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APPLICATIONS OF THE WOODALL-DUCKHAM TWO STAGE COAL GASIFICATION

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What Needs To Be Done NOW

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INTRODUCTION

I value the opportunity to take part in this conference, particularly because of the conference sub-title - "What Needs To Be Done NOW". As Dr. Beroes has underlined NOW, this paper will address itself to - "What Can Be Done NOW".

TWO STAGE GASIFICATION

The principle of W-D two-stage coal gasification is to separate the volatile matter of the coal before subjecting the remainder of the coal to the high temperatures of the gasification reactions. Figure 1 is a schematic view. Coal entering the top of the gasifier meets top gas which leaves the gasifier at about 250°F. As it descends through the fixed-bed gasifier, the coal is dried, then evolves gas, light hydrocarbons and tar, and reaches the bottom of the upper part of the gasifier as semi-coke, or char if the coal is a non-agglomerating coal such as sub-bituminous or lignite. A piece of coal takes several hours to pass through this zone.

The semi-coke or char now passes into the gasification zone, where temperatures rise from about 1200°F at the side gas off-take to typically 2200°F at the final combustion stage. The ash formed is lightly clinkered and is extracted by a rotating grate. The grate also serves as the distributor for the gasification agent, either air/steam or varying proportions of oxygen/air/steam.

The gasification agent converts the semi-coke into carbon monoxide, carbon dioxide, hydrogen, some undecomposed steam and nitrogen. This product is termed clear gas. The sensible heat of this gas at about 1200°F is roughly double that required to dry and distill the volatile matter from the incoming coal. Therefore about half the clear gas is withdrawn at 1200°F, via the side gas offtake.

HOT RAW GAS

Figure 4 shows the two-stage gasifier in its simplest application, as a hot raw gas plant. It also shows some features of the gasifier, such as the coal feed system, consisting of a lock hopper, feed buffer hopper, and flooded feed coal distributor. This system, together with a very low gas off take velocity, keeps fines carryover to a very low level. This feature of very low fines carryover is one of the features of two-stage gasification which distinguish it from single-stage gasifiers. Another is the fact that coal hydrocarbons have not been significantly cracked.

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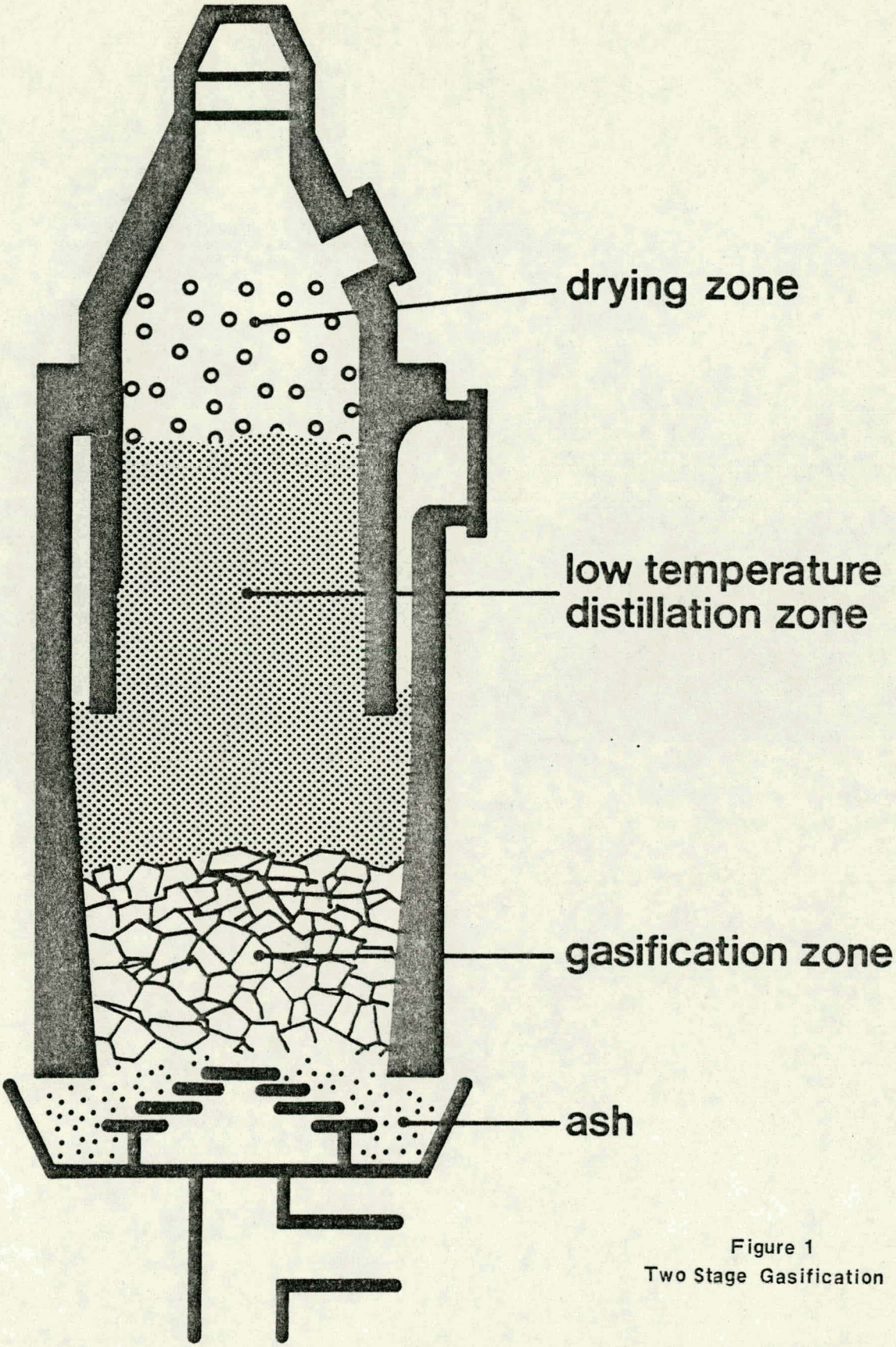


Figure 1
Two Stage Gasification

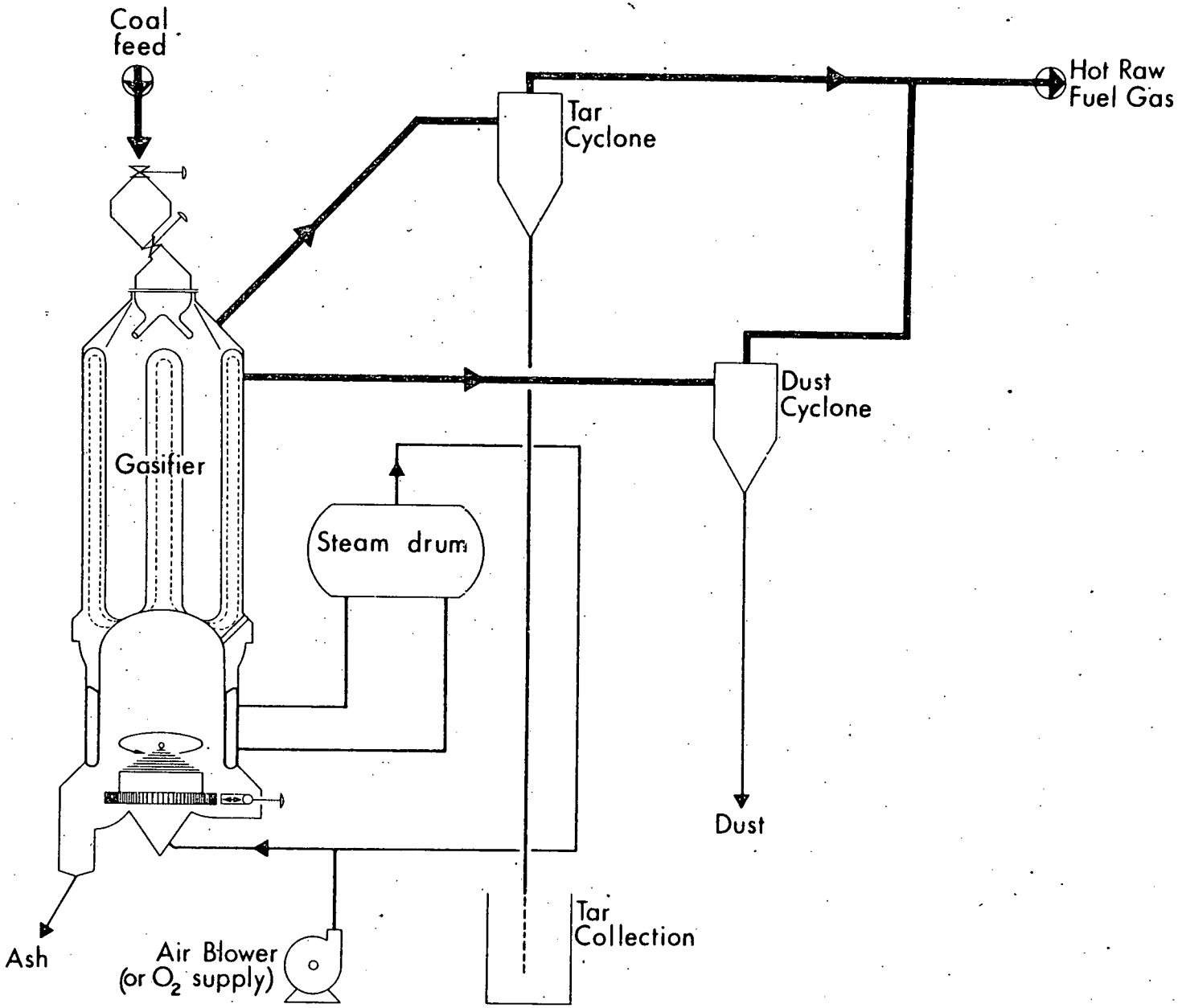


FIGURE 4. WD/GI PROCESS FOR HOT RAW FUEL GAS

The gasifier consists of a steel water-jacket, at high temperature gasification zone surmounted by refractory brickwork. Gasification steam is generated in the water-jacket. The refractory upper section is constructed of various grades of firebrick and contains vertical partitions to ensure good distribution. The grate may be driven either hydraulically or electrically, and the ash is extracted through a simple lock-hopper.

As the gasifier is operated at only a few psi above atmospheric, materials and equipment problems are minimised. This is especially the case at the top of the gasifier, where the temperature is low. Two-stage gasifiers do not accumulate hard carbonaceous deposits in the head and gas off-takes as do many single-stage units.

The Hot Raw Gas plant is widely used for large-scale combustion applications. The two gas streams are combined and the tar and oil mist present in the top gas is re-evaporated. The resultant fuel gas, with an H.H.V. of about 210 Btu/SCF, at about 600°F, is piped directly to major users, such as steel mill furnaces or glass melting. Sulfur removal from this gas has not yet been carried out, a hot gas desulfurisation process is clearly called for so that the high thermal efficiency of these plants (between 90 and 95%) can be maintained. The only such process that has so far achieved commercial use is the Frodingham process, which would complement our gasification process excellently.

DESULFURIZED GAS

Figure 5 shows how two-stage gasification lends itself to simple and effective gas clean-up. No direct contact scrubbing is used. The top gas at 250°F, containing all the coal volatile matter, is electrostatically de-tarred, indirectly cooled, and passed through a second precipitator to ensure full demisting.

The clear gas at 1200°F, containing no volatile matter, is dedusted and cooled in a combination of waste heat boiler and tubular cooler.

The combined gas can be desulfurised by several well-known processes. A Stretford unit is shown here, and this process has been used to desulfurise two-stage gas.

GAS CHARACTERISTICS

Table 1 compares the properties of three well-known industrial fuel gases with cold clean air-blown gas, 176 Btu/SCF, hot raw gas at 207 Btu/SCF, and oxygen-blown gas at 285 Btu/SCF.

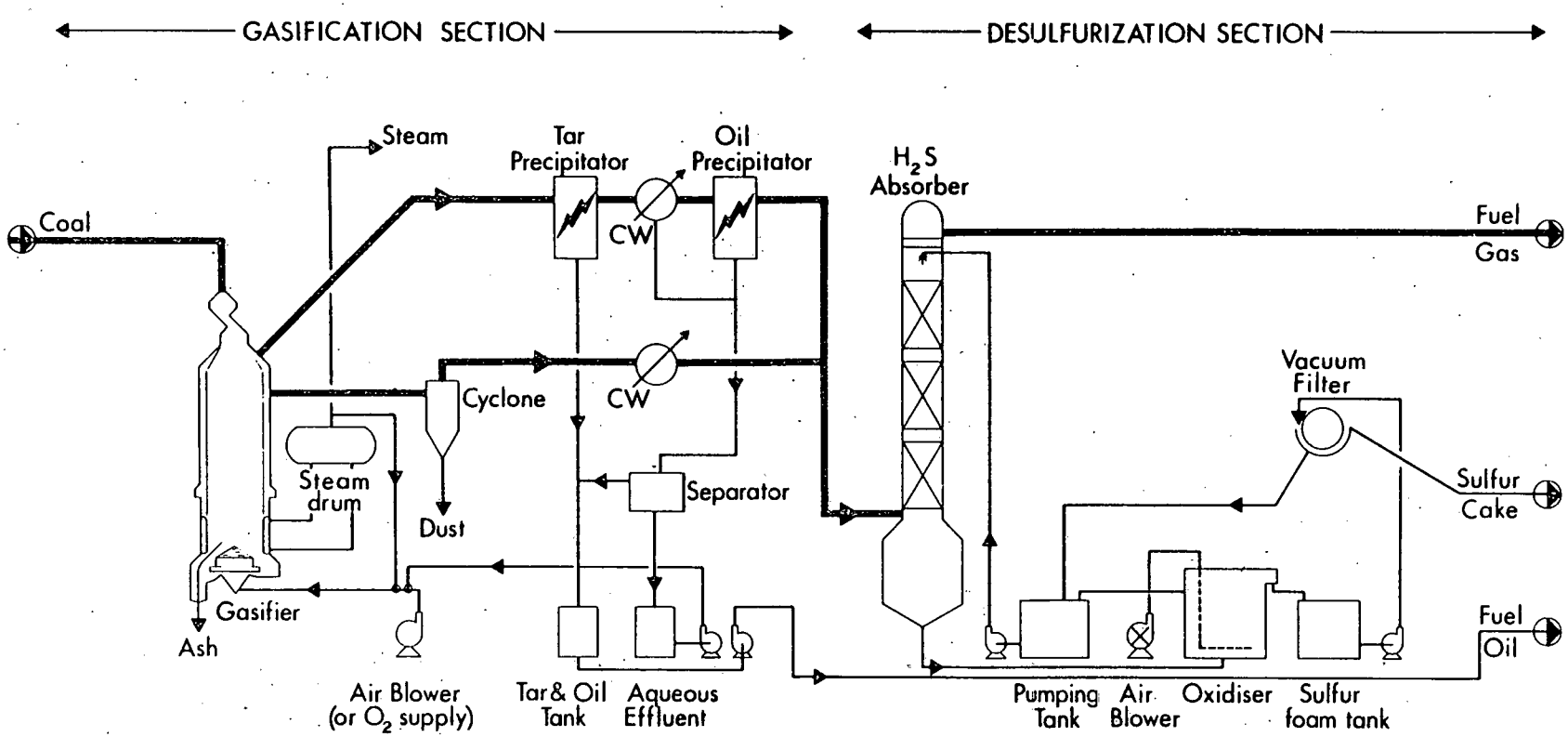


FIGURE 5. WD/GI PROCESS FOR COLD DESULFURIZED FUEL GAS

TABLE 1. PROPERTIES OF CERTAIN FUEL GASES

FUEL GAS	CO_2	CO	H_2	CH_4	ILLUMINANTS	N_2	H.H.V.	S.G.	WOBBE
Blast Furnace Gas	15.6	23.4	1.6	0.1	0	59.3	81	1.04	79
WD Low Btu. Gas	4.5	28.3	17.0	2.7	0.3	47.2	176	0.829	193
WD Hot Raw Gas	4.5	28.2	17.0	2.7	0.5	47.1	207	-	-
WD 300 Btu. Gas	18.0	37.5	38.4	3.5	0.4	2.2	285	0.73	334
WD Cyclic Gas	8.0	28.5	52.2	6.5	0.6	4.2	335	0.52	465
Coke Oven Gas	1.8	4.5	57.9	30.3	3.3	2.2	567	0.36	945
Natural Gas	0.0	0.0	0.0	93.3	4.9	1.8	1036	0.60	1337

Notes:

1. WD gas analyses based on typical coal properties.
2. Properties of other fuel gases are typical values
3. C.V. of hot raw gas is based on a tar and oil content of 0.12 lb./100 SCF, and includes the sensible heat of the gas.

Figure 2 compares the stack gas volumes per unit heat release of air-blown and oxygen-blown fuel gas with stack gas volumes for fuel oil and natural gas.

Figure 3 compares theoretical flame temperatures of air-blown and oxygen-blown fuel gases with natural gas for a range of feed preheat temperatures.

These data demonstrate that combustion performance with two-stage coal gas is closely comparable with natural gas. By using partial oxygen enrichment, it can be identical to natural gas, and when oxygen alone is used, the fuel gas has superior properties to natural gas.

CYCLIC PROCESS

The W-D two-stage process was originally developed for cyclic application, i.e. as an enriched water gas generator. The gas has been widely used for municipal utility applications. It is characterised by an H.H.V. of 330 Btu/SCF, and an unusually high hydrogen: carbon monoxide ratio of around 1.4 : 1. It is a completely proven method of making hydrogen plus carbon monoxide mixtures without the use of oxygen.

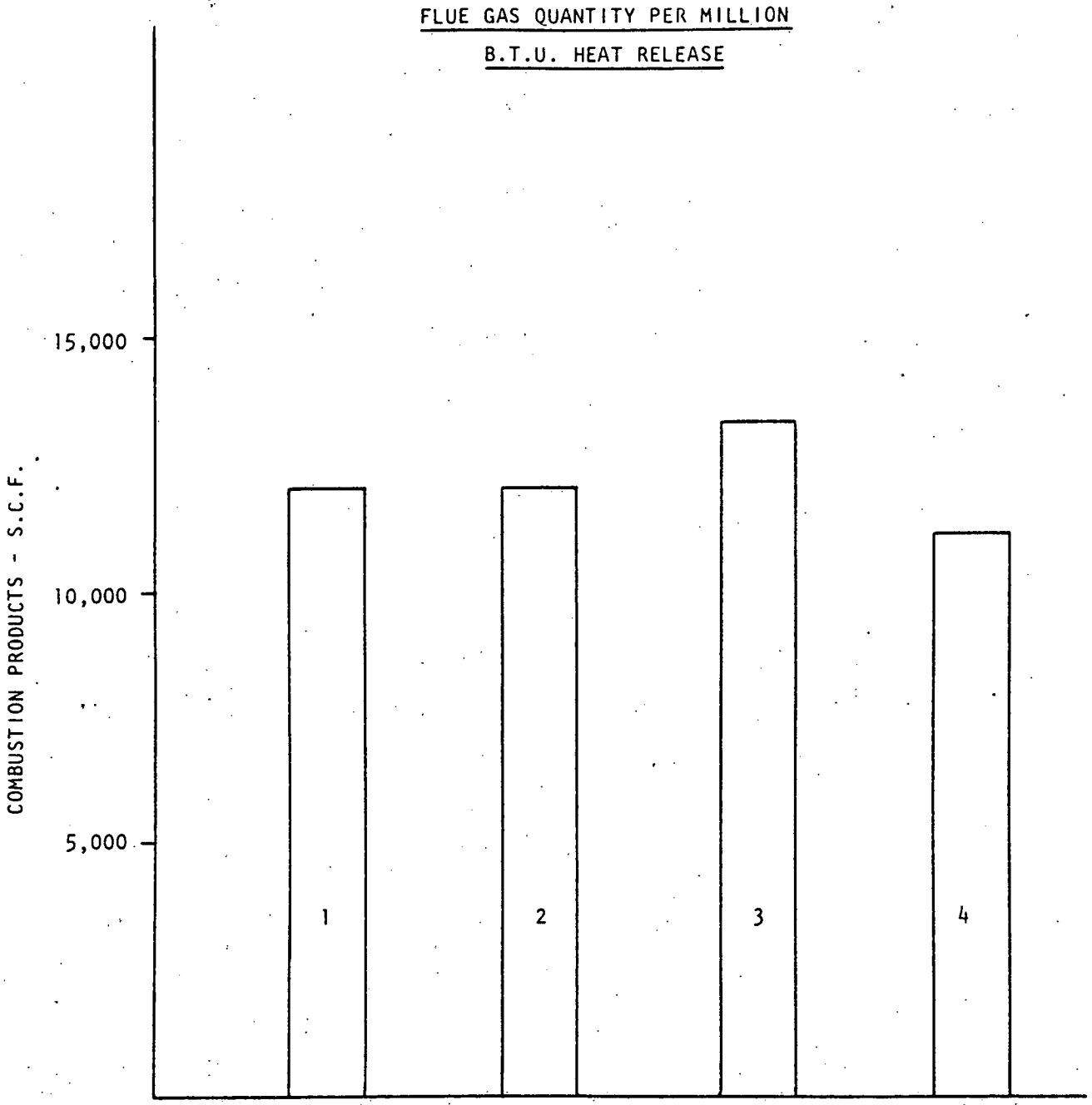
APPLICATIONS

The industrial applications of the W-D two-stage process are shown by slides of the following:

- Plant including 3 two-stage gasifiers, 2 of which supply cold clean gas, one hot raw gas.
- Open hearth furnace burning hot raw gas
- Heat treating and annealing wire
- Forging shop
- Bar heat treating
- Ladle heating
- Controlled atmosphere heat treating - wire carburising
- Boiler fired by gas and coal fines
- Plant control room
- Coal feeder

FLUE GAS QUANTITY PER MILLION

B.T.U. HEAT RELEASE



No. 1 - Typical No. 6 fuel oil with 5% excess air.

No. 2 - Natural gas with no excess air

No. 3 - Air blown WD/GI gas with no excess air

No. 4 - O_2 blown WD/GI gas with no excess air

Figure 2

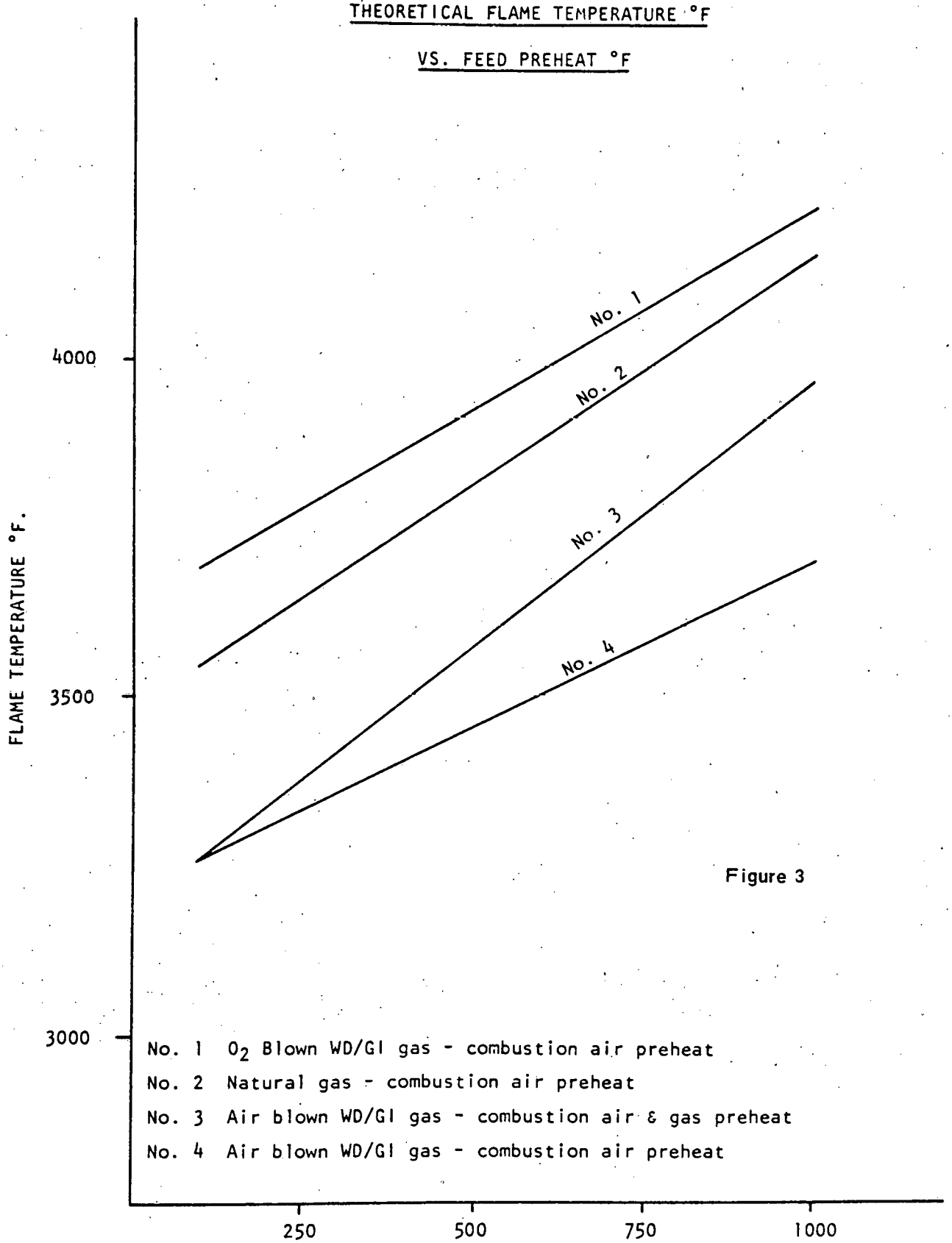

THEORETICAL FLAME TEMPERATURE °F
VS. FEED PREHEAT °F
FLAME TEMPERATURE °F.
4000
3500
3000
No. 1
No. 2
No. 3
No. 4
250
500
750
1000
No. 1 O_2 Blown WD/GI gas - combustion air preheat
No. 2 Natural gas - combustion air preheat
No. 3 Air blown WD/GI gas - combustion air & gas preheat
No. 4 Air blown WD/GI gas - combustion air preheat

Figure 3

The last two slides demonstrate the cleanliness possible in a steel mill coal gasification plant.

- Continuous caster tundish preheating
- Walking beam furnace
- Ladle heating

Although the steel industry is a major user of two-stage gasifiers, other industries such as glass

- Coca Cola bottle line
- Glass melting tank

and ceramics

- Tunnel kiln, enameled bathtubs

are well represented.

The aluminum industry uses W-D two-stage gasifiers, and there are a range of miscellaneous users including paper mills and coke oven underfiring.

An interesting new potential application for two-stage gas are the several regenerative stack gas scrubbing processes. We shall hear the Shell process described tomorrow by Dr. Jack Pohlenz of UOP. This process, in common with others, requires a source of reducing gas. The volume of gas required, composition and delivery conditions are eminently suitable for two-stage gasification.

Suitable applications include all the major fuel using industries of this country. The challenge before us is how to understand the potential user's attitude to "What Needs To Be Done NOW", and to present our processes and plant as a carefully tailored solution to a real need. If we fail in this, industry will decline to authorise capital expenditure, and we shall be totally reliant on Federal funds in our effort to demonstrate that industry in this country can use a lot more coal, NOW.

USER ATTITUDES

While natural gas is still available at around present prices, no industrial user will give it up to replace it by coal gasification. Steel mills and chemical plants are faced with gradual curtailment of natural gas, and they look to gradual substitution of a replacement.

This is one reason why many of the recent large scale plans for industrial fuel gas production are unlikely to be implemented. They attempt to serve a need -- replacement of an entire facility's natural gas by coal gas -- that does not exist, and may not exist until 1986.

The other pitfall to be avoided is overpricing the product for the market. Many plans have been advanced for the production of 300 Btu gas by oxygen/steam gasification, where the end-use has been to fire a variety of furnaces and even boilers! At $3.50 per MMBTU, this is unattractive to a steel mill. To my knowledge, in neither Europe nor South Africa, where both air-blown and oxygen-blown plants have been in regular use, has 300 Btu gas been made on-site for fuel purposes.

The above indicates that initial installations are likely to be of modest size, and predominantly air-blown. Oxygen enrichment may be used to ensure trouble-free retrofit and to facilitate industrial park distribution systems.

By modest size I refer to plants producing 4 billion Btu/day from two standard W-D gasifiers up to say 16 billion Btu/day from eight standard gasifiers. W-D two-stage plants are often expanded by the addition of more gasifiers, and gas purification equipment, so that the initial two or four gasifier plant might end up with twelve gasifiers.

Our largest single plant, serving a large steel pipe mill, grew in just this way to a current total of fourteen gasifiers. Gas desulfurisation is currently being added to this plant using the Giammarco Vetrocoke process, an arsenate-based second cousin to the Stretford process.

CONCLUSIONS

I have tried to point out that if "What Needs To Be Done NOW" is to convert a major portion of gas-and oil-consuming industry to coal, then we already have all the necessary data and experience. Just HOW this should be done is a more subtle problem -- I have suggested one approach, and perhaps Dr. Beroes would consider convening another conference at which alternate suggestions might be put forward!

Conf 760838--

FLUIDISED COAL COMBUSTION -

WHAT CAN BE DONE NOW.

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INTRODUCTION

The purpose of this paper is not to give a review of fluidised bed combustion development, nor to present data on the physical and chemical behavior of the combustion process. Dr. Keairns' paper has provided a comprehensive survey of this field. My aim is to give all of you who are working in the field of coal conversion an appreciation of what can be done today, and its probable impact. For many of you, fluidised coal combustion is a competitive technology - you will therefore be interested in its status! But for many coal gasification, liquefaction and beneficiation processes, fluidised coal combustion is a valuable accessory. It can utilise those awkward plant by-products in an efficient and environmentally acceptable manner; I refer to high-sulfur coal fines, char, residual tars and oils, and particularly the low heating value, high sulfur, reject stream from coal beneficiation.

PROCESS ADVANTAGES

The principal advantages of lean bed fluidised combustion are:

Fuel Flexibility

A fluidised combustor will operate on any fuel which has a net calorific value greater than its sensible heat at bed temperature. Provided that fuel feed facilities are available, a fluidised boiler should give design steam output on any solid, liquid or gas fuel.

Reduced Fuel Preparation

For solid fuels, crushing to achieve the particle size necessary for fluidising is the only preparation required. Large lumps of coal, wood, etc., can also be fed to a bed of smaller particles.

Pollution Control

Addition of limestone or dolomite to the bed enables more than 90% of the sulfur from the fuel to be retained in the bed as calcium sulfate. The quantity of oxides of nitrogen, NO_x, released in this process is also lower than that of other combustion systems, mainly due to the relatively low combustion temperatures employed. SO_3 levels are also kept low.

Controlled Operating Temperatures

An important characteristic of any fluidised bed is the high heat transfer performance that can be obtained with immersed heating surface. By making use of this characteristic or by operating with very high excess air flows, the bed temperature can be controlled within close limits. Normal operation is in the range 1350-1650°F, well below the softening point of the ash. Release of volatile alkali metal salts is also much reduced so that problems usually associated with "difficult" ash fuels may be alleviated.

ATMOSPHERIC PRESSURE APPLICATIONS

As long ago as 1969 Babcock & Wilcox Limited designed a 660MW fluidised bed boiler based on data which had been obtained by the National Coal Board and BCURA from test rigs. The largest of these rigs had a distributor area of some 12 sq. ft., whereas each bed in this large boiler required 1850 sq.ft. Despite the accumulation of over 20,000 hours of test rig operation, it became obvious during this design study that there were problems that could not be resolved without building a pilot plant of some intermediate size.

We therefore converted a boiler in our own boiler fabrication works at Renfrew, Scotland, to fluidised combustion at atmospheric pressure. This unit was commissioned in June 1975 and has since accumulated many months of operation.

As originally built the boiler was a Babcock cross type unit, fired by a spreader stoker, which generated 40,000 lb/hr of steam at 400 psig and 560°F. It had an economiser but no air heater. A mechanical grit collection system was installed and particles greater than 52 bss could be refired if required.

The stoker was removed and the bottom of the furnace chamber reconstructed to accommodate the fluidised bed, which has immersed tube surface. The distributor plate area is about 100 sq. ft. and the normal operating bed depth 3 ft. The coal is crushed to 95%-1/8", metered by rotary feeders, and fed into the bed by means of a pneumatic injection system. The coal nozzles are mounted above the distributor plate and fed from below. The normal operating temperature in the bed is 1550-1650°F. Under these conditions the immersed tubes account for about 60% of the total boiler output. The gases leaving the bed pass upwards through the furnace chamber, which now provides the freeboard or disengagement space. They then go through the original convection banks of the boiler which produce the remaining 40% of the steam and include the superheater and economiser.

In the first test series the fluidising velocity was 4 fps and the total steam production about 20,000 lb/hr. Current tests are being conducted with a fluidising velocity of 8 fps and a steam production of about 40,000 lb/hr.

Experience has already shown that the boiler is easy to operate. Mixing in the bed has been found to be extremely good if appropriate coal feed systems are provided.

The majority of the running to date has been carried out with a local Scottish coal having an ash content of about 20%. However, one test series using an Irish coal containing up to 60% of ash was run without difficulty.

The bed is ignited by means of oil burners directed at the surface and heating is carried out with the bed fluidised. Coal injection commences when the bed temperature reaches 850/900°F and the feed rate gradually increased as the bed temperature rises. The flow of oil is correspondingly reduced and completely turned off when the bed temperature reaches 1300/1400°F, after which all operation is on coal.

This boiler, which is in continuous operation, is the first fluidised combustion unit large enough to make optimisation studies on all the engineering problems associated with scale-up from the original rig facilities.

We have met and solved a variety of problems relating to turn-down, coal feeding and control. Sufficient experience has now been accumulated that we are now able to offer fluidised coal boilers on a commercial basis up to about 500,000 lb/hr. of steam. Utility boilers, e.g. 200MW and up, will have to wait a little longer.

About half the gas and oil used by our major fuel consuming industries - petroleum, petrochemical, paper, stone and ceramics, and primary metals - is burned to generate steam. More than 60% of this steam generation is produced in plants using 100,000 lb/hr. or more - i.e. a sufficiently large scale to justify the installation of a coal fired system.

We are therefore in a position, today, to release most of the gas and oil currently being used for industrial steam generation for other more valuable end-uses. Industrial steam generation is under much greater pressure than are the electric utilities to abide by strict pollution control regulations. Also, fluidised combustion's ability to use low-grade fuel, and to switch fuel sources, is of more importance to an industrial purchasing agent than it is to his counterpart in an electric utility. The utility will generally own large coal reserves or control them under long-term contract.

An additional feature in the industrial scene is our ability to retrofit existing coal-fired boilers to fluidised combustion. Our Renfrew boiler is an example. According to a 1973 survey by the Mitre Corporation there are about 124,000 stoker coal-fired industrial boilers in the United States. While there are definite limitations on the types of boilers that can be so converted, it is a viable method of gaining the advantages of fluidised combustion quickly and at low cost.

PRESSURISED APPLICATIONS

Development of pressurised fluidised combustion has not reached such an advanced stage as for the atmospheric process. The B.C.U.R.A. test rig has been operated at Leatherhead, England, at pressures up to 6 atm.

It has a distributor area of six square feet, a bed depth of four feet and is still the largest pressurised unit in operation.

An important series of tests on Illinois coals have been carried out on this rig, and the work has provided data for commercial plant design.

Many and varied are the schemes being put forward for the application of this pressurised process and all have one feature in common - the integration of fluidised combustion into a gas turbine cycle. The next step in development must be a convincing demonstration that reliable operation of a large utility gas turbine with fluidised combustion is practicable. This is fundamental to a whole chain of other developments leading, ultimately, to high efficiency combined cycle systems.

We have designed such a demonstration plant which would have a rated output of about 70MW electrical. In addition, it could produce a further 118MW of heat for process or district heating purposes.

Our design philosophy has been to seek an existing proven gas turbine design as closely compatible as possible with fluidised coal combustion. We wish to minimise gas turbine development at least until the basic demonstration of a unit has been achieved. For that reason we selected a turbine with a low turbine entry temperature, 1470-1540°F and a high pressure ratio, 16:1.

Our initial idea was to build a pilot plant at about 12MW. Such a plant would of course have virtually no use other than as a pilot plant, and its entire cost must be funded as a development expenditure. Another disadvantage is that utility companies in particular prefer to see a demonstration closer to 100MW before accepting new technology. This would require a further stage of demonstration, and could set true commercialisation back another four or five years.

So we realised that we could take a proven 70MW utility gas turbine, of a design thoroughly compatible with fluidised coal combustion, and sell it as a crude oil or residual oil-fired generator for industrial or small utility service. The cost of adding fluidised coal-firing facilities to this basic power plant is about the same as the entire cost of the 12MW pilot plant. The plant can be oil fired if modifications are necessary to the coal system. The design approach involves a reasonable scale-up, and by using a simple air heater cycle we minimise both particulate carry-over problems and control system development.

We therefore believe that if development money is to be spent on pressurised fluidised combustion, it should be spent so that the results are meaningful on a commercial scale. Also, we believe that this first plant should place simplicity and reliability before a complex and risky chase for ultimate efficiency. Steam generating beds and combined cycles involve significantly greater risks than the basic air cycle which we advocate as the first step.

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SO_2 Control Technologies - Commercial Availabilities and Economics

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SO_2 Control Technologies - Commercial Availabilities and Economics

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What Needs To Be Done Now?

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INTRODUCTION

The demand for electrical energy is projected to increase more than 117% between 1970 and 1985[1]. As a result, the combined usage of coal and oil to generate electricity will almost double in that fifteen year period. The international oil situation dictates that we minimize our dependence on foreign oil as an energy source and utilize our abundant coal reserves; however, this cannot take place without some environmental impact. For example, in 1974 the combustion of coal resulted in the emission of approximately 21 million tons* of SO_2 into the atmosphere[2]. By 2000, power plants alone will emit about 80 million tons of SO_2 per

*It is EPA policy to report measurements in the international system of metric units. For clarity of presentation, units used in this paper are those commonly used in engineering calculations in the U.S. Conversion factors are presented at the end of the paper.

year unless control technology is used to reduce the emissions[3]. Therein lies the complex energy/environment/economic problem: the urgent need to become a coal-fired economy while simultaneously safeguarding the quality of the environment at reasonable cost. The United States must resolve this problem as quickly as possible. This paper addresses the principal SO_2 control technologies that are expected to contribute to the solution. Emphasis is placed on the status of commercial availability and the costs of each alternative technology.

SO_2 Control Technologies

The principal technologies that will be available for continuous SO_2 emission control through 1980 are (1) the use of coal that occurs naturally with an acceptably low sulfur content, (2) physical and chemical coal cleaning, and (3) flue gas desulfurization (FGD) processes. In addition, other technologies currently being developed could make significant contributions to SO_2 control after 1980. These technologies include (1) coal gasification, (2) coal liquefaction, and (3) fluidized bed combustion of coal.

Low-Sulfur Coal

The Technology - Low-sulfur coal may be categorized as coal that occurs naturally with a sulfur content low enough that the coal can be burned without violating SO_2 emission standards. Thus, since SO_2 emission standards vary, the same coal may be considered "low-sulfur" for one application but not for another. The extent to which low-sulfur coal is used to meet national ambient air quality standards (NAAQS) and New Source Performance Standards (NSPS) will depend on the cost and availability of the coal, as well as the capability of a boiler designed for one type of coal to effectively burn a different type of coal. It should also be noted that SO_2 in combustion gases enhances the collection efficiency of electrostatic precipitators, so the use of low-sulfur coal at existing facilities could necessitate modifications to control particle emissions. Also, low-sulfur coals with low ash fusion temperatures may foul existing boilers, reducing capacity and requiring more frequent maintenance.

Commercial Availability - The availability of low-sulfur coal as a control technology depends on the expeditious development of our domestic resources and of the transportation network required to get the fuel to utility markets. In 1974, almost 390 million tons of coal were consumed by electric power plants; about half would comply with SO_2 emission regulations in effect on July 1, 1975[4]. In 1980, utility coal demand is projected to be about 620 million tons.* The basic question of availability, then, is how much of the 620 million ton demand can be supplied with low-sulfur coal. A tentative answer to that question is apparent in the October 1974 report developed by Sobotka and Company, Inc., under contract to EPA. The report suggests that low-sulfur coal production in 1980 could supply less than 44% of the 620 million ton demand. The actual production estimates from the report are summarized in Table 1.

*Basis: 8% annual growth rate for coal usage from 1974 through 1980.

Table 1

Potential Low-Sulfur[a] Coal Production in 1980[5]

Region	Millions of Tons
Appalachian	38
Midwestern	--
Montana/Idaho/Wyoming	194
Rocky Mountain	36
Pacific	--
TOTAL	268

a) Coal that would conform to New Source Performance Standards which limit emissions to 1.2 pounds of SO_2 per 10^6 BTU.

The production capacity for low-sulfur coal and its use in 1980 will, of course, depend on many complex factors, including:

(1) the relative economics of low-sulfur coal in comparison with other control approaches,

(2) political/economic decisions concerning low-sulfur coal recovery and distribution,

(3) the resolution of disputes over compliance schedules and state sulfur regulations, and

(4) the compatibility of low-sulfur coal usage with existing boilers and electrostatic precipitators and limitations on particulate emissions.

Thus, while the 1980 capacity projections may not be exact, it does seem safe to conclude that the supply of low-sulfur coal in 1980 will, at best, satisfy less than half of the total power plant demand.

Costs - The costs of complying with SO_2 standards by using low-sulfur coal include (1) the increased cost of the coal, (2) transportation cost, and (3) the cost of power plant modifications which may be necessary. While low-sulfur coal is generally more expensive than high-sulfur coal, the price differential currently is highly variable. (See Table 8.) Transportation cost often causes significant increases in low-sulfur coal costs. A year ago, some long-term leases for western coal were signed in the \$0.15 to \$0.30/10^6 BTU range, but transportation cost raised the delivered price to \$0.70 to \$1.10 per 10^6 BTU at an eastern plant[6]. In another case about a year ago, low-sulfur Wyoming coal was quoted at \$7/ton, FOB mine. In East St. Louis the delivered cost was about \$18/ton for coal rated at 6,500 to 7,500 BTU/lb. Since almost two tons of the Wyoming coal would be needed to equal the heating value of one ton of eastern coal, the price is raised to about \$36/equivalent ton in East St. Louis. The same equivalent ton of coal would cost \$45 or more in Philadelphia[7].

The capital cost of converting a plant from high-sulfur to low-sulfur coal averages approximately \$20/kw. However, the cost is highly variable depending on the original design of the boiler and the specific characteristics of the coal. This cost includes coal handling, combustion system modifications, and upgrading of the particle emission control system[8]. The incremental cost of compliance by conversion from high-sulfur coal to low-sulfur coal is also highly variable due to differences in coal costs and transportation costs. (See Table 8.)

Coal Cleaning

The Technology - Sulfur occurs in coal in two forms: as organic sulfur chemically bound to the hydrocarbon constituents and as inorganic pyritic and sulfate sulfur in the minerals associated with the coal bed.

For several decades, physical coal cleaning has been conducted using techniques based on differences in the physical properties of the coal and the inorganic impurities. Although many techniques have been evaluated[9], those that rely on differences in the specific gravities of the coal and the pyritic materials have been found to be best for commercial desulfurization[10]. In the cleaning process, the coal is crushed to free the inorganic impurities, sized by screening, and cleaned in equipment which is specifically designed for the size, shape, and specific gravity of the raw coal. Up to 80% of the pyritic sulfur and a significant portion of the ash can be removed by physical coal cleaning techniques[11]. Table 2 presents data on the percentages of various U.S. coal reserves that could be physically cleaned to meet the NSPS limitation for SO_2. It is significant to note that less than 13.5% of U.S. coal reserves can be cleaned to meet an SO_2 emission regulation of 1.2 lb/10^6 BTU[13].

Both inorganic and organically bound sulfur can be removed from coal by chemical cleaning. In chemical coal cleaning, finely ground coal is treated with a reagent under controlled conditions of temperature and pressure. The sulfur removal efficiency is dependent on coal particle size, the physical and chemical properties of the coal, the reagent used, the pressure, the temperature, and reaction time[14]. However, this technique is currently in its development stages and is not in commercial use as a coal desulfurization technology. EPA's Industrial Environmental Research Laboratory at Research Triangle Park, N.C., initiated its chemical coal cleaning development program in 1970. Presently, work is underway on the construction of a 250 pound/hr experimental test facility. Development work needed to bring this technology to the demonstration stage includes additional bench and pilot scale work to define the appropriate combinations of variables for optimum sulfur removal, to evaluate pollutant emissions characteristics, and to evaluate process economics.

Commercial Availability - As indicated in the technology description, physical coal cleaning has been commercially available for quite a few years; chemical coal cleaning is currently developmental and is not likely to be demonstrated on a large scale before 1980.

Most physical coal cleaning facilities are currently operating at or near capacity and are designed primarily for ash removal rather than sulfur removal[15]. Thus, expansion and modification of existing facilities or construction of new facilities will be required before significant quantities of cleaned coal could be delivered for use in power generation. In 1975, only a few companies were actively involved in the design and construction of coal cleaning plants. The design and construction of such a plant takes about 24 months, but this time could easily be lengthened if escalating demand overloads the present designers and equipment suppliers which are active in the field.

Table 2

Percentages of U.S. Coal Reserves Which Could Be Physically Cleaned To Meet NSPS For SO_2 Emissions[12]

Coal Category	Percentage That Could Be Cleaned To Meet NSPS
Lignite	0
Sub-bituminous	2.5
Bituminous	
Pa./W.Va./E.Ky.	5.0
Ill./Ind./W.Ky.	0
Western	2.8
All Other	3.2
TOTAL	13.5

Costs - The capital cost of a physical coal cleaning plant for sulfur removal depends on the type of coal to be cleaned, siting costs, and land costs. Such a plant may cost about $30,000 per ton of raw coal/hr (approximately $12/kw)[16]. This cost includes environmental controls, land for waste disposal, the cleaning plant (including thermal driers and water treatment), and cleaned coal handling facilities. Additional capital costs would be incurred if there were a need to upgrade the power plant's electrostatic precipitators. Operating costs range from $0.10 to $0.20/$10^6$ BTU[16] (approximately 1-2 mills/kwh). Additional costs which can be significant are incurred with carbon losses from the cleaning operation.

Flue Gas Desulfurization

The Technology - Flue gas desulfurization (FGD) is a postcombustion technique for removing SO_2 from combustion gases. FGD processes contact the combustion gases with a sorbent to absorb and react with the SO_2.

The numerous FGD processes are usually categorized as nonregenerable or regenerable processes depending on the fate of the reactive component in the sorbent. In nonregenerable processes the sorbent reacts with absorbed SO_2 and is not regenerated or reused. These processes produce a sludge which consists of a mixture of fly ash and water plus calcium sulfite and calcium sulfate which result from the SO_2/sorbent reactions. Several alternatives for disposition of the sludge are available. The alternatives include ponding of untreated sludge, landfilling of untreated and treated sludge, and commercial utilization. Among the factors to be considered in choosing a disposal alternative for a specific site are land availability, economics, and environmental effects.

Regenerable FGD processes are so named because the reactive compound in the sorbent is regenerated and reused in the process. These processes recover SO_2 from the combustion gases and convert it into marketable by-products such as elemental sulfur, sulfuric acid, or concentrated SO_2 gas.

The best developed and most common FGD processes are the nonregenerable lime and limestone scrubbing processes. In these processes, a lime or limestone slurry absorbs SO_2 from the flue gases, and subsequent reactions produce calcium sulfite and calcium sulfate. A variation of lime/limestone scrubbing, known as the "double alkali" process, uses a clear solution of a soluble salt rather than a slurry to absorb SO_2. Subsequent reactions to produce calcium sulfite and calcium sulfate sludge occur outside the absorber.

The most significant of the regenerable processes are magnesium oxide scrubbing and sodium sulfite scrubbing (the Wellman-Lord process). The magnesium oxide process utilizes an MgO slurry to absorb SO_2 and produce magnesium sulfite and magnesium sulfate. These products are

calcined to regenerate MgO which is reused in the absorption step. The Wellman-Lord process employs a clear solution of sodium sulfite to absorb SO_2 and react with it to form sodium bisulfite. The bisulfite-rich stream can then be processed to produce concentrated SO_2, sulfur, or sulfuric acid and regenerate the sodium sulfite.

Commercial Availability - In the U.S. the first full scale installations of the lime and limestone scrubbing processes were operated in the late 1960's, and the processes that evolved have generally been considered to be commercially available since 1974. Most of the problems that occurred in pilot, prototype, demonstration, and commercial facilities have essentially been overcome as better control of the complex chemical interactions has been achieved and as operational experience has accumulated, permitting improved design and construction.

Regenerable processes have not yet been widely tested on coal-fired boilers. The magnesium oxide process was tested on a 155 MW, oil-fired boiler at Boston Edison's Mystic Station, and the year-long program funded by EPA was concluded as scheduled in June 1974. In addition, a magnesium oxide scrubbing process was operated for about a year during 1974-75 at Potomac Electric's Dickerson Station on a nominal 100 MW boiler with EPA support of the regeneration facility until it was dismantled. Most of the operation at Dickerson was conducted without MgO regeneration. Another MgO system has been installed and operated briefly by Philadephia Electric on a 120 MW coal-fired boiler, and EPA is interested in supporting a test program at that facility.

The Wellman-Lord processes in Japan have cumulatively logged many years of effective operating experience on oil-fired boilers. In the U.S., construction of a Wellman-Lord/Allied Chemical system to service the nominal 100 MW coal-fired boiler at Northern Indiana Public Service Company's Mitchell Station has been completed. This facility, co-funded by EPA and NIPSCo, will log a year of coal-fired operating experience and undergo comprehensive testing by the end of 1977. Also, a western utility company has signed a contract for the construction of two additional systems to control 715 MW of generating capacity, and a letter of intent has been signed for two more Wellman-Lord systems to control another 1000 MW.

While the relationship between the orders placed for the various FGD processes and the commercial availability of the technology is not simple or direct, it is at least somewhat indicative of availability and of trends in utilization of the processes. Table 3 shows the electric generating capacity to be controlled by major FGD systems (systems larger than 50 MW) that will be operating or that are planned for start-up by the early 1980's. The table shows almost 5,000 MW of operational FGD, over 6,600 MW of FGD under construction, and almost 14,000 MW in the planning stages. If all FGD systems are included, the PEDCo survey shows that 108 FGD systems will be in operation by the early 1980's controlling 41,275 MW of generating capacity[18].

Table 3

Major FGD Applications[a] in the U.S.
by the Early 1980's[17]
MW (number of installations)

Non-Regenerable	Operational	Under Construction	Planned	Total
Lime	2064(7)	2436(6)	3135(7)	7635(20)
Limestone	2634(10)	4076(9)	5752(16)	12462(35)
Lime or Limestone	-	-	2500(4)	2500(4)
Subtotals	4698(17)	6512(15)	11387(27)	22597(59)
Regenerable				
Magnesium Oxide	275(2)[b]	-	726(3)	1001(5)
Wellman-Lord	-	115(1)	1715(4)	1830(5)
Subtotals	275(2)	115(1)	2441(7)	2831(10)
Totals	4973(19)	6627(16)	13828(34)	25428(69)

a) Includes systems of 50 MW or greater
b) Includes Boston Edison system which was shut down at end of test program

Costs - The economics of FGD are the most critical factor affecting its widespread use and also the factor that arouses the most debate. Problems of inflation, extent of redundancy, site-specific requirements, vendor optimism/user skepticism, variations in process options, variations in the kinds or quantities of by-products, differences in operating conditions, and other factors have contributed to the complexity of the situation. Figure 1 shows the changes that have occurred over the past nine years in TVA's estimates of the capital cost of the generalized limestone slurry scrubbing process. For a 500 MW application, the cost estimates have escalated from about $18/kw in 1968 to about $72/kw in 1977. Of course, this 300% increase is not wholly attributable to inflation; some of the increase is due to improvement in the knowledge of process requirements. Nevertheless, the figure does underline the difficulty of accurately projecting control costs over a period of a few years.

TVA has recently updated January 1975 generalized, comparative estimates of capital cost and annual revenue requirements for various FGD processes. Table 4 presents the capital costs and Table 5 presents the annual revenue requirements. The systems were estimated using 1973 technology under the premise that the technology was proven with no special provisions for redundancy or bypassing. For an installation designed to remove 90% of the SO_2 from a 500 MW new boiler fired with 3.5% S coal, a lime system is estimated to cost $61.10/kw, and a limestone system is about 12% more expensive at $68.40/kw. By comparison, the regenerable magnesium oxide scrubbing system at $71.70/kw is about 17% more expensive than lime scrubbing, and sodium sulfite scrubbing (Wellman-Lord) at $84.80/kw is about 39% more expensive than lime scrubbing. Table 5 indicates that the average annual revenue requirement for these four systems favors limestone scrubbing (3.41 mills/kwh) while lime, magnesium oxide, and sodium sulfite scrubbing are 7%, 18%, and 57% more expensive at 3.65 mills/kwh, 4.02 mills/kwh, and 5.37 mills/kwh, respectively. These estimates exclude sludge fixation costs and by-product credits. Comparisons of the costs of the four FGD processes for controlling capacities of 200 MW and 1000 MW can also be obtained from Tables 4 and 5.

Coal Gasification

The Technology - Coal to be used for gasification is initially crushed and screened before being transported to a series of gasifiers which operate at temperatures of 1000-3000°F and pressures from atmospheric to 1000 psi. The gasifiers produce a raw product gas containing carbon monoxide, carbon dioxide, hydrogen, methane, ethane, hydrogen sulfide, nitrogen, water, naptha, oil, tar, phenols, and other compounds plus trace elements. The raw product gas is cooled by heating water to generate steam, and quenched gas and by-product liquid streams are produced.

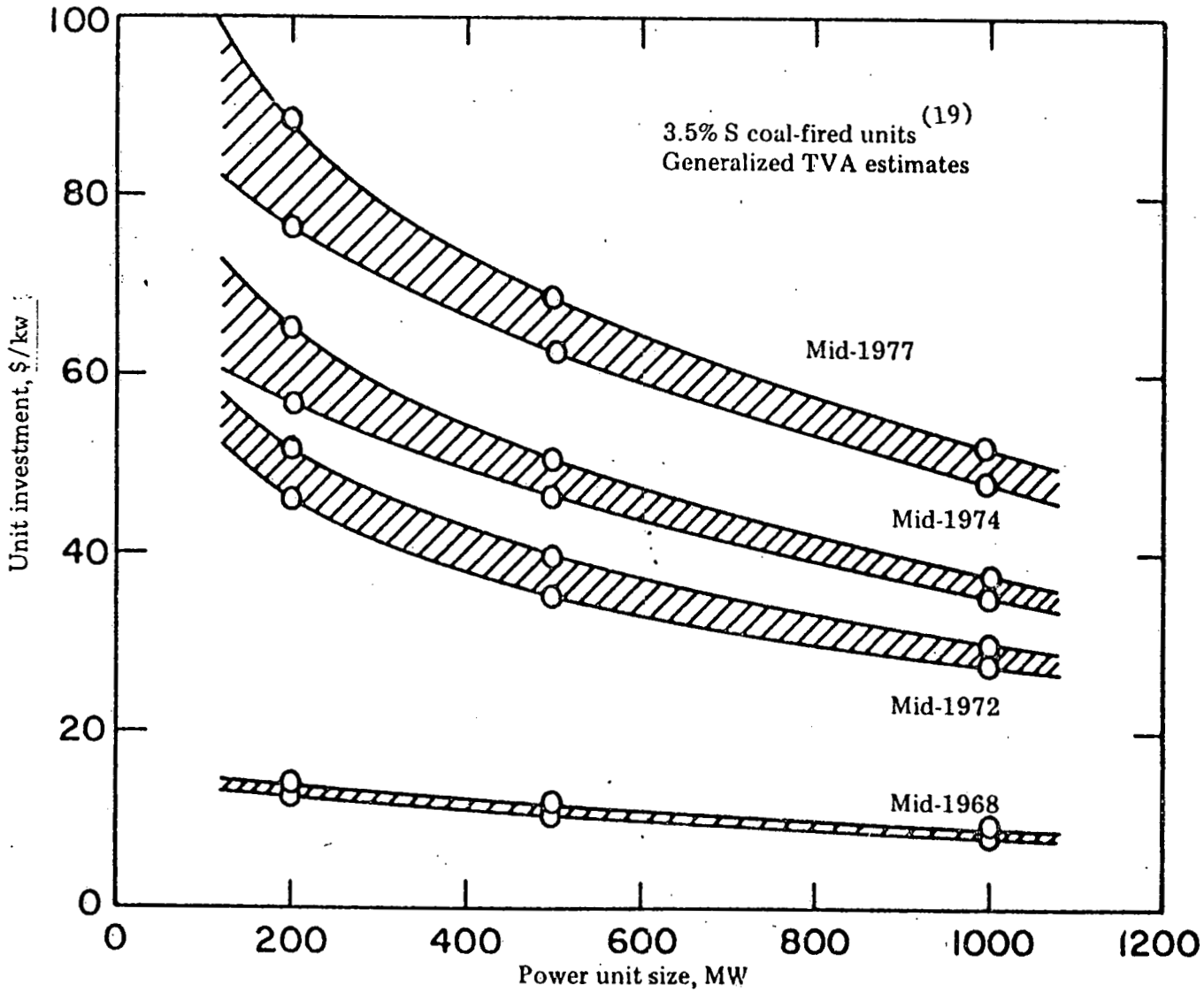


Figure 1. Limestone slurry scrubbing process - range of projected investment estimates.

