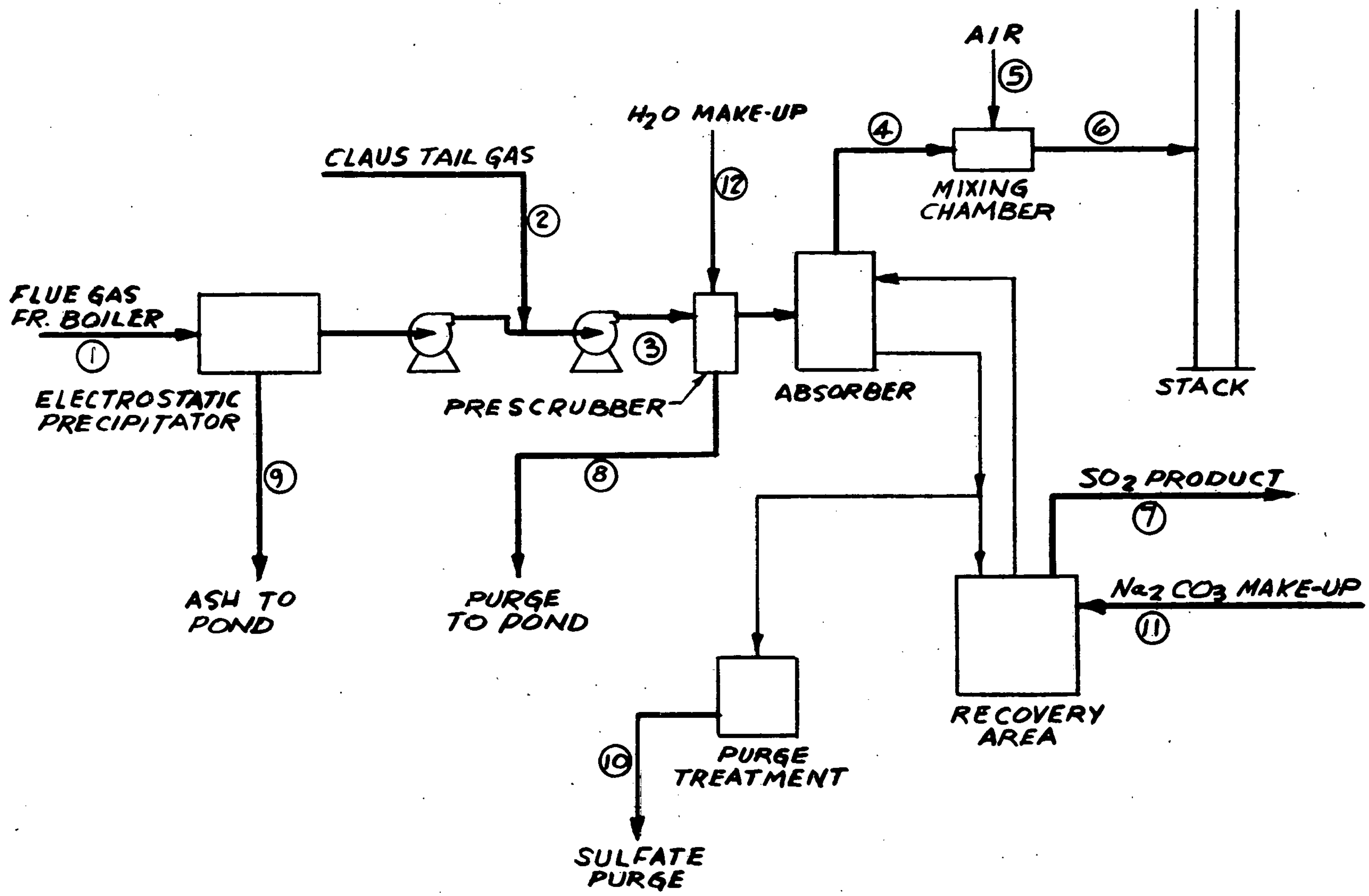

EXHIBIT 11

EBASCO SERVICES INCORPORATED

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

APPROVED

SCHEMATIC FLOW DIAGRAM
AIR QUALITY CONTROL SYSTEM

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

PARTICULATE REMOVAL EQUIPMENT & ID FANS

ANNUAL OPERATING COST

1978 Cost Basis

Item	Unit Cost	Electrostatic Precipitator (ESP) Quantity	$1000	Mechanical Dust Collector (MDC) 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 Unit Cost	Bag House Quantity	Bag House $1000	Cold-Side ESP $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