Comparative Evaluation of High and Low Temperature Gas Cleaning for Coal Gasification–Combined Cycle Power Systems

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Prepared by
Stone & Webster Engineering Corporation
New York, New York

ELECTRIC POWER RESEARCH INSTITUTE

COMPARATIVE EVALUATION OF HIGH- AND LOW-TEMPERATURE GAS CLEANING FOR COAL-GASIFICATION – COMBINED-CYCLE POWER SYSTEMS

EPRI AF-416
(Research Project 243-2)

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Prepared by

Stone & Webster Engineering Corporation
Process Industries Group
One Penn Plaza, New York, New York 10001

Principal Investigators
C. H. Jones
J. M. Donohue

MASTER

Prepared for

Electric Power Research Institute
3412 Hillview Avenue
Palo Alto, California 94304

Project Manager
M. J. Gluckman

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ABSTRACT

The purpose of this screening study was to evaluate the incentives for developing hot gas purification technology for application to coal gasification-combined cycle power generating systems. The iron oxide process currently being developed by the Morgantown Energy Research Center for removal of hydrogen sulfide at high temperature (1,000°F) was selected for this study as it was judged to be in a more advanced stage of development than other high temperature absorption processes currently being developed. Process and economic evaluations were performed for five different coal gasification schemes, i.e. air and oxygen blown dry ash, moving bed Lurgi gasifiers; oxygen blown slagging, moving bed gasification currently being developed by the British Gas Corporation; and oxygen and air blown two-stage entrained gasifiers proposed by Foster Wheeler.

For each of the above gasification schemes, four complete system flowsheets were developed for converting Illinois #6 coal to electricity via combined cycle power generation. Two of these processing schemes were based on high temperature iron oxide technology with gas turbine inlet temperatures of 1,950°F and 2,400°F. The other two flowsheets for each gasifier incorporated the low temperature Benfield process with gas turbine inlet temperatures of 1,950°F and 2,400°F.

The results of this study indicate that there is a large economic incentive for developing hot purification technology for dry ash Lurgi gasification systems. For advanced gasification schemes such as the BGC slagger or the two-stage entrained gasifier, no incentive could be identified for the development of high temperature iron oxide gas cleaning technology. Major technical questions that were raised concerned problems associated with the regeneration of iron sulfide, the fate of trace contaminants such as ammonia and alkali metal vapors in the system, as well as the ability to remove particulates from the gas streams both before and after contact with the iron oxide absorption equipment.

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Allied Chemical Corporation
P.O. Box 1139R
Morristown, NJ 07960

Sulfur Dioxide Reduction
William D. Hunter, Jr., Business
Manager, Sulfur Pollution Control
Services
R. G. Bierbower, Project Manager

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John P. Valentine, P. E.
Manager, Gas Purification Department

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377 Route 17
Hasbrouk Heights, NJ

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Mr. J. T. Gallagher, Vice President

Babcock & Wilcox
Alliance, Ohio 44601
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R. W. Curtis, Manager, Design and
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615 Washington Rd.
Pittsburgh, PA 15228
Benfield Purification Processes
J. H. Field, Vice President

Brunswick Corporation
One Brunswick Plaza
Skokie, IL 60076

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Kenneth R. Mils, Venture Director,
Technical Products Division
John R. Pollack, Manager
Air Pollution Control Engineering
Technical Products Division

City College New York
138th St. and Convent Ave.
NY, NY 10031
Panel Bed Filbration
Professor A. M. Squires
Dr. K. C. Lee, Research Fellow

Ducon Co., Inc.
147 E. 2nd St.
Mineola, NY 11051
Mr. S. M. Khan, Sales Manager

Fluor Engineers and Constructors, Inc.
P.O. Box 4051
Anaheim, CA 92803
Garrett E. Pack
Supervising Process Engineer

Foster Wheeler Energy Corporation
RESOX Technology
R. I. Wheater, District Manager
Technology, 20 William St.
Wellesley, MA 02181
(local address, head office:
Livingston, NJ)

General Electric Company
1 Washington St.
Wellesley, MA 02181
Mr. E. C. Lowe, Power Generation
Sales Engineer, New England

Morgantown Energy Research Center
Morgantown, WV
Hot Purification
Mr. E. C. Oldaker, Program Director

Rexnord Corporation
Air Pollution Control Division
P.O. Box 13226
Louisville, KY 40213
Panel Bed Filtration
J. M. Morris, Division President

United Technologies Research Center
East Hartford, CT 06108
Combined Cycle Technology
Dr. F. L. Robson
Chief Utility Power Systems

Water Purification Associates
238 Main St.
Cambridge, MA 02142
Dr. David J. Goldstein

Woodall Duckam (USA) Limited
200 Manor Oak
1910 Cochran Rd.
Pittsburgh, PA 15220
Appleby Frodingham Process
Mr. A. J. Grant, Manager

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INTRODUCTION AND SUMMARY

INTRODUCTION

Application of high temperature purification of coal derived fuel gas in electrical power generation at near to gasifier exit temperature may lead to reduced energy losses, reduced environmental pollution and reduced capital costs. Such a hot process would replace the established cold scrubbing processes - Benfield, Selexol, Stretford - which require cooling of the fuel gas prior to purification.*

A hot process for removing sulfur-and nitrogen-containing compounds and dust from fuel gas would justify application provided that it improves energy production economics to levels significantly above current experience and that it satisfies environmental regulations. The objective of this study is to identify applications of hot purification technology to combined cycle power generation which may be economically justified and to review and comment on the application of hot purification technology.

Several gasification combined cycle systems have been investigated to determine the extent to which the incentive to adopt hot purification may be a function of the gasifier type. The hot purification process studied is the absorption of sulfur compounds from fuel gas in fixed beds of iron oxide at approximately 1,000°F being developed by the Morgantown Energy Research Center of ERDA. This process was judged to be at a more advanced stage of development than other high temperature absorption processes being developed by Conoco Coal Development Co. (half calcined dolomite), Air Products (calcined limestone) and Battelle (molten carbonates).

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*Unless otherwise explained "cold purification" or the word "cold" (without quotation marks) refers to fuel gas purification in the temperature range 100 F to 300 F using established scrubbing processes such as Benfield, Selexol, etc.

"Hot purification" or the word "hot" (without quotation marks) refers to fuel gas purification at approximately 1,000°F by absorbing sulfur compounds in a fixed bed of solid iron oxide pellets.

In cold purification based schemes, pressurized (300 psig) gasifier product is cooled, either by quenching followed by heat exchanger cooling if it contains tars or cooled in heat exchangers alone if it does not contain tars. The cooled gas passes through a Benfield unit where hydrogen sulfide is removed to the required extent before the gas enters the combustor of the gas turbine. Hot gas leaving the gas turbine enters the heat recovery steam generator where steam is raised for the steam turbine power generation system.

In hot purification schemes, pressurized gasifier product is adjusted to approximately 1,200°F as it leaves the gasifier. The hot gas leaving the gasifier contains appreciable amounts of particulate material which must be reduced in concentration before the gas is introduced to fixed beds of sulfur absorbent to prevent plugging. Desulfurized gases leaving the sulfur absorbent beds may contain particulate material which could be damaging to a gas turbine. A "polishing" filter is therefore assumed to be required downsteam of the sulfur absorption beds. When the absorbent bed becomes saturated with sulfur, it is removed from the process for regeneration and a freshly regenerated bed substituted in its place.

Major topics introduced in the foregoing paragraphs are presented with more detail in the remaining introductory material. Relevant discussion is reserved for the appropriate section of the report.

Gasification Methods

Consistent with the stated objectives of the study, the following gasifiers have been evaluated in association with hot and cold purification processes and varying gas turbine inlet temperatures:

- The Lurgi pressurized, moving bed, air blown gasifier which produces a dry ash residue - designated type MA
- The Lurgi pressurized, moving bed, oxygen blown gasifier which produces a dry ash residue - designated type MX
- The British Gas Corporation pressurized type moving bed gasifier. This gasifier produces a slag rather than a dry ash residue and is oxygen blown - designated type MS
- The air blown entrained bed gasifier currently under development by Foster Wheeler Corporation - designated type EA
- The oxygen blown entrained bed gasifier currently under development by Foster Wheeler Corporation - designated type EX

All related gasifier heat and material balance data used in this study were provided by the Electric Power Research Institute (EPRI).

The addition of "H" or "L" to the above designated type code indicates hot or cold purification, e.g., MSL and MSH refer to slagging gasification with cold and hot purification respectively.

The Lurgi moving bed dry ash producing gasifier is the only available commercial pressure gasifier. Coal is admitted to the top of the Lurgi gasifier through a lockhopper and oxidant and steam are admitted to the bottom of the vessel. A dry ash residue is withdrawn through an ash lock in the bottom of the gasifier. Gas phase and coal encounter each other in countercurrent flow and the residence time of the coal in the vessel is long, which leads to relatively high unit gasification costs. The Lurgi gasifier has been widely described in recent years. A study of oxygen blown Lurgi gasification in application to American coals has been performed by ERDA; the related publication (4) provides background in this technology.

Slagging gasification employs low steam rates in order to establish a base temperature high enough for the production of a liquid slag. Reducing steam supply leads to increased gasifier efficiency and higher capacity per unit gasifier area, as discussed by Hebden (9) (10). This form of gasification is currently undergoing development by British Gas Corporation and fourteen American companies at Westfield, Scotland. Work in slagging gasification also has been reported by Gronhovd (8).

The two stage entrained flow gasifier is being developed by Foster Wheeler but is less advanced than the BGC slagger. The gasifier consists of two stages. In the lower stage, char is gasified at high temperature in entrained flow with air or oxygen and steam. The hot gases pass into the upper stage where they are concurrently quenched with a stream of pulverized coal entrained in a transport gas stream. As quenching takes place, the pulverized coal is devolatilized and converted to char. Gas char mixture leaving the upper stage passes through cyclone separators from which char is removed and returned to the lower part of the gasifier. Separated fuel gas at approximately 1,700°F leaves the cyclone char separator and enters the gas cooling and purification system. Entrained flow gasifiers also discharge ash in the form of a molten slag.

The entrained flow gasifier is not expected to produce tars. Residence time in this gasifier is short resulting in high capacity. Entrained flow gasification data for this study were supplied to EPRI by Foster Wheeler Corporation. Descriptions of the Foster Wheeler BCR entrained flow gasifier have been published (20) (24) but for the most part the information is brief.

Hot Fuel Gas Purification

The hot fuel gas purification process is being developed by the Morgantown Energy Research Center in application with an air blown moving bed gasifier. The Morgantown gasification unit has a stirred bed to enable it to process caking coals and is of the Wellman Galusha type which is similar in many respects to the Lurgi air blown type of gasifier.

In Morgantown work the absorbent is a regenerable iron oxide which has been tested in association with a number of support materials. Of the support materials studied, fly ash and silica appear to lead to composite materials with satisfactory mechanical and chemical resistance for commercial development.

Regeneration of the iron oxide absorbent can be accomplished according to Morgantown work by blowing nitrogen enriched air through the sulfided bed of absorbent. Use of a nitrogen enriched air stream is necessary to avoid fusion of the absorbent bed during regeneration.

A second potential method of sulfided absorbent regeneration is based on passing sulfur dioxide oxygen mixture under pressure through the absorbent.

Study of the Morgantown hot desulfurization process follows recommendations in EPRI Report 243-1 (23).

Sulfur Dioxide Reduction

All methods for regeneration of the sulfided iron oxide produced in hot purification lead to the production of sulfur dioxide for further treatment. There is a possibility that elemental sulfur may be produced by the reaction of sulfur dioxide with iron sulfide. However, it is expected to be necessary to reduce sulfur dioxide to elemental sulfur by methods other than its reaction with iron sulfide. Such reduction may be accomplished employing coal or process gas as the reducing agent, though further development is required to demonstrate this.

The sulfur dioxide reduction processes considered are the process gas based reducing process of Allied Chemical Corporation and the coal based RESOX process of Foster Wheeler Corporation.

Particulate Material Removal from Fuel Gas

Removal of dust from hot high pressure fuel gas (1,000°F 300 psig) is not understood in both the magnitude of the problem and the extent of dust removal required. Processes for accomplishing dust removal under such conditions are so little developed that it is impossible to develop better than rough budget information for use in cost estimations. The level of fuel contamination with dust, specification of levels to be established in clean fuel gas, and the effectiveness and suitability of available dust removal equipment are subject to a high degree of uncertainty in relation to hot purification (20). In cold purification process configurations, a satisfactory degree of dust removal is incidentally achieved by the various scrubbing operations to which the fuel gas is subjected.

Devices considered for removing dust from hot high pressure fuel are:

- Cyclones for preventing the accumulation of dust in beds of solid sulfur absorbents as practised in Morgantown work.
- The panel bed sand filter being developed under the direction of Professor A. M. Squires at City College, New York (22). In this device, the gas to be filtered flows horizontally through a bed of granular solid supported in a tall thin panel between louvered walls. Dust from the treated gas accumulates on the entry surfaces of the panel between the louvers. A sharp surge of reverse flowing gas (the puff-back) is used to clean the entry surface of the panel and in doing so this causes a uniform spill of sand with the dust removed from the panel. In the gas turbine feed line (the exit line from an absorbent bed), the dust load may be small and insufficient to form the filter cake which must be present for the CCNY panel bed filter to be effective. In these circumstances, some other method of cleaning the hot gas turbine feed will probably be required.
- For gas turbine protection, the focus of attention has been on a filter employing a felted woven metal cloth manufactured by Brunswick Corporation.

The actual metallic cloth suitable for hot fuel gas purification service has not been identified through a test program. There is good indication that a satisfactory material can be developed at reasonable cost.

Filters manufactured from this type of cloth have achieved excellent performance on dust emissions from high pollutant potential cement plants, indicating a capability to effectively filter submicron particulates. The operating Brunswick filter has demonstrated performance at atmospheric pressure and about 300°F, though not in a fuel gas environment.

Costs and Thermal Efficiency Determinations

The absolute values of power cost and thermal efficiency for combined cycle systems reported in Sections 4 and 5 are approximate because of the screening nature of the study and different development stages of each gasifier. However, cost differences between hot and cold purification schemes should be realistic.

The major use of the results obtained in this study is to indicate with which types of gasifier the application of hot purification might be justified for combined cycle power generation.

Gas Turbine Protection

Before discussing and analyzing hot purification supported combined cycle units, the gas turbine oriented view of purification requires mention.

Gasifiers followed by quenching and subsequent scrubbing operations probably establish the complete removal of dust and trace vapor phase contaminants occurring in the fuel gas. Chlorine and sodium are of particular concern as dangers to the gas turbine. The high additional degree of purification provided in the cold purification type process configuration may be essential for securing prolonged gas turbine life, in which case hot purification has no future in combined cycle applications.

In meetings in support of this project, General Electric cautioned against the development of gasification-purification schemes without regard for the stringent constraints of the gas turbine on trace materials in the fuel. As things stand, hot purified fuel gases may not be suitable for use in high temperature gas turbines even with near perfect fuel filtration. Establishment of answers to such questions must be through a development program leading to the simultaneous testing of all the component equipment in the hot purification supported combined cycle unit in an integrated system.

Related Topics

The integration of coal gasification, hot fuel gas purification and combined cycle electrical power generation is in an early stage of development. Practical use of hot purified coal derived gas has been made in the Appleby-Frodingham process (early 1960s) which employed a fluidized bed of iron oxide for desulfurization of coke oven gas (see Appendix II).

Figures 4-1, 4-2 and 4-7 through 4-14 illustrate the general gasification, purification, combined cycle flow schemes under consideration.

The hot processes have no capability to remove nitrogen compounds from fuel gas. The cold processes vary in ability to remove ammonia and other nitrogen compounds from fuel gas. The combined cycle NO_x emission problem is not discussed in detail in this report but it may prove to be the most important factor opposing application of hot purification. The topic has been discussed recently by Robson, et al. (21) in work performed under contract for the EPA.

SUMMARY

The purpose of this screening study was to evaluate the incentives for developing hot gas purification technology for application to coal gasification-combined cycle power generating systems. The iron oxide process currently being developed by the Morgantown Energy Research Center for removal of hydrogen sulfide at high temperature (1,000°F) was selected for this study, as it was judged to be in a more advanced stage of development than other high temperature absorption processes being developed by Conoco Coal Development Company (half calcined dolomite), Air Products and Chemicals, Inc. (calcined limestone) and Battelle Northwest (molten carbonates). The Benfield potassium carbonate process was chosen to represent low temperature liquid absorption systems for hydrogen sulfide removal.

As it was anticipated that the advantages to be associated with hot gas purification would depend, to a large extent, on the particular gasification process employed, five different coal gasification schemes were investigated, i.e., air and oxygen blown dry ash, moving bed Lurgi gasifiers; oxygen blown slagging, moving bed gasification currently being developed by the British Gas Corporation; and oxygen and air blown two-stage entrained gasifiers proposed by Foster Wheeler Energy Corporation.

For each of the above gasification schemes, four complete system flow sheets were developed for converting Illinois No. 6 coal to electricity via combined cycle power generation. Two of these processing schemes were based on high temperature iron oxide technology with gas turbine inlet temperatures of 1,950°F and 2,400°F. The other two flow sheets for each gasifier incorporated the low temperature Benfield process with gas turbine inlet temperatures of 1,950°F and 2,400°F. Complete heat and material balance data were generated for each of the systems studied based on an overall net electrical capacity of 1,000 Mw. Plant investment estimates for each system were based on cost data presented by Fluor Engineers and Constructors, Inc. (24) as well as Stone & Webster in-house estimates of the costs associated with the iron oxide system and high temperature particulate removal equipment. Cost of electricity estimates were generated using a set of economic criteria consistent with those used by Fluor (24).

The remainder of this section of the report will be devoted to a review of hot purification technology and summaries of the thermal efficiency and economic studies performed.

Hot Purification Technology Review

Small scale tests conducted by the Morgantown Energy Research Center (MERC) of ERDA have resulted in the following observations concerning the use of supported iron oxide absorbents for hydrogen sulfide capture:

- Maintaining absorption temperatures above 1,000°F avoids plugging of absorbent beds by tar condensation and decomposition.
- Dust trapped in the beds is largely carbonaceous in character and is burned out during regeneration.
- Hydrogen sulfide capture to the level of 10 wt % is feasible.
- Regeneration of sulfided iron oxide absorbent is easily accomplished. A fly ash supported iron oxide absorbent has demonstrated 174 consecutive operation/regeneration cycles without substantial property changes.

Absorbent development work at MERC has been conducted in laboratory scale (1 lb/hr coal) and bench scale (200 lb/hr coal) equipment. This work has demonstrated that fly ash and silica supported iron oxide are suitable for commercial use in fixed bed equipment. Studies conducted by MERC include:

- Effect of temperature on absorbent capacity
- Effect of fuel gas steam content on absorbent capacity
- Absorbent aging effects
- Regeneration with oxygen and nitrogen mixtures

Regeneration processes aimed at producing elemental sulfur by reaction of the sulfided absorbent with sulfur dioxide are being investigated. If nitrogen enriched air is used for regeneration to avoid fusing the absorbent bed, the off-gas tends to be low in sulfur dioxide content and to resemble stack gas in treatment problems. The sulfur dioxide-iron sulfide reaction offers a potential method of avoiding these problems.

A regeneration process based on recycling sulfur dioxide appears to be feasible and to have advantage over using nitrogen enriched air. Such a process is currently under review for potential development by MERC.

Methods of reducing sulfur dioxide to elemental sulfur are available or are being developed. Major problems in regeneration are to produce a steady flow of sulfur dioxide containing gases from the cycling absorbent beds and to control heat release accompanying the regeneration operation. The disposal of sulfur dioxide by converting it to sulfuric acid was not considered.

A review of hot purification process chemistry indicates areas of concern. For example, the ferric to ferrous reduction in a freshly regenerated absorbent bed may impair combined cycle performance and represent a thermal penalty due to the consumption of hydrogen or carbon monoxide. Morgantown results must be regarded as specific to the moving bed, air blown, Lurgi-type of gasifier used as the basis for experimental work. The reason for this belief is that all gasifier products may not be equally amenable to hot desulfurization because of gas composition effects on desulfurization reaction equilibria. Therefore, the assumption made in this study that the Morgantown process can be applied to other gasifier products is subject to some uncertainty.

Moving bed gasifiers produce tars which must be prevented from condensing and plugging the hot absorbent bed. The necessity for providing gasifier product gas heating or other measures to avoid tar condensation in hot absorbent beds has not been proved, but neither has the fact that the phenomenon is not a problem. The BGC slagging gasifier which operates at the lowest gasifier exit temperature of the group studied may be expected to present the greatest problem in tar condensation prevention.

A problem associated with application of the hot iron oxide process to fuel gas desulfurization results from the fact that neither the ammonia produced in the gasifier nor alkali metal and other vapors will be removed from the fuel gas by the solid absorbent. Ammonia in the fuel gas will be burned to NO_x in the gas turbine combustor, contributing greatly to the emission control problem. Robson (21) has indicated that NO_x emissions from a moving bed gasification-combined cycle system with high temperature desulfurization could be 20 times greater than the emission from the same system with low temperature desulfurization. Alkali metal vapors passing through the cleaning system will be harmful to gas turbine machinery.

Morgantown is developing the iron oxide process on the basis that dust will be prevented from plugging the absorbent beds by cyclone dust precipitators* upstream of the beds and that dust trapped in the beds is burnt out during regeneration. It remains to be demonstrated on a large scale that this is a practical commercial design basis.

Dust removal methods other than cyclones for cleaning fuel gas under high pressure hot conditions (300 psig, 1,000°F) are not commercially available. Development is required in this area before combined cycle plants supported by hot purification can become a reality.

Methods for removing dust from high pressure hot sulfur containing fuel gas considered in this study are cyclones, panel bed filters, and filters based on the use of a metallic cloth. Only cyclones are commercially available for the service, but cyclones may not have the capacity to prevent plugging of the absorbent bed with gasifier dust. Cyclones certainly do not have the ability to prevent submicron dust from entering the gas turbine. If submicron dust becomes a problem, some form of stack gas scrubbing or filtration could be required in combined cycle units.

Commercialization of hot fuel gas purification cannot occur without supporting dust removal equipment development. Dust removal equipment development should be accelerated in parallel with any larger scale hot purification development.

Thermal Efficiency Evaluation

With the proviso that technical problems are to be resolved before hot purification can be commercialized and assuming that hot purification is equally applicable to all of the gasifier types considered, Table 1-1 summarizes the thermal efficiencies and heat rates predicted by this study.

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*Since writing, cyclones have been reported by MERC to be unsatisfactory for absorbent bed protection.

Table 1-1

SUMMARY OF ESTIMATED THERMAL EFFICIENCIES

PURIFICATION	COLD BENFIELD	HOT MORGANTOWN	COLD BENFIELD	HOT MORGANTOWN
Gas Turbine Inlet Temperature, F	1,950	1,950	2,400	2,400
	Thermal Efficiency, %* Heat Rate, Btu/KwH			
Lurgi (O_2)				
Thermal Eff.	29.4	35.4	32.4	39.9
Heat Rate	11,628	9,630	10,544	8,558
Lurgi (Air)				
Thermal Eff.	31.0	37.0	34.5	41.2
Heat Rate	10,994	9,223	9,907	8,285
Slagging (O_2)				
Thermal Eff.	36.5	37.5	39.6	40.6
Heat Rate	9,352	9,095	8,624	8,409
Entrained Bed (Air)**				
Thermal Eff.	38.0	38.4	40.8	41.6
Heat Rate	8,982	8,879	8,359	8,215
Entrained Bed (O_2)**				
Thermal Eff.	35.4	36.6	37.8	39.3
Heat Rate	9,641	9,334	9,028	8,688

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* Thermal Efficiency (%) $= \frac{\text{(Delivered Kw) (3412.75) (100)}}{\text{(Coal Lb/Hr) (Coal HHV Btu/LB)}}$

** Foster Wheeler Gasifier. The notation Air or O_2 indicates the oxidant employed in each type of coal gasifier.

Table 1-1 lists the gasifier types studies and the corresponding estimated combined cycle thermal efficiencies associated with hot and cold purification at gas turbine inlet temperatures of 1,950°F and 2,400°F.

The data of Table 1-1 indicate that the greatest thermal benefits are to be derived from applying hot gas purification to Lurgi gasifiers. It is important to remember that the underlying assumption used for these cases is that tars will pass uncondensed through the hot iron oxide beds directly into the gas turbine combustor. It is interesting to note that, for all cases other than the Lurgi dry ash gasifiers, thermal efficiency advantages associated with hot iron oxide purification appear to be marginal. There are two major reasons why high temperature purification proved to be of such benefit to the Lurgi systems and of no advantage to the other gasification systems studied:

- The Lurgi gasifiers were the only systems in this study considered to have a net production of tars. Tars produced by the BGC slagger were separated from the gas stream prior to desulfurization and were recycled to extinction to the gasifier. Tars were also considered to be absent in the crude gas from the entrained gasifier. The presence of tars in a crude fuel gas has major impact on the comparison between high and low temperature cleaning schemes. For the high temperature case, tars are assumed to pass through the iron oxide system and are converted to electricity at combined cycle efficiency (40-50 percent). With low temperature cleaning systems, tars are scrubbed from the crude gas by direct quench and are eventually converted to electricity at only the steam cycle efficiency (30-40 percent).
- The Lurgi gasifiers consume large quantities of steam to prevent ash matter from clinkering in the bottom. Most of this steam passes through the gasifier unconverted and is condensed in the gas quench operation necessary for low temperature gas cleaning. Therefore, if a quench is necessary, most of the sensible heat in the steam is unavailable for power generation. If high temperature desulfurization is employed, the steam passes through the iron oxide beds, and its sensible heat is converted into electricity in the combined cycle plant.
- The BGC slagger requires only 13 percent of the steam required by the dry ash oxygen blown Lurgi gasifier due to the higher bottom temperature required for ash slagging. It also converts approximately 90 percent of the gasifier steam to hydrogen and carbon monoxide. Therefore, steam losses due to cooling of the slagging gasifier effluent in the quench operation are negligibly small. Steam consumed by both the air and oxygen blown entrained gasifier is approximately the same as that consumed by the slagger resulting in the same negligibly small steam losses on gas cooling.

An important reason why hot fuel gas purification shows smaller thermal efficiency advantage than might be expected on superficial analysis is the thermal cost of hot sulfur absorbent regeneration. This amounts to a loss of approximately 1.5 net thermal efficiency percentage points for the coal studied. This loss represents the process gas consumption required for reduction of the sulfur dioxide produced during regeneration, after adjustment for waste heat recovery. The major uncertainty in the thermal efficiency determinations is the thermal cost of sulfur absorbent regeneration. Difficulties in this area relate to the production of a sulfur dioxide stream of satisfactory composition for treatment at a steady controlled rate.

Economic Evaluation

The incentive for pursuing thermal efficiency increases is to obtain attractive returns on the capital required to achieve these increases. A broad and approximate survey was conducted comparing gasification-combined cycle economics in which hot and cold purifications are practiced.

Order of magnitude capital requirements and costs of electricity estimates were prepared for all of the cases, using a consistent set of economic criteria. These criteria, detailed elsewhere in this report, are summarized as follows:

- Mid-1975 dollars with no escalation
- 36-month construction period
- 10 percent construction loan interest, compounded quarterly
- Coal cost of $1.00/MMBtu
- 70 percent operating load factor
- 25-yr plant life
- 50:50 debt: equity ratio
- Eight percent bond interest, compounded semiannually
- 12 percent return on equity after taxes, compounded semiannually

Total capital requirements for each system were determined by adding capital related charges such as state sales taxes, preproduction costs, paid-up royalties, initial chemical and catalyst costs, construction loan interest and working capital to the estimated plant investments. Plant investments included a 15 percent contingency which was an allowance to account for the undeveloped state-of-the-art.

The costs of services were calculated by adding capital related charges such as depreciation, bond interest, return on equity and income taxes to the operating charges.

Table 1-2 summarizes the order of magnitude capital requirement estimates for the gasification-combined cycle systems studied.

The data presented in Table 1-2 demonstrate the major conclusions of this study, i.e., dry ash Lurgi gasification would benefit greatly from the development of a hot gas cleaning system capable of passing all normally liquid hydrocarbons and gasifier steam through to the gas turbine. The reductions in capital requirements associated with the application of hot gas purification to all four of the Lurgi systems considered are approximately 35 percent, as explained in a following paragraph. For all of the other gasification technologies investigated, Table 1-2 indicates that there is no capital reduction incentive to be derived from the application of the iron oxide high temperature sulfur removal system.

Table 1-2

SUMMARY OF ESTIMATED CAPITAL REQUIREMENTS FOR 1,000 MW
GASIFICATION - COMBINED CYCLE POWER PLANTS

PURIFICATION	COLD BENFIELD	HOT MORGANTOWN	COLD BENFIELD	HOT MORGANTOWN
Gas Turbine Inlet Temp., F	1,950	1,950	2,400	2,400
	Capital Requirements $/kW*			
Lurgi (O_2 Dry Ash)	1,117	739	1,046	703
Lurgi (Air Dry Ash) Total Capital	1,000	667	936	642
Slagging (O_2) Total Capital	643	629	629	606
Entrained Bed (Air)** Total Capital	619	616	604	597
Entrained Bed (O_2)** Total Capital	670	679	658	657

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*All capital estimates are based on mid-1975 dollars with no escalation. Capital Requirements ($/kW) correspond to Plant Investment with adjustment for Illinois Sales Tax, Preproduction Costs, Royalty Payments, Initial Catalyst and Chemicals, Construction Loan Interest and Working Capital - summarized from Tables 5-1A and B. The Capital Requirements tabulated above represent an increase of 34 percent over the Plant Investment estimates due to inclusion of these allowances.

**Foster Wheeler gasifier.

Table 1-3 presents costs of electricity estimates for all of the systems considered. Examination of the data presented in Table 1-3 indicates the same conclusions as discussed above for the capital requirement data of Table 1-2. Once again, the only major incentive for the development of iron oxide high temperature sulfur removal scheme results from application to dry ash Lurgi gasification systems.

Table 1-3

SUMMARY OF ESTIMATED COSTS OF ELECTRICITY FOR GASIFICATION-COMBINED CYCLE POWER PLANTS OF 1,000 MW CAPACITY

PURIFICATION	COLD BENFIELD	HOT MORGANTOWN	COLD BENFIELD	HOT MORGANTOWN
Gas Turbine Inlet Temperature, F	1,950	1,950	2,400	2,400
	Power Costs, Mills/kWh*			
Lurgi (O_2)	48.3 (60.2)**	33.8 (43.7)	44.9 (55.7)	31.6 (40.4)
Lurgi (Air)	43.8 (55.0)	31.1 (40.5)	40.6 (50.7)	29.3 (37.7)
Slagger (O_2)	30.4 (39.9)	29.7 (39.0)	29.2 (38.0)	28.2 (36.8)
Entrained*** (Air)	29.2 (38.4)	29.0 (38.1)	28.1 (36.7)	27.7 (36.1)
Entrained*** (O_2)	31.5 (41.4)	31.5 (41.1)	30.5 (39.8)	30.2 (39.1)

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* Based on: Delivered coal costs of $1.00/MMBtu. Mid-1975 dollars with no escalation included and an operating load factor of 70 percent.

** Numbers in parentheses represent Power Costs related to a delivered coal cost of $2.00/MMBtu.

***Foster Wheeler gasifier.

Table 1-4

SUMMARY OF ESTIMATED CAPITAL REQUIREMENTS FOR THE PURIFICATION SYSTEM COMPONENTS OF EACH CONFIGURATION INVESTIGATED

PURIFICATION	COLD BENFIELD**	HOT MORGANTOWN***	COLD BENFIELD**	HOT MORGANTOWN***
Gas Turbine Inlet Temperature, F	1,950	1,950	2,400	2,400
	Capital Requirements $/kW*			
Lurgi (O_2)	141	115	129	103
Lurgi (Air)	126	121	114	109
Slagging (O_2)	52	87	48	80
Entrained (Air)	102	106	95	99
Entrained (O_2)	86	99	80	93

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* Based on mid-1975 dollars with no escalation included.

** Cold purification costs include the costs of the H_2S scrubbing system, the Claus plant, the tail gas treating system as well as crude gas cooling equipment.

***Hot purification costs include the costs of the iron oxide system, the SO_2 recycle and sulfur recovery systems, and the high temperature particulate removal devices upstream and downsteam of the iron oxide system.

Careful examination of the detailed cost breakdowns presented in the body of this report for each system studied indicates the major reasons for the results obtained. For the Lurgi dry ash cases, the cost reductions associated with the application of hot fuel gas purification are, to a great extent, the result of the elimination of equipment necessary in the low temperature purification case. When low temperature gas cleaning is applied to Lurgi gasification, a separate tar fired boiler with stack gas scrubber and a large water treatment facility to treat the condensed gas liquor are required. For the high temperature Lurgi purification cases, it has been assumed that both the tars, the steam, and other condensable components in the crude gas pass as vapors through the cleaning system directly into the gas turbine. This would eliminate the necessity for the tar boiler and water treatment facilities resulting in a capital requirement reduction of approximately 20 percent. Actual differences in plant investment costs for the high and low temperature cleaning systems are negligibly small as shown in Table 1-4. The remainder of the economic differences between Lurgi systems associated with either high or low temperature purification schemes can be attributed directly to the large heat rate advantages (approximately 17 percent improvement) associated with systems employing high temperature gas purification. These substantial heat rate improvements result from the assumption that all condensable hydrocarbons pass through the hot absorbent beds and are converted to electricity at combined cycle efficiency. Also, the large quantity of water vapor in the crude gas is not condensed in the hot purification system and, therefore, adds to the mass of high temperature working fluid passing through the gas turbine machinery.

The equipment eliminations plus the major heat rate improvements resulting from the application of high temperature gas cleaning technology to dry ash Lurgi systems do not occur for any of the other gasification technologies investigated. This is due to the fact that no net production of tars or large quantities of steam in the crude gas have been assumed for these other gasification devices. The BGC slagger does produce tars, oils, and phenols. It has been assumed, however, that all of the hydrocarbon by-products can be recycled to extinction in the gasifier. This implies that the hydrocarbon by-products eventually enter the gas turbine as fuel in both the high and low temperature cleaning configurations. Also, as only 10 percent of the steam introduced into the slagging gasifier is unconverted, the losses due to quenching the crude gas in the low temperature cleaning case are small. For these reasons, the screening economic projections developed do not indicate any incentive for the development of iron oxide hot purification technology for the BGC slagging gasifier.

Similar arguments to those presented for the slagger can be made to explain the absence of incentive to develop iron oxide purification technology for entrained flow gasifiers. It has been assumed that tars will not be produced in entrained flow devices and that their consumption of steam is similar to that of the slagger and is much lower than the steam consumption of a dry ash Lurgi.

In summary, it can be stated that this study indicates that economic incentives exist for the development of high temperature iron oxide purification technology only for those gasifiers having a net production of tars or large quantities of water vapor in the crude gas product. This conclusion should be treated with caution for the following reasons. If, for example, techniques already under development for recycling tars to a Lurgi gasifier were to be demonstrated for a variety of U.S. coals or if the steam requirement of a Lurgi gasifier could be substantially reduced, the incentives for the development of high temperature iron oxide purification technology to be used in conjunction with Lurgi gasification would be greatly reduced and probably disappear completely.