Table 4

Summary of Total Capital Investment Requirements
Four Leading Flue Gas Desulfurization Processes[20]
1975-78 Cost Basis

Case	Years Life	Total capital investment[a,b,c] Limestone $/kw	Lime $/kw	Magnesia $/kw	Sodium Sulfite $/kw
Coal-fired power unit					
90% SO_2 removal; on-site solids disposal[d]					
200 MW, new, 3.5% S	30	88.4	79.9	95.6	112.0
500 MW, new, 3.5% S	30	68.4	61.1	71.7	84.8
1,000 MW, new, 3.5% S	30	51.4	44.9	53.0	64.2

a) Midwest plant location represents project beginning mid-1975, ending mid-1978. Average cost basis for scaling, mid-1977. Minimum in process storage; only pumps are spared. Investment requirements for disposal of fly ash excluded. Construction labor shortages with accompanying overtime pay incentive not considered.
b) These investment costs depend heavily on project definition.
c) Working capital has been included.
d) Sludge disposal pond with clay liner.

Table 5

Summary of Total Average Annual Revenue Requirements
Four Leading Flue Gas Desulfurization Processes(21)
1978 Cost Basis

Case	Years Life	Total average annual revenue requirements[a,b] Limestone Mills/kwh	Lime Mills/kwh	Magnesia Mills/kwh	Sodium Mills/kwh
Coal-fired power unit					
90% SO_2 removal; on-site solids disposal					
200 MW, new, 3.5% S	30	4.20	4.54	5.03	6.60
500 MW, new, 3.5% S	30	3.41 (4.01)[c]	3.65 (4.15)[c]	4.02 (3.23)[d]	5.37 (4.99)[e]
1,000 MW, new, 3.5% S	30	2.74	2.94	3.26	4.46

a) Power unit on-stream time, 7,000 hr/yr. Midwest plant location, 1978 revenue requirements. Investment and revenue requirements for disposal of fly ash excluded.
b) These revenue requirements reflect capital investments shown in Table 4 (updated); by-product credit and sludge fixation costs excluded.
c) Assumes sludge fixation service fee by contract treating sludge in the utility's pond at $10/ton of 100% solids.
d) Includes by-product credit for 100% H_2SO_4 at $25/ton.
e) Includes by-product credit for elemental sulfur at $40.18/ton.

In production of pipeline quality gas, the quenched gas stream is subjected to a catalyzed reaction in a shift converter to produce the appropriate CO/H_2 ratio for subsequent reaction to methane. After shift conversion and cooling and before the methanation reaction, the gas is purified by scrubbing to remove sulfur compounds and carbon dioxide. When the methanation reaction is completed, the product gases are compressed for distribution and use.

Depending on application and end product, specific gasification processes are usually described by one or more of the following terms: low BTU, intermediate BTU, high BTU, fixed bed, entrained bed, fluidized bed, slagging, air blown, oxygen blown, atmospheric, pressurized, retrofit, and combined cycle.

Gasifier systems like Lurgi, Winkler, Wellman-Galusha, Koppers-Totzek, HyGas, Bi-Gas, Synthane, CO_2-Acceptor, Texaco, Riley-Morgan, U-Gas, and COGAS are familiar to those interested in coal gasification. In addition, sulfur cleanup systems such as Stretford, iron oxide, Benfield, MDEA, molten salt, fluidized bed dolomite, Trencor, Selexol, Sulfinol, and Rectisol are also well known to the industry.

Commercial Availability - Considerable research, development, and engineering have already taken place in coal gasification, and some systems are commercially available and are being used around the world. Actual installations of the Lurgi, Winkler, Koppers-Totzek, and Wellman-Galusha gasifiers are too numerous for mention here and these are considered commercial. In utility applications, however, new combinations of these gasifiers, along with H_2S and particulate matter removal systems need to be coupled in practice and considerable additional research and development are required to reach commercial practice. The other gasification systems named fall under the classification of advanced technology and their final development is not expected for quite a while.

EPA's Industrial Environmental Research Laboratory at Research Triangle Park has studied and prepared reports on several high-BTU gasification processes. The Laboratory will also evaluate the environmental impact of emissions from the processes using an in-house test facility. One particular development of coal gasification currently receiving widespread attention by EPA and others is the application of gasification to combined cycle power generation. EPRI in combination with Commonwealth Edison and others is currently funding the design, construction, and testing of a fixed-bed, low-BTU, pressurized gasifier system combined with a gas turbine to produce 20 MW of power. This installation is part of a large proposed gasifier component test facility at Commonwealth Edison's Powerton Station in Pekin, Illinois. This facility, costing over $100 million, is expected to begin operation by 1978 with a three-year test program planned. ERDA is also presently negotiating participation in the program.

Costs - Economics of the leading coal gasification systems are available from a number of studies completed during recent years. TVA has conducted considerable work in evaluating the economics of coal gasification and H_2S removal systems. Results of evaluations for retrofitting power plants with low-BTU and intermediate-BTU gasifier systems are shown in Tables 6 and 7[22]. These costs are based on 1975 estimates. Additional economic data, covering only battery limits of H_2S removal systems for a 500 MW power unit with sulfur production of about 218 tons per day, indicate the following 1975 direct capital cost and annual revenue requirements for 7,000 hours/year of operation:

Process	Direct Capital Investment		Annual Revenue Requirements
	Without Claus	With Claus	
Stretford	$18,000,000	not needed	$3,400,000
Benfield	5,000,000	$10,000,000	3,600,000
Trencor		10,000,000	3,800,000
Selexol	12,000,000	17,000,000	3,500,000

Total project cost would be somewhat higher than the direct battery limit costs indicated.

Economics for new combined cycle plants depend a great deal on the state of the art of gas turbine operation[23]. For existing gas turbine technology where the maximum temperature is around 1950°F, coal gasification is more expensive than conventional boilers with FGD. If the advancement of the technology permits turbine temperatures up to 2500°F, the advantage could possibly switch to combined cycle units.

One major hurdle confronting application of coal gasification technology is financing for the highly capital intensive systems. Because the economics are presently questionable in comparison with other available fuel and control options, there is a reluctance on the part of private industry to pursue this technology. Several proposals for public financial backing have been sent to Congress by the Administration; to date, progress has been slow and the future remains uncertain. If the technology continues to advance with the development of gasifiers with greater throughput, better H_2S removal systems, and higher turbine temperatures, sufficient economic incentive could exist for the industry to use the technology without the need for public financial support.

Coal Liquefaction

The Technology - Liquid fuels can be produced from coal by four general methods. First, the coal can be pyrolyzed to break it down into a synthetic crude oil, a by-product gas, and a char residue. The char residue must be used in some way since it represents a significant fraction of the total heating value of the coal. Second, the coal can be

Table 6

Total Capital Investments of Coal Gasification
Systems For Retrofitting a 500 MW Power Unit
(Derated to 475 MW for the Low-BTU Gasification Systems)

	Total capital investment	
	10^6 $	$/kw
Fixed-bed, low-Btu gasification systems (475 MW)		
Wellman-Galusha/iron oxide	161.0	339
Wellman-Galusha/iron oxide/fines gasification	201.8	425
Lurgi/Benfield	211.3	443
Wellman-Galusha/Stretford	221.1	465
Lurgi/Stretford	234.8	494
Lurgi/iron oxide	234.9	495
Fixed- and entrained-bed intermediate-Btu gasification systems (500 MW)		
Oxygen-blown Lurgi/Benfield	213.1	426
Koppers-Totzek/MDEA	229.8	460

Table 7

Total Average Annual Revenue Requirements of
Coal Gasification Systems for Retrofitting a 500 MW Power Unit
(Derated to 475 MW for the Low-BTU Gasification Systems)

	Total average annual revenue requirements	
	10^6 $	mills/kwh[a]
Fixed-bed, low-Btu gasification systems (475 MW)		
Wellman-Galusha/iron oxide	38.9	11.71
Wellman-Galusha/iron oxide/fines gasification	48.3	14.53
Lurgi/Benfield	52.4	15.75
Wellman-Galusha/Stretford	52.9	15.91
Lurgi/iron oxide	55.4	16.65
Lurgi/Stretford	56.3	16.94
Fixed- and entrained-bed, intermediate-Btu gasification systems (500 MW)		
Oxygen-blown Lurgi/Benfield	52.9	15.12
Koppers-Totzek/MDEA	52.4	14.98

a) based on 7000 hrs of operation/year

dissolved in an organic solvent, treated with hydrogen, and filtered to remove the non-combustible minerals. The combustible fraction filtered out with the non-combustibles must be reclaimed and used if the process is to be economically competitive with alternative technologies. Third, the coal can be gasified to produce carbon monoxide and hydrogen which can subsequently be converted to methanol or other liquid hydrocarbon fuels. Fourth, pulverized coal can be contacted with hydrogen in the presence of catalysts to form fractions of liquid and gaseous hydrocarbons.

It would appear that the pyrolysis processes are more suited to multiproduct coal refinery concepts whereas the dissolution schemes could be designed specifically as sources for boiler fuel. The coal dissolution processes can be characterized by the use of direct or solvent supplied hydrogenation to decrease the carbon to hydrogen ratio to about 6-7 to 1 for liquefaction (normally 15 to 1 for bituminous coal)[24]. Systems that have been studied intensively are COED, Occidental, and Toscoal which are pyrolysis processes and solvent refined coal (SRC), H-Coal, and Synthoil which are considered to be hydrogenation processes.

Commercial Availability - Of these processes, the SRC and COED systems apparently have been developed further and have more operating experience than the others. The SRC system has been operated since 1973 at the 6 ton/day level in a Wilsonville, Alabama, prototype unit sponsored by Southern Services, Inc., and EPRI. A larger unit processing 50 tons/day has been in service since 1974 at a facility near Tacoma, Washington, sponsored by ERDA. The COED unit, a development of FMC Corporation, is located at Princeton, New Jersey, and processed about 36 tons of coal per day. The COED development program met its objectives and was terminated in 1974. The SRC process produces a clean (0.9% S and low ash) coal residue as a fuel with some net liquid output and the COED process produces a fuel gas, a syncrude liquid, and a char which can all be used for fuel after desulfurization.

Development of the H-Coal process is now beginning to accelerate rapidly with Hydrocarbon Research, Inc., and ERDA currently building a 600 ton/day plant at Catlettsburg, Kentucky. This unit is expected to be started up in March 1978 with a 2-year test program planned. H-Coal produces liquid hydrocarbons which can be upgraded to higher quality fuels.

Synthoil is a process under development by the Bureau of Mines using mild hydrogenation in a reactor in which slurried coal is pushed by hydrogen through a fixed bed of catalyst. The prototype located at Pittsburgh, Pennsylvania, will convert 10 tons of coal per day into 30 barrels of oil per day. Startup of this process is expected about September 1977.

The Occidental coal pyrolysis process has also shown considerable promise. The 3-1/2 tons/day test unit at La Verne, California, has been operated to produce a crude liquid hydrocarbon or gaseous fuel and a char which can be burned to provide fuel for the pyrolysis step.

The three processes probably nearest to commercial application are the SRC, H-Coal, and COED systems. In the case of SRC, consideration has been given to building a commercial-size plant (2000 ton/day range); however, details have not been made public at this time. With adequate funding, commercial coal liquefaction systems could be available in the mid-1980's.

Costs - The economic aspects of coal liquefaction are not well defined. Some studies were made earlier in the development of the technology, but these estimates were optimistic and are generally not regarded as accurate. The preparation of additional estimates is now underway on many of these systems; Ralph M. Parsons is currently preparing a detailed evaluation of SRC for EPRI. In general, the cost of coal liquefaction is comparatively high; base load power generation using fuels from these processes is probably not economically competitive with low-BTU coal gasification, fluidized bed combustion, or conventional boilers with FGD. Coal liquefaction will undoubtedly find application for peaking service such as providing gas turbine fuels, where it is much more competitive with other technologies.

The limited cost data available indicate that fuels from a 20,000 ton/day coal liquefaction unit will cost about \$3.50-\$5.00/10^6 BTU using coal at \$20.00/ton. This quantity of coal would supply a 1500-2000 MW unit. Fuel cost to a base load boiler would be about 33 mills/kwh.

Fluidized Bed Combustion

The Technology - In the fluidized bed combustion process, coal is burned in a bed of granular, non-combustible material which serves as the reactant for the capture of sulfur dioxide. Air is passed up through a grid plate beneath the bed to fluidize the bed and to support combustion of the coal. Coal is injected near the base of the fluidized bed, above the grid plate. The coal burns rapidly in the bed, and the carbon content of the bed during operation is usually less than one percent by weight. The bed temperature, usually controlled by a steam-generating heat exchanger placed in the bed, is maintained between 1400 and 2000°F, well below the furnace temperature in conventional, pulverized-coal boilers.

Limestone, dolomite, or coal ash may serve as the non-combustible bed material and react with SO_2 in the combustion zone. The spent bed material may be continuously withdrawn, discarded, and replaced ("once-through") or continuously regenerated to recover sulfur and produce

reusable sorbent ("sorbent regenerative"). Flue gases leaving the bed pass over a heat transfer surface and through a gas turbine (high pressure boilers) before being exhausted.

Fluidized bed combustion (FBC) is one of the most promising routes to the effective use of coal as a boiler fuel. This process offers great promise for (1) increasing boiler heat flux and combustion rate thereby reducing vessel size, (2) using lower grade coals (reducing the cost of fuel), and (3) controlling particulate, SO_2, and NO_x emissions.

Commercial Availability - The FBC process is being studied intensively by several groups including boiler manufacturers such as Foster Wheeler Corporation, Babcock and Wilcox Company, and Combustion Engineering. Research and design activities are also underway at Argonne, Battelle, Westinghouse, General Electric, Curtiss-Wright, Combustion Power Company, Combustion Systems Ltd., Exxon, and Pope, Evans and Robbins. Both private and public funds are being used in the development of the technology with ERDA (which has primary responsibility for the development of FBC for the government), EPA, and EPRI all contributing to various phases of work.

EPA has a number of projects under contract to evaluate the environmental requirements and economics of FBC technology[25]. TVA is evaluating for EPA the chemistry and physical properties of the solid effluent from FBC beds and the economics of FBC systems as compared to conventional boilers with FGD; for ERDA, TVA is evaluating the agronomic potential of the bed material as a lime and sulfur source for crops.

There are two basic designs for fluidized bed boiler systems: atmospheric fluidized bed (AFB) and pressurized fluidized bed (PFB) combustion. Of the two, AFB systems are much further advanced with a 30 MW prototype unit scheduled for startup during the summer of 1976 at the Rivesville Station of Monongahela Power Company. The successful operation of the 30 MW prototype will be a major milestone in the orderly development of the technology, but it must be demonstrated at a larger scale before commercial availability can be claimed. ERDA currently has plans to design, construct, and operate a unit in the 200 MW range in the early 1980's; even so, some groups have shown an interest in trying to speed up the schedule and are pushing for a faster pace. Prospects are good for commercialization of this technology by 1984-1986.

PFB boilers are definitely in an earlier stage of development with several key problems yet to be resolved. Particle emission control, gas turbines, and solids handling will need considerable additional development if this technology is to emerge in the next few years. As compared with conventional and AFB units, PFB has the advantage that higher overall boiler-turbine efficiencies of 39-40% can be achieved whereas 36% is achieved with AFB, and 32-36% is achieved with conventional units.

Both AFB and PFB pilot studies indicate excellent SO_2 removal efficiencies. Limestone addition to AFB units at two-to-four times stoichiometric amounts has shown SO_2 removal efficiencies of 90% and better. In PFB units, SO_2 removal efficiencies of 90% or better have been achieved using dolomite to provide Ca/S ratios of 1.5 to 2. Similar usage of limestone results in removal efficiencies around 70%.

Currently the PFB units are in the bench scale and pilot plant stage; the largest units are the 0.63 MW miniplant which has been built at Exxon Research and Engineering, a unit of similar size at the British Coal Utilization Research Association (BCURA), and a 1.8 MW unit under study at Combustion Power Company. ERDA has plans to fund a 14 MW unit which Curtiss-Wright will construct and test in the early 1980's. Commercialization could come in the late 1980's.

Costs - Economic data for AFB and PFB systems are just now beginning to emerge as developers evaluate the technology. Because the entire boiler-turbine-environmental control system must be evaluated as a unit, comparisons with other technologies must be made with care. One difficulty in making such comparisons is the state of development for FBC technology. Contingencies for FBC systems obviously should be higher than for conventional boilers with FGD.

Cost projection results to date indicate that AFB systems may be less expensive than PFB as well as conventional boilers with FGD. Capital costs for FBC systems are available, though details are lacking. Because of this, the confidence levels of most estimates are not very high. Even so, considering that conventional coal-fired boilers with turbines will cost $500-700/kw and FGD systems probably $100/kw additional, projected costs for AFB systems are competitive at $632/kw[26]. PFB systems at $723/kw[26] appear more costly than AFB systems. From an operating standpoint, in one unpublished study, power from an atmospheric fluidized bed unit is predicted to cost around 25-35 mills/kwh whereas PFB and conventional boilers with FGD systems may be 3-8 mills/kwh higher. Additional economic data probably will be available soon so that a more meaningful comparison can be made[27].

Comparison of SO_2 Control Alternatives

The projected increase in the use of coal over the next quarter of a century mandates the widespread application of SO_2 control technologies, singly and in combination, to prevent extensive degradation of the environment. Thus, an SO_2 control program which permits timely and economic attainment of the regulatory requirements is essential. As suggested earlier, such a program may comprise various control options including naturally occurring low-sulfur coal, coal cleaning and flue gas desulfurization separately or in combination, coal gasification, coal liquefaction, fluidized bed combustion of coal, and others.

The choice of an appropriate SO_2 control strategy for a specific site is often complex due to the interaction of such factors as commercial availability of competing systems, reliability and operability of the systems in other applications, capital costs and annual revenue requirements, fluctuation in the availability and cost of low-sulfur fuels, and uncertainty about pollution emission regulations. Table 8 summarizes the commercial availability and cost data for the major SO_2 control alternatives presented earlier. These data are intended only to give a general comparison of relative costs; it is unfortunately necessary to so qualify the table since economic and availability data were obtained from several different sources and the degree of refinement of such data varies from technology to technology depending on the status of development. Using the available data, Figure 2 shows the effect of the application of various control technologies on the total cost of power. In using the data presented it is important to remember that the costs for the technologies are not simply and directly comparable and that the relationships depicted for the various technologies will vary as better, more accurate cost data are accumulated through additional experience with the technologies.

Table 8

Availability/Cost Estimates for SO_2 Control Alternatives

SO_2 Control Alternative	Date of Commercial Availability	Potential Applicability by 1985, % of Total Power Plant Fuel			Capital Costs $/kw	Coal Cost, Mills/kwh	Control Technology Annual Revenue Requirement,[a] Mills/kwh	Effective Total Cost of Fuel (Coal Cost + Annual Revenue Requirement), Mills/kwh	Average Total Cost of Power Mills/kwh
		Retrofit	New Boilers	Total					
Conventional Coal-Fired Boiler	--	--	--	--	500-700	7.9[b]	16.1-31.1[c]	24-39[d]	31.5
High-Sulfur Coal	--	--	--	--	--	7.9[b]	--	7.9	--
Low-Sulfur Coal	Current	--	--	<50	20	8[e] 14.4-17.2[f] 20-25[g]	0.5	8.5 14.9-17.7 20.5-25.5 (Ave. =16)	39.6
Physical Coal Cleaning	Current	--	--	<14	12	7.9[b]	1-2	8.9-9.9	33.0
Chemical Coal Cleaning	1980	30	30	30	75(28)	7.9[b]	4.5[h]	12.4	36.0
Flue Gas Desulfurization[i]									
Limestone	Current	70	100	85[j]	54.7[k]	7.9[b]	3.21[k]	11.1	34.7
Lime	Current	70	100	85[j]	48.9[k]	7.9[b]	3.32[k]	11.2	34.8
Magnesium Oxide	Current	70	100	85[j]	57.4[k]	7.9[b]	2.58[k]	10.5	34.1
Sodium Sulfite	Current	70	100	85[j]	67.8[k]	7.9[b]	3.99[k]	11.9	35.5
Advanced Processes[l]	1981	--	--	--	65-95	7.9[b]	2.5-3.0	10.4-10.9	34.3
Coal Gasification	1980-85	--	--	<10[m]	339-495	7.9[b]	6.8-7.6	14.4-15.5	38.7
Coal Liquefaction	1980-85	0	0	0	(n)	7.9[b]	(n)	(n)	--
Fluidized Bed Combustion									
Atmospheric	1984-86	0	0	0	632(26)	7.9[b]	17.1-27.1[c]	25-35	30.0
Pressurized	mid-1980's	0	0	0	723(26)	7.9[b]	20.1-35.1[c]	28-43	35.5

a) Basis: 7000 hr of operation/yr; includes labor, raw materials, maintenance, utilities, overhead, amortization and return on rate base (allowable profit). Includes sludge fixation costs of $10/ton of 100% solids for limestone and lime processes, a by-product credit for magnesium oxide of $25/ton of 100% H_2SO_4, and a by-product credit for sodium sulfite of $40.18/ton of elemental sulfur

b) Basis: $21/ton(21); 12,000 BTU/lb; 1.33 kwh/lb; 38% conversion efficiency; 3-3.5% S

c) Basis: Obtained by subtracting coal cost from effective total cost of fuel

d) Basis: AFB at 25-35 mills/kwh; boiler plus scrubber 3-8 mills/kwh higher (28-43 mills/kwh); FGD average (excluding advanced processes) 4 mills/kwh

e) Basis: $14.40/ton ($0.90/10^6 BTU(6)); 8,000 BTU/lb; 0.9 kwh/lb; includes transportation costs

f) Basis: $26-31/ton(21); 8,000 BTU/lb; 0.9 kwh/lb; includes transportation costs

g) Basis: $36-45/ton(7); 8,000 BTU/lb; 0.9 kwh/lb; includes transportation costs

h) Basis: Derived from an equivalent cost of $12 per ton of coal processed(27)

i) All costs given for FGD are for a new 500 MW, coal-fired power unit burning 3.5% S coal with 90% SO_2 removal

j) This is an "if necessary" number; obviously other technology alternatives will meet some of the needs

k) Reduced by 20% from costs presented in Tables 4 and 5 to convert costs to 1975 basis; other costs in Table 8 are based on 1975 dollars

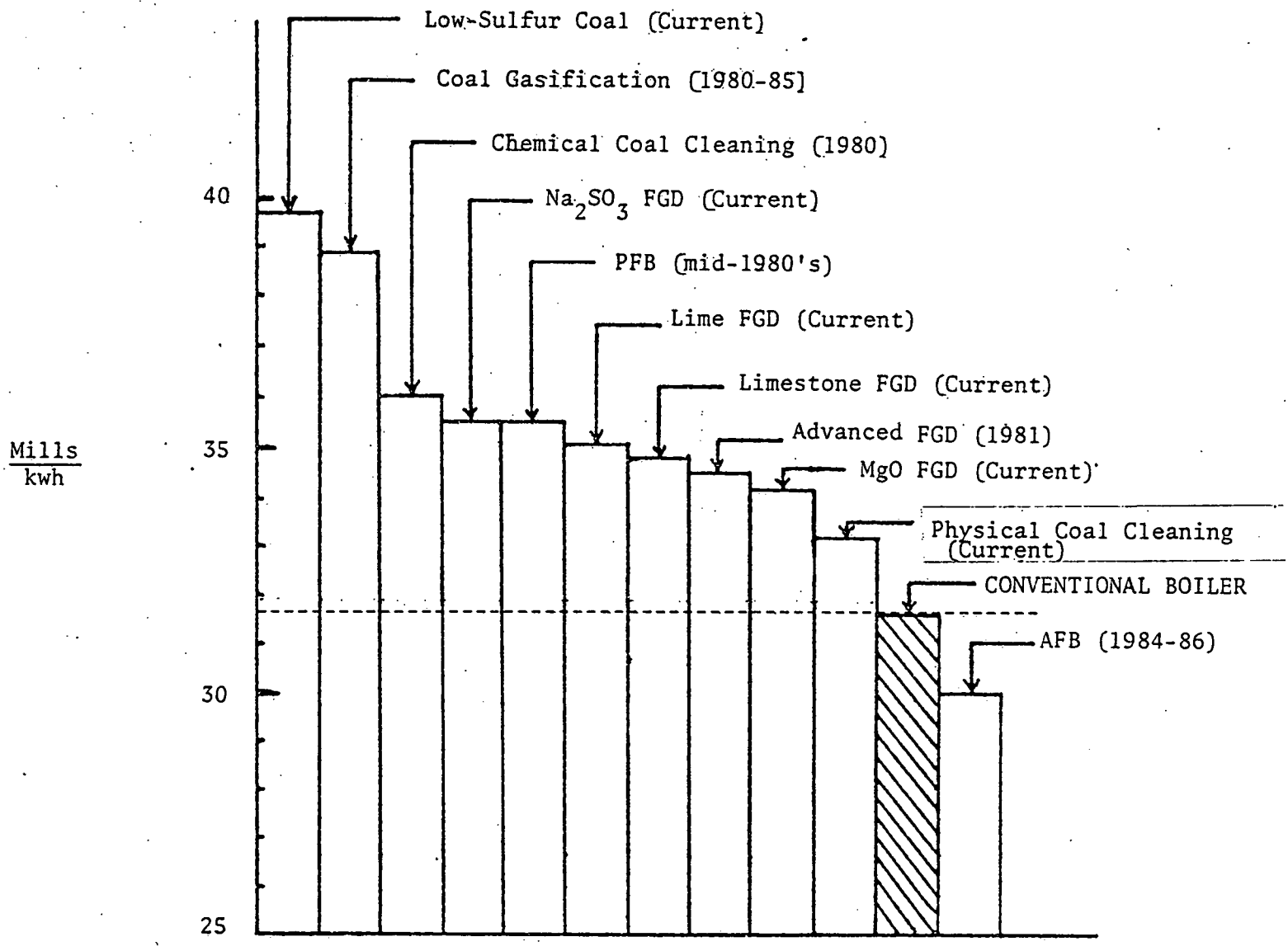
l) Cost ranges given for advanced FGD processes are based on preliminary cost projections for the aqueous carbonate process, the double alkali process, and the citrate process all of which will be demonstrated at full scale through jointly funded EPA/utility or EPA/USBM programs

m) Based on 1973 Mitre study with Energy/Fuel Projections to 2000

n) No recent data currently available; expected to be equal to or greater than the range for gasification

Figure 2

Effect of Control Technology on
Total Cost of Power
(Availability)



Conclusions

As suggested earlier, SO_2 emission control technologies can essentially be subdivided into two categories: (1) those that are currently available and will be in widespread use by 1980 and (2) those which are expected to achieve significant use after 1980. But commercial availability of control technologies is only one of many complex factors that will interact to determine the nature and extent of the use of the technologies for SO_2 control. In addition to commercial availability and economics--the primary emphases of this paper--consideration must be given to the types and quantities of energy required in the processes, the comparative energy penalties associated with competing processes, the actual and potential environmental effects of extensive utilization of the processes, and the variation in control regulations throughout the country.

While many factors may affect the eventual balance of usage between the competing control technologies, commercial availability and economics are so dominant in evaluating and comparing control technologies that valid conclusions can be drawn from the foregoing discussion of these two essential factors. Among these conclusions are:

1. Essentially all coal-fired electric power plants could achieve compliance operation between now and 1985 through the application of currently available continuous SO_2 emission control technologies, singly or in combination.

2. Between now and 1985, the availability of low-sulfur coal supplies will restrict use of this fuel to less than half of the total projected power plant coal usage. In addition, based on the average cost from Table 8, the effective total cost of low-BTU, low-sulfur coal is about double the normal high-sulfur coal cost. It is likely that this disparity will grow substantially if low-sulfur coal is substituted for presently used high-sulfur coal.

3. Between now and 1985, the only significant control alternatives permitting compliance with current SO_2 regulatory requirements are the use of low sulfur coal, coal cleaning, and the application of FGD technology (or FGD in combination with coal cleaning).

4. Capital costs associated with FGD are 3-4 times those associated with low-sulfur coal usage, but, based on the average cost from Table 8, the effective total cost of fuel is about 45% greater for low-sulfur coal than for FGD.

5. In comparison to the use of a single approach, FGD/coal cleaning and FGD/low-sulfur coal combinations may be economically and environmentally superior for achieving compliance operation between now and 1985.

6. Coal gasification and liquefaction technologies, projected to be available for commercial application in the early- to mid-1980's, are expected to be comparatively expensive.

7. Fluidized bed combustion technology is expected to be economically competitive with conventional boilers with FGD; commercial availability is expected about 1985 or later.

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CONVERSION FACTORS

Although EPA's policy is to use the metric system in all of its reports, this paper uses units that are commonly used in engineering calculations in the U.S. The conversion factors needed to convert the units used in the paper to metric units are presented below.

1 kg = 2.2 lbs

1 kg = 0.0011 ton

$$°C = \frac{°F-32}{1.8}$$

kcal = 3.97 BTU

APPLICABILITY OF THE MEYERS PROCESS FOR DESULFURIZATION OF U.S. COAL (A SURVEY OF 35 COAL MINES)

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for presentation at the

Third Annual International Conference on

Coal Gasification and Liquefaction

University of Pittsburgh
August 3-5, 1976

EV 77-X-05-0840 Rec'd 5-11-77

APPLICABILITY OF THE MEYERS PROCESS FOR DESULFURIZATION OF U. S. COAL (A SURVEY OF 35 COAL MINES)

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The Meyers Process (1, 2) utilizes a regenerable aqueous ferric sulfate leaching unit to chemically convert and remove the pyritic sulfur content of coal as elemental sulfur and iron sulfate. In addition, the ash content of the coal is decreased by 10 to 50 percent and as a result the heat content per unit weight increases by as much as 13%.

In this process, aqueous ferric sulfate (a mild but effective oxidizing agent) selectively oxidizes the pyritic sulfur content of coal to form free sulfur and sulfate (Equations 1 and 2). Iron sulfate dissolves in the aqueous solution. The free sulfur may then be removed from the coal matrix by vaporization or solvent extraction (Equation 3) and the oxidizing agent may be regenerated and recycled by reaction with air or oxygen (Equation 4).

$$2Fe^{+3} + FeS_2 \rightarrow 3Fe^{+2} + 2S \quad (1)$$

$$14Fe^{+3} + 8H_2O + FeS_2 \rightarrow 15Fe^{+2} + 2SO_4^{-2} + 16\,H^{+} \quad (2)$$

$$S \cdot Coal \longrightarrow S + Coal \quad (3)$$

$$3Fe^{+2} + 3/2\,[O] \rightarrow 3Fe^{+3} + 3/2\,[O^{-2}] \quad (4)$$

The detailed chemistry, reaction kinetics and process engineering and cost estimates have been published (3-7) or are in preparation (8) and a test plant is being built. This paper presents a discussion and experimental results directed to the applicability of the process for meeting pollution control standards by desulfurization of raw run-of-mine coal. This work was sponsored by the Environmental Protection Agency (9, 10).

A summary of sulfur analyses of run-of-mine coals utilized in this study is shown in Figure 1. The sulfur content of the coals is reasonably representative of the sulfur distribution in U. S. coal.

A summary of the results to date for chemical removal of pyritic sulfur and the optimal results for conventional coal washing (float-sink evaluation) are shown in Table 1 in terms of total sulfur changes. The table describes the results obtained on coals which contained sufficient pyritic sulfur for accurate sulfur removal determination (i. e. > 0. 25% w/w). Three of the mines sampled, the Edna, Belle Ayr and Walker mines were below this limit and therefore do not appear in the table. Actual total sulfur values before and after chemical removal are shown in Columns 4 and 5. These may be compared with Column 6, which shows sulfur values

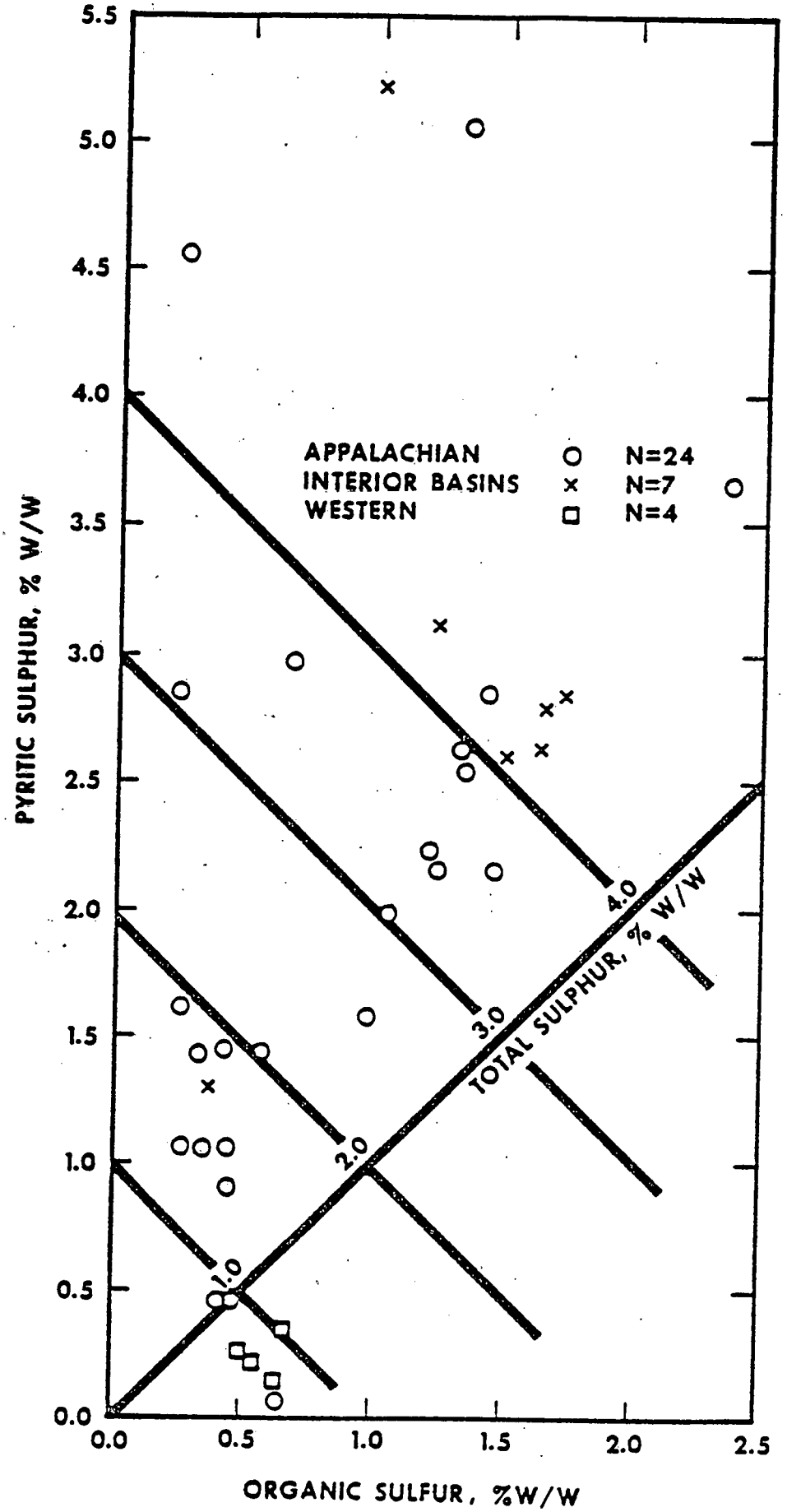


FIG. 1. Distribution of sulfur forms (dry moisture free basis) in run-of-mine U. S. coals

Table 1

SUMMARY OF PYRITIC SULFUR REMOVAL RESULTS

Mine	Seam	State	% Total Sulfur w/w in Coal[a]			Meyers Process Pyrite Conversion % w/w	Meyers Process Total Sulfur Decrease % w/w	% Sulfur in Coal[b] After Float-Sink
			Initial	After Meyers Process				
				Current Results	95% Removal[c]			
Navajo	Nos. 6, 7 & 8	N. Mexico	0.8	0.6	0.5	90	25	---
Kopperston No. 2	Campbell Creek	W. Virginia	0.9	0.6	0.5	92	33	0.8
Harris Nos. 1 & 2	Eagle & No. 2 Gas	W. Virginia	1.0	0.8	0.5	94	23	0.9
Colstrip	Rosebud	Montana	1.0	0.6	0.7	83	30	---
Warwick	Sewickley	Pennsylvania	1.4	0.7	0.3	92	54	1.0
Marion	Upper Freeport	Pennsylvania	1.4	0.7	0.5	96	50	1.2
Matnies	Pittsburgh	Pennsylvania	1.5	0.9	0.5	95	36	1.7
Isabella	Pittsburgh	Pennsylvania	1.6	0.7	0.6	96	54	1.5
Orient No. 6	Herrin No. 6	Illinois	1.7	0.9	0.4	96	44	1.4
Lucas	Middle Kittanning	Pennsylvania	1.8	0.6	0.4	94	64	0.7
Jane	Lower Freeport	Pennsylvania	1.8	0.7	0.5	91	63	0.8
Hartinka	Lower Kittanning	W. Virginia	2.0	0.6	0.7	92	70	0.8
North River	Coruna	Alabama	2.1	0.9	0.7	91	55	2.2
Humphrey No. 7	Pittsburgh	W. Virginia	2.6	1.5	1.1	91	42	1.9
No. 1	Mason	E. Kentucky	3.1	1.6	1.2	90	48	2.3
Bird No. 3	Lower Kittanning	Pennsylvania	3.1	0.8	0.4	96	75	1.5
Williams	Pittsburgh	W. Virginia	3.5	1.7	1.4	96	50	2.3
Shoemaker	Pittsburgh	W. Virginia	3.5	1.7	1.4	96	51	3.6
Meigs	Clarion 4A	Ohio	3.7	1.9	1.6	93	48	2.8
Fox	Lower Kittanning	Pennsylvania	3.8	1.6	0.8	93	57	2.0
Dean	Dean	Tennessee	4.1	2.1	1.6	94	49	3.0
Powhattan No. 4	Pittsburgh No. 8	Ohio	4.1	1.9	1.7	99	53	3.1
Eagle No. 2	Illinois No. 5	Illinois	4.3	2.0	1.8	94	54	2.9
Star	No. 9	W. Kentucky	4.3	2.5	1.9	91	43	3.0
Robinson Run	Pittsburgh	W. Virginia	4.4	2.2	1.6	97	50	3.0
Homestead	No. 11	W. Kentucky	4.5	2.4	1.5	93	47	3.2
Camp Nos. 1 & 2	No. 9 (W. Ky.)	W. Kentucky	4.5	2.0	1.8	99	55	2.9
Ken	No. 9	W. Kentucky	4.8	2.8	2.1	91	42	3.5
Belmont	Upper Freeport	Pennsylvania	4.9	1.0	0.6	96	80	2.1
Muskingum	Meigs Creek	Ohio	6.1	3.2	2.6	94	47	4.4
Weldon No. 11	Des Moines No. 1	Iowa	6.4	2.2	1.4	92	65	3.9
Egypt Valley No. 21	Pittsburgh No. 8	Ohio	6.6	2.7	1.7	93	59	4.6

[a] Dry, moisture-free basis.

[b] 1.90 Float material, 14 mesh x 0, is defined here as the limit of conventional coal cleaning (See Section 4.4)

[c] Sulfur content of coal at 95% pyrite removal and no increase in sulfate or measured organic sulfur content.

which could be obtained for full process optimization (at 95 percent pyrite removal with no increase in starting sulfate or measured organic sulfur content). Thus, for example, although 99 percent pyrite conversion was obtained for Camp Nos. 1 and 2 mines, the total sulfur was reduced to 2.0 percent, not the theoretical 1.8 percent, due to a slight measured increase in other sulfur forms.

Because of the widespread application of physical cleaning techniques for removal of non-combustible rock (which includes varying amounts of pyrite) from coal (along with some carbon), float-sink fractionation was performed in order to define the relative utility of washing and chemical desulfurization for each coal. The results which are shown in Table 1, indicate that: a) the Meyers Process, at its current state of development, removed 83 to 99 percent of the pyritic sulfur content of the 32 coals studied, resulting in total sulfur content reductions of 25 to 80 percent, b) eleven (34 percent) of the coals were reduced in sulfur content to the 0.6 to 0.8 percent sulfur levels generally consistent with the federal standard for New Stationary Sources and many state standards, c) in all cases, the Meyers Process removed significant to very large increments of sulfur over that separable by physical cleaning, and d) in two cases, the Corona and Mathies mines, coal cleaning resulted in a sulfur content increase.

State emission regulations for discharge of sulfur oxides from utility and large industrial power plants can also be met by application of the Meyers Process. For example, the Pennsylvania state standard for eight air basins is approximately 1.1 percent sulfur, for coal of 25 x 10^6 btu/ton. Several of the tested Pennsylvania coal mines (Marion, Mathies, Isabella, Bird No. 3 and Delmont) meet this standard after chemical desulfurization but do not meet the standard after efficient physical cleaning. These coals could also be transported to Michigan, New Jersey or New York to meet their state standards of approximately 1.0 percent and 1.3 percent and 2.4 percent sulfur, respectively. Two of the Ohio coal mines (Meigs and Powhattan No. 4) would meet the "28 county standards" of approximately 2 percent sulfur for the state of Ohio after treatment by the Meyers Process, whereas efficient cleaning of these coals reduces their sulfur content to only 2.8 percent and 3.3 percent.

The Orient No. 6 mine of Illinois meets the Chicago area standard of 1.29 percent sulfur after chemical desulfurization but does not meet the standard after physical cleaning. The Camp Nos. 1 and 2 mines in Western Kentucky meet the state standard for "Priority 3" regions of less than 2.3 percent sulfur after treatment by the Meyers Process, whereas physical cleaning reduces the total content of this coal to 2.9 percent. The Humphrey No. 7 mine is reduced to 1.5 percent sulfur, which meets the West Virginia standards for "Regions 2 and 3" of 1.7 and 2 percent respectively, whereas physical cleaning reduces the sulfur content to 1.9 percent. The Weldon mine in Iowa is reduced to 2.2 percent sulfur by the Meyers Process which meets the state requirement of approximately 3.1 percent sulfur. Physical cleaning does not meet the standard, reducing the sulfur content to 3.9 percent.

Process improvements, such as more efficient residual sulfur and sulfate removal, will cause most coals to be further reduced in sulfur

content to the "95 percent removal" level shown in Column 6 of Table 2.

In the production of clean fuel using commercial practices, it is very likely that an optimum process cost and product will be obtained by physically cleaning coal prior to ferric sulfate leaching, in order to remove rock and some of the larger pyrite particles. There are preliminary indications that the efficiency of the Meyers Process may be enchanced by utilization of physically cleaned coal, resulting in faster rates, greater total removal, and reduced ash dissolution

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SYNTHOIL PROCESS

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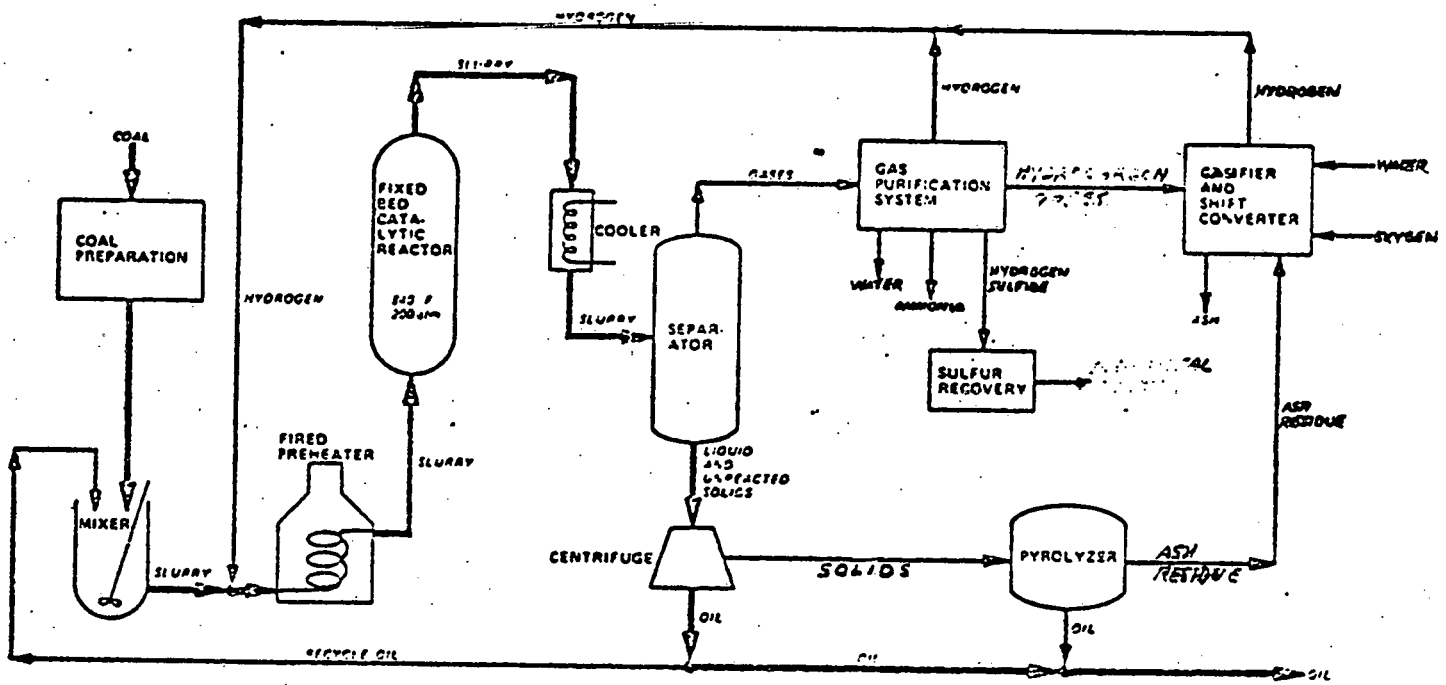
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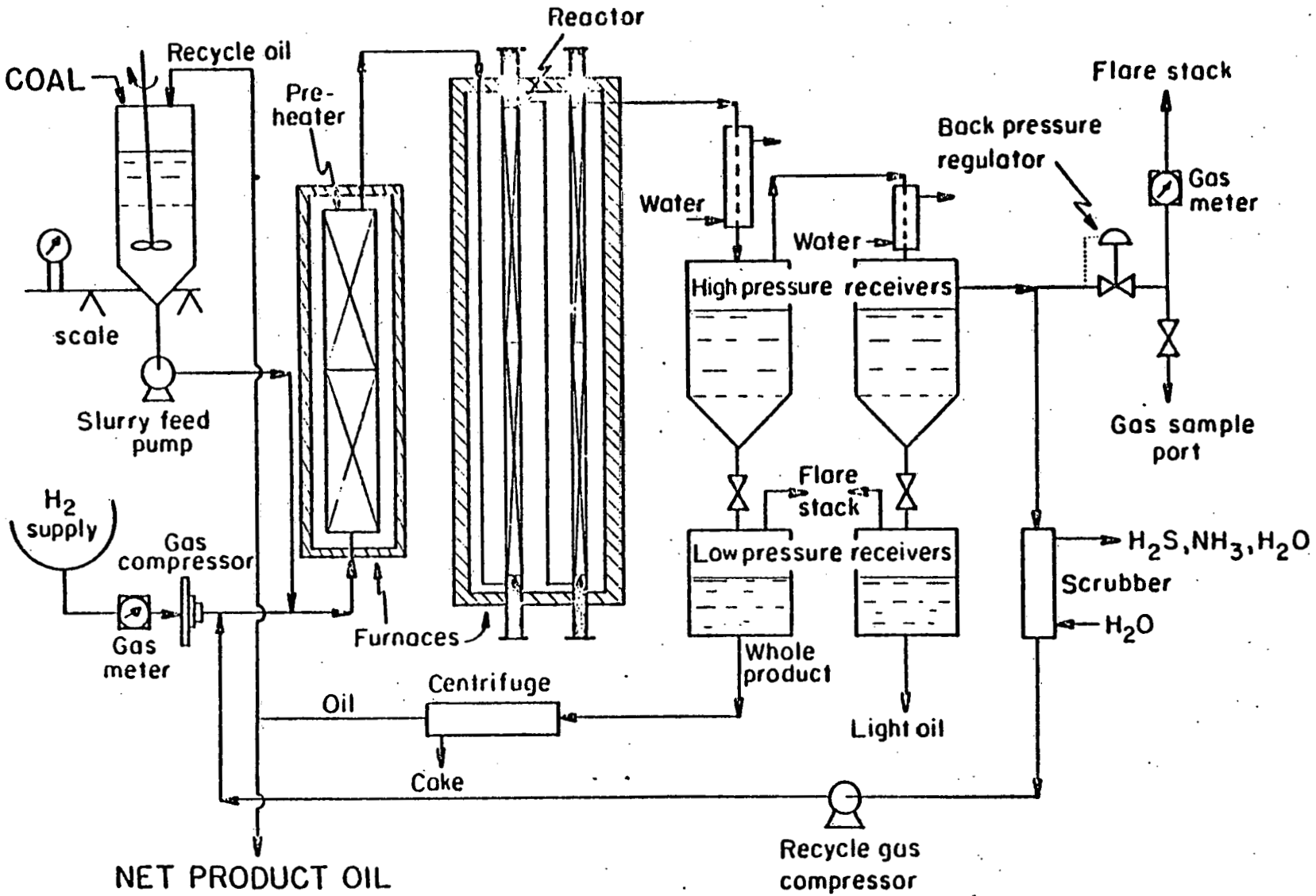
COAL LIQUEFACTION PROGRAM STRATEGY

I. IMMEDIATE GOAL: PRODUCE LOW-SULFUR, LOW-ASH FUEL OIL FOR ELECTRIC POWER GENERATION: RESERVE THE REPLACED EQUIVALENT PETROLEUM AND NATURAL GAS FOR MOTOR AND DOMESTIC USE.

II. SECONDARY GOAL: UPGRADE THE FUEL OIL TO MOTOR FUELS IN READINESS FOR FURTHER OIL DEPLETION.



SYNTHOIL PROCESS



SYNTHOIL PILOT PLANT FLOW SHEET

L-13694

PROCESSING OF A KENTUCKY COAL TO SYNTHOIL

RESULTS

SULFUR IN PRODUCT OIL, WT %	0.2
ASH IN PRODUCT OIL, WT %	0.1-0.2
VISCOSITY OF PRODUCT OIL, SSF @ 77° F	26-440
SPECIFIC GRAVITY OF PRODUCT OIL, 60° F/60° F	1.020-1.082
CALORIFIC VALUE OF PRODUCT OIL, BTU/LB	17,400
SOLIDS, WT PCT OF WHOLE PRODUCT	11.7
OIL YIELD FROM PYROLYSIS OF SOLIDS, WT %	28.4
SULFUR IN PYROLYSIS OIL, WT PCT	0.4
ASH IN PYROLYSIS OIL, WT PCT	0.000

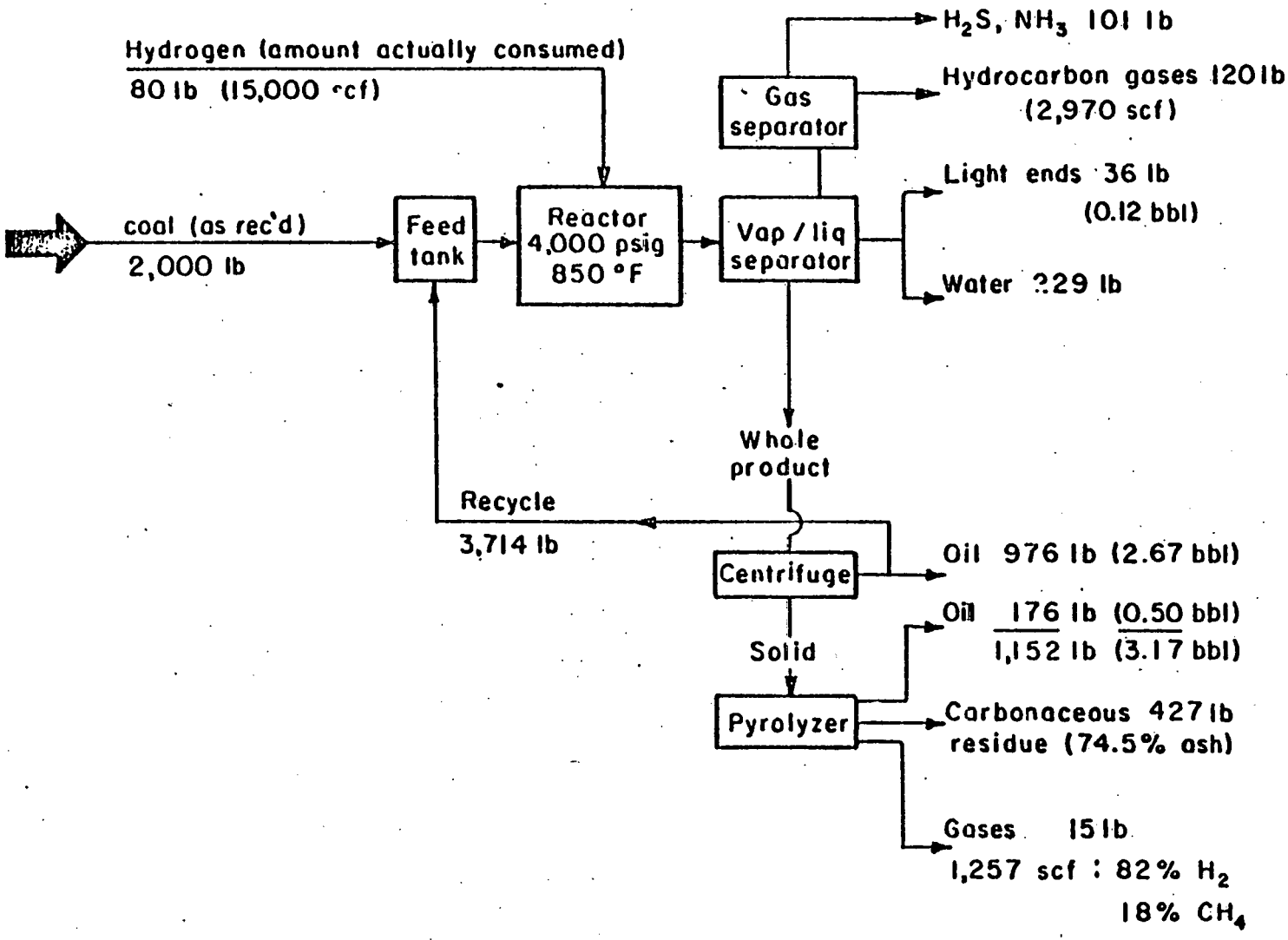


Figure —Material balance for Synthoil run with Western Kentucky coal.