A surprising result of this study is the relatively small incentive predicted for the development of the 2,400°F gas turbine shown in Table 1-3. This table indicates that, for all cases except the dry ash Lurgi, the economic benefit of changing turbine inlet temperature from 1,950°F to 2,400°F is in the range of 1-2 mills/kWh irrespective of the purification temperature employed. For the Lurgi cases, this incentive appears to range between 2 and 3.5 mills/kWh for hot and cold purification cases, respectively.

This result must be treated with caution.

It should not be used as a basis for making technology development decisions without further investigation for the following reasons:

- The major purpose of this study was to investigate the incentives for developing high temperature gas purification systems. Therefore, the bulk of the engineering effort was concentrated on this section of the plant.
- Detailed steam system design optimization for each case was not carried out. Although uniform stack gas temperatures and a reasonable pinch point approach to each case was established, similar steam cycles were employed for both the 1,950 and 2,400°F cases.
- It was also necessary to estimate costs for gas turbine technology not yet available. For this preliminary study, a nominal value of $200/kW was made for the combined cycle portion of the plant, independent of gas turbine inlet temperature.

- The use of such methods was consistent with the primary objective of this screening study. However, they could have indicated a lower incentive for the development of a 2,400°F turbine than really exists.

- It is therefore strongly recommended that further detailed studies be conducted to provide a basis for establishing the benefits to be associated with the development of high temperature gas turbines.

Section 2

PURIFICATION OF FUEL GAS AT HIGH TEMPERATURES

ERDA HOT SULFUR ABSORPTION PROCESS (THE MORGANTOWN PROCESS)

The Morgantown Energy Research Center (MERC) of the United States Energy Research and Development Administration (ERDA) is developing regenerable solid absorbents for removal of sulfur compounds from fuel gas at temperatures above 1,000°F. Key Morgantown observations in the laboratory scale treating of hot fuel gases are:

- Maintaining absorption temperatures above 1,000°F avoids plugging of fixed beds of absorbent by tar condensation and decomposition.
- Dust trapped in the fixed absorbent bed during sulfur absorption is burned out on regeneration. As a result, pressure drop through the absorbent bed is restored to its initial design value when the subsequent absorption cycle is commenced.
- Sulfur accumulations at the commercially useful level of about 10 weight percent are possible in the supported iron oxide absorbents under study.
- Regeneration of the absorbents by direct oxidation with an oxygen containing gas is readily accomplished.
- Absorbents can be developed which are capable of sustaining a large number of absorption/regeneration cycles without suffering loss of mechanical or chemical properties.

The sulfur absorbent used in Morgantown work is iron oxide which has been evaluated with a range of different support materials in a program to develop a mechanically stable and chemically active product which is suitable for commercial use. Absorbent development has involved:

- Experimental screening of absorbents with the selection of fly ash supported iron oxide and silica supported iron oxide as being the most suitable for commercial use in fixed bed application.
- Study of effect of changing temperatures on sulfur absorption capacity.
- Evaluation of the effect of changing fuel gas steam concentration on absorbent performance.
- Study of absorbent aging as it undergoes multiple absorption/regeneration cycles.

- Absorption of sulfur compounds from simulated and commercial type fuel gases. Commercial type fuel gases were obtained from the stirred, fixed bed, air blown gasifier installed at the Morgantown facility.
- Evaluation of differing absorbent production techniques on absorbent performance.
- Absorbent regeneration studies employing a range of oxygen nitrogen mixtures (from 100 percent oxygen to 10 percent oxygen).
- Absorbent regeneration based on reacting sulfur dioxide with the sulfided absorbent to reconstitute the iron oxide with the direct production of elemental sulfur.

Some useful papers by the Morgantown Energy Research Center are listed under Reference 18 and summarized in Appendix A.

Current Morgantown Development Activity

The Morgantown program is approaching hot purification and moving bed coal gasification as an integrated development, rather than developing hot purification in isolation for later integration with gasification. Development is in an early stage but sufficient information is published on sulfur absorption performance to permit an order of magnitude evaluation of its commercial value.

Appendix A is a summary of Morgantown papers recording progress to date in process development.

An integrated moving bed gasification and fixed bed hot fuel gas desulfurization pilot unit is undergoing initial operations. Testing of the system is in progress gasifying anthracite which is low in volatile matter and which will not cause excessive carbon laydown in the event of operating problems. The capacity of the unit is for gasification of 200 lb/hr coal producing 12,000 SCFH of low Btu gas. Fuel gas leaving the gasifier is subject to cyclone dust removal prior to entering the fixed bed sulfur absorption vessels. The system operates at up to 15 psig and will yield data which, though useful, are not directly applicable to combined cycle power generation where approximately 300 psig gasifier pressure is required. The schedule calls for completion of first phase test work late in 1976.

Air Products and Chemicals Inc. (APCI) is carrying out contracted practical and theoretical evaluations of iron oxide absorbents as an extension of former work with limestone as the hot sulfur absorbent. APCI objectives are to provide absorbent development support, definition of process scale-up criteria, and commercial process design. APCI has prepared a mathematical model to simulate the dynamics of absorption and regeneration which has been reported in published project documentation (15).

Replacement of the fly ash support by silica in the iron oxide absorbent is undergoing laboratory investigation. The silica supported absorbent is the more rugged, attrition resistant and temperature resistant material and it may be suitable for use in moving bed or fluidized bed process configurations which could be the topic of future programs.

In regeneration, processes aimed at the production of elemental sulfur by reaction of the sulfided absorbent with sulfur dioxide are being investigated. The reaction involved is approximately equivalent to

$$4FeS_{1.5} + 3SO_2 = 2Fe_2O_3 + 9S$$

Where nitrogen enriched air is employed for regeneration in order to avoid the possibility of fusing the absorbent bed, the regeneration product tends to be low in sulfur dioxide concentration and to resemble stack gas in its treating problems. The sulfur dioxide iron sulfide reaction offers a potential method of avoiding such problems.

Process Chemistry

In applications of hot fuel gas purification, where the involved fuel gas is of appreciably different composition from the gas studied by Morgantown, the possibility exists that absorbent performance may be different from that reported by Morgantown. In addition, Morgantown work has been on a small scale and in large scale applications the possibility of behavior which was not recognized in small scale work must be considered. A review of the chemistry provides indicators of possible importance.

Under high temperature conditions in an oxidizing atmosphere, ferric oxide (Fe_2O_3) is the stable form of iron oxide which is probably the principal form present in the freshly regenerated absorbent bed. Employed for desulfurization in a reducing atmosphere, ferric oxide undergoes the reaction:

$$Fe_2O_3 + 2H_2S + H_2 = 2FeS + 3H_2O \qquad \text{(A)}$$

This reaction is cause for concern since it suggests hydrogen consumption. However, Morgantown report that the empirical sulfided bed composition is $FeS_{1.3}$, suggesting that the sulfided bed contains a mixture of FeS and FeS_2, the latter being formed by the following reaction which indicates that some of the hydrogen consumed in Reaction A is again liberated into the fuel stream:

$$FeS + H_2S = FeS_2 + H_2 \qquad (B)$$

There are two concerns in connection with the applicability of the above chemistry. It is necessary to determine whether gases containing a high proportion of steam are less amenable to hot purification than the lower steam containing gases studied by Morgantown, because of the reversing effect on the chemical equilibrium (Reaction A). In cases where the gas contains a higher proportion of the reducing gases carbon monoxide and hydrogen than the gas used at Morgantown, the effect on Reaction B must be known in order to determine if Reaction B is appreciably reversed. Such reversal effects may imply significant reduction in sulfur absorption capacity and consequent net fuel loss.

Note that carbon monoxide may react similarly to hydrogen in Reaction A with a similar fuel consuming effect.

$$Fe_2O_3 + 2H_2S + CO = 2FeS + 2H_2O + CO_2 \qquad (C)$$

Morgantown work has indicated between about 6 percent volume and 20 percent by volume steam content in the fuel gas has little influence on absorbent performance at atmospheric total pressure. However, there does not appear to be any information available based on published experimental data which describes the effect of reducing gases concentration on Reaction B.

Ferric oxide is converted to ferrous oxide under reducing conditions, and potential problems exist because of the reactions.

$$Fe_2O_3 + H_2 = 2FeO + H_2O \qquad (D)$$

$$Fe_2O_3 + CO = 2FeO + CO_2 \qquad (E)$$

The concern here is that a massive ferric to ferrous oxide reduction at the start of absorption may cause a variation in fuel gas production (5) and quality which impairs combined cycle power unit performance. Reactions D and E are probably the initial part in the mechanism of Reactions A and C, ferrous oxide being subsequently converted to the sulfide by reaction with the hydrogen sulfide.

Utlimately, convincing demonstration of general applicability of the Morgantown process must be based on systematic experimental work. Equilibria studies are encouraging and are a useful guide but they are not certain indicators of process viability.

Referring to Table 2-1 and using the nomenclature indicated in Section 1:

- Case MXH. This gas contains a lower concentration of reducing components than the MERC gas (RH column) and a considerably higher concentration of steam. Conclusion: any projections of associated hot purification application are speculative because of possible steam suppression of absorbent performance.
- Case MAH. Lower than MXH (above) in steam but still above the MERC level. MERC process is probably satisfactory for Case MA gas but it should not be applied to the gas of Case MA without a test.
- Case MSH. Low steam and high reducing gases concentration may inhibit sulfur absorbent performance by chemical equilibrium effects.
- Case EXH. On steam content comparison with MERC, this gas would be satisfactory. However, this gas contains a high concentration of reducing gases which may impair absorbent performance, as for MSH.
- Case EAH. This gas is similar to the MERC gas and may be satisfactory for hot purification application.

The gases containing a high proportion of CO_2 relative to MERC may tend to reverse Reaction C and limit sulfur absorption by this mechanism.

The general conclusion is that process development should cover a sensitivity study on the interaction of gas composition and absorbent performance for the full range of gas compositions expected from each gasifier to which application is proposed.

Ammonia with iron oxide chemical reaction is not known to occur, but some adjustment of ammonia concentration may occur by equilibrium decomposition effects, depending on the composition of the fuel gas. The feasibility of such effects may be investigated in any case using published equilibrium data (25). For carbonyl sulfide removal, iron oxide based hot desulfurization appears to be satisfactory (21).

Table 2-1

COMPOSITION OF GASES ENTERING HOT DESULFURIZATION VESSELS

CASE*	MXH	MAH	MSH	EXH	EAH	MERC
Component	Vol%					
CH_4	4.28	2.84	7.89	5.98	3.35	-
C_2H_6	0.31	0.07	0.58	-	-	-
H_2	20.86	17.12	26.94	30.08	13.90	13.9
CO	7.47	11.91	54.36	34.79	28.17	21.3
CO_2	15.24	9.68	1.81	12.62	3.40	4.1
H_2O	50.33	27.66	4.94	14.39	2.23	18.0
H_2S	0.59	0.52	1.38	0.97	0.65	1.2
COS	0.03	0.02	0.06	0.10	0.07	-
N_2	0.18	29.55	0.38	0.44	47.81	41.5
NH_3	0.40	0.36	0.94	0.63	0.42	-
Tar(MW125)	0.31	0.27	0.72	-	-	-
Total	100.00	100.00	100.00	100.00	100.00	100.00

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*Nomenclature as indicated in Section 1.

The general chemistry of hot desulfurization is summarized in Table 2-2. MERC reports that absorbent testing was accompanied by a water gas shift effect to near equilibrium in the gas leaving the absorber.

Morgantown Process Design Discussions

Operating Temperature and Tar Formation. A review of Morgantown work suggests that inlet temperature to a bed of sulfur absorbent be set at 1,200°F with normal operation being in the range 1,000°F to 1,400°F. At 1,200°F inlet temperature, margin exists to accommodate upward swings in gasifier exit temperature and up to about 150°F temperature rise by exothermic heat of reaction effects without causing fusion of the absorbent bed. The lower limit on absorption operating temperature is 1,000°F, below which tar condensation and consequent absorbent deactivation are a possibility in the moving bed gasifier cases.

In the entrained bed cases, the gasifier product temperature is in the range 1,700°F to 1,900°F and no tar should be produced. The entrained bed gasifier, therefore, produces a gas which after cooling is better suited for application of hot purification than the other gasifiers included in the studies, assuming that there are no other chemical problems associated with it.

The requirement to avoid tar condensation in cases where tar is produced indicates that the gas exit temperature should be above the dew point of contained tars. The Morgantown ERDA work recognizes that gasification and hot purification are closely coupled by such connections and is developing an integrated gasification-purification process system. The Morgantown result will be a process combining air blown dry ash gasification with hot purification and employing an absorbent developed for this application.

Success of the Morgantown project would make it reasonable to assume that hot purification can be connected to a Lurgi air blown gasifier operating at a similar exit temperature to that in the Morgantown work, provided that the steam level in the gas is about the same as in the Morgantown case. It would also be reasonable to anticipate that an entrained bed gasifier product after cooling would be satisfactory for hot purification, provided that the particulate content and particulate character in the gas can be established at levels similar to those encountered in Morgantown work. The similar anticipation for the oxygen blown dry ash producing Lurgi gasifier would also be reasonable since it can operate at similar exist temperatures to the air blown case. However, the possibility of chemical problems (see page 2-3) must be kept in mind in the oxygen blown Lurgi and entrained bed gasifier cases.

Morgantown work may conclude that some means of gasifier exit temperature adjustment is required in the case of the type of gasifier used in that work. Moving bed gasifiers can experience swings in gasifier exit temperature. It may not always be practical to adjust gasifier steam and oxidant supply to simultaneously satisfy ash quality condition and exit temperature constraints. For this reason, in the process flow diagrams for moving bed gasifier cases, a temperature controlling action poses a development question to be answered by future programs; all that is possible here is to call attention to the potential problems.

Table 2-2

HOT FUEL GAS DESULFURIZATION WITH IRON OXIDE

General Chemistry - Summary

Sulfur Absorption Processes

$Fe_2O_3 + 2H_2S + CO = 2FeS + 2H_2O + CO_2$

$Fe_2O_3 + 2H_2S + H_2 = 2FeS + 3H_2O$

$FeS + H_2S = FeS_2 + H_2$

Average Process Representation

$Fe_2O_3 + 3H_2S = 2FeS_{1.5} + 3H_2O$

Average Process Representation

$Fe_2O_3 + 3H_2S + CO = 2FeS_{1.5} + 2H_2O + CO_2 + H_2$

Absorbent Regeneration Processes

$2FeS + 3O_2 = 2FeO + 2SO_2$

$2FeS_2 + 5O_2 = 2FeO + 4SO_2$

Average Regeneration Representation - FeO Production

$2FeS_{1.5} + 4O_2 = 2FeO + 3SO_2$

Average Regeneration Representation - Fe_2O_3 Production

$4FeS_{1.5} + 9O_2 = 2Fe_2O_3 + 6SO_2$

Ferrous Oxide Oxidation

$4FeO + O_2 = 2Fe_2O_3$

Direct Elemental Sulfur Production

$2FeS_{1.5} + SO_2 = 2FeO + 4S$

The preceding discussion of practical matters suggests that a moving bed coal gasifier supporting hot desulfurization be operated at a higher gas production temperature than need be employed where cold purification is practiced. An ideal comparative evaluation would be one in which each gasifier is operated so that no extraordinary process measures, such as gasifier exit temperature adjustment, are required. This would probably impose additional penalties on hot purification. For this study, the same gasifier operating bases are employed for hot and cold case thermal analyses, since insufficient gasification data were available to permit any broader approach to evaluation. However, a thermal efficiency penalty is introduced in moving bed gasification schemes employing hot purification by using a heater to achieve exit gas temperature control.

Several means of achieving exit temperature control on moving bed gasifiers were considered in addition to the heater which has little to commend it from a practical point of view. It was proposed to burn part of the hot producer gas stream under pressure and to inject the hot products of combustion into the gasifier effluent to achieve the necessary control. Practical problems make this unattractive. It was proposed to meter air or oxygen into the gasifier exit line in the hope that the exothermic combustion process in the line would boost the temperature by the required amount. This produced a divergence of expert opinion and cannot be recommended though some precedent exists for it in the Appleby-Frodingham process experience (Appendix B).

The unsolved problems connected with gasifier exit temperature control introduce uncertainties to the conclusions of this study since the thermal and capital cost of eliminating problems can be only guessed before the problems are actually solved.

It is axiomatic that with cold purification a gasifier should be operated at maximum cold gas efficiency to achieve maximum overall combined cycle efficiency. With hot purification this is not so. The gasifier should be operated in a manner consistent with avoiding practical hot purification problems because of tars and with obtaining good desulfurization performance. This may impose an additional penalty on hot purification as a result of operating the gasifier less efficiently. There is also a possibility that the action may be beneficial if temperature elevation is achieved by reducing steam consumption. The state of the published arts does not permit full analysis of the interaction of gasifier operation with purification operation and combined cycle efficiency. The uncertainty with respect to integrating gasification with purification would have been reduced had gasifier models been available to assist the study. The chemical uncertainties in using hot purification with gasifiers other than the type studied by Morgantown were pointed out previously (see page 2-3).

The introduction of silica supported absorbents may lead to the establishment of higher permissible normal operating temperatures than the 1,200°F judged to be practical for this work.

Sulfur Absorption and Distribution

Sulfur Absorption. In establishing the design value for sulfur pick-up by the Morgantown supported iron oxide, factors to be taken into account are:

- Expected sulfur pick-up by fresh absorbent from dry simulated fuel gas
- Depression of sulfur pick-up by reversing effect of steam, if any
- Depression of sulfur pick-up by aging of absorbent
- Reduction in effective sulfur pick-up because of the impossibility of accomplishing complete regeneration of the sulfur absorbing iron oxide
- Chemical equilibrium constraints, if any

Hot desulfurization technology has not been advanced to where the sum of the factors affecting sulfur pick-up can be predicted satisfactorily for commercial design. For the purposes of current studies, sulfur pick-up levels in the range of 5 to 10 weight percent were anticipated.

Based on early laboratory work reported in Summaries A-1 and A-2 of Appendix A, it is indicated that normal fuel gases containing between 10 and 20 volume percent steam will experience low sulfur pick-up (4 weight percent or less) because of the effect of steam. In that laboratory work a steady falloff in performance was observed as steam concentration increased to 10 volume percent. Hence, the fall-off is predicted to be larger for higher steam concentrations (partial pressures).

In subsequent short term tests with a commercial type of producer gas (reported Appendix A, Summaries A-3 and A-4), better performance was observed with smaller steam effect on absorption capacity (see also page 2-3).

Based on short term testing with a commercial type of producer gas, an optimistic sulfur pick-up may be anticipated with the assumptions that steam has no significant effect and that an absorbent is developed in which the aging effect is not significant.

With the above rationale the range of sulfur pick-up expected is in the range of 5 to 10 weight percent which determines the investment requirement range for the hot desulfurization vessels.

Sulfur Distribution. The distribution of sulfur to various process destinations corresponding to sulfur pick-up in the 5 to 10 weight percent range is as follows:

- 5.0 percent of sulfur in coal appears in slag or ash.
- 90.0 percent of sulfur in gasifier effluent is absorbed by iron oxide.
- 90.0 percent of sulfur desorbed from sulfur absorbent converted to elemental sulfur.

With this distribution, 18 percent of the sulfur in the coal feed appears in the stack gases. If the slag or ash is sulfur free the stack receives 19 percent of the sulfur in the coal feed.

This distribution established the sulfur dioxide emission based on Illinois No. 6 coal containing 3.7 weight percent sulfur, having a heating value of 12,235 Btu/lb (HHV), at 1.09 lb SO_2/MM Btu with 5 percent of the sulfur in the ash. With zero sulfur in the slag or ash, the sulfur dioxide emission increases to 1.15 lb/MM Btu.

Corresponding to the above factors, with 5 percent of the sulfur diverted to the slag or ash, the sulfur load on the iron oxide beds at coal consumption rate of 850,000 lb/hr (1,000 Mw) is approximately 27,000 lb/hr.

By operating at a lower sulfur pick-up than the 5 to 10 weight percent range, it should be possible to absorb more than 90.0 percent of the sulfur in the gasifier effluent material and to improve the gas turbine feed quality. The upper limit on sulfur removal efficiency may be a function of gasifier type and gas composition.

Conversion of the sulfur dioxide desorbed from the iron oxide absorbent during regeneration to elemental sulfur appears to be limited to about 90 percent with the balance entering the tail gas incineration system. To accomplish smaller sulfur emissions from the system, some form of regeneration vent gas scrubbing would be required. One of the stack gas sulfur dioxide scrubbing processes would probably be suitable for this purpose.

By a combination of accepting reduced sulfur pick-up and adding regeneration vent gas scrubbing facilities (both increasing capital cost and process complexity) the hot iron oxide process can be improved in performance. The combination, 90 percent sulfur absorption from the fuel gas on iron oxide and 90 percent reduction of sulfur dioxide produced in regeneration to elemental sulfur, is a judgment selection. To obtain higher sulfur absorption from fuel gas would lead to reduced sulfur pick-up by the absorbent and increased capital costs. An optimization study to determine the best combination of fuel gas sulfur removal and regeneration vent cleanup facilities in order to satisfy present and projected environmental emissions regulations would be useful. However, sulfur absorbent performance data are as yet not complete enough to justify such a study.

Materials of Construction - Hot Sulfur Absorption

In a process scheme of the type involved in hot fuel gas purification, equipment downstream of the gasifier encounters gases containing hydrogen, hydrogen sulfide, ammonia, carbon dioxide, sulfur dioxide during regeneration of the sulfur absorption vessels, water vapor and particulate matter. The material of construction must be able to accommodate process streams of the above type to about 1,500°F and 300 psig.

High Temperature Range. The materials used must have adequate strength at elevated temperature, must resist hydrogen embrittlement and must resist corrosion by hydrogen sulfide.

Hydrogen sulfide at elevated temperatures is corrosive to carbon and low alloy steels. Steels containing 9 percent and more chromium are somewhat more resistant than carbon and low alloy steels. 18/8 Cr/Ni austenitic stainless steels offer more resistance to hydrogen sulfide corrosion but these materials are rapidly attacked under severe conditions of concentration and temperature. Aluminum coatings offer good resistance to hydrogen sulfide corrosion. Actual commercial experience with materials in high temperature hydrogen sulfide environments is limited to 1,000°F.

A laboratory and pilot plant study is under way by the Metals Properties Council (MPC) in collaboration with American Gas Association (AGA) and ERDA for the purpose of evaluating metals and refractories for pilot and commercial gasification plants.

The study by MPC covers testing at three hydrogen sulfide levels (0.1, 0.5 and 1.0 mol percent) at temperatures of 900°F, 1,500°F, 1,800°F and 1,000 psig pressure.

Low Temperature Range. Some of the gas constituents, primarily H_2S, CO_2, and SO_2, cause severe corrosion of carbon and some alloy steel when dissolved in condensate on a metal surface.

Erosion of exposed metal surfaces in high velocity areas such as cyclones and transfer lines is a possibility at all temperatures.

The MPC studies and some limited pilot experience show that, with 1.0 percent hydrogen sulfide at 1,500°F, corrosion rates are below 0.020 in./yr for 309 s.s., 310 s.s., 310 s.s. aluminized, Incoloy 800, Incoloy 800 aluminized, and a few other alloys. Types 304 and 316 s.s. proved to be unsatisfactory for this environment. Results of studies with 0.1 percent hydrogen sulfide at 1,500°F indicate good performance of 304 s.s. and 310 s.s.

Those of the above noted materials which will provide the necessary resistances may be selected and applied in hot purification with a reasonable corrosion allowance. Aluminum coatings will probably provide protection against corrosion, but limitations may be found in the size of equipment which can be aluminized and fabrication problems will have to be resolved. A suitable refractory liner is required where erosion is a problem.

Another approach to corrosion prevention uses sufficient internal refractory insulation to maintain metal temperature low enough to permit use of less expensive materials; e.g., carbon steel. Approximate corrosion rates for carbon steel exposed to 1.4 mol percent H_2S at 300 psig at

- 500°F is 0.005 in./yr
- 600°F is 0.020 in./yr
- 700°F 0.040 in. +/yr, minimum

The problem with this design is that any spalling off of the refractory could result in hot spots with the possibility of corrosion by hydrogen sulfide or catastrophic failure by hydrogen embrittlement. One solution to this is a water jacket on the outside of the vessel shell to ensure that the shell metal temperature is kept low. Before a jacket is considered, problem areas to be investigated are:

- Temperature of internal surface of metal shell must be above the water dew point of the gases since condensation would result in severe corrosion. In this case a stainless steel internal cladding is required to resist corrosion.
- Corrosion of carbon steel water jacket and external surface of the steel pressure shell by boiling water and oxygen pickup from the atmosphere.
- The location of hot spots where refractory failure takes place is difficult on a jacketed vessel.

In this study the atmospheric pressure jacket was used. Another possibility would be to equalize jacket and process pressures and thereby reduce the thickness of the alloy jacket/vessel interface wall. With this approach the cost of hot purification may be reduced compared to the situation where the alloy vessel wall must be designed for the full design pressure.

Sulfur Absorption Vessel Design

Sulfur absorption vessels have been cost estimated based on employing a single bed of absorbent approximately 15 ft deep, an internally insulated shell and a water jacket.

Because of the large number of these vessels that are required, consideration was given to combining two absorbent beds into one vessel which would require a structural support for the upper bed. This support would consist of beams resting on seats attached to the pressure shell. The seats would be difficult to design to avoid heat flow from the inside of the vessel to the shell. In addition the beam would be unreasonably large because of the low stresses at the internal design temperature. Therefore, single bed sulfur absorption reactors with maximum bed depth set by absorbent strength were employed as the basis for cost estimation of these reactors.

Costs of desulfurization vessels were developed assuming the shell to be internally clad with stainless steel and based on all carbon steel construction. The clad vessel costs approximately 70 percent more than the carbon steel version. The clad item is to be preferred because these vessels are alternately operated and regenerated and there will be times, between the active periods, when a highly corrosive wet environment could develop between the vessel wall and the refractory linings.

Regeneration

Morgantown development indicates that:

- Absorbent destruction by fusion of the bed occurs at approximately 1,500°F and 1,700°F for fly ash and silica supported iron oxides, respectively.
- A continuously varying sulfur dioxide concentration (typically between 1.0 percent and 12.0 percent) is encountered in the regeneration product gases when air or other oxygen nitrogen mixture is employed. Sulfur dioxide reduction to elemental sulfur must be carried out in equipment designed to accomodate this variation, or some method of avoiding the variation problem must be found.
- Oxygen, when passed into a hot (1,000°F) sulfided laboratory scale bed of absorbent, reacts to extinction and the product is essentially only sulfur dioxide during the regeneration period. Only low regeneration rates are permitted when using oxygen because of the possibility of damaging the bed if temperature becomes too high.
- By injecting sulfur dioxide into a hot sulfided bed of absorbent, it may be possible to produce elemental sulfur directly with the elimination of a subsequent sulfur dioxide reduction process. The chemical reaction involved is:

$$4FeS_{1.5} + 3SO_2 = 2Fe_2O_3 + 9S$$

Regeneration development to date is outlined in the referenced summaries of Morgantown work and a number of regeneration process schemes are being evaluated in order to identify the best commercial approach.

The main approach to regeneration in most Morgantown experimental work involves passing oxygen, air, or oxygen nitrogen mixtures into the bed of sulfided absorbent. The ultimate process may well be based on this approach, but it has many disadvantages and few advantages, as follows:

- The use of oxygen or oxygen rich regeneration gas tends to cause excessive temperature rise causing damage to the absorbent bed.
- The use of air or nitrogen rich regeneration gas produces an effluent from the regenerating bed which contains a low (approx. 12.0 vol percent) sulfur dioxide concentration. Further, this sulfur dioxide concentration is variable, typically falling to 1 vol. percent when regeneration is 80 percent complete.
- The low and fluctuating sulfur dioxide level may require a sulfur dioxide concentrating device to produce a satisfactorily stable feed to a unit where the reduction of sulfur dioxide to elemental sulfur is carried out. Studies of sulfur dioxide concentration systems indicate that the capital cost of such facilities would approach the capital cost of a cold desulfurization unit. They would also carry a substantial thermal efficiency penalty and the operating problems associated with varying feed sulfur dioxide

concentration would be significant. Further, a vent gas of low sulfur dioxide content would be produced which is analogous to stack gas and would require additional expensive treatment. Water was employed as the concentrating solvent in a conventional sulfur dioxide absorber stripper combination for order of magnitude studies. It was concluded that a suitable sulfur dioxide concentration unit cannot be designed for the service described based on current data for sulfur absorbent regeneration.

- In another concept of regeneration, a gaseous recycle is established using a recycle compressor. The recycle gas would be an oxygen nitrogen mixture in which sulfur dioxide would be permitted to rise to a level sufficient to eliminate the need for a separate sulfur dioxide concentration unit. The necessity to circulate large gas volumes in order to limit the temperature rise across the regenerating bed leads to excessive compression power consumption. There is scope for optimization in this type of system, and the gas recirculation rate is a function of the temperature gradient permissible across the regenerating bed which has not been established. However, the minimum thermal efficiency penalty associated with recycle gas compression for this type of system is of the order of 1.0 percent, and it could be higher in a conservative design. The thermal efficiency penalty associated with reduction of sulfur dioxide to elemental sulfur would be additional to the above compression penalty.
- All of the processes which are available for scrubbing boiler stack gas and smelting process gases are potentially applicable for treating streams containing from 1 percent to 12 percent sulfur dioxide, as would be produced using air for sulfur absorbent regeneration on a once-through basis. Alternatively, the solvents used in some of the available treating processes may find use in sulfur dioxide concentration as previously described with water as the assumed concentrating solvent. Thus, on the advantage side, there are many processes which, with development and adaptation, may be effective in the treatment of air blown regeneration vent gases.

The conversion of sulfur dioxide in dilute gaseous concentrations to sulfuric acid, as practiced in the Appleby-Frodingham process, is also a possibility. It may be that in certain circumstances it is more economical to transport sulfuric acid away for neutralization, rather than to transport limestone or other material to the power station site.

Of the approaches considered in this study, the most attractive is that based on sulfur dioxide recirculation through the sulfided bed of absorbent which is undergoing regeneration. However, without experimental support, it cannot be claimed that this method is feasible, though support for feasibility is provided by Morgantown work so far completed. Mechanical and materials problems may render the idea impractical.

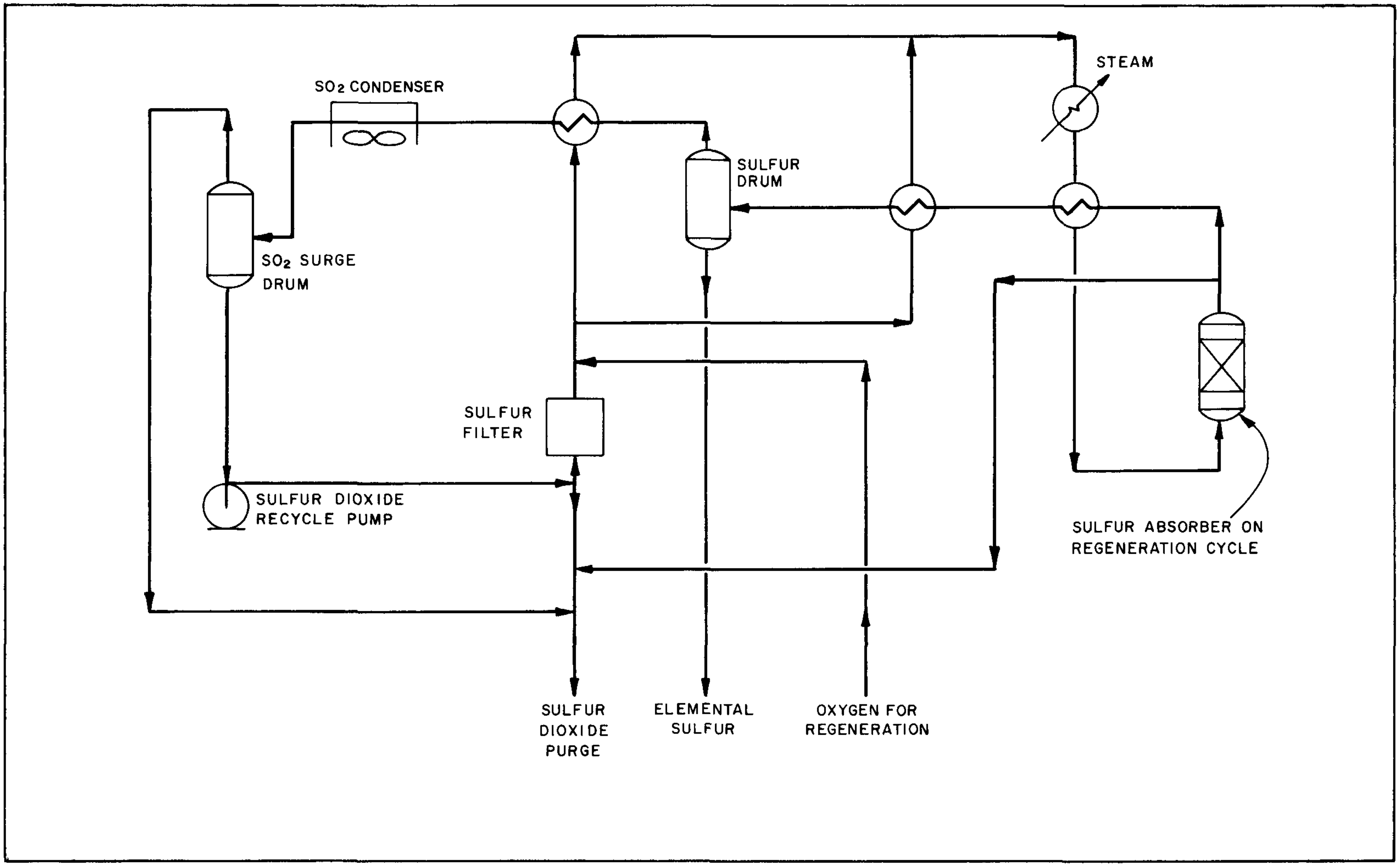

Figure 2-1. Regeneration Scheme Based on SO_2 Recycle

One scheme of the sulfur dioxide recirculation process is shown in Figure 2-1. In this scheme, an oxygen sulfur dioxide mixture at approximately 1,000°F is injected into the sulfided bed of absorbent. Oxygen flow is set by the period of time available to complete regeneration of the absorbent bed. Sulfur dioxide flow is set by the constraint that the regenerating absorbent bed temperature cannot exceed the limit above which danger of absorbent fusion occurs.

Two processes occur in the regenerating bed of Figure 2-1; these are the reaction of sulfided absorbent with oxygen to produce sulfur dioxide and regenerated iron oxide and the reaction of sulfur dioxide with sulfided absorbent to produce elemental sulfur and regenerated iron oxide. The exit stream leaving the regenerating bed is cooled by exchange with the incoming stream and other streams. In the exit stream cooling operation, sulfur is condensed and withdrawn from the process after which the sulfur dioxide is condensed and recycled by pumping back to the regenerating reactor. The process is carried out at approximately 200 psig so that sulfur dioxide can be condensed without refrigeration. The liquefied sulfur dioxide is revaporized on return to the reactor by exchange with the cooling regenerating reactor effluent. Uncondensed oxygen rich gas in the recycle loop can be recycled to the regenerating reactor or recycled to the gasifier or injected to a sulfur dioxide regeneration unit where elemental sulfur is produced.