L-14007

Thermal efficiency analysis of the SYNTHOIL Process

Input Energy	B.t.u./Ton Coal (as received)
Coal Input	22,400,000
Makeup Hydrogen	8,200,000
Net Process	600,000
Fuel Fired	400,000
Coal Pulverization	300,000
Total In	31,900,000
Output Energy	
Heavy Oil Product	19,700,000
Light Oil Product	1,200,000
Hydrocarbon Gases	3,000,000
Total Out	23,900,000

$$\text{Efficiency} = \frac{23,900,000}{31,900,000} \times 100 = 74.9\%$$

SYNTHOIL PROCESS ECONOMICS
FOR
98,000 B/CD PLANT

INVESTMENT COST: $575,000,000

OPERATING COSTS (NO PROFIT OR TAXES):

COAL COST, $/T	$/BBL FUEL OIL
8	4.75
10	5.56
12	6.37

SELLING PRICE:

DCF RATE, PER CENT	$/BBL WITH COAL AT $8	$10	$12
12	7.52	8.33	9.14
15	8.35	9.16	9.97
20	9.82	10.63	11.43

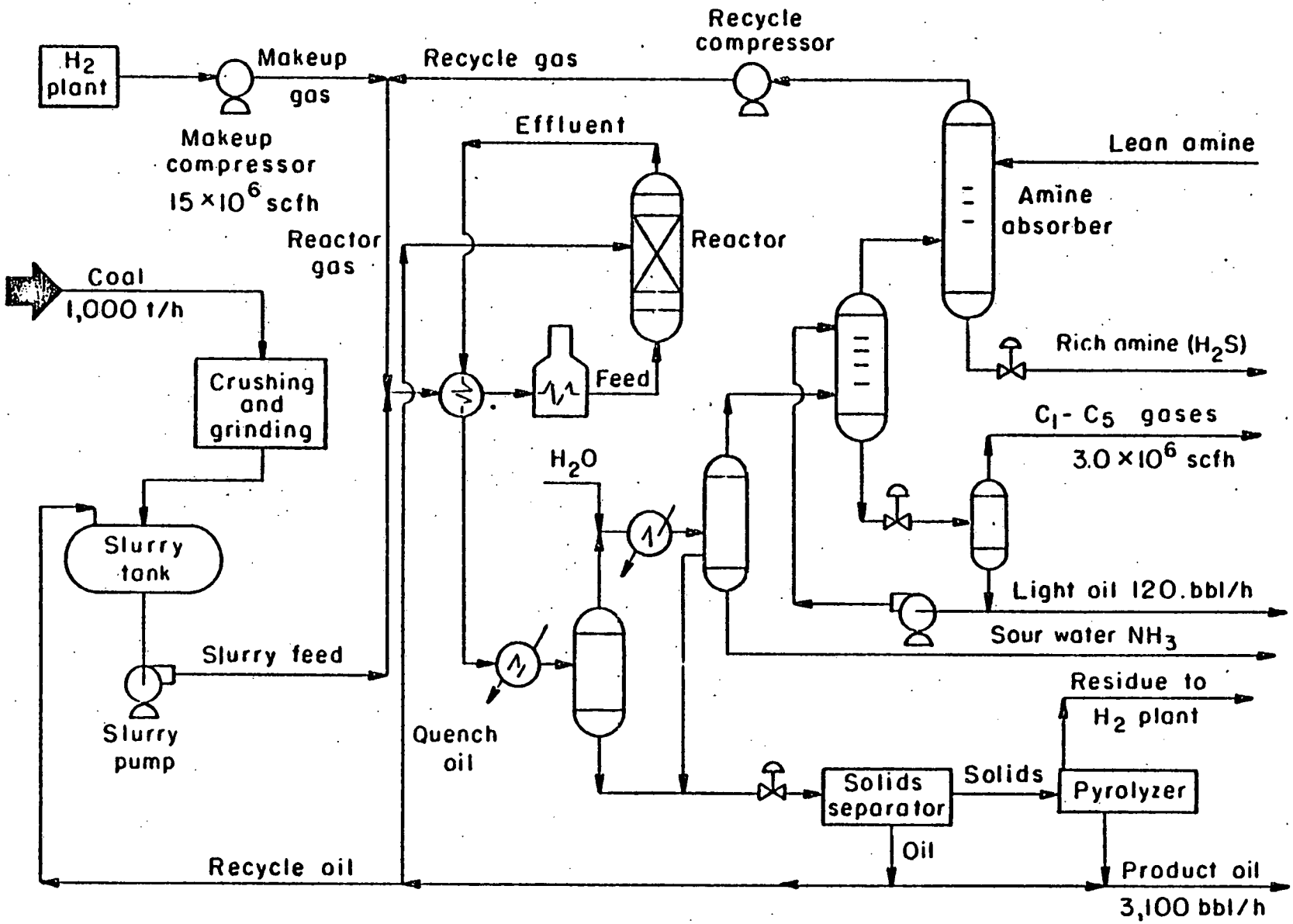


Figure Flow diagram for a future commercial SYNTHOIL plant.

L-14647

SYNTHOIL DEVELOPMENT SCHEDULE

PDU SIZE	REACTOR ID	COMMENTS
100 LB/DAY	5/16"	OPERATED FEBRUARY 1971-DECEMBER 1972
1/2 TON/DAY	1"	CONSTRUCTED NOVEMBER 1972-MARCH 1973 OPERATE APRIL 1973-JUNE 1977
10 TON/DAY	6" (AND 8")	CONSTRUCT OCTOBER 1975-JANUARY 1977 OPERATE JANUARY 1977-JULY 1979
700 TON/DAY	36"	PRELIMINARY DESIGN APRIL 1977-JULY 1978 DESIGN FOR CONSTRUCTION AUGUST 1978-AUGUST 1979 CONSTRUCT SEPTEMBER 1979-JANUARY 1981

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COMPARISON OF COAL CONVERSION PROCESSES

FOR ELECTRIC POWER GENERATION

By

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COMPARISON OF COAL CONVERSION PROCESSES FOR ELECTRIC POWER GENERATION

Introduction

The purpose of this paper is to examine the relative economic positions of some of the competing technologies for producing electric power from coal and coal-derived fuels. Among these concepts are:

- coal liquids from processes which could produce a range of heavy and light low sulfur, low ash boiler fuels, depending on the severity and nature of treatment;
- solvent refined coal (SRC) which would be a low sulfur, low ash solid which would meet existing limits for these pollutants;
- coal gasifiers with integrated advanced power systems including (1) combined cycles (i.e., gas turbines followed by steam turbines which utilize the waste heat of the flue gas), (2) fuel cells, and (3) magnetohydrodynamics (MHD). Additionally, fluid bed boilers could be integrated with gas turbines;
- conventional steam/electric boilers with flue gas desulfurization (FGD);
- gas turbines which, to minimize capital investment and to maximize siting flexibility, would not be integrated into a combined cycle;
- fuel cells which, for reasons similar to gas turbines, would stand alone with the necessary clean fuel processing equipment; and
- coal cleaning processes in which sulfur is removed by physical

beneficiation or leaching. The product fuel would have a higher ash content than SRC, but being preground, may be suitable for pulverized coal-fired boilers with no further treatment.

Economics are presented for new large-scale (1000 Mw) installations in the time frame when the technologies become fully commercially available, projected to be 1980-1985 for most of these systems. Since MHD is unlikely to reach commercial availability in this period, it is specifically excluded from the analysis. The coal cleaning processes have not been included because of their early stage of development which make cost predictions very difficult. It should be noted, however, that the process developers predict the costs of coal cleaning to be competitive with FGD costs.

Methodology

The economics of power generation were developed based on a model which assumed the use of an Eastern or Midwestern high sulfur coal (11,000 Btu/lb, 3% sulfur). For secondary fuels such as SRC and coal liquids, the conversion step was assumed to take place at the mine mouth since the product transportation charge would be less costly than that of the as-mined coal. In every case the fuel, solid or liquid, was transported an assumed 500 miles to the load centers for generation, rather than generating the power at the mine for subsequent transmission.

Energy Cost

The sensitivity of the results with respect to energy cost was tested at three mine mouth coal prices. They were $20, $25 and $30 per ton which bracket the range of coal prices available in 1976.

Transportation Costs

The fuel transportation costs, assuming all-rail movement, were determined as follows for the 500-mile distance:

	Unit Cost $/T-mile	Transportation Charge $/MMBtu-500 miles
Coal	.015	.34
SRC	.015	.24
Oil	.0072	.10
Methanol	.0059	.15

Any sensitivity of the resulting power costs to fuel transportation charge or distance is similar to the sensitivity with respect to mine mouth coal price described above, and therefore was not carried out as a separate analysis.

Structure of Coal Conversion/Power Plant Venture

Each power generation concept was taken as a separate venture beginning with the mine mouth coal purchase and including fuel conversion, if any, contract transportation, and generation. It was assumed that the generating company would wholly own the power plant and would have a pro rata interest in the operation of the fuel conversion facility, based on its average fuel take. Thus, in viewing the venture, the company would look at the combined profitability of its generating station and its share in the mine mouth converter, taking the same capital charge rate (or DCF rate) for both plants in determining the cost of the final product--electric power. Although many ventures would not be structured this way, this provides a consistent method for assessing the relative economic benefits of the options.

Costs for fuel, transportation, operation and capital for 1980-1985 technology were estimated on the basis of purchase in 1976 in current dollars. Therefore, all comparisons avoid the complication of predicting escalations for various cost components over the next ten years.

Economics

The estimated direct capital costs for the power generating options are presented in Table 1 along with the owner's costs and allowance for funds during construction (AFDC). Note that since the costs are all in 1976 dollars, there is no escalation included, a factor which would also impact on the AFDC figures.

Capital costs and thermal efficiencies for the coal conversion facilities are shown in Table 2. Add-on factors for owner's costs and AFDC are comparable to those for the generating plants.

Table 3 summarizes the overall energy efficiency of the generating schemes expressed as weight of mined coal per kilowatt-hour. Note that the inefficiency introduced by the coal conversion step has serious impact on the overall energetics and ultimately on the overall economics.

The delivered fuel costs at the generating site are developed in Table 4. As expected, the derived fuels are more economical to transport than the raw coal.

Findings

Based on the economics and efficiency data discussed above, plots of power cost versus load factor (average annual percent of capacity operated) were prepared for each overall conversion scheme. For example, Figure 1 is for a coal price of $25.00/ton at the mine and a 15% capital charge rate. Similar plots, Appendix Figures A-1 through A-4, show the

Table 1

CAPITAL INVESTMENT SUMMARY - GENERATING PLANTS

$/Kw (1976)

	Fixed Plant Investment[1]	Owner's Costs[2]	Allowance for Funds During Construction (AFDC)[3]	Total Investment (Unescalated)
Conventional Generation w/FGD	400	32	79	511
Advanced Power Cycle	400	32	79	511
Conventional w/SRC (solid)	300	24	59	383
Conventional w/Coal Liquids	255	20	50	325
Fuel Cell/Methanol	250	20	19	289
Fuel Cell/Coal Liquids	225	18	17	260
Gas Turbine/Methanol	180	14	13	207
Gas Turbine/Coal Liquids	180	14	13	207

[1] Costs for 1000 Mw conventional and advanced plants, and multiple 20-100 Mw units for fuel cells and gas turbines.

[2] Based on 8% of fixed plant investment.

[3] Based on 10% interest and five-year construction periods for the conventional and advanced power cases and two years for the others.

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Table 2

CAPITAL INVESTMENTS AND THERMAL EFFICIENCIES OF

COAL CONVERSION PLANTS

Capacity - 20,000 TPD Coal Input

(90% Load Factor)

	Fixed Plant Investment $MM	Thermal Efficiency* %
Solvent Refined Coal	370	70
Coal Liquids-Heavy	489	67
Coal Liquids-Light	489	65
Methanol	688	55

* Thermal Efficiency here is defined as the energy content of the solid or liquid products (as higher heating value) divided by the higher heating value of the coal input.

Table 3

MINED COAL REQUIREMENTS FOR POWER GENERATING SCHEMES

Description of Energy System	Location*	Coal Conversion Eff.%	Thermal/Elec. Heat Rate Btu/Kwh	Transported Fuel Form	Transported Fuel lbs/Kwh	Mined Coal lbs/kwh
Conventional Generation w/FGD	urban	-	9500	coal	0.86	0.86
Coal Liquids/Conventional	MM/urban	67	9000	oil	0.50	1.22
SRC (Solid)/Conventional	MM/urban	70	9000	SRC	0.56	1.17
Advanced w/Gasification	urban	**	8300	coal	0.75	0.75
Methanol/Fuel Cell	MM/dispersed	55	6500	methanol	0.67	1.07
Coal Liquids/Fuel Cell	MM/dispersed	65	7500	oil	0.68	1.05
Coal Liquids/Gas Turbine	MM/dispersed	65	11,300	oil	0.63	1.58
Methanol/Gas Turbine	MM/dispersed	55	11,300	methanol	1.16	1.87

* MM - mine mouth

** Coal to power heat rate assumes integrated gasifier.

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Table 4

DELIVERED FUEL COSTS FOR VARIOUS POWER GENERATING SCHEMES

	Mined Coal Cost		Product Fuel	Product Fuel Charge*	Transportation Cost	Total Fuel Charge	
	$/Ton	$/MMBtu		$/MMBtu	$/MMBtu	$/MMBtu	$/MMBtu Coal Mined
Conventional	20.00	.91	coal	.91	.34	1.25	1.25
	25.00	1.14		1.14		1.48	1.48
	30.00	1.36		1.36		1.70	1.70
Advanced	20.00	.91	coal	.91	.34	1.25	1.25
	25.00	1.14		1.14		1.48	1.48
	30.00	1.36		1.36		1.70	1.70
SRC	20.00	.91	SRC	1.30	.24	1.54	1.08
	25.00	1.14		1.63		1.87	1.31
	30.00	1.36		1.94		2.18	1.53
Coal Liquids-Heavy	20.00	.91	oil	1.36	.10	1.46	.98
	25.00	1.14		1.70		1.80	1.21
	30.00	1.36		2.03		2.13	1.43
Coal Liquids-Distillates	20.00	.91	oil	1.40	.10	1.50	.98
	25.00	1.14		1.75		1.85	1.21
	30.00	1.36		2.09		2.19	1.43
Methanol	20.00	.91	methanol	1.65	.15	1.80	.99
	25.00	1.14		2.07		2.22	1.22
	30.00	1.36		2.47		2.62	1.44

* Product fuel charge equals coal cost divided by conversion efficiency.

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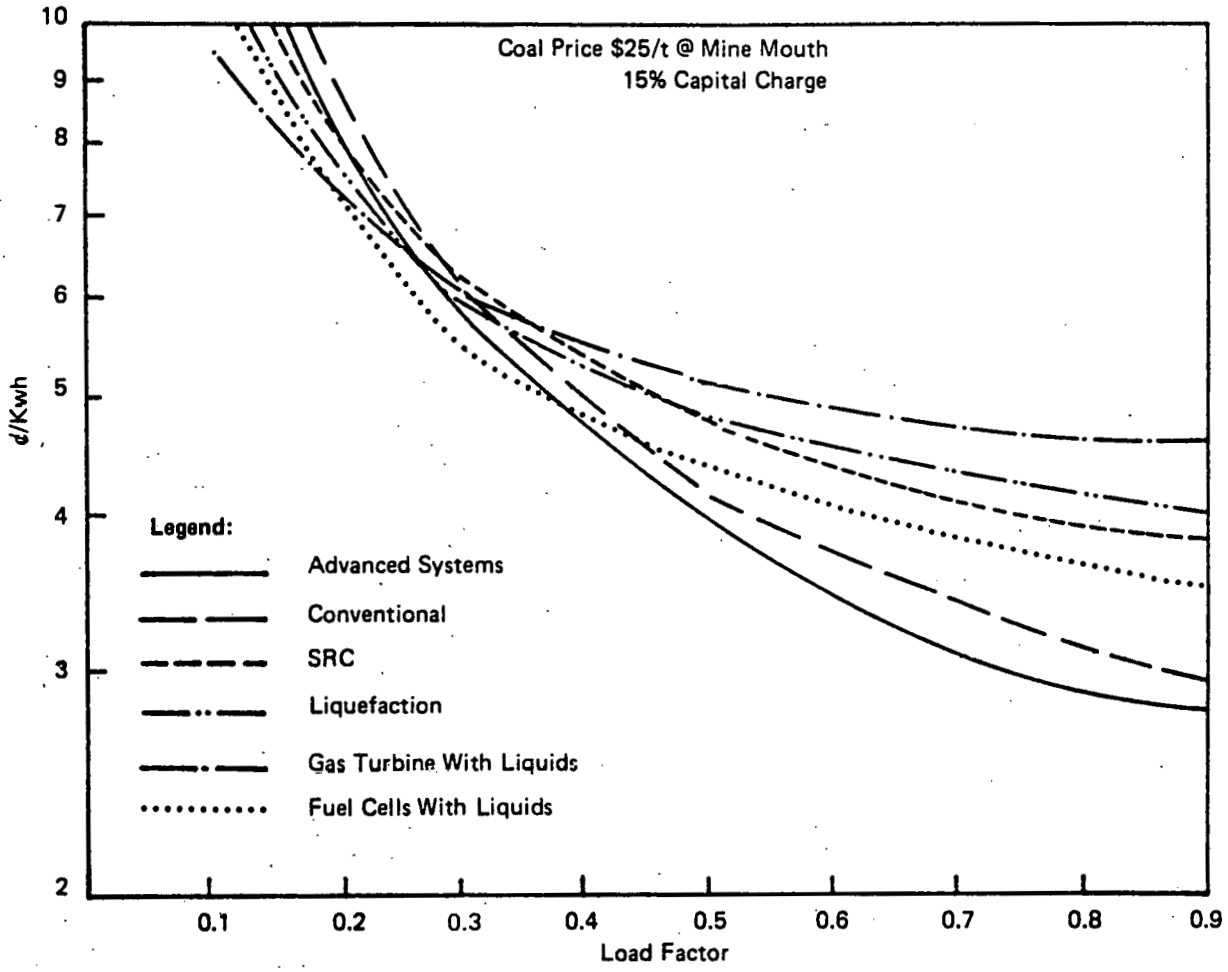


FIGURE 1 POWER GENERATION COSTS VS. LOAD FACTOR

sensitivity of the result with respect to the capital charge rate and the coal price. The same ranking prevails in each range of load factor. For simplicity, the methanol cases for fuel cells and gas turbines have not been shown because they typically lag the comparable coal liquids case by about 1¢/kwh and 2¢/kwh, respectively.

Table 5 summarizes the relative capital intensities of the overall conversion options per unit of base load capacity (90% load factor). Also summarized is the rate of coal consumption (as mined) per kwh of output. It is interesting to note the direct correlation between energy consumption and capital investment. The intermediate step to produce clean fuels impacts importantly on both the total required investment and the total required energy, much of which is lost in this step.

Although conclusions drawn from a general study of this type will vary in certain instances--each utility has its own unique fuel supply and power generating situation--the following discussions will have widespread applicability for utilities who are planning their generating mix for the 1980's around coal.

1. Advanced power systems now under development appear to have only slight potential advantage over conventional steam/electric generation with FGD. Despite the apparent risks of flue gas desulfurization, principally system reliability, it would appear that the risks of advanced systems, which revolve mainly around coal gasification, high temperature sulfur removal, and gas turbine life and reliability for base load operations at high temperatures (and/or fuel cell life), are at least equivalent or more serious. Beyond 1985 it is probable that further improvements in advanced system designs will improve their overall reliabilities and, to

Table 5

SUMMARY CAPITAL INVESTMENTS AND COAL

CONSUMPTIONS OF COAL-TO-POWER OPTIONS

	Fixed Plant Investment* $/Kw	Coal Mined lbs/kwh
Conventional Generation with FGD	400	0.86
Advanced Power Cycle	400	0.75
Conventional with SRC (solid)	560	1.17
Conventional with Coal Liquids	613	1.22
Fuel Cell/Coal Liquids	568	1.05
Gas Turbine/Coal Liquids	670	1.58

*Includes both coal-to-fuel conversion plant, if any, and power plant, at a load factor of 90% for each plant.

some extent, their energy efficiencies.

2. Dispersed fuel cells appear to have the most promise in the intermediate load range. Despite their relatively attractive heat rates and capital costs, dispersed fuel cells (i.e., those not integrated with coal gasifiers) suffer in the base load range from the fact that the coal-derived fuels are expensive to produce.

Although this study is concerned with new installations, it is worth noting that in the intermediate load range, fuel cells will be competing not only with new systems using other technologies but also with older conventional boilers, which have previously been used in base load service, written off for accounting purposes, and relegated to intermediate or peak service.

3. Gas turbines based on coal-derived distillates are attractive in the peak load range.

4. Solvent refined coal for feed to a new conventional power plant is less economical than the above discussed options. However, for feed to an existing power plant, the SRC route may have some advantage over a flue gas desulfurization retrofit in the peak-to-intermediate range.

5. Heavy coal liquids have a similar position to SRC. More expensive to produce, the heavy liquids are more attractive than SRC only in the peaking range where the lower power plant investment prevails. Further, liquids may be more appropriate for existing installations where conversion to burning a solid fuel may be infeasible.

6. Methanol from coal has little attractiveness for large scale utility use when compared with other clean fuels from coal and direct fired coal with flue gas desulfurization.

A P P E N D I X

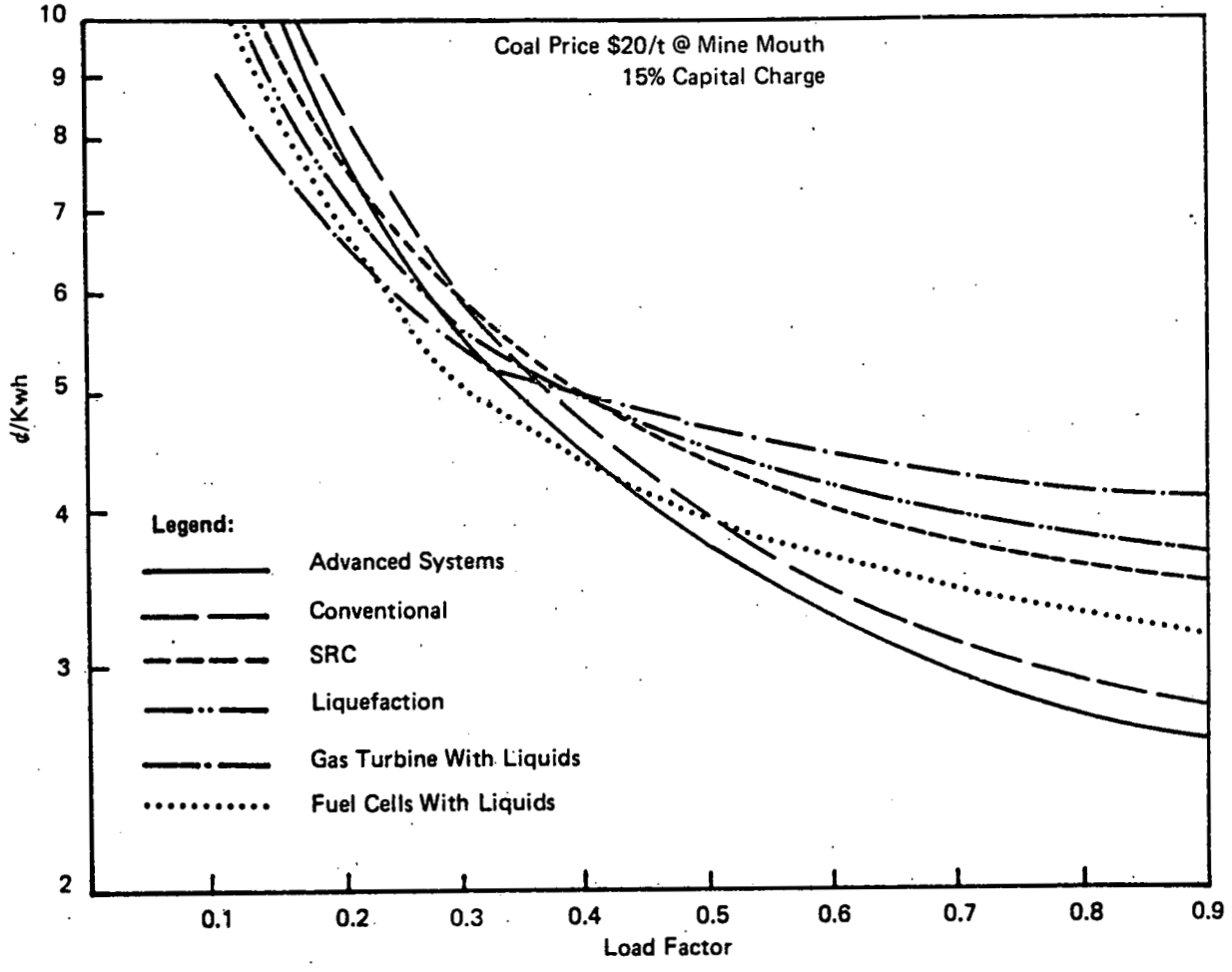


FIGURE A–1 POWER GENERATION COSTS VS. LOAD FACTOR

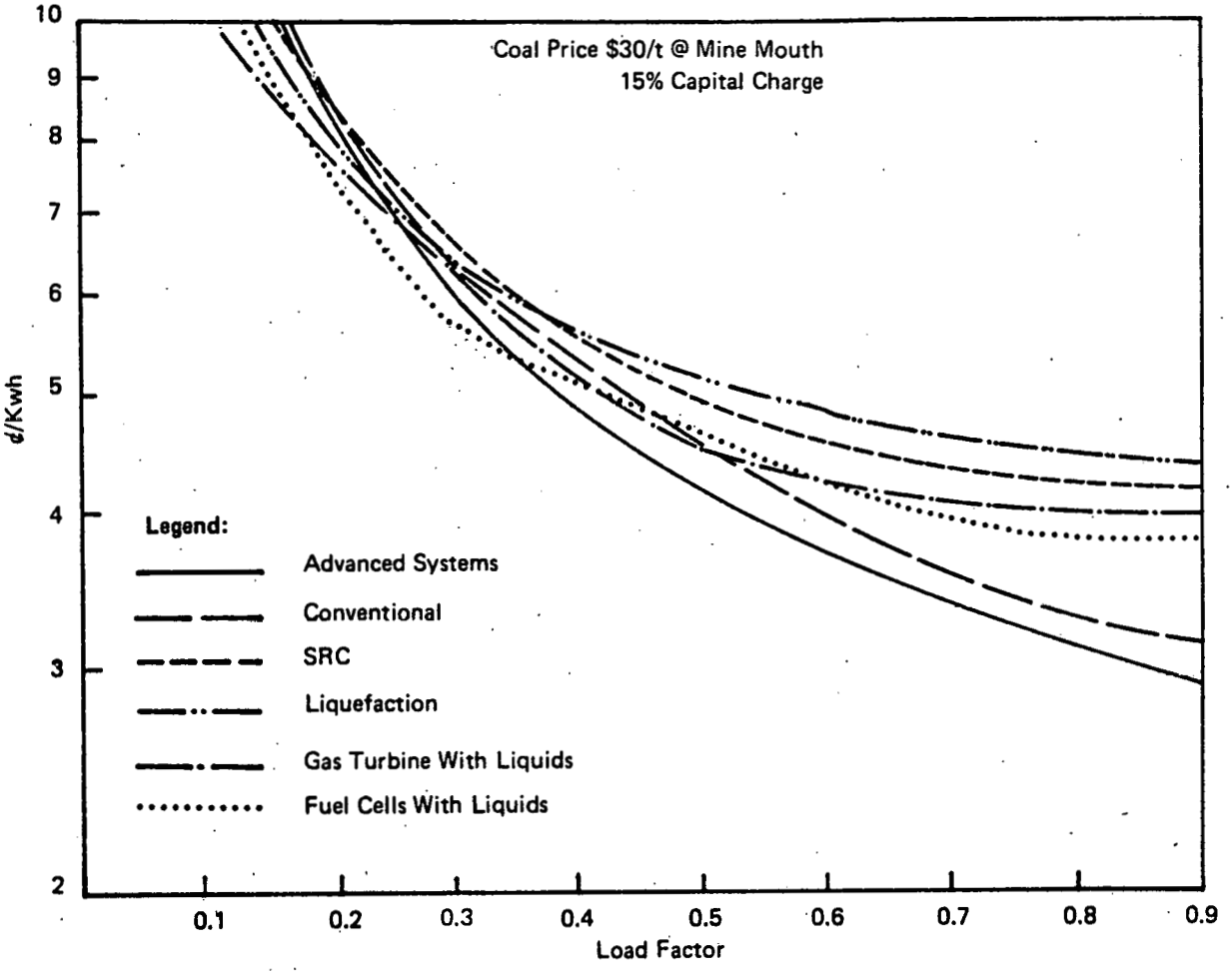


FIGURE A–2 POWER GENERATION COSTS VS. LOAD FACTOR

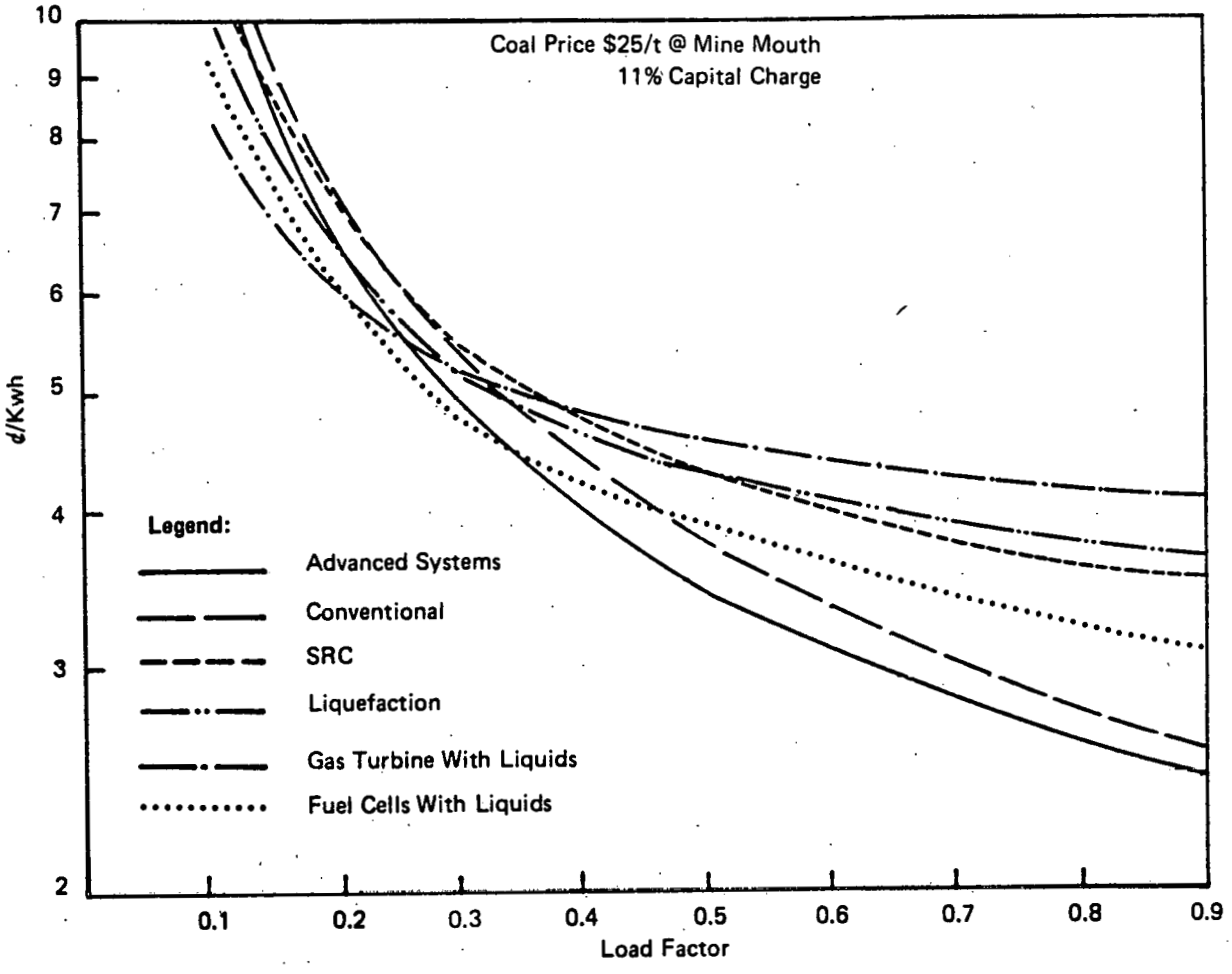


FIGURE A–3 POWER GENERATION COSTS VS. LOAD FACTOR

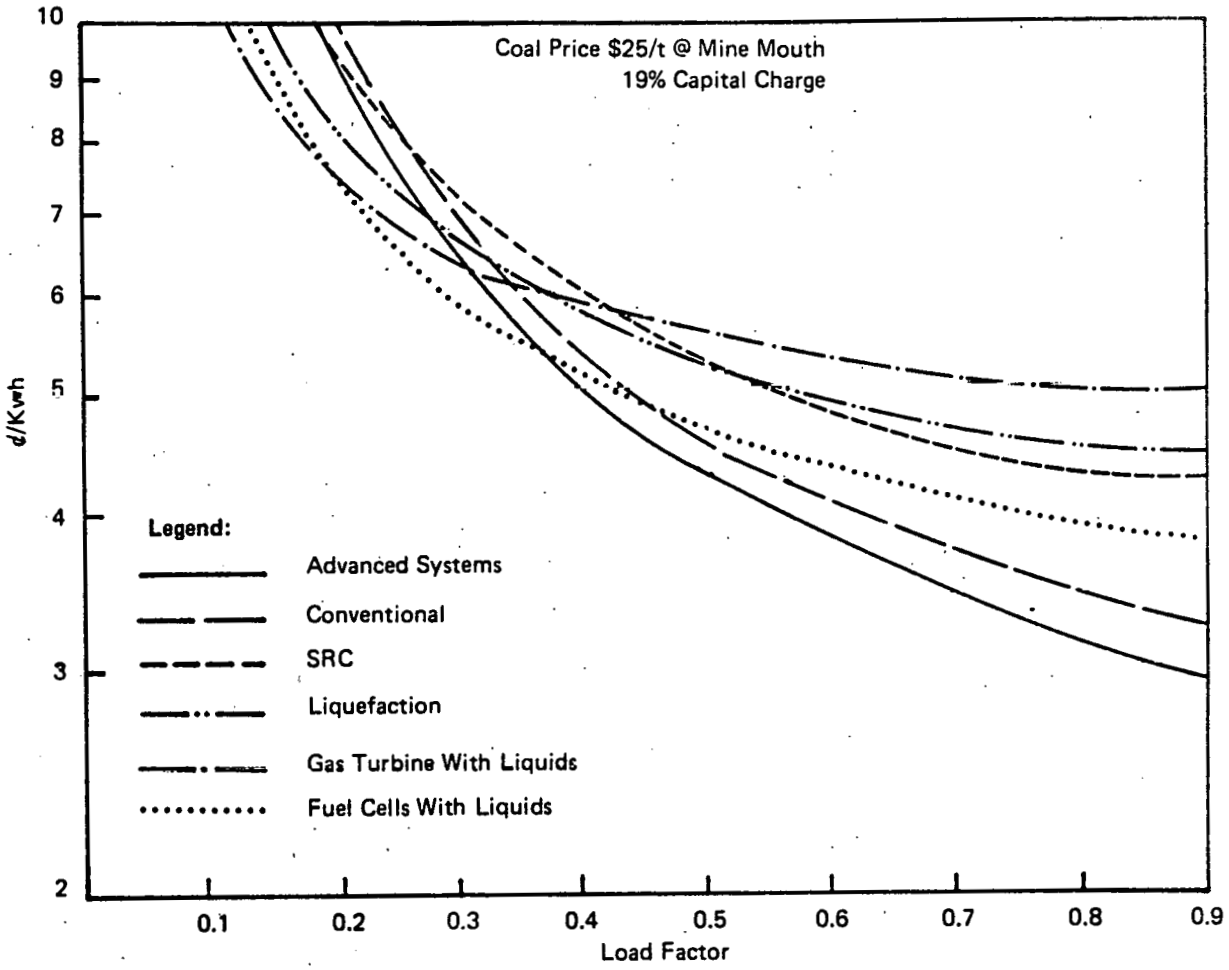


FIGURE A–4 POWER GENERATION COSTS VS. LOAD FACTOR

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FLUIDIZED BED COMBUSTION

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ABSTRACT

Fluidized bed combustion processes can utilize a wide range of fossil fuels for electric power generation, steam production or heat utilization to achieve lower costs compared with competitive systems and to meet environmental requirements. No technical problems have been identified which preclude commercialization. The primary need is for integrated programs to implement the development and commercialize fluidized bed combustion processes.

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INTRODUCTION

National energy programs and energy projections call for a greater use of coal.[1,2] However, many uncertainties are associated with achieving this increased use of coal. Coal extraction and supply, for example, are critical areas of concern. The increased use of coal is also limited by the high cost and uncertainties associated with the utilization; the environmental impact from direct utilization being a primary concern. Currently about 85% of the coal utilization is by direct combustion for central station electric generation (65%) and industrial electric generation or heat (20%). The increased demand in these areas and the need for substitution of coal for oil and gas make it important to investigate improved technology for direct utilization. Fluidized bed combustion is one process concept which offers the opportunity to realize improvements in cost, fuel resource utilization, and environmental impact.

Fluidized bed combustion processes considered in this document will be limited to concepts which recover heat for industrial or commercial use, steam production, or electric power generation. Fluidized bed combustion processes are commercially used in over 300 installations in the U.S. for roasting pyrite ores, incineration of industrial wastes, incineration of sludge from sewage plants, and related applications. A high degree of reliability has been demonstrated with these systems. Many of these systems use water cooled tubes to control bed temperature. This experience is valuable for the development of coal fired fluidized bed combustion systems.

Historical perspective on the development of fluidized bed combustion is available from a number of documents.[3-6] In addition, several conferences have been held which provide a broad scope of the work which has been performed and the work being carried out.[7-11] The

primary emphasis here will be a review of process concepts and an overview of fluidized bed combustion development.

PROCESS CONCEPTS

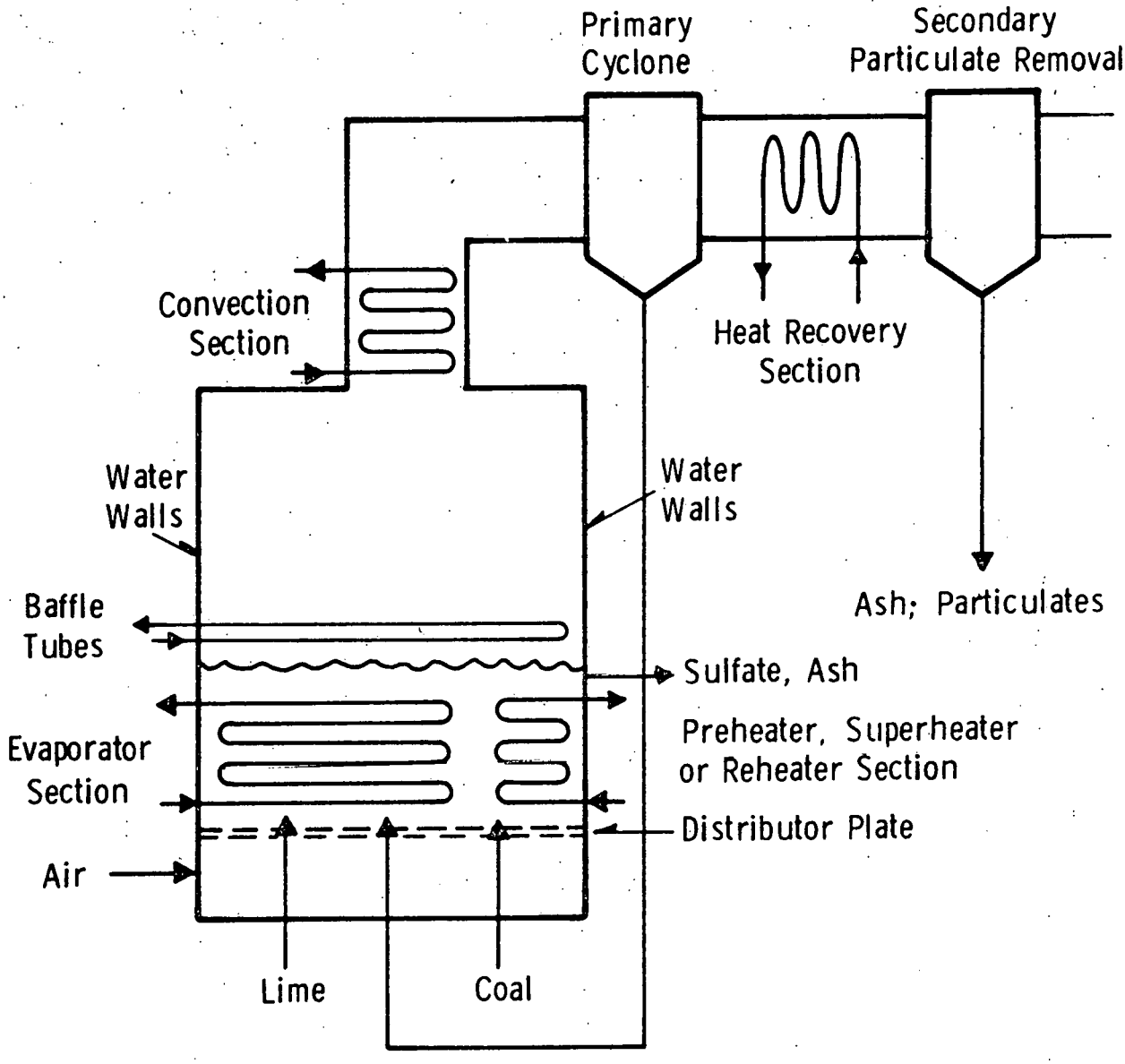
Fluidized bed combustion units which recover heat for industrial use or electrical energy generation are now being designed for commercial operation. The basic concept is to provide sufficient air to the fuel in a fluidized bed to complete the combustion of carbon to CO_2 and hydrogen to H_2O. Both reactions release large quantities of heat which can be absorbed by generating steam or transferred to a gas.

Sulfur in the fuel is converted primarily to SO_2. A limestone/dolomite sorbent can be utilized in fluidized bed combustion to remove this pollutant from the combustion gases:

$$\begin{pmatrix} CaCO_3 \\ CaO \end{pmatrix} + SO_2 + 1/2\ O_2 \rightarrow CaSO_4 + \begin{pmatrix} CO_2 \\ - \end{pmatrix}$$

The $CaSO_4$ can either be disposed of as a solid waste or it can be processed to regenerate $CaCO_3$ (or CaO) sorbent and to recover sulfur as a useful product -- elemental sulfur, S, or sulfuric acid, H_2SO_4.

A schematic diagram of a simple coal-burning fluidized bed combustion boiler concept is shown in Figure 1.[3] It is formed by an enclosure usually consisting of waterwalls -- abutting boiler tubes in which water and steam flow. The pressure in this enclosure is in the 1 to 25 atm range. A flat distributor plate with holes or bubble cap devices distributes the air flow uniformly over the base of the enclosure. The air then passes at velocities in the range of 2 to 15 ft/sec through a bed of particles at temperatures from 1400 to 2000°F. These particles, whose top size may be 1/16 to 1/4 in., comprise ash or other inert material, lime sorbent, and small quantities (less than 1%) of unburned coal or



Pressure: 1 - 25 atm
Coal Size: pf - 1/4 in.
Air Flow: 2 - 15 ft/sec
Temperature: 1400 - 1900°F
Surface: Water Walls, Horizontal, and Vertical Tubes in Bed

Sulfur removal: $CaO + SO_2 + \frac{1}{2} O_2 \rightarrow CaSO_4$

-Fluidized bed combustion boiler

carbon. Coal and sorbent are fed to the bed through the air distributor or the waterwalls. From 60 to 70% of the heat released in burning the fuel with air is transferred to the water/steam in the tubes surrounding and submerged in the bed. Various tube configurations have been proposed -- horizontal tubes extending through the bed, horizontal tubes with serpentine bends in vertical platens or planes, and vertical tube walls passing up through the distributor plate, slicing the bed into narrow segments. Convection heat transfer surface can be located above the bed to recover heat from the sensible heat of the gases.

The sulfur removal system must be compatible with the total process. In order to achieve this, factors such as sulfur removal, attrition, trace element release, and spent sorbent disposition must be assessed. First generation systems are projected to use once-through operation.[12,20] Molar ratios of limestone calcium feed to sulfur in the coal will vary from 1.2 to 3.0 at the design bed temperature for > 80% sulfur removal. The technical or economic feasibility of sorbent regeneration with low sorbent makeup rates (Ca/S << 1) has yet to be demonstrated for atmospheric or elevated pressure systems. A regenerative system with low makeup sorbent rates is attractive since it would minimize the spent sorbent disposition problem. Thus, development work on regenerative systems should continue.

This basic fluidized bed combustion concept can be utilized in a number of different power cycles and thermal applications. The particular use requires selection of a number of critical process operating, system configuration, and hardware design parameters. The major process distinctions between alternatives being considered are the selection of the working fluid and the utilization of the working fluid. For example, heat may be transferred to water/steam cooled tubes for use in a steam turbine or as industrial heat, heat may be transferred to air cooled tubes for use in a closed cycle gas turbine, or heat may be transferred to the combustion gas for expansion through a turbine expander. Design and operating conditions are selected to meet the particular process objectives. Other design concept variations are being considered, such as the use of

"fast fluidization" or recirculating beds.[12,13] These will not be discussed.

FLUIDIZED BED COMBUSTION PROCESS FEATURES

The primary potential advantages of fluidized bed combustion systems for electric power generation, steam production, or heat utilization are:

- Reduced volume and modular construction. Because combustion rates are more intense in fluidized beds than in the fire box of a pulverized fuel furnace, and because heat transfer surface can be placed in the bed, fluidized bed boilers are more compact than conventional coal-fired boilers. Boiler modules or sections can be fabricated in shops and assembled at a plant site.
- Reduced heat transfer surface requirements. Because heat transfer coefficients are an order of magnitude greater in fluidized beds ($\sim$ 50 Btu/hr ft^2°F) than in the fire box of a conventional boiler, less heat transfer surface is required in the heat exchanger. In a pressurized combustor or heat exchanger less heat is extracted from the combustion gases after they leave the fluidized bed because they are cooled by expansion in a gas turbine. Further large reductions in heat transfer surface are thus possible.
- Reduced fuel costs and increased flexibility. Because fluidized beds can readily burn crushed coal (fine grinding is not required), char, coal with a high ash content, and a variety of miscellaneous combustibles (from sewage sludge, municipal solid wastes, paper mill liquid wastes, and oil wastes), fluidized bed combustors can utilize cheap fuels and a wide variety of fuels to generate power, steam, or heat for utilization. Combustion efficiencies with coal from 95 to 99% can be achieved. High efficiencies are achieved by either recycling carbon carryover back to the combustor or by utilizing a second stage carbon burnup cell.

- Reduced emissions of SO_2 and NO_x. Because a limestone/dolomite sorbent can be utilized in a fluidized bed combustor, SO_2 reductions of 90 to 95% can be economically achieved. The low combustion temperature at which the bed operates minimizes the formation of NO_x by fixation of atmospheric nitrogen. Pressurized fluidized bed combustion systems, in general, result in lower nitrogen oxide formation than atmospheric pressure fluidized bed combustion.
- Reduced environmental impact and maintenance. Because fluidized bed combustors operate at a maximum combustion temperature far below that in a conventional boiler, volatilization of minor elements and corrosive contaminants such as alkali metal compounds and fusion of ash are projected to be reduced. If these reductions are realized, the environmental impact and corrosion of heat transfer surface will be minimized.

These features of fluidized bed combustion systems result in the wide range of concepts and applications being developed throughout the world.

ATMOSPHERIC PRESSURE SYSTEMS

The atmospheric pressure concept receiving primary attention is the boiler concept for industrial steam or power generation. The prototype fluidized bed combustion boilers which are operational and are being constructed utilize the basic concept illustrated in Figure 1. Representative operating conditions projected for these plants are summarized in Table 1. A 30 MW_e (135,000 kg steam/hr) multicell fluidized bed steam generator at the Monongahela Power Company's Rivesville Plant in West Virginia is scheduled to begin operation this year.[14] Babcock and Wilcox, Ltd., with Combustion Systems Ltd., have converted a 20,000 kg steam/hr cross-type, water tube boiler to a fluidized bed combustion system and operated the facility since mid-1975.[15] The unit is located

TABLE 1

Range of Operating Conditions for Atmospheric Pressure
Fluidized Bed Boiler Prototype Plants

Fluidizing Velocity, ft/sec	2.5 - 13 (0.8 - 4 m/sec)
Bed Temperature, °F	1550 - 1650 (840-900°C)
Excess Air, %	10 - 20
Bed Depth, ft.	2 - 4 (0.6 - 1.2 m)
Freeboard, ft.	4 - 10 (1.2 - 3 m)
Bed Material	limestone/coal ash
Particle Size, in. μm	-1/16 to -1/8 (-1500μ to -3200μ)

at Babcock and Wilcox's works at Renfrew, Scotland. Bharat Heavy Electricals Ltd. in India have completed the construction of a 9,000 kg steam/hr fluidized bed boiler which is scheduled for operation this year.[16] Smaller scale atmospheric pressure fluidized bed combustion test facilities have been successfully operated beginning in the early 1960's which have provided the basis for the current plants. These units include the fluidized bed combustion 5000 kg steam/hr shell boiler at BCURA, Ltd., England which was successfully operated for about 5 years and the 2200 kg steam/hr fluidized bed boiler built in 1967 by Pope, Evans and Robbins, Inc.

The most attractive applications for fluidized bed combustion boilers are utility power plants and relatively large (e.g. > 150,000 lb steam/hr) new boilers for industrial steam generation. Small industrial boilers or conversion of packaged gas or oil fired boilers to fluidized bed coal combustion are not projected to be attractive markets. In general, low sulfur fuel used in conventional boilers is projected to be preferred over the use of high sulfur coal for industrial applications.

There are other atmospheric pressure fluidized bed combustion concepts which are attractive. Two of these are the indirect air cooled concepts such as that being developed by Oak Ridge National Laboratory and the large coal/shallow bed concepts being developed by the Coal Research Establishment, National Coal Board, England.

The Oak Ridge National Laboratory have been evaluating coal-fueled modular integrated utility systems (MIUS) for providing all or several of the utility services for a community.[17] An indirect air cooled fluidized bed coal combustion system coupled to a closed cycle gas turbine is recommended as the most promising system. The system would convert about 30% of the energy to electricity and about 50% into heat for domestic hot water, space heating, and air conditioning, The design power output ranges from 0.5 to 1.0 MW_e per unit. The reference fluidized bed combustor design operates with 10% excess air, 900°C bed temperature, 0.8 m/sec gas velocity, 0.6 m bed depth, 400 μm mean particle size, and 4 m^2 bed area.

The National Coal Board have successfully carried out tests in shallow fluidized beds, 0.15 m deep, with coal particles as large as 50 mm.[18,19] A 600 kw hot water boiler and a 1.2 MW shell boiler have been operated utilizing this concept. The concept is also being used for the production of a hot gas for drying applications. A 5 MW fluidized bed furnace for drying grass is being commercially operated. Other units are planned.

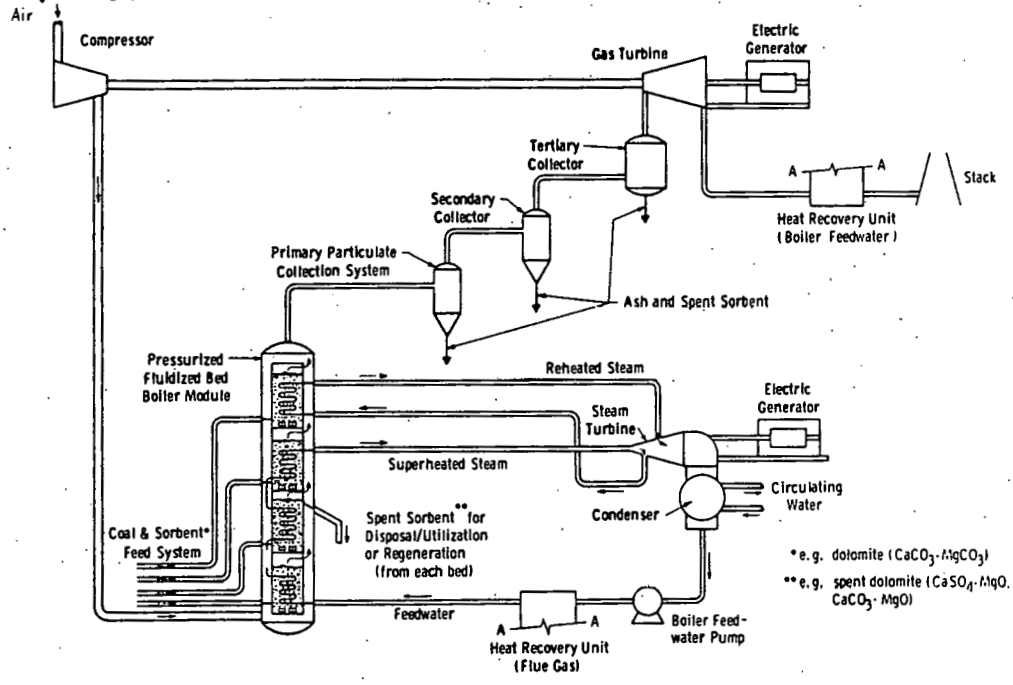
PRESSURIZED SYSTEMS

Pressurized fluidized bed combustion system concepts are essentially all being developed for electric power generation. Two basic pressurized fluidized bed combustion power plant systems, shown in Figure 2, have received primary attention. The general range of operating conditions is shown in Table 2.

Pressurized fluidized bed boiler power plant concepts have been reported by Westinghouse,[3,20,21] General Electric,[22] and the British Coal Utilization Research Association (BCURA) and Combustion Systems, Ltd.[23,24] The effect of boiler operating conditions and design parameters on economics and performance has been parametrically evaluated; the optimum power plant cycle conditions such as gas turbine pressure ratio, compressor intercooling, and steam conditions have been identified; and integrated plant operating procedures developed.[3,12,21,22] A 0.56 m^2 pressurized fluidized bed combustor pilot plant has been successfully operated at BCURA in Leatherhead, England since 1969 to obtain process data.[23,25] The Exxon pressurized fluidized bed unit called the miniplant has been operating since 1975.[9]

The requirement for internal heat transfer surface in fluidized bed combustion can be eliminated by increasing the excess air to a bed operating at 927°C (1700°F) to approximately 300 percent. An adiabatic coal-fired fluidized bed combustor is applicable to gas-turbine or combined gas-turbine/steam-turbine cycles. Combustion Power Company has operated a pilot plant adiabatic fluidized bed combustor-gas turbine on

Pressurized Fluidized Bed Boiler-Power Plant



Pressurized Fluidized Bed Adiabatic Combustor-Combined-Cycle Power Plant

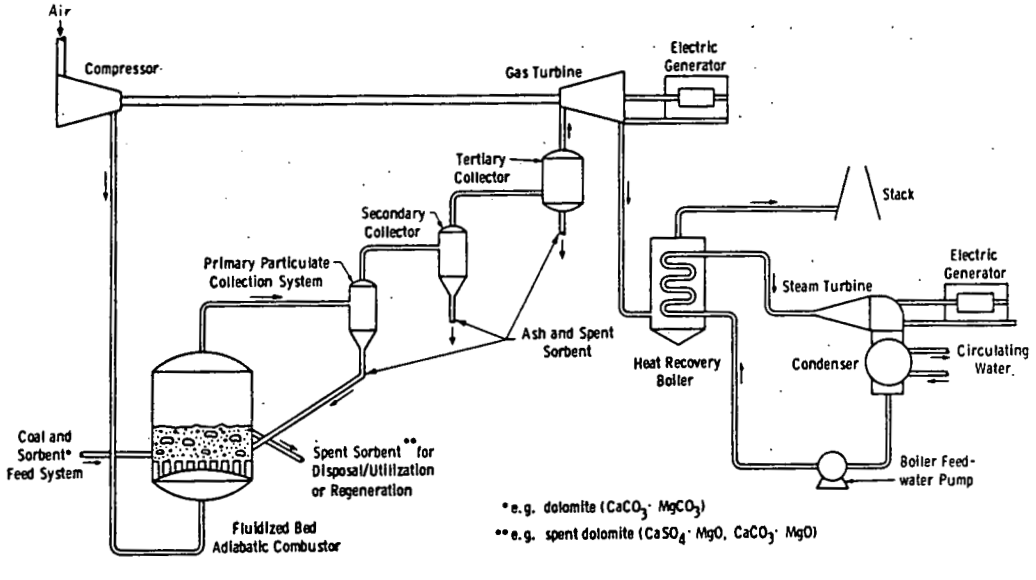


FIG. 2 - Pressurized Fluidized Bed Combustion Concepts

TABLE 2

Range of Operating Conditions for Pressurized Fluidized Bed Combustion Systems

Pressure, atm	7-13	(700 - 1300 kPa)
Fluidizing Velocity, ft/sec	2.5-10	(0.8 - 3 m/sec)
Bed Temperature, °F	1500-1850	(810 - 1010°C)
Excess Air, %	10-300	
Bed Depth, ft.	4.5-15	(1.5 - 5 m)
Freeboard, ft	5-35	(1.5 - 11 m)
Bed Material	limestone; dolomite; ash	
Particle Size, in.	-1/8 to -1/2	(-3200μ to -12,700μ)

prepared solid waste and coal.[26] The pilot plant combustor is 2.7 m (9 ft) in diameter operating at 414 kPa (61 psi) with gas velocities up to 2.1 m (7 ft)/sec and 0.6 m (2 ft) bed depths. The product gas goes to a Rustin TA 1500 gas turbine. Design, performance and economic studies have been carried out to assess this concept.[12,24,26]

One variation of the adiabatic combustor concept is the utilization of the excess air required to maintain the bed temperature in air-cooled heat transfer surface in the bed. This open-cycle gas turbine with a pressurized fluidized bed combustor/air heater is being developed by Stal-Laval.[9,27] Curtiss Wright have recently initiated work under contract to ERDA to develop this pressurized fluidized bed combustion concept. This concept permits part of the air (about 70%) from the compressor to by-pass the combustor gas flow, pass through the heat transfer surface immersed in the fluidized bed, where it approaches the bed temperature, and be mixed with the products of combustion after they have been cleaned of particulates. This alternative is less demanding of the particulate control system than the pressurized boiler and adiabatic combustor base designs. The cycle performance will be slightly lower than the adiabatic combustor system for a given bed temperature, since the excess air will only approach the bed temperature and result in a slightly lower gas-turbine inlet temperature.