Purges from the system are elemental sulfur, sulfur dioxide, and a gas purge to remove accumulating inert components. The gas purge is rich in sulfur dioxide and is fed to the sulfur dioxide regeneration unit. If possible, the total sulfur content of the sulfided bed will be removed as elemental sulfur with zero sulfur dioxide purge. If development indicates that 100 percent elemental sulfur production is not feasible, it would be necessary to purge the net sulfur dioxide production and to reduce this to elemental sulfur by other means, which will be discussed under Removal of Particulate Material from Gasifier Effluent beginning on page 2-23.

In a balanced sulfur dioxide recycle system, half of the absorbent would be reduced by oxygen and half by sulfur dioxide via the reactions:

- $4FeS_{1.5} + 6O_2 = 2Fe_2O_3 + 3SO_2$
- $4FeS_{1.5} + 3SO_2 = 2Fe_2O_3 + 9S$

For current estimating purposes, it is felt that the assumption of zero elemental sulfur production should be made, thereby calling for maximum sulfur dioxide reduction capacity in the estimated plant.

The sulfur dioxide recirculation process is the basis for quoted hot purification costs in Section 5. It offers a potential method for direct production of elemental sulfur in the regenerating bed, and failing this, the ability to produce a concentrated sulfur dioxide for feed to another process unit for producing elemental sulfur. It would probably represent a considerable saving if an air based regeneration case can be satisfactorily developed which does not require a concentration unit or a vent gas scrubbing unit.

Sulfur Dioxide Reduction

Iron oxide regeneration may be developed based on either passing an oxygen nitrogen mixture or a sulfur dioxide oxygen mixture through the bed of spent absorbent. In both cases a sulfur dioxide stream is produced for reduction to elemental sulfur or other treatment.

Oxygen Nitrogen Regeneration. In the case of using an oxygen-nitrogen mixture for regeneration, present indications are that a gas which is low in sulfur dioxide concentration (7 percent vol) must be produced if fusion of and damage to the absorbent bed are to be avoided. Further, the sulfur dioxide concentration falls off as regeneration progresses and has been quoted to be 1 percent (vol) when regeneration is 80 percent complete - implying a substantial unconverted oxygen content in the vent gas towards the end of the regeneration period.

The RESOX process of Foster Wheeler Corporation employs low volatility coal or anthracite for the reduction of sulfur dioxide to elemental sulfur. However, the RESOX process at this time is not able to reduce feeds containing below 20 percent vol sulfur dioxide to the 90 percent minimum conversion level believed to be required for hot purification to be acceptable. It is doubtful if a sulfur dioxide concentration unit can be designed to produce a feed satisfactory for the RESOX process because regeneration gas flow tends to be discontinuous as reactors are switched back and forth from operation to regeneration and sulfur dioxide effluent concentration is variable within wide limits, as previously stated.

The introduction of silica supported absorbents offers the possibility of relief in this situation since higher temperatures can be sustained by these materials without damage to them. However, the RESOX requirement to produce a stream containing 20 percent (vol) sulfur dioxide is severe, because a yield of less than one part sulfur dioxide per part of oxygen is produced. To produce a gas containing 20 vol percent sulfur dioxide in nitrogen, assuming that regeneration is represented by:

$$4\ \mathrm{Fe\,S_{1.5}} + 9\mathrm{O_2} = 2\ \mathrm{Fe_2O_3} + 6\ \mathrm{SO_2}$$

the regeneration gas would have to contain 27.3 percent (vol) oxygen 72.7 percent (vol) nitrogen. Such a gas corresponds to using oxygen enriched air and it is believed that such a gas would lead to impractically high temperatures, even for silica supported absorbents.

The requirement to produce a low oxygen content gas of high (20 percent vol) sulfur dioxide content, at a steady flow rate, as feed to the RESOX process remains a problem connected with air related regeneration.

The Morgantown finding that oxygen consumption is a function of extent of regeneration when a nitrogen oxygen mixture is employed for the purpose requires analysis because it may be misleading. When oxygen rich gas (99 percent vol) is used (at low rate to avoid bed fusion) for regeneration, the Morgantown experience is that it reacts essentially to completion with the production of sulfur dioxide rich gas.

This difference between the oxygen rich and air experiences is an anomaly; nitrogen is an inert component and it should not prevent the reaction proceeding to completion as observed in the oxygen rich case.

It is thought that the explanation lies in the temperature established within the regenerating bed, which should be higher in the oxygen rich case. In all probability, if the nitrogen rich gas is supplied to the regenerating bed at temperatures encountered in the oxygen rich case, the oxygen will be found to react to extinction, as in the oxygen rich case. Thus, by using a high enough or programmed increasing inlet temperature to the regenerating bed, it should be possible to stabilize the exit gas composition from the bed and to produce a more stable sulfur dioxide concentration in the effluent gas. As bed inlet temperature is increased, the nitrogen "ballast" flow must ultimately be increased to avoid causing fusion of the absorbent bed and again tending to depress regeneration effluent sulfur dioxide concentration.

Discussion in this area is limited by the proprietary nature of RESOX process details.

Sulfur Dioxide Recycle Regeneration. In this case a gas of high sulfur dioxide concentration (near 100 percent) is anticipated as the regeneration product. Regeneration temperature is to be selected such that oxygen conversion is complete in the regenerating bed. Sulfur dioxide oxygen mixture composition is to be set such that the temperature in the regenerating bed cannot exceed the safe level to avoid damaging the bed. In this circumstance, it would be possible to dilute purged sulfur dioxide with nitrogen in order to produce a satisfactory RESOX process feed. It may develop that the RESOX process is capable of processing streams of high sulfur dioxide concentration.

As the reducing medium for a gas rich in sulfur dioxide, process gas (tar free gasifier effluent) is a possibility. In support of this study Allied Chemical Corporation found the gas produced by the slagging gasifier to be suitable for the purpose, involving a thermal efficiency penalty of about 1.5 percent when Illinois No. 6 coal is employed. However, sulfur dioxide reducing process chemistry is complex and all gasifier produced fuel gases may not be suitable for the purpose. Any effort to develop a sulfur dioxide recycle regeneration process using process gas for the reduction of purged sulfur dioxide should be supported by a study of the suitability of gas from the particular gasifier.

The Allied Chemical Corporation process employing coal derived fuel gas for reducing sulfur dioxide to elemental sulfur is supported by laboratory study information and by commercial experience with the use of natural gas.

Lepsoe (13, 14) reported on the kinetics and thermodynamics of sulfur dioxide reduction by process gas components.

A commercial example of the Allied Chemical sulfur dioxide reduction process based on the use of coal derived reducing gas is not available. A process profile of Allied Chemical SO_2 Reduction Technology is provided in Appendix D.

Establishment of Steady Feed to Sulfur Dioxide Reduction Unit. Both process gas and coal based sulfur dioxide reduction processes require a steady controlled feed situation which will be difficult to sustain in connection with iron oxide regeneration. Average continuous sulfur dioxide rates were employed as the basis for discussions with process licensors. With several absorbent vessels in use, it appears to be possible to program regeneration to approach this situation. Capital costs and thermal costs of sulfur dioxide reduction have been estimated on the continuous feed basis. They may be higher, and unconverted sulfur dioxide may be higher if feed supply is unstable.

The reduction processes will be able to accommodate normal feed rate and composition variations consistent with general process experience. It should not be anticipated that "on-off" feeds or high amplitude oscillating feed situations will be satisfactory. Therefore, an area of uncertainty exists in relation to establishing a satisfactory continuous and stable regeneration rate.

Related Topics Summary. A major area of uncertainty in hot purification related evaluations is the energy cost of iron oxide bed regeneration. Where air derived regeneration gas is used, it must be nitrogen enriched in order to avoid meltdown of the absorbent bed during regeneration. Use of nitrogen enriched gas tends to produce a regeneration vent gas of low and fluctuating sulfur dioxide content, typically between 1 percent and 7 percent (vol) sulfur dioxide. Sulfur dioxide in such a gas may be converted to sulfuric acid as was practiced in the Appleby-Frodingham process. This possibility has not been examined in the current project.

If a coal based sulfur dioxide regeneration unit is employed, the feed to it must contain at least 20 percent (vol) sulfur dioxide in order to achieve satisfactory conversion to elemental sulfur. Therefore, where nitrogen enriched air is used for regeneration, some form of sulfur dioxide concentration unit would be required. Design of such a sulfur dioxide concentration unit to accommodate a possibly intermittent feed of varying composition was judged to be virtually impossible.

The possibility exists for carrying out regeneration based on recycling a hot sulfur dioxide oxygen mixture over the regenerating bed of iron oxide. This process, if it were developed, offers the possibility of producing a feed satisfactory for coal based sulfur dioxide reduction.

A process based on employing fuel gas as the reducing agent is a feasible and cleaner route to reducing sulfur dioxide to elemental sulfur. Allied Chemical Corporation supported this conclusion based on preliminary analysis of the slagging gasifier data of this study.

More investigation would be required to ascertain that all the fuel gases involved in this study are satisfactory for use in the Allied Chemical SO_2 reduction process.

The thermal penalty associated with sulfur dioxide reduction to elemental sulfur from Illinois No. 6 coal is projected to be approximately 1.5 percentage points. This arises as a result of the process gas consumed for sulfur dioxide reduction with appropriate corrections for waste heat recovery in the reduction unit.

Removal of Particulate Material from Gasifier Effluent

Plugging of a hot sulfur absorption vessel with particulate material (dust) entrained from a gasifier is a potential problem. Dust removal facilities installed between the gasifier and the sulfur absorption vessel must protect against plugging of the absorbent bed in operation when the loading of dust in the fuel gas is normal and in the upset periods when the fuel gas carries an abnormally high loading of dust. The practical problems involved will be more appreciated as hot purification progresses through pilot and demonstration plant stages.

An idea of the magnitude of the dust problem on the commercial scale can be obtained from data published by ERDA in which operation of a stirred bed gasifier on Western Kentucky No. 9 and Upper Freeport coals produced a fuel gas containing approximately 0.5 lb of dust per 1,000 scf. A similar dust loading in the gas leaving a commercial slagging gasifier would have a potential dust deposition rate of about one ton per hour in the hot sulfur removal vessel. Similar magnitude dust deposition rates would be expected for other types of gasifier. On the commercial scale, experience tends to support the above order of gasifier effluent dust loading which, unchecked, would rapidly cause plugging of the sulfur absorbent bed.

In practice, dust loading is a function of gasifier type, gasifier operation, and coal properties, and it will vary within wide limits. Entrained bed gasifiers may be expected to present the most difficult fuel gas dust removal problems because of the severe attrition processes which occur within them.

Morgantown development work is showing that particulate material accumulating in the absorbent bed in sulfur absorption can be burnt out during regeneration, restoring the bed to its fresh condition. Current work at near atmospheric pressure anticipates protecting against high levels of dust contamination of the absorbent bed using cyclone collectors. It seems clear that some such protection is necessary.

A major barrier to particulate removal analysis is lack of data on dust size and loading in the fuel gases leaving gasifiers and sulfur absorbent beds.

In these studies, two methods which are potentially suitable for avoiding dust accumulation in the sulfur absorbent beds are considered.

- Commercial refractory lined two stage cyclones
- Positive sand filters incorporating a pulsed reverse flow (puff-back) cleaning of the type being developed by City College of New York

Related methods which may have potential applicability would be a renewable panel filter of the type suggested by Rexnord Corporation, Combustion Power Company, and Ducon Company.

Cyclones for Absorbent Bed Protection. Employment of cyclones for protection of absorbent beds against dust plugging is attended by major questions. Thus, the satisfactory Morgantown experience in cyclone application on the small scale may not be repeated on the commercial scale. Information is not available defining dust particle size and character to indicate that the dust is amenable to separation by cyclone. Similarly, no information exists to define to what level dust contamination must be reduced in order that absorbent bed performance remain unimpaired throughout the entire sulfur absorption cycle. Problems related to tar fouling, dust collection, and transportation systems have an unknown impact on the design and operation of these systems.

Tars tend to condense from Lurgi-type fuel gases in the dust beds in lockhoppers and cyclones and in dust transfer lines. It is to be anticipated that steps must be taken to exclude tar containing gases from such locations though Morgantown work has not identified such problems in small scale experimentation.

Ducon Corporation advises that cyclones have been built for coal gasification, designed for internal pressure, and ASME code stamped. Sample drawings from a cyclone vendor were used as a basis for estimating the cost of cyclones required based on the slagging gasifier gas flow data.

The cyclones are internally lined for insulation and erosion protection and water jacketed for safety using a similar rationale to that discussed for the sulfur absorbent vessels.

The two-stage pressure cyclone configuration is preferred in comparison to the system where the cyclones are placed in a pressure containment vessel, since elimination of the pressure containment vessel approximately halves the delivered cost of the cyclones. In addition, the individual pressure cyclones conveniently discharge dust into separate lockhoppers through separate cyclone dust legs which can be kept purged with hot gas or steam to avoid fouling them with condensed tars. It would be difficult, if not impossible, to avoid tar fouling of the lower area of a cyclone containment pressure vessel.

The main concern in connection with cyclones is the potential inadequacy of performance. Using a dust particle size distribution and loading derived from Morgantown data, and assuming it to be present in fuel gas from a slagging gasifier at approximately 300 psig, 1,200°F, Ducon estimated a maximum removal factor of 60 percent using a two-stage system. Provision for three stages of cyclone separation made little improvement. The poor performance was attributable to the effect of a high gas density in reducing the ability of the system to separate the finer dust fractions. With a 1.0 ton/hr dust loading suggested previously and 60 percent removal, the potential accumulation rate in the sulfur absorbent vessels would be 0.4 ton/hr, or 3.2 ton/8 hr cycle. The removal of 3.2 tons of dust by combustion and entrainment from the sulfur absorbent vessel during regeneration is difficult to accept as a practical objective.

If all the dust leaving the cyclone were to pass through the absorbent bed, it would clearly endanger the gas turbine at the above rates.

It cannot be claimed that these findings are scientific, being based on so few data. However, the concerns are the result of consultation and rough estimation but they are plausible and require investigation before cyclones can be applied with confidence to hot fuel gas purification on the commercial scale.

One aspect of cyclone application not investigated but relevant is use of techniques to promote agglomeration in order to increase average dust particle size and thereby improve cyclone performance. The success of such a device could eliminate the need for more complex filters.

A final possibility for cyclones is to follow them with a particulate polishing device. Rexnord Corporation has suggested a panel filter/cyclone combination which takes advantage of promoted agglomeration.

Panel Bed Fuel Gas Filter. In this device, which has been investigated on the laboratory scale under the direction of Prof. A. M. Squires at the City College of New York (CCNY), the gas to be treated flows in a horizontal direction across a bed of granular solid supported in a tall, thin panel. The bed is supported in place by louvered walls resembling Venetian blinds rigidly fixed in an open position. See figures in Appendix C. Dust removed from the treated gas accumulates on the entry surfaces of the panel between the louvers which comprise the walls.

CCNY has developed a technique termed "puff-back" for removing solids from the gas-entry surfaces of such a panel. It is accomplished by applying a surge of backflowing gas from the clean side of the panel. The sharp surge of gas produces a substantially uniform spill of solid from each gas entry surface when applied in reverse direction across the panel, provided that the surge has the proper intensity and the panel is of suitable design.

At CCNY, efficiencies of beyond 99.9 percent have been achieved for removal of redispersed power station fly ash at normal stack gas loadings in atmospheric air.

Although CCNY development has been at atmospheric temperature, the panel bed could possible be designed to provide performance at about 1,000°F or above which would be adequate to avoid dust accumulation in a sulfur absorption vessel. However, a design for such high temperature operation has not been developed except in a conceptual form suitable for approximate economic evaluation. Tests of a laboratory panel bed at about 1,000°F at Morgantown have shown that finer sands are required to achieve a given filtration efficiency as the temperature is increased.

Dr. K. C. Lee of CCNY (12) has outlined an idea of how the panel bed may be incorporated into a unit for the filtration of the fuel gas leaving a gasifier. In the Lee design, four square panel bed elements are enclosed in a single pressure vessel as shown in the diagrams collected into Appendix C. The square panel bed element is suggested in EPRI report 243-1 prepared by Stone & Webster Engineering Corporation. The multiple-element-in-pressure-vessel configuration is an outgrowth of this study and it is probably not the optimum design. A better configuration may be to enclose each square panel bed element in its own small diameter pressure vessel, in which case the cost of the panel bed unit may be much reduced.

In the Lee design, over the top of each gas exit nozzle from the central clean gas duct of each square panel bed element, there will be a cup-sleeve. The cup-sleeve will be moved up during filtration by a specific panel bed element and will be moved down to cover the clean gas exit nozzle when applying puff-back through the cup-sleeve. See diagrams in Appendix C.

With the moving cup-sleeve concept, switching valves in the main hot gas flow lines are not required.

The Lee proposal suggests the design of a filter sand recycling system and aims that the sand be kept under pressure at all times. This system is not reported in Appendix C because it is not essential to the successful application of the filter. For this study, it is assumed that the sand/fly ash mixture will be withdrawn from the vessel, dust will be elutriated from the mixture and returned to the gasifier oxidation zone, and the used sand will be recycled or disposed of.

A new air cannon, product of Martin Engineering Co., Neponset, Illinois, is proposed by Dr. Lee as the source of motive gas for puff-back operation. This device is scheduled for testing in CCNY laboratories.*

The CCNY panel bed device is a filter which requires a filter cake on the dirty gas inlet surface to be effective. It is most effective in applications where the dust load is large enough to permit the rapid formation of such a cake. Thus, the device is proposed for use between the gasifier and hot sulfur absorption vessels where higher dust loads are expected. The CCNY panel is not proposed for gas turbine protection unless the dust load in the turbine feed gas is high. In most cases this would seem to be unlikely since the sulfur absorption vessels themselves are likely also to provide assisting filtering action.

In comment, the CCNY panel filter unit is a complex operation which requires a great deal of development in order to become commercially practical for hot purification. The complicated puff-back system which must be designed for severe differential pressures and high temperatures represents an untried system. In addition the panel bed system as proposed by Dr. Lee was estimated to be approximately twice the delivered cost of a two-stage cyclone in pressure vessel for similar service. The cost factor on the panel bed compared to a two-stage pressure cyclone system - without containing pressure vessel - would be closer to a cost factor of five.

In fairness to the CCNY panel bed, costs may be reduced by installing each panel element in its own pressure vessel rather than enclosing multiple elements in a single large vessel. The complexity is for the most part confined to external systems for puff-back control and sand and dust handling. Removal of switching valves from the main gas lines at the cost of providing the moving cup-sleeves would seem to make the latter worth pursuing, although the moving cup-sleeves themselves present mechanical design problems. The performance of the CCNY panel bed promises to surpass that of any conventional cyclone for protection of hot

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*Since writing, the Martin Air Cannon has been reported to have demonstrated excellent performance in application to panel bed puff-back service.

sulfur absorbent beds against accumulating excessive dust deposits during operation. Cyclones may not be adequate for sulfur absorbent bed protection and almost any better device capable of significant filtration duty will be subject to "complexity" criticism. Thus, it is concluded that continuation of CCNY panel bed filter development is justified by its potential for use in hot fuel gas purification as well as by its potential in other applications.

Removal of Particulate Material from Gas Turbine Fuel

As scale increases, development will indicate the problems to be anticipated downstream of the sulfur absorbent beds. It is reasonably certain that cyclones will not be adequate and that some form of polishing filter will be necessary to protect the gas turbines. For estimation purposes, it is only possible to make an allowance for some device yet to be developed.

A CCNY panel bed unit would not be satisfactory if the dust load is low, as may be the case. A renewable panel of a type promoted by Rexnord and which operates as a depth filter could be satisfactory since the sulfur absorbent bed and its upstream particulate protection act as supplementary filters for the Rexnord system.

For gas turbine protection, the main focus of attention in this study has been a filter based on use of a felted woven metal cloth manufactured by Brunswick Corporation. The fibres of such a cloth are fine (a few microns diameter) and vulnerable to disintegration through corrosion. Brunswick has identified a potentially suitable material for construction for the proposed application and is considering a test program to prove its applicability.

It is visualized that a filter of Brunswick metallic cloth would be contained in an unlined vessel, since spalling refractory from the wall of this vessel would pass directly into the gas turbine. The cost and complexity of such a policy are considerable. A single alloy vessel, 6 ft diam x 18 ft length, would cost about \$0.5MM designed for 345 psig and 1,400°F, design conditions which are consistent with those of the sulfur absorption vessels. One such vessel would be required in operation at all times downstream of each sulfur absorbent vessel. For each operating filter vessel, a clean filter vessel must be available which can be switched into service when the operating filter becomes fully loaded with dust. Thus, on the basis of slagging gasification (six gasifiers, 1,000Mw) about \$6MM would be required to purchase pressure vessels (without installation and piping) for turbine protecting filter containment. Use of better proportioned 9 ft diam x 12 ft vessels would increase cost to about double that of the above 6 ft diam vessel.

These costs are not optimized. By using multiple filter vessels of smaller diameter, employing refractory linings, and using other innovations, costs may be reduced.

To clean a Brunswick filter, the cloth must be vibrated and backflushed with inert gas. Before development of the hot fuel gas filter cleaning system, its cost can only be guessed at. Cleaning costs may be expected to have a nondecisive influence in the overall cost of hot purification.

In operation, the Brunswick system would require use of switching valves in the main gas lines to bring units in and out of service - difficulties with hot switching valves should be anticipated.

Based on low pressure low temperature experience, Brunswick Industries has detailed knowledge of dust loading and superficial gas velocity which will probably apply in hot purification related operation.

Section 3

PURIFICATION OF FUEL GASES AT LOW TEMPERATURE

FACTORS IN PROCESS EVALUATION

No single recommendation as to the best low temperature sulfur removal process can be made. The selection is situation dependent involving the following factors:

- Ability to remove all sulfur compounds
- Selectivity of process for sulfur compounds removal in the presence of carbon dioxide. Generally with lower carbon dioxide removal, more gas is left for expansion through the combined cycle unit gas turbine and a better Claus (sulfur recovery) unit feed gas is produced
- Tendency to physically absorb combustible fuel gas components and therefore to reduce the power generating capacity of the system
- Tendency of solvent to react chemically with fuel gas components and to reduce the power generation capacity of the system or to become deactivated
- Tendency of absorbent to contaminate the fuel gas with substances which would be harmful to gas turbine equipment
- Flexibility - ability to operate with changing feedstock conditions and more stringent product quality specifications
- Energy consumption for stripping, remembering that saved energy must be at a level where it can be usefully employed for other purposes
- Environmental compatibility of the solvent. Is it harmless, poisonous, biodegradable, etc.
- Capability to remove nitrogen compounds which cause NO_x pollution on combustion
- That the process be commercially proven in the capacity range proposed

Both Benfield Corporation (Benfield Process) and Allied Chemical Corporation (Selexol Process) demonstrated exceptional chemical engineering ingenuity and insight in support of the current project. It is particularly regretted that for reasons of confidentiality, it is not possible to publish in full the designs developed by Benfield Corporation. The Selexol process demonstrated competitive designs in the cases studied.

The conclusion reached is that for any actual combined cycle project, management is advised to make a detailed low temperature process evaluation study taking into account the selection criteria listed.

The situation dependent nature of cold purification process selection is best understood through example. Thus, a slagging gasifier producing a gas low in carbon dioxide content would not involve selectivity for hydrogen sulfide as a significant factor. An oxygen blown Lurgi-type gasifier producing a fuel gas with high carbon dioxide content would call for careful attention to this topic. The same cold process would not necessarily be selected for both of the preceding applications.

It is important not to reject any process without reviewing the situation with the licensor. As a result of process flowscheme modifications, the licensor is often able to meet specifications which would be uneconomical in a simple absorber stripper combination.

In the context of a hot and cold purification comparison, a detailed study of cold purification processes is not required. What is required is to be relatively sure that a reference cold process is used which is economic, consistent with the general expectation for such processes.

Cost estimates are to be found in various publications which strongly favor one or other of the cold processes. Such quoted data can be misleading and should not be used as the basis for selection. Order of magnitude curve derived cost estimates from different sources are generally not satisfactory for comparison purposes.

SULFUR DISTRIBUTION IN LOW TEMPERATURE PURIFICATION COMBINED CYCLE UNITS

Removal of 90 percent of the sulfur compounds from the fuel gas by low temperature purification is achieved comfortably. A Benfield study of slagging gasification was furthermore able to increase sulfur absorption from 90 to 95 percent while maintaining tower sizes in the region of 20 ft diam, etc. (single train). Conversion of 90 percent of the hydrogen sulfide produced by the acid gas removal unit to elemental sulfur in a Claus plant is comfortably achieved. Thus, the low temperature processes are capable of achieving the performance specified to be acceptable from hot purification (Section 2, p. 2-7).

In addition, by increasing sulfur absorbent circulation rate, use of maximum efficiency Claus technology, and the addition of Claus tail gas treatment, sulfur emissions from cold purification process configurations can be reduced to lower levels than were assumed for this study. The maximum feasible extent of sulfur emission reduction was not investigated.

Section 4

THERMAL EFFICIENCY STUDY

BACKGROUND OF THERMAL EFFICIENCY STUDY

Development of a process to purify fuel gas at near to gasifier exit temperature may increase combined cycle power generation efficiency to above the level feasible based on purification of a cooled coal derived gas. The objective in this section is to test and quantify the above hypothesis for the types of coal gasifiers previously identified.

In studying thermal efficiency, it is important not to lose sight of the practical state of the art and not to be misled into anticipating performance gains which may not be practically feasible. A significant theoretical thermal efficiency advantage for a given type of gasifier supplying a combined cycle power unit and supported by hot purification conveys little information about the difficulty of filtering dust from that gasifier effluent gas. A high cold gas efficiency for a given type of gasifier conveys nothing regarding the problems of liquid slag removal. For these reasons, it is useful to review the general state of the practical art before discussing thermal efficiency in the context of this study.

Fuel gas manufactured from coal for consumption in combined cycle power plants leaves the gasifier at typically 1,000°F to 2,000°F and 300 psia. The gasifier effluent is contaminated with particulate matter (dust), nitrogen compounds (ammonia), and sulfur compounds (hydrogen sulfide and carbonyl sulfide) which must be reduced in concentration in order to protect the environment and the gas turbine before burning the gas to generate electrical power. Available gas purification technology requires cooling the gas to about 250°F for scrubbing out sulfur compounds. A high degree of dust removal is incidentally achieved in gas cooling and scrubbing. The fact that gas is cooled provides the opportunity to scrub ammonia from fuel gases and therefore to control NO_x emission levels from the combined cycle power plant.

The current necessity to cool gasifier produced fuel prior to its purification creates the problem of making effective use of the heat removed from the gas. It is logical that purifying hot (at near gasifier exit temperature) would eliminate much of the thermal and power loss associated with fuel gas cooling. Thus, hot purification is pursued on the premise that the thermal losses associated with it are considerably lower than the losses associated with cooled gas purification. The major losses taken into account in hot purification efficiency estimation are associated with regeneration of the solid sulfur absorbent used in the process.

Five gasification cases have been considered in this work to represent current, near term future, and advanced technology. Air and oxygen blown Lurgi gasifier cases represent current technology. An oxygen blown, slagging gasifier (British Gas) represents near term future technology. Air and oxygen blown entrained bed gasifiers (Foster Wheeler Corporation) represent advanced technology. In the material balances of Section 4, p. 4-8, each gasifier is defined by a design basis which was established by EPRI in consultation with the licensors concerned.

Each of the five gasifiers has been considered in association with hot and cold fuel gas purification. For each gasifier-purification combination, combined cycle power plants were considered based on 1,950°F and 2,400°F gas turbine inlet temperatures.

Commercial cooled fuel gas desulfurization processes offered by Benfield Corporation (the Benfield process) and Allied Chemical Corporation (the Selexol process) were studied in the cooled gas cases. Results are based on application of the Benfield process. Process performance data indicated Selexol to be economically similar to Benfield and that, depending on circumstances, it may be the preferred choice. Choice of low temperature process is discussed qualitatively in Section 3.

It is not the intent in this study to prejudice competition between the low temperature gas desulfurization processes. The intent here is to assess the relative value of hot purification, not to distinguish narrowly between established cooled gas purification processes. The possibility exists for obtaining different results in terms of the relative value of hot purification by comparing it with other cooled gas purification processes which may be better than those considered. Because of the latter and because the hot purification process will continue to improve with development, the exercise of comparing the economics of hot and cooled gas purification is expected to continue.

The process for the absorption of sulfur on supported iron oxide at 1,000°F to 1,500°F being developed by the United States Energy Research Administration (the Morgantown process) is the method for hot fuel gas desulfurization which has been evaluated.

It is also not the purpose in this work to study in detail the technical and economic benefits to be anticipated with gasification combined cycle power systems. The incentives for the introduction of such systems must be measured in studies which address detailed practical and theoretical aspects of integrated gasification and gas turbine technology. Thus, this work employs only a realistic simulation of combined cycle power unit performance and the same simulation is employed for both hot and cold case analyses. It yields a reliable indication of the difference between the cases for a given gasifier, though the absolute calculated performances are not precise.

In relation to the different types of gasifier studied, attention should be focused on comparing hot and cold gas purification for a given gasifier, as discussed in Section 1 . Comparison between gasifiers would require a detailed study of each gasifier and would focus on comparing optimized operations. While every effort has been made to select best commercial expectations for each gasifier studied, the detailed studies necessary for comparative gasifier evaluation have not been carried out.

POWER UNIT CALCULATION BASES

The combined cycle power generation simulation employed in this study is realistic but not precise, as is consistent with the requirements of a comparative study of hot and cold purification. For the purposes involved, it was unreasonable to request gas turbine system manufacturers to perform an analysis of all cases under study. General Electric Corporation (GE) supplied data in support of early work which were later discarded as the related gasifier bases were improved. Subsequently, the GE data were employed to assist in setting the parameters of an approximate steady state model of the gas turbine and air compressor system of Table 4-1.

The interest and assistance of GE in this project are gratefully acknowledged. However, the possibly oversimplified gas turbine simulation employed is not a representation of GE's expectations in this area or in any way a responsibility of GE.

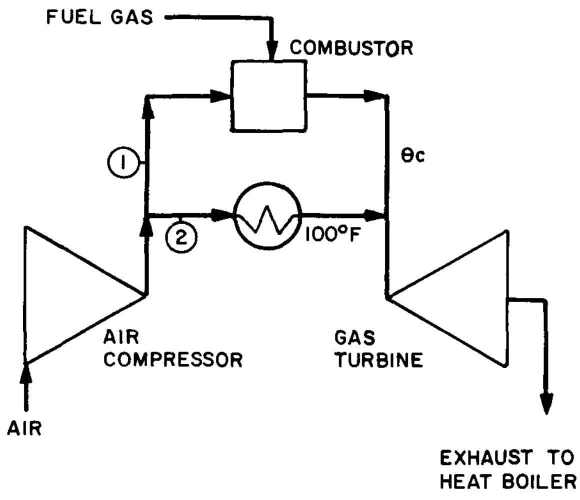
The simplified simulation has been used in this study in the belief that the inaccuracies in it largely cancel out in calculating efficiency differences between systems employing different purification methods. As a result, such efficiency differences are attributable to purification system differences. The possibility that hot purified fuel gases will incur special penalties not incurred by cold purified gases when employed as gas turbine fuel has not been considered in this study but cannot be ruled out.

For up-to-date combined cycle performance data, maintenance of contact with gas turbine manufacturers is recommended. Recent publications may be helpful in this regard (1, 21).

In designing the steam system, steam produced by cooling the turbine exhaust gas is integrated with the steam production and requirements of the fuel gas production system to develop the overall design. Figures 4-1 and 4-2 are examples of integrated power and process system diagrams related to the slagging gasifier cases.