Another system which has received considerable attention is the use of liquid metal topping cycles with fluidized bed combustion. Investigations of liquid-metal topping cycles coupled with pressurized fluidized bed fired heaters have been made by Oak Ridge National Laboratory,[28,29] General Electric,[22,30] and by Westinghouse.[21]

SYSTEM PERFORMANCE

The national average heat rates for fossil fuel steam electric plants have leveled off at 11.1 MJ/kWh (10,500 Btu/kWh). The most efficient plants average around 9.4 MJ/kWh (8900 Btu/kWh) and do not include sulfur removal or nitrogen oxide minimization. Conventional plants which meet environmental standards have heat rates near 9.6 MJ/kWh (9100 Btu/kWh). Coal-fired units now being built or on order are all 12.411 MPa/538°C/538°C (1800 psi/1000°F/1000°F), 16.547 MPa/538°C/538°C (2400 psi/1000°F/1000°F), and 24.132 MPa/538°C/538°C (3500 psi/1000°F/1000°F).

Atmopsheric pressure fluidized bed combustion boiler plants are projected to achieve a heat rate of 9440 Btu/kWh. This corresponds to a 16.5 MPa/538°C/538°C steam cycle with a 1-1/2 in. Hg back pressure and a boiler efficiency of 86%.

The first-generation pressurized fluidized bed boiler or adiabatic combustor combined-cycle power plants utilizing state-of-the-art power generation equipment are projected to match the most efficient current plants. Plant heat rates of 9.04 MJ/kWh (8570 Btu/kWh) are projected for the pressurized fluidized bed boiler plant operating at 17.5 percent excess air. The adiabatic combustor base plant heat rate is 9.6 MJ/kWh (9100 Btu/kWh). The steam cycle is the same as that assumed for the atmospheric pressure system. Closed-cycle gas-turbine, air-cooled heat transfer surface and related applications offer potential advantages by avoiding or minimizing possible problem areas in the reference concepts and should thus be investigated.

ENVIRONMENTAL IMPACT

The Environmental Protection Agency (EPA) is carrying out a program aimed at the complete environmental characterization of fluidized bed coal combustion processes.[31] The work includes environmental assessment and control technology development. The EPA program is part of a continuing effort which was initiated in 1967.

Available data demonstrate the ability of fluidized bed combustion systems to meet proposed sulfur dioxide, nitrogen oxide, and particulate emission standards. Minor element and trace element emissions have not yet been completely characterized and a basis for assessing the environmental impact is premature. Initial results are encouraging. Thermal emissions are reduced 15 to 25% in pressurized systems. Solids from the plant operated with once-through sorbent include coal ash and dry, granular calcium sulfate/calcium carbonate -- both suitable for disposal without further processing.[32] Spent stone may be regenerable, in which case the quantity of solids would be reduced. Thus, regenerative systems should continue to be studied and developed.

ECONOMIC ASSESSMENT

Cost estimates for new processes are always subject to a high degree of uncertainty. The range of economic parameters (e.g. interest during construction, escalation, regional differences, fuel cost, etc.) and technical assumptions (e.g. waste disposal requirements, reliability, equipment performance, construction time, etc.) makes it difficult to compare the cost estimates which have been projected for fluidized bed combustion systems. The results of the Phase I Energy Conversion Alternatives Study by NASA/ERDA/NSF illustrate this difficulty.[33]

Cost estimates have been prepared based on the assumption that commercial plants will achieve performance based on projections from pilot scale and laboratory test results. The capital costs for the atmospheric pressure boiler, pressurized boiler, and adiabatic fluidized bed combustion power plant systems prepared on this basis are estimated to be from 10% to 25% less than conventional power plants with stack gas cleaning. Energy cost savings are projected to range from 5% to 15%.[20] The pressurized fluidized bed combustion combined cycle power plant offers the greatest potential for economic savings. A summary of the cost projections and the basis is shown in Table 3.

TABLE 3

Economic Comparison of 600 MW Plants

(3% Sulfur Coal, 90% Sulfur Removal)[a]

	Conventional steam power plant with limestone scrubbing	Atmospheric pressure boiler once-through limestone sulfur removal system	Pressurized fluid bed boiler combined-cycle once-through dolomite sulfur removal system[b]	Adiabatic combustor combined-cycle once-through dolomite sulfur removal system
Plant cost, $/kw[c]	530[d]	410	395	470
Energy cost, mills/kWh[e]	22.4	19.8	19.0	21.4

[a] 15%/year
70% capacity factor
Coal @ 80¢/GJ (10^6 Btu)
1977 operation of PFBB and adiabatic combustor
1978 operation of conventional plant and APFBB

[b] 17.5 excess air, three stages particulate removal including a granular bed filter

[c] based on total costs for November 1974 start of project, and including allowances for 6% contingency, 7.5% interest rate during construction, 7.5% escalation during construction, all plus 2% G&A

[d] $115/kw for limestone scrubbing

[e] dolomite or limestone supply and disposal cost, $10/mg

DEVELOPMENT NEEDS

Three factors will determine the commercial use of fluidized bed combustion systems: the ability to implement integrated programs, the resolution of technical developments, and the action taken by government on fuel use and environmental regulations.

The greatest development need is the need to implement integrated programs with responsibility and authority given for producing a commercial process. There have been and continue to be important projects to investigate specific problem areas, to operate specific test facilities, to assess market applications, and to plan development programs. Component test facilities, process development units, and demonstration plants are planned and cooperative efforts with other countries are underway.[9] In order to efficiently and economically produce a commercial system, these segmented efforts must be effectively integrated and directed.

The Energy Research and Development Administration, the Electric Power Research Institute, and the Environmental Protection Agency have presented programs on fluidized bed combustion and are evolving working relationships.[9] The resultant relationship between these parties will play an important role in the effectiveness of the commercial utilization of fluidized bed combustion in the United States.

No problems have been identified which preclude commercialization or meeting environmental standards. The technical uncertainties are specific for each fluidized bed combustion process concept. Particular concerns for the three reference concepts discussed include:

Atmospheric Pressure Fluidized Bed Boiler -- the principal uncertainties are associated with the reliability of the coal and limestone feed system, the life of boiler tube materials, the control and turndown capability, and the environmental impact. The coal feed system is of concern due to the large cross-sectional area required for utility power plants. Approximately 500 feed points would be required for a 1000 MW_e plant based on the available data. Boiler tube materials corrosion is an

area of concern. Due to the lower combustion temperatures the corrosion may be reduced from that in conventional plants. However, the specific corrosion behavior must be determined since local cyclic oxidizing/reducing atmospheres or the release of corrosive contaminants (e.g. sodium and potassium) from the sorbent may result in a corrosive environment which will require higher grade alloys for heat transfer surface. The projected techniques to meet turndown and load follow requirements must be demonstrated. For example, the use of bed temperature to control load will require a variation in the molar limestone calcium to sulfur ratio from $\sim$ 2.5 to > 5 to maintain > 80% sulfur removal efficiency.[20] The impact of this technique on spent sorbent disposition must be assessed. Spent sorbent disposal is subject to regulations which may require further processing. While available data do not indicate a problem, there are no standards or criteria which have been established to permit a final assessment of the spent sorbent.

Pressurized Fluidized Bed Combustion Boiler -- three primary development areas are identified as immediate needs: experimental tests on a commercial scale test facility with integrated auxiliary components, experimental and analytical studies to establish gas turbine performance, and experimental and analytical programs to ensure compliance with environmental regulations. The work to establish gas turbine performance must include investigation of combustor emissions (particulates and trace elements), investigation of alkali metal and particulate control, and gas turbine analyses and tests. Systems studies indicate that low temperature particulate removal (e.g. water scrubbing) can be economically attractive.[20] This would be an alternative to high temperature particulate control and is significant since it may offer improved power plant reliability. Control and turndown of the integrated combined cycle system must be demonstrated. A particular area of concern is to develop control techniques to protect the gas turbine. If dolomite is used as the sorbent and bed temperature is used for control, data indicate the molar calcium to sulfur feed ratio can be maintained approximately constant from 1.5 to 2.0. This is expected to improve the ability to control trace element release

and to meet environmental constraints. Heat transfer surface corrosion is not as critical in the pressurized system since the total surface is small and is thus not as sensitive to tube material costs. Reliable coal and sorbent feeding must also be demonstrated. The pressures (7-15 atm) proposed for pressurized fluidized bed combustion systems do not result in any engineering problems or safety concerns.

Pressurized Adiabatic Combustor -- the primary concerns with the adiabatic combustor are similar to those for the pressurized boiler. Problems which may result from the heat transfer surface in the bed are eliminated. The particulate control to achieve gas turbine performance must be carried out at elevated temperature for this case.

The fluidized bed air heater systems operating at atmospheric pressure and elevated pressure have an additional concern to those associated with the reference designs. The higher heat transfer surface temperatures (e.g. up to 800-850°C vs. 600°C) may limit the tube life due to corrosion by the combustion gases. This must be balanced against the minimization of particulates to the gas turbine in the closed cycle concept and the reduced particulate loading in the pressurized air-cooled concept.

Once-through limestone/dolomite systems are projected for first generation fluidized bed combustion systems requiring sulfur removal. Work is recommended to confirm the ability to achieve high utilization of the limestone/dolomite on commercial scale plants and to develop regenerative calcium-based sorbent processes to minimize the sorbent requirements for second generation systems.

Government action on energy policy and environmental constraints are required in order to guide the development of fluidized bed combustion systems. Action to stimulate coal and formulate energy policy have direct impact on the application of fluidized bed combustion. The importance given to the use of low grade fuels, nitrogen oxide minimization, the ability to achieve high SO_2 removal efficiencies (e.g. > 90%), the land use impact from spent sorbent, and related areas will impact the commercial use of fluidized bed combustion and the development program requirements.

ASSESSMENT

Atmospheric and pressurized fluidized bed combustion systems offer the opportunity to increase our ability to utilize our fossil fuel resources while reducing the environmental impact and reducing cost for the specific applications of electric power generation, steam production, or heat utilization. No technical problems have been identified which preclude commercialization. The primary need is for integrated programs with accepted responsibility and authority to commercialize a complete system. Government action on fuel use and environmental regulations will have a significant impact on the commercial use of fluidized bed combustion, particularly for industrial applications.

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MEYERS PROCESS: PLANT DESIGN, ECONOMICS AND ENERGY BALANCE

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I. INTRODUCTION

Coal can be desulfurized prior to combustion using the Meyers Process (1, 2) to meet governmental requirements for sulfur oxide emissions.

The physical form of the coal remains unchanged, only sulfur and some inorganic materials are removed. The process removes up to 80 percent of the total sulfur content of coal through chemical leaching of 90 to 95 percent of the pyritic sulfur contained in the coal matrix with aqueous ferric sulfate solutions at temperatures of 90° to 130°C. The ferric sulfate content of the leach solution is regenerated at similar temperatures using air or oxygen, and elemental sulfur and iron sulfates are recovered as reaction products or alternatively gypsum can replace a portion of the iron sulfates as a product.

The process consists of several steps including crushing, chemical treating, sulfur removal and solution regeneration (Fig. 1). While the process is new, the unit operations are based on various existing technologies such as: processes for the heap leaching of copper (operation 1), (Fig. 1), regeneration of steel mill waste pickle liquor (operation 3) and the recovery of elemental sulfur from volanic ash (operation 2). The chemistry of the process is represented by the treating step (Equation 1) and the solution regeneration (Equation 2):

$$FeS_2 + 4.6\ Fe_2(SO_4)_3 + 4.8\ H_2O \longrightarrow 10.2\ FeSO_4 + 4.8\ H_2SO_4 + 0.8S \qquad (1)$$

$$2.4\ O_2 + 9.6\ FeSO_4 + 4.8\ H_2SO_4 \longrightarrow 4.8\ Fe_2(SO_4)_3 + 4.8\ H_2O \qquad (2)$$

A set of products which can be obtained from a 440 tph (6400 hrs/yr operation) Meyers Process plant removing 1.0 percent by weight pyritic sulfur from coal are shown below:

Product	Tons/year
Elemental Sulfur	11254
$CaSO_4 \cdot 2H_2O$	15137
$FeOH\ SO_4$	74509

The process applies primarily to coals rich in pyritic sulfur rather than those with mainly organic sulfur. Such coal is found in the Appalachian region of the United States which now supplies 60 percent of the current U. S. production. This production can be lowered to sulfur contents of 0.6 to 0.9 percent for an estimated one-third of the production, at which level the emission requirements for new power plants can be met (3). Most remaining Appalachian coal and Eastern Interior Basin coal can meet state and local standards for existing power plants after treatment by the Meyers Process (4).

II. TEST PLANT

The Process is under development for the Environmental Protection Agency. Because of the success of extensive laboratory and bench-scale testing (over 200 separate material balanced leach and regeneration runs on over 30 different coals), the process is now in scale-up. A Reactor Test Unit (RTU) sized to process up

to 8 tons per day of coal is being built, under the sponsorship of EPA at TRW's Capistrano Test Site (Fig. 2). The RTU facility will be capable of on-line evaluation of the following critical process operations.

- Leaching of pyritic sulfur from 10 to 100 mesh top-size coal,
- Regeneration of ferric sulfate both separately, for processing of larger top-size coal and in a single vessel with the leaching step for processing of suspendable coal,
- Filtration of leach solution from reacted coal,
- Washing of residual iron sulfate from the coal.

Iron sulfate crystallization, elemental sulfur recovery and coal-drying unit operation will be evaluated in an off-line mode. Coarse coal processing (1/4 to 3/8 inch top-size) has been very promising in laboratory tests (5). If this approach proves out in bench-scale evaluation, coarse coal leaching units may be added to the RTU.

Processing fine coal allows the highest rate of pyritic sulfur removal; while processing coarse coal, although slower, allows lower cost coal dewatering units and the direct shipping of desulfurized coal product without need for pelletizing.

III. ENGINEERING DESIGN AND COST ESTIMATES

Conceptual designs or cost estimations of full-scale plants have recently been prepared by a number of organizations including TRW (5), Dow Chemical USA (6), Exxon (7) and EPRI - U. of Michigan (8). They show that the process is economically competitive with projected costs for flue-gas scrubbing, particularly for small installations or the average coal-fired electric utility where test load factors are less than 90 percent.

Some of the process variations which have been tested and engineered include the following: 1) leaching and regeneration temperatures of 70 to 130°C, 2) leaching and regeneration in the same or separate vessels, 3) removal of generated elemental sulfur by vaporization or solvent extraction, 4) air vs oxygen for regeneration, 5) feed coal top-sizes of 1/4 inch, 14 mesh, 100 mesh and near 200 mesh, 6) slurry solids content from 20 to 33 percent, 7) oxygen partial pressures of 1 to 8 atm, 8) combinations of the process with coal cleaning plants and 9) colocation with the power plant at a mine or at a separate site.

A processing cost range (investor financed) of 7 - $14/ton of coal is reported by Dow (6) for treatment of fine coal, with a thermal efficiency of 90 percent from run-of-mine coal crushing to desulfurized and crushed product coal. A processing cost range (utility financed) of $6 to $10/ton is reported by TRW (5) for treatment of fine coal and $3 to $6 for treatment of coarse 1/4 or 3/8 inch top-size coal. Coal yield, on a btu basis, is 94 percent for fine coal and 90 to 95 percent for coarse coal. Overall process thermal efficiency based on an Exxon developed flowsheet was recently calculated at 92.1 percent (7).

A TRW developed flow diagram for treatment of fine coal (14 mesh top-size) by the Meyers Process is shown in Figure 3. In this design, leaching and regeneration is accomplished in the same vessel at a temperature of 120°C and oxygen partial pressure of 1 to 5 atm. Approximately 85 percent of the pyrite is reacted in the first vessel. Reaction is completed, to the 90 to 95 percent level, in the second reactor. The coal is then washed free of residual iron sulfate and conveyed to a dryer where generated elemental sulfur is vaporized and the coal is dried. The products of this particular design are those described above. The capital cost for a coal desulfurization plant to provide low sulfur utility coal,

which was developed from a detailed flow diagram and equipment list, is $100/KW power plant name plate capacity (5). This is for removal of 3 percent sulfur from raw run-of-mine coal. For lower pyritic sulfur coals or cleaned coal, the capital cost is diminished. The relative cost of each section is shown in Figure 3.

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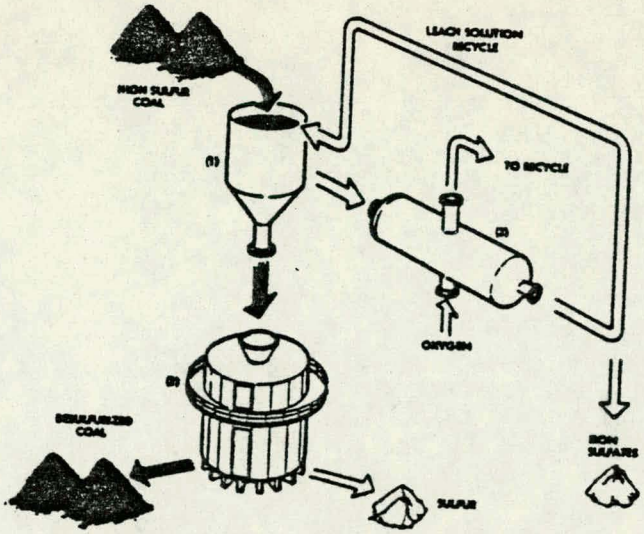


Fig. 1. Schematic of Meyers Process for Coal Desulfurization

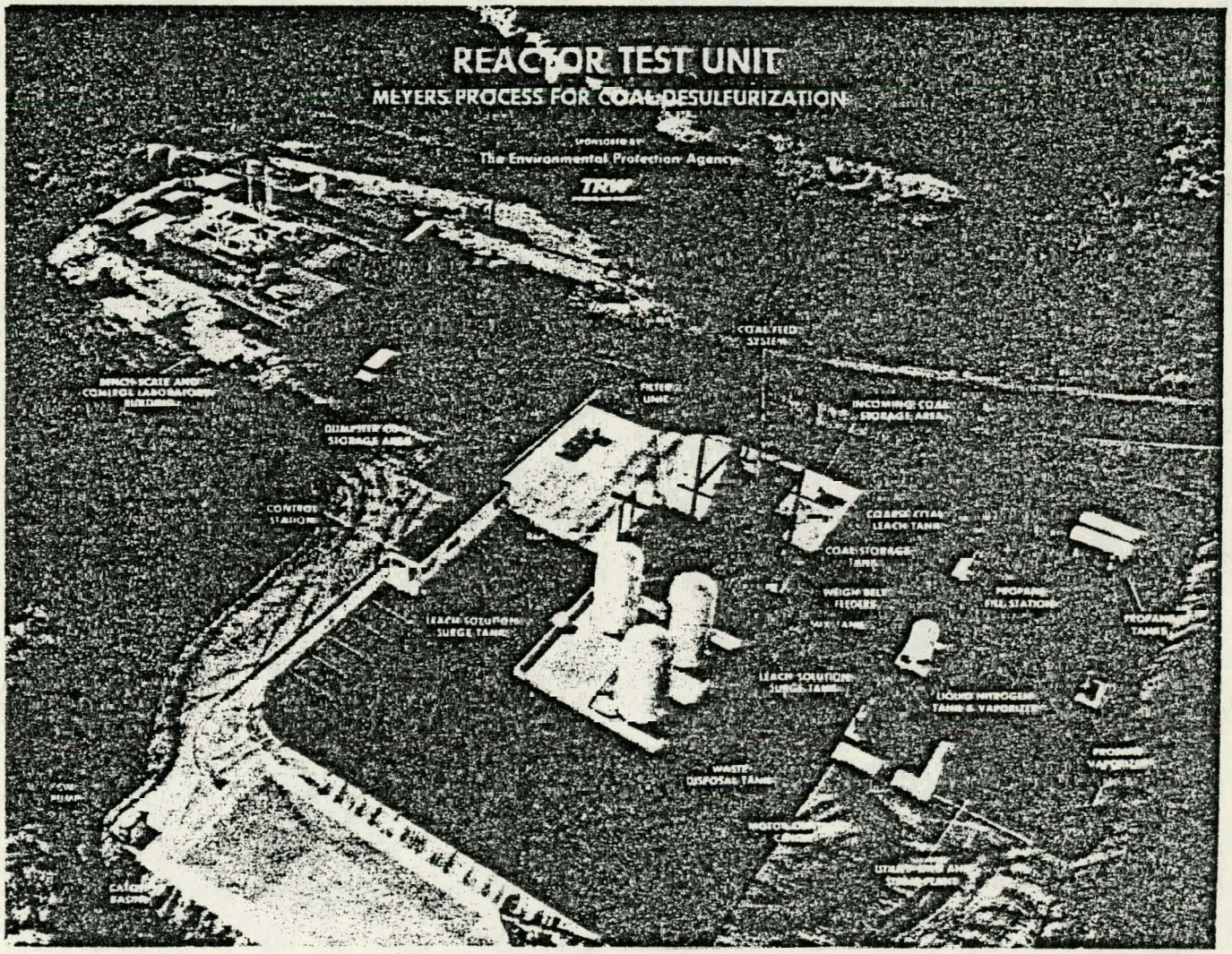


Fig. 2. Reactor Test Unit

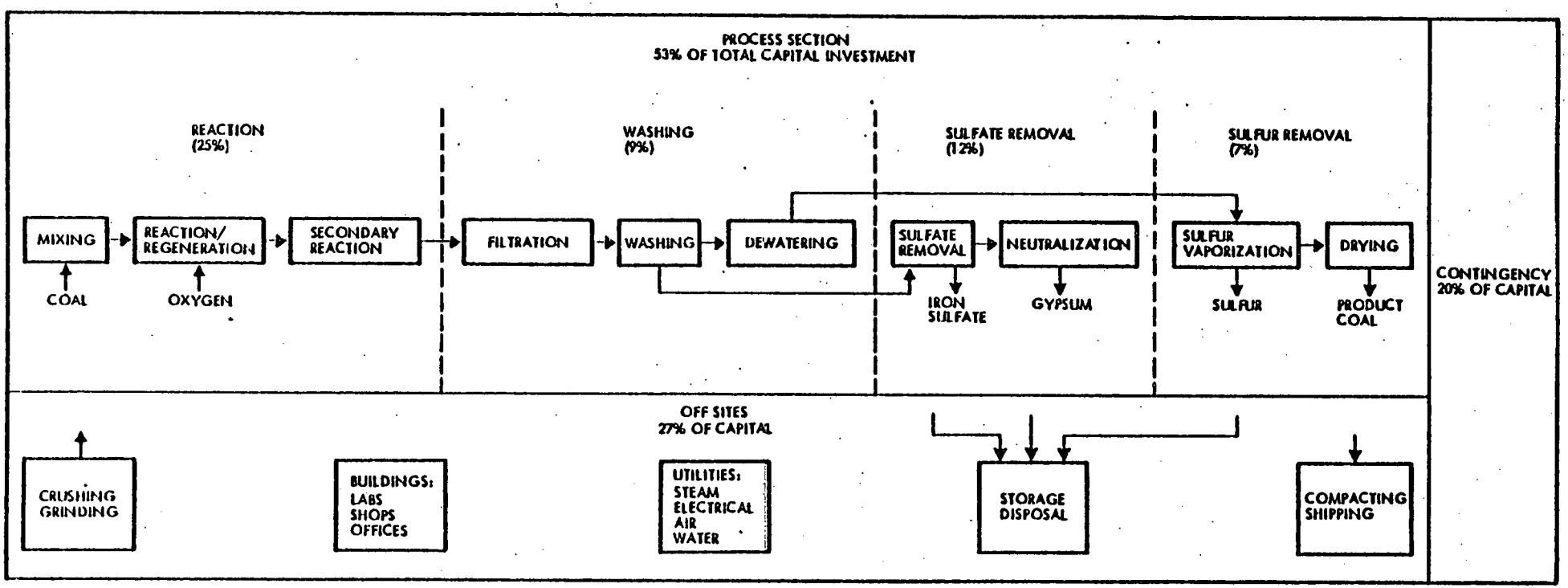


Fig. 3. TRW Process Flow Diagram for Fine Coal Processing

THE H-COAL® PROCESS

By

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Hydrocarbon Research, Inc.

A subsidiary of Dynalectron Corporation

As Presented At

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ABSTRACT

Since our last report in November 1975, three additional H-Coal® PDU runs have been completed. PDU Run 2 achieved all of the planned operating conditions, and demonstrated that very little catalyst breakdown and catalyst loss occurred at commercial gas velocity. PDU Run 3 demonstrated that under planned Pilot Plant operating conditions for the boiler fuel mode, the fuel oil product would contain 0.5% sulfur from an Illinois No.6 coal containing 3.5% sulfur. PDU Run 4 demonstrated that design conditions could be achieved in the syncrude mode with recycle of hydroclone overflow containing residual oil and solids to the reactor and result in an increased yield of distillate oil. A precipitation system on solid-liquid separation has been developed which produces a 500°F+ fuel oil product containing 0.065% ash from a fresh hydroclone underflow feed obtained on-line from the PDU.

It appears that the major technical hurdles in the H-Coal Process have been cleared, and that the emphasis is shifting to technology verification and product application studies.

INTRODUCTION

The H-Coal Process has been under development for over ten years and has been supported by government and industry. Work has been carried out on bench-scale units handling 25 to 100 pounds of coal per day and in a Process Development Unit (PDU) handling three tons per day. Some fourteen coals, including bituminous, subbituminous, lignite and brown coal, have been tested.

In September, 1974, Hydrocarbon Research, Inc. received a first-phase contract to design a nominal 600 ton per day Pilot Plant to demonstrate the H-Coal Process. The contract is jointly funded two-thirds by government and one-third by industry. At the present time, two oil companies and Electric Power Research Institute are industry representatives. The State of Kentucky also provides financial funding.

The Pilot Plant will be located at Catlettsburg, Kentucky, adjacent to the Ashland Oil Refinery. It is designed for two modes of operation: (1) to process 600 tons of coal per day to produce 1,820 barrels of fuel oil containing no more than 0.7 W % sulfur from coal containing more than 3 W % sulfur (fuel oil mode); and (2) to process 200 tons of coal per day to produce 705 barrels day of synthetic crude (syncrude mode). The design basis for

the fuel oil mode of operation is based on bench-scale experimentation. The current laboratory program has been directed towards confirming in the PDU the design basis for the Pilot Plant.

An uninterrupted run of twenty-three days (PDU Run 1) was completed in November, 1974. This run closely simulated the Pilot Plant mode of operation and did achieve the target level of 0.7% sulfur in the fuel oil product. Experimental results were reported at the 68th Annual Meeting of the A.I.Ch.E., Los Angeles, California in 1975. Catalyst deactivation curves were described in terms of changes in coal conversion, hydrogen consumption, product distribution and sulfur content of the fuel oil with catalyst age.

Since the completion of PDU Run 1, HRI has continued two major areas of activity: Pilot Plant engineering, and supporting research at the Trenton Research and Development Center. Long-lead equipment has been ordered for the Pilot Plant, and site preparation has begun at Catlettsburg.

Recent correlations of the data predict that catalyst consumption to achieve 0.7% sulfur fuel oil will be 0.75 pound per ton of coal compared to the design value of 1.0 pound per ton. Hydrogen consumption was 3600 SCF per barrel of products compared to the target value of 4000 SCF per barrel. Performance requirements set initially for the fuel oil mode have been met and exceeded in the run.

RECENT EXPERIMENTAL RESULTS

The heart of the PDU is the ebullated bed reactor shown in Figure 1. After the completion of PDU Run 1, an external gas-liquid separator was installed as shown, and the recycle tube was modified to facilitate gas-liquid disengagement. These modifications were essential to the successful execution of PDU Run 2 to demonstrate that only an insignificant amount of catalyst attrition and breakdown and catalyst carryover occurred at the high gas velocity chosen for the Pilot Plant and intended for the commercial unit.

PDU Runs 2, 3 and 4 have been carried out with the recycle of hydroclone overhead (residuum recycle), as shown in Figure 2. A precipitator has been incorporated into the Process Development Unit so that this solid-liquid separation unit can be operated with the fresh hydroclone bottoms and underflow as feed with the precipitant derived from the H-Coal operation.

Illinois No.6 coal from the Burning Star Mine, containing 3.5% sulfur, was used for the three PDU runs. The catalyst used was a commercial cobalt molybdenum hydrodesulfurization catalyst. The operating conditions and process performances are described below.

1. PDU Run 2 - Catalyst Attrition

The overall objective of this run was to test for catalyst attrition, catalyst breakdown, and catalyst carryover under fluid dynamic conditions as severe as any conditions planned for the Pilot Plant. The following targets were established:

- Commercial gas velocity
- On-stream time - Seven days of steady operation
- Equal to Pilot Plant design

The catalyst recovered from this operation was 97.6 W %. Representative samples of fresh and recovered catalysts were obtained by thieving and rifling. The length of one-hundred particles was measured. The data are presented in Figure 3. The measurements indicate a slight breakdown of particles longer than 6 mm to shorter particles, and a loss of catalyst shorter than 2 mm, about 2.4 W % of the total. The results of this run showed that catalyst degradation is negligible compared to the normal catalyst use rate due to deactivation.

2. PDU Run 3 - Pilot Plant Demonstration Run

The objective of this run was to demonstrate the boiler fuel mode operation under process conditions which duplicated as closely as possible the planned Pilot Plant operation. The goal was to achieve no more than 0.7 W % sulfur in the fuel oil product at a space velocity equal to that of the Pilot Plant, and with a feed slurry composition equal to that of the Pilot Plant. All of the

objectives of this PDU run were reached. The intended eleven-day run length was completed without interruption. The sulfur content of 400°F+ fuel oil product is plotted versus catalyst age and presented in Figure 4. At a catalyst age of 980 pounds coal per pound catalyst, the sulfur content of the fuel oil product was analyzed to be 0.43%. However, the smoothed curve for the entire period of operation, 1,500 pounds coal per pound of catalyst, indicates a sulfur content of 0.49% at 980 pounds per pound catalyst age. This is the predicated sulfur content based on bench-scale data for the Pilot Plant design in 1974. This low sulfur concentration indicates that additional process design freedom is available if the desirable product sulfur content is 0.7%.

3. PDU Run 4 - Syncrude Mode with Hydroclone Recycle

The objective of the run was to demonstrate the syncrude mode of operation in which severity is increased to a point at which the residual oil plus unconverted oil just supplies the hydrogen feedstock required for the process. The net product is thus all distillate, which is of particular interest to some of the sponsors of the H-Coal Process.

The yield of distillate oil is maximized by recycling residual oil and solids to the reactor via the hydroclone overflow, and this had never been demonstrated in the PDU.

PDU Run 4 was successfully completed in June 1976 to the planned catalyst age of 1500 pounds coal per pound catalyst. In the process, the hydroclone recycle rate was increased to achieve the solids to oil ratio in the reactor feed which is planned for the Pilot Plant. The result was a substantial increase in net production of middle and heavy distillate oil.

SOLID-LIQUID SEPARATION

A workable and economical solid-liquid separation system is required for the fuel oil mode. In this mode of operation reactor severity is reduced and throughput is increased such that the residual oil and unconverted coal are in excess of the amount required for hydrogen generation.

As a consequence, the plant produces net residual oil in the product, and this product must meet sulfur and particulate specifications, both of which require separation of solids from the net residual oil. No external materials are used and as a result the operation of the precipitator is simplified and its economics are improved.

This unit was placed on line with the PDU near the end of Run 3. The objective of the solid-liquid separation process is to obtain ash content equivalent to 0.11% in the final blended fuel oil product.

Filtration, magnetic separation and centrifugation have all been evaluated as part of the H-Coal program. None has appeared to be truly desirable for commercial operation. Instead, a precipitation unit has been developed at HRI which features reliable operation with a high yield of specification product and which uses a process-derived precipitant.

The data on fresh feed from Run 3 indicate that this separation scheme is capable of producing a 500°F+ fuel oil product containing 0.065% ash.

One-line operation is important to the development and demonstration of solid-liquid separation processes because the feed material definitely changes and becomes easier to separate, both with age and on re-melting or reconstitution.

SUMMARY

The past eight months have been very productive ones for the H-Coal program, both in the laboratory and in the field. The questions which were to be answered by the POU program have been answered affirmatively. We believe we have demonstrated a viable fuel oil mode operation of the reactor itself and made a significant contribution to the important companion technology of solid-liquid separation.

Design and advance ordering for the H-Coal Pilot Plant are proceeding on schedule. The technical tasks now are to support that program and begin the application studies which will be required to integrate coal-derived liquid fuels into the energy economy of the nation.

ACKNOWLEDGEMENT

The authors wish to acknowledge the financial and technical assistance given to this program by the U.S. Energy Research and Development Administration, Electric Power Research Institute, Ashland Oil, Inc., the Commonwealth of Kentucky, and by prior participant oil companies.

FIGURE 1

EBULLATED BED REACTOR

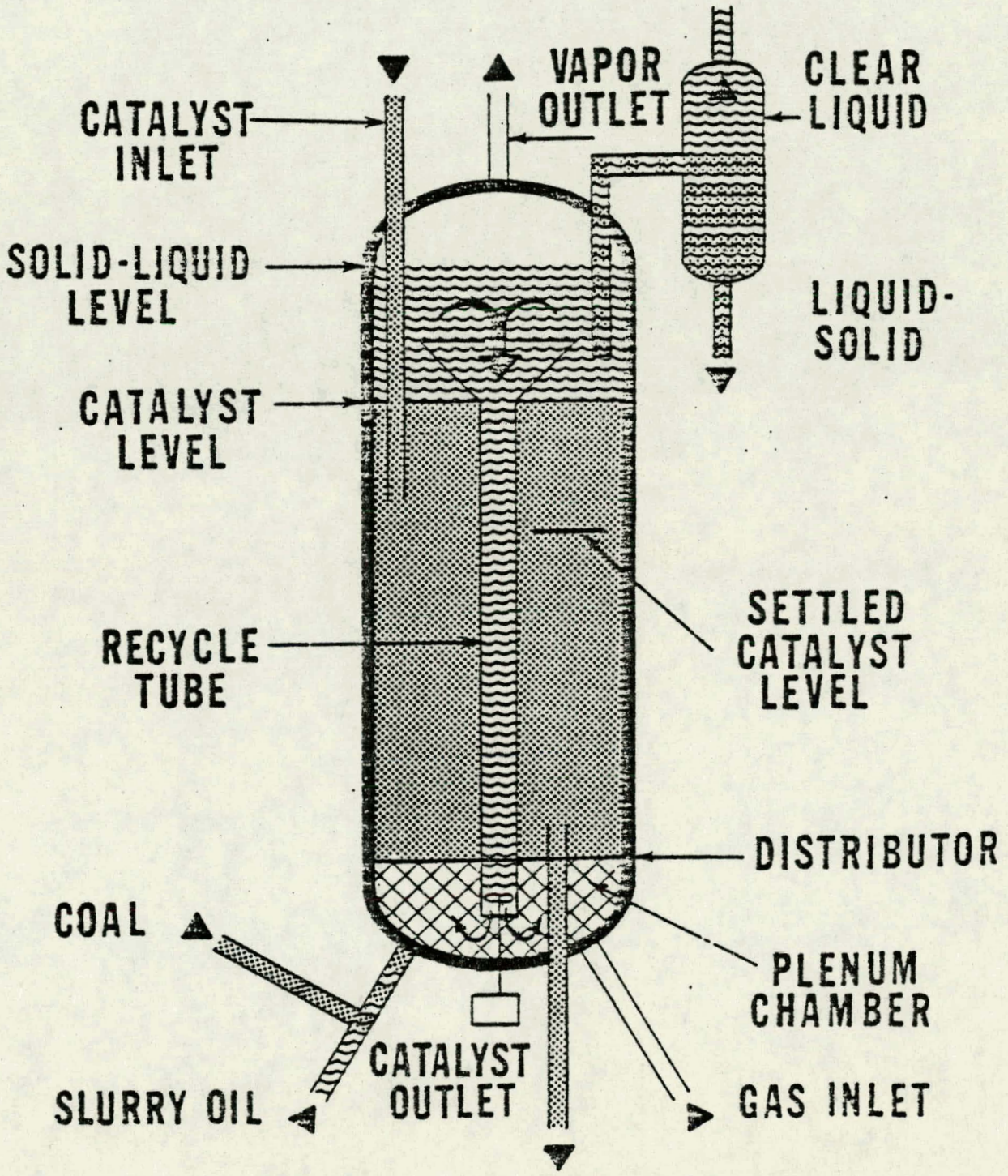
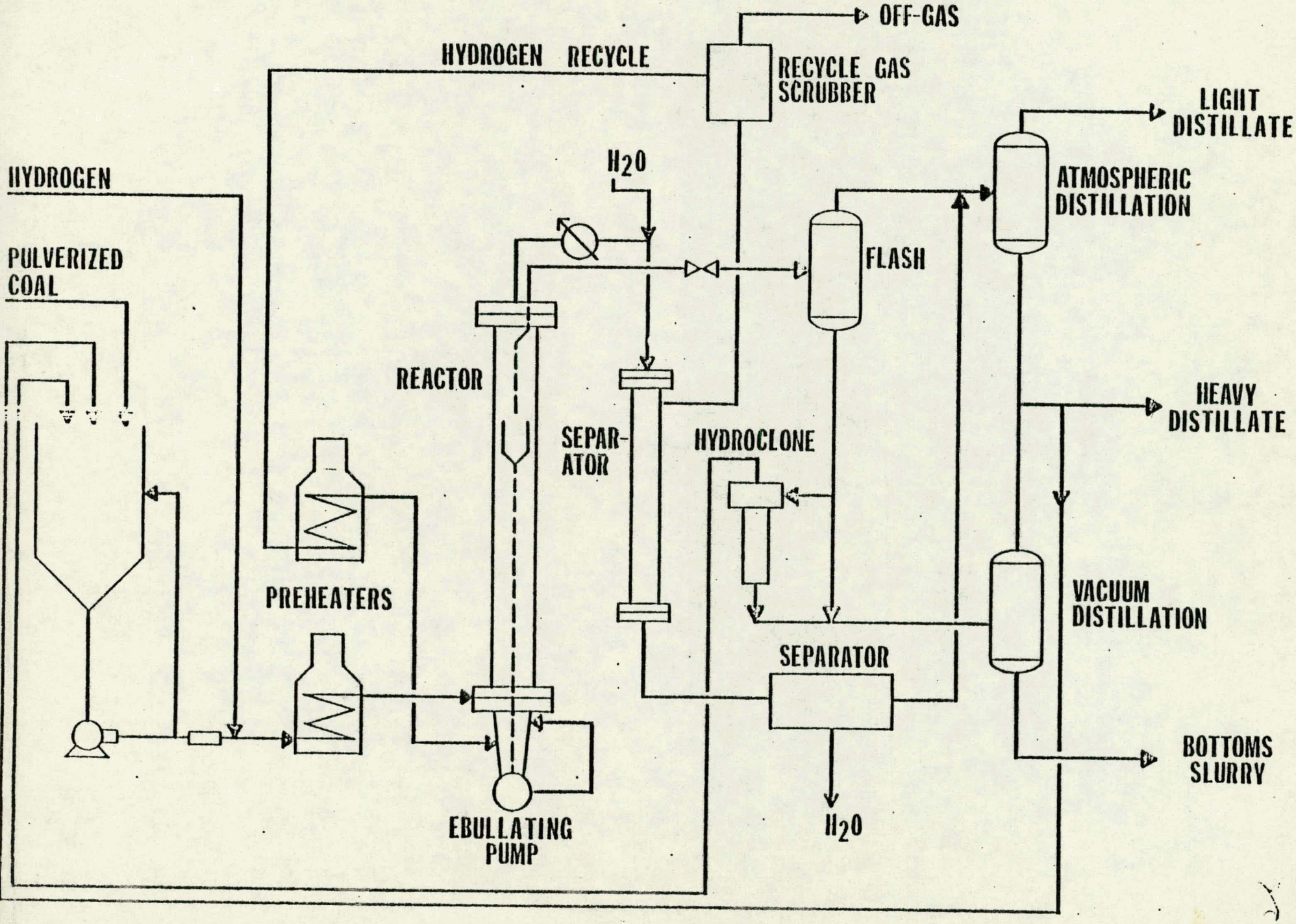


FIGURE 2 - H-COAL PROCESS DEVELOPMENT UNIT



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FIGURE 3

CATALYST PARTICLE SIZE DISTRIBUTION

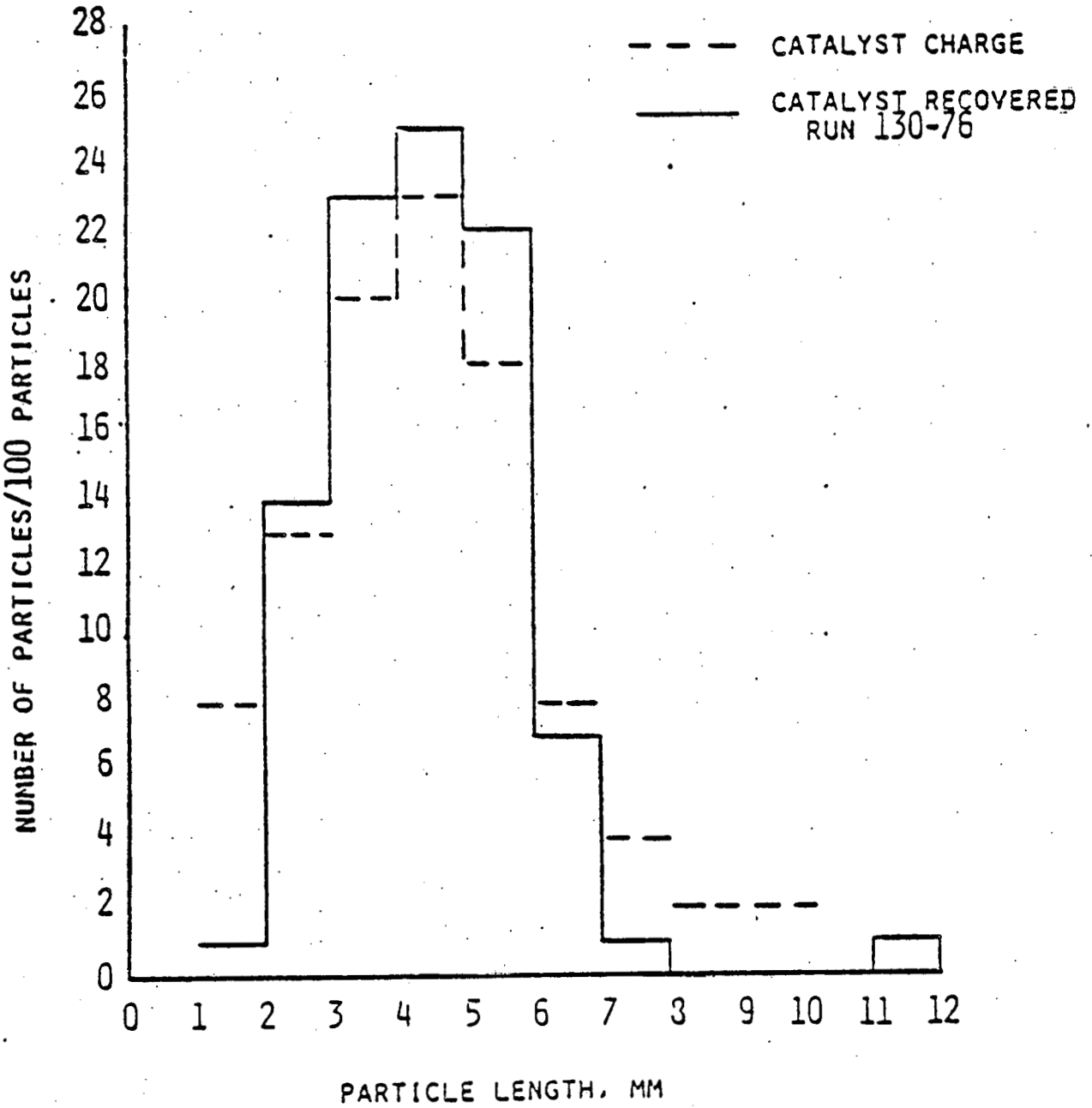
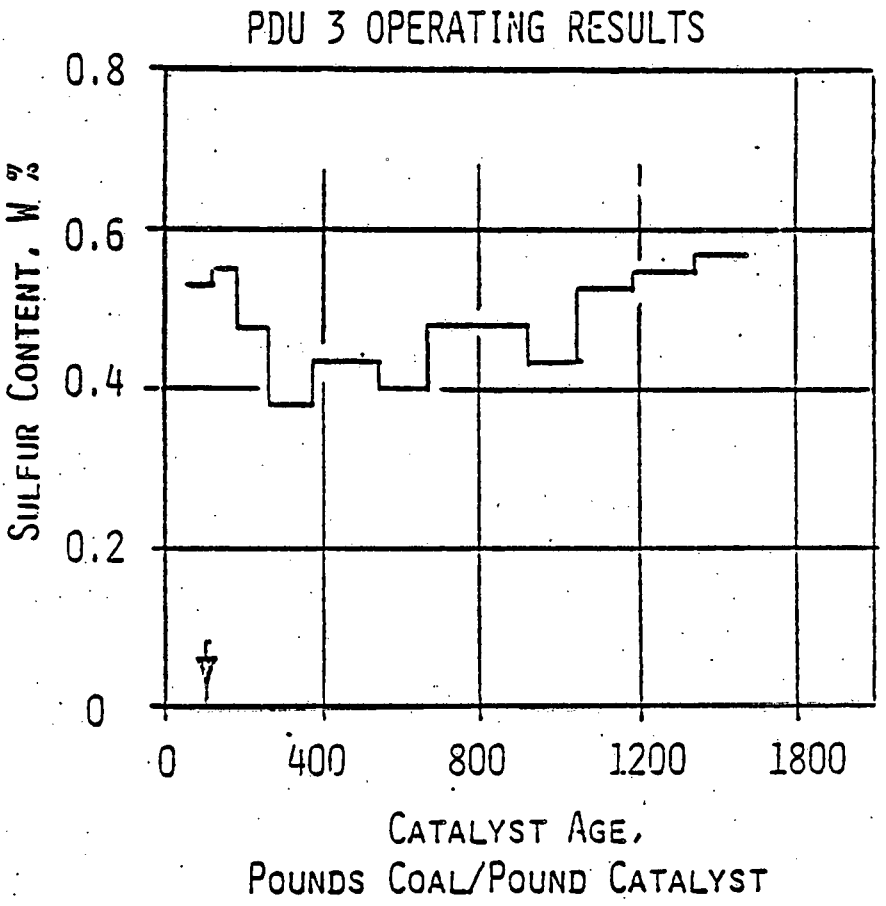


FIGURE 4


PDU 3 OPERATING RESULTS
0.8
0.6
0.4
0.2
0
SULFUR CONTENT, W %
0
400
800
1200
1800
CATALYST AGE,
POUNDS COAL/POUND CATALYST

A NEW CONCEPT FOR THE MANUFACTURE OF LOW SULFUR FUELS AND CHEMICALS FROM COAL

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UNIVERSITY OF PITTSBURGH,
PITTSBURGH, PA.

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BY

L. H. UDANI, R. R. MADDOCKS AND C. A. STOKES
(WEDNESDAY, 4 AUGUST 1976, 1 PM)

ERRATA

PAGE NO.	PARAGRAPH NO.	LINE NO.	INCORRECT	CORRECT
1	2	9	distillation	distillations
1	3	13	plant	product
4	1	4	by-partial	by partial
5	5	5	enforced,	enforced
8	5	2	$6/bbl	$10-12/bbl
8	5	3	$35	$60-70

A NEW CONCEPT FOR THE MANUFACTURE OF LOW SULFUR FUELS AND CHEMICALS FROM COAL

In this paper, we want to put in perspective for you an approach to coal processing which we shall, for convenience, call the "coalfinery". This concept is completely analogous to the well known "fuels refinery" in petroleum processing. It would be surprising if others have not used our term, "coalfinery", so we apologize if we have appropriated, and have perhaps even misused, someone else's pet term. We want to show you that this kind of a simple coal refinery makes good engineering and economic sense on a commercial scale if one can put the low sulfur fuel into power and the less pure fuel cut into chemicals, just as is done now with the analogous petroleum cuts.

In the petroleum industry, we find two extremes of refinery practices for turning crude oil into finished fuels and petrochemical feedstocks. At one extreme is the so-called "fuels refinery" and at the other the high conversion "gasolene refinery". The first of these is concerned with transforming crude oil into naphtha, jet fuel, diesel fuel, home heating oil, and heavy fuel oils largely by simple distillations and, nowadays, by additional hydrodesulfurizing although seldom is the bottom of the barrel desulfurized. The second type of refinery includes highly sophisticated catalytic processes which follow atmospheric and vacuum distillation. The processes of fluid catalytic cracking, alkylation, reforming, hydrodesulfurization, and hydrocracking may all be used in a single large refinery. Gasolene yield may be 50 - 85% of the barrel and #6 fuel only 5%, or even down to zero, if coking is used. In contrast, the fuels refinery may make 30 - 60% residual fuel depending on the crude used.

To build a fuels refinery costs about $1 to $2 per annual million btu's of output. Capital charges (12.5% return after tax and 7% depreciation) and processing costs total about $0.50 to $0.80 per MM btu. With crude oil at the current U.S. composite price and 93% yield on a btu basis, products must sell for an average of $2.4 - $2.7/MM btu to justify a new plant. With crude at world prices rather than U.S. controlled prices, the range for a new plant would be $2.9 - $3.2 per MM btu. These economics are on an instant basis, i.e. if the plant could be built at today's costs. Also, they do not include deep desulfurizing such that all #6 fuel would be 1% max. sulfur content. If we add on the construction cost the increase due to inflation for the 4 or 5 years needed to get approvals and to build a new grass roots refinery, the increase in crude oil prices over 4 or 5 years and the increased cost for desulfurizing the bottom of the barrel, then the ~~plant~~ product cost will be at least in the $3.50 - $4.00/MM btu range.

This is the real world of tomorrow as nearly as anyone can project it. It is within this sort of economic environment that we must project a coal refinery to make products to substitute for those we can make from heavy, high sulfur crudes in 1980 - 81. The "coalfinery" will cost more than a petroleum fuels refinery -- about $6 per annual MM btu, but the "crude", i.e. coal, will go in at perhaps $1.2/MM btu in 1980 - 81 versus at least $2.6/MM btu

for crude petroleum at that time. As pointed out above the petroleum fuels refinery capital cost must be escalated to 1980 - 81 and corrected upwards for bottom of the barrel desulfurization which is not being done in the U.S. today. There will, undoubtedly, remain a capital cost difference in favor of the petroleum fuels refinery and this difference must be traded off against the lower cost of input btu's from coal. So far, in the uncertain world we live in, no one has been willing to gamble the extra capital dollars to gain this long-term input btu cost advantage and the expectation of a lower overall cost long term.

Today, the refiner struggles mightily to avoid deep desulfurization, i.e. taking the sulfur out of the bottom of the barrel. He tries to sell off a high sulfur #6 fuel to cement plants, ships' bunkers, and to chemical plants in order to balance his refinery sales with as little hydrogen treating as possible. But we can follow the old cliche' - "What's sauce for the goose is sauce for the gander" - and process liquefied coal by hydrodesulfurization and distillation in reverse order to make a clean fuel (0.7% S, 0.1% or less ash) and a dirty fuel (high sulfur, high ash) just as is done in the petroleum industry. This involves a series of steps that are commercial in the sense of being offered by large, reliable engineering companies who will guarantee the operability of such a plant even though full throughput and efficiency guarantees may have to await the experience from first large commercial unit. While our definition of "commercial" may not satisfy everyone, recall that we are here today to talk about what can be and should be done. If it had already been done, we wouldn't be here. We would be talking to the investment analysts instead. A process can be considered commercially ready when it meets our prestablished criteria provided also that it will be suitably profitable within a reasonable time frame if not in the first year. In one of the other papers[1] the effects of cost inflation on the profitability of such projects were discussed pointing out the advantages of investing sooner rather than later in order to arrest the inflation of the capital charge component of total costs.

To be specific about the concept of "coalfinery and chemicals plant", we selected the hydroliquefaction step used in the Solvent Refined Coal (SRC) process.[2] We are most familiar with this process at Catalytic. Since 1970, Catalytic, Inc. has been actively involved in developing processes for obtaining clean fuels from coal via SRC technology, having designed, constructed, and operated the six ton-per-day pilot plant in Wilsonville, Ala.

We believe that the selection of this process is a good one for our discussion here because:

1. it is a simple process

2. it generates its own solvent; it is solvent sufficient

3. it is the oldest process and the one with greatest experience. The Wilsonville SRC Pilot Plant has operated over 2-1/2 years, including a 75-day continuous run and 85% on-stream factor in recent months. The ERDA Fort Lewis, Wash., 50 ton-per-day pilot plant is now operating for more than a year.

4. it doesn't rely on a catalyst, which may become contaminated. Ash is known to act as a catalyst, but no external catalyst is needed.

5. it requires minimum hydrogen addition

6. it permits minimum capital investment per unit throughput

7. it makes the product that meets EPA emission standards for utility fuels for sulfur ($< 0.9\%$) and mineral matter ($< 0.1\%$).

PROCESS DESCRIPTION

Figure 1 shows the conceptual "coalfinery and chemicals plant" process scheme. The primary process steps are:

Hydroliquefaction of coal

Solvent de-ashing

Partial oxidation of coal residue containing the bulk of the impurities

Ammonia from synthesis gas

The process begins with run-of-the-mine coal which is prepared and hydroliquefied in the usual fashion as in the SRC process.(2) The reaction mixture is flashed to remove excess hydrogen and other vapors and gases. Hydrogen and hydrocarbon gases are recovered after gas clean up and recycled or used as plant fuel. The flashed liquid is further stripped to remove a light oil fraction and some solvent. The light distillates are also used as clean plant fuel.

The stripped bottoms are treated in a solvent de-ashing unit.(3) Here a paraffinic solvent of a narrow boiling range around 350°F is added to the slurry of undissolved coal and ash in coal solution to promote the separation of the solids by gravity settling in a series of vessels.

The overflow is 70% of the charge and is essentially solid-free product. It is fractionated to separate the de-ashing solvent and process solvent from the SRC product. The de-ashing solvent and process solvent are recycled. The remaining 30% of the charge constitutes the underflow, which may have 6% solids. This is concentrated further by stripping to a pumpable consistency. It is then fed to a gas plant, which includes a gasifier and a gas clean-up train, to manufacture syn-gas by partial oxidation.

Syn-gas would be the source of hydrogen for the SRC process and ammonia synthesis. Co-products from the gas plant would be elemental sulfur, phenolics, CO_2 and slag. For start-up of the gas plant petroleum resid may be used.

The product slate of such a coal processing plant would include SRC, low BTU fuel gas, heavy and light distillates, and ammonia. The process makes excess solvent, which may be used as internal combustion turbine fuel or as a feed stock for carbon black. The SRC product would be used as a premium utility fuel or as a feed stock for needle or anode coke. It can also be converted to binder's pitch. Uses for ammonia are legion.

All the major process steps discussed above are currently practiced commercially. In regard to use of the underflow, for example, Figure 3 shows a plant VEBA CHEMIE[5] has on stream for making 450 tons-per-day of methanol and 1250 tons-per-day of ammonia by partial oxidation of petroleum resids. This plant is completely analogous to what we describe. The resids are similar to the de-ashing underflow of the SRC process.

To illustrate the economics of this concept, a 10,000 TPD (coal feed basis) plant is considered. The process conditions used are shown in Table I. A mid-western bituminous coal is the feed. Figure 2 shows a material balance of the process. Coal is ground to 14 mesh and dried. (Note: the power required for coal preparation would be considerably less than that required for pulverizing coal for direct firing). The coal is slurried with a recycle solvent (solvent-to-coal ratio is 2.0:1), and hydroliquefied at 850°F and 1700 PSIG pressure in presence of hydrogen at a space velocity of 50 to 75 lbs./hr. x cu. ft. About 90 - 95% coal goes into solution. H_2 uptake is about 2.0% MAF coal or 7000 to 8000 SCF per ton of MAF coal. For comparison, desulfurizing a high sulfur reduced petroleum crude requires 2000 - 6000 SCF per ton and pressures are similar.

The reaction product is flashed and degassed. After gas clean up a hydrogen rich gas stream is recovered and fortified with 166 TPD of make-up hydrogen for recycle. Some 330 TPD of clean fuel gases are also obtained as well as 150 TPD of elemental sulfur. The light distillates and some of the solvent are stripped off from the degassed reaction product to give an ash concentration of 4%. Together with the unconverted carbon the total solids concentration is about 8% in the resulting slurry.