Table 4-1

GAS TURBINE CYCLE - SIMULATION PARAMETERS



Gas Turbine Inlet Temp. θc	Compression Ratio
1,950°F	10:1
2,400°F	16:1

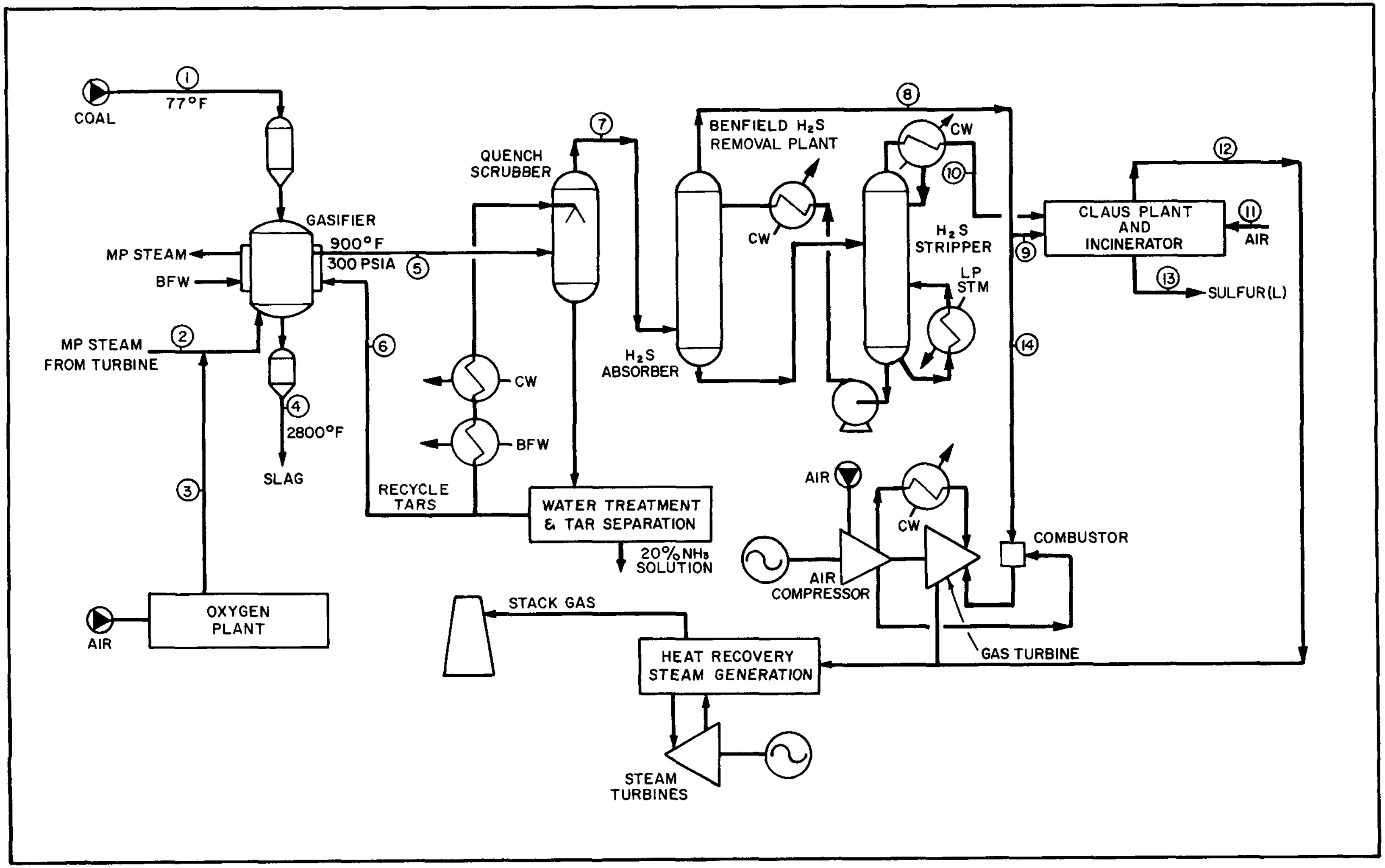


Figure 4-1. Slagging Gasification Study - Cold Purification Case - Schematic (MSL)

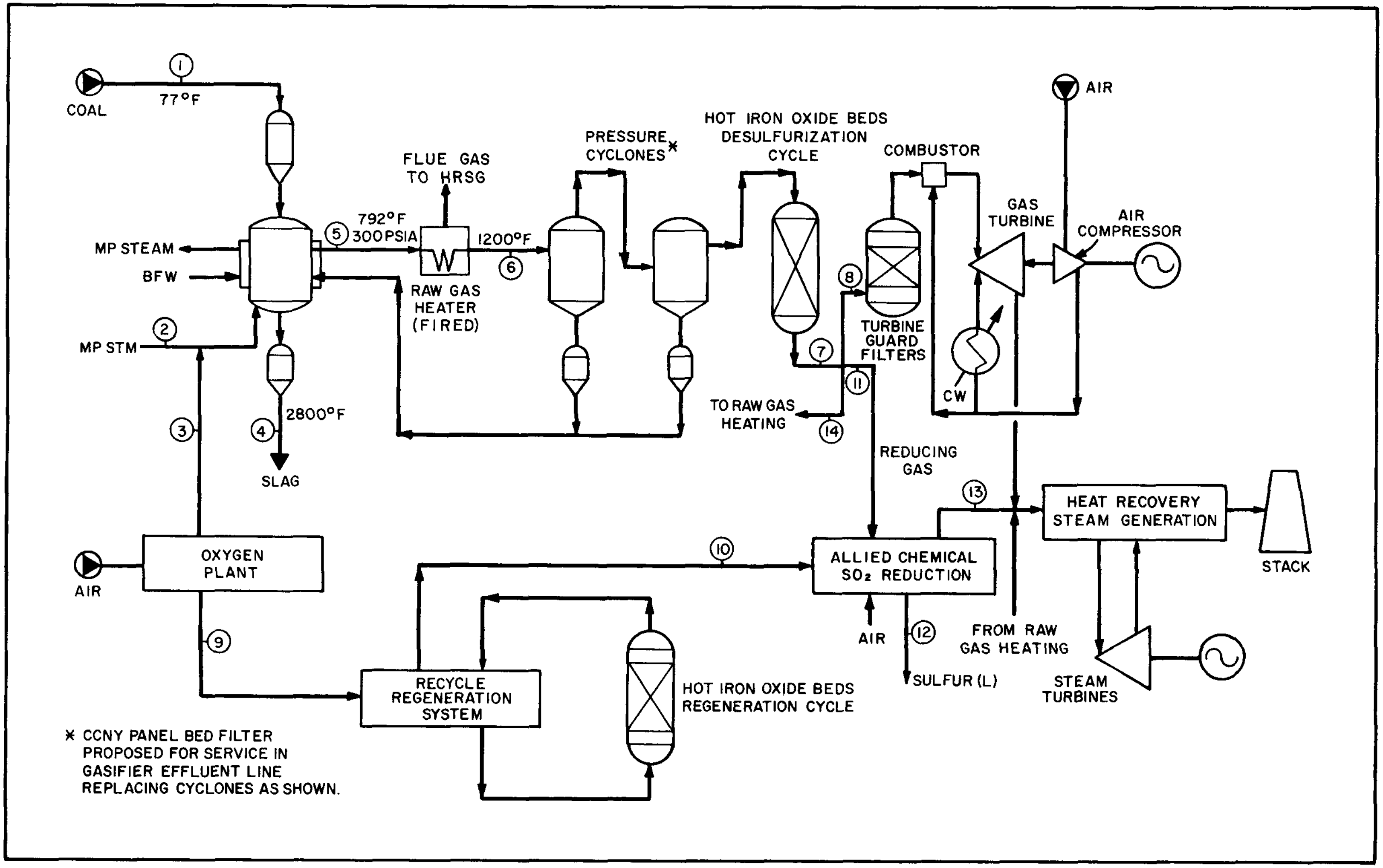


Figure 4-2. Slagging Gasification Study - Hot Purification Case - Schematic (MSH)

Table 4-2

NOMENCLATURE FOR IDENTIFICATION OF CASES

GASIFIER	TYPE	PURIFICATION	GAS TURBINE INLET TEMPERATURE
M = Moving Bed (Lurgi)	X = Oxygen Blown	L = Low Temp (e.g. Benfield)	1,950
E = Entrained Bed	A = Air Blown	H = High Temp (Morgantown Process)	2,400
	S = Slagging/ Oxygen Blown		

Example: Case MXL 1950 refers to a Lurgi moving bed gasifier, oxygen blown with low temperature fuel gas desulfurization supplying fuel to a gas turbine with an inlet temperature of 1,950°F.

SUPPORTING INFORMATION - FLOWSCHEMES, MATERIAL BALANCES AND HEAT BALANCES

With the nomenclature of Table 4-2, the cases studied are as follows:

- Cases MSL 1950 and MSL 2400, corresponding to a slagging gasification combined cycle power plant employing low temperature fuel gas desulfurization with gas turbine inlet temperatures of 1,950°F and 2,400°F respectively.
- Cases MSH 1950 and MSH 2400, corresponding to a slagging gasification fueled combined cycle power plant employing hot fuel gas desulfurization with gas turbine inlet temperatures of 1,950°F and 2,400°F respectively.
- Cases MAL 1950, MAL 2400, MAH 1950, and MAH 2400, defining air blown Lurgi cases in similar manner to the above MS series.
- Cases MXL 1950, MXL 2400, MXH 1950 and MXH 2400, defining oxygen blown Lurgi cases in similar manner to the above MS series.
- Cases EAL 1950, EAL 2400, EAH 1950 and EAH 2400, defining air blown Foster Wheeler entrained bed gasification cases in similar manner to the above MS series.
- Cases EXL 1950, EXL 2400, EXH 1950 and EXH 2400, defining oxygen blown Foster Wheeler entrained bed gasification cases in similar manner to the above MS series.

In all cases, the type of coal studied is as specified by the following analysis:

TYPE	ILLINOIS NO. 6
PROXIMATE ANALYSIS	
Moisture	4.2 wt %
Ash	9.6
Fixed Carbon	52.0
Volatile Matter	34.2
	100.0
ULTIMATE ANALYSIS - DAF COAL	
Carbon	77.26 wt %
Hydrogen	5.92
Oxygen	11.14
Nitrogen	1.39
Sulfur	4.29
	100.00

TYPE	ILLINOIS NO. 6
HEATING VALUE - AS RECEIVED	
Higher Heating Value (HHV)	12,235 Btu/lb
Net Heating Value (LHV)	11,709 Btu/lb
AS PURCHASED	
Moving Bed Cases	Washed, sized 1 1/2 x 1/4, unit train delivered to plant battery limits
Fluid and Entrained	Washed, sized 1 1/2 x 0, unit train delivered to plant battery limits

Tables 4-3A and B summarize material balances for cases MSL and MSH respectively. Table 4-3A defines the fuel gas produced to supply the gas turbine for both cases MSL 1950 and MSL 2400 based on processing one million pounds per hour of moisture and ash free Illinois No. 6 coal. Table 4-3B defines the fuel gas similarly for cases MSH 1950 and MSH 2400. Stream numbers in Table 4-3A correspond to numbers indicated on Figure 4-1. Stream numbers on Table 4.3B correspond to numbers indicated on Figure 4-2.

Figures 4-1 and 4-2 show the configuration of major power generation equipment, process equipment and satellite process units associated with cases MSL and MSH respectively.

Figures 4-3 and 4-4 show the steam system in schematic form estimated for cases MSL 1950 and MSL 2400 respectively and corresponding to the material balance Table 4-3A, Figures 4-5 and 4-6 show the steam system in schematic form estimated for cases MSH 1950 and MSH 2400 respectively and corresponding to the material balance Table 4-3B. Tables 4-3C to 4-3F summarize overall energy balances for MSL 1950, MSL 2400, MSH 1950 and MSH 2400, corresponding to the material balances of Table 4-3A and B respectively. Tables 4-3A-F combined with Figures 4-1 - 4-6 document the results of slagging gasifier studies.

Other gasifier studies are documented in abbreviated form by Figures 4-7 - 4-14 and by the corresponding short form material balances, Tables 4-4 - 4-7 respectively.

In order to determine overall combined cycle thermal efficiency, it is necessary to perform overall energy and material balances on the fuel gas production and power generation units. In order to calculate the overall balance it is necessary to divide the plant into sections and to perform separate balances round each section, which in turn details the losses from each section. The component section balances are then combined into the overall heat balance of the plants leading to the determination of thermal efficiency. The major plant sections considered are:

- Gasifier
- Oxygen Plant
- Gas Quench and Cooling
- Tar and By-product Separation
- Desulfurization
- Sulfur Recovery
- Gas Turbine
- Waste Heat Recovery/Steam Generation

Comprehensive heat and material balance information is presented for the slagging gasifier (Case MS), as explained in the foregoing presentation of figures and tables. Outline information is presented corresponding to the other gasifier cases MA, MX, EA and EX.

The schematic system diagrams, Figures 4-3 - 4-14, show in general terms a gasifier followed by hot or cold fuel gas desulfurization to produce a clean gas turbine fuel. The gas turbine exhaust gas enters a heat recovery steam generator which supplies the steam turbine driven power generation unit and the balance of process steam requirements. The gas turbine drives the combustion air compressor and a second power generation unit. Gasification oxygen and booster air compressor steam turbine drive requirements are supplied by the heat recovery steam generation unit.

In the cold purification cases, fuel gas leaving the gasifier is cooled in a train of exchangers prior to entering a Benfield desulfurization unit. The Benfield unit is shown in simplified representation as a conventional absorber/regenerator (stripper) system.

In the hot purification cases, fuel gas leaving the gasifier passes through a cyclone dust removal unit or a CCNY panel bed type filter, is heated or cooled to the temperature required for hot sulfur removal and enters the iron oxide filled vessels. The function of this heater is to increase gas temperature to above the level where condensation of tars in the absorption vessels can occur. The gas temperature elevation required may possibly be accomplished in other ways, as discussed in Section 2, p. 2-6, and questions of a substantial nature remain to be answered with respect to the need for and methods of tar condensation prevention.

The entrained bed gasifiers with hot purification require a steam boiler and superheater system in the gasifier exit line in order to reduce the temperature of the gasifier exit gas from 1,700°F to the operating temperature of the hot iron oxide vessels (1,200°F). The hot purification cases achieve gas turbine protection by a Brunswick metallic cloth filter, shown in schematic representation in the various figures.

Table 4-3A

CASE MSL MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed To Gasifier		Steam Feed To Gasifier		Oxidant To Gasifier		Slag		Gas From Gasifier		Recycle Tars, Phenols, Etc.		Gas From Quench Scrubber	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04									7,220	115,809			7,220	115,809
Ethane	30.07									531	15,967			531	15,967
Carbon Monoxide	28.01									53,997	1,512,456			53,997	1,512,456
Carbon Dioxide	44.01									1,800	79,218			1,800	79,218
Water	18.02	2,704	48,726	19,205	346,074					4,625	83,343			3,757	67,701
Hydrogen Sulfide	34.08									1,282	43,691			1,282	43,691
Carbonyl Sulfide	60.08									56	3,364			56	3,364
Ammonia	17.03									843	14,356				
Tars	125.0									647.2	80,900	647.2	80,900		
Ash		-	111,370					-	111,370						
Sulfur Dioxide	64.06														
Carbon	12.01	64,324	772,531					189	2,270						
Hydrogen	2.02	29,365	59,317							28,292	57,150			28,292	57,150
Oxygen	32.00	3,481	111,392			16,703	534,496								
Nitrogen	28.02	496	13,898			341	9,555			416	11,656			416	11,656
Sulfur	32.06	1,338	42,896												
Totals		-	1,160,130	19,205	346,074	17,044	544,051	-	113,640	99,709.2	2,017,910	647.2	80,900	97,351	1,907,012
Temperature, F		77		750		392		2,800		900		350		200	
Pressure, Psia				380										280	
Heat Content - Sensible/Latent		-0-		465.682		38.500		93.499		706.777		12.667		158.219	
HHV		14,193.701		-0-		-0-		31.997		14,907.647		1,270.802		13,497.927	
Total Heat - MM Btu/hr		14,193.701		465.682		38.500		125.496		15,614.424		1,283.469		13,656.146	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Slagging gasifier - cold purification
Stream numbers refer to Fig. 4.1

Table 4-3A (Continued)

Stream No.		8		9		10		11		12		13		14	
Component/ Properties	Mol Wt	Clean Gas From Benfield		Fuel to Claus Incinerator		Claus Feed Acid Gas		Air To Claus Plant		Claus Off-gas To Boiler		Sulfur Product		Net Fuel To Combustor	
		Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04	7,220	115,809	24	385									7,196	115,424
Ethane	30.07	531	15,967	2	60									529	15,907
Carbon Monoxide	28.01	53,997	1,512,456	181	5,070									53,816	1,507,386
Carbon Dioxide	44.01	1,021	44,934	3	132	835	36,748			1,047	46,078			1,018	44,802
Water	18.02	2,822	50,852	9	162	879	15,840			2,241	40,383			2,813	50,690
Hydrogen Sulfide	34.08	134	4,567			1,204	41,032							134	4,567
Carbonyl Sulfide	60.08														
Ammonia	17.03														
Tars	125.0														
Ash										60	3,844				
Sulfur Dioxide	64.06														
Carbon	12.01	28,292	57,150	95	192									28,197	56,958
Hydrogen	2.02							855	27,360						
Oxygen	32.00	416	11,656	3	84			3,216	90,112	3,219	90,196	1,144	32,055	413	11,572
Nitrogen	28.02														
Sulfur	32.06														
Totals		94,433	1,813,391	317	6,085	2,918	93,620	4,071	117,472	6,567	180,501	1,144	32,055	94,116	1,807,306
Temperature, F		194		194				77		1,200		300		194	
Pressure, Psia		278		278										278	
Heat Content - Sensible/Latent		131.092		0.437		18.801		-0-		104.280		2.367		130.655	
HHV		13,206.918		44.023		291.359		-0-		-0-		147.973		13,162.895	
Total Heat - MM Btu/hr		13,338.010		44.460		310.160		-0-		104.280		150.340		13,293.550	

Table 4-3B

CASE MSH MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed To Gasifier		Total Steam Feed To Gasifier		Oxidant Feed To Gasifier		Slag		Gas From Gasifier		Raw Gas		Clean Fuel Gas Net to Power	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04									7,067	113,355	7,067	113,355	7,067	113,355
Ethane	30.07									518	15,576	518	15,576	518	15,576
Carbon Monoxide	28.01									48,700	1,364,087	48,700	1,364,087	45,965	1,287,480
Carbon Dioxide	44.01									1,623	71,428	1,623	71,428	4,401	193,688
Water	18.02	2,704	48,726	17,188	309,719					4,423	79,702	4,423	79,702	2,807	50,582
Hydrogen Sulfide	34.08									1,237	42,157	1,237	42,157	118	4,021
Carbonyl Sulfide	60.08									54	3,244	54	3,244	11	661
Ammonia	17.03									843	14,356	843	14,356	843	14,356
Tars	125.0									647	80,900	647	80,900	647	80,900
Ash		-	111,370					-	111,370						
Sulfur Dioxide	64.06														
Carbon	12.01	64,324	772,531					189	2,270						
Hydrogen	2.02	29,365	59,317							24,141	48,765	24,141	48,765	26,876	54,290
Oxygen	32.00	3,481	111,392			14,946	478,272								
Nitrogen	28.02	496	13,898			299	8,378			341	9,555	341	9,555	341	9,555
Sulfur	32.06	1,338	42,896												
Totals		-	1,106,750	17,188	309,719	15,245	486,650	-	113,640	89,594	1,843,125	89,594	1,843,125	89,594	1,824,464
Temperature, F		77		750		392		2,800		792		1,200		1,270	
Pressure, Psia				380						300		290		270	
Heat Content - Sensible/Latent		-0-		416.767		34.436		93.499		604.323		917.801		955.464	
HHV		14,193.701		-0-		-0-		31.997		13,673.653		13,673.653		13,396.056	
Total Heat - MM Btu/hr		14,193.701		416.767		34.436		125.496		14,277.976		14,591.454		14,351.520	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Slagging gasifier - hot purification
Stream numbers refer to Fig. 4.2

Table 4-3B (Continued)

Stream No.		8		9		10		11		12		13		14	
Component/ Properties	Mol Wt	Clean Fuel Gas Net to Power Gen.		Oxidant To Iron Oxide Regen.		Sulfur Dioxide To Reduction		Full Gas to Allied SO_2 Reduction		Sulfur Product		SO_2 Reduction Off-gas		Fuel Gas To Raw Gas Heating	
		Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04	6,669	106,971					184	2,951					214	3,433
Ethane	30.07	489	14,704					13	391					16	481
Carbon Monoxide	28.01	43,390	1,215,354					1,182	33,108					1,393	39,018
Carbon Dioxide	44.01	4,155	182,862					113	4,973			1,505	66,235	133	5,853
Water	18.02	2,650	47,753					72	1,297			1,206	21,732	85	1,532
Hydrogen Sulfide	34.08	111	3,783					3	102					4	136
Carbonyl Sulfide	60.08	11	661												
Ammonia	17.03	795	13,539					22	375					26	443
Tars	125.0	611	76,367					17	2,082					20	2,451
Ash															
Sulfur Dioxide	64.06					1,162	74,438					60	3,844		
Carbon	12.01														
Hydrogen	2.02	25,371	51,249					691	1,396					814	1,644
Oxygen	32.00			1,743	55,776							80	2,560		
Nitrogen	28.02	322	9,022	36	1,009	36	1,009	9	252			1,358	38,051	10	280
Sulfur	32.06									1,105	35,426				
Totals		84,574	1,722,265	1,779	56,785	1,198	75,447	2,306	46,927	1,105	35,426	4,209	132,422	2,715	55,271
Temperature, F		1,270		400		1,284		1,270		300		1,000		1,350	
Pressure, Psia		270				160		270						270	
Heat Content - Sensible/Latent		904.130		4.123		16.627		24.561		2.286		53.693		26.773	
HHV		12,641.200		-0-		-0-		346.775		141.139		-0-		408.081	
Total Heat - MM Btu/hr		13,545.330		4.123		16.627		371.336		143.425		53.693		434.854	

Table 4-3C

OVERALL ENERGY BALANCE
CASE MSL/1950
DATUM 77 F, $H_2O(1)$,10^6 LB/HR MAF COAL

	SENSIBLE & LATENT HEAT MMBTU/HR	HHV MMBTU/HR	TOTAL HEAT MMBTU/HR	PERCENT	POWER KW
Feeds------------------					
Coal @ 77°F	0	14193.701	14193.701		
Combustor Air @ 60°F	-132.033	0	-132.033		
Feeds Subtotal			14061.668	100.00	
Products---------------					
Slag @ 2,800°F	93.499	31.997	-125.496		
Sulfur @ 300°F	2.367	147.973	-150.340		
Ammonia Soln @ 77°F	0	136.918	-136.918		
Stack Gas @ 316°F	2960.658	0	-2960.658		
Products Subtotal			-3373.412	23.99	
Cooling Duties---------					
Quench Trim Cooler			-52.959		
Benfield Plant			-260.916		
Process Water Plant			-208.298		
Steam Turbine Condenser			-4145.219		
Oxygen Plant Cooling			-439.176		
Cooling Subtotal			-5106.568	36.31	
Work------------------					
Gas Turbine			-3878.890		+1135839
Steam Turbine (Net)			-1818.488		+532500
Oxygen Plant			+477.676		-139876
Auxiliaries			+37.025		-10842
Work Subtotal			-5182.677	36.86	1517621
Losses----------------					
Gasifier			-103.008		
Gas Turbine			-296.003		
Losses Subtotal			-399.011	2.84	
Total			0	100.00	

Table 4-3D

OVERALL ENERGY BALANCE
CASE MSH/1950
DATUM 77 F, $H_2O(1)$, 10^6 LB/HR MAF COAL

	SENSIBLE & LATENT HEAT MMBTU/HR	HHV MMBTU/HR	TOTAL HEAT MMBTU/HR	PERCENT	POWER KW
Feeds------------------					
Coal @ 77°F	0	+14193.701	+14193.701		
Combustor Air	0	-137.022	-137.022		
Feeds Subtotal			+14056.679	100.00	
Products---------------					
Slag @ 2,800°F	93.499	31.997	-125.496		
Sulfur @ 300°F	2.367	147.973	-143.425		
Tar from Reducing Gas	0	136.918	-55.490		
Stack Gas	2902.813	0	-2902.813		
Products Subtotal			-3227.224	22.96	
Cooling Duties---------					
Gasifier Cooling			-56.775		
Oxygen Plant			-429.952		
Recycle Regeneration			-474.411		
Steam Turbine Condenser			-4168.999		
Cooling Subtotal			-5130.137	36.50	
Work-------------------					
Gas Turbine			-3990.520		1168527
Steam Turbine (Net)			-1842.734		539600
Oxygen Plant Compressors			+468.511		-137192
Auxiliary Power			+35.256		-10324
Work Subtotal			-5329.487	37.91	1560611
Losses-----------------					
Gasifier			-45.591		
Gas Turbine			-307.189		
SO_2 Reduction			-17.055		
Losses Subtotal			-369.835	2.63	
Total			0	100.00	

Table 4-3E

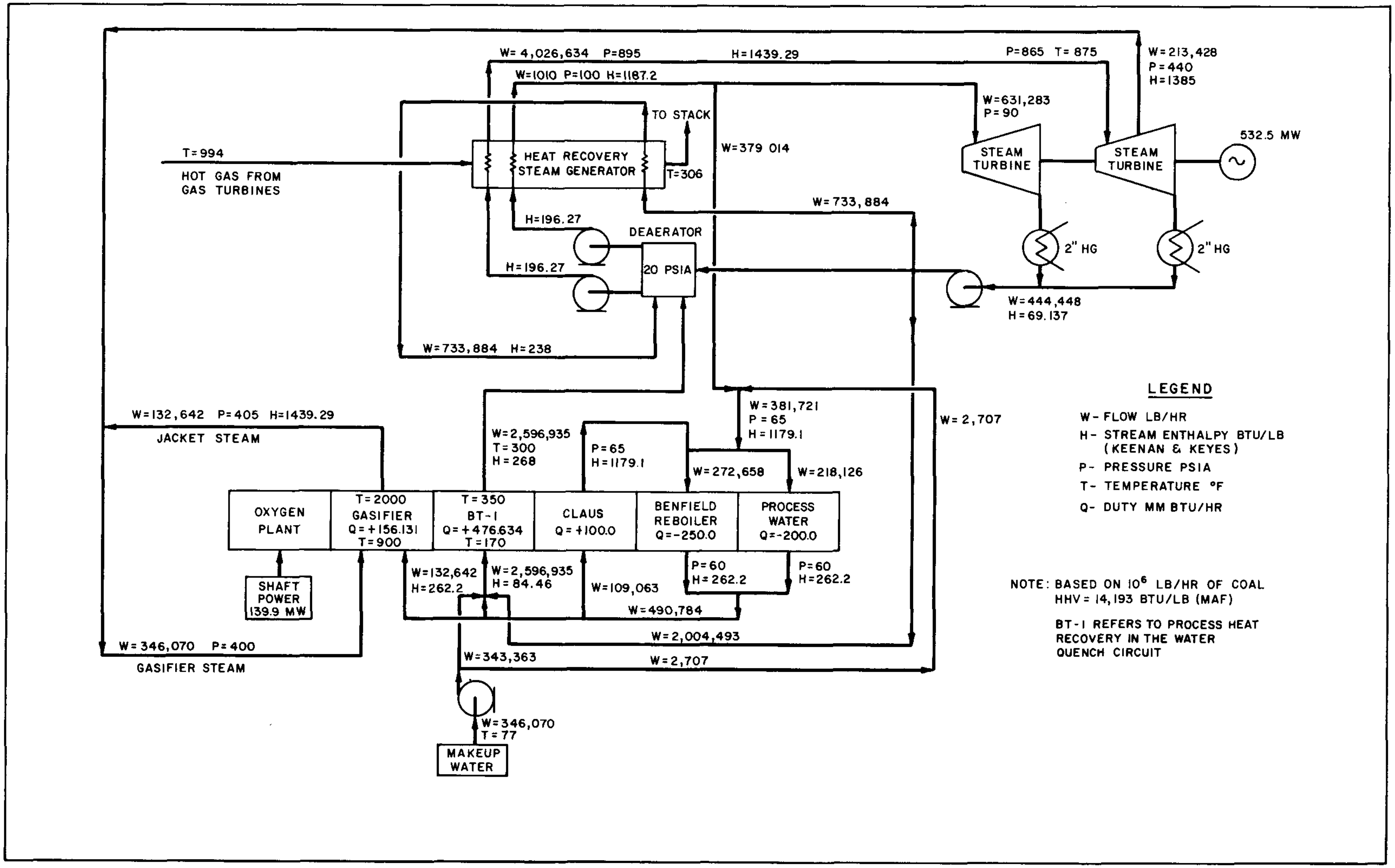
OVERALL ENERGY BALANCE
CASE MSL/2400
DATUM 77, F, H_2O(1), 10^6 LB/HR MAF COAL

	SENSIBLE & LATENT HEAT MMBTU/HR	HHV MMBTU/HR	TOTAL HEAT MMBTU/HR	PERCENT	POWER KW
Feeds------------------					
Coal @ 77°F	0	14193.701	14193.701		
Combustor Air @ 60°F	-110.818	0	-106.495		
Feeds Subtotal			14087.206	100.00	
Products---------------					
Slag @ 2,800°F	93.499	31.997	-125.496		
Sulfur @ 300°F	2.367	147.973	-150.340		
Ammonia Soln 77°F	0	136.918	-136.918		
Stack Gas	2579.513	0	-2579.513		
Products Subtotal			-2992.267	21.24	
Cooling Duties---------					
Quench Trim Cooler			-52.959		
Benfield Plant			-260.916		
Process Water Plant			-208.298		
Steam Turbine Condenser			-3941.347		
Oxygen Plant Cooling			-439.176		
Cooling Subtotal			-4902.696	34.80	
Work-------------------					
Gas Turbine			-4203.701		+1230952
Steam Turbine (Net)			-1931.524		+565600
Oxygen Plant			+477.676		-139876
Auxiliaries			+37.025		-10842
Work Subtotal			-5620.524	39.90	+1645834
Losses-----------------					
Gasifier			-103.008		
Gas Turbine			-468.711		
Losses Subtotal			-571.719	4.06	
Total			0	100.00	

Table 4-3F

OVERALL ENERGY BALANCE
CASE MSH/2400
DATUM 77 F, $H_2O(1)$, 10^6 LB/HR MAF COAL

	SENSIBLE & LATENT HEAT MMBTU/HR	HHV MMBTU/HR	TOTAL HEAT MMBUT/HR	PERCENT	POWER KW
Feeds------------------					
Coal @ 77°F	0	+14193.701	+14193.701		
Combustor Air @ 60°F	-119.818	0	-110.818		
Feeds Subtotal			14082.883	100.00	
Products---------------					
Slag @ 2,800°F	93.499	31.997	-125.496		
Sulfur @ 300°F	2.367	147.973	-143.425		
Tar from Reducing Gas	0	136.918	-55.490		
Stack Gas	2554.047	0	-2554.047		
Products Subtotal			-2878.458	20.44	
Cooling Duties---------					
Gasifier Cooling			-56.775		
Oxygen Plant			-429.952		
Recycle Regeneration			-474.411		
Steam Turbine Condenser			-3929.487		
Cooling Subtotal			-4890.625	34.73	
Work-------------------					
Gas Turbine			-4323.711		+1266094
Steam Turbine Net			-1943.476		+569100
Oxygen Plant			+468.511		-137192
Auxiliary Power			+35.257		-10324
Work Subtotal			-5763.419	40.92	1687678
Losses-----------------					
Gasifier			-45.591		
Gas Turbine			-487.736		
SO_2 Reduction			-17.055		
Losses Subtotal			-550.382	3.91	
Total			0	100.00	



LEGEND

W- FLOW LB/HR
H- STREAM ENTHALPY BTU/LB (KEENAN & KEYES)
P- PRESSURE PSIA
T- TEMPERATURE °F
Q- DUTY MM BTU/HR

NOTE: BASED ON 10^6 LB/HR OF COAL
HHV= 14,193 BTU/LB (MAF)

BT-1 REFERS TO PROCESS HEAT RECOVERY IN THE WATER QUENCH CIRCUIT

Figure 4-3. Slagging Gasification Study - Cold Purification Case, Steam System - MSL 1950

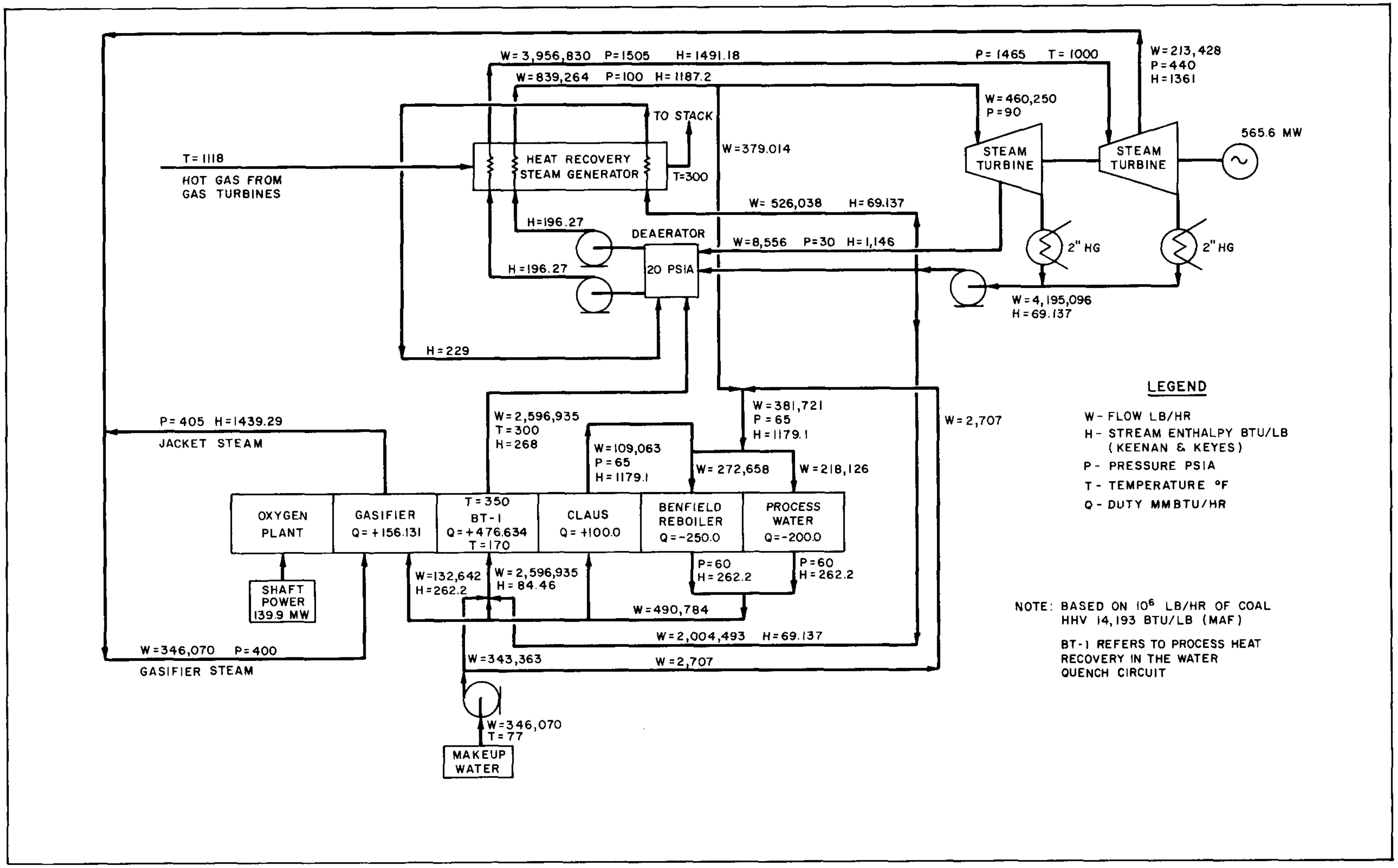


Figure 4-4. Slagging Gasification Study - Cold Purification Case, Steam System - MSL 2400

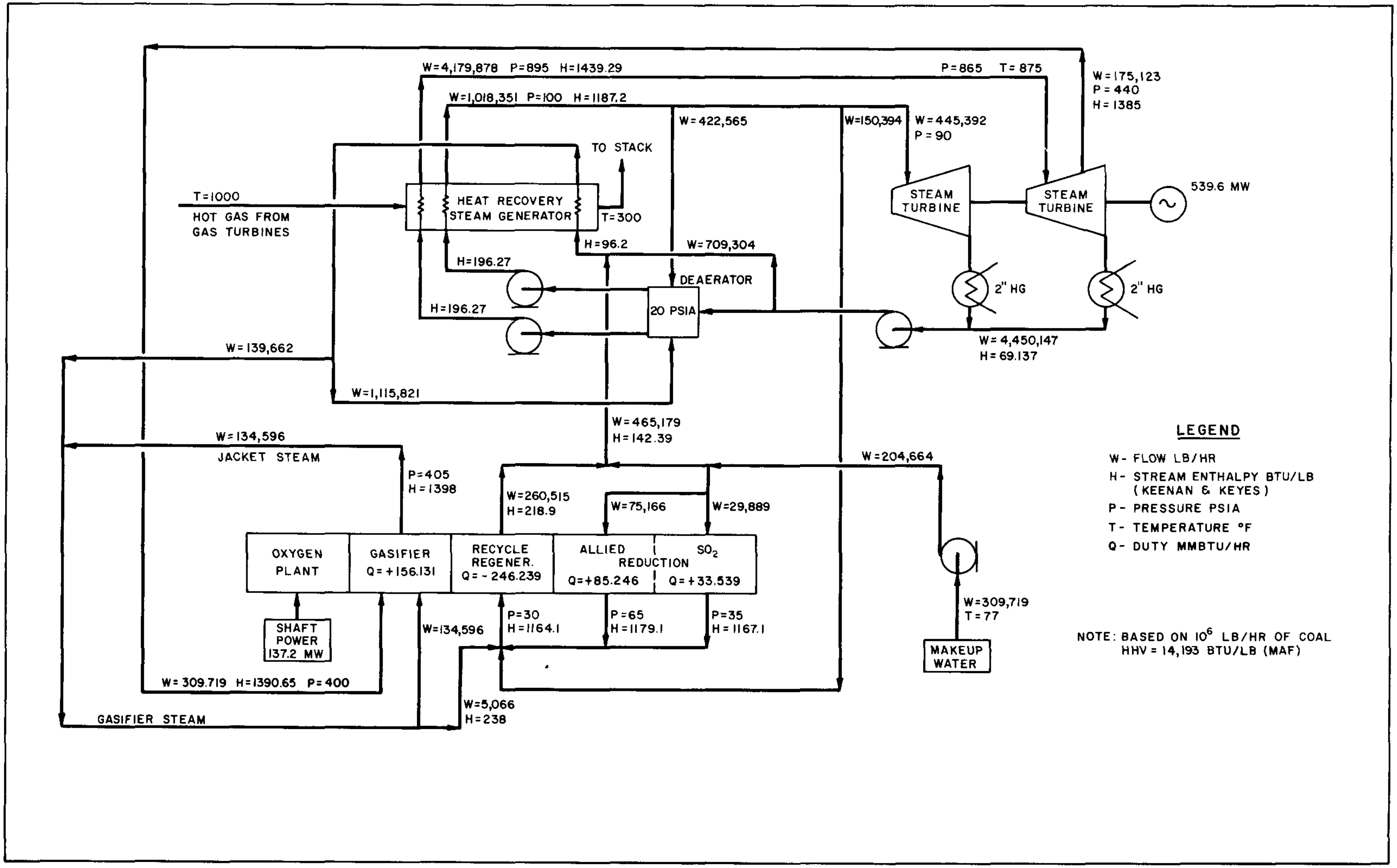


Figure 4-5. Slagging Gasification Study - Hot Purification Case, Steam System - MSH 1950