The slurry is then charged to a de-ashing unit where equal weight of a light solvent is added.[3] The mixture is allowed to settle. The overflow constituting 70% of the feed to the de-ashing unit has a solids content of 0.01% and is fractionated to separate light solvent, process solvent (450 - 800°F), and SRC. Ash content of SRC would be less than 0.1%. Nearly 4080 TPD of SRC product is obtained from this stream.

The underflow is fractionated to recover light solvent and process solvent. The bottoms are 38% SRC, 34% unconverted carbon and 28% ash. At the bottoms temperature of about 550°F this mass would be pumpable. It serves as the gasifier feed.

The energy required to recover the light solvent from both overflow and underflow is estimated to be equal to or less than that for operating a precoat filter under pressure. Cost of make-up light solvent at 0.05% process loss is lower than cost of precoat. The process solvent is recovered in three separations. A net solvent excess of 833 TPD is realized.

By-product steam from the gasifier waste heat boiler and oxygen from an air separation plant are supplied to the gasifier to produce 4063 TPD (or 55.5 MMM BTU/day) syn-gas that is 46% H_2 and 47% CO (by volume). For producing 166 TPD of make-up hydrogen required in hydroliquefaction 40% of the syn-gas produced is utilized. The balance of the syn-gas is used to produce 1100 TPD of ammonia.

All the power requirements of the plant and process fuel requirements are met by using by-product steam, fuel gases and light and heavy distillates. Net product slate would, therefore, be as shown in Table II. The fuel oil equivalent of the product slate is 35,200 bbl/day of which 28,800 bbl/day would be clean utility fuels.

Commercialization of processes for the conversion of coal into clean boiler fuel has been inhibited by just two points:

1. The economics are always viewed looking backwards, so to speak, instead of looking ahead to what the competitive economics will be at on stream time and thereafter for 15 years or so. There is no practical significance in trying to prove that clean fuels made by coal conversion can compete with clean petroleum fuels now.

2. The process as a whole has, heretofore, not been put together in a series of steps that are fully acceptable from a technological standpoint by those who have to judge such ventures.

The scheme we have proposed overcomes both of these barriers, we believe. That doesn't say that industry will jump in to take the risk until there is more assurance that there will not suddenly be a break in petroleum prices or some last minute reprieve on burning high sulfur coal which will always be cheaper than upgraded coal. There is every reason to suspect that OPEC economists will attempt to keep crude oil price just low enough to discourage coal conversion plants. They can do the simple arithmetic as well as we. We, therefore, have a situation where some degree of Government/industry sharing of the risk is necessary. There are many ways this could be done, but that is a subject covered by other papers and events on the program.

As we have suggested, in looking at the economics of the combined "coalfinery" and chemicals plant, it is important to make this evaluation in the real world that will probably exist by the early 1980's. We will give our assumptions in some detail. First, we assume that the sulfur dioxide emission standards of 1.2 lbs. of sulfur dioxide per million btu's will be rigidly enforced along with EPA's current policy interpretations regarding new plant siting. Second, we assume that U.S. crude oil will be decontrolled by 1980 as promised by current legislation. Third, we assume that by that time, the United States will have run out of an available supply of 1%-sulfur fuel oil,

or fuel oil of such sulfur content as to meet the emission standards. This means that new supplies of such fuel oil will have to come from brand new processing facilities built to make the product from high sulfur crudes. For the necessary price of 1%-sulfur fuel oil, we are taking the objective evaluation of a major oil company[4] as an independent standard of fuel oil price. This value is $3.7 per million btu's f.o.b. refinery based on crude at $15/bbl. delivered.

We use high sulfur coal as feedstock. This means that in a competitive coal market, such a coal will be depressed in price below a coal which is practical to use directly in a utility boiler furnace without flue gas scrubbing. Hardly anyone would argue about the availability of such coals from existing mines, let alone from new mines that could be developed. Nevertheless, we are taking a price ($30 per ton) that reflects development of a new mine.

In the case of ammonia, we assume that by the early 80's, there will not be large enough amounts of natural gas available in the U.S. on a long-term contract basis to justify building a world scale ammonia plant regardless of the price that might be paid for this gas. This means that new supplies of ammonia manufactured in the United States must come either from the dregs of the crude oil barrel or from coal. We, therefore, take the selling price of ammonia as the price that would prevail if vacuum high sulfur resid were used to make ammonia. This is very conservatively estimated as $180 per ton f.o.b. plant.

Since the only two products leaving the plant are ammonia and low sulfur fuels (SRC and distillates) for power plant use and since we have fixed the value of these product streams, then it is possible to calculate a profitability for the overall venture. The distillate fuels that can be produced are priced out at the same btu value as the heavy fuel simply for convenience. Actually, the distillate fuels are more valuable. It is visualized that these distillate fuels would be used in internal combustion turbines for peak shaving power generation purposes.

As to profitability, we want to see if the venture has a chance to meet or exceed a standard profitability which we next define. It is logical to assume that if the Government would guarantee the repayment of 50% debt, then it would be possible to secure a 20 year loan at 10% interest rate. Because of this guaranteed debt situation and the uniqueness of the venture, we further assume that equity holders would be content with 12.5% after tax return on equity. An additional payoff to the equity holders is the know-how that they would gain by participating in such a venture. This know-how could lead equity holders to an opportunity to participate in many such ventures on their own once this concept has been proven successful. The depreciation on a 20 year basis would be sufficient to pay back the 50% debt in 10 years rather than 20, and thus there is a safety factor. From these assumptions, it is then possible to calculate a total capital charge factor for equity return, interest, and depreciation, the result being a charge of 20% of total capital per year on the average which is in the "utility return" category.

It might be argued that these assumptions are tailored to make the project look good. A little closer examination will show that they are perfectly reasonable assumptions in line with the concept that some minimum Government, but only a minimum, support is necessary. The assumptions concerning the competitive environment are alwyas subject to argument, but we believe ours are in line with the FEA and the major energy companies.

It is highly unlikely that this plant will operate, as assumed, at full capacity 330 days per year in the first few years. But, it is equally likely that it will far exceed this capacity in later years when the plant personnel have learned to cope with the process and have made the inevitable bottleneck removals. These two likelihoods taken together justify, we believe, an assumption of 330 days operation per year. The processes used in this plant are not very sophisticated, and their counterparts have been operated in one form or another in many plants for many years. For example, the solvent precipitation step is the analog of solvent deasphalting of petroleum which has been carried out on even larger scale than contemplated here. The hydroliquefaction of coal was carried out in Germany in the late 30's and through the early 40's. It is being operated by ERDA on the scale of 50 tons a day or more at the Fort Lewis, Washington, pilot plant. It is a thermal process with no catalysts present other than possibly some minor catalytic effect of the coal ash. The partial oxidation of heavy liquefiable hydrocarbon residues under high pressure is old art. Likewise, the gas processing and clean-up are old art, and certainly the synthesis of ammonia is not subject to question.

Lest it be overlooked, let us again stress a key point. If in coal processing, we can handle all streams after the initial liquefaction as liquids or gases under high pressure, the costs, both capital and operating, are greatly reduced. The "coalfinery" does just this as does the chemical plant which is satellite to it. It is not necessary to have a chemical plant if industrial fuel gas could be sold as the co-product. We have chosen ammonia because it is now made from natural gas and because the largest concentrated market for ammonia in the world lies in the eastern coal region of the U.S. Methanol could be made instead of or in addition to ammonia. While the chemical methanol market is limited, it is growing relatively fast.

Let us now turn to some results. Table III shows the total capital requirement for the "coalfinery" and ammonia plant based on 10,000 tons of coal charged per day. As a reference point, the fixed capital cost per annual MM btu of output for the "coalfinery" itself is approximately $6. By 1981, a grass roots petroleum fuels refinery will cost, we believe, about $2 to $3 per annual MM btu. If we had built and operated as many "coalfineries" and on as large a scale as petroleum refineries, we believe these two figures would long since have equated. This point needs to be stressed over and over again. We can never get the valuable learning curve cost reduction until we build commercial plants, as one of the keynote speakers has stressed.

In Table IV, the cash costs of operations are tabulated. Excluding coal, the really large items are capital related: 4-1/2% or $20.7 MM for maintenance and 2% or $9.2 MM for taxes and insurance. Coal cost is only 35% of revenues (Table VI). In a fuels refinery, crude oil cost would be more like 75% of revenues. These facts again emphasize that we need to learn how to process coal at a lower capital cost.

In Table V, we set up the financial assumptions and return standards that we believe to be reasonable for a plant of this kind if the debt were guaranteed in a fashion such as that envisioned in the U.S. ERDA plan pending Congressional approval. This table shows that we need $96 MM spread between revenues and cash cost on the average and $108 MM the first year to satisfy our return criteria.

The results in Table VI show that we have met the targeted return with room to spare. While this is highly gratifying, it does not give us room to be complacent about the difficulties in selling this concept to the investment community. What we have proved is that if we compare fuels from coal conversion and petroleum on the same basis, then we are ready to go ahead commercially.

If we use up the first year cash flow safety factor to reduce the clean fuel price, this price can be reduced $2.6/bbl or down to $20.8/bbl. If the 20 year average cash flow safety margin is used similarly, we can reduce the clean fuel price by $4 or down to $19.4/bbl.

If the plant produces 10% more than design, the incremental costs are essentially only the cost of the additional coal used which would amount to some $~~8~~ 10-12/bbl of clean fuel and about $~~35~~ 60-70 per ton of ammonia. These are strikingly low incremental costs and they illustrate the kind of improvements that are almost sure to be realized once commercial plants are operated. We cannot stress too much the need to realize the learning curve effect. With an overall thermal efficiency near 80%, and coal representing only about 1/3 of the revenue dollar, research to up the efficiency is not going to produce much effect although every bit helps. The big factor is learning how to produce more products from a given plant and, at the same time, going to world scale plants, i.e., 100 - 200,000 bbl FOE per day versus the mere 35,000 bbls/day used in this paper.

We have not looked at the effect of investment credit and accelerated depreciation which would greatly increase the cash flow over the first five years. We have preferred instead to look at the project on conventional long-range accounting which more properly reflects the kind of transfer prices that will prevail over the long run. If a project looks reasonably sound on this basis, then all the allowed tax incentives can be applied to make the front end economics even better. However, like death, taxes for the most part are not avoided -- they are simply deferred.

The flies in the ointment are numerous:

1. U.S. crude prices are controlled below world levels, and there is no great confidence that controls will actually be removed by 1980.

2. The oil companies will very likely not be able to charge replacement prices for low S #6 fuel in 1980 - 81 but will be forced by competition and/or price controls to reflect a pool price based on old and new facilities and crude prices below world prices.

3. Ammonia may be imported from Canada, Mexico, the Caribbean, or even the Middle East at lower prices reflecting lower gas costs. Further, if interstate gas is controlled in the U.S., intrastate prices will be lower because of a supply/demand effect. This will affect the pool price on U.S. ammonia. The controlled price of crude will also tend to lower the replacement cost of ammonia from residual fuel oil.

With these uncertainties plus the ever possible threat of capricious oil price actions by OPEC, it is doubtful if private capital will move into coal conversion without some additional incentives, such as a floor on clean fuel price sufficient to service the debt and give the equity holders assurance of getting all of their capital back eventually with enough return to at least compensate for inflation.

Our purpose has been achieved. We have illustrated a perfectly straightforward way to use coal to replace crude oil, and we have used only technology that is commercializable right now. While we have used ammonia as a sink for the bottom of the "coalfinery's" barrel, industrial fuel gas would serve almost as well even if taken off at $3.7/MM btu as used for the low sulfur fuel.

We recommend that industry look hard at this approach to commercialize three separate aspects at once.

(1) Low sulfur utility fuel from coal to supplant low sulfur petroleum fuel.

(2) Chemicals from coal to supplant natural gas.

(3) Industrial fuel gas from coal to supplant petroleum fuels and natural gas.

This technology would be furnished through, and engineered and constructed by, large U.S. engineering and construction companies throughly familiar with such technology. We, therefore, feel that the time is ripe for a consortium of industry to undertake a project to bring this concept to reality.

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TABLE I

BASIC PROCESS CONDITIONS

Coal	Midwestern Bituminous (R.O.M.) Feed Rate 10,000 TPD Moisture 10% Sulfur 3.2% Ash 6.6% Heating Value 11,500 BTU/lb.
Coal Particle Size	14 Mesh
Solvent/Coal Ratio	2:1
H_2 - Uptake	2% MAF coal
Coal Conversion	90%
Net Make of Process Solvent	10% MAF coal
Light Oils Produced	10% MAF coal
Deashing Solvent:Coal Solution	1:1
Deashing Overflow: Underflow	70:30
% Ash in Underflow	4% (before light solvent removal)
% Ash in Overflow	0.01% (before light solvent removal)
Light Solvent Loss in Deashing	0.05%
% Ash in Feed to Gasifier	28%

TABLE II

PRODUCTS MADE

	TPD	MM BTU/D	*F.O.E./bbl PER DAY
SRC	4084	131	21,000
Distillates	1292	49	7,800
Ammonia	1100		6,400
			35,200

*6.25 MM BTU/bbl

TABLE III

CAPITAL INVESTMENT (1980 DOLLARS)

	$MM
I. PLANT FACILITIES INVESTMENT	
Coal Prep. & Slurry Handling[6]	6.8
Hydroliquefaction[6]	82.5
Flash & Degas, Acid Gas Removal[6]	20.3
Deashing[9]	40.3
Stripping[6]	24.2
Gasification, O_2 Plant, Gas Clean Up[6][7][8]	104.7
Waste Treatment[6]	14.0
Steam Generation[6]	5.2
Power Generation[6]	34.5
General Facilities[6]	57.5
Sub-Total	390.0
1100 TPD Ammonia Synthesis Loop[10]	70.0
Total Plant Facilities Investment	460.0
II. WORKING CAPITAL	
30 Days' Cash Operating Cost	4.0
30 Days' Coal Supply	8.0
30 Days' Receivables at Cash Cost	14.0
Total Working Capital	25.0
III. TOTAL CAPITAL INVESTMENT	485.0

TABLE IV

CASH COSTS OF OPERATIONS

Plant Facilities Investment (P.F.I.)	$460 MM
Operating Costs Other Than Raw Materials	
	$MM/YR.
Operating Labor ($17500/Yr., 50 Men/Shift)	3.78
Supervision, (15% Oper. Lab.)	0.57
Maintenance Labor (1.5% P.F.I.)	6.90
Administrative (20% Oper. Lab., Super., Maint. Lab.)	2.25
Payroll Burden (35% all Labor)	3.74
Total Labor	17.24
Chemicals, Catalyst, Light Solvent	4.33
Maintenance Mat'l. (3% P.F.I.)	13.80
Operating Supplies (30% Oper. Labor)	1.13
Total Material	19.26
Water (40¢/1000 gal.)	1.15
Fuel	Internal
Power	Internal
Total Utilities	1.15
G.&A. (2% Revenues)	5.70
T.&I. (2% Cap. Inv.)	9.20
Total Fixed Costs	14.70
TOTAL CASH COST OF OPERATIONS OTHER THAN COAL	52.35
Coal @ $30/Ton	100.00
TOTAL CASH COSTS OF OPERATIONS	152.35

TABLE V

FINANCING ASSUMPTIONS
AND RETURN TARGETS

50% Debt, 10% Interest, 20 Year Term
20 Year Depreciation for Book Purposes
12.5% Return on Equity
50% Income Tax

Average Capital Charge: -

	% of Total Capital Investment	$MM/YR.
Average Interest	2.5	12
Depreciation (Book)	4.7	23
Equity Return	12.5	61
	19.7	96

Capital Charge 1st. Year: -

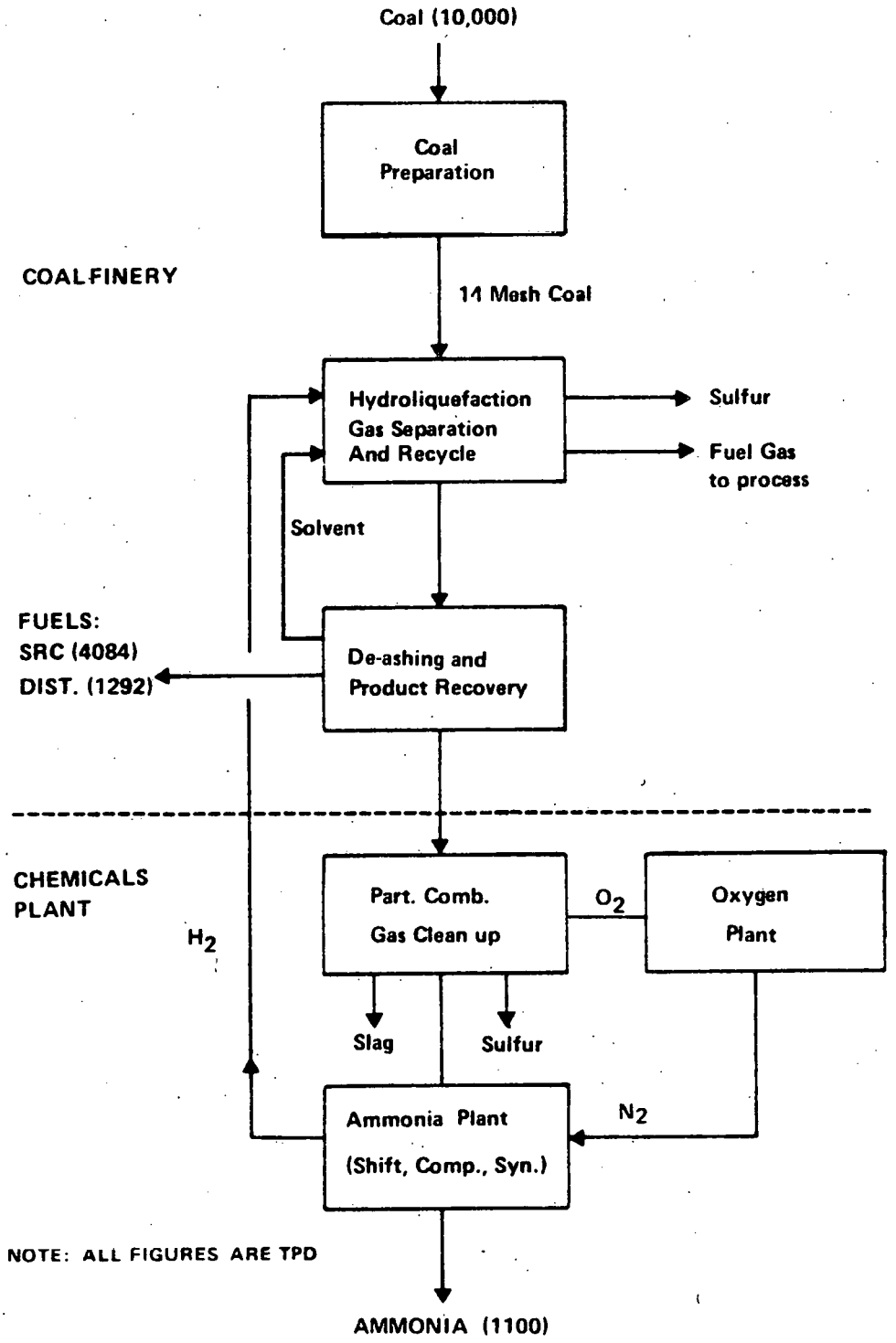
Interest	5.0	24
Depreciation (Book)	4.7	23
Equity Return	12.5	61
	22.2	108

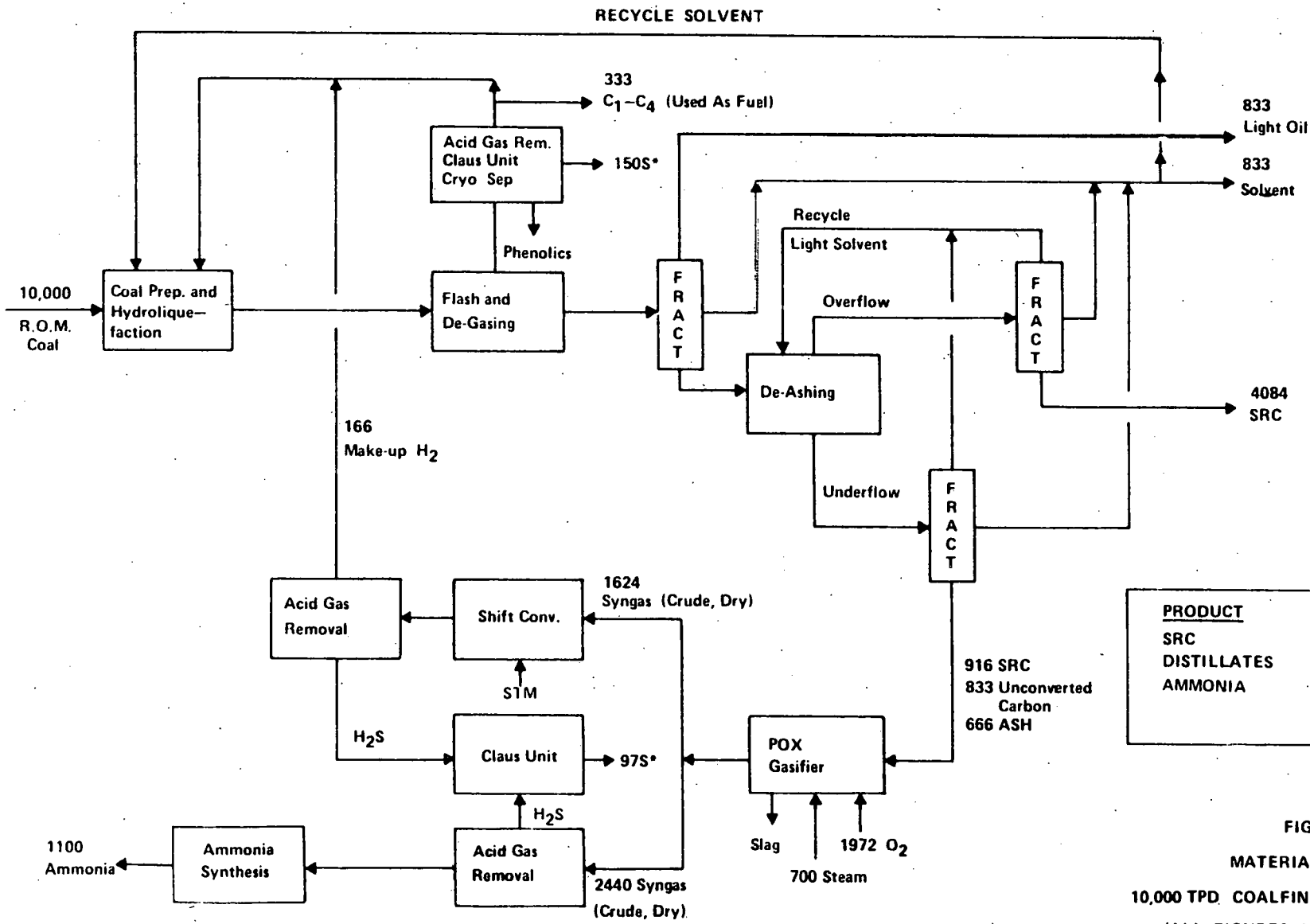
TABLE VI

ACTUAL VERSUS REQUIRED RETURN

	$ MM/YR.
REVENUES	
9,500,000 bbls. FOE at $23.4/bbl	222
363,000 tons NH_3 at $180	65
	287
Cash Operating Cost	(152)
Balance available for interest, taxes, equity return, and depreciation	135
FIRST YEAR CASH FLOW AFTER INTEREST AND TAXES BASED ON NORMAL DEPRECIATION AND WITHOUT INVESTMENT CREDIT	$ MM
Depreciation	23
Return After Tax & Interest	44
	67
Cash Flow Required First Year	
Debt Payment	12
Dividends (Return on Equity After Tax)	30
	42
Safety Factor First Year	25

FIGURE 1
COAL FINERY AND CHEMICALS PLANT





PRODUCT	T/D	MMMBTU/D
SRC	4084	130.7
DISTILLATES	1270	48.3
AMMONIA	1100	

FIGURE 2

MATERIAL BALANCE

10,000 TPD COALFINERY AND CHEMICALS PLANT

(ALL FIGURES ARE TONS PER DAY)

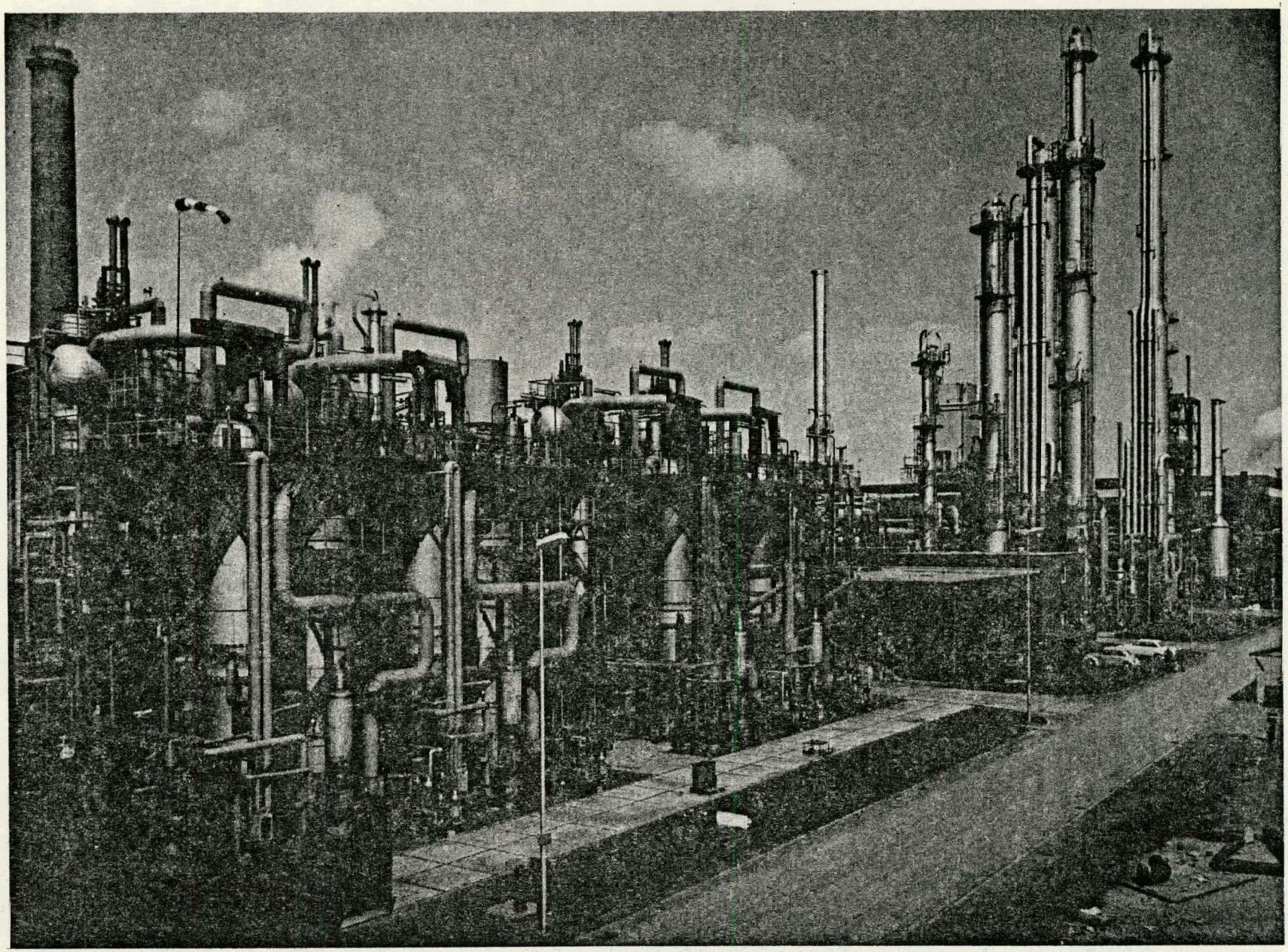


Figure 3

SHELL PARTIAL OXIDATION AND LURGI RECTISOL WASH IN

VEBA–CHEMIE AMMONIA/METHANOL PLANT

OPERATIONS AT THE WILSONVILLE SRC PILOT PLANT

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I. Introduction

The major sources of energy for the generation of electricity are water for hydroelectric generation, nuclear fuels, and the fossil fuels - gas, oil, and coal. Several facts about these energy resources are clear:

1. There are few suitable sites for hydroelectric generation which have not been developed.

2. Nuclear energy facilities cannot be provided rapidly enough to satisfy the need.

3. Natural gas is disappearing from the utility market.

4. Oil will be an uncertain resource which will be either completely unavailable or available only at a very high cost.

5. Coal is the only available choice from among the fossil fuels for developing long-term plans.

Much of the coal which is reasonably available to electric utilities in the eastern United States contains sulfur and ash at levels which make it unacceptable as a fuel. It is clear that major pollution abatement efforts are necessary and that two options are available: (1) to clean the coal prior to combustion or (2) to clean the flue gas after combustion. A great effort has been made to develop commercially acceptable processes for removing pollutants from flue gas. Although these efforts have encountered many technical problems and rapidly increasing costs, it is expected that a few of the processes will emerge with adequate reliability and within acceptable limits of cost. Even then, the inclusion of such a process as part of electric generating facilities will entail a number of disadvantages. For this reason, there is a strong justification to develop the available options for making a clean utility fuel from coal. The solvent refining process is a most promising option for making clean fuel from coal.

This paper describes operations at the 6-ton per day pilot plant at Wilsonville, Alabama where the solvent refined coal process is being investigated.

II. Background

The basic technology for the solvent refined coal (SRC) process was developed in Germany soon after World War I. Two German scientists, Pott and Broche, patented the basic process for dissolving coal and reducing its ash content in 1932.

Bench scale work on the present SRC process was carried out from 1962 to 1965 by the Spencer Chemical Company, under the sponsorship of the Office of Coal Research of the U. S. Department of the Interior. The Spencer Chemical Company was acquired by Gulf Oil Corporation and work on the SRC process was continued by the Pittsburg & Midway Coal Mining Company, a Gulf subsidiary. Under the sponsorship of the Energy Research and Development Administration, Pittsburg & Midway is operating a 50-ton per day SRC pilot plant at Ft. Lewis, Washington.

The Edison Electric Institute and The Southern Company system began a joint pilot plant project to study the key steps in the solvent refining process in March, 1972. The 6-ton per day pilot plant was built on the site of Alabama Power Company's Ernest C. Gaston Steam Plant located near Wilsonville, Alabama.

Catalytic, Inc. designed, built, and is operating this pilot plant, which began operation in January, 1974. The Electric Power Research Institute (EPRI) assumed the responsibilities of the Edison Electric Institute for utility industry project supervision in April, 1973. The U. S. Energy Research and Development Administration (ERDA) recently joined EPRI as a joint sponsor of the pilot plant operation.

III. Process Description

The following description of the solvent refining process is based on the simplified flow diagram shown in Figure 1 for the 6-ton per day facility at Wilsonville, Alabama.

Pulverized coal is mixed with two or three parts of a solvent fraction that is generated internally in the process. This recycled solvent fraction has a boiling range of about 350-800^{o}F. Hydrogen is added to the slurry of coal and solvent and the mixture is preheated and fed into a single-stage reactor or dissolver. The reactor operates at a temperature of 800-875^{o}F, and a total pressure of about 1700 psig. Under these conditions, approximately 93% of the carbonaceous material in the coal feed is dissolved during the residence time of approximately 30 minutes in the reactor. About 60% of the organic sulfur in the coal is converted to hydrogen sulfide. The hydrogen consumption is about 2-3% of the weight of the coal processed.

The effluent from the reactor is passed to a high pressure separator where the liquid and gas phases are separated. The liquid slurry is then subjected to a mineral separation step where the undissolved solids are removed. Filtration is the method of mineral separation initially used in the pilot plant. Other techniques for mineral separation such as the use of hydroclones and centrifuges also are being investigated. The filtrate from which the solids have been removed is flashed in a vacuum distillation column. Process solvent is recovered overhead and recycled to slurry the coal feed. Bottoms from the vacuum column form the solvent refined coal product which solidifies at about 350°F. The product solidification technique that was initially used in the pilot plant involved freezing the SRC by direct quenching in a water bath from which it was removed by a conveyor. The SRC product is now solidified on the water-jacketed plates of a vibrating conveyor.

In the pilot plant, the gas phase from the high pressure separator is scrubbed with a 20% caustic solution to remove the hydrogen sulfide produced in the reactor. In a commercial plant, one of the many commercial absorption processes would be used and the sulfur would be recovered for sale as elemental sulfur.

The solid residue from the filter is discharged to a rotary dryer. The filter residue contains a substantial quantity of wash solvent as an absorbed liquid. The wash solvent is removed as vapors from the rotary dryer, condensed to a liquid, and recycled to the process. The dried solids (mineral residue) which contain most the mineral matter and some undissolved carbon from the coal feed are stored for use in gasification studies. In a commercial plant, it is expected that the mineral residue will be fed to a gasifier to produce the hydrogen required in the process.

IV. Pilot Plant Operations

The pilot plant was completed in August, 1973. A series of pre-operational tests was begun, such that each item of equipment and all piping systems were checked for conformance with design requirements and expected operating conditions. Integrity and performance of the entire plant was checked with water and nitrogen prior to actual loading of solvent and coal. In January, 1974, using commercial creosote oil as the initial solvent, operations with coal were commenced. Solvent for sustaining the operation is derived from the feed coal in the process. The design value of the solvent-to-coal ratio (three-to-one) was reached in mid-February and sustained plant operations for a period of 45 days began early in April, 1974. The plant was then shut down for modifications and, upon starting up again in June, it achieved the first steady production of fuel product meeting the design specifications. Since then a number of significant goals have been achieved.

By the end of 1975 the pilot plant had been in operation with coal for more than 7800 hours, including periods of 45 and 75 days of sustained operation. Four bituminous coals (a mixture from Kentucky 9 and 14 seams, Pittsburgh seam, and two from Illinois No. 6 seam) and a Wyoming sub-bituminous coal have been tested. Typical analyses of these five coals are given in Table 1. Solvent refined coal product meeting the plant specifications of 0.16% maximum ash and 0.96% maximum sulfur has been produced from each coal. A fuel meeting these specifications and having a heating value of 16,000 Btu per pound could be used to achieve compliance with most emission limits and to comply with the EPA "new source performance standards." The operating conditions used to produce SRC meeting the plant specifications are given in Table 2. Typical analyses of the solvent refined coal produced from each coal are given in Table 3. The data in Table 3 show that the type of feed coal had very little effect on the properties of the solvent refined coal product. Bench-scale studies have shown that lignites, sub-bituminous, and bituminous coals all yield refined products having essentially the same characteristics.

A significant achievement was the production of specification grade SRC from the Illinois No. 6 (Monterey mine) coal which contained about 2.9% organic sulfur and 4.3% total sulfur. This indicates that the process can treat coals with organic sulfur contents greater than those found in most coals.

The production of process solvent from each coal was sufficient to maintain the required solvent inventory without the addition of make-up solvent during normal operations. However, losses of solvent during emergency shutdowns and problems with the filter cake dryer have required the addition of make-up solvent. The plant has operated for more than five months without adding any new solvent to the system.

The design capacity of the pilot plant is 500 lb/hr (28 lb/hr-ft^3) of coal fed as a 25% slurry in process solvent. Successful operation has been achieved with coal feed rates as high as 1000 lb/hr and with coal concentrations as high as 40% in the feed slurry. It is desirable to operate with high concentrations of coal in the feed slurry in order to reduce the amount of process solvent which must be pumped, heated, and distilled per unit of coal processed.

The yields of SRC are generally 50 to 70% of the moisture and ash-free (MAF) coal feed. Higher temperatures and longer reactor residence times cause a decrease in the SRC yield and an increase in the yield of gases and liquids. Coals with higher oxygen contents give lower yields of SRC.

Operating pressures have ranged from 1000 to 2400 psig, with the corresponding hydrogen partial pressure being 850 to 2000 psig. Operation at a pressure of 1000 psig gave a coal conversion of about 80% compared to about 93% at an operating pressure of 1700 psig. Operation at a pressure of 2400 psig also gave a coal conversion of about 93%, while

the yield of SRC decreased from about 60% to 50% of the MAF coal and the yield of organic liquids increased from about 10% to 20% of the MAF coal. The hydrogen consumption increased from about 2.3% to about 3.2%. Operations at a pressure of 1000 psig were unstable and resulted in some coking in the slurry preheater. The preheater was cleaned by a normal steam-air decoking procedure. The overall indications are that the hydrogen partial pressure should be at least 1000 psig.

Accumulations of fine, "sandy" fluidizable solids have been found in the dissolver at some conditions with several coals. Problems have been encountered in the letdown valves and vessels downstream of the dissolver when these solids are carried from the dissolver. The reasons for the accumulation of solids and methods of coping with the problems are being studied.

The temperature in the dissolver has been varied over the range of 775-870°F. Dissolver temperatures of at least 815°F are required to obtain the desired conversion, yield of solvent, and sulfur removal with the coals that have been tested. An increase in the dissolver temperature generally causes a decrease in the yield of SRC, an increase in the yield of gases and a decrease in the sulfur content of the SRC.

The primary method used for solids removal at Wilsonville has been filtration with a pressure leaf filter. This batch precoat filter has consistently produced SRC containing less than 0.16% ash. The amount of diatomaceous earth filter aid used is about 10 to 15% of the weight of ash removed from the coal. Although filters do perform the required separation to meet product specifications, they are expensive to install and operate. Alternate methods of solids separation are being investigated. Short-term tests showed that hydroclones operating at a pressure drop of 40-90 psi removed about 75% of the ash from the dissolved coal stream. The product from this stream contains about 3% ash and 1% sulfur compared to the present SRC specifications of 0.16% maximum ash and 0.96% maximum sulfur. A disc-nozzle centrifuge and a solid bowl centrifuge have been installed. The two centrifuges will be tested singly and in combination to determine the potential of centrifugation for replacing filtration in the mineral separation step.

V. Mechanical Performance

A number of problems have arisen, mainly in connection with pumps and valves, but key items of process equipment have performed reasonably well. In view of the developmental nature of the plant, the problems encountered are not judged to be extraordinary either in kind or complexity, and good progress has been made in developing solutions which will be useful in designing future systems. The mechanical performance of some specific items of equipment is discussed below.

A. Pumps

Centrifugal pumps handling process slurries have generally given satisfactory performance except for casing wear. The use of replaceable wear plates and hard coatings has been necessary. Double mechanical seals have proved satisfactory when they are properly supplied with flushing oil. Single mechanical seals did not give good service.

Check valve seats made of tungsten carbide have given excellent performance in the positive displacement pumps used in high pressure slurry service. The present seats have been in service for more than 22 months.

The original gear pumps which handled the molten SRC from the vacuum column failed in less than one month. They were replaced with centrifugal pumps with double mechanical seals. These centrifugal pumps have given satisfactory performance.

B. Valves

Letdown valves with tungsten carbide trim and liners have given 3 to 4 months of service. Most of the letdown valve failures have been due to mechanical damage from hard foreign objects which were caught in the valve trim. Tungsten carbide trim has given satisfactory service in valves handling process slurries. No other satisfactory trim materials have been found.

C. Filter

The pressure leaf filter has generally performed reliably. Most of the problems encountered in the filtration area have been due to problems with valves associated with the filter rather than with the filter.

The original 120 x 330 mesh filter screens gave a maximum service life of 150 filter cycles. These screens were replaced with 60 x 60 mesh, 316 SS screens made of larger diameter wire. The larger mesh screens have been used for about 1500 filter cycles.

D. Mineral Residue Dryer

A jacketed rotary dryer was used to recover solvent from the filter cake under vacuum. Many problems were encountered with the dryer. The filter cake became very sticky during the drying cycle and the accumulation of solids on the dryer body and the agitator caused the dryer to stall. The solvent is now recovered from the filter cake by pulling a vacuum on the filter body.

E. Slurry Preheater

The slurry preheater has given good performance. There have been few problems with coking in the preheater, and most of these occurred during

operations at unusual conditions. In each case the coil was cleaned by a normal steam-air decoking procedure.

Periodic ultrasonic thickness measurements have indicated that there has been very little erosion in the 316 SS preheater coil. These measurements indicated that the wall thickness had decreased only 10-20 mils in over 7800 hours of operation with coal.

F. Flow Rate Measurement

Venturi meters and orifice flanges are used for slurry flow rate measurements. The major problems with these instruments have been erosion of the orifices and venturi throats and plugging of the pressure taps.

Table 1. Typical Analyses of Coals Used

Source					
State	Kentucky	Illinois	West Virginia	Illinois	Wyoming
Seam	No. 9 and 14	No. 6	Pittsburgh	No. 6	Roland-Smith
Mine	Colonial	Burning Star No. 2	Loveridge	Monterey	Belle Ayr
Company	Pittsburg & Midway Coal Mining Co.	Consolidation Coal Co.	Consolidation Coal Co.	Monterey Coal Company	Amax Coal Co.
Proximate Analysis (dry basis), wt%					
Volatile Matter	38.03	38.27	39.98	44.04	54.16
Fixed Carbon	53.04	50.93	51.74	45.81	38.07
Ash	8.93	10.80	8.28	10.15	7.77
Moisture	4.21	5.80	6.50	6.22	6.36
Heating Value, Btu/lb	13,060	12,602	13,856	12,215	11,307
Ultimate Analysis, wt%					
Carbon	72.94	70.94	75.71	64.80	67.95
Hydrogen	5.27	5.06	5.12	5.39	4.94
Nitrogen	1.14	1.25	1.32	1.20	1.06
Chlorine	0.13	0.09	0.11	0.07	0.01
Sulfur	3.07	3.23	2.63	4.25	0.62
Ash	8.93	10.80	8.28	10.15	7.77
Oxygen	8.52	8.63	6.83	14.14	17.65
Sulfur Forms, wt%					
Pyritic	1.12	1.03	0.89	1.14	0.06
Sulfate	0.10	0.15	0.04	0.19	0.03
Organic	1.85	2.05	1.70	2.92	0.53
Iron in Ash, wt%	17.36	10.25	8.61	11.86	7.28

Table 2. Typical Operating Conditions and Results

Feed Coal					
State	Kentucky	Illinois	West Virginia	Illinois	Wyoming
Seam	9 and 14	No. 6	Pittsburgh	No. 6	Roland-Smith
Mine	Colonial	Burning Star No. 2	Loveridge	Monterey	Belle Ayr
Conditions					
Coal Feed, MF, lb/hr	500	460	500	500	370
Coal in Feed Slurry, wt%	24	25	25	25	20
Gas Feed Rate, scfh	5600	5000	5000	10,000	10,000
Hydrogen in Feed Gas, vol%	86	88	87	86	86
Pressure, psig	1700	1800	1700	2400	2400
Temperature, ^{o}F	830	820	855	855	825
Results					
Conversion, % of MAF Coal	92	90	91	95	88
Hydrogen Consumed, % of MAF Coal	2.2	2.3	2.2	3.1	3.2
Yields, % of MAF Coal					
Solvent Refined Coal	66	66	74	52	50
Organic Liquids					
Boiling below 350^{o}F	8	3	1	4	3
Boiling 350-800^{o}F	6	12	7	24	17
Gas	11	11	11	13	14
Water	1	1	3	4	8

Table 3. Typical Analyses of Solvent Refined Coal

Feed Coal

State	Kentucky	Illinois	West Virginia	Illinois	Wyoming
Seam	9 and 14	No. 6	Pittsburgh	No. 6	Roland-Smith
Mine	Colonial	Burning Star No. 2	Loveridge	Monterey	Belle Ayr
Proximate Analysis (dry basis), wt%					
Volatile Matter	70.6	68.5	56.6	72.1	51.7
Fixed Carbon	29.3	31.4	43.3	27.8	48.2
Ash	0.1	0.1	0.1	0.1	0.1
Heating Value, Btu/lb	15,700	15,500	15,900	15,800	15,900
Ultimate Analysis, wt%					
Carbon	87.2	88.5	87.9	87.1	88.4
Hydrogen	5.6	5.3	5.7	6.0	5.6
Nitrogen	1.8	1.9	1.6	1.9	1.7
Chlorine	Trace	Trace	Trace	--	--
Sulfur	0.8	0.8	0.9	0.8	0.1
Ash	0.1	0.1	0.1	0.1	0.1
Oxygen	4.5	3.4	3.8	4.1	4.1
Sulfur Forms, wt%					
Pyritic	0.05	0.03	0.01	0.07	0.01
Sulfate	0.00	0.06	0.02	0.01	0.00
Organic	0.75	0.68	0.86	0.75	0.11
Melting Point, ^{o}F	360	380	340	300	335

SOLVENT REFINED COAL PROCESS

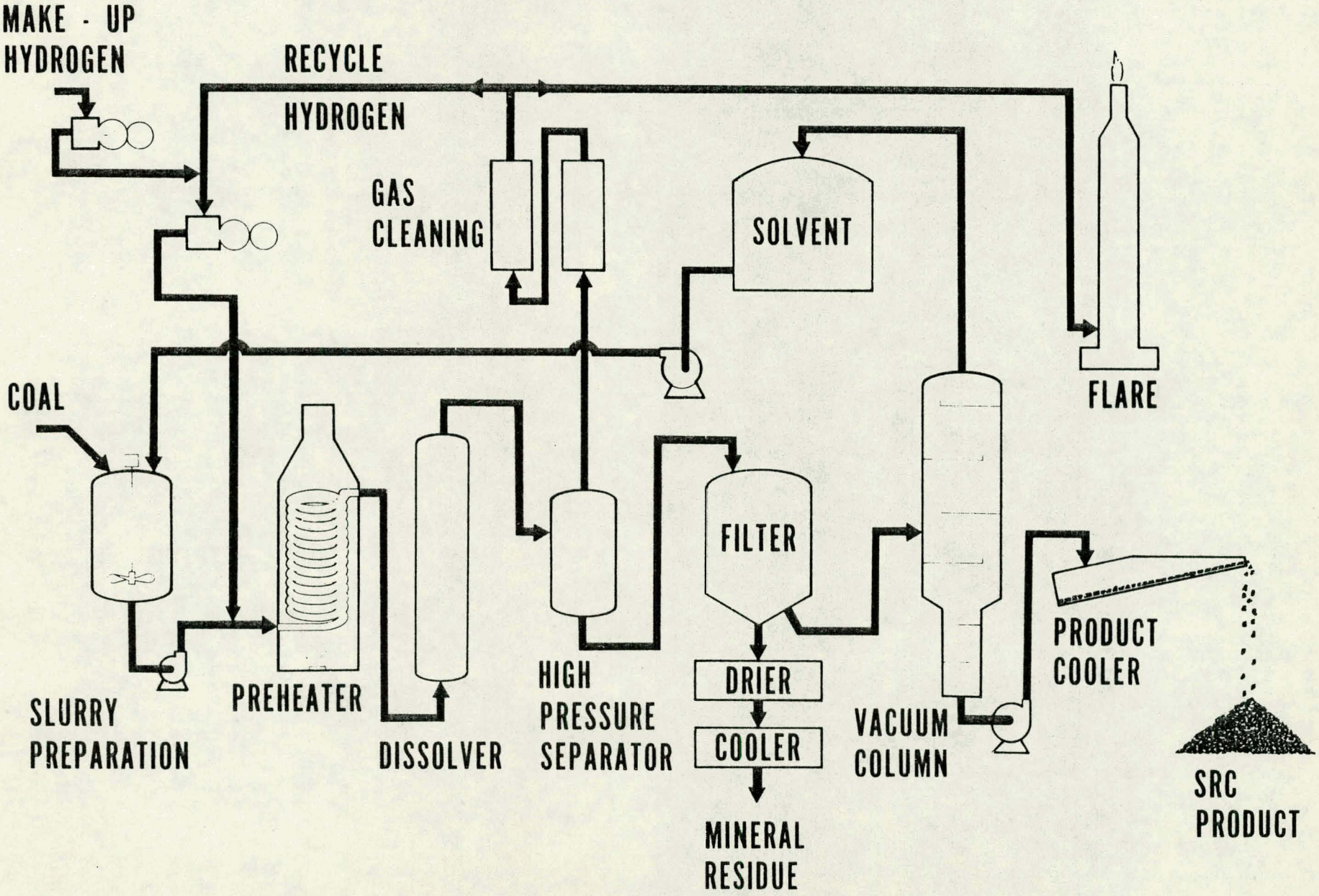


Figure 1

SOLVENT REFINED COAL PROCESS

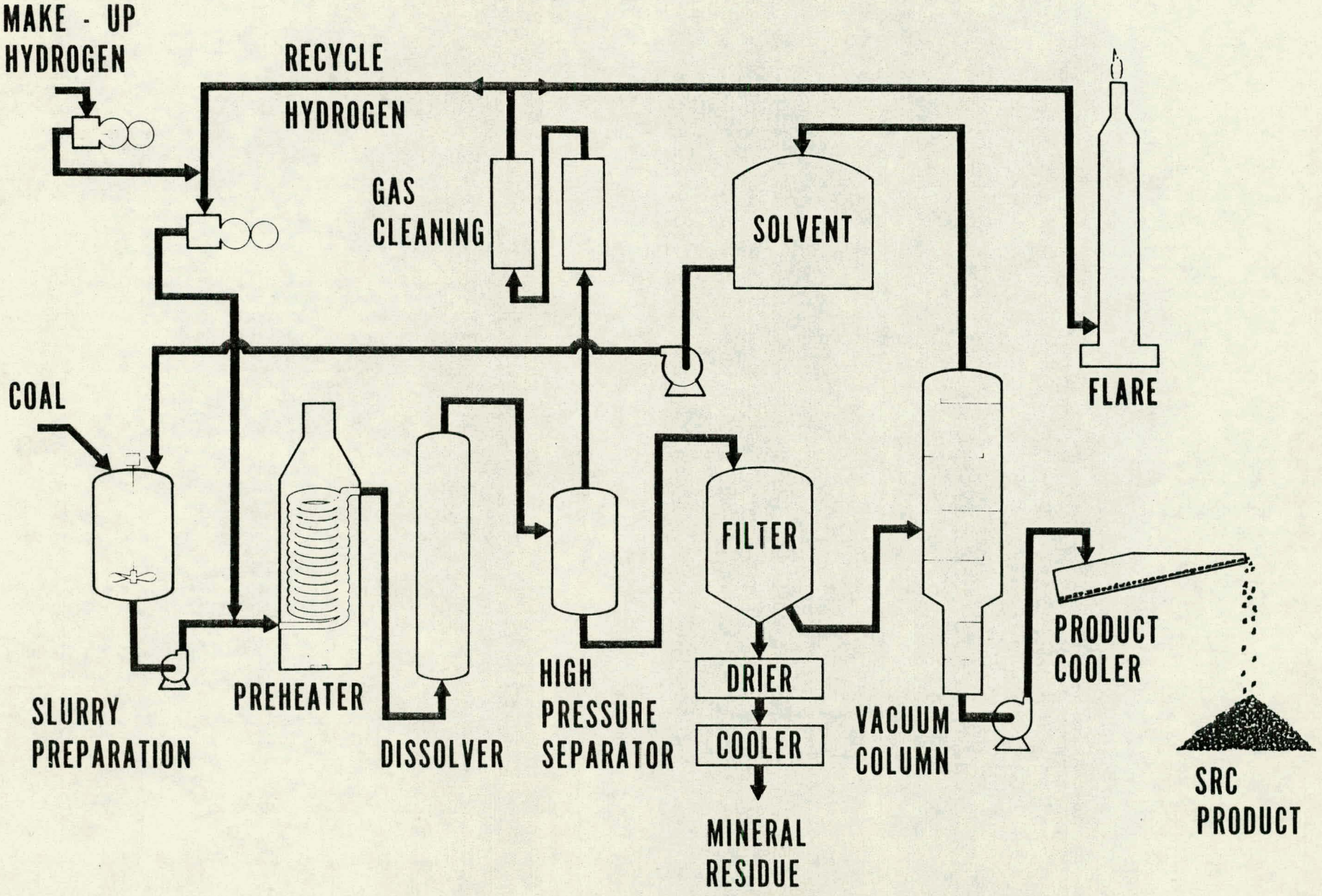


Figure 1

THE SHELL FLUE GAS DESULFURIZATION PROCESS

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INTRODUCTION

The Shell Flue Gas Desulfurization (SFGD) process has been discussed in detail in the published litersture.(2,3,4,5,6,7,8,9) A summary description of the process follows.

This is a dry process with two or more reactors in swing operation for acceptance of SO_2 and regeneration of the acceptor. Each reactor, containing a static acceptor bed, is operated throughout the cycle at essentially the same temperature -- 400°C; heating or cooling of the acceptor is therefore unnecessary. Contact between the flue gas and the acceptor is established in such a way that reactor pressure drop is not affected by fly ash. This is achieved by the application of the parallel-passage design which causes the gas to flow alongside the acceptor mass rather than through it. When the acceptor becomes loaded with SO_2 to such an extent that the limiting removal efficiency is reached, the flue gas flow is switched to a regenerated reactor and the spent acceptor is regenerated with fuel gas. The regeneration off-gas contains sulfur dioxide, water vapor and traces of unreacted fuel gas. Before and after switching from acceptance to regeneration and back to acceptance, the reactor is purged to separate the oxidizing (flue gas) and reducing (regeneration) atmospheres.

The reaction sequence is as follows:

Oxidation

$$Cu + \tfrac{1}{2} O_2 \longrightarrow CuO \quad \Delta H_{400^\circ C} = -37{,}300 \text{ Kcal/Kg-mole Cu}$$

Acceptance

$$CuO + \tfrac{1}{2} O_2 + SO_2 \longrightarrow CuSO_4 \quad \Delta H_{400^\circ C} = -74{,}600 \text{ Kcal/Kg-mole Cu}$$

Regeneration

$$CuSO_4 + 2H_2 \longrightarrow Cu + SO_2 + 2H_2O \quad \Delta H_{400^\circ C} = -5{,}600 \text{ Kcal/Kg-mole Cu}$$

$$CuO + H_2 \longrightarrow Cu + H_2O \quad \Delta H_{400^\circ C} = -20{,}900 \text{ Kcal/Kg-mole Cu}$$

PROCESS CHARACTERISTICS

The acceptor material is copper on a special alumina support which provides physical and chemical stability. In each of the chemical steps of the cycle experienced by the copper -- oxidation, sulfation and reduction -- heat is released and absorbed by the processing gas. The amount is equivalent to the heat of combustion of the fuel gas used in the regeneration. By locating the SFGD unit upstream of the air-preheater the heat of reaction absorbed by the flue gas is totally recoverable and accordingly reduces the fuel required in the boiler. The copper-on-alumina acceptor also removes SO_3 with the result that the lowered sulfuric acid dew point of the treated gas permits stack temperatures equivalent to those for flue gas from low sulfur fuels.

A detailed analysis of the energy requirements of the SFGD process has been reported. (9) Dry flue gas desulfurization (FGD) processes as a class have a basic potential advantage over wet processes in permitting additional heat recovery from the flue gas and not requiring re-heat energy. The electrical energy requirement for a SFGD unit, expressed as a fraction of the thermal energy released in the boiler, is less than 0.5%. The net sum of electrical, steam and fuel requirements, expressed in the same manner, is approximately equal numerically to the weight-percent sulfur in the fuel. For example, the net requirements to effect 85% desulfurizstion on the flue gas from a 150 MW boiler fired with coal containing 3.9 wt-% sulfur is 3.6% of the higher heating value of the fuel. It is interesting to note that half of this quantity of energy is released with the oxidation of the coal sulfur, i.e., pyrites to iron oxide and sulfur oxides.

The kinetics of the reactions which occur around the cycle also provide the process with several interesting characteristics. The oxidation of the copper to copper oxide occurs very rapidly and the oxidation wave passes through the reactor in a few minutes. The acceptance reaction is first order with respect to sulfur dioxide concentration so that percent desulfurization is substantially independent of inlet SO_2 concentration and, therefore, not related to SO_2 partial pressure as in wet scrubbing. Acceptance reaction rate is also high; acceptance is accomplished with a gas residence time in the reactor of less than half a second. The reactor internals are so designed that the temperature rise resulting from the oxidation and acceptance reactions will not be additive at oxygen concentrations in the flue gas greater than 0.5%. The copper sulfate at 400°C is an excellent scavenger for hydrogen and, therefore, affords a means for regeneration with only small quantities of reductant slipping through the reactor and appearing in the regeneration off-gas. Another characteristic of the regeneration chemistry occurs at the beginning of the regeneration when hydrogen is introduced to the purged reactor. As the copper sulfate is reacted at the inlet to the bed, the released SO_2 is re-accepted by unreacted cupric oxide. All the copper oxide is reduced before SO_2 breaks through and suddenly appears in the off-gas. The SO_2 production rate change is equally abrupt at the end of regeneration.

The acceptor material possesses catalytic properties which permit the reduction of nitrous oxide with ammonia to be selective in the presence of

oxygen and sulfur oxides, yielding nitrogen and water vapor. The system can be designed to operate with almost any combination of NOx-SOx reduction required. The ammonia injection equipment can be added to an existing SFGD unit at any time that NOx control becomes necessary.