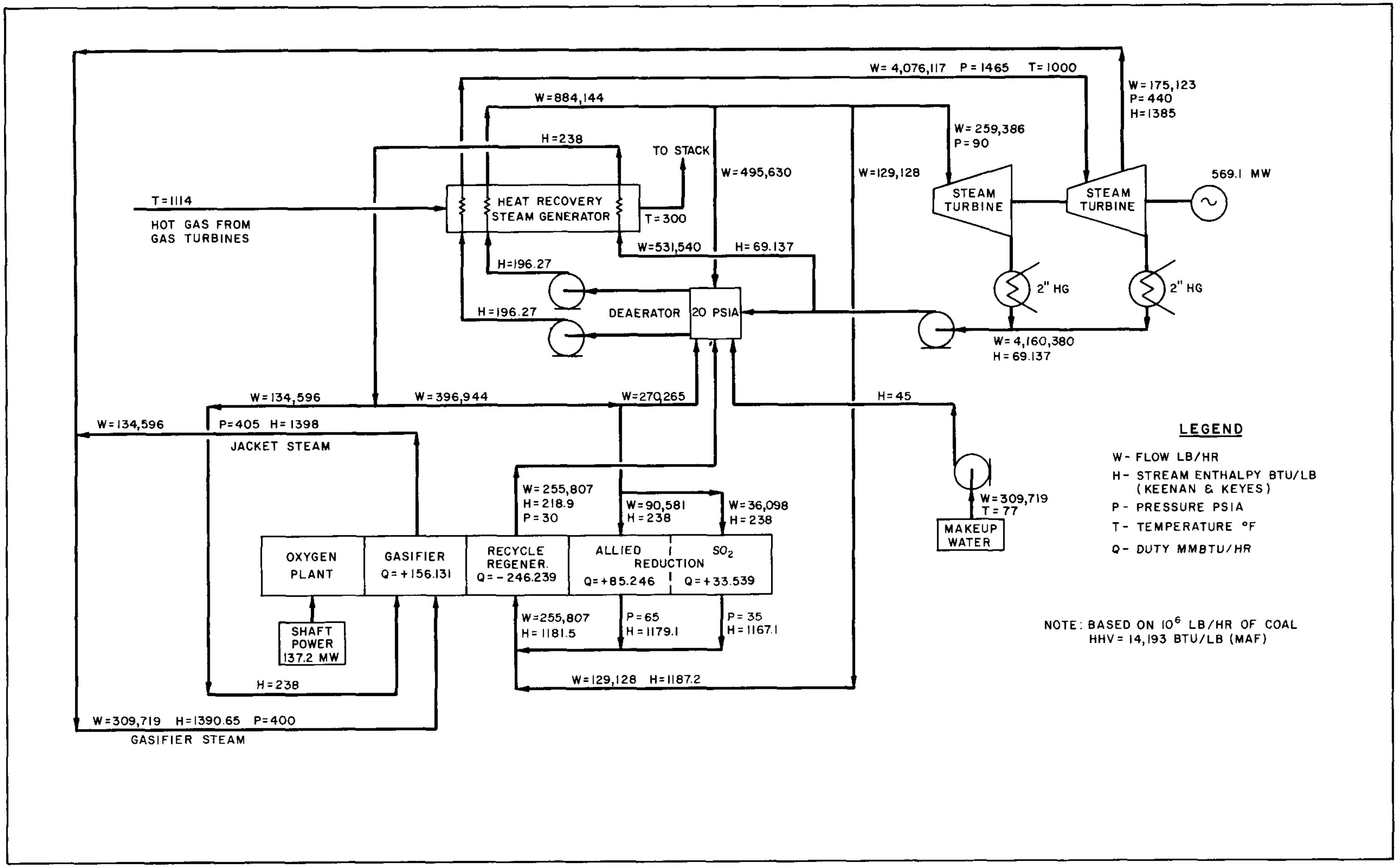


Figure 4-6. Slagging Gasification Study - Hot Purification Case, Steam System - MSH 2400

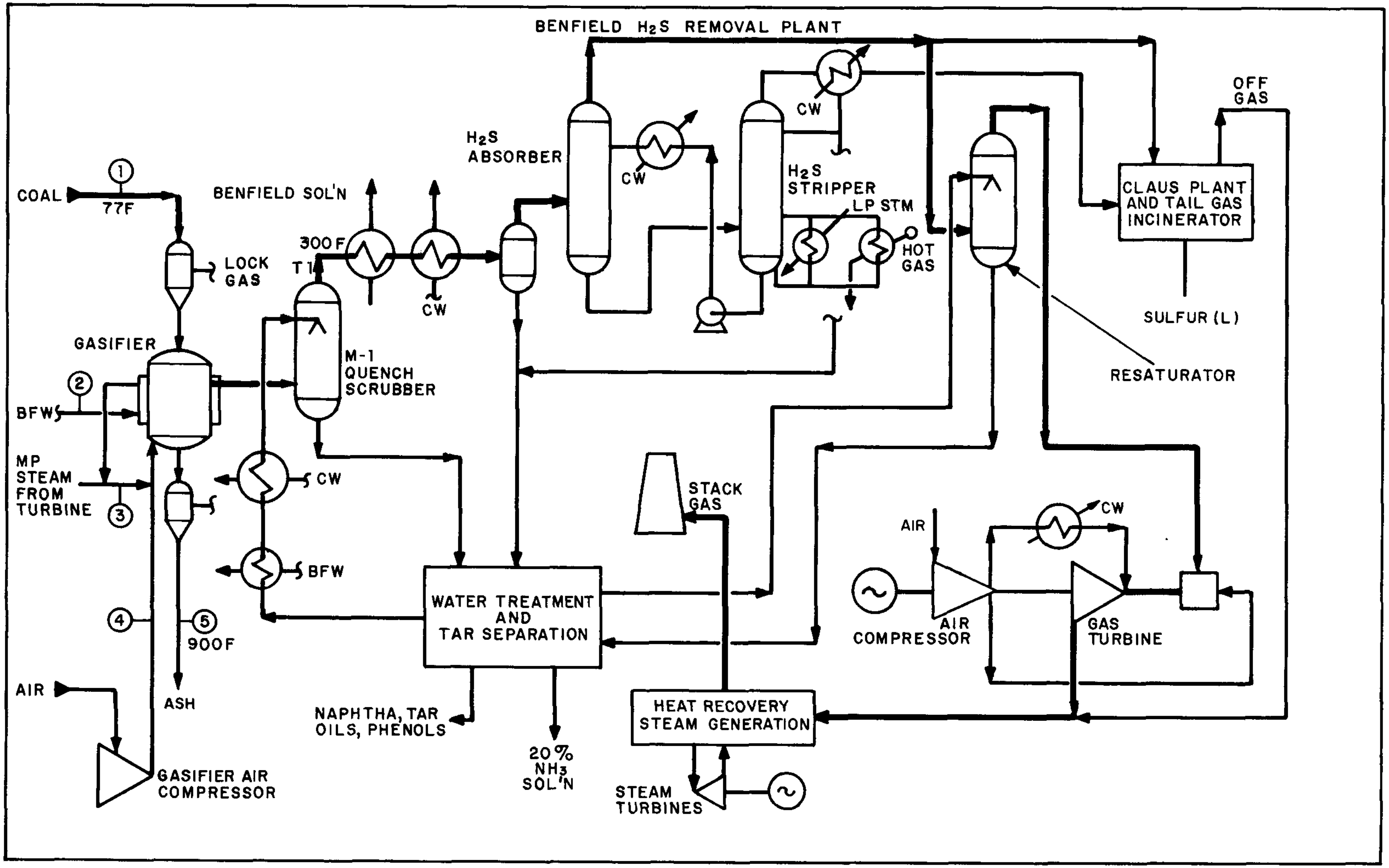


Figure 4-7. Lurgi Air Blown Gasifier Study - Cold Purification Case (Case MAL)

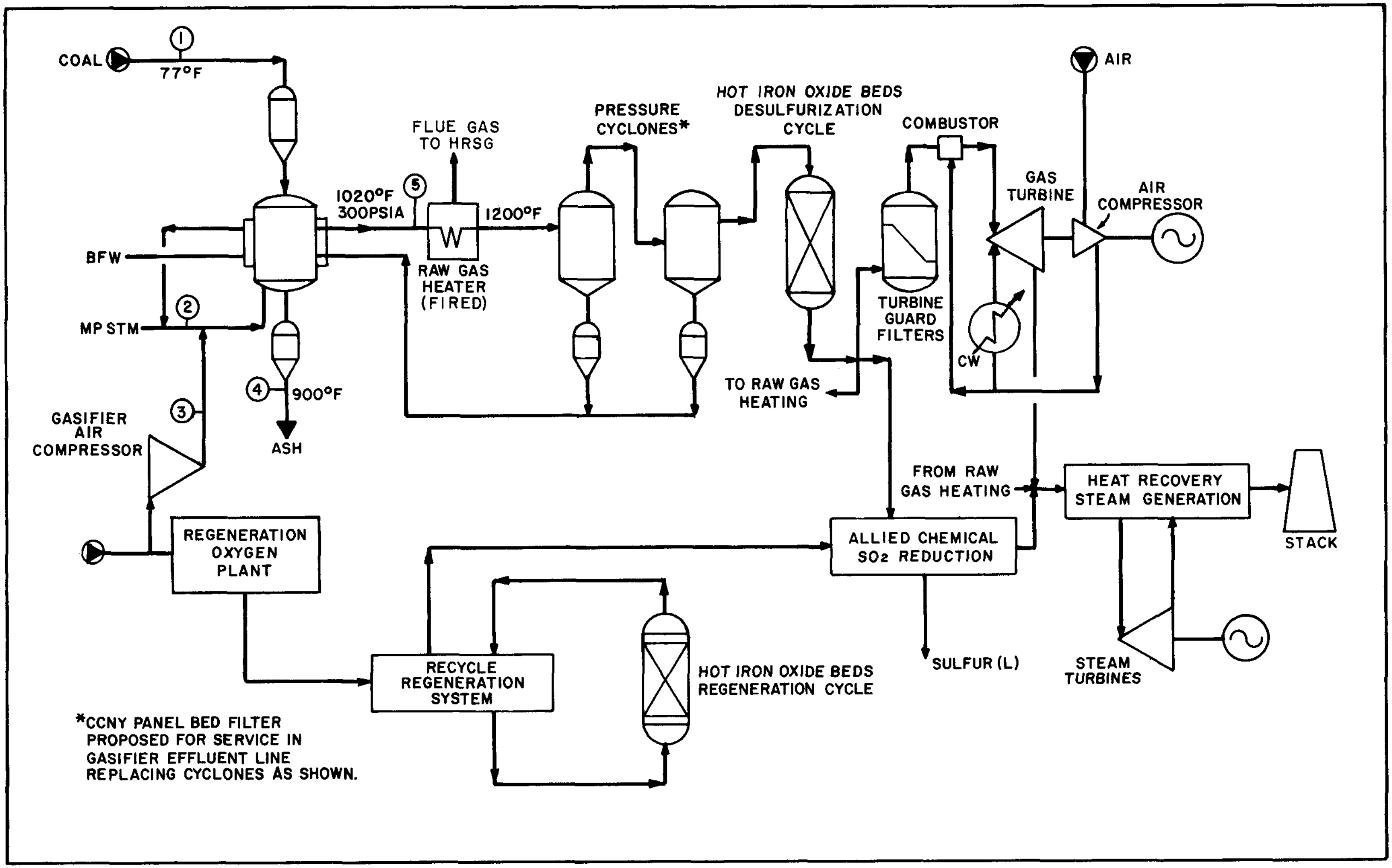


Figure 4-8. Lurgi Air Blown Gasifier Study - Hot Purification Case (Case MAH)

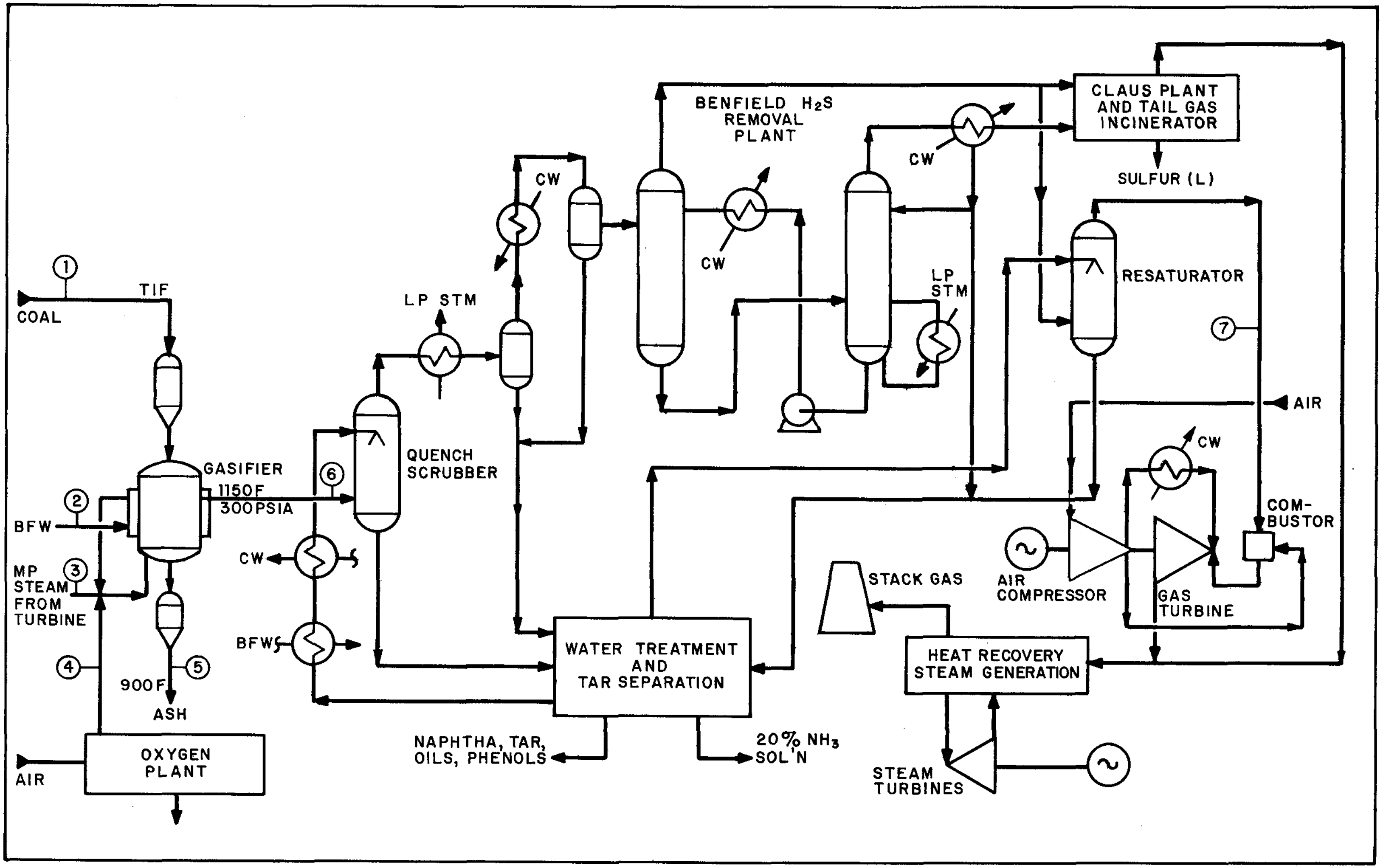


Figure 4-9. Lurgi Oxygen Blown Gasifier Study - Cold Purification Case

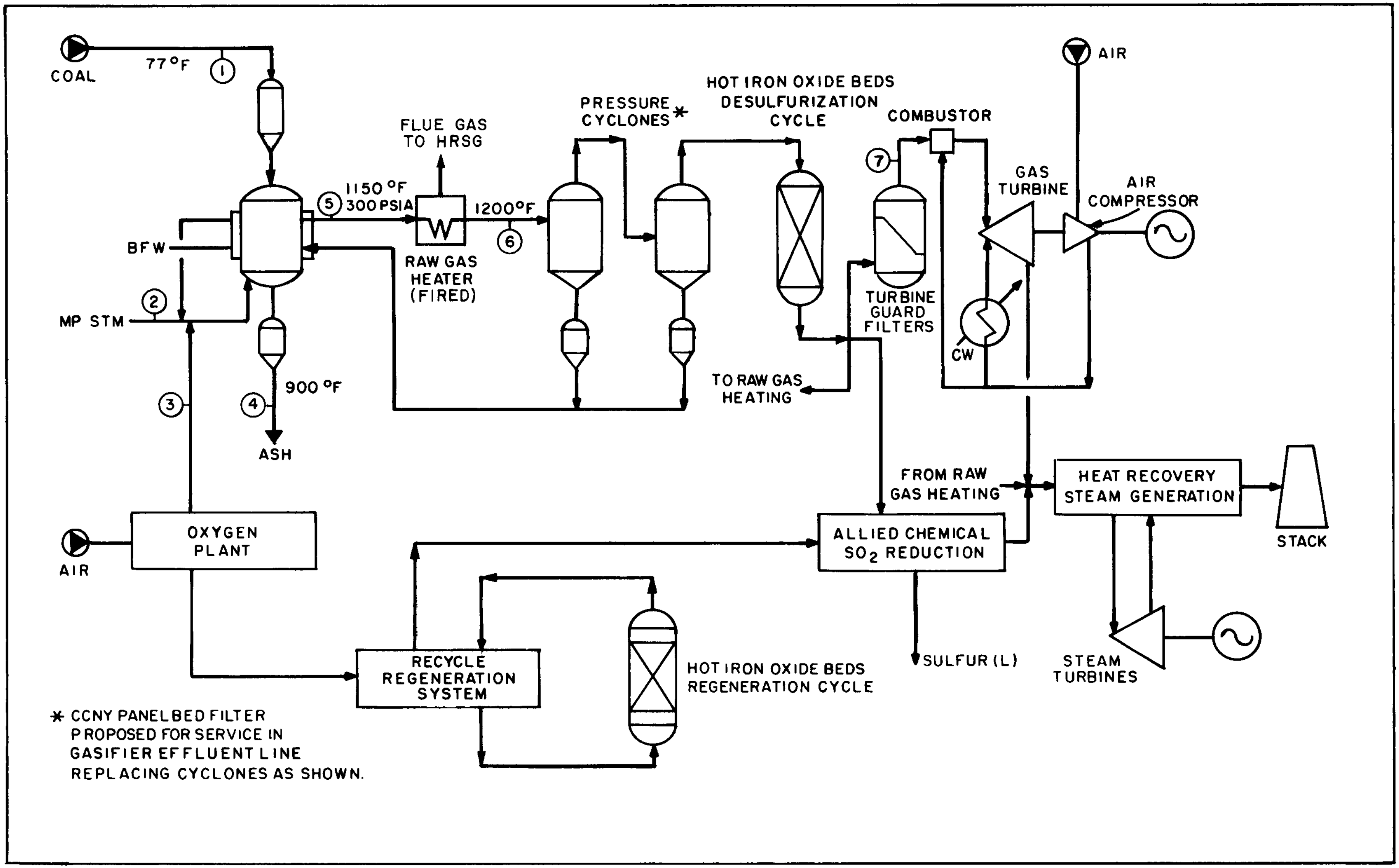


Figure 4-10. Lurgi Oxygen Blown Gasifier Study - Hot Purification Case

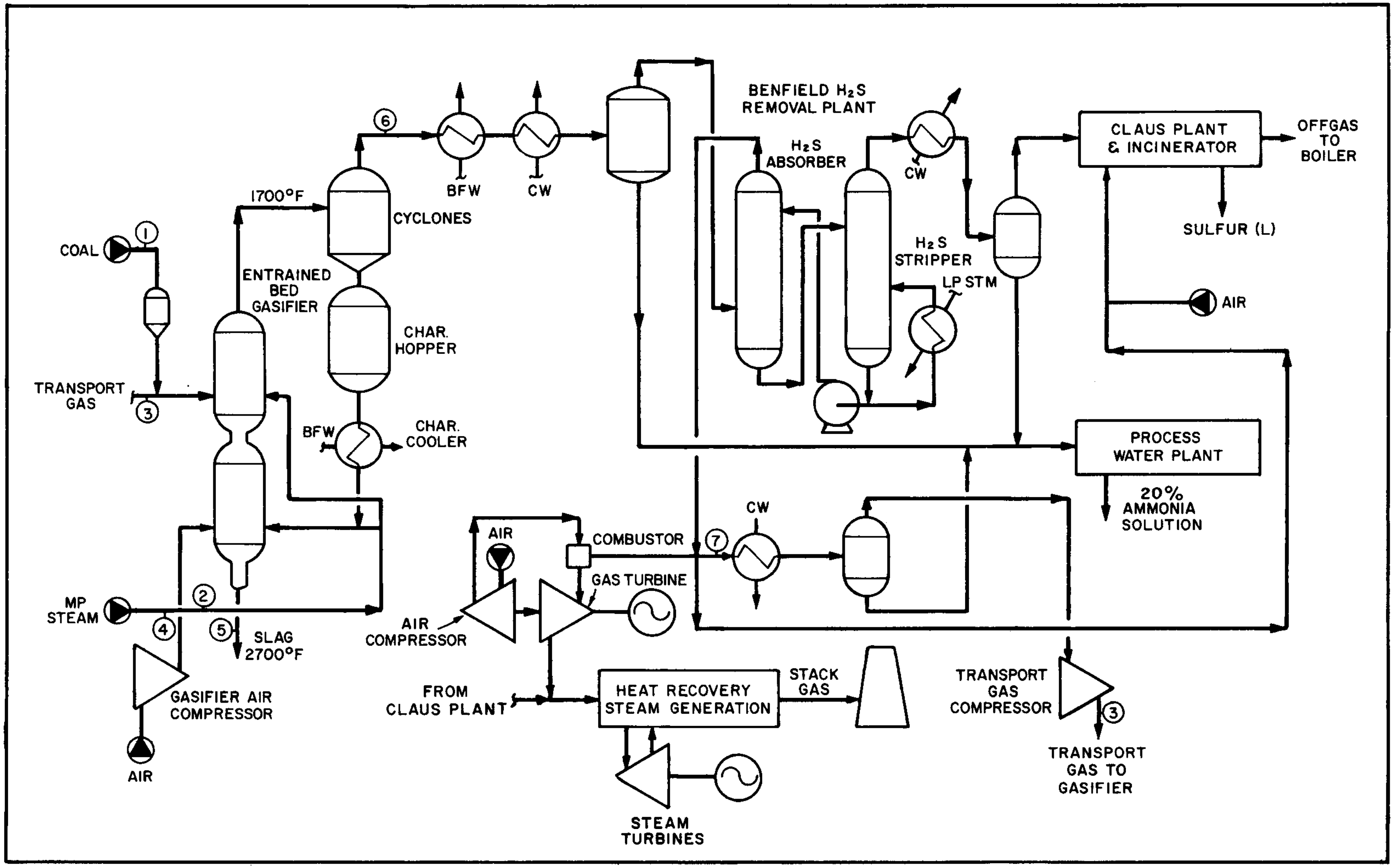


Figure 4-11. Air Blown Entrained Bed Gasifier Study - Cold Purification Case

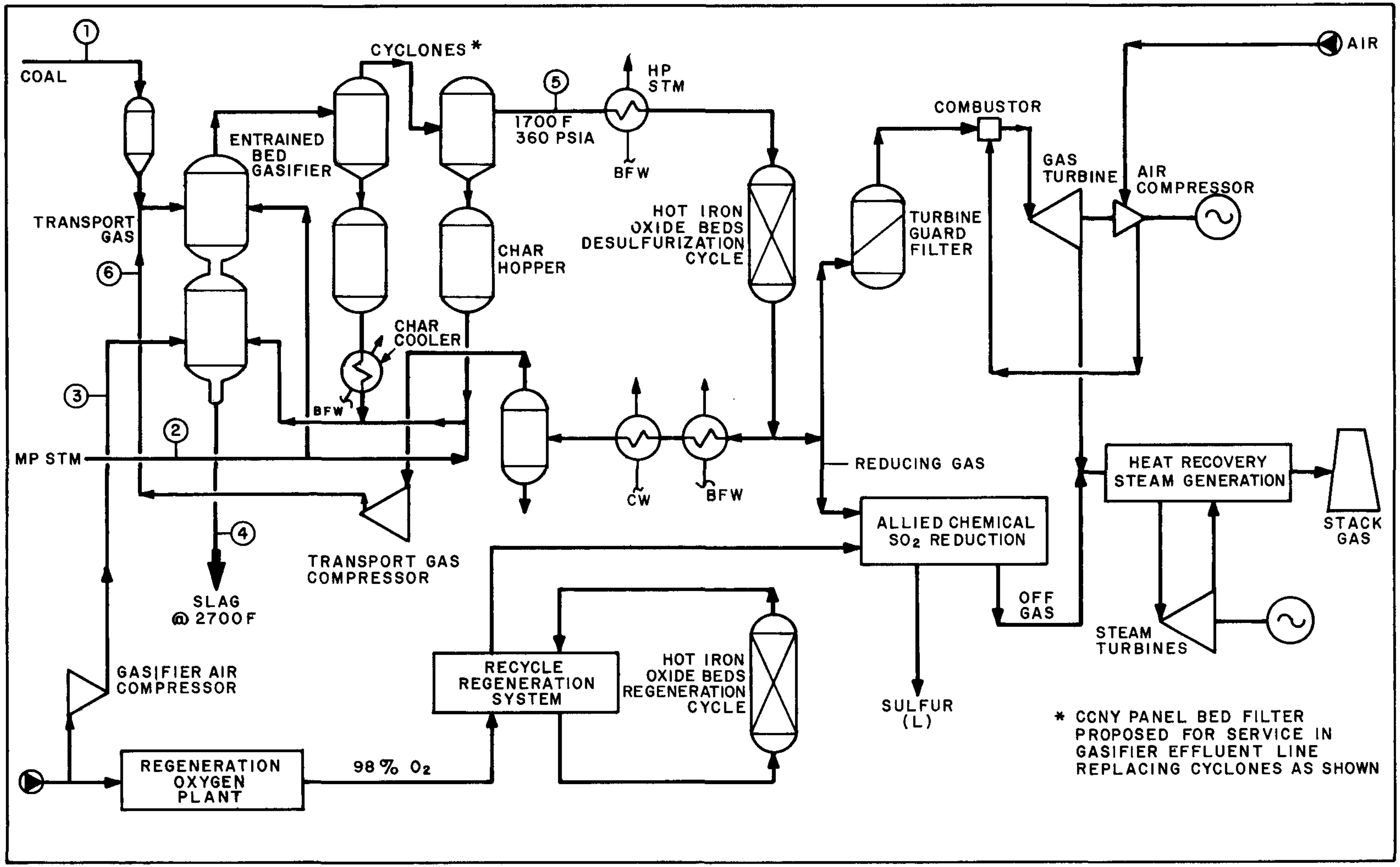


Figure 4-12. Air Blown Entrained Bed Gasifier Study - Hot Purification Case

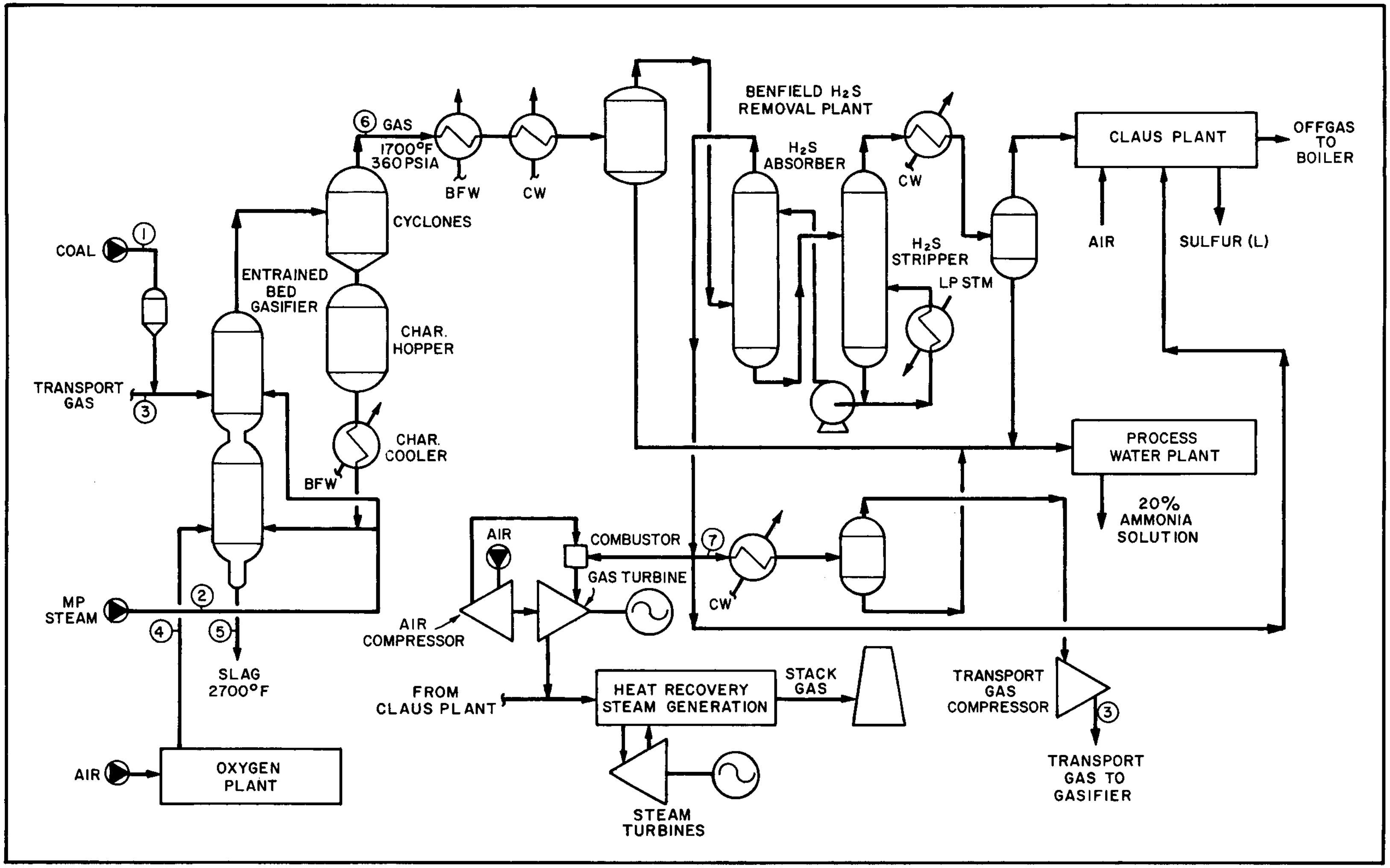


Figure 4-13. Oxygen Blown Entrained Bed Gasifier Study - Cold Purification Case

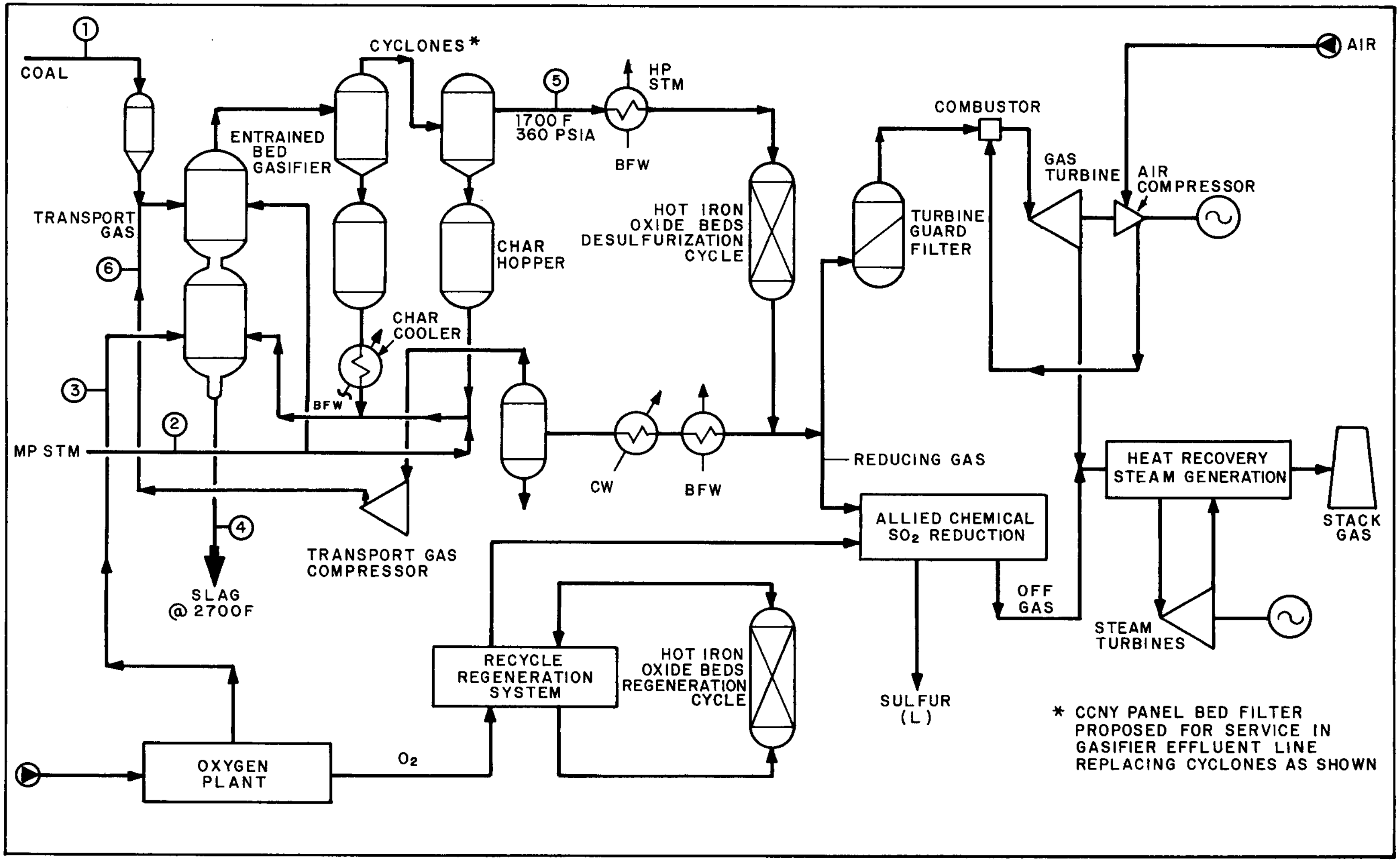


Figure 4-14. Oxygen Blown Entrained Bed Gasifier Study - Hot Purification Case

Table 4-4A

CASE MAL - ABBREVIATED MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed To Gasifier		BFW To Gasifier		Steam Feed To Gasifier		Oxidant To Gasifier		Ash		Gas From Gasifier		Clean Gas From Resaturator	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04											6,734	108,013	6,640	106,506
Ethane	30.07											167	5,022	165	4,962
Carbon Monoxide	28.01											28,222	790,498	27,827	779,434
Carbon Dioxide	44.01											22,938	1,009,501	16,983	747,422
Water	18.02	2,704	48,726	26,637	480,000	65,189	1,174,700	1,312	23,642			65,553	1,181,265	3,948	71,143
Hydrogen Sulfice	34.08											1,237	42,157	88	2,999
Carbonyl Sulfide	60.08											54	3,244	39	2,343
Ammonia	17.03											843	14,356		
Tars	125.0											647	80,900		
Ash		-	111,370							-	111,370				
Sulfur Dioxide	64.06														
Carbon	12.01	64,324	772,531							386	4,636				
Hydrogen	2.02	29,365	59,317									40,568	81,947	40,001	80,802
Oxygen	32.00	3,481	111,392					18,601	595,232						
Nitrogen	28.02	496	13,898					69,974	1,960,671			70,016	1,961,848	69,037	1,934,417
Sulfur	32.06	1,338	42,896												
Totals		-	1,160,130	26,637	480,000	65,189	1,174,700	89,887	2,579,545	-	116,006	236,979	5,278,751	164,729	3,730,028
Temperature, F		77		212		752		392		900		1,020		170	
Pressure, Psia						370		380				300		250	
Heat Content - Sensible/Latent		-0-		64.824		1,581.381		224.332		19.458		3,072.431		186.965	
HHV		14,193.701		-0-		-0-		-0-		65.347		12,837.402		10,990.926	
Total Heat - MM Btu/hr		14,193.701		64.824		1,581.381		224.332		84.805		15,909.833		11,117.891	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Lurgi air blown gasifier - cold purification
Stream numbers refer to Fig. 4.7