The acceptor material is relatively inert to chlorine/hydrogen chloride so that flue gas from coal with high chloride content can be successfully processed.

Because the water content of the flue gas is unchanged through the SFGD reactor and because the desulfurization performance is insensitive to sulfur-oxide concentration, a full line-size flue gas by-pass system around the SFGD unit can be used which serves to isolate it from the boiler. With this by-pass the turndown of the system is from rated-load to zero, recycling or by-passing flue gas as the case may be irrespective of boiler output.

The control system on the SFGD reactors is a digital sequence unit which, with the appropriate SO_2 analyzers, can provide completely automatic control of the process cycle, adjusting regeneration frequency as required to provide a fixed sulfur removal independent of boiler load.

REGENERATION OFF-GAS -- PROCESS OPTIONS

The workup of the SO_2-rich stream may take various forms depending upon the pass-through of inerts in the reducing gas and the desired form of the sulfur product. Since the nature of the SFGD process is cyclic, the production of off-gas is also cyclic and these flow fluctuations must be dampened before subsequent processing. Further, the regeneration off-gas contains water vapor, the bulk of which should be condensed. Thus, the usual design includes a waste heat boiler to recover sensible heat and a direct contact cooler to remove heat at temperatures below dew point. The simplest system for dampening the flow is a gas-compressor/gas-holder from which the concentrated SO_2 is released on flow control.

Since sulfur is used in the oxidized form of sulfuric acid, it is expedient to produce it in oxidized form, e.g., as a concentrated stream of SO_2 to the effluent stream from a waste acid burner in an existing sulfur plant to the feed preparation section of a sulfuric acid plant, or as anhydrous liquid sulfur dioxide. Processing to elemental sulfur through an existing or new Claus unit is also an available option. The route chosen in any application will depend upon specific economics; in the event that elemental sulfur is chosen as the desired form for the sulfur product, it should be remembered in the economic analysis that fuel is required to reduce sulfur dioxide, and the thermal value of this fuel is released when the sulfur is subsequently re-oxidized for sulfuric acid manufacture.

OPERATING EXPERIENCE

The reduction to practice has been demonstrated with two pilot plants and one full-scale unit.

Pernis Pilot Plant (3)

This unit was operated for a 4-year period and through more than 20,000 cycles of operation with several experimental and commercial acceptors. The flue gas processed was a slip stream from an oil-fired refinery process heater. This plant was designed so that scaling-up to a full-scale unit would be possible in one step. The nature of the process permits such a procedure, since any size reactor can be based on a unit cell of internals, the number of which is determined by the required capacity. An adiabatic reactor was applied with a geometry representative of a large-scale design.

Tampa Electric Company (TECO) Pilot Plant (7)

This plant has been operated for 3 years on a slip stream from the flue gas from either No. 1 or No. 2 boiler at TECO's Big Bend Station which fires Western Kentucky coal (3.5% sulfur). Further the flue gas may be taken from the main ducts either upstream or downstream of the cold electrostatic precipitators. The design uses a single cell-stack of commercial reactor internals. The reactor is adiabatically insulated.

Showa Yokkaichi Sekiyu (SYS) 40 MW Unit (6)

The SYS unit went on stream August, 1973, processing the flue gas from a refinery boiler fired with residual fuel oils. This is a 2-reactor system, each reactor of which can accept the full output of the boiler (130,000 NM^3/hr). The reactor internals are identical with those in the Tampa plant. The captured SO_2 is released through a feed preparation section into an existing Claus unit and converted to elemental sulfur. This plant has been retrofitted for simultaneous NOx and SOx reduction.

Based on the experience gained from these three units an assessment of the process and mechanical design includes the following:

- The results obtained with computer programs developed to assist with design and optimization agree well with commercial performance.
- Engineering scale-up consists of paralleling standardized unit cells -- Tampa has one cell stack in horizontal cross-section, whereas the reactors at SYS each have 68 identical stacks in cross-section.
- The commercial acceptor has excellent physical and chemical stability.
- The unit cells are easy to handle and easy to load into the reactors. Resistance to corrosion and erosion has been established.
- The automatic sequence controller operates and safeguards the unit by

setting the length of the acceptance and regeneration cycles and by continuously surveying all valve positions, flow rates and temperatures.

- The required attention of operators, after start-up and initial run operation, is almost nil. No extra manpower is required at SYS.

- The special valves used to isolate a reactor during regeneration have operated satisfactorily in oil and coal service, with high and low fly ash loadings.

- The "open by-pass" functions as conceived which is to isolate the operation of the SFGD and boiler sections from each other.

The only real technical problem to emerge from the operating experience is the application of the parallel-passage principle to the reactor design so that fly ash will cause no interference to gas flow or SO_2 acceptance. Since fouling of the reactor internals is a cumulative phenomenon, a successful application of the principle requires that 100% of the fly ash will pass through the reactor 100% of the time.

In the Pernis pilot plant, the system worked almost perfectly -- there was no discernible decline in performance during runs of more than 3000 cycles. At SYS the nature of the fly ash (high vanadium and sodium) resulted in gradual fouling of internals, limiting runs to one or two months. At Tampa, the fouling rate was very high resulting in a noticeable fall-off in performance in just a few days.

At both SYS and Tampa, initial attempts to cope with ash deposition were directed toward adjusting the characteristics of the operation to permit the parallel-passage design to function as at Pernis, that is, to make the system work with total pass through all of the time. When it became apparent that such a course of action would require customizing the conditions for every potential application, it was decided to alter the strategy, to accept the fact that perfect performance is not realistically attainable, and to provide for continuous or intermittent cleaning of fly ash from reactor internals in situ and during normal operation.

Several such procedures have been developed and tested at Tampa and SYS. Using one such procedure, requiring no physical alterations to the pilot unit, stable performance has been demonstrated at Tampa on flue gas with high loadings of fly ash. Operating on flue gas taken upstream of the electrostatic precipitator, with ash loadings of 10 gr/SCF (higher during soot blowing), pressure drop across the reactor and desulfurization performance were unaffected -- pressure drop of 5-6 in. water and 90% SO_2 removal on an acceptor with over 13,000 redox cycles.

Facilities for in situ cleaning of the reactor internals are to be incorporated in the SYS reactors.

ECONOMICS

The classic study of FGD economics by the engineers in the chemical development division at TVA has been reported in detail; (1) any attempt at comparative economics of the SFGD process is beyond the scope of this paper. However, even the TVA study is applicable in only a limited number of situations, the results being very sensitive to individual and local conditions. Hence there is no shortcut to the usual procedures used in process selection employing case studies.

One set of economics as applied to an industrial coal-fired boiler is presented in Table 1. This study is not necessarily typical, but it is real.

###

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Table 1

JANUARY 1976 COST PROJECTIONS

FLUE GAS DESULFURIZATION PROCESS

Basis: 70 MW Coal-fired Industrial Boiler
12,000 Btu/lb, 4% Sulfur Coal
82% Desulfurization, 8,000 hr/yr
Hydrogen Plant is Available
Elemental Sulfur Produced

Installed Capital Cost	
Acceptor Section	$5,280,000
Workup Section	2,070,000
Modified Claus Section	930,000
Total Capital Cost	$8,280,000

Annual Operating Cost	$1,000/yr
Electricity, $0.015/kwhr	105
Steam, $1.50/1000 lb	75
Hydrogen, H_2/CO =3/1, $1.25/1000 SCF	1023
Labor, $10/hr	20
Maintenance, 2% on capital	165
Acceptor Replacement, $3.25/lb	177
Capital Charges, 20% on capital	1656
Fuel Credit, $1.50/MMBtu	(247)
Sulfur Credit, $45/ton	(313)
Total Operating Cost	2661

FLUE GAS DESULFURIZATION BY THE THIOSORBIC® PROCESS

by

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Vice President, Research
Dravo Lime Company

Presented at the
Third Annual International Conference

on

COAL GASIFICATION AND LIQUEFACTION: WHAT NEEDS TO BE DONE NOW!

University of Pittsburgh
School of Engineering
August 3-5, 1976

Rec'd 5-11-77

EY-77-X-05-0840

FLUE GAS DESULFURIZATION BY THE THIOSORBIC®* PROCESS

INTRODUCTION

The facts that calcium-based alkalis (i.e., lime and limestone) are the lowest cost basic compounds and that their reaction products with the combustion products of the various sulfur compounds contained in the fossil fuels are solids, led to the earliest utilization of these bases for flue gas desulfurization. The reactions in the scrubber are not limited to interaction between SO_2-water and the calcium-based alkali, however, due to the presence of other substances in the flue gas, the base and the scrubbing liquor. A rather complex series of reactions are, therefore, taking place governed by the kinetics of these reactions and the even more complex equilibrium relationships. The equilibrium between two or more species is much more extensively studied and understood than the kinetics of the respective reactions.

Two of the most important solid species, calcium carbonate ($CaCO_3$) and calcium sulfite ($CaSO_3.½H_2O$) have rather low solubilities as compared to others (for example, calcium sulfate, gypsum ($CaSO_4.2H_2O$), calcium chloride ($CaCl_2$) or lime ($Ca(OH)_2$). This has several important ramifications:

First, lime slurry as a scrubbing agent absorbs not only the oxides of sulfur but the less acidic carbon dioxide, with the consequent problem of calcium carbonate scaling. For this reason, in general, lime is used in FGD processes indirectly to neutralize the acidified exhausted scrubbing liquor. The other significance arising from the low solubility of calcium sulfite and carbonate is the low availability in the scrubbing liquor of weakly basic soluble species such as sulfite, bicarbonate and carbonate. Therein lies directly or indirectly almost all of the problems experienced with conventional calcium-based FGD processes, namely, limited SO_2 removal efficiency, high energy consumption, scaling of the scrubbers and water balance problems. In order to overcome these problems and utilize the lime as a low cost base for FGD, the Thiosorbic process was developed.

DESCRIPTION OF THE THIOSORBIC PROCESS

The heart of the process is based on the fact that the soluble alkaline species ($SO_3^{=}$, HCO_3^{-}, $CO_3^{=}$) form much more soluble salts with magnesium than with calcium. Magnesium salts, therefore, accumulate in the scrubbing liquor according to the supply and blowdown relationship in the system. This allows higher concentrations of sulfite ions, which, according to the solubility product principle, reduces the concentration of calcium ions in the solution. Magnesium becomes by far the most abundant cation and the scrubbing reaction can be schematized as follows:

$$SO_2 + H_2O + MgSO_3 \longrightarrow Mg(HSO_3)_2$$

The exhausted scrubbing solution is regenerated in the reaction or mix tank with hydrated lime containing a small amount of magnesium hydroxide:

$$m\ Mg(HSO_3)_2 + m\ Ca(OH)_2 \longrightarrow m\ MgSO_3 + m\ CaSO_3.½H_2O + 1½mH_2O$$

$$n\ Mg(HSO_3)_2 + n\ Mg(OH)_2 \longrightarrow 2n\ MgSO_3 + 2nH_2O$$

where m + n = 1 and m and n are, respectively, the calcium and magnesium mole fractions both in the lime and in the waste product at steady state condition.

In a nutshell, each mole of SO_2 converts one mole of magnesium sulfite to bisulfite in the scrubber, which is, in turn, regenerated in the mix tank to soluble magnesium sulfite and precipitated calcium sulfite.

A continuous blowdown of the scrubbing slurry is clarified in a clarifier thickener. The thickener underflow is taken to sludge disposal and the clear overflow is returned to the process, usually as wash water to the mist eliminators.

The Thiosorbic scrubbing liquor is capable of carrying 10-15 times as much dissolved alkalinity – or reacting basic species – in solution than the conventional calcium-based process liquors. This represents several important advantages, some of which have been discussed in greater detail in previous papers:

1) Lower liquid flow rate or liquid-to-gas ratios can be used. This results in lower pipeline and pumping costs.

2) The liquor is more reactive, reducing the liquid phase resistance, resulting in a combination of higher SO_2 removal efficiency and/or lower mechanical energy for the generation of inter-

* Patented; registered Trade Mark of Dravo Corporation

faces such as gas or nozzle pressure drop in the form of fan and pump horsepower, respectively.

3) The reactions are limited to their respective vessels, each designed for its purpose (such as: in the scrubber vessel only gas-liquid reactions are taking place, and liquid-solid reactions are limited to the reaction or mix tank).

4) Both the fresh and exhausted scrubber liquors are subsaturated for gypsum, which is the nemesis of conventional scrubbing processes, and the system is free from scale.

5) The thickener overflow is also subsaturated for solids, as opposed to conventional calcium-based processes, and can be used for washing the mist eliminator free of sludge solids collected from the scrubbing slurry. This not only relieves the demisters from scaling, but also allows the operation of the scrubbing system in a more tightly closed-loop condition.

6) The scrubbing liquor is well buffered, allows the operation within narrower pH limits, which reduces corrosion considerations on the low pH end; and, most importantly, results in a lower amount of CO_2 absorption, hence, lower lime consumption, better economy.

7) SO_2 removal efficiency is nearly independent of fuel sulfur content and can be controlled with the process chemistry.

OPERATING EXPERIENCE

Earlier papers[1,2] on this subject gave the results of several pilot plant experiments with various scrubbing apparatus, where the scrubbers operating in the Thiosorbic mode produced 2 to 2.7 times as many equivalent transfer units as compared to conventional calcium-based systems. In these comparisons, the equivalent transfer units were expressed as

$$Nt = \ln \frac{pSO_2 \text{ in}}{pSO_2 \text{ out}}$$

The comparisons were made under identical hydrodynamic conditions (i.e., same gas and liquid flow, pressure drop, etc.). The highly reactive liquor reduced the liquid phase resistance particularly when the gas SO_2 contents were high. Gas phase resistance became the overwhelming factor when SO_2 contents dropped below about 100-200 ppm which are below the SO_2 emission values in most local situations. In practical terms, when removal of 90% of the SO_2 was required and achieved in the Thiosorbic mode, only 55-68% was obtained with limestone or conventional limes. Coupled with the higher removal efficiency, the reagent consumption was considerably reduced in the Thiosorbic process. The removal efficiency remained nearly independent of the inlet SO_2 concentration under the same hydrodynamic conditions, but could be varied by controlling the process chemistry.

In the following are several examples with the process which have not been reported elsewhere by this author:

I. Duquesne Light Company - Phillips Station

During the months of October, November and December of 1975, a 2½-month test was conducted in two of the 450,000 CFM single-stage venturi scrubbers with Thiosorbic lime. Due to the loss of magnesium values caused by high blowdown rates experienced during these tests, the MgO content of the lime had to be increased. A quotation from the report of R. Gordon Knight and Steve L. Pernick, Jr., to the 3rd EPA Flue Gas Desulfurication Symposium sums up the results:

"It was determined that the required degree of SO_2 removal - approximately 83% - can be obtained with an MgO content of 8-10% in the lime with single stage scrubbing. This will eliminate the necessity to install any additional stages or scrubber vessels at both the Phillips and Elrama Stations. This will save approximately $10 million in capital costs.

"The two and a half month test with modified lime resulted in 1,612 service hours on one train and 1,309 on another. After this length of service, the throat damper pressure drop in both trains was still 6 inches, the same as with newly cleaned dampers. With the normal high calcium lime, the pressure drop would have been at or approaching 12 inches, and an outage for cleaning would have been necessary. Inspection showed negligible deposit accumulation in all accessible locations, and the deposit on some previously scaled surfaces had been removed. For example, the maximum flow that could be obtained through the bleed line on one scrubber prior to the test was 420 gpm. This flow increased during the test to a value of 720 gpm at termination. On the basis of SO_2 removal efficiency and freedom from deposits, it is unfortunate that we cannot continue the use of the modified lime . . ."

II. TVA - Shawnee Station

As part of the EPA experimental flue gas desulfuri-

zation program at the Shawnee Station, a venturi-spray tower combination has been operated with lime for more than two years. During a period in October-November of 1974, magnesium oxide was added to the scrubbing slurry to produce a nominal 3,000 ppm magnesium concentration. A summary diagram from the EPA report is enclosed[3]. For the sake of comparison, a diagram of one of the conventional lime scrubbing tests is also shown.

The major differences were as follows:

1) SO_2 removal efficiency ranged between 96 and 99% for the Thiosorbic mode vs. 70 to 98% for the conventional mode.

2) SO_2 removal efficiency was independent of the SO_2 inlet concentration for the Thiosorbic mode and varied inversely in the other mode.

3) Alkali feed as a percentage of stoichiometric amount varied between 93 and 105% (average 102%) for the magnesium modified case and between 107 and 132% (average 115%) for the conventional lime case.

4) Inlet/outlet pH values of 7.0/5.85 gave a pH differential of 1.15 unit for the Thiosorbic mode, and for the other mode, 7.8/5.3 resulted in a pH differential of 2.5.

5) A portion of SO_2 absorbed was oxidized to sulfate and that was about the same for both runs (15-28% for the Thiosorbic, and 12-28% for the conventional), yet the Thiosorbic process liquor ran at 45% of gypsum saturation value whereas the conventional averaged 125% saturation, or 25% supersaturation for gypsum.

At the conclusion of the test the scrubber was inspected and the following comments reported:

"Old scale deposits remaining after the previous cleaning had slowly disappeared . . . . . Calculated average sulfate saturations of 45% of the scrubber inlet liquor and 95% for the scrubber outlet liquor confirmed that the system was operating unsaturated with respect to sulfate. The apparent descaling effect observed during the system inspection underscores the beneficial effect of unsaturated operation . . ."

III. Pennsylvania Power Company - Bruce Mansfield Unit I

Dravo Lime Company has a long-term contract to supply Thiosorbic lime to the FGD systems of Units I and II of this station. Unit I, an 880 MW capacity station, is the first unit where the Thiosorbic process has been applied on a continuous basis. Keith H. Workman, Superintendent of the Bruce Mansfield Station, is presenting a paper at this symposium, and will describe the system in greater detail.

Since not all of the scrubbers were available at the time Unit I was ready for firing, operation began at reduced load in December of 1975, with only two scrubber trains. Capacity was increased as additional scrubbers were available. The final or sixth scrubber train was on-line in June 1976. In the interim period, station load often exceeded the design capacity of the scrubbers which created several problems, primarily with the mist eliminators. In view of all of the activity going on during a start-up, it is not surprising that optimization of scrubber operation and scrubber performance tests have not been done as yet. In spite of this, however, encouraging observations may be made. Without undue attention and control of the scrubber chemistry, the scrubbers have not developed scaling problems, contrary to conventional lime and limestone systems operating on high sulfur coal where scaling of the scrubbers is responsible for high maintenance cost and reduced station availability.

PROCESS CHEMISTRY

The solubility of gases and solids and the various equilibria existing in the aqueous phase were investigated by several workers. Lowell et al.[4] reviewed the extensive literature and determined several of the missing data. This author has reported in detail the solubility of SO_2[5] and gypsum[6] in aqueous electrolytes. As indicated in the introduction, we are dealing with a rather complex equilibrium but the clarity improves without a great loss of accuracy if we concentrate on the more important equilibrium conditions only. The various constants in the following table are given at 50°C which is quite typical of FGD processes.

TABLE I

Gas-Liquid Equilibrium

		Constants
1)	$\frac{H_2SO_3}{pSO_2}$	5.6×10^{-1}
2)	$\frac{H_2CO_3}{pCO_2}$	1.97×10^{-2}

TABLE II

Dissociation Constants of Soluble Species

		Constants
3)	$H^+ \times OH^-$	5.5×10^{-14}
4)	$\frac{H^+ \times HSO_3^-}{H_2SO_3}$	7.8×10^{-3}
5)	$\frac{H^+ \times SO_3^{=}}{HSO_3^-}$	4.3×10^{-8}
6)	$\frac{H^+ \times SO_4^{=}}{HSO_4^-}$	4.8×10^{-3}
7)	$\frac{H^+ \times HCO_3^-}{H_2CO_3}$	5.2×10^{-7}
8)	$\frac{H^+ \times CO_3^{=}}{HCO_3^-}$	6.7×10^{-11}

TABLE III

Solubility Products of Solid Species

		Constants
9)	$Ca^{++} \times SO_3^{=}$	8.4×10^{-8}
10)	$Ca^{++} \times CO_3^{=}$	1.8×10^{-9}
11)	$Ca^{++} \times SO_4^{=}$	2.2×10^{-5}
12)	$Ca^{++} \times (OH^-)^2$	5.7×10^{-6}
13)	$Mg^{++} \times (OH^-)^2$	9.4×10^{-12}
14)	$Mg^{++} \times CO_3^{=}$	1.2×10^{-5}
15)	$Mg^{++} \times SO_3^{=}$	5.3×10^{-5}

TABLE IV

Complexes and Ion Pairs

		Constants
)	$\frac{Ca^{++} \times SO_3^{=}}{CaSO_3^o}$	3.0×10^{-4}
17)	$\frac{Ca^{++} \times CO_3^{=}}{CaCO_3^o}$	4.7×10^{-4}
18)	$\frac{Ca^{++} \times HCO_3^-}{CaHCO_3^+}$	4.6×10^{-2}
19)	$\frac{Ca^{++} \times SO_4^{=}}{CaSO_4^o}$	3.5×10^{-3}
20)	$\frac{Mg^{++} \times SO_3^{=}}{CaSO_3^o}$	9.3×10^{-4}
21)	$\frac{Mg^{++} \times HCO_3^-}{MgCO_3^o}$	9.6×10^{-2}
22)	$\frac{Mg^{++} \times SO_4^{=}}{MgSO_4^o}$	3.0×10^{-3}
23)	$\frac{Mg^{++} \times CO_3^{=}}{MgCO_3^o}$	3.0×10^{-4}

Examining the modified Henry's constants of equations 1 and 2, we find that SO_2 is about 28 times more soluble than CO_2. Since, however, the CO_2 partial pressure in the flue gas is 30 to 300 times greater than that of SO_2, equal or greater amounts of molecular CO_2 will be absorbable in water than SO_2. Comparing the first dissociation constants, however, 7.8×10^{-3} and 5.2×10^{-7}, respectively, for SO_2 and CO_2 we find that SO_2 is four orders of magnitude stronger acid than CO_2 which makes the selective absorption of SO_2 possible. In the pH range of 6-7, the SO_2 solubility exceeds CO_2 solubility by about three orders of magnitude.

Table IV lists several strong complexes and ion pairs for both calcium and magnesium ions with the important anions $SO_3^{=}$, $SO_4^{=}$ and $CO_3^{=}$. The bicarbonate complexes are by at least one order of magnitude less stable. The fact that magnesium pairs up with sulfate ion to an appreciable degree makes magnesium a unique cation in reducing the active sulfate ion concentration and, therefore, saturation concentration for gypsum.

The solubility product of magnesium sulfite is 631 times as large as that of calcium sulfite. This means that a given sulfite ion activity (the activity of magnesium ions) may be 631 times as high as that of the calcium ion before the system becomes saturated for both calcium and magnesium sulfite. For the carbonate ion the difference is even greater. It is not difficult to visualize, therefore, that a magnesium sulfite-containing solution will be saturated for calcium sulfite and that results in the depression of calcium ion. In turn, if we examine the solubility product for gypsum, we will see that the water will

be saturated for both gypsum and calcium sulfite when the activity of sulfate ion exceeds the activity of sulfite ion 262-fold. Therefore, it takes only a rather small amount of sulfite ion activity to reduce the calcium ion concentration to a level below what is required for gypsum saturation.

Let us take, for example, a solution with a rather low sulfite ion activity of 10^{-4} mole/liter (8 mg/l). Let us further assume, for the sake of simplicity, that in view of complexes and a typical ionic strength the concentration of the ions exceeds the free ion activities four-fold. If the system is in equilibrium with calcium sulfite, the calcium ion activity will be 8.4×10^{-4} mole (concentration: 134.4 ppm). In order for this solution to become saturated for gypsum, the sulfate ion activity has to be no less than 2.62×10^{-2} mole ($SO_4^=$ concentration: 10,057 ppm). This was a very conservative calculation for calcium sulfate saturation since considering the $CaSO_4^o$ ion pair itself should contain 250 ppm Ca ion at saturation, which is nearly twice as much as the 134.4 total calcium ion assumed. It is, therefore, sufficient to maintain a small sulfite ion concentration in all parts of the system to depress the calcium ion concentration and gypsum saturation conditions and scale forming can be avoided. The ability of the Thiosorbic process to run undersaturated for gypsum explains how in the trials old gypsum scale could be dissolved.

Sulfate ions which accumulate in the process liquor via SO_3 absorption and the oxidation of sulfite ions are purged by two mechanisms. One is the blowdown of some of the liquor with the sludge; the other is the solid solution phenomenon. We have rediscovered what the British found more than two decades earlier[7], that sulfate ions substitute for sulfite ions in the calcium sulfite crystal. We have noted[2] that the chemical analysis of the sludge solids may indicate a considerable amount of sulfate ions, yet no gypsum were found by X-ray. The absence of gypsum was further justified by the fact that during crystal formation the solution was undersaturated for gypsum. The British work suggests that gypsum saturation was reached when more than 19% of the absorbed sulfur was oxidized to sulfate. In our experience, gypsum saturation can be avoided at much higher oxidation levels. This is possible because the British experience was based on limestone scrubbing, at lower pH and with calcium as the major cation. In their case, a host of other anions besides sulfite and sulfate ions such as bisulfite, bicarbonate and chloride necessitated a high calcium ion concentration for the sake of electro-neutrality. At higher calcium concentration, lower sulfate ions will result in supersaturation for gypsum.

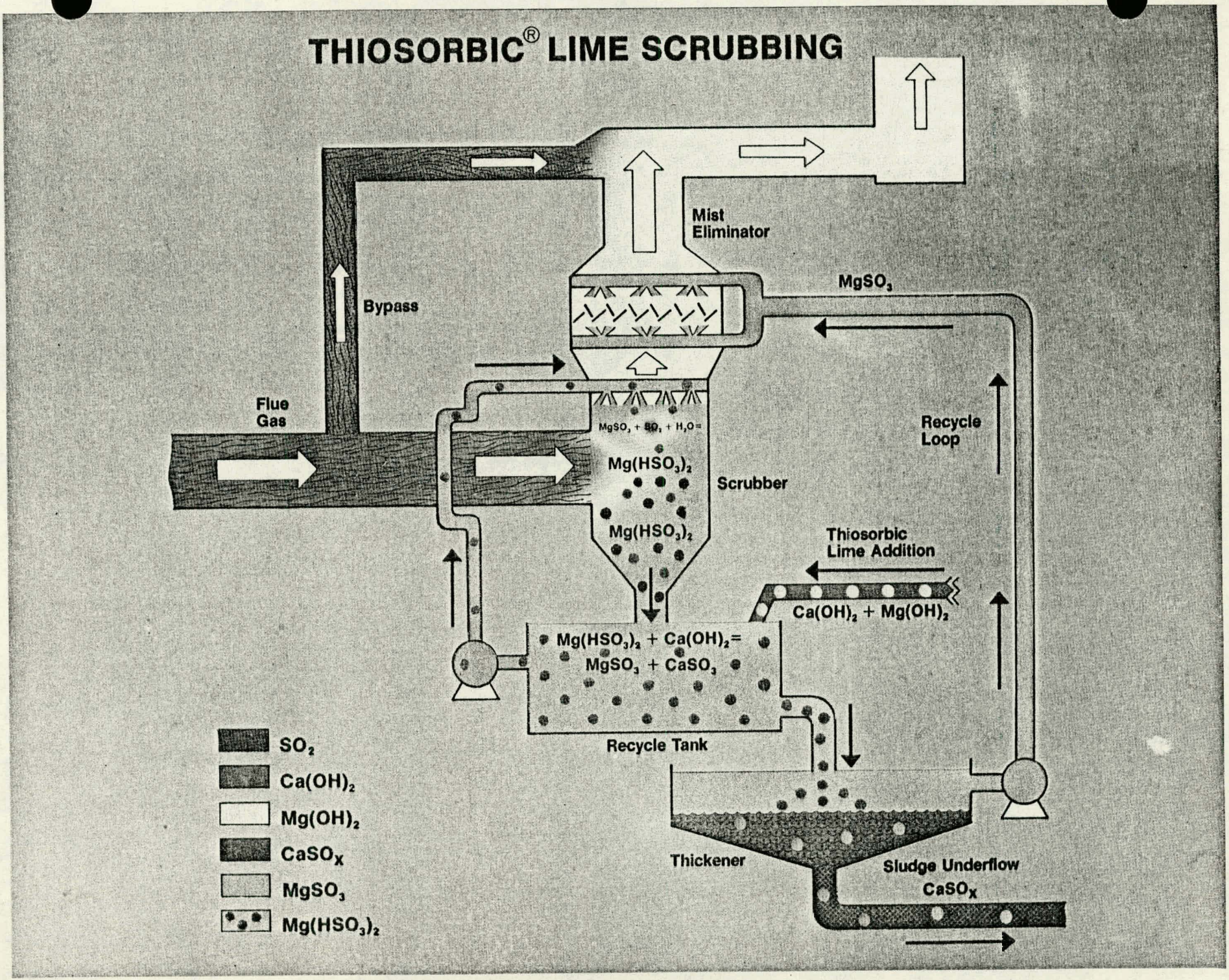
The magnesium ion contained in the sludge liquor is made insoluble by the addition of lime, most of which is available in the grits from the lime slaker:

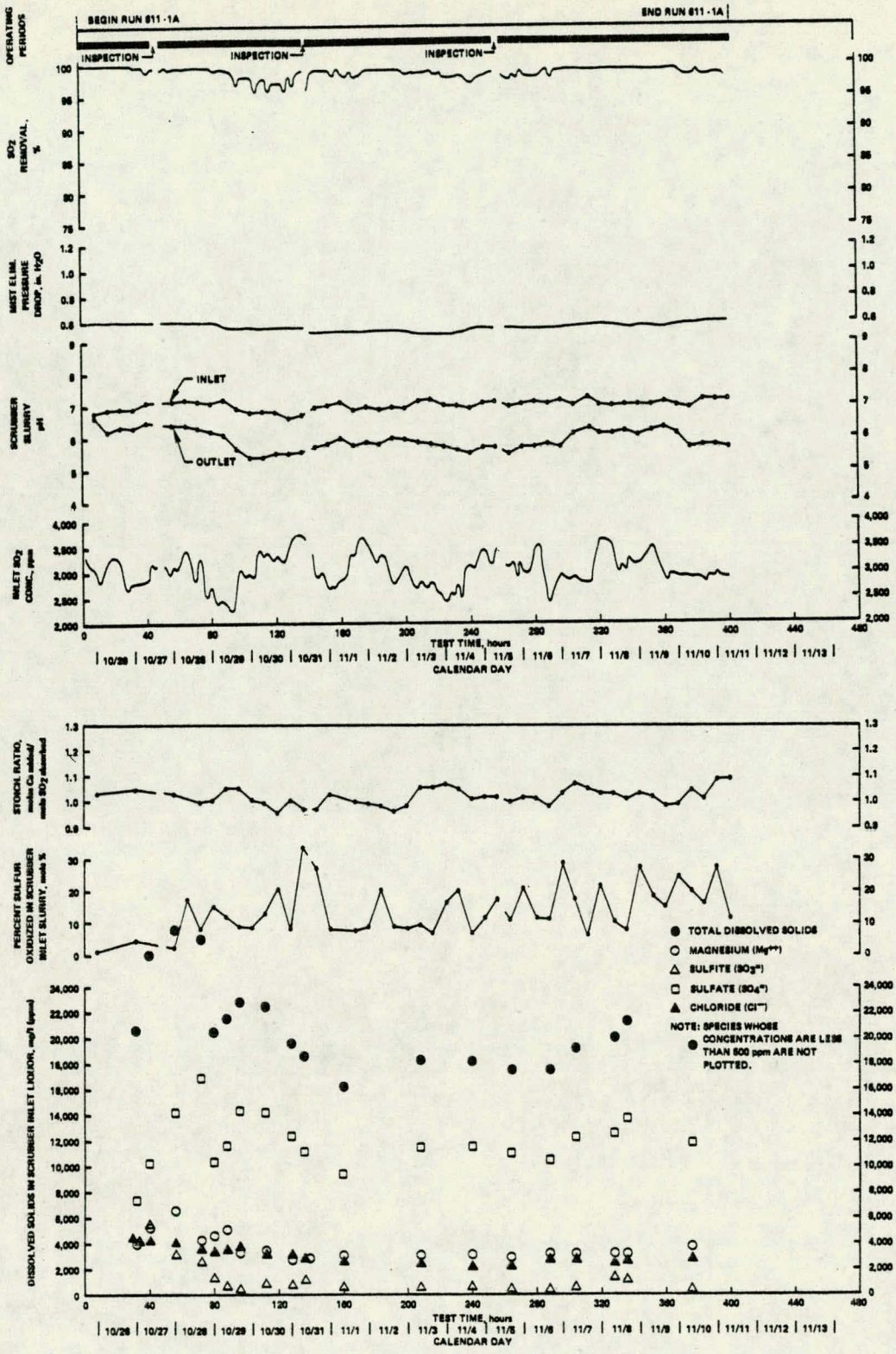
$$MgSO_3 + Ca(OH)_2 \longrightarrow Mg(OH)_2 + CaSO_3.\tfrac{1}{2}H_2O + \tfrac{1}{2}H_2O$$

$$MgSO_4 + Ca(OH)_2 + 2H_2O \longrightarrow Mg(OH)_2 + CaSO_4.2H_2O$$

Due to the low solubility of magnesium hydroxide, these reactions go to near completion at pH 10.0. The dissolved solids content of the sludge water thereby is essentially the same as that of conventional lime scrubbing processes.

THIOSORBIC® LIME SCRUBBING

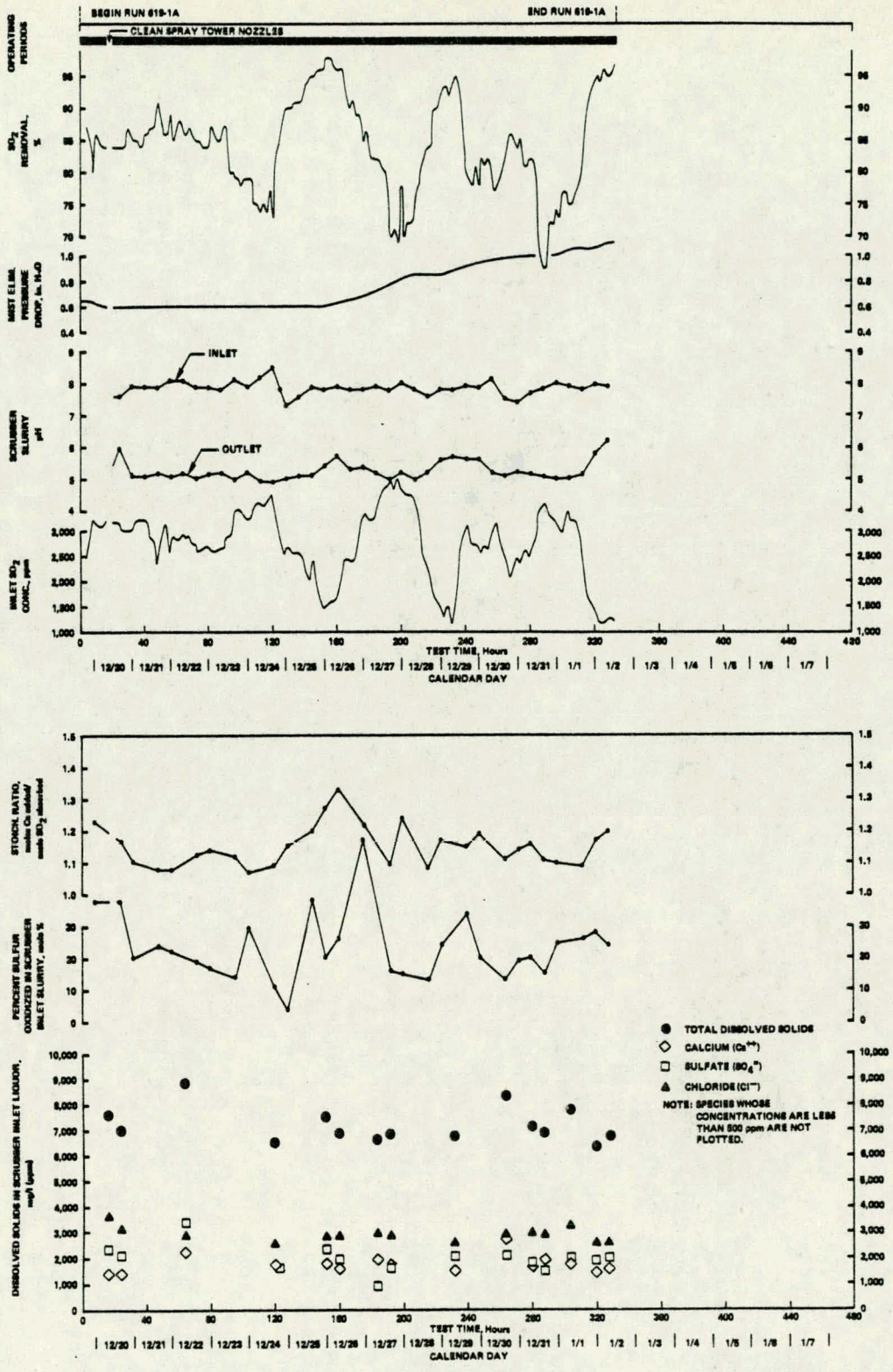




Gas Rate = 25,000 acfm @ 330° F
Liquor Rate to Venturi = 600 gpm
Liquor Rate to Spray Tower = 1200 gpm
Venturi L/G 30 gal/mcf
Spray Tower Gas Velocity = 6.7 ft/sec
No. of Spray Headers = 4
EHT (Sealed) Residence Time = 6 min

Percent Solids Recirculated = 8.0-9.3 wt %
Venturi Pressure Drop = 9 in. H_2O
Total Pressure Drop, Excluding Mist Elim. = 10.4-12.0 in H_2O
Scrubber Inlet Liquor Temperature = 127-131 ° F
Liquid Conductivity = 9,200-13,000 μ mhos/cm
Discharge (Clarifier and Filter) Solids Concentration = 44-48 wt %
Lime addition to Scrubber Downcomer

Figure D-3. Operating Data for Venturi/Spray Tower Run 611-1A



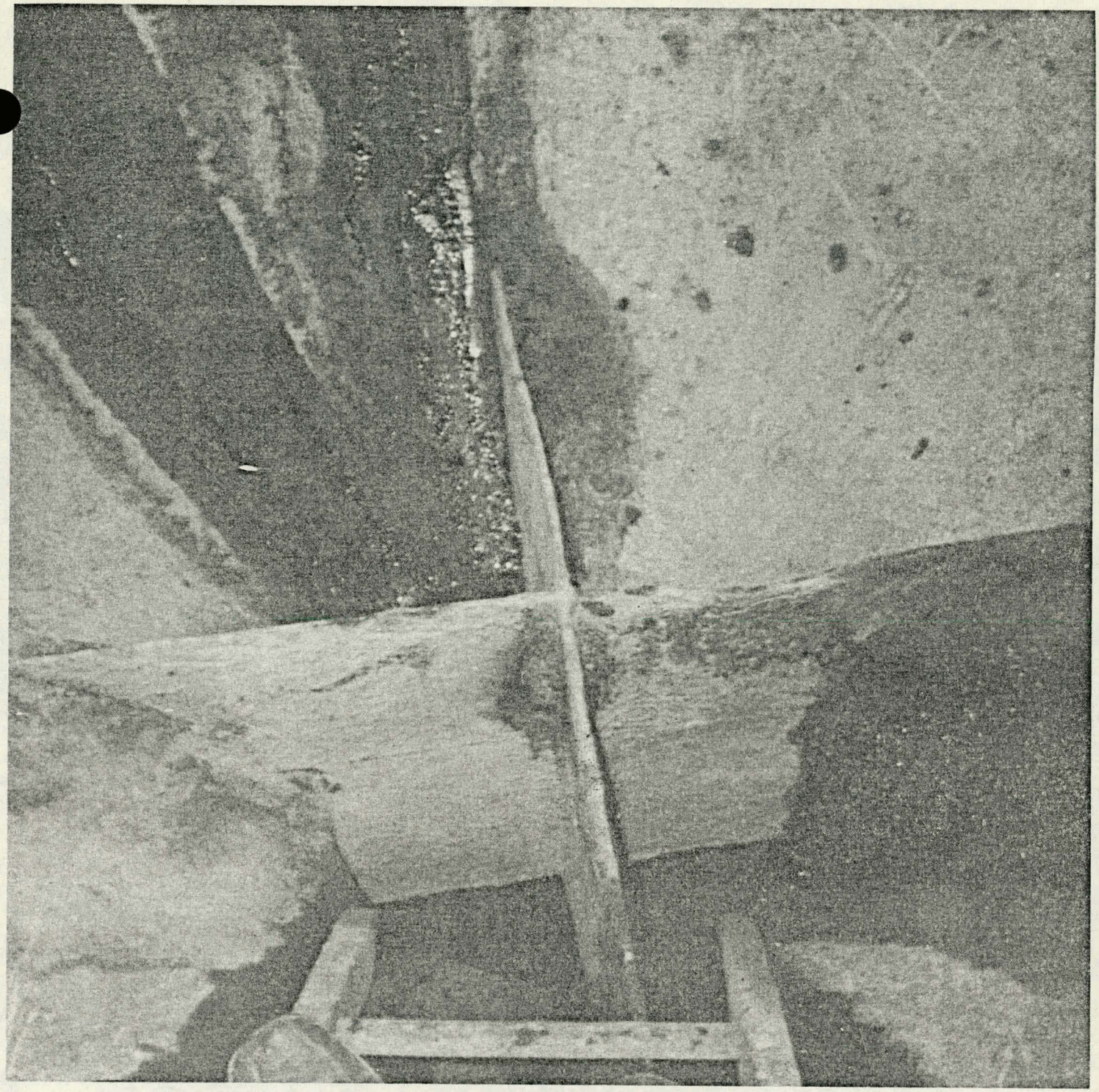
Gas Rate = 25,000 acfm @ 330 °F
Liquid Rate to Venturi = 600 gpm
Liquor Rate to Spray Tower = 1200 gpm
Venturi L/G = 30 gal/mcf
Spray Tower L/G = 60 gal/mcf
Spray Tower Gas Velocity = 6.7 ft/sec
No. of Spray Headers = 4
EHT Residence Time = 12 min

Percent Solids Recirculated = 7-10 wt %
Venturi Pressure Drop = 9 in H_2O
Total Pressure Drop, Excluding Mist Elim. =10.8-12.0 in H_2O
Scrubber Inlet Liquor Temperature = 118-128 °F
Liquid Conductivity = 3,700-7,600 μ mhos/cm
Discharge (Clarifier and Filter) Solids Concentration = 53-60 wt %
Lime Addition to Scrubber Downcomer

Figure D-5. Operating Data for Venturi/Spray Tower Run 619-1A



Venturi Scrubber bottom in conventional lime scrubbing mode.



Venturi Scrubber bottom after Thiosorbic® test.

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UNIVERSITY OF PITTSBURGH

DILUTE-PHASE HYDROGASIFICATION PROCESS FOR SNG PRODUCTION

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DILUTE-PHASE HYDROGASIFICATION PROCESS FOR SNG PRODUCTION

INTRODUCTION

A dilute-phase hydrogasification (DPH) process is under exploratory development by ERDA at the Pittsburgh Energy Research Center (PERC) for the production of pipeline quality substitute natural gas (SNG). The process has the potential of SNG production by a system of high reliability and mechanical simplicity as demonstrated in several long duration experiments. Results are reported from a test series conducted to both demonstrate operability of the DPH process and obtain engineering design data for scale-up to a PDU.

THE DPH PROCESS

Hydrogasification is the name given to conversion of coal to methane by direct reaction with hydrogen according to,

(1) $\text{Coal} + \lambda H_2 \rightarrow CH_4 + \text{Char}$

as opposed to methanation of synthesis gas by,

(2) $CO + 3H_2 \rightarrow CH_4 + H_2O$.

In the DPH process, raw coal is fed directly to the reactor so that the particles free-fall in a dilute cloud through the hydrogen atmosphere. During free-fall particles are heated very rapidly forming a nonagglomerating and highly reactive char, thus eliminating the requirement for coal pretreatment. Since it is estimated that 9% of the volatile matter and 13% of the coal weight may be lost in pretreatment,[1] this approach will result in decreased hydrogen and coal requirements in SNG production. Another attractive feature of the process is that nominally 90% of the methane product is produced directly in the reactor, requiring only a light methanation step to remove the approximately 5% CO remaining in the effluent gas.

A block diagram for the DPH process is shown in figure 1. Coal and hydrogen are fed to the DPH reactor with char and raw gas as products. Char from the reactor may be used either in generation of hydrogen for the process or as a fuel for the plant. Downstream of the reactor the raw product gas passes through purification and light methanation steps to produce a pipeline quality SNG.

LABORATORY DPH FACILITY

A schematic drawing of the facility with instrumentation and sampling points is shown in figure 2. The reactor and experimental procedure used in the DPH experiments have been described in detail previously.[2] Briefly, the reactor consists of an electrically heated pipe of 3.26 inches internal diameter, 5 feet in length which is enclosed in a 10-inch-diameter pressure vessel. Pulverized coal is fed from the hopper by a rotary vane feeder through a water-cooled nozzle to the reactor at rates controllable from 5 to 18 lb/hr. Hydrogen is heated by passing through a helical coil of tubing located in the annulus between the hot reactor and pressure vessel. The hydrogen is then injected at the entrance of the reactor concurrently downward with the coal. Char is collected at the base of the reactor in air-cooled receivers pressurized with nitrogen. The two char receivers are alternately filled and emptied during the test. Tests duration is limited by the capacity of the single high-pressure feed hopper, about 100 lb of coal. Product gas samples are withdrawn at 10-minute intervals directly to a gas chromotograph for analysis. A PDP-11 computer is used for recording, tabulating and storing all relevant variables during an experiment.

EXPERIMENTAL PROGRAM

The objectives of this test series were to demonstrate system operability and reliability within the 100 lb coal feed hopper limitation, to measure the

yields of coal hydrogasification products and to initiate a thorough process design data base. Results are reported in detail of four tests of 5.5 to 8 hours duration. Illinois coal was used in three of these tests and North Dakota lignite in the fourth. It is anticipated that only the more reactive lower rank coals will have adequated carbon conversion for the production of high Btu gas in this single-stage operation. Typical analyses of the coal and lignite used in these experiments are given in table 1.

All tests were conducted at 900° C and 1,000 psig using a 5-foot heated reactor section. The 900° C temperature is that of the reactor wall. No internal thermocouples were located below the point of coal injection to eliminate any possibility of particles sticking to them. Thermocouples were located to measure the hydrogen temperature immediately prior to the point of contact with the coal. These temperatures were approximately 650° C in all tests.

Table 2 summarizes results of these tests in terms of operating test conditions, product yield, conversion, and mass balance. Included also are calculated heating value and composition for the product gas, both as produced and after an assumed light methanation. The Illinois coal and lignite used in these tests were pulverized to 80 pct minus 200 mesh. Average particle size for the lignite and coal was determined by screen analysis to be 73.4 microns and 82.6 microns, respectively.

Experiment #120 was conducted with North Dakota lignite at steady state conditions for 5.5 hours, during which 89.6 lb of lignite were fed. Carbon and maf coal conversions of 40.9 and 56.0 pct by weight were calculated. Product yields of carbon monoxide and water were higher in this experiment than those conducted with Illinois coal due to the high oxygen content of the

lignite. Condensed water from test #120 was 0.7 lb/lb coal fed, where in tests with Illinois coal, it was .03-.04 lb/lb. Light oils were produced only in trace amounts in all tests. Figure 3 shows hydrogen and methane content of the product gas as a function of time during the test. After 40 minutes of operation, the product gas contained about 35 pct methane for the duration of the test. Average particle size of the lignite char was 67.6 microns, approximately the same as the pulverized lignite feed.

Tests #122, #124, and #128 were conducted with Illinois #6 coal. To determine its effect upon coal conversion and product gas heating value, the feed hydrogen to coal ratio was varied to cover the range of 7.93 to 15.24 scf/lb. Both carbon conversion and methane yield were found to vary directly with hydrogen-to-coal ratio, while product gas heating value varied inversely due to the dilution effect of the hydrogen. Char particle size from these tests showed a significant size increase from the coal and averaged 588 microns by screen analysis. The size appeared independent of test condition.

In experiment #122, 72.33 lb of coal were fed in 5.93 hours. A hydrogen-to-coal ratio of 15.2 scf/lb was used which yielded a product gas of 700 Btu/scf. Test #124 was conducted at two hydrogen-to-coal ratios, 9.31 and 7.93 scf/lb, reported as #124A and #124B, respectively, in table 2. Total time of test #124 was 5.93 hours, with #124A lasting 4.5 hours. A total of 85.5 lb of coal were fed in #124. The low hydrogen-to-coal ratios of #124 yielded a higher Btu product gas but with a lower carbon conversion than #122. The slightly higher nitrogen content of the product gas in #124B was due to an operational change in which the char receiver was pressurized with nitrogen before being brought back on stream. Both #122 and #124 were terminated

due to char removal problems. Experiment #128 was conducted in two parts, reported as #128A and #128B, using hydrogen-to-coal ratios of 13.78 and 12.10 scf/lb. A total of 70.4 lb of coal were fed in 7.92 hours. Figure 4 shows the methane and hydrogen content of the product gas in run #128. The results are typical of all the tests with Illinois coal. After attaining steady state conditions, the methane content of the product gas varied less than $\pm$ 2.5% for the duration of the test. The effect upon product gas dilution and carbon conversion of hydrogen feed gas-to-coal ratio for the Illinois coal tests is shown in figure 5.

Table 3 is a comparison showing analysis of the char produced in these experiments with Illinois coal and North Dakota lignite.

CONCLUSIONS AND FUTURE PLANS

The present test series has demonstrated successful continuous operation of a single-stage dilute-phase hydrogasification process in a series of long duration experiments with Illinois hvCb coal and North Dakota lignite. Carbon conversion of up to 36 pct with hvCb coal and 44 pct with lignite has been achieved. The process has yielded a product gas with heating values of 892 Btu/scf with hvCb coal and 938 Btu/scf from lignite, based on the assumption of light methanation.

Future plans include installation of a second coal lock hopper for extended continuous operations. A parametric test program is planned to obtain more thorough process design data and better define product yield and distribution.

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TABLE 1.- Typical analyses of coal and lignite used in this study

	Illinois #6 hvCb coal	North Dakota Lignite
Proximate Analysis (wt pct)		
Moisture	1.0	7.9
Volatile matter	35.2	38.3
Fixed carbon	53.7	42.9
Ash	10.1	10.9
Ultimate Analysis (wt pct dry)		
Hydrogen	4.84	4.15
Carbon	71.82	63.30
Nitrogen	1.72	0.87
Oxygen (by difference)	10.01	21.00
Sulfur	1.41	1.30
Ash	10.20	9.38
	100.00	100.00

HvCb coal from Orient #3 mine, Freeman Coal Co., Waltenville, Ill.
Lignite from Beulah seam, North Dakota.

TABLE 2.- Hydrogasification of coal in dilute phase

Experiment No.	120	122	124A	124B	128A	128B
Test Conditions[1]						
Temperature, (° C)	900	900	900	900	900	900
Pressure (psig)	1000	1000	1000	1000	1000	1000
Time (hour)	5.50	5.93	4.5	1.43	3.75	4.17
Coal	Lignite	Illinois	Illinois	Illinois	Illinois	Illinois
Coal rate (lb/hr)	17.96	12.3	15.0	15.0	10.03	10.03
H_2/coal (scf/lb)	10.97	15.24	9.31	7.93	13.78	12.10
Product Yield[1]						
Gas/coal (scf/lb)	12.37	14.93	9.75	8.03	14.53	12.17
CH_4 (scf/lb)	4.35	6.32	4.77	4.69	6.68	6.31
C_2H_6 (scf/lb)	0.09	0.12	0.15	0.03	0.0	0.0
CO (scf/lb)	1.53	.60	.52	.53	.82	.76
CO_2 (scf/lb)	0.95	.06	.19	.15	.06	.07
Char (lb/lb)	0.497	.625	.635	.668	.595	.643
Product Gas Composition						
H_2 (vol %)	43.6	52.3	42.1	32.5	47.4	40.1
CH_4 (vol %)	35.2	42.3	48.9	58.4	46.1	52.1
C_2H_6 (vol %)	0.7	0.8	1.5	0.4	trace	trace
CO (vol %)	12.4	4.0	5.3	6.2	5.7	6.3
CO_2 (vol %)	7.7	0.4	2.0	1.9	0.4	0.6
N_2 (vol %)	0.4	.3	0.2	0.7	.3	.3
Conversion						
Maf coal	56.0	43.4	40.6	37.0	45.1	39.7
Carbon (wt %)	40.9	33.7	31.6	27.2	36.0	29.8

Continued on next page.