Table 4-4B

CASE MAH - ABBREVIATED MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed To Gasifier		BFW To Gasifier		Steam Feed To Gasifier		Oxidant To Gasifier		Ash		Gasifier Effluent Gas		Clean Fuel Gas Net To Power	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04											6,734	108,013	6,559	105,206
Ethane	30.07											167	5,022	163	4,901
Carbon Monoxide	28.01											28,222	790,498	18,432	516,280
Carbon Dioxide	44.01			-								22,938	1,009,501	49,880	2,195,219
Water	18.02	2,704	48,726	26,637	480,000	65,189	1,174,700	1,312	23,642			65,553	1,181,265	55,874	1,006,849
Hydrogen Sulfide	34.08											1,237	42,157	123	4,192
Carbonyl Sulfide	60.08											54	3,244	3	180
Ammonia	17.03											843	14,356	821	13,982
Tars	125.0											647	80,875	630	78,797
Ash		-	111,370							-	111,370				
Sulfur Dioxide	64.06														
Carbon	12.01									386	4,636				
Hydrogen	2.02	29,365	59,317									40,568	81,947	48,569	98,109
Oxygen	32.00	3,481	111,392					18,601	595,232						
Nitrogen	28.02	496	13,898					69,974	1,960,671			70,016	1,961,848	68,196	1,910,852
Sulfur	32.06	1,338	42,896												
Totals		-	387,599	26,637	480,000	65,189	1,174,700	89,887	2,579,545	-	116,006	236,979	5,278,726	249,250	5,934,567
Temperature, F		77		212		752		392		900		1,020		1,250	
Pressure, Psia						370		380				300			
Heat Content - Sensible/Latent		-0-		64.824		1,581.381		224.332		19.458		3,072.431		3,394.784	
HHV		14,193.701		-0-		-0-		-0-		65.347		12,837.402		12,237.169	
Total Heat - MM Btu/hr		14,193.701		64.824		1,581.381		224.332		84.805		15,909.833		15,631.953	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Lurgi air blown gasifier - hot purification
Stream numbers refer to Fig. 4.8

Table 4-5A

CASE MXL - ABBREVIATED MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed To Gasifier		BFW To Gasifier		Steam Feed To Gasifier		Oxidant To Gasifier		Ash		Gas From Gasifier		Clean Gas To Combustor	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04											9,026	144,777	8,936	143,333
Ethane	30.07											662	19,906	655	19,696
Carbon Monoxide	28.01											15,752	441,214	15,594	436,788
Carbon Dioxide	44.01											32,126	1,413,865	25,444	1,119,790
Water	18.02	2,704	48,726	24,972	450,000	118,002	2,126,400					106,122	1,912,318	11,438	205,933
Hydrogen Sulfide	34.08											1,237	42,157	85	2,897
Carbonyl Sulfide	60.08											54	3,244	43	2,583
Ammonia	17.03											813	13,845		
Tars	125.0											647	80,875		
Ash		-	111,370												
Sulfur Dioxide	64.06														
Carbon	12.01	64,324	772,531												
Hydrogen	2.02	29,365	59,317									43,978	88,836	43,538	87,947
Oxygen	32.00	3,481	111,392												
Nitrogen	28.02	496	13,898									385	10,788	381	10,676
Sulfur	32.06	1,338	42,896												
Totals		-	1,160,130	24,972	450,000	118,002	2,126,400		551,019		116,006	210,802	4,171,825	106,104	2,029,643
Temperature, F		77		212		752		392		900		1,150		245	
Pressure, Psia						512		380				300		245	
Heat Content - Sensible/Latent		-0-		60.772		2,847.888		38.992		19.458		4,022.742		465.497	
HHV		14,193.701		-0-		-0-		-0-		65.347		12,948.606		11,145.169	
Total Heat - MM Btu/hr		14,193.701		60.772		2,847.888		38.992		84.805		16,971.348		11,610.666	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Lurgi oxygen blown gasifier - cold purification
Stream numbers refer to Fig. 4.9

Table 4-5B

CASE MXH - ABBREVIATED MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed To Gasifier		BFW To Gasifier		Steam Feed To Gasifier		Oxidant To Gasifier		Ash		Raw Gas From Gasifier		Net Clean Fuel Gas To Power Gen.	
Component/ Properties	**Mol Wt**	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04											9,026	144,777	8,791	141,008
Ethane	30.07											662	19,906	645	19,395
Carbon Monoxide	28.01											15,752	441,214	11,146	312,199
Carbon Dioxide	44.01											32,126	1,413,865	35,537	1,563,983
Water	18.02	2,704	48,726	24,972	450,000	118,002	2,126,400					106,122	1,912,318	100,248	1,806,469
Hydrogen Sulfide	34.08											1,237	42,157	124	4,226
Carbonyl Sulfide	60.08											54	3,244	2	120
Ammonia	17.03											843	14,356	821	13,982
Tars	125.0											647	80,900	630	78,797
Ash		-	111,370							-	111,370				
Carbon	12.01	64,324	772,531							386	4,636				
Hydrogen	2.02	29,365	59,317									43,978	88,836	47,031	95,003
Oxygen	32.00	3,481	111,392					16,919	541,408						
Nitrogen	28.02	496	13,898					343	9,611			385	10,788	375	10,508
Sulfur	32.06	1,338	42,896												
Totals		-	1,160,130	24,972	450,000	118,002	1,126,400	17,262	551,019	-	116,006	210,832	4,172,361	205,350	4,045,690
Temperature, F		77		212		752		392		900		1,150		1,240	
Pressure, Psia						512						300		270	
Heat Content - Sensible/Latent		-0-		60.772		2,847.888		38.992		19.458		4,022.742		4,058.931	
HHV		14,193.701		-0-		-0-		-0-		65.347		12,948.606		12,343.484	
Total Heat - MM Btu/hr		14,193.701		60.772		2,847.888		38.992		84.805		16,971.348		16,402.415	

Notes: **Basis 1,000,000 lb/hr MAF coal, Illinois No. 6**
Enthalpy datum 77 F, H_2O (L)
Lurgi oxygen blown gasification - hot purification
Stream numbers refer to Fig. 4.10

Table 4-6A

CASE EAL - ABBREVIATED MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed		Gasifier Steam		Transport Gas To Gasifier		Oxidant To Gasifier		Slag		Gas From Gasifier		Net Clean Fuel Gas	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04					158	2,534					6,268	100,539	6,069	97,347
Ethane	30.07														
Carbon Monoxide	28.01					1,327	37,169					52,754	1,477,640	57,078	1,598,755
Carbon Dioxide	44.01					16	704					6,188	272,334	3,110	136,871
Water	18.02			11,124	200,454	13	234	927	16,705			4,163	75,017	4,495	81,000
Hydrogen Sulfide	34.08					3	102					1,219	41,544	109	3,715
Carbonyl Sulfide	60.08					1	60					123	7,390	20	1,202
Ammonia	17.03														
Tars	125.0														
Ash		-	111,362							-	111,362				
Sulfur Dioxide	64.06														
Carbon	12.01	64,320	772,483							488	5,861				
Hydrogen	2.02	29,364	59,315			654	1,321					25,995	52,510	25,169	50,841
Oxygen	32.00	3,481	111,392					23,164	741,248						
Nitrogen	28.02	496	13,898			2,249	63,017	87,139	2,441,635			89,488	2,507,455	80,647	2,259,729
Sulfur	32.06	1,338	42,896												
Totals		-	1,111,346	11,124	200,454	4,421	105,141	111,230	3,199,588	-	117,223	186,198	4,534,429	176,697	4,229,460
Tamperature, F		200		709		320		800		2,700		1,700		194	
Pressure, Psia				500		450		450				360		328	
Heat Content - Sensible/Latent		47.002		273.024		7.840		594.979		93.849		2,467.267		225.428	
HHV		14,193.700		-0-		303.269		-0-		82.600		12,474.542		11,669.060	
Total Heat - MM Btu/hr		14,240.702		273.024		311.109		594.979		176.449		14,941.810		11,894.488	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Entrained bed (air) gasification - cold purification
Stream numbers refer to Fig. 4.11

Table 4-6B

CASE EAH - ABBREVIATED MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed		Gasifier Steam		Oxidant To Gasifier		Slag		Gas From Gasifier		Transport Gas To Gasifier		Net Clean Fuel Gas	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04									6,262	100,442	152	2,438	5,923	95,005
Ethane	30.07														
Carbon Monoxide	28.01									52,680	1,475,567	1,253	35,097	48,854	1,368,401
Carbon Dioxide	44.01									6,350	279,464	177	7,790	7,083	311,723
Water	18.02			11,124	200,454	927	16,705			4,163	75,017	13	234	4,006	72,188
Hydrogen Sulfide	34.08									1,219	41,544	3	102	116	3,953
Carbonyl Sulfide	60.08									122	7,330	0.3	18	10	601
Ammonia	17.03									794	13,522	0.1	2	757	12,892
Tars	125.0														
Ash		-	111,362					-	111,362						
Sulfur Dioxide	64.06														
Carbon	12.01	64,320	772,483					488	5,861						
Hydrogen	2.02	29,364	59,315							25,995	52,510	654	1,321	25,554	51,619
Oxygen	32.00	3,481	111,392			23,164	741,248								
Nitrogen	28.02	496	13,898			87,139	2,441,635			89,406	2,505,156	2,167	60,719	84,556	2,369,259
Sulfur	32.06	1,338	42,896												
Totals		-	1,111,346	11,124	200,454	111,230	3,199,588	-	117,223	186,991	4,550,552	4,419.4	107,721	176,859	4,285,641
Temperature, F		200		709		800		2,700		1,700		1,400		1,400	
Pressure, Psia				500		450				360		330		330	
Heat Content - Sensible/Latent		47.002		273.024		594.979		93.849		2,468.325		7.833		1,889.991	
HHV		14,193.700		-0-		-0-		82.600		12,463.167		291.886		11,513.042	
Total Heat - MM Btu/hr		14,240.702		273.024		594.979		176.449		14,931.492		299.719		13,403.033	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Entrained bed (air) gasification - hot purification
Stream numbers refer to Fig. 4.12

Table 4-7A

CASE EXL - ABBREVIATED MATERIAL BALANCE

Stream No.		1		2		3		4		5		6		7	
		Coal Feed To Gasifier		Total Steam To Gasifier		Transport Gas To Gasifier		Oxidant To Gasifier		Slag		Gas From Gasifier		Net Fuel Gas	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04					419	6,721					7,558	121,230	7,082	113,595
Ethane	30.07														
Carbon Monoxide	28.01					2,443	68,428					44,080	1,234,681	41,300	1,156,813
Carbon Dioxide	44.01					555	24,426					15,489	681,671	9,840	433,058
Water	18.02	2,704	48,726	38,470	693,230	16	288					18,105	326,252	2,451	44,167
Hydrogen Sulfide	34.08					6	204					1,222	41,646	108	3,681
Carbonyl Sulfide	60.08					3	180					125	7,510	16	961
Ammonia	17.03											794	13,522		
Tars	125.0														
Ash		-	111,362							-	111,362				
Sulfur Dioxide	64.06														
Carbon	12.01	64,320	772,483							488	5,861				
Hydrogen	2.02	29,364	59,315			2,098	4,238					37,862	76,481	35,475	71,660
Oxygen	32.00	3,481	111,392					20,789	665,248						
Nitrogen	28.02	496	13,898			31	869	424	11,880			554	15,523	519	14,542
Sulfur	32.06	1,338	42,896												
Totals		-	1,160,072	38,470	693,230	5,571	105,354	21,213	677,128	-	117,223	125,789	2,518,516	96,791	1,838,477
Temperature, F		200		709		320				2,700		1,700		194	
Pressure, Psia				500								360		327	
Heat Content - Sensible/Latent		47.002		944.210		10.344		33.652		93.849		2,100.917		126.486	
HHV		14,193.700		-0-		718.097		-0-		86.600		13,373.254		12,132.730	
Total Heat - MM Btu/hr		14,240.702		944.210		728.441		33.652		176.449		15,474.172		12,259.216	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Entrained bed oxygen - cold purification
Stream numbers refer to Fig. 4.13

Table 4-7B

CASE EXH - ABBREVIATED MATERIAL BALANCES

Stream No.		1		2		3		4		5		6		7	
		Coal Feed		Gasifier Steam		Oxidant To Gasifier		Slag		Gas From Gasifier		Transport Gas To Gasifier		Net Fuel Gas	
Component/ Properties	Mol Wt	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr	Lb Mol/hr	Lb/hr
Methane	16.04									7,526	120,717	388	6,224	6,914	110,901
Ethane	30.07														
Carbon Monoxide	28.01									43,751	1,225,466	2,114	59,213	37,646	1,054,464
Carbon Dioxide	44.01									15,874	698,615	940	41,369	17,251	759,217
Water	18.02			38,470	693,230					18,105	326,252	16	288	15,104	272,174
Hydrogen Sulfide	34.08									1,223	41,680	7	239	118	4,021
Carbonyl Sulfide	60.08									122	7,330	0.4	24	3	180
Ammonia	17.03									794	13,522	0.4	7	730	12,432
Tars	125.0														
Ash		-	111,362					-	111,362						
Sulfur Dioxide	64.06														
Carbon	12.01	64,320	772,483					488	5,861						
Hydrogen	2.02	29,364	59,315							37,843	76,443	2,079	4,200	37,307	75,360
Oxygen	32.00	3,481	111,392			20,789	665,248								
Nitrogen	28.02	496	13,898			424	11,880			552	15,467	28	785	507	14,206
Sulfur	32.06	1,338	42,896												
Totals		-	1,111,346	38,470	693,230	21,213	677,128	-	117,223	125,790	2,525,492	5,572.8	112,349	115,580	2,302,955
Temperature, F		200		709		300		2,700		1,700		1,400		1,400	
Pressure, Psia				500		450				350		330		330	
Total Content - Sensible/Latent		47.002		944.210		33.652		93.849		2,103.223		10.564		1,508.173	
HHV		14,193.700		-0-		-0-		82.600		13,318.025		662.869		11,968.446	
Total Heat - MM Btu/hr		14,240.702		944.210		33.652		176.449		15,421.248		673.433		13,476.619	

Notes: Basis 1,000,000 lb/hr MAF coal, Illinois No. 6
Enthalpy datum 77 F, H_2O (L)
Entrained bed (oxygen) - hot purification
Stream numbers refer to Fig. 4.14

THERMAL EFFICIENCY DISCUSSION

Tables 4-8A and B summarize thermal efficiency and its contributing components for the cases studied. Table 4-9 is a summary of the results of Table 4-8 collected for convenience in a single array. Table 4-10 is a summary of the differences between hot purification and cold purification thermal efficiencies for each combination of gasifier and gas turbine inlet temperature involved in Table 4-9. Caution must be exercised in arriving at conclusions based on the above tables because substantial technical uncertainty tends to be obscured in the numerical presentation of the results, as previously discussed.

Cases MX (Lurgi oxygen blown) and MA (Lurgi air blown) show the largest incentive to apply hot fuel gas purification, as indicated by the differences in Table 4-10. The cause of the larger incentive in Cases MX and MA relates in part to the losses of power generating capacity as a result of quenching the gasifier effluent, which in turn is dictated by practical considerations. If the quenching operation could be replaced by a heat exchange scheme capable of raising high pressure steam for use in power generation, the cold purification supported efficiencies of Cases MX and MA would be increased and the incentive to adopt hot purification would be reduced in them. The effect can be seen roughly by comparing the higher efficiencies of Cases EAL 1950 and EXL 1950, where no tar is produced and the gasifier effluent can be used for raising steam, with those of Cases MAL 1950 and MXL 1950 where quenching is employed.

Bearing in mind the need for caution in comparing methods of gasification, it may be noted that MA type efficiencies (Lurgi, air blown) are generally higher than those in the MX cases (Lurgi, oxygen blown), indicating a preference for the air blown case (see Table 4-9). The higher efficiency of Lurgi air blown gasification (MAL) for combined cycle application relates to the work producing capacity of diluent nitrogen in the gas turbine. Steam is the diluent controlling gasification (rather than nitrogen) in Lurgi oxygen blown gasification (MXL). The diluent steam is largely condensed in preparing MXL type fuel gas for cold purification and it can only be partially recovered in the fuel gas by resaturation. As a result, in oxygen blown gasification, steam is less effective than the nitrogen diluent of air blown gasification in supplying working fluid to the turbine system. The fact that Case MAH efficiency is greater than Case MXH efficiency indicates that subtracting steam from the steam cycle in order to supply the increased gasifier need of MXH leads to losses of power. The latter effect can be seen comparing the contribution of the steam system to total power generation in Cases MXH and MAH in Table 4-8A.

Table 4-8A

THERMAL EFFICIENCIES - SUMMARY
1,950°F CASES
1,000,000 LB/HR MAF COAL

	MXL	MXH	MAL	MAH	MSL	MSH	EXL	EXH	EAL	EAH
Gasifier Oxidant MW	-139.1	-154.2	-168.7	-168.7	-139.9	-137.2	-171.0	-185.9	-209.3	-209.3
Gas Turbine Net MW(1)	+946.4	+1285.3	+992.5	+1287.7	+1135.8	+1168.5	+1038.5	+1142.1	+1091.0	+1227.0
Steam Turbine Net MW(2)	+262.1	+342.7	+326.0	+429.0	+521.7	+529.3	+604.7	+564.5	+698.5	+580.9
Tar Combustion @ 9000 Btu/kW	+141.2		+141.2							
Total Power MW	+1220.6	+1473.8	+1291.0	+1539.0	+1517.6	+1560.6	+1472.2	+1520.7	+1580.2	+1598.6
Heat Rate(3) Btu/kWh	11600	9600	11000	9200	9300	9100	9600	9300	9000	8900
Thermal Eff'y %	29	35	31	37	37	38	35	37	38	38

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1. Gas turbine power less combustor air compressor power.

2. Steam turbine power after gasifier steam, other process debits and credits, and auxiliary power.

3. Coal HHV is 14,194 Btu/lb (MAF basis).

Table 4-8B

THERMAL EFFICIENCIES - SUMMARY
2,400°F CASES
1,000,000 LB/HR MAF COAL

	MXL	MXH	MAL	MAH	MSL	MSH	EXL	EXH	EAL	EAH
Gasifier Oxidant MW	-139.1	-154.2	-168.7	-168.7	-139.9	-137.2	-171.0	-185.9	-209.3	-209.3
Gas Turbine Net MW(1)	+1032.0	+1414.6	+1084.6	+1413.5	+1230.9	+1266.1	+1127.5	+1243.2	+1200.5	+1347.3
Steam Turbine Net MW(2)	+302.1	+398.1	+375.6	+468.3	+554.8	+558.8	+615.6	+576.4	+706.9	+574.7
Tar Combustion @ 9000 Btu/kW	+141.2		+141.2							
Total Power MW	+1346.2	+1658.5	+1432.7	+1713.1	+1645.8	+1687.7	+1572.1	+1633.7	+1698.1	+1727.7
Heat Rate(3) Btu/kWh	10500	8600	9900	8300	8600	8400	9000	8700	8400	8200
Thermal Eff'y %	32	40	34	41	40	41	38	39	41	42

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1. Gas turbine power less combustor air compression power.

2. Steam turbine power after gasifier steam, other process debits and credits, and auxiliary power.

3. Coal HHV is 14,194 Btu/lb (MAF basis).

Table 4-9

SUMMARY OF ESTIMATED THERMAL EFFICIENCIES

PURIFICATION	COLD BENFIELD	HOT MORGANTOWN	COLD BENFIELD	HOT MORGANTOWN
Gas Turbine Inlet Temp.	1950°F	2400°F	1950°F	2400°F
	Thermal Efficiency %* Heat Rate Btu/kWh			
Lurgi (O_2)				
Thermal Efficiency	29.4	35.4	32.4	39.9
Heat Rate	11628	9630	10544	8558
Lurgi (Air)				
Thermal Efficiency	31.0	37.0	34.5	41.2
Heat Rate	10994	9223	9907	8285
Slagging (O_2)				
Thermal Efficiency	36.5	37.5	39.6	40.6
Heat Rate	9352	9095	8624	8409
Entrained Bed (Air)**				
Thermal Efficiency	38.0	38.4	40.8	41.6
Heat Rate	8982	8879	8359	8215
Entrained Bed (O_2)**				
Thermal Efficiency	35.4	36.6	37.8	39.3
Heat Rate	9641	9334	9028	8688

$$\text{*Thermal Efficiency (\%)} = \frac{\text{(Delivered kW) (3412·75) (100)}}{\text{(Coal Lb/Hr) (Coal HHV Btu/lb)}}$$

**Foster Wheeler Gasifier.

Table 4-10

SUMMARY OF THERMAL EFFICIENCY DIFFERENCES

PURIFICATION	COLD BENFIELD	HOT MORGANTOWN	COLD BENFIELD	HOT MORGANTOWN
Gas Turbine Inlet	1950°F	1950°F	2400°F	2400°F
Cases				
MX Lurgi O_2	MX 1950	= 6.1 (Note 1)	MX 2400	= 7.5
MA Lurgi Air	MA 1950	= 6.0	MA 2400	= 6.7
MS Slagging	MS 1950	= 1.0	MS 2400	= 1.0
EA Entrained/Air	EA 1950	= 0.4	EA 2400	= 0.7
EX Entrained/O_2	EX 1950	= 1.2	EX 2400	= 1.4

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1. Example MX 1950 = Thermal efficiency difference between hot and cold purification based on 1950°F gas turbine inlet temperature and taken Table 4-9.

2. Differences are calculated based on heat rates summary of Table 4-9.

Reviewing the data of Tables 4-8, 4-9 and 4-10 for the group of gasifiers studied it appears that the Morgantown iron oxide process is being developed with the type of gasifier to which it is best applied as measured by thermal efficiency improvement. The use of Morgantown hot iron oxide purification with Lurgi air blown gasification in application to combined cycle power generation results in a thermal efficiency that is as high as that in any case studied. Once again bearing in mind the need for caution, use of hot purification appears to have the function of "equalizing" gasifiers where a use exists for hot purified gas without cooling it. Apparently it provides a means by which currently commercial gasifiers may be used at near the maximum efficiency attainable with the introduction of advanced gasifier technology to combined cycle operations. The Lurgi dry ash producing type of air blown gasifier is unlikely to be superceded for some time to come and the incentive to apply hot purification will certainly remain until this happens.

Tars are not recycled to extinction in Cases MAL and MXL and are credited in the thermal efficiency computation with a value of 9,000 Btu/kWh. This credit may be optimistic because tars possibly may not be used at the same boiler efficiency as low sulfur containing heavy fuel oils. It is important to keep in mind the fact that these tars contain approximately 2 percent sulfur. Local emission control restrictions will probably require that any boiler fired with these tars be equipped with stack gas scrubbing devices. Use of tars at a somewhat higher heat rate does not alter conclusions related to Cases MAL and MAX. The tar problem would be alleviated if techniques for recycling tars to extinction were to be developed.

The slagging gasifier (Cases MSL and MSH) is shown by the data in Table 4-10 to promise a smaller incentive to apply hot purification than the dry ash producing gasifiers. The reason for this is that, although the slagging gasifier involves quenching with its irreversibilities and also produces tars, the tars are recycled to the gasifier and the steam is almost completely converted to carbon monoxide and hydrogen resulting in an extremely high cold gas efficiency. As a result the maximum loss associated with quenching has to be smaller than in the other quenched gasifier cases where the cold gas efficiency is lower. Cold gas efficiencies in the various cases are summarized as follows:

Gasifier	Cold Gas Efficiency
MX Lurgi, oxygen blown, dry ash	91%
MA Lurgi, air blown, dry ash	90%
MS slagging, oxygen blown	96%*
EX Entrained bed, oxygen blown	89%
EA Entrained Bed, air blown	86%

$$\text{where Cold Gas Efficiency} = \frac{\text{Heating value of product gases, tars and oils}}{\text{Heating value of coal feed}}$$

*It should be noted that "cold gas efficiency" as used in this study represents an efficiency of converting potential energy (HHV) in the coal to potential energy in the gas. It is not a gasifier efficiency as it does not include thermal losses associated with compressing the oxidant to gasifier pressure or the energy content of the steam feed to the gasifier. Therefore, for all cases, actual gasifier efficiency will be substantially lower than the cold gas efficiency.

The cold gas efficiency predicted for slagging gasification is substantially greater than that predicted for Lurgi dry ash gasification. As a result, the observed better efficiency of the combined cycle power unit fueled by slagging gasification and supported by cold purification is to be expected.

An additional effect increasing slagging gasification thermal efficiency is its lower consumption and almost total conversion of steam consistent with running a high slag producing temperature at the bottom of the gasifier. This leads to smaller cooling losses in Case MSL than in Case MXL on cooling the fuel gas to cold purification temperature. In addition, lower gasifier steam consumption in slagging gasification renders relatively more steam available for power generation, again tending to increase thermal efficiency. Tars are recycled to extinction in the slagging gasifier and uncertainty as to the tar heat rate and disposition of tar sulfur does not occur.

For slagging gasification with hot fuel gas purification, the previously recognized "equalizing" effect is observed. That is, the efficiencies of CASES MSH 1950 and MSH 2400 are similar to the efficiency of the other gasifier cases at corresponding gas turbine inlet temperatures (Table 4-8).

It is concluded, provided that the slagging gasifier achieves the promised high cold gas efficiency that application of hot purification is probably not justified on the grounds of combined cycle thermal efficiency improvement. This is particularly so when the uncertainties in estimated hot purification requirements are considered. Since the capital cost of a hot purification combined cycle system may be lower than that of a cold purification system, there may still be an economic incentive to apply hot purification to slagging gasification as explained in the following discussion of economics.

The entrained flow gasifiers which are the basis of Cases EA and EX show a similarly decreased incentive to adopt hot purification. In this case, the reduction in incentive does not arise as a result of a high cold gas efficiency.

The entrained flow gasifiers have lower values of this parameter than Lurgi-type gasifiers. Instead, the effect arises from the fact that the entrained flow gasifier does not produce tars, thereby enabling high level steam energy to be recovered in the cooling of gasifier effluent to cold purification temperature. With hot purification, the entrained flow gasifiers show the similar order of "equalized" high efficiency to the other types of gasifier discussed earlier.

Differences between Cases EA and EX can be understood in terms of the earlier diluent steam related discussions. Use of diluent steam to control gasification in Case EX leads to loss of turbine driving medium as it is condensed in cooling gasifier effluent to cold purification temperature. Also, use of diluent steam in the gasifier of Case EX has the net effect of reducing the availability of steam for the generation of power, with a resulting reduction in combined cycle thermal efficiency to below that in Case EA.

The thermal efficiency of the EX cases was determined using the methods illustrated for Case MS. Because of constraints on time and budget, more approximate methods were used for estimating the thermal efficiency of the EA cases. Using approximate methods it is difficult to distinguish between the efficiency of Cases EAH 1950 and EAL 1950 and both are shown as being at the 38 percent level.

The greatest uncertainty in these thermal efficiency evaluations which may be detrimental to the conclusions is the thermal cost of iron oxide regeneration, as discussed previously.

Uncertainties as a result of assumptions which cancel in calculating differences are less serious and easier to recognize. A large error in the estimated thermal cost of regeneration as a result of "State of the Art" misjudgment could erase any advantage hot purification may have in some cases. In the Lurgi cases it is unlikely that an error large enough to cancel the apparent advantages of hot purification can be made.

Factorization of the components of thermal efficiency difference between air and oxygen blown gasification in order to understand them is difficult to achieve. This discussion ascribes the difference to two major causes:

- Losses of gas flow as the diluent steam added to oxygen blown gasifiers is condensed in cooling the gasifier effluent to cold purification temperature.
- Loss of the power generating capacity of the additional steam which must be added to the oxygen blown gasifier in order to control gasification.

Other factors influence the analysis such as the difference between air and oxygen plant feed compression energy requirements and whether or not the gasifier produces slag. Answers to many questions in this area are suggested in the results obtained based on gasifier data supplied by EPRI. However, the results obtained are based on point operating data and are only suggestive of the differences to be expected between air and oxygen blown gasification for Illinois No. 6 coal in application to combined cycle power generation.

Section 5

ECONOMICS

BACKGROUND OF ECONOMIC STUDY

An approximate economic survey comparing combined cycle economics in which hot and cold purification are practiced has been prepared. A general study of the cases covered in the thermal efficiency studies was considered to be necessary since there is no incentive to pursue thermal efficiency increase if the capital cost of achieving it would increase the price of electricity. Also, where there is negligible thermal efficiency advantage, capital cost incentives to apply hot purification may exist such that the cost of electricity can be reduced. The range of possible situations leads to the conclusion that it is not possible to eliminate cases from economic studies on thermal efficiency grounds. However, while hot purification technology is under development and the objective is more one of relative situation recognition than of absolute cost determination, it is reasonable that the economic survey be approximate.

A comprehensive study of the economics of fuel gas production is being developed by Fluor Engineers & Constructors, Inc. (24), under contract to EPRI, employing cold Selexol purification and all of the gasifiers of interest in this study.* The Fluor data have been employed to provide the consistent reference basis for the order of magnitude economics developed in this study. The Fluor cost data have an estimated accuracy of ±40 percent and are unsatisfactory for the prediction of small order economic effects. Other economic studies have been published by Ahner (1) and by Robson (21) which could have been used in a similar way.

The capital cost of the fuel gas production facilities in each cold purification case has been determined by prorating Fluor data to the coal feed rate corresponding to 1,000 MW capacity. The required adjustment of the Fluor data to the 1,000 MW capacity was generally small. Integrated combined cycle capital costs have been added to the fuel gas production costs in order to establish the total plant cost. By combining capital return factors with operating costs, the cost of electrical energy production was estimated.

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*Except the slagging gasifier which is taken to be equal in cost to the dry ash producing gasifier.

Electrical energy costs in hot purification cases were derived by substituting hot purification capital costs for the cold purification and related cost information in the Fluor data. The cost of electrical energy was subsequently determined as noted above for the cold purification case.

The same combined cycle unit capital cost ($200 KW) was employed for the cases of 1,950°F and 2,400°F gas turbine inlet temperature. Thus, the difference between corresponding case economics (e.g., MSL 1950 and MSL 2400, MAH 1950 and MAH 2400, etc.) is principally due to thermal efficiency differences between the cases.

Formal study of the sensitivity of conclusions to variation of the component plant capital costs has not been carried out in this study. However, the economic results are presented so that any particular sensitivity can be investigated. The cost component of most interest, hot purification, is subject to a high level of uncertainty. This, however, is not a limitation on the ability to arrive at tentative conclusions, as discussed in Section 5, page 5-12.

Criteria used in the development of economics are as follows:

Economic Criteria for Hot vs. Cold Gas Cleaning Study

General Economic Factors

- Estimate based on mid-1975 dollars
- Chicago area plant location
- 70% operating load factor
- Debt to equity ratio = 50/50
- Return on equity after taxes = 12% compounded semiannually
- Bond interest = 8% compounded semiannually
- 25 year bond life
- Income tax rate = 52%
- Straight line depreciation
- 25 year plant life
- Interest during construction = 10% compounded quarterly
- 36 month construction period for all cases
- Capital expenditures disbursed equally over the construction period
- Illinois sales tax = 5% of total materials
- Paid-up royalties = 2.5% of plant investment
- Raw makeup water = 20¢ per 1,000 gallons
- Administrative and support labor = 30% of operating plus maintenance labor
- General and administrative expenses = 2% of plant investment

- Ash disposal cost = \$1.00 per ton
- Clear and level site
- Annual property taxes and insurance = 2.5% of plant investment
- 50% labor burden
- Annual maintenance expense = 4.5% of onsite and 2.5% of offsite plant investment
- Maintenance labor:materials split
  - Onsite units = 40:60
  - Offsite units = 50:50
- No credit for sulfur or ammonia

Preproduction and Start-up Costs

The sum of:

- One month variable operating costs excluding the cost of coal
- Two months fixed costs excluding income taxes
- 25% of full load coal costs for one month
- 2% of total plant investment

Working Capital

The sum of:

- 1.5 months of accounts receivable (gross income)
- One month of coal supply

Accounts receivable would be equivalent to total cost of services after all byproduct credits have been removed, plus the byproduct credits.

Contingency

For consistency with the work of EPRI, a contingency of 15 percent has been added to the estimated capital requirements. The philosophy for inclusion of this term is to maintain consistency in EPRI work rather than to take care of specifiable uncertainty.

Accuracy

The accuracy of the reference work from which subject economics are prorated is ±40 percent (24). In view of the prorating procedure used, the high uncertainty in hot purification costs, the assumption of costs for unavailable combined cycle power generation systems, and other uncertainties, it is concluded that the accuracy of the economics presented in this report is not specifiable. Order of magnitude economics are presented for the purpose of identifying important differences which are significant in identifying "best" possible future applications for hot purification. Absolute values of prices, investments, etc., as quoted, are subject to all of the previously stated adjustments in order to attempt to orientate the reader realistically in combined cycle economics.