Experiment No.	120	122	124A	124B	128A	128B
Mass Balance[1]						
Overall	92.2	98.3	92.9	94.7	95.1	98.3
Carbon (wt %)	96.6	97.4	94.4	97.3	97.8	102.0
Hydrogen (wt %)	86.3	99.7	93.4	91.5	99.0	94.7
Ash (wt %)	104.6	113.8	100.0	100.0	100.0	100.0
Product Gas After Methanation[2]						
H_2 (vol %)	11.8	46.1	32.0	17.6	37.0	22.9
CH_4 (vol %)	36.9	53.0	66.2	81.9	63.0	77.1
C_2H_6 (vol %)	1.3	0.9	1.8	0.5	0.0	0.0
Heating value (Btu/scf).	938	700	803	892	755	852
CH_4 made in reactor (%).	74.0	91.4	90.2	90.4	89.0	89.2

1 Coal as received.
2 Calculated value assuming methanation step.

TABLE 3.- Typical analysis of char from DPH experiments

	Char From Illinois Coal	Char From North Dakota Lignite
Proximate Analysis		
Moisture	1.2	3.2
Volatile matter	7.2	10.1
Fixed carbon	74.9	67.7
Ash	16.7	19.0
Ultimate Analysis (Dry Basis)		
Hydrogen	1.8	1.3
Carbon	76.3	75.3
Nitrogen	1.5	0.6
Oxygen (by difference)	2.2	1.9
Sulfur	1.3	1.3
Ash	16.9	19.6
	100.0	100.0

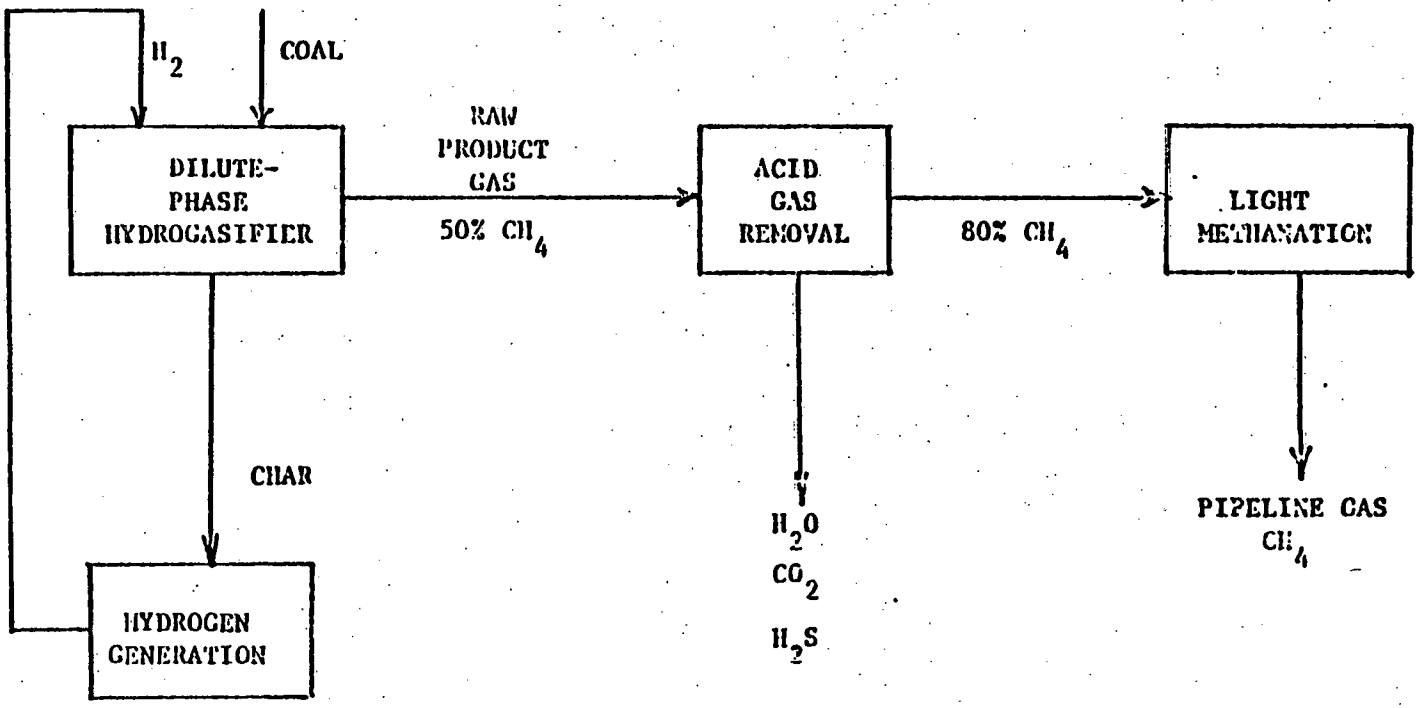


FIGURE 1.- DPH PROCESS

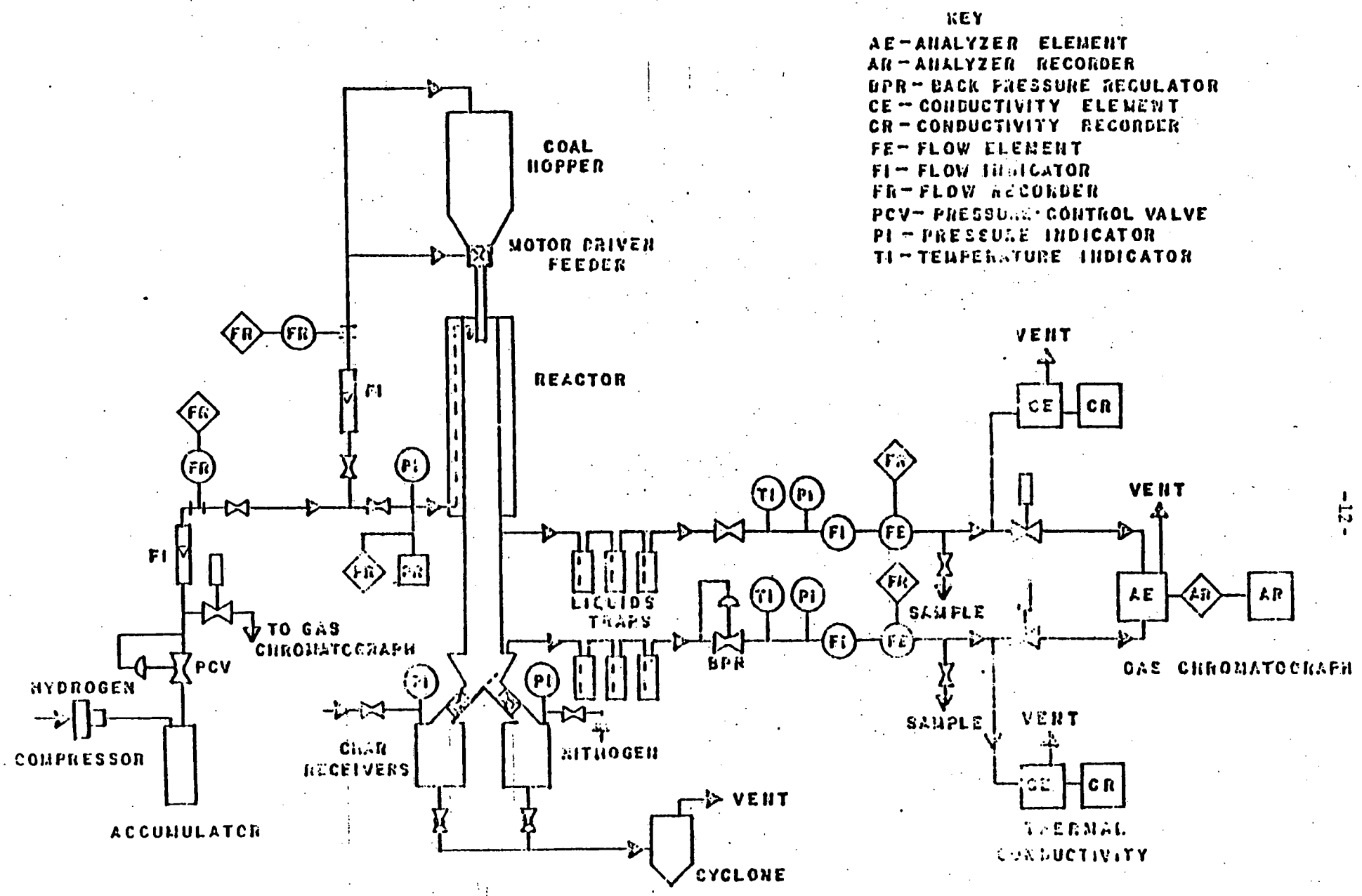


FIGURE 3 - FLOW DIAGRAM- DILUTE PHASE HYDROGASIFICATION

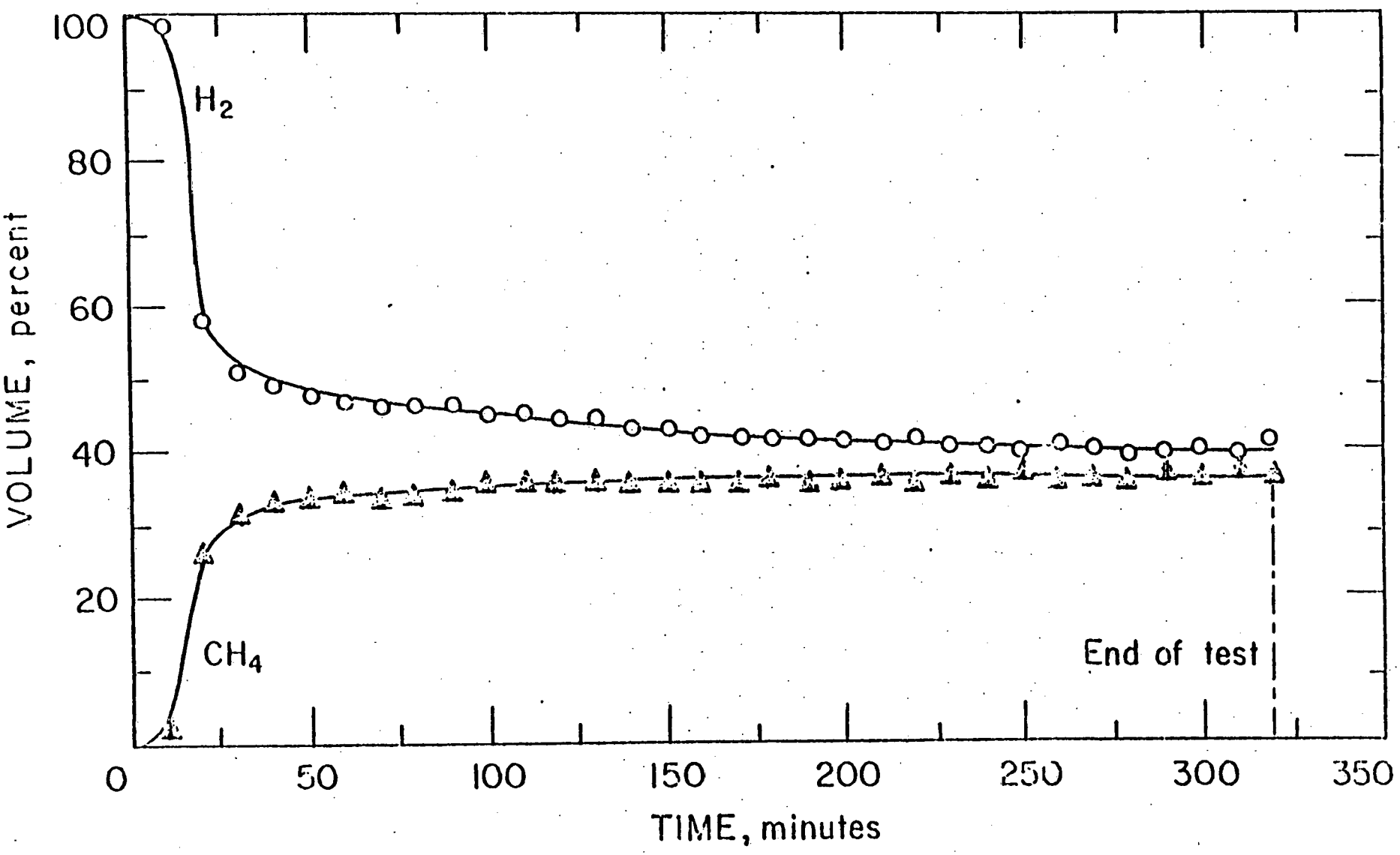


Figure 3-Product gas composition vs time, experiment No. 120, North Dakota lignite.

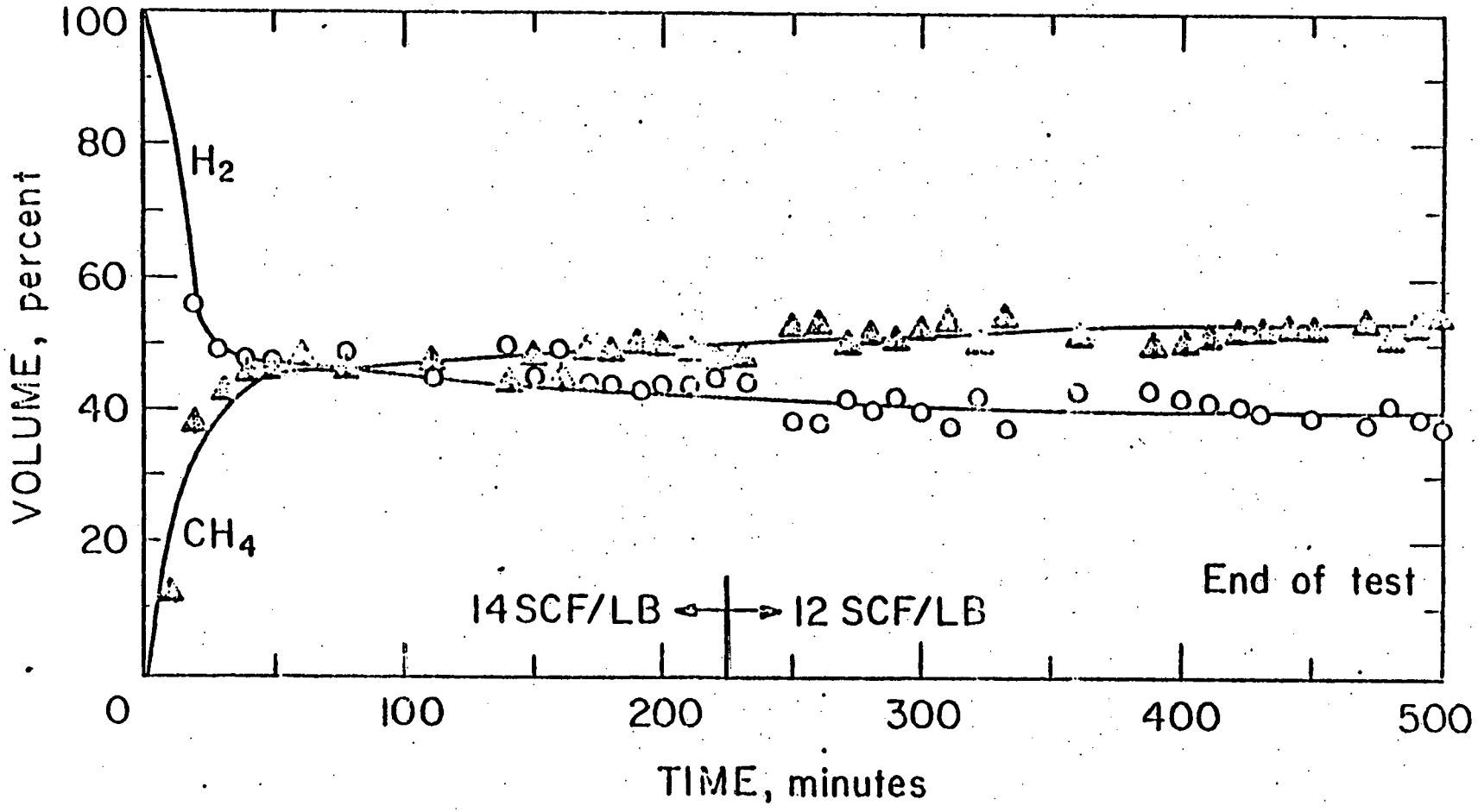


Figure 4 – Product gas composition vs time, experiment No. 128, Illinois No. 6 coal.

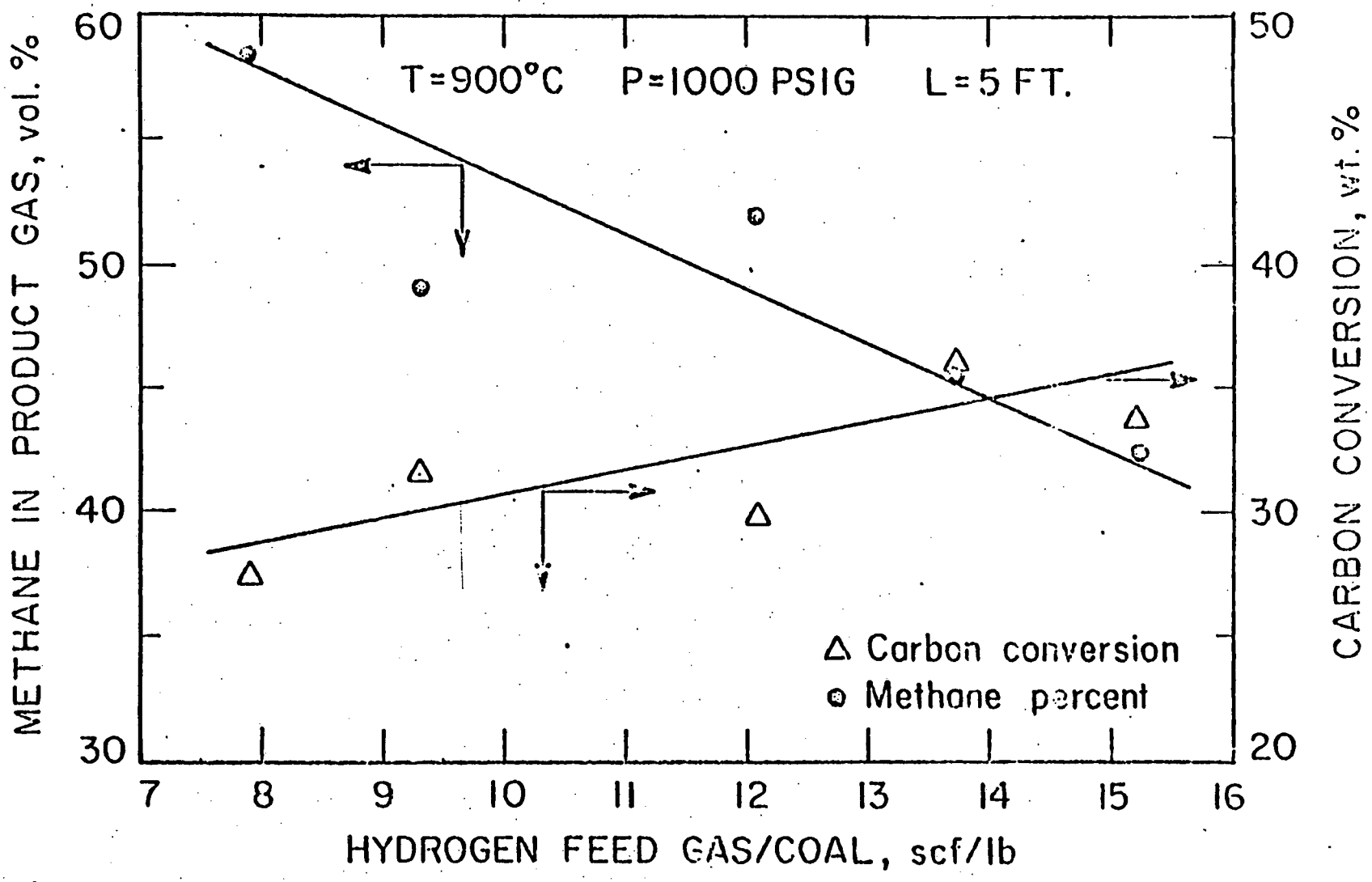


Figure 5—Product gas dilution and carbon conversion dilute phase hydrogasification of Illinois coal.

IMPLICATIONS OF UTILIZING SYNTHETIC FUELS IN COMBINED CYCLES

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IMPLICATIONS OF UTILIZING SYNTHETIC FUELS IN COMBINED CYCLES

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Introduction

The strong implications that coal and various coal derived synthetic fuels will find an increasing role in the generation of domestic power places a new perspective on the basic design approaches of future gas turbines and combined cycle systems.

The characteristics of the fuel to be handled, combusted and expanded in gas turbine cycles has a major influence from the standpoint of equipment reliability, environmental performance, efficiency, and economics. These characteristics effect the basic component design approach, materials selection, selection of the thermodynamic cycle, and the operating conditions, and thus performance levels which may be achieved in the basic gas turbine and correspondingly in the combined cycle.

The following discussion indicates the potential of the combined cycle and how the prospect of utilizing coal or coal derived fuels impact some of the basic considerations in the equipment configuration, design and performance.

The gas turbine has traditionally been an extremely versatile and useful prime-mover concept for power generation. Its application flexibility is demonstrated by the manner in which it can be configured and tailored to meet various but specific application requirements. Simple and regenerative cycle arrangements to supply power for short duration, peak and intermediate duration generation loads results in a very low cost per KW power source. With clean oil and gas fuels, this type of plant minimizes siting problems and costs due to clean emissions, quiet operation, compact size, short procurement and installation cycle, plus no requirement for cooling water.

The combined gas and steam turbine cycles offer increases of almost 50% in efficiency over simple Brayton cycles by the addition of a simplified waste heat recovery steam cycle with minimal water requirements.

These features will be increasingly important considerations, in addition to high efficiency, in meeting the various generation requirements of tomorrow.

As you are aware, gas turbines operating on oil or natural gas fuels and integrated with heat recovery steam cycles have the highest conversion efficiencies in the industry. These cycles offer the potential of significantly higher levels, in addition to improved total plant economics, as turbine operating conditions are increased. Current industrial turbines operate in the turbine inlet temperature range of 1800° to 2000°F while the theoretical limits approach 4000°F; thus, we're only halfway there. The challenge now being faced is to continue to achieve this performance potential while meeting the changing situation of fuel availability with high performance reliability.

Fig. 1 shows the chronological history of turbine inlet temperatures as the advanced technology in materials and cooling design is extrapolated into industrial power generation application. It can be noted that by the late

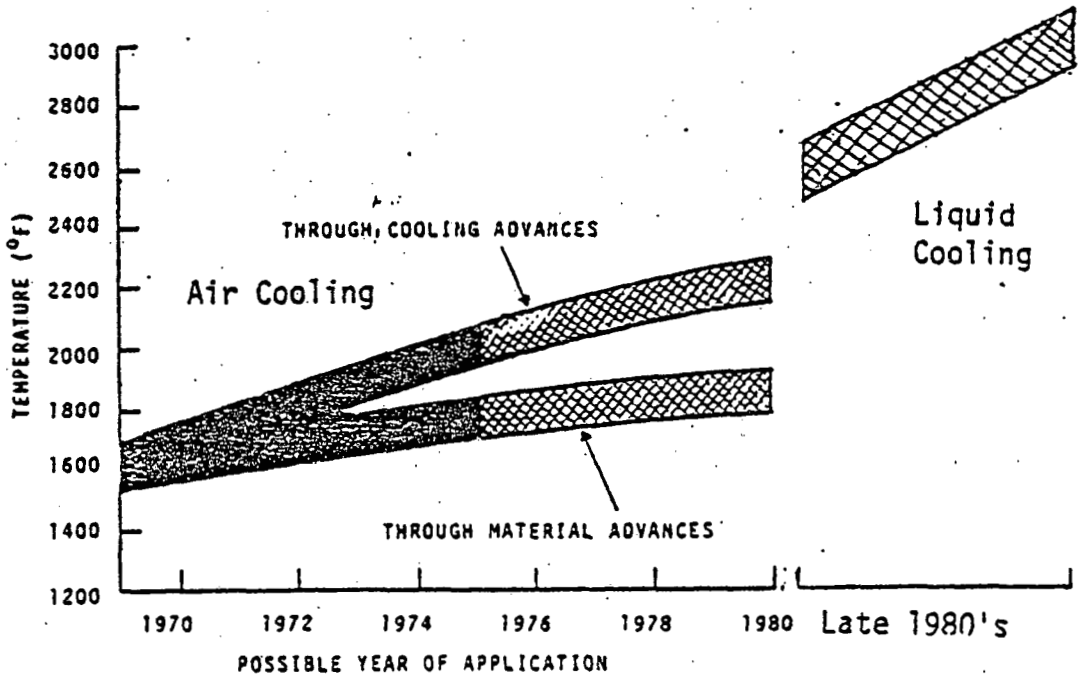


Figure 1 Gas Turbine Firing Temperature Trend

'70's turbine peak inlet temperature approaching 2200°F can be developed primarily through advances in air cooling and materials with future prospects of 3000°F levels as liquid cooling technology is developed through the '80's.

As you are aware, turbine inlet temperature is the key variable in the design and resulting performance of power generation gas turbines. Fig. 2 and 3 indicate the strong relationship of turbine inlet temperature and the output and efficiency of the various cycle configurations. The combined cycle performance is tabulated below:

Combined Cycle Performance (Synthetic Liquid Fuels)

Turbine Inlet Temp °F	Heat Rate BTU/KW/HR
2000	8100
2400	7800
3000	6700

These advancements produce combined cycle efficiencies of 48 to 51% with individual unit combined cycle sizes of 300 to 350 MW. Although the thermodynamic performance of these advanced systems may stir the imagination, the aspect of systems reliability and environmental performance must be given priority in pacing the technological development to insure an overall performance level consistent with the application needs. The pace of the development of this technology will certainly depend on the characteristics of the fuels to be utilized.

Future gas turbines must have fuel burning flexibility ranging from low-grade crudes, residuals and synthetic liquid products (resulting from various coal refining concepts) through a spectrum of low and intermediate specific heat content gases produced by coal gasification processes.

Turbine component cooling and advanced materials technology which has been incorporated in today's and to an increasing degree in tomorrow's gas turbines is of great advantage in utilizing lower grade fuels by reducing corrosion attack while permitting increases in turbine inlet temperatures.

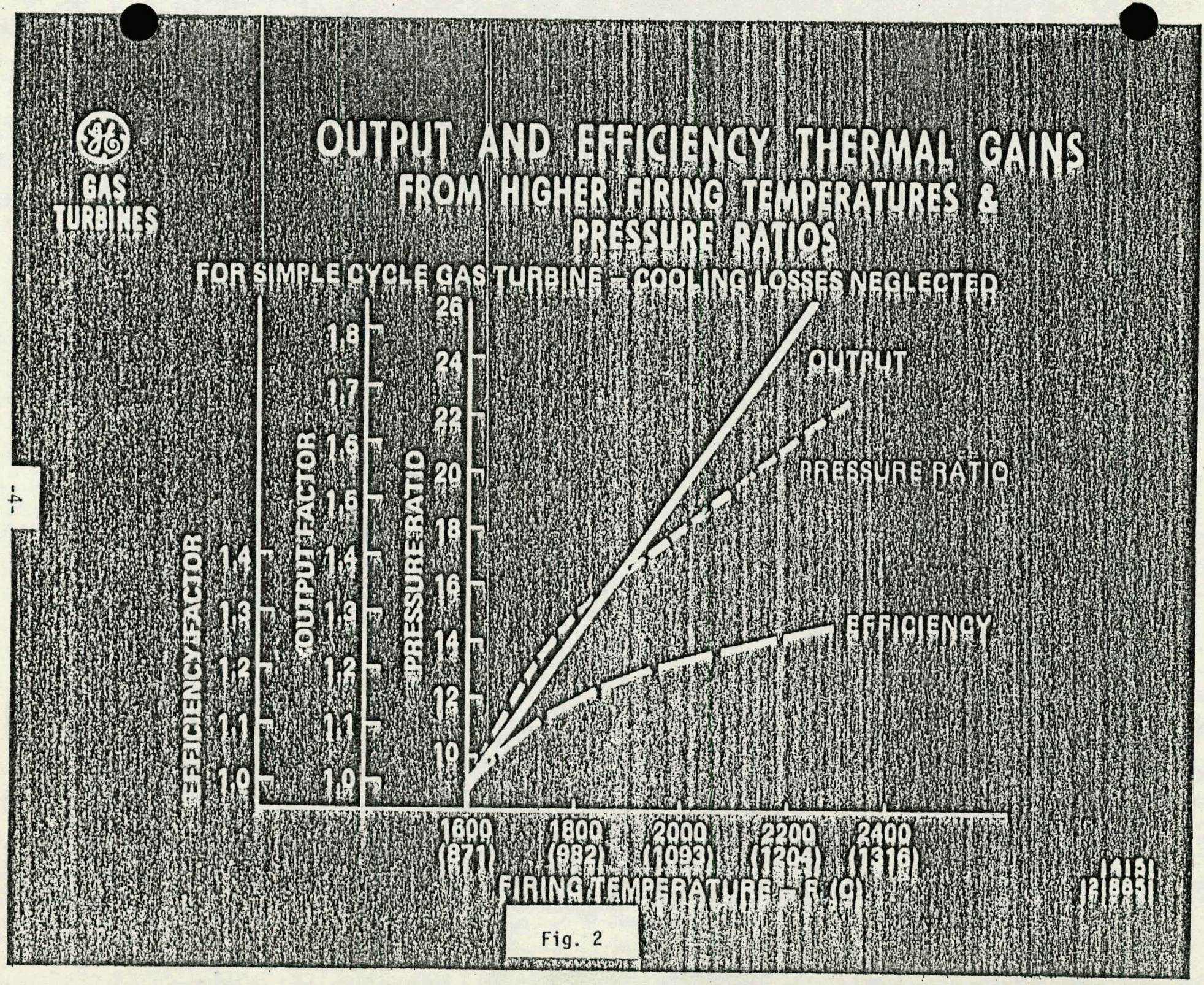

GAS TURBINES
OUTPUT AND EFFICIENCY THERMAL GAINS
FROM HIGHER FIRING TEMPERATURES &
PRESSURE RATIOS
FOR SIMPLE CYCLE GAS TURBINE – COOLING LOSSES NEGLECTED
OUTPUT
PRESSURE RATIO
EFFICIENCY
EFFICIENCY FACTOR
1.4
1.3
1.2
1.1
1.0
OUTPUT FACTOR
1.8
1.7
1.6
1.5
1.4
1.3
1.2
1.1
1.0
PRESSURE RATIO
26
24
22
20
18
16
14
12
10
1600 (871)
1800 (982)
2000 (1093)
2200 (1204)
2400 (1316)
FIRING TEMPERATURE – R (C)

Fig. 2

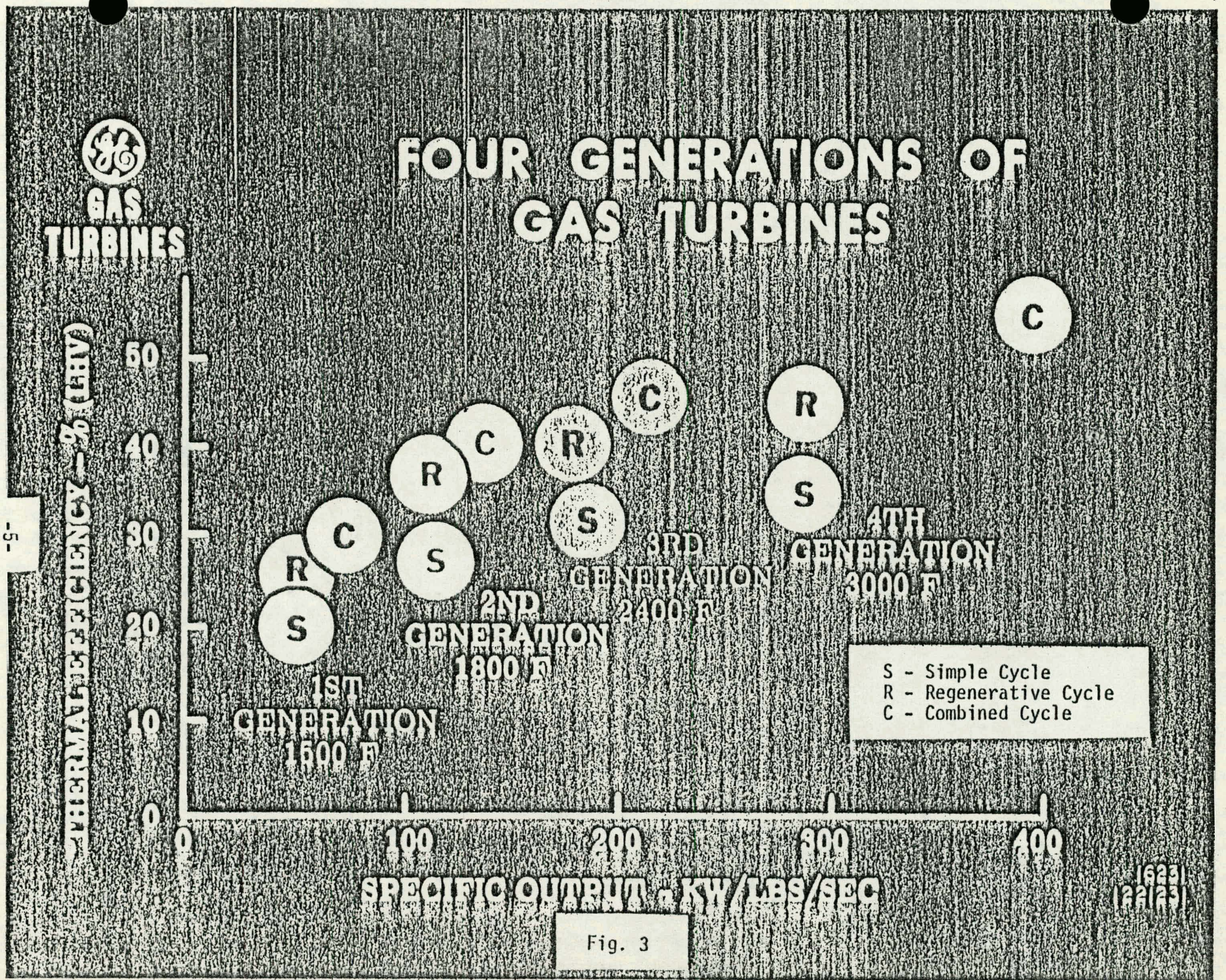

GAS TURBINES
FOUR GENERATIONS OF GAS TURBINES
THERMAL EFFICIENCY - % (LHV)
50
40
30
20
10
0
0
100
200
300
400
SPECIFIC OUTPUT - KW/LBS/SEC
(162.3)
(122.3)
S
R
C
1ST GENERATION 1500 F
2ND GENERATION 1800 F
3RD GENERATION 2400 F
4TH GENERATION 3000 F
S - Simple Cycle
R - Regenerative Cycle
C - Combined Cycle

Fig. 3

The type of fuels to be utilized in combined cycles can impact the basic equipment configuration employed.

Liquid Fuels from Coal

Fig. 4 indicates a conventional type of combined cycle arrangement which is employed with heavy residual or synthetic oils from coal liquefaction. A fuel treatment function is indicated which is primarily for fuel preheat, washing of soluable alkali metal compounds and inhibition of vanadium. Such systems have been developed and commercially applied with residual and crude oil. Much of this work will be applicable in utilizing the lower grade oil fuels produced by various coal liquefaction processes now being developed.

Although such fuels have the potential of achieving the 50% thermal efficiency levels indicated in Fig. 3, there are several important aspects with respect to thermal and environmental performance of these synthetic liquids which should be noted.

Significance of Fuel Properties

For gas turbine application, critical fuel properties can be divided into three groups:

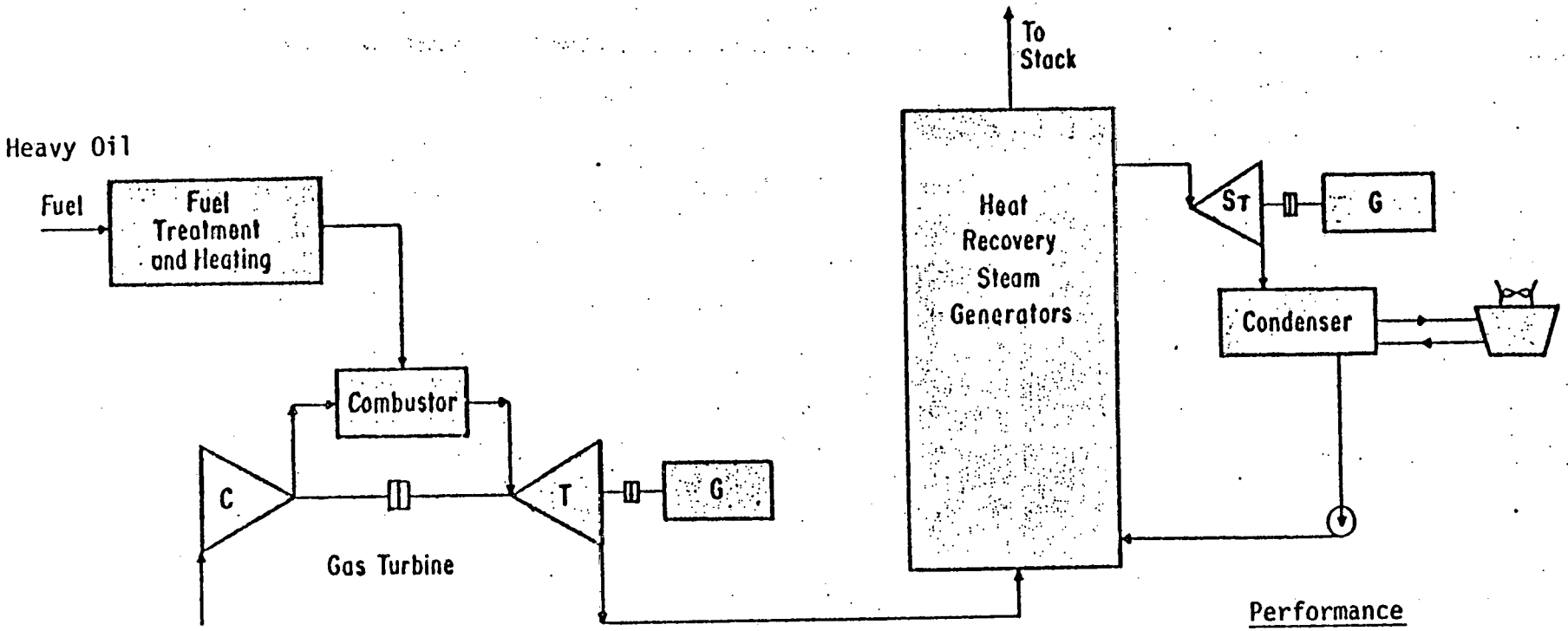
1) Ash-forming trace contaminants
2) Fuel chemical properties
3) Fuel physical properties

The ash-forming contaminants can have a pronounced effect on turbine reliability, maintenance and availability since certain of these contaminants can cause erosion, hot corrosion and ash deposit buildup.

Chemical properties of liquid fuels affect the overall combustion process including the exhaust emissions.

Physical properties also have an effect on combustion as well as fuel handling requirements.

COMBINED CYCLE UTILIZING
HEAVY LIQUID FUELS



Performance

Turbine Temp °F	Heat Rate BTU/KW/HR
2000	8100
2400	7800
3000	6700

Fig. 4

The following table outlines the significance of critical fuel properties on gas turbine system performance.

INFLUENCE OF FUEL PROPERTIES ON GAS TURBINE SYSTEM PERFORMANCE

	Fuel Handling	Fuel Combustion	Turbine Maintenance & Reliability	Stack Emissions
Ash-Forming Trace Contaminants			x	x
Chemical Properties				
H/C Atom Ratio (Aromatic Content)		x		x
Nitrogen Content				x
Sulfur Content	x			x
Carbon Residue		x		x
Thermal Stability	x	x		
Fuel Compatibility	x			
Physical Properties				
Viscosity	x	x		
Distillation Range		x		
Specific Gravity	x			
Flash Point	x	x		
Pour Point	x			
Filterable Dirt	x			

Ash-Forming Trace Contaminants

Ash-forming trace contaminants are important from a turbine corrosion, erosion and deposition standpoint. Superalloys in high-temperature service oxidize uniformly and predictably because of the formation and retention of protective oxides. Hot corrosion, however, is initiated by the destruction of protective oxides via reactions with S, S/alkali compounds, V, Pb, and other contaminants. It culminates in accelerated oxidation of the alloy, resulting in a catastrophic rate of material destruction.

The level of sulfur required to initiate this reaction is very low and the reaction quickly becomes strongly dependent primarily on the levels of alkali metal compounds present. Thus, the alkali metal content is considered the dominant and controlling factor.

Four methods have been used to combat hot corrosion:

-- Development and use of hot corrosion resistant alloys and coatings
-- Specifications to control contaminant levels in fuels to appropriate levels
-- Fuel treatment to remove or inhibit contaminants
-- Reduction of metal surface temperatures to a point where corrosion proceeds at an acceptably low rate

There are performance/cost trade-offs associated with each of the methods. This optimization can only be made after the fuel characteristics and their behavior with the various approaches has been tested and analyzed.

Erosion resulting from the impact of particles entrained in a gas flow upon a surface is mainly a function of the following variables:

1) Velocity, hardness, size, shape, concentration, and angle of impact of the particles, and
2) hardness of the material surface.

In the turbine gas stream, large slow moving particles can be impacted on both stationary and highly rotating parts, causing erosion.

Although the severity of the erosion depends to a great extent on the physical characteristics of the particles (sharp edged dense particles being more severe than soft power ash or spheroidal configurations), a general rule is that particles in the 2 μ to 10 μ range should be kept below 0.01 GR/SCF with no appreciable quantity above 10 μ and minimum limits on particles less than 2 μ.

For larger particle sizes, the amount of metal removed is directly proportional to the weight of the particles in the gas stream and is independent of particle size.

Centifugal forces and secondary flows in a turbine can result in concentration of the particles at either or both the I.D. or O.D. of the turbine gas path, producing concentrated erosion. This is a major

consideration with the larger particles and can result in significant aerodynamic shape changes and/or special material selections.

Analysis of the coal-derived liquid (CDL) fuels to be utilized in the turbine is necessary to establish the basic characteristics and quantity of this type of ash and its possible impact on turbine design and performance.

Deposits from the ash fuels burned in gas turbines can adhere to the hot-gas path parts and restrict the flow or change the direction of the hot gas, thereby reducing the output and efficiency of the turbine in addition to the possibility of exceeding the compressor stall limits. The rate at which these deposits accumulate and the ability to remove them will depend upon a number of factors, such as fuel composition, treatment effectiveness, turbine design, and firing temperatures.

A critical aspect of this fouling is that associated with advanced component cooling techniques. Fouling of small air cooling openings in critical nozzle and bucket components can cause loss of cooling with resulting major material damage. This effect could prohibit the use of some advanced air cooling concepts, necessitate reduction in firing temperatures, or require development of a new cooling technology with attendant effects on unit performance levels. Internal cooling techniques, such as use of liquid coolants, could alleviate this problem but requires extensive development before commercialization.

At the present state of development of coal liquefaction processes, it is not possible to define the exact properties and composition of future commercial coal-derived liquid (CDL) fuels. The limited data currently available on these fuels is based on pilot plant or bench scale processing and may not be necessarily typical of ultimate large scale commercial production. However, this data can be used to estimate typical fuel profiles for the future commercial coal-derived liquids.

Of the various types of liquid fuels which can be produced in the coal liquefaction processes, the fuels which seem most likely to be available for gas turbines are:

1) Boiler grade low sulfur fuel
2) Heavy distillate fuel

Such intermediate grade products are characteristics of currently proposed liquefaction processes, such as H-coal and synthoil.

The H-coal process involves the direct catalytic hydrogenation of coal and is operated in one of two alternate modes: syncrude and low sulfur boiler fuel. In the syncrude mode, three boiling ranges of distillate (low ash) fuels can be produced comprising naphtha, light distillate and heavy distillate. In the boiler fuel mode, the main product is a heavy fuel with some ash content, somewhat analogous to petroleum residual fuel. The principal operating mode for the ultimate commercial plant will depand on operating experience of the demonstration plant. The syncrude mode offers higher quality products but at the expense of higher hydrogen consumption and lower product output rate.

The ERDA synthoil process is also a direct catalytic hydrogenation process with syncrude as the initial product which can be further processed into a series of distillate liquids and possibly a residual fuel.

The boiler-grade fuel has a low H/C ratio, high nitrogen content, less than 1% sulfur and ash carry-over from the parent coal. The target ash content is 0.1% maximum, although most pilot plant batches to date have exceeded this. The final level depends on improvement in the solids removal step in the process. The viscosity of the boiler grade is camparable to the heaviest residual petroleum fuels. The specific gravity of this fuel is greater than 1.0 (heavier than water).

The heavy distillate fuel would be ash-free if a sophisticated vacuum still were used; but if a simple flash still were used, the heavy distillate could contain small amounts of ash. The H/C ratio and nitrogen level are comparable to the boiler grade fuel but the viscosity would be lower.

An estimate of the ash limit of CDL's is 0.1% (1000 ppm). This is the target goal in some of the processes and is also consistent with the ash limit based on stack particulate emissions.

The chemical composition of the ash in the CDL will reflect the composition of the mineral matter in the parent coal. The chemical elements in the ash-forming contaminants will not be in exactly the same ratio as in the coal since there is usually selective removal of some of the elements during the overall processing from the coal preparation of the final CDL product.

The ash-forming material remaining in the CDL product would have a smaller average particle size than the solid product removed but the chemical composition should be nearly the same.

The content of alkali metals is expected to be significantly greater than that experienced with conventional distillate oils. Such CDL fuels will probably require a water wash or similar treatment to reduce the total alkali content to less than 1 ppm as required by current high temperature turbines.

Although particulate content is not anticipated to cause significant erosion, the buildup of ash deposits and possible approaches for their periodic removal must be investigated by tests with CDL synthetic fuels.

Chemical Properties

Coal-derived liquid fuels contain a high percentage of aromatic hydrocarbons as the principle fuel component in contrast to petroleum fuels where the percentage of aromatic components is generally low. Coal-derived liquids also have higher contents of complex organic compounds of sulfur, nitrogen and oxygen.

The high aromatic hydrocarbon content means that a large fraction of the heat released during combustion would be in the form of radiant energy; and this could have a significant effect on combustor design, cooling and performance.

The relatively large concentration of complex nitrogen and sulfur compounds can result in high levels of NO_x and sulfur oxides respectively in the turbine exhaust gases. Oxygenated compounds in the fuel may degrade the thermal stability.

The H/C ratios of CDL fuels, both boiler grade and heavy distillate, appear to be typically 1.0-1.1. This is in contrast to ratios of 1.5-2.0 with good petroleum fuels and 1.35 for the poorest petroleum residual fuel burned successfully to date.

CDL fuels may be upgraded for gas turbine use, most probably by hydrotreating which raises the H/C ratio. The critical factor in upgrading may be the nitrogen content. If the nitrogen is lowered to lower NO_x emissions, the H/C ratio will increase since hydrogen addition proceeds more readily than de-nitrogenation.

Carbon residue is a measure of the degree to which fuels will pyrolyze to solid carbonaceous residues on exposure to high temperatures in the combustion system. It can lead to fuel nozzle fouling and possibly smoke particles in the exhaust gases. In heavy-duty gas turbines, petroleum residual fuels have been burned with satisfactory smoke emission levels at carbon residues of 3-4%. CDL gas turbine fuels could have carbon residues ranging from 3-7%. Up to 10% might be burned successfully but the smoke levels may be excessive.

Fuel-bound nitrogen is one source of NO_x in the exhaust gases, the other source being the direct combination of atmospheric nitrogen and oxygen in the combustion process. The nitrogen in CDL fuels is not only at higher levels than in the petroleum counterparts, but also it occurs combined within very stable compounds. The primary liquids resulting from coal liquefaction have nitrogen contents of 1% or higher. Upgrading by hydrogenation can reduce this, but the level which can be justified economically still must be resolved. This is a problem common to most of the CDL processes.

The maximum allowable fuel-bound nitrogen for today's heavy-duty gas turbines is less than 0.1% if the turbine system is to meet strict NO_x levels. An economic breakthrough in de-nitrogenation appears required to meet the necessary levels or the development of a new type of staged combustion system.

The level of sulfur could be expected to be reasonably low (less than 0.5%) and is not expected to be a critical problem in CDL in terms of emissions or corrosion. Any de-nitrogenation by hydrogenation of the fuel would result in an even greater decrease in the sulfur level. The sulfur remaining in the product fuel after processing should be in a relatively inert form and would not be expected to cause corrosion in the fuel system or turbine.

Heavy fuels tend to undergo thermal breakdown at high temperatures resulting in the formation of deposits of sludges. If these are formed in the fuel system, they can foul fuel nozzles and fuel lines. In the application of heavy petroleum fuels to gas turbines this can be avoided by keeping the fuel temperature under a maximum limit. The equivalent temperature for CDL fuels has not been established. The high aromatic content could help thermal stability, but the presence of oxygen compounds in CDL's could make them less stable. This property will have to be established by a test program.

Physical Properties

Of the physical properties of gas turbine liquid fuels the most significant are viscosity, viscosity-temperature relationship, fuel distillation range, specific gravity, flash point and pour point.

Viscosity level and viscosity-temperature relationships of a liquid fuel determine fuel heating requirements for fuel pumping and fuel atomization. CDL fuels tend to have higher viscosity levels than their petroleum counterparts, but the viscosity of CDL fuels decreases more rapidly as the temperature is increased (lower viscosity index). The proper handling temperature for CDL fuels must be established by a test program.

The temperature range over which the liquid fuel volatilizes affects ignition requirements and ignition reliability. Very low volality fuels (light naphthas) and most heavy fuels require a secondary fuel (usually No. 2 fuel) for light-off. Boiler grade fuels have a high percentage of heavy fuel components which do not volatilize even at 1000^{o}F.

In advanced combustors designed for low NO_x formation, it may be necessary to volatilize the liquid fuel before it enters the combustor. This could require a new combustion design for boiler grade fuels.

Specific gravity is important when the fuel requires water washing to remove water-soluble ash-forming contaminants (sodium and potassium). Available washing hardware requires that the fuel be several percent less dense than water at the operating temperature of the washing equipment. In CDL fuels, the boiler grade would need cleanup for use in conventional heavy-duty gas turbines since the current maximum level of alkali metals is less than 1 ppm while CDL fuels may have more than an order of magnitude in excess of this. Heavy distillates may not require washing unless ash-forming contaminants were carried over in the vacuum distillation equipment. Since only highly hydrogenated (upgraded) CDL fuels would be lighter than water, CDL fuels available for gas turbines requiring washing to remove water soluble contaminants would require the development of new washing techniques and/or hardware.

Low flash points are a problem in light distillate fuels or blended fuels with light components. Heavy distillate grade and boiler grade CDL fuels should not present a problem.

The viscous pour point would be of concern with CDL since it indicates the congealing of hydrocarbon fuel components as the fuel temperature is lowered and can lead to problems in line fouling and pumping.

Particulate matter is largely removed by a final fuel filter. This particulate matter may be carry-over from the original fuel processing, or part of it may be solid products formed after processing or aging of the fuel. This may require particular attention with CDL since a higher degree of polymerization may result during storage. The significance is one of maintenance or the frequency at which filters must be serviced.

These many facets of the impact of the fuel characteristics on gas turbine design and performance are important considerations in developing and optimizing those coal liquefaction processes which may find commercial applications for power generation fuels.

Gas Fuels from Coal

Fig. 5 indicates an integrated low BTU/combined cycle configuration which could find application with the various air-blown low BTU fuel gas systems now being contemplated for development.

The integrated gasification/combined cycle shown in Fig. 5 is a concept of particular potential in converting low-grade fuels to electrical energy. Direct integration of the fuels plant with the power plant affords significant opportunity to optimize in matching process needs with power plant capabilities. This reduces the overall cost and increases the fuel to power efficiency by taking advantage of the high conversion efficiency and low cost of combined cycle power generation.

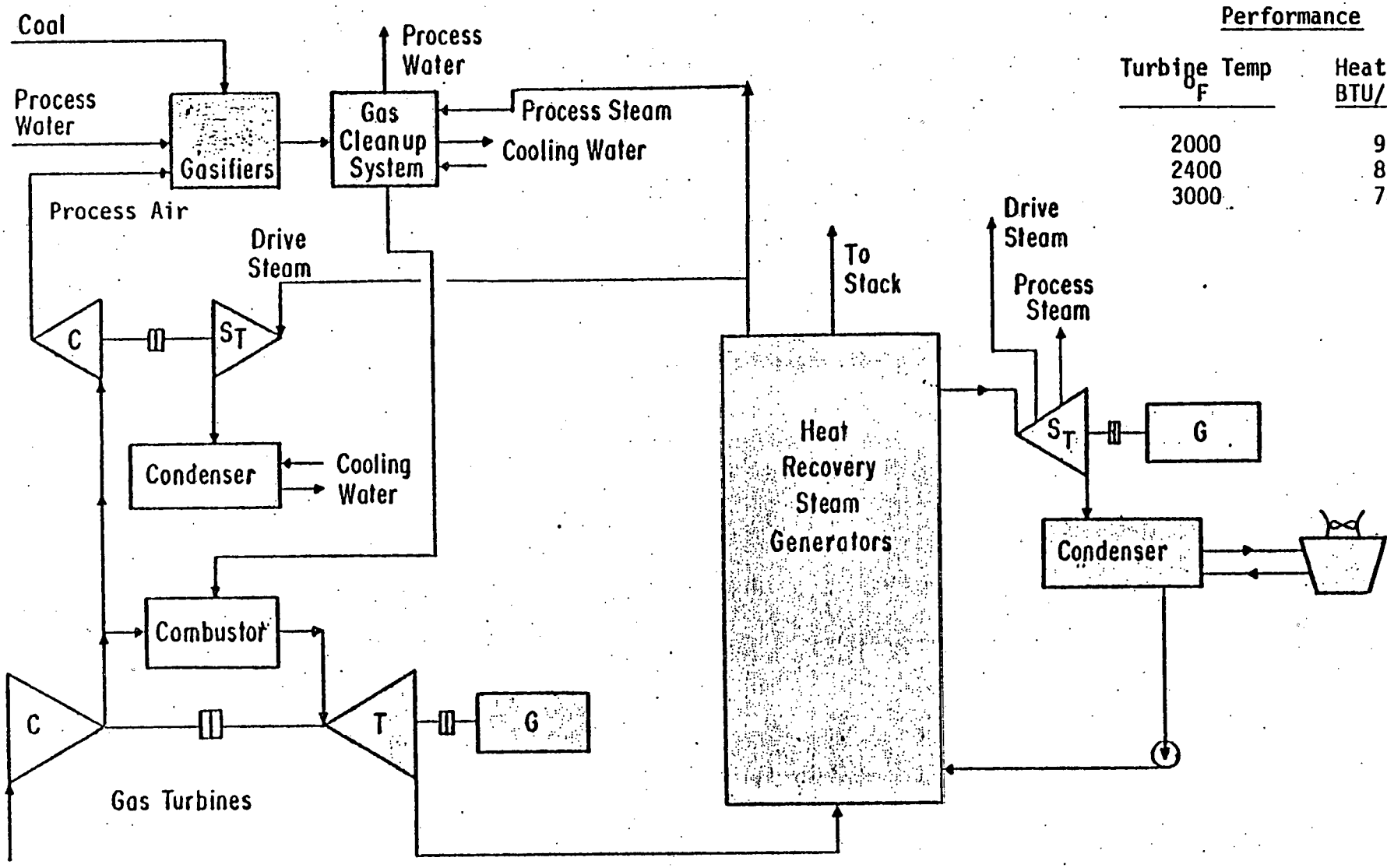
There are a variety of gasifier designs, fixed bed, fluid bed and suspension bed, each with its particular operating characteristics, including pressurized, atmospheric, low or high temperature, etc., and thus having different integration requirements and performance in power generation configurations.

In the process of Fig. 5, coal is gasified at pressure. High-pressure steam and air required for process reactants are extracted from readily available sources within the combined cycle boiler and steam and gas turbines.

Gasification is a partial oxidation process whereby fuel and air or oxygen are reacted in a vessel under reducing conditions. Partial combustion supplies heat for subsequent endothermic reactions in the gasifier reducing zones. A combustible gas containing carbon monoxide and hydrogen is thus produced. Methane and volatile vapors may also be formed in low temperature gasification processes.

Although the heating values will vary with feedstock, a typical range of 250 to 350 BTU/SCF and 100 to 150 BTU/SCF gas is produced with oxygen and air-blown gasifiers, respectively. The sulfur in the coal is converted to hydrogen sulfide in the gasifier reducing atmosphere which, in addition to particulate matter, can be readily removed from the fuel gas stream by various commercial chemical wash processes and water washing respectively.

INTEGRATED LOW BTU GASIFICATION/COMBINED CYCLE



Performance

Turbine Temp °F	Heat Rate BTU/KW/HR
2000	9700
2400	8600
3000	7800

Fig. 5

Of course, with the use of oxygen and additional gas processing equipment, including shift conversion, CO_2 removal and methanation, synthetic natural gas (SNG) is produced. However, these complex and costly steps are not required for a fuel gas which is to be used directly and not transported long distances or stored.

The estimated integrated combined cycle performance of such systems is indicated below:

Integrated Combined Cycle Performance (Synthetic Low BTU Gas Fuels)

Turbine Inlet Temp °F	Heat Rate BTU/KW/HR
2000	9700
2400	8600
3000	7800

Combustion tests conducted to date with low BTU gas have indicated very desirable combustion characteristics in terms of ignition, flame stability, turndown, and emissions.

These combustion characteristics are an important aspect for the success of the integrated plant concept. This is significant from the standpoints of environmental and reliability performance, handling and combustion of the fuel, and possible operating constraints which may be imposed by use of these fuels, since the composition and contamination levels of these fuels is a major factor in turbine design and performance.

Since permissible contaminant levels in the gas fuel will depend on its specific constituency, the levels to be indicated in the discussion can only be considered approximate and are referenced to the low BTU gas fuels of ~130 BTU/SCF.

Chemical Content

The presence of fuel-bound nitrogen (FBN) in the gas, usually in the form of NH_3, will result in an increase in the emissions of NO_x in the gas turbine exhaust. Tests indicate that the conversion rate of ammonia (NH_3)

to nitrogen oxide (NO) is very high in turbine combustors where the presence of very large amounts of excess air increase the yield. This is in addition to the thermal NO_x which would be produced during the combustion of the fuel even in the absence of any FBN. Small amounts of fuel moisture can significantly reduce the thermal NO_x component of the emission but has a deleterious effect on NO_x yield from FBN.

An emission limit of 100 ppm of NO_x by volume would indicate an equivalent NH_3 fuel gas content of less than about 250 ppm in the fuel gas and, without fuel gas moisture or moisture injection in the turbine combustors, fuel-bound nitrogen should be eliminated from the gas to meet emission requirements.

Although CO concentration in the exhaust gases increases substantially when excessive fuel moisture is present, particularly in the form of droplets, results show that up to 30% H_2O by volume could be tolerated in a low BTU gaseous fuel without significantly affecting the combustion efficiency and the CO emission levels. Where the gases in question are dry when burned, the effect of H_2O on CO emissions is not of concern. Small percentages of H_2O are known to increase the reaction rate of CO. Thus, a slightly "moist" gas may be more desirable than a dry gas especially if the H_2 content is near the lower limit.

Tests indicate that CH_4 has an adverse effect on the combustion of CO in low BTU flames. The degree to which the CH_4 inhibits the CO reaction is not presently known and requires tests to establish permissible levels. However, the fact that fuel gases from pressurized gasifiers may contain considerable CH_4 is of interest and should be investigated further.

Efficient combustion of low BTU fuel will require a minimum H_2 content of about 10% by volume. This requirement is necessary in order to assure efficient consumption of the CO in the gas since the reaction rate of CO (which is normally very slow) is dependent on the present of H_2. Most low BTU fuels being considered can meet this requirement.

Another important factor is the H_2/CO ratio. Based on available data, it appears a 1:2 ratio may be acceptable which also meets most low BTU fuel gas compositions.

Most low BTU fuels have a ratio of upper to lower flammability limit of about 4:1. This is well above gas turbine specified minimum and thus should pose no problem. This factor is important in obtaining reliable turndown without the complicated use of supplemental fuel.

As a result of the low heating value and inert content of many low BTU gases, the combustion system component requirements becomes greatly increased since more gas is needed to maintain the gas turbine firing temperature, particularly at the advanced firing temperatures of future units.

The stoichiometric temperature rise of fuel gases as low as 100 BTU/SCF will permit continual advance of turbine inlet temperatures to 3000°F. However, fuel gases much below this level will require compromise in thermal performance and significant hardware design difficulties in fuel handling, valving, control, combustion, and hot gas handling of such large volumes of gases. Fuel gases with low heating values also result in large fuel air ratios as compared with use of oil or natural gas fuels, whereas the fuel flow constitutes about 2% of the turbine flow with conventional fuels, low BTU fuels represent about 20%. This effect can cause large flow mismatches between compressor and turbine components, normally designed for operation on high specific energy fuels which must be compensated by new turbine component designs, reduced firing temperatures or use of turbine air extraction for process reactant feed. It should be noted that there are also attendant new turbine control and equipment protection considerations associated with such significant cycle changes.