Catalyst and Chemicals

No specific allowance has been made for the cost of the supported iron oxide sulfur absorbent in the hot purification cases as the material is not commercially available. Catalyst and chemicals charges in the hot purification cases are held in the same ratio to capital investment as in the cold purification cases. The conclusions are not sensitive to these charges. The dominant operating cost is that of coal which is set by thermal efficiency.

Operating Labor Requirements

Operating labor costs are generally based on data of Fluor. Conclusions are not sensitive to these charges. The dominant operating cost is that of coal which is set by thermal efficiency.

In order to establish approximate overall capital costs in the hot purification area and other areas not covered by the Fluor study, the following information was employed:

Sulfur Absorbent Vessels

Costs of desulfurization vessels assume refractory lines steel construction. Sufficient absorbent volume is provided for the operating cycle to be of 8 hr duration based on a sulfur saturation level (pick-up) of 10 percent of the fresh absorbent charge. The cost per absorbent vessel is estimated to be of the order of $500,000 installed; each vessel is 13.5 ft internal diameter, refractory lines, containing a 15 ft deep bed of absorbent. The provision of an internal stainless cladding between the vessel wall and the refractory lining would increase the cost of each vessel by about 40 percent.

Design conditions upon which quoted costs are based are as follows:

	Operation	Regeneration
Pressure, psig	345	250
Temperature, F	1,400	1,600 (max)
H_2 psig (partial)	70	0
H_2S (% vol)	1.5	0
SO_2 (% vol)	0	90.0

For developing the basic economics (Tables 5-1 and 5-2), all-steel sulfur absorbent vessel construction is assumed. Vessel design is discussed in Section 2.

The design sulfur pick-up level may be set by the process developer to be lower than 10 weight percent used as the basis for this work.

Particulate Removal Equipment

Costs of particulate removal equipment for the proposed service are subject to substantial uncertainty, and the problem becomes one of making an informed allowance for such facilities in the economic investigations. Estimated costs of particulate removal equipment, at a unit capacity of 10,000 ACFM and based mainly on pricing the required pressure containment vessels for operating conditions of 300 psig and 1,200°F are as follows. No cost of development allowance is made in the following cost approximations:

- Cyclones

  Two-stage refractory lined

  Separate lockhopper for each stage

Installed cost	- cyclones	\$40 ACFM
	- lockhoppers	\$30 ACFM
	Total	\$70/ACFM

- Brunswick Felted Metallic Cloth Filter

  Cost - \$100/ACFM

  Design: Panels of metallic cloth supported in an unlined pressure vessel. Use of an unlined pressure vessel is preferable because of possible gas turbine problems which result from spalling of refractory.

- Note: A metallic cloth suitable for application is not developed.
- Panel Bed Filters (CCNY)

  Cost - \$100/ACFM

  Design: Multiple panels in a refractory lined pressure vessel.

Two of the foregoing filters are required in a Morgantown-type iron oxide unit, for protecting the absorbent bed against plugging with dust entrained from the gasifier and for protection of the gas turbine. The filter capacity for cost estimation purposes was set at 10,000 ACFM which corresponds to the volumetric output of a typical moving bed gasifier.

The actual equipment to be provided for particulate collection is a function of the magnitude and nature of the removal problem which is largely unknown. For example, specific filter units for gas turbine protection may not be required if the sulfur absorbent beds themselves are found to provide sufficient filtering action.

The judgment made for developing economics in this study was to provide a filter equipment cost allowance of $200/ACFM. This allowance leads to a filter equipment cost of the same order as that of the Benfield or Selexol unit in the cold process configuration. Physical situations can be imagined where the above filter cost factor varies from half to twice the quoted value of $200/ACFM and scope for further study exists in this area.

Sulfur Dioxide Reduction

The cost of sulfur dioxide reduction and tail gas incineration is estimated to be approximately $13.0 MM at 1,000 MW combined cycle capacity based on Illinois No. 6 coal. The above sulfur dioxide reduction plant capital cost corresponds to using process gas for sulfur dioxide reduction. Based on using coal for sulfur dioxide reduction, a lower but similar order capital requirement is predicted.

Sulfur Dioxide Recycle Regeneration

An allowance of $10 MM is made for the cost of heat exchange and sulfur dioxide circulation equipment connected with hot sulfur absorbent regeneration. This cost is arrived at by approximately sizing and costing the heat exchange equipment required for the purpose. Considerable technical uncertainty exists in the regeneration equipment costs. However, the situation is such that at their expected level these costs are believed not to have significant influence on the economic comparison between hot and cold purification.

The allowance made should be more than sufficient to cover the cost of air regeneration equipment, if air regeneration is proved to be feasible, and provided that a sulfur dioxide concentration unit is not required.

Fired Tar Boilers

For the small (ca 100 MW) tar fired boilers with stack gas scrubbing in cases MAL and MXL, a capital allowance of $1,000/kW is made.

Cold Purification Costs

Cold purification cost data used as the reference for this work were developed for EPRI by Fluor and are based on application of the Selexol purification process.

Cost studies in support of this project indicated that at the ±40 percent accuracy levels involved, the Benfield and Selexol process capital requirements are similar enough to be interchangeable in the economic evaluation without alteration of conclusions with respect to hot purification.

Cold purification process performance data in thermal efficiency evaluations were based on information provided by Benfield Corporation.

ECONOMICS RESULTS AND DISCUSSION

Results of Economic Study

Summaries of component and total capital costs for each of the cases included in the thermal efficiency studies are presented in Tables 5-1A and B; Table 5-1A is based on 1,950°F gas turbine inlet temperature and Table 5-1B on 2,400°F gas turbine inlet temperature. Tables 5-2A and B summarize the estimated delivered price of electrical power corresponding to the capital costs of Tables 5-1A and B and the thermal efficiencies of Tables 4-11A and B, respectively. The general procedure for developing these costs was explained previously (Section 5, page 5-1). The price of electricity in Tables 5-2A and B represents an averaged price during the depreciable 25 year life of the project. Subsequent tables summarize results in more concise form.

Table 5-3 summarizes estimated electrical power costs (mills/kWh) based on Illinois No. 6 coal at a coal cost of $1.0/MM Btu. Corresponding power cost based on a coal cost of $2/MM Btu is indicated in Tables 5-2A and B.

Table 5-4 summarizes estimated total capital requirements ($/kW) for each case of interest.

Table 5-5 summarizes the plant investment cost allowance for purification and related gas cooling facilities incorporated in the capital cost summary of Tables 5-4. In Table 5-5 cold purification costs are based on data prorated from information developed by Fluor. Hot purification costs have been estimated by Stone & Webster based on allowances discussed previously in Section 5.

Table 5-6 is a special purpose summary of the plant investment costs of equipment which is deleted from the cold purification flow scheme as a result of the introduction of hot purification.

Table 5-1A

SUMMARY OF ESTIMATED COMPONENT CAPITAL COSTS 1,950°F GAS TURBINE INLET TEMPERATURE

Case		MXL	MXH	MAL	MAH	MSL	MSH	EXL	EXH	EAL	EAH
Output Mw (100%)		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
Coal M Lb/Hr		950.4	787.1	898.6	753.8	764.4	743.4	788.0	762.9	734.1	725.7
Coal Preparation	$MM	14.189	11.751	13.415	11.254	11.412	11.098	11.764	11.389	10.960	10.834
Gasification	$MM	69.932	57.928	66.121	55.471	25.904	25.519	23.358	25.000	28.555	32.254
Gas Cooling/Heating	$MM	53.793	1.543	34.275	5.166	7.289	4.194	24.436	8.065	28.387	7.400
Oxygen Unit	$MM	69.644	63.934	-	-	56.322	53.719	70.958	74.704	-	-
Air Compression	$MM	-	-	28.343	23.776	-	-	-	-	28.772	28.391
Water Treatment	$MM	79.836	-	75.482	-	26.602	-	7.943	-	7.400	-
Cold Purification (1)	$MM	37.840	-	47.338	-	26.753	-	31.373	-	38.003	-
Hot Purification	$MM	-	73.021	-	72.968	-	52.349	-	56.148	-	61.386
Combined Cycle	$MM	176.900	200.000	178.146	200.000	200.000	200.000	200.000	200.000	200.000	200.000
Offsites	$MM	109.204	70.480	97.389	63.182	61.383	60.095	63.225	64.059	58.487	58.173
Tar Fired Power Unit	$MM	115.658		109.369							
Subtotal	$MM	726.996	478.656	649.878	431.817	415.665	406.974	433.057	439.365	400.514	398.438
Contingency	$MM	109.050	71.798	97.480	64.772	62.350	61.046	64.959	65.905	60.077	59.766
Plant Investment	$MM	836.050	550.454	747.358	496.589	478.015	468.020	498.016	505.270	460.591	458.204
Illinois Sales Tax	$MM	20.901	13.761	18.684	12.415	11.950	11.701	12.450	12.632	11.515	11.455
Preproduction Costs	$MM	48.421	32.022	43.397	28.983	27.977	27.380	29.129	29.472	26.951	26.801
Royalties	$MM	20.901	13.761	18.684	12.415	11.950	11.701	12.450	12.632	11.515	11.455
Catalyst & Chemicals	$MM	4.180	2.752	3.737	2.483	2.390	2.340	2.490	2.526	2.303	2.291
Construction Interest	$MM	144.168	94.920	128.874	85.632	82.429	80.705	85.878	87.129	79.424	79.013
Total Depreciable Capital	$MM	1,074.621	707.672	960.734	638.516	614.712	601.847	640.414	649.660	592.299	589.218
Working Capital	$MM	42.948	30.859	39.168	28.520	28.057	27.390	29.102	28.945	26.995	26.781
Total Capital	$MM	1,117.569	738.531	999.902	667.036	642.769	629.236	669.516	678.606	619.294	615.999
$/KW Plant Investment		836	550	747	497	478	468	498	505	461	458
$/KW Total Capital		1,117	739	1,000	667	643	629	670	679	619	616

Table 5-1B

SUMMARY OF ESTIMATED COMPONENT
CAPITAL COSTS 2,400°F GAS TURBINE INLET TEMPERATURE

Case		**MXL**	**MXH**	**MAL**	**MAH**	**MSL**	**MSH**	**EXL**	**EXH**	**EAL**	**EAH**
Output, Mw (100%)		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
Coal, M Lb/Hr		861.8	699.5	809.9	677.2	704.9	687.4	737.9	710.1	683.2	671.5
Coal Preparation	$MM	12.865	10.443	12.088	10.110	10.523	10.262	11.017	10.601	10.199	10.024
Gasification	$MM	63.421	54.902	59.556	53.118	28.837	21.681	26.683	23.580	28.347	29.870
Gas Cooling/Heating	$MM	48.784	1.370	30.872	4.637	6.720	3.712	22.886	7.507	26.419	6.853
Oxygen Unit	$MM	63.146	56.814	-	-	51.935	49.674	66.449	69.537	-	-
Air Compression	$MM			25.540	21.359						
Water Treatment	$MM	72.387	-	68.017	-	24.530	-	7.438	-	6.886	-
Cold Purification	$MM	34.309	-	42.656	-	24.669	-	29.379	-	35.365	-
Hot Purification	$MM	-	65.445 -	-	66.061	-	48.783	-	52.609	-	57.172
Combined Cycle	$MM	179.000	200.000	180.300	200.000	200.000	200.000	200.000	200.000	200.000	200.000
Offsites	$MM	102.241	67.035	91.119	60.794	60.092	57.849	62.099	62.018	57.011	56.359
Tar Fired Power Unit	$MM	104.918	-	98.539							
Subtotal	$MM	681.071	456.009	608.687	416.078	407.216	391.961	425.951	425.852	390.954	386.547
Contingency	$MM	102.160	68.401	91.300	62.412	61.082	58.794	63.893	63.878	58.643	57.982
Plant Investment	$MM	783.231	524.410	699.987	478.490	468.299	450.755	489.844	489.730	449.597	444.529
Illinois Sales Tax	$MM	19.581	13.110	17.497	11.962	11.707	11.269	12.246	12.243	11.240	11.113
Preproduction Costs	$MM	45.293	30.392	40.569	27.814	27.531	26.306	28.567	28.499	26.232	25.927
Royalties	$MM	19.581	13.110	17.497	11.962	11.707	11.269	12.246	12.243	11.240	11.113
Catalyst & Chemicals	$MM	3.916	2.622	3.500	2.392	2.341	2.254	2.449	2.449	2.248	2.223
Construction Interest	$MM	135.060	90.429	120.708	82.511	80.753	77.728	84.469	84.449	77.528	76.655
Total Depreciable Capital	$MM	1,006.662	674.074	899.758	615.132	602.338	579.580	629.821	629.613	578.086	571.560
Working Capital	$MM	39.781	28.594	36.175	26.700	26.787	25.920	28.030	27.588	25.818	25.461
Total Capital Requirement	$MM	1,046.443	702.668	935.933	641.832	629.125	605.500	657.851	657.201	603.904	597.020
$/KW Plant Investment		783	524	700	478	468	451	490	490	450	445
$/KW Total Capital		1,046	703	936	642	629	606	658	657	604	597

Table 5-2A

ESTIMATED COST OF ELECTRICAL POWER AT 1,950°F GAS TURBINE INLET TEMPERATURE AND 70 PERCENT LOAD FACTOR

Case	**MXL**	**MXH**	**MAL**	**MAH**	**MSL**	**MSH**	**EXL**	**EXH**	**EAL**	**EAH**
Output (70%), Mw	700	700	700	700	700	700	700	700	700	700
Heat Rate, Btu/KWHR	11,628	9,630	10,994	9,223	9,352	9,095	9,641	9,334	8,982	8,879
Total Capital Requirement,$MM	1,117.569	738.531	999.902	667.036	642.769	629.236	669.516	678.606	619.294	615.999
Operating Charges $MM/Yr										
Coal at $1/MM Btu	71.303	59.051	67.415	56.555	57.346	55.770	59.119	57.236	55.078	54.446
Operating Labor	9.196	6.055	8.221	5.462	5.258	5.148	5.478	5.558	5.066	5.040
Catalyst and Chemicals	0.585	0.385	0.523	0.348	0.335	0.328	0.349	0.354	0.322	0.321
Maintenance Labor	13.795	9.082	12.331	8.194	7.887	7.722	8.217	8.337	7.600	7.560
Materials	19.229	12.660	17.189	11.422	10.994	10.764	11.454	11.621	10.594	10.539
Administrative and Support Labor	6.897	4.541	6.166	4.097	3.944	3.861	4.109	4.168	3.800	3.780
General and Admin. Expense	16.721	11.009	14.947	9.932	9.560	9.360	9.960	10.105	9.212	9.164
Ad Valorem Taxes and Ins.	20.901	13.761	18.684	12.415	11.950	11.701	12.450	12.632	11.515	11.455
Utilities	0.175	0.253	0.092	0.156	0.037	0.030	0.046	0.066	0.024	0.018
Ash Disposal	0.279	0.231	0.265	0.221	0.225	0.219	0.232	0.225	0.216	0.213
Total Operating Charge										
Capital Charges $MM/Yr										
Depreciation	42.985	28.307	38.429	25.541	24.588	24.074	25.616	25.986	23.692	23.569
Avg. Bond Interest, 8%	22.800	15.066	20.398	13.608	13.112	12.836	13.658	13.844	12.633	12.566
Avg. Return on Equity, 12%	34.198	22.599	30.597	20.411	19.669	19.255	20.487	20.765	18.950	18.850
Avg. Income Tax, 52%	37.036	24.475	33.137	22.106	21.301	20.853	22.188	22.489	20.523	20.414
Total Capital Charge										
Cost of Services	296.073	207.477	268.394	190.466	186.208	181.922	193.364	193.387	179.226	177.935
Cost of Power, Mills/KWHR Coal = $1/MM Btu	48.3	33.8	43.8	31.1	30.4	29.7	31.5	31.5	29.2	29.0
Cost of Power,Mills/KWHR Coal = $2/MM Btu	60.2	43.7	55.0	40.5	39.9	39.0	41.4	41.1	38.4	38.1

Table 5-2B

ESTIMATED COST OF ELECTRICAL POWER AT 2,400°F GAS TURBINE INLET TEMPERATURE AND 70 PERCENT LOAD FACTOR

Case	**MXL**	**MXH**	**MAL**	**MAH**	**MSL**	**MSH**	**EXL**	**EXH**	**EAL**	**EAH**
Output (70%), Mw	700	700	700	700	700	700	700	700	700	700
Heat Rate, Btu/KWHR	10,544	8,558	9,907	8,285	8,624	8,409	9,028	8,688	8,359	8,215
Total Capital Requirement $MM	1,046.443	702.67	935.933	641.83	629.125	605.5	657.85	657.20	603.90	597.02
Operating Charges $MM/Yr										
Coal at $1/MM Btu	64.656	52.478	60.750	50.804	52.882	51.564	55.360	53.275	51.257	50.374
Operating Labor	8.616	5.768	7.700	5.263	5.151	4.958	5.388	5.387	4.946	4.890
Catalyst and Chemicals	0.548	0.367	0.500	0.335	0.328	0.316	0.343	0.343	0.315	0.311
Maintenance Labor	12.923	8.653	11.550	7.895	7.727	7.437	8.082	8.080	7.418	7.335
Materials	18.014	12.061	16.100	11.005	10.771	10.367	11.266	11.264	10.341	10.224
Administrative and Support Labor	6.462	4.326	5.775	3.948	3.863	3.719	4.041	4.040	3.709	3.667
General and Admin. Expense	15.665	10.488	14.000	9.570	9.366	9.015	9.797	9.795	8.992	8.890
Ad Valorem Taxes and Insurance	19.581	13.110	17.497	11.962	11.707	11.269	12.246	12.243	11.240	11.113
Utilities	0.159	0.225	0.082	0.140	0.035	0.028	0.043	0.062	0.023	0.016
Ash Disposal	0.254	0.206	0.238	0.199	0.208	0.202	0.217	0.209	0.201	0.198
Capital Charges $MM/Yr										
Depreciation	40.266	26.963	35.990	24.605	24.094	23.183	25.193	25.185	23.123	22.862
Avg. Bond Interest, 8%	21.347	14.334	19.093	13.093	12.834	12.352	13.420	13.407	12.320	12.179
Avg. Return on Equity, 12%	32.021	21.502	28.640	19.640	19.251	18.528	20.130	20.110	18.479	18.269
Avg. Income Tax, 52%	34.679	23.286	31.017	21.270	20.849	20.066	21.801	21.780	20.013	19.785
Cost of Services	275.191	193.769	248.932	179.730	179.066	173.005	187.328	185.180	172.378	170.115
Cost of Power, Mills/KWHR Coal = $1/MM Btu	44.9	31.6	40.6	29.3	29.2	28.2	30.6	30.2	28.1	27.7
Cost of Power, Mills/KWHR Coal = $2/MM Btu	55.7	40.4	50.7	37.7	38.0	36.8	39.8	39.1	36.7	36.1

Economics Discussion

Examination of the summaries of Tables 5-3 and 5-4, which are abstracted from the more general Tables 5-1A and B and 5-2A and B, indicates that dry ash, oxygen and air blown Lurgi-type gasification shows the greatest benefit from the replacement of cold with hot purification in the combined cycle power plant. Thus, the air blown, dry ash, Lurgi based process with cold purification (Benfield or similar scrubbing type) produces power at estimated costs of 44 mills/kWh (coal at $1/MM Btu).* With hot purification, the same gasifier produces power at an estimated cost of 32 mills/kWh (coal at $1/MM Btu).* Similar differences are estimated for the oxygen blown gasifier, though the actual power cost in the oxygen cases is estimated to be higher than that in the air blown cases.

In the Lurgi dry ash cases, the cost reductions associated with application of hot fuel gas purification are in large measure the result of the deletion of equipment necessary in the cold purification case but not in the hot. The magnitude of the deleted equipment costs in each case is summarized in Table 5-6. Thus, a tar fired steam power system (with stack gas scrubbing) and a large water treatment facility are not required in the Lurgi hot purification cases, since all tars and water vapor pass uncondensed to the gas turbine combustor with the fuel gas. Additional cost savings occur with the use of hot fuel gas purification in the Lurgi dry ash cases due to the significantly higher thermal efficiency which leads to additional coal feed savings and equipment savings (e.g., to reduced gasifier costs). The net effect of thermal efficiency related savings together with the equipment deletions is to establish the incentive to apply hot purification shown in Table 5-3 for the Lurgi dry ash cases.

Slagging gasification is estimated to show a small cost saving in terms of reduced capital requirements (Table 5-4) and reduced cost of power (Table 5-3) with the application of hot purification. In the cold purification slagging cases, tar is recycled and gasified and the thermal efficiency increase which results from application of hot purification with slagging gasification is smaller than in the Lurgi cases. These effects contribute to the smaller cost savings associated with the application of hot purification to slagging gasification. The use of hot purification in association with slagging gasification does, however, result in a water treatment cost saving.

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*Referring to 1,950°F gas turbine inlet temperature.

Table 5-3

SUMMARY OF ESTIMATED COSTS OF ELECTRICITY FOR GASIFICATION-COMBINED CYCLE POWER PLANTS OF 1,000 MW CAPACITY

PURIFICATION	COLD BENFIELD	HOT MORGANTOWN	COLD BENFIELD	HOT MORGANTOWN
Gas Turbine Inlet Temperature, F	1,950	1,950	2,400	2,400
	Power Costs, mills/kWh*			
Lurgi (O_2)	48.3 (60.2)**	33.8 (43.7)	44.9 (55.7)	31.6 (40.4)
Lurgi (Air)	43.8 (55.0)	31.1 (40.5)	40.6 (50.7)	29.3 (37.7)
Slagger (O_2)	30.4 (39.9)	29.7 (39.0)	29.2 (38.0)	28.2 (36.8)
Entrained*** (Air)	29.2 (38.4)	29.0 (38.10)	28.1 (36.7)	27.7 (36.1)
Entrained*** (O_2)	31.5 (41.4)	31.5 (41.1)	30.5 (39.8)	30.2 (39.1)

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*Based on: Delivered coal costs of $1.00/MM/Btu.
Mid-1975 dollars with no escalation included and an operating load factor of 70 percent.

**Numbers in parentheses represent Power Costs related to a delivered coal cost of $2.00/MMBtu.

***Foster Wheeler gasifier.

Table 5-4

SUMMARY OF ESTIMATED CAPITAL REQUIREMENTS** FOR 1,000 MW GASIFICATION - COMBINED CYCLE POWER PLANTS

PURIFICATION	COLD BENFIELD	HOT MORGANTOWN	COLD BENFIELD	HOT MORGANTOWN
Gas Turbine Inlet Temp., F	1,950	1,950	2,400	2,400
	Capital Requirements $/kW*			
Lurgi (O_2 Dry Ash)	1,117	739	1,046	703
Lurgi (Air Dry Ash)	1,000	667	936	642
Slagging (O_2)	643	629	629	606
Entrained Bed (Air)***	619	616	604	597
Entrained Bed (O_2)***	670	679	658	657

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*All Capital Estimates are based on mid-1975 dollars with no escalation.

**Capital Requirements ($/KW) correspond to Plant Investment with adjustment for Illinois Sales Tax, Preproduction Costs, Royalty Payments, Initial Catalyst and Chemicals, Construction Loan Interest and Working Capital - summarized from Tables 5-1A and B. The Capital Requirements tabulated above represent an increase of 34 percent over the Plant Investment estimates due to inclusion of these allowances.

***Foster Wheeler gasifier.

Table 5-5

SUMMARY OF ESTIMATED CAPITAL REQUIREMENTS FOR
THE PURIFICATION SYSTEM COMPONENTS
OF EACH CONFIGURATION INVESTIGATED

PURIFICATION	COLD BENFIELD **	HOT MORGANTOWN ***	COLD BENFIELD **	HOT MORGANTOWN ***
Gas Turbine Inlet Temp., F	1,950	1,950	2,400	2,400
	Capital Requirements $/kW*			
Lurgi (O_2)	141	115	129	103
Lurgi (Air)	126	121	114	109
Slagging (O_2)	52	87	48	80
Entrained (Air)	102	106	95	99
Entrained (O_2)	86	99	80	93

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*Based on mid-1975 dollars with no escalation included.

**Cold purification costs include the costs of the H_2S scrubbing system, the Claus plant, the tail gas treating system as well as crude gas cooling equipment.

***Hot purification costs include the costs of the iron oxide systems, the SO_2 recycle and sulfur recovery systems, and the high temperature particulate removal devices upstream and downstream of the iron oxide system.

Table 5-6

EQUIPMENT DELETION SAVINGS
SUMMARY OF TAR FIRED BOILER PLUS WATER TREATMENT CAPITAL COST COMPONENTS OF COMBINED CYCLE CAPITAL REQUIREMENTS WHICH ARE SUMMARIZED IN TABLE 5-4

PURIFICATION	COLD BENFIELD	HOT MORGANTOWN	COLD BENFIELD	HOT MORGANTOWN
Gas Turbine Inlet Temp., F	1,950	1,950	2,400	2,400
Capital Cost, $/kW				
Lurgi (O_2 Dry Ash)	225	0	204	0
Lurgi (Air Dry Ash)	212	0	192	0
Slagging*	27	0	25	0
Entrained Bed (Air)**	8	0	8	0
Entrained Bed (O_2)**	8	0	8	0

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*Tar recycled to gasifier.

**No tars produced, water treatment requirements in the entrained bed gasification cases are insignificant compared to other cases.

All costs are in mid-1975 dollars with no escalation.

Above quoted costs are investment costs including 15 percent contingency as derived from Tables 5-1A and B. They include no allowance for Illinois Sales Tax, Preproduction Costs, etc., as explained in relation to determination of Capital Requirements in the notes of Table 5-4.

The lower capital requirements of the slagging gasification based combined cycle unit compared to the corresponding Lurgi unit is a result of equipment savings associated with the higher thermal efficiency of the slagging case of the different equipment deletions situation, and of reduced gasifier requirements. The slagging gasifier is predicted to have three to four times the coal processing capacity of a similar diameter Lurgi gasifier and a lower cost gasification area is, therefore, predicted for the slagging case as shown in Tables 5-1A and B.

If the cost of the required hot purification equipment is higher than that estimated and if the thermal penalty associated with hot purification regeneration is shown by development to be greater than that employed for this study, it is conceivable that the small advantage shown for hot purification in Table 5-3 could disappear or become a disadvantage in the case of the slagging gasifier.

The gas production temperature from the slagging gasifier is predicted to be lower (870°F) than that from the Lurgi dry ash producing gasifiers (1,000°F plus), and the related problem of preventing tar condensation in the hot sulfur absorbent bed was discussed previously. This is an important unresolved technical question in proposing to link hot purification to slagging gasification. In addition, slagging gasification produces a gas with a higher concentration of reducing components (CO and H_2) which may impair the effectiveness of the sulfur absorbent through chemical reaction equilibrium effects.

For the entrained bed gasification cases, the thermal efficiency improvement associated with replacement of cold with hot purification is small. In addition, entrained bed gasifiers should not produce a tar related water contamination problem and significant equipment deletions are not possible with the application of hot purification. The conclusion is that the difference between hot and cold purification combined cycle economics is too small and the uncertainties in them too large to be interpreted as presenting an incentive to apply hot purification in the entrained bed cases. The attractiveness of entrained bed, cold purification, combined cycle economics is apparent based on projections from the current stage of gasifier development. However, previous cautions should be remembered regarding absolute costs being unreliable and the importance of arriving at conclusions on the basis of difference between corresponding hot and cold purification processes.

The problem introduced because of uncertainty in the capital investment requirements of hot purification calls for review of sensitivity of power cost to hot purification investment. With the economic criteria of this study, at the investment levels of Table 5-5, hot purification capital expenditure contributes typically between 1.0 and 2.0 mills/kWh to the cost of electrical power. Therefore, increasing the investment allowance for hot purification by 50 percent above that in Table 5-5 will increase the cost of power in the hot purification cases by between 0.5 and 1.0 mills/kWh, which is insufficient to change the relative status of hot and cold purification in Lurgi cases.

The provision of a separate tar fired power generation unit may be considered to be an unfair burden on the Lurgi dry ash cases. The effect of exporting tar rather than providing the tar burning unit is outlined in the following tabulation. The tabulation shows that, if tar products are exported at \$1/MM Btu credit value (equal to the price of coal), the corresponding adjusted cost of power (49 mills/kWh) is very close to the price in Table 5-2A (48 mills/kWh). If the value of tar is doubled, there is a reduction in adjusted cost of power of approximately 1 mill/kWh, but the situation is basically unaltered - essentially no change in the cost of power. There is, however, a considerable reduction in required capital where tar is exported as a byproduct. The economics of exporting tar will depend on the price that is established for it. Tar could be a valuable source of aromatic chemicals.

	Case	MXL	MAL
	Gas Turbine Inlet Temp., F	1,950	1,950
		\$MM	\$MM
1.	Base Case Plant Investment (Table 5-1A)	726.996	649.878
2.	Tar Fired Power Unit (Table 5-1A)	115.658	109.369
	Adjusted Plant Investment (1)-(2)	611.338	540.509
	Adjusted Capital Charges	115.300	98.000
	Adjusted Operating Charges	145.000	130.300
	Tar Credit (\$/MM Btu)	(6.4)	(6.100)
	Adjusted Cost of Services	253.9	222.2
	Adjusted Capacity (6,132 hr/yr)	884 MW	890 MW
	Adjusted Cost of Power Coal = \$1/MM Btu	47 mills/kWh	41 mills/kWh
	Base Case Cost of Power (Table 1-5A) Coal = \$1/MM Btu	48 mills/kWh	41 mills/kWh

The results suggest that, in developing hot purification with dry ash air blown, Lurgi type gasification, Morgantown is investigating the application with the greatest potential for economic gain in combined cycle electrical energy generation. The same conclusion is suggested by the thermal efficiency study.

One of the surprising results of this study is the relatively minor incentive predicted for the development of the 2,400°F gas turbine as shown in Table 5-3. This table indicates that for all cases except the dry ash Lurgi, the economic benefit of changing turbine inlet temperature from 1,950 to 2,400°F is in the range of 1-2 mills/kWh, irrespective of the purification temperature employed. For the Lurgi cases, this incentive appears to range between 2 and 3 mills/kWh for hot and cold purification cases respectively.

This result must be treated with caution.

It should not be used as a basis for making technology development decisions without further investigation, for the following reasons:

The major purpose of this study was to investigate the incentives for developing high temperature gas purification systems. Therefore, the bulk of the engineering effort was concentrated on this section of the plant.

Detailed steam system design optimization for each case was not carried out.

Although uniform stack gas temperatures and a reasonable pinch point approach to each case was established, similar steam cycles were employed for both the 1,950 and 2,400°F cases.

It was also necessary to estimate costs for gas turbine technology not yet available. For this preliminary study, a nominal value of $200/kW was made for the combined cycle portion of the plant, independent of gas turbine inlet temperature.

The use of such methods was consistent with the primary objective of this screening study. However, they could have indicated a lower incentive for the development of a 2,400°F turbine than really exists.

It is therefore strongly recommended that further detailed studies be conducted to provide a basis for establishing the benefits to be associated with the development of high temperature gas turbines.

Appendix A

ERDA HOT SULFUR ABSORPTION PROCESS DEVELOPMENT PROGRESS SUMMARY

Summary A-1	RI 7947 Removal of Hydrogen Sulfide from Hot Producer Gas
Summary A-2	MERC/TPR-75/1 Removal of Hydrogen Sulfide from Hot Low Btu Gas with Iron Oxide Fly Ash Sorbents
Summary A-3	MERC/TPR-75/2 Hydrogen Sulfide Removal from Hot Producer Gas with a Solid Fly Ash-Iron Oxide Absorbent
Summary A-4	Regenerable Iron Oxide Silica Sorbents for Removal of H_2S from Producer Gas (Summary of paper presented at 4th Energy Resources Conference, University of Kentucky, January 6-7, 1976)

SUMMARY A-1

Summary of Report of Investigation No. 7947 1974

Morgantown Energy Research Center, ERDA

Title: Removal of Hydrogen Sulfide from Hot Producer Gas

Abel, W. T., F. G. Shultz, and P. F. Landon

A sintered blend of fly ash and iron oxide (FAIO) had the best mechanical and chemical properties of 48 materials tested for use in the desulfurization of hot producer gas. Tests of FAIO were performed through 174 consecutive absorption regeneration cycles in the temperature range 1,000°F to 1,500°F using dry and wet simulated producer gas (see Figure A-1). The test absorbents were contained in an electrically heated stainless steel reactor in which wall effects could be significant.

In extended test, absorbent capacity was lower than observed for freshly sintered absorbent:

- Because of reducing pore volume as number of completed cycles increases. Effect observed to be complete after 30 cycles.
- Because of inhibiting effect of steam in desulfurization process.

Performance of absorbent was examined in desulfurization of a commercial type of coal gasifier produced fuel gas. The gas contained approximately

Tar	1.0 lb/1,000 scf
Dust	0.5 lb/1,000 scf
Steam	5.0 lb/1,000 scf

Over a 60 hour absorption period, 95 percent desulfurization of gas was accomplished. In above test lower absorbent capacity than in wet simulated gas test was observed which was attributed to the higher steam content of the commercial type gas.

Figure A-1 summarizes observed FAIO capacity in reported tests. Changes in composition of the gas undergoing desulfurization occurred because of shift reaction effects catalyzed by iron oxide absorbent. Study reports that effect could be beneficial where the objective is to produce pipeline gas. Reaction by which H_2S is removed from fuel gas is approximated by:

$$Fe_2O_3 + 3H_2S = 2\ FeS_{1.5} + 3H_2O$$

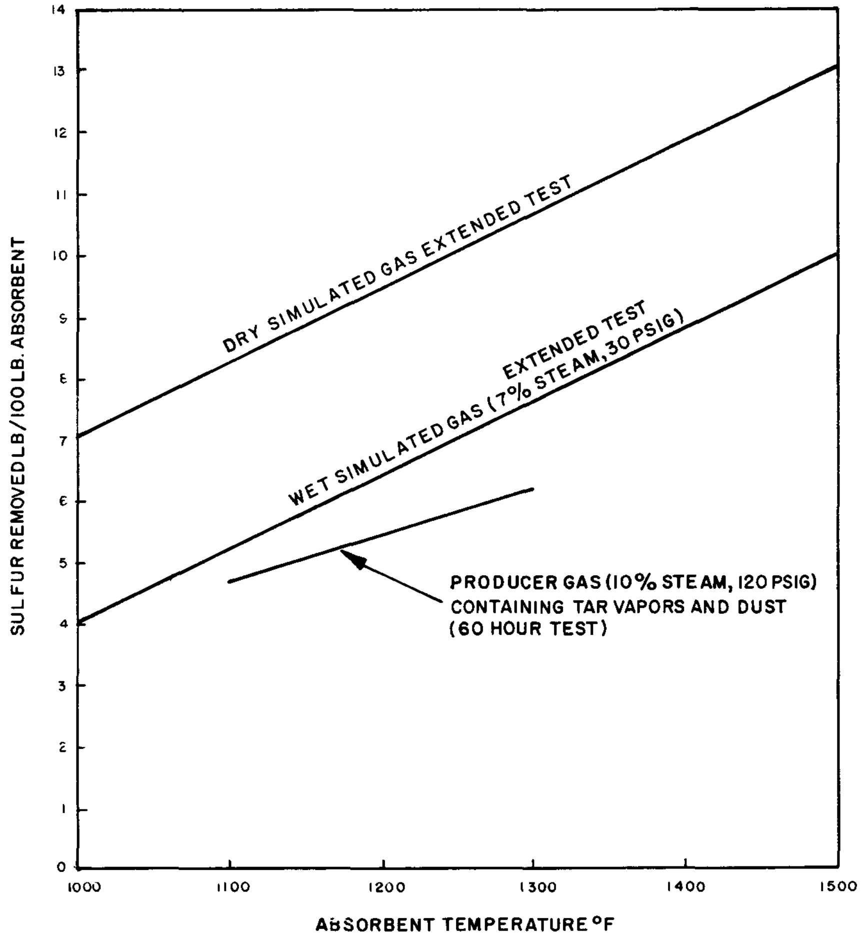


Figure A-1. Absorption of Sulfur by Fly Ash Iron Oxide Absorbent

Complete reaction of hydrogen sulfide with iron oxide would lead to absorption of 1.5 pounds sulfur per 100 pounds of FAIO absorbent. Capacity would be slightly higher if credit is allowed for indigenous iron oxide in fly ash support material.

At 1,300°F under producer gas conditions approximately 42 percent of theoretical capacity observed to be attainable.

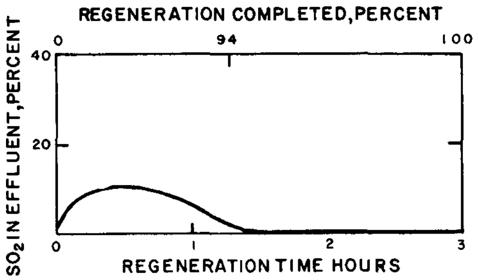
Regeneration of the sulfided bed was accomplished by passing ambient air into the hot absorbent bed.

In regeneration, an hourly space velocity of 1,000 was found not to cause the maximum temperature limitation (1,500°F) on the absorbent to be violated and produced an effluent gas containing between 6 and 10 percent sulfur dioxide during a half hour regeneration period. Sulfur dioxide decreased to about 1 percent after one hour when regeneration was over 80 percent complete.

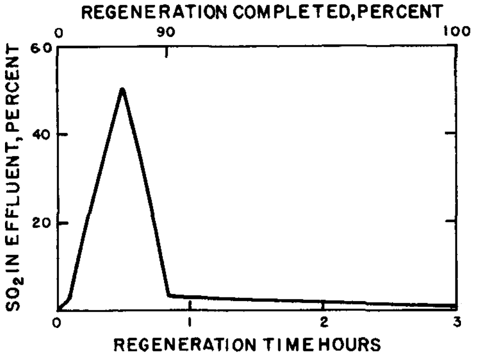
The major constraint in regeneration was avoiding fusion of the absorbent pellets because of allowing temperature to become too high. High purity sulfur dioxide could be produced using oxygen only for regeneration, but such a regeneration would be difficult to commercialize because of the high temperature bed fusion problem. The regeneration process is approximated by:

$$4\ FeS_{1.5} + 9.0_2 = Fe_2O_3 + 6.0_2$$

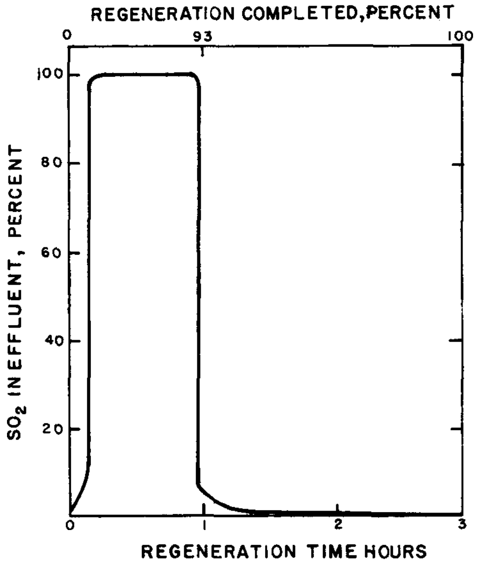
Regeneration experience is summarized in Figure A-2. This figure is reproduced because Morgantown regeneration experience has important relevance in the engineering design of commercial hot purification units, particularly related to the inability to produce a high concentration sulfur dioxide stream without endangering absorbent properties.



REGENERATION OF SULFIDED ABSORBENT IN AIR (AIR FLOW =10 SCHF, INITIAL BED TEMPERATURE = 1,000°F, MAXIMUM BED TEMPERATURE = 1,410°F).



REGENERATION OF SULFIDED ABSORBENT IN 50 PERCENT AIR-O_2 MIXTURE (AIR-OXYGEN FLOW RATE =6 SCHF, INITIAL BED TEMPERATURE =840°F, MAXIMUM BED TEMPERATURE = 1,410°F).



REGENERATION OF SULFIDED ABSORBENT IN O_2 (O_2 FLOW RATE = 3 SCHF, INITIAL BED TEMPERATURE = 850°F, MAXIMUM BED TEMPERATURE = 1,350°F).

Figure A-2. Morgantown Regeneration Experience Summary

SUMMARY A-2

Summary of Report No. MERC/TPR-75/1 1975

Morgantown Energy Research Center, ERDA

Title: Removal of Hydrogen Sulfide from Hot Low Btu Gas with Iron Oxide-Fly Ash Sorbents

Authors: E. C. Oldaker, A. M. Poston, and W. L. Farrior

Work outlined in Summary A-1 is continued in this paper. Tests are described demonstrating removal of H_2S from commercial type producer gas at 1,000°F. A cyclone cleaned commercial type of producer gas was contacted with FAIO absorbent at an hourly space velocity of 1,900 v/v/hr. Observed absorbent capacity averaged 8.25 percent (wt) during four 15 hour absorption periods with 90 to 94 percent of hydrogen sulfide being removed from the producer gas. The tested absorbents were contained in an electrically heated stainless steel reactor in which wall effects could be significant.

Carbonaceous material deposited in the absorber bed during absorption was observed burned out during regeneration with air.

Average H_2S removal from the producer gas was a function of the concentration of H_2S in the treated gas, as follows:

Treated Gas H_2S Content	H_2S Removal from Fuel Gas
150 grains/100 scf	83 percent
50 grains/100 scf	93 percent

Quotation:

> "The main problem appearing while performing the tests was controlling regeneration temperatures to protect the sorbents from excessive temperature exposure. Research is now underway to determine the best method of regeneration, including dilution of the air used for regeneration, lower iron oxide content in the sorbents, and/or a different support material for the iron oxide which would permit higher temperatures during regeneration."

A summary of the experimental results of this work is provided in the following table which also indicates the composition of the fuel gas studied by Morgantown.

Table A-1

RUN DATA DURING 4 ABSORPTION PERIODS

RUN	1	2	3	4	
Space Velocity	1910	1910	1910	1910	V/V/Hr
Inlet Steam Vol %	8.5	3.3	3.1	6.3	% Vol
Bed Temperature	1085	1085	1085	1085	F
H_2S Absorbed @ 1.5 Gr/scf = Saturation	83	83	81	86	%
H_2S Absorbed @ 0.5 Gr/scf = Saturation	92.9	91.4	90.7	94.2	%
Absorption Capacity	9.19	6.17	8.35	9.29	% Wt
Producer Gas (Dry)					
CO	17.3	16.2	23.2	20.9	% Vol
CO_2	13.0	12.2	6.7	8.1	% Vol
H_2	17.8	15.7	13.5	13.5	% Vol
N_2	46.8	52.5	52.5	56.1	% Vol
CH_4	3.6	3.6	3.2	3.1	% Vol
H_2S	0.48	0.58	0.52	0.53	% Vol

SUMMARY A-3

Summary of Report No. MERC/TPR-75/2 1975

Morgantown Energy Research Center

Title: Hydrogen Sulfide Removal from Hot Producer Gas with a Solid Fly Ash-Iron Oxide Absorbent

Authors: E. C. Oldaker, A. M. Poston, and W. L. Farrior

Paper continues recording FAIO development described in Summaries A-1 and A-2. The emphasis is on discovering binding materials which combine high crushing strength with high sulfur pick-up. Satisfactory binders are found to be solid bentonite and sodium silicate. The latter is precipitated from solution by evaporation of the solvent from the saturated absorbent. As a result of this development, crushing strength of the absorbent has been increased to 50 lb/cm length and absorption capacity increased to 12-14 weight percent in the region of 1,200°F, values believed to be adequate for commercial application.

A most important result of this development is evidence that absorbents can be designed to be independent of steam content of producer gas; though some steam depression of performance can be seen in the results. (Comment: more investigation is required before an absorbent which is capable of processing Lurgi gas containing 50 percent [vol] steam can be claimed.)

The regenerative technique concluded to protect the absorbent from excessive temperature exposure is use of air diluted with nitrogen.

Effects of tests of recycling the absorbents imply that the aging effect observed earlier has been largely overcome. However, the aging tests were over 30 cycles and the effect may have been incomplete when the test was terminated.

On the basis of this paper it appears reasonable to predict a design sulfur pick-up of 8 wt percent in a commercial application operated at 1,200°F.

SUMMARY A-4

Summary of Paper Presented at Fourth Energy Resources Conference

University of Kentucky

January 6-7, 1976

Title: Regenerable Iron Oxide Silica Sorbents for the Removal of H_2S from Hot Producer Gas

Authors: W. L. Farrior, A. M. Poston, Jr., and E. C. Oldaker

The abstract of this paper summarizes the general advances described in it.

> "Results of research towards development of a more efficient solid regenerable sorbent for the removal of hydrogen sulfide from hot (1,000°F-1,500°F) producer gas is discussed. Different solvents were prepared by mixing, extruding and sintering iron oxide using inexpensive silicas of different particle size distributions as support material. Different binders and burnout materials were investigated. Hydrogen sulfide sorption increased with iron oxide content up to the 45 percent level, the highest tried, whereas a 25 percent additon is optimum for fly ash supported sorbents. Contrary to results obtained with fly ash supported sorbents, bentonite additions, by the methods used produced little increase in physical strength while sodium silicate not only increased strenth but sorption capacity as well. One such sorbent proved to have higher sorption capacity, crushing strength and apparent higher heat resistance than the fly ash supported sorbent."

Most significant is that silica supported sorbents can be regenerated at higher temperatures than FAIO developed in earlier work.

The favorable experience quoted in the subject paper leads the investigators to predict the development of supported iron oxide absorbents rugged enough for fluidized or moving bed application.

The silica-iron oxide absorbents are likely to operate above 1,500°F without penalty.

The silica based absorbents are the newest and most promising hot absorbent so far studied. Studies are at an early stage.

Appendix B

OTHER HOT FUEL GAS PURIFICATION PROCESSES

Iron oxide, zinc oxide, and limestone are well known desulfurization agents discussed in various texts and standard sources. Certain historical and current activities with these absorbents are of interest in the present context, as follows:

- Appleby-Frodingham Fluidized Bed Desulfurization Process
- Hot Fuel Gas Purification with Zinc Oxide
- Hot Fuel Gas Purification with Limestone

Appleby-Frodingham Process

Bureau and Olden (3) described a commercial hot coke-oven gas desulfurization process based on reacting hydrogen sulfide with iron oxide in a fluidized bed of 16 to 100 mesh material at 650°F to 750°F. In this process, sulfided iron oxide passes to a regenerator where it is reoxidized in an air fluidized operation. Regeneration produces a vent gas containing sulfur dioxide, and the heat of regeneration is utilized to make the process thermally independent - hot regenerated oxide being returned to the desulfurization reactor. A unit of this type was installed at the Exeter Works of the SW Gas Board (UK) in which gases from regeneration passed through a catalytic converter where sulfur dioxide was oxidized to sulfur trioxide and the latter was subsequently absorbed in 98 percent sulfuric acid washing operation. The process is usually referred to as the Appleby-Frodingham (A-F) process.

The Exeter A-F unit was closed in 1965 for economic reasons. In parallel with A-F development the naphtha reforming processes had become a source of less expensive town gas and the solid fuels conversion processes of that time became obsolete. Also, sulfuric acid revenues were lower than expected and required staffing was greater than expected. Bureau and Olden (3) discuss the local problems of the Exeter operation and say "Publicity of failure is not a common event, but the authors feel that much interest and perhaps even value to others could be lost in this instance if success is the sole criterion for publication."

In 1976 we have reason to thank Mr. Bureau and Dr. Olden for the service of recording adverse experience. There are many data in their paper to assist today's hot purification student. It is also conceivable that in today's energy and environment-conscious world the A-F process will be revived.

The A-F operating temperature (650°F to 750°F) is probably too low where tars are a component of the gas to be purified. This may limit the applicability of the process with moving bed gasifiers which produce tars and where the gas must be maintained above 1,000°F to avoid tars condensation.

Secondly, A-F experience with oxygen in the crude gas is notable, quoted as follows from Bureau and Olden (3):

> "The presence of oxygen in the crude gas results in a loss of combustible constituents during its passage through the absorber. Attempts were made to check if this loss was limited to hydrogen or possibly carbon monoxide but, while inlet and outlet gas analyses showed minor variations, the results were inconclusive. Apart from experimental error, the results were possibly affected to some degree by the reduction of ferric oxide by hydrogen in the gas. The oxygen in the crude gas undoubtedly had some influence on the bed temperature in the absorber due to internal combustion and it appeared a possibility that attainment of working temperature could be accelerated in the final stages of a start-up by the controlled increase of oxygen in the crude gas but there were no facilities to do this on the Exeter plant."

One of the problems of today's hot fuel gas purification process application is to elevate the temperature of the gas leaving certain moving bed gasifiers to above the tar's dew point (1,000°F+). A method proposed for accomplishing this is to inject air or oxygen to the cool stream (c 850°F) and to rely on internal combustion to achieve the desired temperature increase. The A-F experience tends to support the feasibility of such an operation.

In addition, the foregoing quotation from Bureau and Olden mentions the reduction of ferric oxide by hydrogen in the gas. In the massive hot purification units contemplated for use in combined cycle power generation, such an effect should not go unaccounted. The possibility of the effect is again discussed in this report under Chemistry in Section 2, page 2-3.

A-F process technology may be discussed in the United States with Woodall Duckam representatives in the Pittsburgh office of that company. Woodall Duckham was the construction contractor involved with the process in the United Kingdom.

Hot Fuel Gas Purification with Zinc Oxide

Zinc oxide reacts with hydrogen sulfide to form zinc sulfide. It is used to absorb sulfur in the protection of hydrocarbon reforming catalysts which are deactivated by sulfur levels above about 0.2ppm in the material passing over them. Phillipson (5) reports that zinc oxide is superior to iron oxide for sulfur absorption as a result of the lesser effect of water vapor and reducing gases on the related hydrogen sulfide absorption equilibria. He explains, in particular, why iron oxide is not a satisfactory absorbent for the protection of hydrocarbon reforming catalysts while zinc oxide is so. However, the order of equilibrium hydrogen sulfide in the treated product above sulfided iron oxide quoted by Phillipson would be adequate to satisfy environmental protection regulations where the treated material is a fuel gas. Phillipson also reports adverse equilibria for sulfided iron oxide in a reducing atmosphere. However, the equilibrium hydrogen sulfide levels above iron sulfide in the presence of hydrogen once again appear to be satisfactory where the treated material is to be used as a fuel gas. The gas turbine developers will ultimately decide this.

IFP (11) are developing a regenerable zinc oxide absorbent for hot desulfurization use. The absorbent is stated to be able to withstand the temperature swings involved in the process without damage. It is also stable at considerably higher temperatures than the limit temperatures on the Morgantown supported iron oxide absorbent discussed in this report.

Hot Fuel Gas Purification with Limestone

Use of dolomitic limestone for hot purification has been reviewed by Oliver (20). He concludes that the benefits in economy and efficiency of energy conversion from developing hot limestone purification are potentially substantial, but the development cost would be high. The gains in economy, variously estimated to be from 5 percent to 15 percent coal savings, would be accompanied by greater emissions of SO_2 and probably NO_x than are anticipated from current cooled gas cleanup technology. It is further observed by Oliver that the limestone process would be more efficient than iron oxide cleanup because it operates at higher temperature. However, with the more severe temperature conditions, more chemical and mechanical development difficulties are to be anticipated. Oliver's evaluation relates to the proposed Conoco fluidized bed hot limestone process.

In ascribing advantage to the Conoco process over cold purification, Oliver points out that the comparison is between developmental and established processes. Until the development is complete, the possibility of inaccuracy in that comparison must be recognized. For iron oxide, the lower temperature (1,200°F compared to 1,600°F for limestone) is stated to imply less of an efficiency advantage for that process. Here, the comparison is between two processes both in the early stages of development and the possibility of erroneous conclusion exists. Iron oxide hot purification meshes better with Lurgi ash producing and slagging technology than the limestone process. Limestone hot purification meshes better on a temperature basis with entrained bed gasification, which is also not commercialized.

Air Products and Chemicals (APCI) has found the regenerative use of limestone to be impractical because of poor regenerability of the limestone, rapid coke deposition in the limestone bed, high energy consumption, the current lack of suitable materials of construction, and limestone properties variation with source (19).

The APCI fixed bed process shows higher estimated operating costs than the fluidized bed Conoco process. A substantial part of the difference arises because the processes have different chemical schemes, as outlined in the following:

AIR PRODUCTS PROCESS CHEMISTRY

Process $CaO + H_2S = CaS + H_2O$

Regeneration $CaS + H_2O + CO_2 = CaCO_3 + H_2S$

Calcination $CaCO_3 = CaO + CO_2$

Absorbent is limestone in the APCI process.

Endothermic heat of calcination is supplied by natural gas.

CONOCO PROCESS (6)

$$MgO \cdot CaCO_3 + H_2S \underset{400°F}{\overset{1600°F}{\rightleftharpoons}} MgO \cdot CaS + CO_2 + H_2O$$

Absorbent is dolomitic limestone; sulfur absorption occurs at 1,600°F via the forward component of the above reaction, and regeneration occurs at 400°F via the reverse component of the above reaction.

In the APCI case, the endothermic heat of calcination is supplied in a separate natural gas fueled operation. In the Conoco case, the more reactive dolomitic limestone is in effect calcined in situ by process heat and the irreversibilities associated with the separated calcination are largely reduced. The Conoco process has not been penalized in published work for absorbent loss by attrition. The APCI process suffers excessive absorbent loss through attrition, which has a substantial effect on energy requirements since only regenerated absorbent may be discharged.

The net effect of the differences is that the APCI process has about three times the energy consumption of the Conoco process. The Conoco process has a reported energy-related cost of three percent of the coal (four percent sulfur in the coal) and APCI about nine percent of the coal.

The following table summarizes operating costs for APCI single and multiple cycle processes and for the Conoco process as abstracted from published work. Unrecovered energy from the calcination operations accounts for a large part of the indicated cost differences.

In an extensive study of power generation, Robson (12) has evaluated the application of limestone to the desulfurization of fuel gas produced by air blown entrained bed coal gasification. Robson (21)(page 201) indicates that the combined cycle thermal efficiency corresponding to 2200°F gas turbine inlet temperature with air blown entrained bed gasification and hot limestone desulfurization is 37.6% (based on coal HHV). This compares favorably with the efficiency determined for the Morgantown system in this work which is approximately 38% at 1950°F gas turbine inlet temperature. Robson (21)(page 201) also shows that the entrained bed system with cold (Selexol) purification has a thermal efficiency of 31.9% which is to be compared with the value of 38% determined for the Benfield supported system in this work. Robson (21)(page 186) explains the discrepancy (hot 37.6% to cold 31.9%) as being due to a more stringent sulfur removal duty designed into the Selexol system. The latter explanation also accounts for the difference between the cold purification cases of this work (38%) and Robson (31.9%). In this work, purification systems were required to remove 90% of the fuel gas contained hydrogen sulfide (consistent with current EPA regulations applied to the whole plant) compared to the nearly total removal required in Robson's work (Table 18).

Other hot limestone uses are for in situ desulfurization in fluidized bed combustion (boilers) by Westinghouse and fluidized bed coal gasification by Foster Wheeler Corporation. Both applications are potentially successful major applications of hot purification which are being developed at the time of writing.

Battelle Northwest Institute is employing molten carbonates for scrubbing fuel gas in its hot purification process. SRI points out that this process does not merit consideration for combined cycle systems until it is proved that the molten salts can be employed for the purpose without hazard to the gas turbine.

Table B-1

COMPARISON OF HOT LIMESTONE H_2S REMOVAL PROCESSES AND UTILITY COSTS AT 1,000 MW BASIS (6,132 HR/YR)

		AIR PRODUCTS REGENERATIVE		AIR PRODUCTS SINGLE CYCLE		CONOCO REGENERATIVE	
UNIT	RATE COST/UNIT	UNITS/HR	$/YR	UNITS/HR	$/YR	UNITS/HR	$/YR
Dolomite	$3/T	39.5	$ 726,642	44.0	$ 809,424		
Half Calcined Dolomite	$10/T					7.6	$ 466,032
Waste Disposal	$2/T	39.5	$ 484,428	44.0	$ 539,616	7.6	$ 93,206
LP Steam	$1.30/MLbs	49.0	$ 390,608	104	$ 827,452	56	$ 445,612
Electricity	$0.02/kWh	32942	$4,040,007	37686	$4,621,811	9189	$1,126,939
Natural Gas (for calcination)*	$2/MMBtu	398	$4,874,940	312	$3,832,500	-	-
Cooling Water	$0.024/MGAL	1232	$ 181,311	1485	$ 218,544	1217	$ 179,103
Utilities Total			$10,697,936		$10,849,347		$2,310,892
Mills/kWh			1.74		1.77		0.377

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*Based on no credit for heat recovery in the calcination-regeneration-absorbent disposal systems.

Appendix C

DIAGRAMS OF PANEL BED FILTER BEING DEVELOPED BY CITY COLLEGE NEW YORK

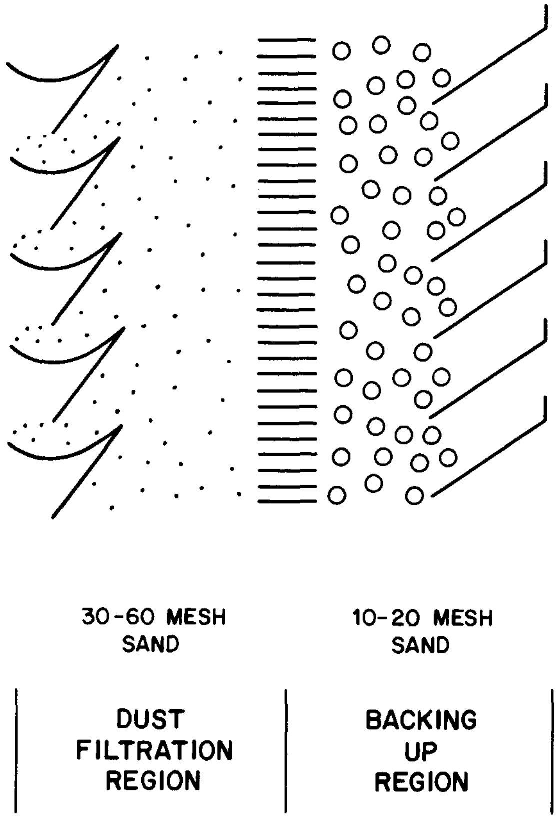
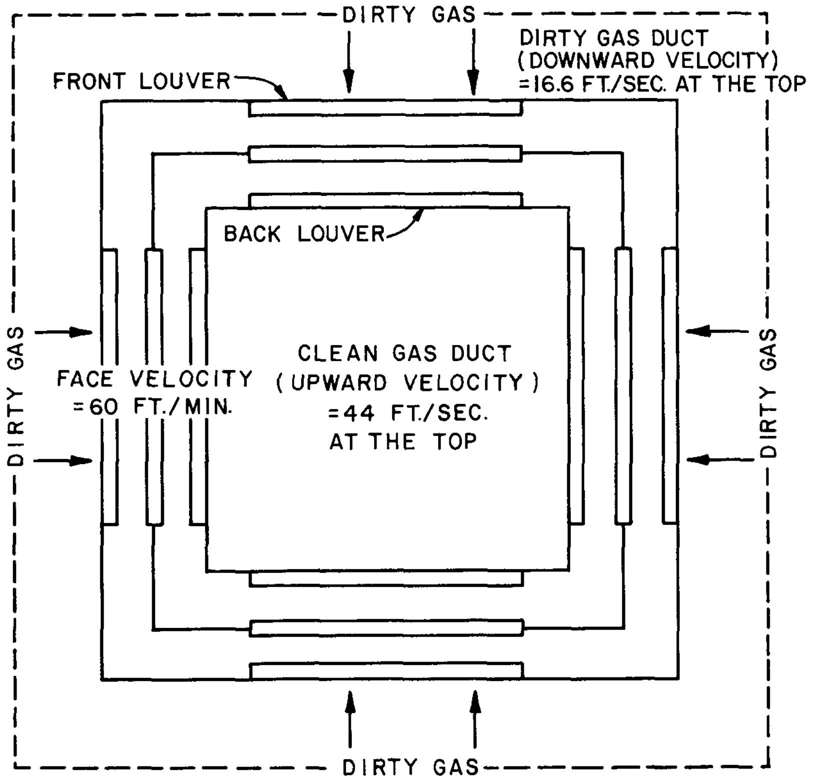
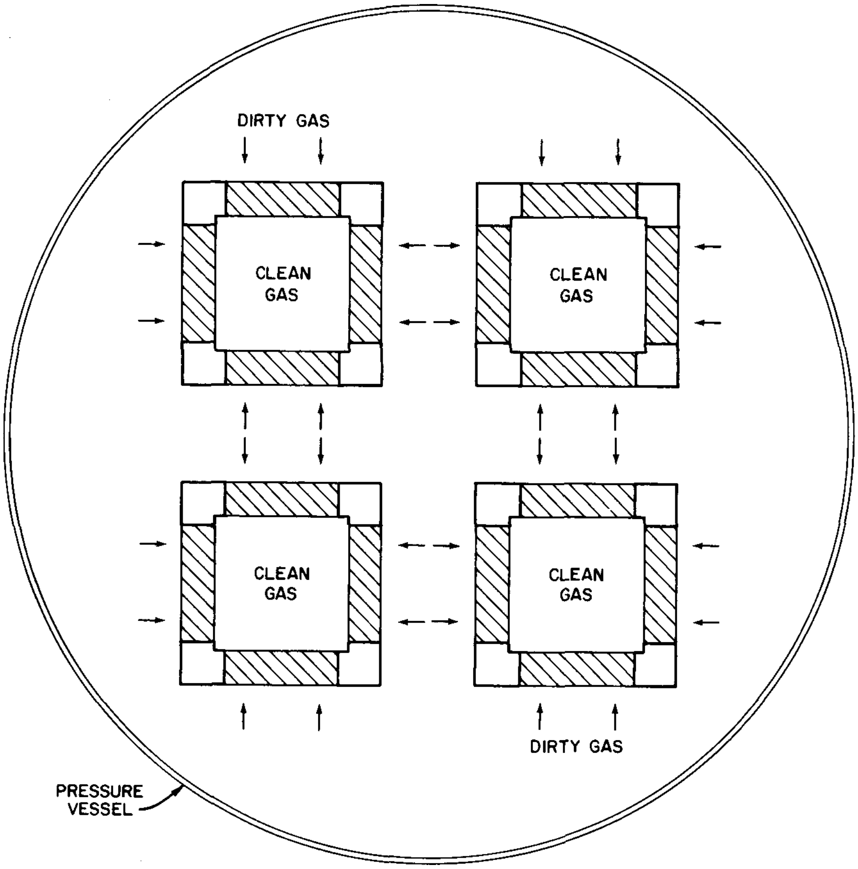


Figure C-1. Panel Bed Filter Louvers Arrangement



FILTERING SURFACE WIDTH = 1.0 FOOT
FILTERING SURFACE HEIGHT = 11.0 FEET
CAPACITY PER ELEMENT = 2,500 ACFM

Figure C-2. Panel Bed Filter Element



NOTES
ENCLOSURE DIAMETER = 7.3 FT.
CAPACITY PER UNIT (FOUR SQUARE ELEMENTS) = 10,000 ACFM.

Figure C-3. Panel Bed Filter Unit, Top View

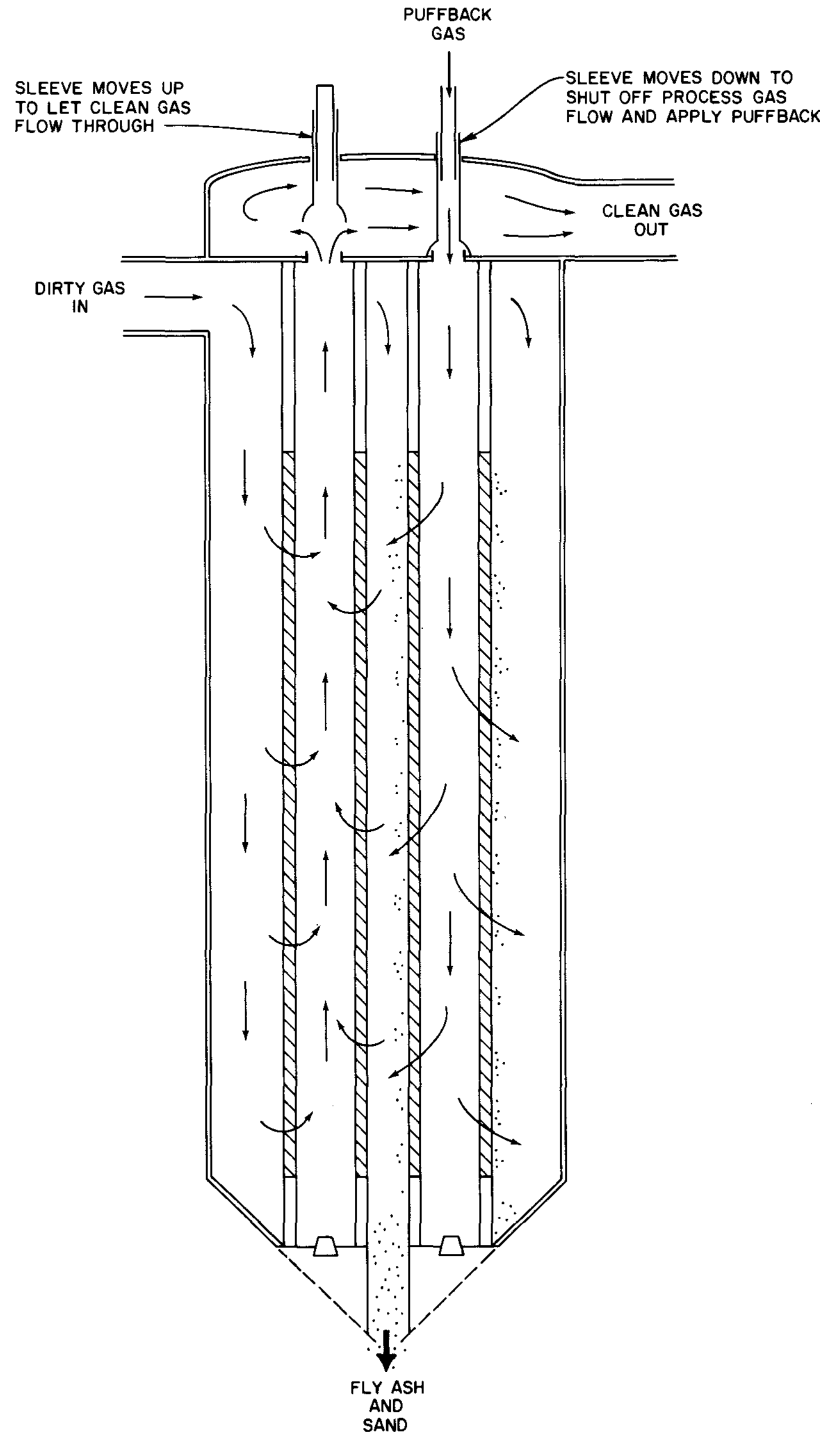


Figure C-4. Panel Bed Filter Unit, Side View

Appendix D

PROCESS PROFILE: ALLIED CHEMICAL SO_2 REDUCTION TECHNOLOGY

Process Profile: Allied Chemical SO_2 Reduction Technology
Stone & Webster: Fuel Gas Purification Study
Electric Power Research Institute

This process profile describes Allied Chemical SO_2 Reduction Technology for recovery of elemental sulfur from SO_2 gases utilizing CO/H_2 containing fuel gas as the reductant. In this case, the SO_2 feed gas and the reducing gas will be free of dust, tars, and oils.

In the first, or reduction system, a portion of the sulfur dioxide in the feed gas is reacted with the CO/H_2 reductant, in a fixed bed catalytic reactor system, yeilding a mixture of elemental sulfur, hydrogen sulfide and some unreacted SO_2, as well as CO_2 and water vapor according to the following equations.

$$2CO + SO_2 \longrightarrow 2CO_2 + S$$

$$2H_2 + SO_2 \longrightarrow 2H_2O + S$$

$$3H_2 + SO_2 \longrightarrow 2H_2O + H_2S$$

In the second, or Claus system of the process, the H_2S and unreacted SO_2 from the reduction system react to give additional elemental sulfur and water vapor as shown in the equation:

$$2H_2S + SO_2 \longrightarrow 2H_2O + 3S$$

The operating conditions in the reduction system are controlled to maintain the proportions of H_2S and SO_2, in the outlet gas, very close to the stoichemetric ratio of 2 to 1 which results in optimum removal of SO_2 in the Claus system and maximum recovery of sulfur for the process.

Since the reactions in the reduction system are exothermic, heat is removed from the gas in waste heat boilers. In this case, 50 psig steam is produced although the boilers could be designed for operation at higher steam pressures. The gas leaving the reduction system is maintained at a constant temperature.

Gas from the reduction system flows to the first sulfur condenser where the stream is cooled and a portion of the sulfur is condensed and removed as molten sulfur product. Steam at 50 psig is produced in this condenser. The temperature of the gas leaving the first condenser is increased by mixing in a by-pass stream of hot gas from the reduction system outlet.

The re-heated stream flows to the first Claus converter catalyst bed where H_2S and SO_2 react to produce an equilibrium mixture of these reactants plus sulfur and water vapor according to the equation given previously.

Leaving the first Claus converter the gas stream next flows to the second sulfur condenser which functions in the same manner as the first. Operating at a lower temperature, the second sulfur condenser generates 30 psig steam.

The cooled gas from the second sulfur condenser is also mixed with a by-pass stream of hot gas from the reduction system outlet to raise its temperature, and is then passed through the second Claus converter and the third sulfur condenser which generates 20 psig steam.

Gas from the final condenser is passed through a mist eliminator to remove any entrained liquid sulfur, and then through a tail gas incinerator where unreacted H_2S is oxidized to SO_2. The small quantity of other combustible components formed by side reactions (e.g., CS_2, CO, COS, and H_2) are also oxidized in the incinerator as are traces of elemental sulfur not removed in the final condenser and mist eliminator. The incinerated tail gas stream is delivered to an offsite's heat recovery system for generation of additional steam and then vented to a stack.

Molten sulfur from the sulfur condensers and the mist eliminator is collected in a sulfur pit from which it is periodically transferred to sulfur storage facilities.

Appendix E

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