An additional effect of these large fuel/air ratios is the decrease in permissible contaminant levels in the fuel gas. Essentially the contaminant concentrations must be reduced by a factor of ten to produce the equivalent emission, corrosion, erosion and fouling performance previously discussed under liquid fuels. Such degrees of cleanliness may require development of fuel gas cleaning concepts and equipment and turbine component design compromises to be commercially viable.

Many of the comments with respect to ash-bearing trace contaminants and their impact on turbine design made previously concerning coal-derived liquids are also directly applicable in the case of low BTU fuel gases.

The multiple washing steps possible, and necessary, during the fuel gas cleaning are a significant factor anticipated to reduce problems associated with corrosive and particulate matter.

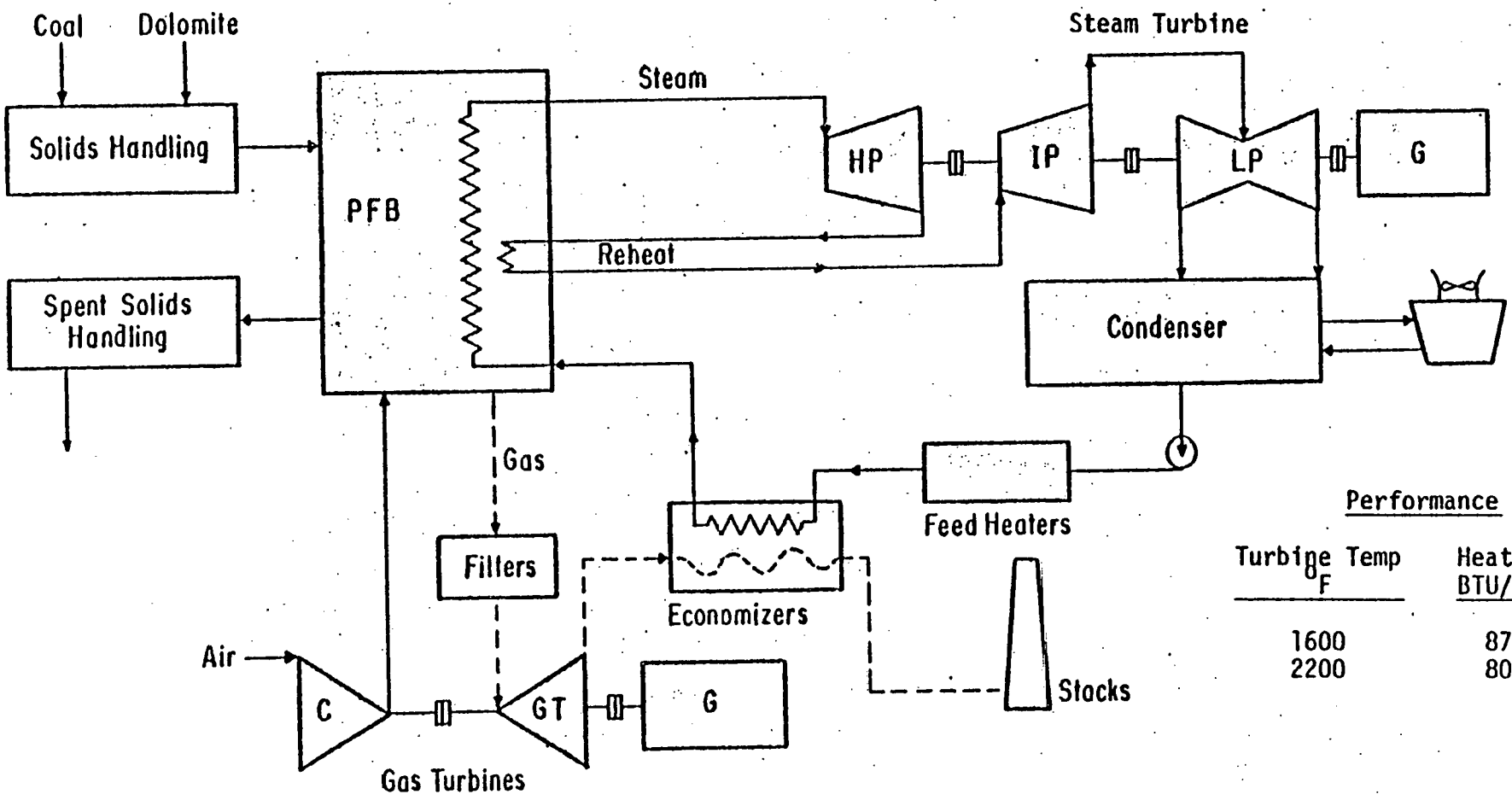
It appears that fuel gas treatment systems may be designed to meet strict environmental emission requirements for particulate, sulfur and nitrogen compounds. This feature makes this power generation fuel approach particularly attractive for advanced power generation systems.

Coal Based Solid Fuels Integrated with Combined Cycles

Fig. 6 indicates a combined cycle approach contemplated for application with direct coal combustion or possibly with "clean" solid fuels processed from coal, such as solvent refined coal.

Direct combustion of coal, or solid processed coal products, under pressurized fluid-bed conditions in a combined cycle configuration offers a potentially attractive clean power approach. In this process, combustion air and pulverized coal is injected at a sufficient velocity to suspend a mixture of inert bed material and pulverized solid fuel, creating a fluid bed. Sulfur may be removed in the bed by the addition of dolomite which reacts with the SO_2 formed during combustion, and the sulfate products discharged with ash from the particulate separators and bed material withdrawal purge. The hot gases are then directed back to the gas turbine for expansion. With uniform, relatively low, bed temperatures, the ash fines carried over with the gas steam have a soft characteristic which is anticipated to greatly reduce turbine bucket erosion and fouling problems evident in previous experience with direct high temperature slagging coal firing in gas turbine combustors. Feedwater is heated and steam generated in the gas turbine exhaust and in the heat exchanger surface buried in the fluid bed. This steam powers a steam turbine, producing about 70% of the plant output, while the gas turbine produces about 30%. The very high heat-transfer rates possible in the

DIRECT COMBUSTION COMBINED
CYCLE -- PRESSURIZED FLUIDIZED BED



Performance

Turbine Temp °F	Heat Rate BTU/KW/HR
1600	8700
2200	8000

Fig. 6

fluid bed offers the potential of reduced boiler surface requirements. Investigation of this concept is proceeding, including the study of particulate carry-over effects on the gas turbine and advanced methods for their removal from the gas stream.

Although there may be maximum gas temperature limitations of 1600 to 1700°F associated with the fluid bed sulfur removal and ash fusion with direct coal combustion, this combined cycle with steam conditions possible in such configurations results in plant heat rates of about 8700 BTU/KW/HR (HHV).

Of course, the major consideration in such direct fired cycles is ash contaminant carry-over to the turbine. The previously discussed areas of corrosion, deposition and emission are particularly crucial in this case.

Since cooling of the combustion gases for cleaning is not possible, considerable advanced development is needed in the relatively new technology area of hot gas particulate cleanup before such systems can achieve the required order of magnitude improvements in particulate cleaning and become commercially practical.

The requirement of cleaning the total combustion gas flow makes the permissible particulate content of~.01 GR/SCF and the alkali metal content of about~0.2 ppm difficult to achieve. This is an area where compromises between turbine component designs and combustion gas cleanup technology will probably be required to realize a practical overall design.

The benefits of advanced turbine component liquid cooling and coating technologies, critical to exploiting higher gas turbine firing temperature capability, will be of direct benefit in utilizing this concept since the resulting lower turbine metal temperatures will minimize possible corrosion effects associated with alkali metals in the hot gas stream.

Utilizing a processed "clean" coal-derived fuel may alleviate some of the problems and constraints with coal by increasing permissible cycle operating temperature, reducing total contaminants to the turbine and elimination of the disposal problems of contaminated bed material.

Since sulfur retention would no longer be required and the "de-ashed" solid fuel would minimize slagging problems, higher permissible bed operating temperatures could open possibilities for cleaning smaller volumes of gases and reduction of contaminant levels by partial air bypass, with subsequent mixing while maintaining acceptable thermal performance.

Without bed temperature limitations, such cycles can achieve heat rates of about 8000 BTU/KW/HR at turbine inlet temperatures of 2200^{o}F. The combustion concept is also applicable for gas turbine combustion systems with conventional waste heat recovery combined cycles.

Summary

Considering the many aspects of the implication of synthetic fuels for combined cycle application, the following points may be summarized:

1) The combined cycle concept offers the application flexibility to utilize the fuels from, and efficiently integrate with, a wide spectrum of synthetic fuel from coal processing alternatives.

2) The characteristics of these various fuels will have a significant impact on the development requirements for new turbine/combined cycle systems and their thermal performance levels.

3) Consideration of the impact of synthetic fuel characteristics on turbine development needs and performance should be factored into the overall economics, development needs, and optimization of processing concepts directed at producing fuels for combined cycle power generation.

4) Depending on the specific nature of the fuel characteristics, long-term turbine development programs may be necessary to obtain a reliable system with improved thermal performance utilizing synthetic fuels from coal.

5) Advancements in turbine cooling technology will enhance the use of these synthetic fuels while allowing a continual improvement in cycle operating conditions. However, specific characterization of the types of fuels to be used for future power generation is definitely required before component cooling, combustion and basic configuration designs can be finalized.

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AN OVERVIEW OF THE U.S. ENERGY DILEMMA

BY

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Princeton, New Jersey

To Be Presented At:

THE THIRD ANNUAL INTERNATIONAL CONFERENCE ON COAL GASIFICATION AND LIQUEFACTION
WHAT NEEDS TO BE DONE NOW!

August 3-5, 1976

UNIVERSITY OF PITTSBURGH
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The purpose of this paper is to assess where we are today in respect to commercializable processes to greatly increase the use of coal to replace petroleum and natural gas and to recommend what should be done from the technological, regulatory, and Government financial assistance standpoints to speed up the program. This is not a matter of Sunday morning quarterbacking. There are serious deficiencies in our coal utilization effort which, in some cases, are apparent merely by the application of common sense. Nor will we merely criticize. We will offer positive recommendations as to how to get something done.

It is most appropriate that we follow on the same program with Dr. McCormick of ERDA who has the heavy responsibility of pushing the early commercialization of such coal conversion processes that are both ready and needed in our natural energy picture.

Our viewpoint is that of a project and venture planner. We have been trying for several years to actually bring together need, technology, and capital to do something in fact. This has included a large scale methanol venture and any number of gasification ventures. Not one has succeeded although serious industrial backers were found who spent a great deal of time and money reaching negative conclusions. In one case, an important backer of a methanol-from-coal project withdrew and went back to petroleum resid. In another

CASE, WE ARE HAPPY TO SAY, A BACKER OF A COAL GAS PLANT PROPOSAL HAS SUBMITTED A CAREFULLY STRUCTURED PROPOSAL TO ERDA FOR A DEMONSTRATION PLANT TO MAKE 240 BTU GAS. THE GAS WOULD GO TO TWO INDUSTRIAL PLANTS AND AN INDUSTRIAL PARK. ANOTHER STUDY SHOWED THAT A COAL GAS PLANT BASED ON ANTHRACITE WAS TECHNOLOGICALLY FEASIBLE IN AN INDUSTRIAL PARK, BUT THE VENTURE WOULD HAVE PROVIDED GAS AT A HIGH PRICE COMPARED TO THE VERY LOW CONTROLLED PRICE OF NATURAL GAS IN AN ADJACENT INDUSTRIAL PARK. SO FAR, THIS SITUATION IS ON DEAD CENTER.

JUST NOW, WE ARE WORKING ON A CASE IN WHICH A LARGE CHEMICAL PLANT WANTS TO CONVERT FROM NATURAL GAS TO COAL FOR FUEL, AND POSSIBLY EVEN FOR FEEDSTOCKS, IF ECONOMICALLY ATTRACTIVE ON A PROJECTED BASIS USING ASSURED TECHNOLOGY. WE ARE LOOKING BOTH AT DIRECT BURNING WITH FLUE GAS SCRUBBING, WHICH MEANS SCRAPPING PRESENT BOILERS, AND AT GASIFICATION TO A MORE EXPENSIVE FUEL BUT WHICH CAN BE USED IN PRESENT BOILERS WITHOUT DOWN RATING.

THUS, WE ARE NOT TALKING FROM A HYPOTHETICAL POINT OF VIEW BUT FROM THE VIEWPOINT OF ONE WHO HAS TRIED SEVERAL TIMES AND FAILED -- AND KEEPS TRYING.

IN THE COURSE OF THIS PAPER, WE MAY APPEAR TO PASS OVER OR SLIGHT A PARTICULAR PROCESS OR APPROACH FOR WHICH WE APOLOGIZE IN ADVANCE. THERE IS SO MUCH THAT COULD BE COVERED WE HAVE HAD TO SELECT ONLY THE FEW CLOSE AT HAND, COMMERCIALLY READY METHODS OF COAL UTILIZATION WHICH WE FEEL ARE READY TO GO NOW GIVEN THE RIGHT GOVERNMENT ACTIONS AT THE REGULATORY AND FINANCIAL ASSISTANCE LEVELS.

WE HAVE CALLED THIS PAPER "AN OVERVIEW OF THE U.S. ENERGY DILEMMA." THE WORD, DILEMMA, IMPLIES THAT WE DON'T KNOW WHICH WAY TO TURN, AND THAT IS EXACTLY WHAT THE SITUATION IS.

IT ALL STARTS WITH THE IRRATIONAL PRICING STRUCTURE FOR NATURAL GAS. TO A LESSER BUT NONE THE LESS IMPORTANT EXTENT, THE CONTROLLED CRUDE

OIL PRICE WORKS AGAINST US IN THE SAME WAY. TO PUT IT BLUNTLY, THE POLITICIANS ARE TRYING EVERY WAY IN THE WORLD TO CONVINCE THE PUBLIC THAT THEY CAN SOFTEN THE ENERGY PRICE BURDEN WHEN ALL THEY ARE ACCOMPLISHING IS TO MAKE MORE CERTAIN AN EVENTUAL LARGE PRICE DISCONTINUITY. LOOK AT THE FIGURES. IT WILL COST $3.75 ± 0.25 PER MM BTU TO PRODUCE METHANE FROM COAL IN THE WEST. THIS WILL DELIVER TO THE HOUSEHOLDER OR INDUSTRY IN THE EAST FOR ABOUT $5.25 ± 0.25 PER MM BTU.

YET, WE PERSIST IN SELLING NATURAL GAS TO INDUSTRY AND FOR HOME HEATING USE AT WELL UNDER $2 PER MM BTU DELIVERED IN THE EAST. IT IS THE GREATEST SATURDAY SPECIAL IN ENERGY IN THE HISTORY OF THE WORLD.

IF WE MAKE 300 BTU GAS FROM COAL IN THE EAST, IT MUST SELL FOR $3.50 ± 0.25 PER MM BTU. WITH NATURAL GAS PRICED AS IT IS, WHO IN INDUSTRY IS GOING TO PICK UP THE DIFFERENCE OF $2 OR SO PER MM BTU? AND, LET US NOT SAY THE NATURAL GAS ISN'T AVAILABLE TO FIRM CONTRACT USERS. (TRUE, INTERRUPTIBLE GAS IS RAPIDLY DISAPPEARING.) IT HAS TO BE AVAILABLE UNDER THE PROTECT-THE-PUBLIC PRICING SYSTEM, OR THE PIPELINE AND DISTRIBUTION COSTS WOULD SKYROCKET SINCE INDUSTRY AND UTILITIES NATIONWIDE USE ROUGHLY 2/3RDS OF THE GAS. WE AREN'T EVEN SAYING THAT THE NATURAL GAS SHOULD SUDDENLY BE DENIED TO ANYONE SELECTIVELY EXCEPT INTERRUPTIBLE CUSTOMERS. WE ARE ONLY SAYING THAT ITS PRICE MUST BE ALLOWED TO FLOAT FREELY. OTHERWISE, THE SUBSIDIES REQUIRED TO BRING ABOUT AN INDUSTRIAL FUEL GAS FROM COAL INDUSTRY, LET ALONE A METHANE-FROM-COAL INDUSTRY, WILL BE SO MASSIVE THAT NO CONGRESS WILL EVER GRANT THEM.

ON THE OTHER HAND, IF NATURAL GAS WERE DECONTROLLED WE PROJECT IT WOULD GO FROM THE PRESENT $2 WELL HEAD LEVEL FOR UNCONTROLLED GAS TO THE $3 - $3.50 LEVEL BY 1980-81. THIS WOULD MEAN A DELIVERED PRICE $4 - $4.50 WHICH WOULD MAKE INDUSTRIAL FUEL GAS FROM COAL QUITE ATTRACTIVE TO INDUSTRY AND TO POWER PLANTS FOR PEAK SHAVING. IT WOULD BACK NATURAL GAS OUT OF BOILERS VERY QUICKLY.

WHEN INDUSTRY IS CURTAILED ON GAS, IT DOES THE NEXT BEST THING. IT SWITCHES TO LOW SULFUR OIL, EITHER #6 OR #2 DEPENDING UPON THE EQUIPMENT IT HAS. THE COST TODAY IS ON THE ORDER OF $2 - 2.50 PER MM BTU DELIVERED BASED ON U.S. CONTROLLED CRUDE OIL PRICES.

WE ARE TODAY IN FACT PREPARED TO MAKE INDUSTRIAL FUEL GAS FROM COAL ON SMALL OR LARGE SCALE. THE INVESTMENT REQUIRED WILL BE ON THE ORDER OF $6 - 12 PER ANNUAL MM BTU, DEPENDING UPON BTU LEVEL FROM 160 TO 300, FOR A 1981 ON STREAM DATE. AS INDICATED ABOVE, THE GAS MUST SELL FOR ABOUT $3.50 ± .25 PER MM BTU. SINCE CAPITAL CHARGES ARE 1/2 TO 3/4 THE COST OF GAS, OBVIOUSLY THERE MUST EITHER BE A LARGE CAPITAL SUBSIDY OR SOME ASSURANCES THAT THE PRICE WILL INDEED BE HIGH ENOUGH TO SUPPORT THE PLANT. IF WE WAIT FOR FURTHER TECHNOLOGICAL PROGRESS, WE ARE KIDDING OURSELVES. THERE IS A CEILING ON THE IMPROVEMENT IN THERMAL EFFICIENCY SET BY THE LAWS OF NATURE, BUT INFLATION CONTINUES ON A COMPOUND BASIS. THE NET EFFECT IS AN EVEN GREATER DISPARITY BETWEEN THE COST OF MANUFACTURED GAS AND THE RIGGED PRICE OF NATURAL GAS AND PETROLEUM.

IT IS QUITE REASONABLE TO PROJECT PETROLEUM PRICES BY 1981 WHICH WOULD MAKE MANUFACTURED GAS COMPETITIVE, BUT WHAT INVESTOR WILL GAMBLE TH- ENORMOUS SUMS REQUIRED ON THE HOPE THAT SUCH PROJECTIONS ARE CORRECT OR THAT THE PRICES WILL NOT BE RIGGED AGAINST HIM BY ONE GOVERNMENT OR ANOTHER?

WHAT INDEED IS THE U.S. ENERGY DILEMMA? TO PUT IT IN SIMPLEST TERMS, WE ARE SUFFERING FROM THE DELUSION THAT WE HAVE GREAT ENERGY SUPPLIES IN THE FORM OF COAL THAT SOMEHOW WILL BAIL US OUT WHEN WORST COMES TO WORST. THAT IS THE PREVAILING PUBLIC AND CONGRESSIONAL VIEWPOINT, AND I'M AFRAID IT IS SOMEWHAT NUTURED BY THE MANY SPEECHES AND PAPERS WE ALL GIVE. THE PLAIN FACT OF THE MATTER IS THAT COAL IS ALMOST NO DAMN GOOD FOR ANYTHING OTHER THAN MAKING STEEL OR BURNING IT TO RAISE STEAM FOR POWER GENERATION. BUT, IF THE EPA ENFORCES THE LAWS AND REGULATIONS ON SCHEDULE, WE CAN CROSS OFF MUCH OF THE DIRECT BURNING OF COAL. IN OUR GOVERNMENT ACTIONS TO DATE, WE HAVE SIMPLY REFUSED TO RECOGNIZE THAT

IF COAL IS GOING TO BAIL US OUT EVEN PART WAY ON OUR NEEDS FOR GROWTH ENERGY, THEN WE NEED TO PUT FIRST THINGS FIRST AND DO SOME OF THEM IN PARALLEL WITH RESEARCH, NOT IN SERIES WITH NEVER ENDING RESEARCH AND DEVELOPMENT AND DEMONSTRATION.

WE NEED TO LEARN TO PROCESS COAL TO PRODUCE A CLEAN COMBUSTION RESULT AT COSTS WHICH WILL ALLOW US TO COMPETE WITH PROJECTED PRICES OF PROCESSED PETROLEUM. THERE IS NO HOPE AT ALL TO COMPETE WITH OIL AT TODAY'S PRICES PARTICULARLY IN THE CONTROLLED PRICE ATMOSPHERE IN THE U.S. OF THESE COSTS, THE LIMIT ON LOSS OF ENERGY DURING PROCESSING IS SET BY THE LAWS OF THERMODYNAMICS, NOT BY MAN, AND WHILE WE CAN APPROACH THIS LIMIT MORE CLOSELY, I DOUBT IF THE DIFFERENCE BETWEEN CURRENT ART AND THE THEORETICAL MINIMUM LOSS WILL EVER BE CRITICAL IN OUR DECISION FOR OR AGAINST COAL. DIRECT LABOR COSTS IN LARGE FUEL PROCESSING PLANTS ARE NOT REALLY SIGNIFICANT, AND WE HAVE A SURPLUS OF LABOR FOR THE FORESEEABLE FUTURE. THIS LEAVES US WITH CAPITAL RELATED CHARGES AS THE REAL BUGABOO. THESE ARE:

DEPRECIATION
RETURN BEFORE FEDERAL TAX
MAINTENANCE
LOCAL TAXES AND INSURANCE

THE HISTORY OF THE PROCESS INDUSTRIES HAS SHOWN US OVER AND OVER AGAIN THAT THE REAL WAY TO REDUCE CAPITAL CHARGES IS TO OPERATE COMMERCIAL SCALE PLANTS. RESEARCH MERELY SHOWS US HOW TO GET STARTED; IT DOES NOT SHOW US HOW TO DO ANYTHING WELL COMMERCIALLY. DOING THINGS REALLY WELL COMMERCIALLY COMES FROM HARD EXPERIENCE ON THE LARGEST SCALE WE CAN ACHIEVE. UNLESS WE GET STARTED WITH MORE AMBITIOUS COMMERCIAL COAL PROCESSING PROJECTS SOON, WE WILL FIND THAT OUR TARGET IS ALWAYS MOVING JUST BEYOND OUR REACH SUCH THAT MORE AND MORE RESEARCH WILL BE THOUGHT NECESSARY WHICH WILL PUSH COMMERCIALIZATION EVEN FARTHER INTO THE FUTURE.

COMMON SENSE WOULD SAY THAT DIRECT BURNING OF COAL WITH FLUE GAS DE-SULFURIZATION IS THE CHEAPEST AND QUICKEST ANSWER. THIS HAS NOT TURNED OUT TO BE AS CLEAR CUT AS MANY HAD ASSUMED AND PROBABLY FOR THESE

REASONS:

(1) THE TECHNICAL PROBLEM SOLVING HAS BEEN UNBELIEVABLY FRAGMENTED HAVING BEEN GIVEN TO A REGULATORY AGENCY, NOT AN ENERGY TECHNOLOGY DEVELOPMENT AGENCY LIKE ERDA. WE HAVE DOZENS OF PROCESSES AND NO CONSENSUS ON TWO OR THREE GOOD THROW-AWAY PROCESSES NOR ANY CONSENSUS ON ONE OR TWO GOOD REGENERATIVE PROCESSES. WE SHOULD HAVE HAD THESE BY NOW. EVIDENTLY WE HAVE CONSTANTLY EXPANDED THE PROBLEM RATHER THAN CONCENTRATING ON ITS SOLUTION BY A FEW REASONABLE METHODS.

(2) THE PROBLEM WAS HARDER THAN WE THOUGHT.

(3) APPARENTLY, FLUE GAS DESULFURIZATION ON A SCALE BELOW ABOUT SEVERAL HUNDRED MW (700,000 POUNDS OF STEAM PER HOUR IS ABOUT 100 MW) IS NOT COMPETITIVE WITH RELATIVELY LOW PRICED, LOW SULFUR WESTERN COAL OR PREMIUM PRICED, LOW SULFUR PETROLEUM FUELS WHERE THESE CAN BE USED. WE SAY APPARENTLY BECAUSE NO ONE SEEMS REALLY TO KNOW, AND INDUSTRY (AS OPPOSED TO UTILITIES) ISN'T GENERALLY OFFERING TO BE THE GUINEA PIG.

(4) THROW AWAY PROCESSES ARE NOT A SATISFACTORY LONG-TERM SOLUTION AND COULD WELL CREATE IN THE LONG RUN AS MANY PROBLEMS AS THEY SOLVE.

(5) THERE HAS BEEN TOO LITTLE CONCENTRATION OF GOVERNMENT HELP ON INDUSTRIALLY DEVELOPED REGENERATIVE PROCESSES THAT HOLD THE POTENTIAL OF A COMPLETE LONG-TERM SOLUTION TO DIRECT BURNING OF COAL WITH RECOVERY OF EVERY POUND OF SULFUR IN USABLE FORM.

IT IS RECOMMENDED THAT ERDA TREAT THIS CASE AS ANY OTHER FUEL CONVERSION ALTERNATIVE AND OFFER INCENTIVES TO INDUSTRY FOR THE COMMERCIALIZATION OF A SOUND, DRY, OR WET, FULLY REGENERATIVE PROCESS

SUCH AS THE UOP-SHELL (DRY) OR THE CATALYTIC-IFP OR THE STAUFFER PROCESSES WHICH ARE BOTH WET. EPA IS FINALLY BACKING THE WELLMAN POWER GAS WET REGENERATIVE PROCESS. AFTER ALL, THE OPERATION OF SUCH A COMPLICATED CHEMICAL PROCESS IS JUST ANOTHER WAY TO CONVERT COAL TO CLEAN FUEL. THE FACT THAT A BOILER IS PART OF THE SYSTEM IS NEITHER HERE NOR THERE. WE SHOULD LOOK AT THE CAPITAL INVESTED PER ANNUAL MILLION BTU OF INPUT ENERGY TO THE BURNER, AS THIS IS EQUIVALENT TO THE OUTPUT OF A CLEAN FUELS PLANT, AND WE SHOULD LOOK AT THE OVERALL COST PER MM BTU FIRED. THESE TWO FIGURES FOR THROW AWAY PROCESSES ARE RESPECTIVELY ABOUT $0.6 TO 1.2 PER ANNUAL MM BTU AND 40 - 60¢ PER MM BTU. IT CAN READILY BE SEEN THAT THESE FIGURES, IF ANYWHERE NEAR CORRECT, ARE MORE FAVORABLE THAN ANY LIKELY TO BE ACHIEVED BY FIRST PROCESSING COAL TO A LOW SULFUR FUEL AND THEN BURNING IT. TO MAKE COAL INTO GAS, WE MUST FIRST INVEST $6 - 8 PER ANNUAL MM BTU. THE PROBLEM IS INDUSTRY IS NOT GOING YET TO FLUE GAS SCRUBBING, EITHER THROW AWAY OR REGENERATIVE. UNTIL THESE PROCESSES ARE USED COMMERCIALLY ON WIDE SCALE BY INDUSTRIAL PLANTS, NO ONE WILL BELIEVE ANY FIGURES. UTILITY EXPERIENCE IS NOT ENOUGH.

COAL BENEFICIATION, WHICH IN MANY CASES COULD BE THE CHEAPEST ALTERNATIVE OF ALL, HAS GOTTEN VERY LITTLE ATTENTION. IT HAS MERIT AS IS AND IN SERIES OR SERIES-PARALLEL WITH OTHER ALTERNATIVES. FOR EXAMPLE, ONE CUT COULD GO TO POWER BOILERS AND THE OTHER CUT TO GASIFICATION OR TO A FLUID BED BOILER WHICH CAN BURN NEARLY ANYTHING.

VERY FORTUNATELY, ERDA HAS PUSHED HARD ON THE FLUID BED BOILERS WITH LIMESTONE ADDITION TO REMOVE SULFUR WHICH IS ANOTHER WAY TO BURN COAL DIRECTLY. THIS TECHNOLOGY IS SO NEAR COMMERCIAL THAT WE ARE WILLING TO CALL IT CERTAIN AS TO OPERABILITY AND AS TO A LOWER OVERALL COST THAN A CONVENTIONAL BOILER WITH FLUE GAS SCRUBBING. WE, THEREFORE, RECOMMEND THAT INDUSTRIAL FLUID BED BOILER INSTALLATIONS BE CONSIDERED IN THE ERDA COMMERCIALIZATION PROGRAM RIGHT ALONG WITH FUEL CONVERSION PROCESSES. WHY NOT? THE FLUID BED BOILER WOULD SOLVE THE PROBLEM WE'RE TRYING TO SOLVE, I.E. MORE USE OF COAL DIRECTLY. BUT UNTIL SOME INDUSTRIAL PLANT PROVES IT COMMERCIALLY AND COMPETITIVELY, NO ONE WILL

LISTEN. THE FLUID BED BOILER COULD TURN OUT TO BE THE SINGLE, MOST IMPORTANT DEVELOPMENT FOR THE NEXT DECADE OR SO FROM THE STANDPOINT OF VOLUME OF COAL USED BY INDUSTRY AS OPPOSED TO UTILITIES, SURPASSING EVEN FLUE GAS SCRUBBING WITH ALL OF THE RETROFITTED AS WELL AS NEW INSTALLATIONS ON INDUSTRIAL BOILERS.

NEXT, WE COME TO INDUSTRIAL FUEL GAS. WE HAVE ALREADY LAID THE BACKGROUND IN OUR DISCUSSION OF THE NEED TO DEREGULATE NATURAL GAS AND CRUDE OIL PRICES TO GIVE INDUSTRIAL FUEL GAS A CHANCE TO COMPETE. THIS WILL NOT BE DONE QUICKLY ENOUGH YOU MAY BE SURE, SO WE ARE FACED WITH THE SUBSIDY APPROACH IF THE JOB IS TO GET DONE. LOOK AT THE BARE FACTS. AN IFG PLANT OF LARGE SIZE WILL COST $6 - 8/ANNUAL MM BTU IN CAPITAL. INDUSTRY WILL WANT AT LEAST 10% AFTER TAX AND AT LEAST 7% OR SO PER YEAR DEPRECIATION. THIS MEANS A CAPITAL CHARGE FACTOR OF 27% BEFORE TAX, OR $1.6 - 2.2/MM BTU OF OUTPUT. TO THIS, WE CAN ADD COAL AT ABOUT $1 PER MM BTU DELIVERED TO INDUSTRIAL AREAS IN 1980 REGARDLESS OF THE COAL. AT 75% GASIFICATION, EFFICIENCY COAL COST IS $1.3/MM BTU. OTHER COSTS WILL RANGE $0.5/MM BTU FOR SMALL PLANTS DOWN TO $0.3 FOR LARGER ONES:

CAPITAL	$1.6 - 2.2
COAL	1.3
OTHER	0.3 - 0.5
	$3.2 - 4.0

IF THESE FIGURES LOOK HIGHER THAN WHAT YOU HAVE SEEN, RECALL THIS IS INDUSTRIAL, NOT UTILITY, ECONOMICS WITH NO GIMMICKS. THE RETURN IS SET SO LOW THAT LEVERAGE DOESN'T HELP THAT MUCH. EVEN WITH ALL THE GIMMICKS WE CAN THINK OF, THE COST ALWAYS SEEMS TO COME IN THE RANGE OF $3.5 ± 0.25/MM BTU. WHO IS COURAGEOUS ENOUGH TO START A PLANT TODAY WITH PRIVATE CAPITAL WITH NO GOOD IDEA WHERE 1% S FUEL PRICE WILL BE IN 1980, LET ALONE WITHOUT HIS OWN COAL SUPPLY? IF, FOR EXAMPLE, UNCLE SAM WOULD PICK UP HALF THE CAPITAL BY SUBSIDY AND GUARANTEE THAT THE TOTAL TRANSFER OF GAS WOULD NOT EXCEED 10% OVER THE PRICE OF LOW S #6 FUEL, THERE WOULD BE PLENTY OF TAKERS.

SUCH PLANTS CAN BE BUILT RIGHT NOW BY THE WINKLER OR THE KOPPERS TOTZEK PROCESS. THE FORMER IS PROBABLY OPTIMUM FOR LIGNITE AND

WESTERN COAL AND THE LATTER FOR EASTERN COALS. BUT ONCE INSTALLED COMMERCIALLY, THERE WOULD BE A SWIFT RACE TO IMPROVE EACH PROCESS TO USE ANY COAL AND GENERALLY SURPASS THE OTHER AND THUS GET THE LION'S SHARE OF FUTURE PLANTS. THIS IS A COMPETITIVE PROCESS WE MUST GET STARTED IN THIS COUNTRY.

NOTE THAT WE ARE SILENT ON METHANE FROM COAL. THIS IS FOR TWO REASONS. FIRST, WE ARE NOT WISE ENOUGH TO SUGGEST WHEN SUCH PLANTS SHOULD BE BUILT OR BY WHAT PROCESS. IF WE FOLLOW SENSIBLE NATURAL GAS POLICIES, WE CAN EASILY BUY THE TIME TO USE 2ND GENERATION TECHNOLOGY VIA A TIME CONSUMING BUT BENEFICIAL DEVELOPMENTAL SEQUENCE OF STEPS INTERMEDIATE TO A FULL SCALE COMMERCIAL PLANT. SECOND, WE ARE CONCERNED MAINLY WITH THE PROBLEMS OF INDUSTRY AND UTILITIES WHO ARE THE REALLY LARGE FUEL USERS. WE ARE NOT WILLING TO BELIEVE THAT WE ARE GOING TO SYNTHESIZE METHANE FROM COAL IN ORDER TO REFORM IT FOR AMMONIA OR TO BURN IT TO RAISE STEAM. IF METHANE IS SYNTHESIZED, IT WILL BE PURELY TO SUPPORT OUR EXTENSIVE DISTRIBUTION SYSTEM TO HOUSEHOLDS AND COMMERCIAL ESTABLISHMENTS WHICH TODAY CONSUME TOGETHER ABOUT 1/3 OF OUR NATURAL GAS. THERE IS EVERY REASON TO SUPPORT THIS MARVELOUS DISTRIBUTION SYSTEM INDEFINITELY AS IT PROBABLY REPRESENTS THE CHEAPEST WAY WE WILL EVER DEVISE TO GET ENERGY TO THE SMALL CONSUMER.

WE HAVE SUGGESTED HOW NATURAL GAS CAN BE BACKED OUT OF INDUSTRIAL AND UTILITY USE BY GAS FROM COAL IF THE POLITICIANS WILL ALLOW NATURAL GAS TO SEEK ITS TRUE VALUE PRICE LEVEL. WE HAVE SUGGESTED HOW COAL MIGHT BE BURNED DIRECTLY IN NEW INSTALLATIONS USING THE FLUID BED BOILER AND THAT WE MAY BE CLOSE TO SOME ANSWERS FOR FLUE GAS CLEAN UP ON BOTH A RETROFIT AND NEW PLANT BASIS IF THE 2ND GENERATION REGENERATIVE PROCESSES CAN BE ENCOURAGED. THIS LEAVES US IN DIRE NEED OF:

(1) A CLEAN LIQUID OR SOLID BOILER FUEL FROM COAL TO USE WHERE RETROFITTING IS IMPRACTICAL OR WHERE WESTERN COAL CANNOT BE OBTAINED ECONOMICALLY

AND

(2) A LIQUID FUEL, HIGH SULFUR OR NOT, FOR REPLACING HEAVY PETROLEUM FUEL IN THE MANUFACTURE OF CHEMICALS MADE VIA HYDROGEN AND/OR SYNTHESIS GAS FROM PARTIAL COMBUSTION.

BOTH OF THESE OBJECTIVES ARE WELL WITHIN OUR GRASP AS WILL BE DESCRIBED IN THE PAPER(1) BY DR. L. H. UDANI AND DR. C. A. STOKES LATER IN THE PROGRAM.

SUFFICE IT TO SAY HERE THAT ONE OF THE OLDEST COAL CONVERSION TECHNOLOGIES WE HAVE IS THE SIMULTANEOUS, HIGH PRESSURE, NON-CATLAYTIC LIQUEFACTION AND HYDRODESULFURIZATION OF COAL IN THE PRESENCE OF A RECYCLED SOLVENT MADE IN THE PROCESS. THIS PROCESS, DESCRIBED IN THE PAPER BY UDANI ET AL, IS ALREADY TO USE RIGHT NOW IF THE PROJECT COULD BE GIVEN THE RIGHT KIND OF FINANCIAL ASSISTANCE. IT TAKES HIGH SULFUR COAL, PROJECTED TO COST AROUND $0.80 -1.00/MM BTU, MINE MOUTH BY 1980 - 81 AND TURNS IT INTO A LOW SULFUR, LOW ASH, BOILER FUEL AND INTO A 300 BTU CO/H_2 MIXTURE. THE FORMER CAN GO ACROSS THE FENCE TO A LARGE UTILITY, THE LATTER IS PARTLY USED FOR THE PROCESS HYDROGEN REQUIREMENTS AND THE REST EITHER FOR CHEMICALS OR INDUSTRIAL FUEL GAS WITH PARTIAL COMBUSTION UNDER 800 PSIG + AS THE KEY CONVERSION STEP. THE ONLY OTHER PRODUCT IS PURE SULFUR AND, OF COURSE, ASH FOR DISPOSAL. WHEN WE SAID READY TO USE, WE MEAN THAT LARGE, RELIABLE ENGINEERING COMPANIES ARE WILLING TO BUILD COMMERCIAL PLANTS. OBVIOUSLY, NO ONE IS GOING TO MAKE TIGHT GUARANTEES ON THE PROCESS AS THE PROCESS TODAY BELONGS LARGELY TO THE PUBLIC. THERE IS NO INCENTIVE TO DO SO. BUT, THE ENGINEERING COMPANIES WILL GUARANTEE A WORKABLE PLANT AND PROBABLY SOME MINIMUM OPERATING FACTOR AND CONVERSION EFFICIENCY.

WE WILL NOW TRY TO SUMMARIZE WHAT WE HAVE RECOMMENDED IN A SERIES OF 4 SLIDES. IN A SHORT PRESENTATION OF THIS KIND, WE CAN'T POSSIBLY BACK UP THESE RECOMMENDATIONS WITH FACTS AND FIGURES. BUT, SUFFICE IT TO SAY, WE HAVE BEEN POURING OVER SUCH FIGURES FOR SEVERAL YEARS TO BE ABLE TO REACH THESE RECOMMENDATIONS.

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(1) "NEW CONCEPT FOR THE MANUFACTURE OF LOW SULFUR FUELS AND CHEMICALS FROM COAL."

IN FINAL SUMMARY, WITH THE VERY FEW APPROACHES DESCRIBED, WE COULD UNLOCK ENOUGH COAL USE TO EXCEED OUR ABILITY TO PRODUCE THE COAL. WE DON'T NEED TO MAKE SYNTHETIC CRUDE OIL OR GASOLENE OUT OF COAL RIGHT NOW, OR TO MAKE VAST AMOUNTS OF METHANE IN ORDER TO SOLVE OUR MAJOR PROBLEM WHICH IS TO REDUCE CRUDE OIL IMPORTATION WITH ITS CONSEQUENT STRAINS ON OUR SECURITY AND SOLVENCY. WE ONLY NEED TO BURN COAL DIRECTLY TO MAKE STEAM, TO GASIFY COAL FOR INDUSTRY'S CRITICAL FUEL AND FEEDSTOCK NEEDS AND TO MAKE A LOW S, LOW ASH BOILER FUEL FROM COAL WITH METHANOL AND/OR HIGH PRESSURE SYNTHESIS GAS AND/OR INDUSTRIAL FUEL GAS AS A CO-PRODUCT TO USE THE DREGS OF THE LIQUEFIED BARREL OF COAL. THE OTHER OBJECTIVES OF MOTOR FUEL AND METHANE ARE VERY WORTHY OF DEVELOPMENT, BUT LARGE SCALE COMMERCIAL PRODUCTION IS ANOTHER MATTER. WE MAY NEED THEM EVENTUALLY, AND WE SHOULD HAVE THE BEST POSSIBLE TECHNOLOGY TO DO IT. BUT, RIGHT NOW, WE MUST CONCENTRATE ON A FEW REASONABLE, ATTAINABLE COMMERCIAL OBJECTIVES, AND GET ON WITH ACCOMPLISHING THESE COMMERCIALLY NOT JUST FOREVER TALKING AND RESEARCHING WHILE THE TAXI METER OF INFLATION TICKS AWAY AT 5 OR 6% PER YEAR OR MORE.

COMPARISON OF CONVENTIONAL STACK GAS SCRUBBING PROCESSES

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COMPARISON OF CONVENTIONAL STACK GAS SCRUBBING PROCESSES

As is usual with a new technology, a great many routes have been followed in trying to develop viable processes for flue gas desulfurization. The number of projects that have been carried to the relatively expensive level of pilot-plant testing is remarkable, due in part, no doubt, to the pressure exerted by regulatory authorities and the resulting commercial promise that process developers have generally thought they could see in the field.

Although this profusion of methods still exists to a considerable extent, a conventional technology is clearly emerging. The trend has been in the direction of process simplicity, use of readily available raw materials, and production of a waste product rather than one for sale.

The process that fits in with these trends best is lime-limestone scrubbing. The flowsheet is relatively simple, limestone is a prevalent natural material, and the product, although undesirable in some respects, can be disposed of without the problems involved in creating and maintaining a market. These factors have led to wide adoption, to the extent that most utilities in the U. S. choose lime-limestone when it becomes necessary to select an FGD process.

This trend has become so pronounced that conventional technology can be said to be made up of only one process, and that process differences come about only because process developers follow different routes in trying to solve the several unit problems encountered at various points in the lime-limestone flowsheet. The present paper identifies these problems and evaluates the several approaches used for solving them.

Process Variations in Lime-Limestone Scrubbing

The basic lime-limestone flowsheet shown in Fig. 1 is quite simple. The flue gas (after particulate removal) passes through a scrubber where it contacts a circulating slurry of lime or limestone, converting the absorbent to solid calcium sulfite and sulfate. A portion of the slurry is drawn off and pumped to a pond where the solids settle out; the supernatant liquor is recycled.

The system installed by Montana Power at the Colstrip station probably comes as close as any to this basic design; the only major difference is omission of the dust removal step before the scrubber. Most of the other installations, although following in general the basic design, have significant variations.

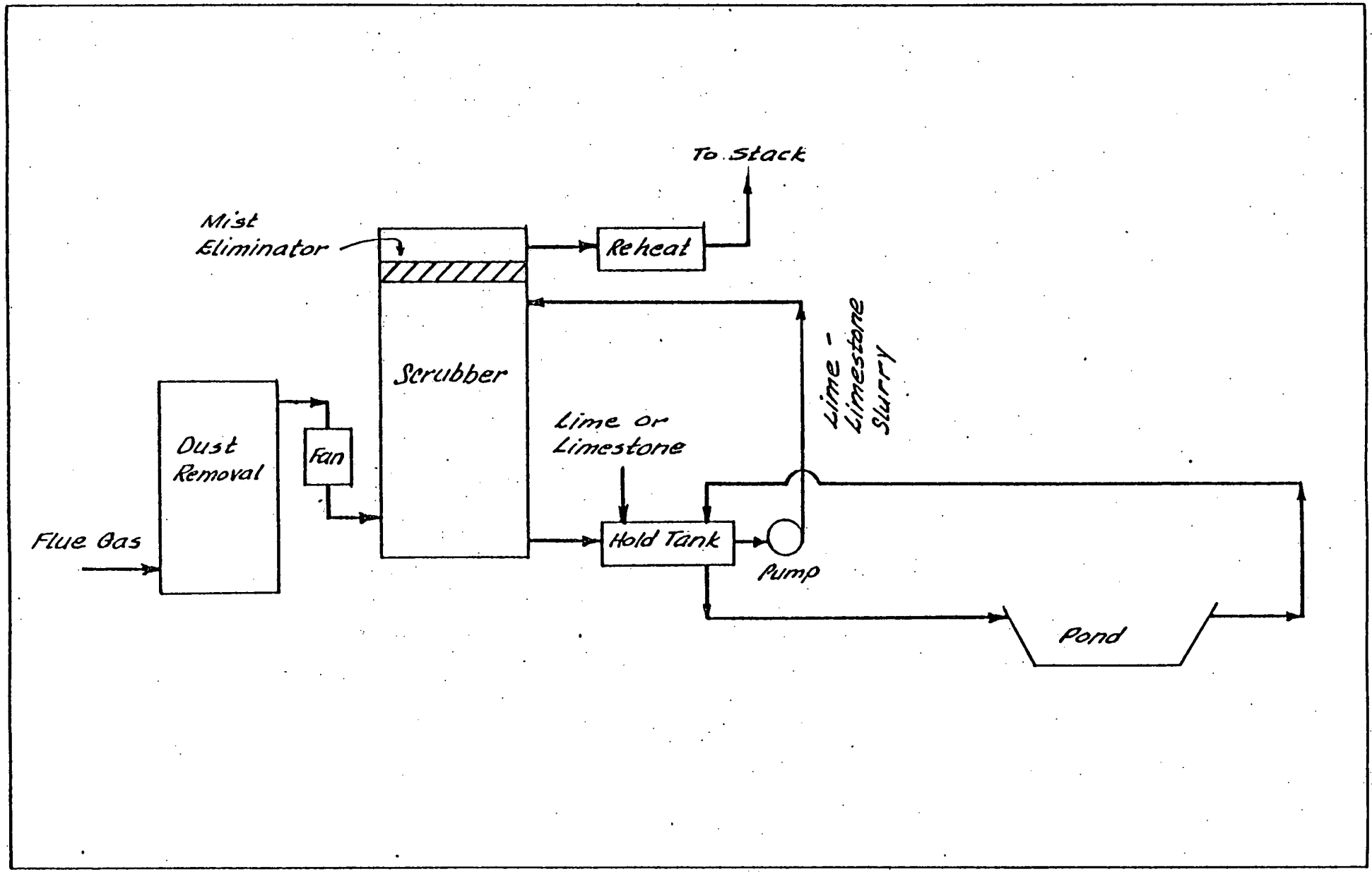


Figure 1

Lime - Limestone Scrubbing (Simplified Flow Sheet)

1. LaCygne (Kansas City Power and Light) and Will County (Commonwealth Edison) have two scrubbers in series, the first a venturi and the second a sieve-tray tower. These systems, 820 and 170 mw respectively, use limestone as the absorbent and were installed by Babcock and Wilcox (B and W).

2. Cholla (Arizona Public Service; 115-mw limestone system installed by Research-Cottrell) also has two scrubbers (venturi followed by packed tower) and in addition the main scrubber is arranged to provide two separate circulation loops.

3. Paddy's Run (Louisville Gas and Electric) is a 65-mw test installation designed by Combustion Engineering. The absorbent is lime and the single scrubber (per train) is a mobile bed type (glass spheres). The sludge is dewatered before disposal by a thickener followed by a filter.

4. Mansfield (Penn Power; 830-mw lime installation by Chemico) has two venturis in series with a wet fan between. There is no dust removal before the scrubbers and the scrubber loop has no hold tank. The sludge is dewatered in a thickener before disposal.

5. Phillips and Elrama. These Duquesne Light stations have several Chemico single-venturi trains that were installed for particulate removal only but now are operating with lime addition for SO_2 removal also. All have precipitators ahead of the scrubber and both plants dewater in thickeners (plus a filter at Elrama).

6. Mohave (170-mw test system installed by a group of southwestern utilities including Southern California Edison and Salt River Project) had a single mobile-bed scrubber (TCA; hollow plastic balls) designed by UOP. A thickener was used instead of the pond, the underflow was further dewatered by filtration, and the sludge was stabilized by the IUCS process.

7. Mohave also had a test system (170-mw) in which a horizontal spray scrubber developed by Southern California Edison (now offered by Pullman Kellogg) was tested. This system also used a thickener; the underflow was stabilized by the Dravo process. The Mohave scrubbers (TCA and spray type) were operated both with lime and limestone.

8. Colstrip (Montana Power; 360-mw lime-ash system installed by CEA) has only a single venturi to remove both ash and SO_2.

9. St. Clair (Detroit Edison; 180-mw limestone installation by Peabody) has a vertical spray tower preceded by a venturi.

10. Sherburne County (Northern States; 680-mw system installed by Combustion Engineering) has two scrubbers in series per train, the first a venturi and the second a mobile bed (glass spheres). The absorbent is ash-limestone.

11. Green River (Kentucky Utilities; 65-mw lime system by American Air Filter) uses a special scrubber with entrained packing (rubber balls) preceded by a venturi.

12. Duck Creek (Central Illinois Light; 100-mw limestone scrubber designed by Riley Stoker) has a "rod-packed" scrubber, plus the unique distinction that the mist eliminator is outside the scrubber in the exit duct.

Although there are other installations operating or in the design stage, these appear to be the more significant ones from the standpoint of operating experience and will be used as the basis for the analysis presented in this paper.

Particulate Removal

One of the main design problems is deciding how much dust (fly ash), if any, to remove in the scrubber. One of the factors in this is that the first scrubbers installed in the U. S. were designed for particulate removal only. This turned out to be much more difficult than expected, to the extent that the many troubles encountered gave scrubbers a bad reputation as dust collection devices. There is growing evidence, however, that if designed properly scrubbers can serve the double function without the earlier troubles. For example, Colstrip seems to be working well (full particulate removal), although the low-sulfur coal makes the results questionable for the eastern situation. For the latter, Mansfield (also full particulate removal) has not been operating long but appears to have no problems directly associated with particulate removal. Perhaps a better example is the EPA-TVA-Bechtel test facility at Shawnee, where the full dust loading is accepted in two different types of scrubbers. After some initial troubles, operation now seems to be acceptable.

All the other systems listed above have some type of dry dust collection equipment ahead of the scrubber. The amount of particulate left for the scrubber to remove varies widely, from practically none (highly efficient precipitator) to as much as half the total loading (e.g., Widows Creek system being installed by TVA).

Removal of particulate as well as SO_2 is a worthy objective because the investment is much lower than for the combination of a highly efficient precipitator followed by an SO_2 scrubber. In some of the recent system planning a compromise has been used, installing a not-so-efficient

precipitator (say 90-95%), or even a cyclonic type, and designing the scrubber to take care of the remainder.

Wet-Dry Interface

The evaporation that takes place where the hot gas first contacts the scrubber slurry can produce troublesome deposits. One of the primary reasons for the "pre-scrubber" used by some designers is to separate the problem from the main scrubber, humidifying the gas in the first unit in such a way as to avoid deposits. A venturi-type prescrubber is normally used.

Although this approach has been fairly effective, addition of the prescrubber complicates the system and adds cost. An acceptable alternative is to spray slurry into the inlet duct to the main scrubber (or use sootblowers), which has been done successfully at Mohave (TCA unit) and at Shawnee.

Scaling

Since calcium sulfate (gypsum) has a long history of causing scaling wherever it has been involved in chemical processes, it would not be expected that lime-limestone scrubbing would be an exception. For some time scaling was considered to be the main problem in lime-limestone systems and although the situation has improved somewhat, designers still must give careful attention to avoiding scale deposits that can accumulate and finally make the system inoperable.

There are three major approaches to preventing scaling.

1. Controlled precipitation. A time-honored way of avoiding scale formation is to make the dissolved salt precipitate on crystals in a controlled way, and usually in a vessel other than the one subject to scaling. This is common practice in crystallizers and concentrators, such as in the phosphoric acid industry, and the English applied it to lime-limestone scrubbing some 30 years ago. Three design factors are involved: (1) high recirculation rate to minimize crystallization in the scrubber, (2) high concentration of sulfate crystals to provide crystallization sites, and (3) extended retention time in the hold tank to cause precipitation outside the scrubber.

   This approach, the one generally taken in current systems, has been reasonably effective. Most of the installations are operating with no significant scaling in the scrubber proper, even at LaCygne where scale deposits were a severe problem in early

operation. The main exceptions are Phillips and Four Corners, where lack of hold tanks and low slurry solids content are the probable reasons for scrubber scaling.

2. Use of Mg^{++}. Another well-established remedy is to carry dissolved magnesium salts in the scrubber loop. Some 25 years ago, the National Lime Association recommended use of dolomitic lime for operations such as acidic waste water neutralization where calcium sulfate scaling was a problem.

   The use of Mg^{++} to control scaling is just coming into the FGD picture. Dravo is contributing by installing facilities to produce dolomitic lime for use in the Ohio Valley. Tests at Phillips with MgO addition resulted in dissolving scale that had already formed, and at Mansfield the use of lime containing MgO has apparently prevented scale in a system of such design that heavy scaling would be expected.

   The situation in regard to limestone scrubbing remains uncertain. Pullman Kellogg has reported good results from adding Mg^{++} in pilot-plant work but large scale tests have not been made in the U. S. In Japan, however, Kawasaki Heavy Industries has been using magnesium-promoted scrubbing for several years and has two full-scale installations in which the absorbent is magnesium-promoted limestone.

   Use of Mg^{++} to control scaling will probably increase in the future, although there are areas where dolomitic limestone is not readily available. A magnesium material (MgO, $MgSO_4$) could be purchased separately in such situations but would complicate supply and storage problems.

   One drawback to use of magnesium is the adverse effect on crystal size, which makes dewatering more difficult.

3. Unsaturated sulfate operation. EPA work at the Research Triangle Park laboratories has shown that if oxidation of sulfite in the scrubber loop can be held to about 30% or lower, the solution stays unsaturated in calcium sulfate because all the sulfate formed precipitates in the sulfite crystals. Thus no gypsum can precipitate and cause scaling.

   The unsaturated sulfate mode does not seem to have played much of a part yet in system design, mainly because oxidation is not very amenable to design factors. The amount of excess air in the flue gas usually depends on factors over which the scrubber designer has no control. Steps can be taken to reduce oxidation

in the scrubber loop, such as protecting equipment from air and designing hold tanks for plug flow, but these seem to have received little attention. Thus unsaturated operation may be obtained incidentally, as a result of low oxygen in the gas or presence of magnesium in the lime, but the tendency is to use controlled precipitation when this fortunate circumstance of unsaturation is not involved.

4. Forced oxidation. Another facet of the scaling problem, and one that indicates the frontier nature of this design area, is use of forced oxidation to prevent scaling--as opposed to inhibition of oxidation as discussed above. Oxidation systems have been in use in Japan for some time and the subject has been studied by EPA at RTP. This approach, like several others in lime-limestone scrubbing, is inter-related with other problem areas such as limestone utilization and sludge properties (see later).

   In the U. S., Combustion Engineering has studied forced oxidation in limestone operation (at Paddy's Run) and has installed it at Sherburne County. Since the system only started up recently, good data are not yet available.

Removal Efficiency

It is not difficult to obtain any desired SO_2 removal efficiency if enough mass transfer is designed into the system and if some excess of absorbent is used. However, it is quite difficult to assess the trade-offs and get the most economical design and operation. It should be noted that the operating installations are not much help in this respect; because of the long construction period (usually at least 4 years from initial decision to start-up), all of them were designed years ago and therefore are somewhat obsolete. The new systems under construction are probably more efficient but this will not be known with certainty until they operate.

The main factors affecting removal efficiency are as follows.

1. Scrubber type. Scrubber designers have a difficult choice between the simpler and less expensive types of scrubbers (also less prone to plugging) and, on the other hand, types with internal elements that promote mass transfer but also cost more, are more susceptible to plugging, and more difficult to maintain.

   The best examples of the more efficient types are the mobile bed (CE and UOP), entrained packing (AAF), and fixed packing /Research-Cottrell and MHI (Japan)/. Of these, CE has offered another type recently, there is the problem of ball wear with UOP (although

progress has been made), the AAF scrubber is somewhat complicated internally, and the fixed packing is especially susceptible to plugging. The MHI design appears superior in this respect because the gas flow is downward (cocurrent with the liquor) and thus tends to keep the packing clean.

The venturi is to some extent in a class by itself. Originally intended for particulate removal only, it is relatively simple and has some advantage in regard to mist removal but mass transfer is poor. For the high inlet SO_2 concentration at Phillips only about 50% removal has been obtained. However, for low-sulfur coal, as at Colstrip, a single venturi is adequate.

The venturi is superior for particulate removal, which is an important design factor, of course, when the scrubber is designed to take care of part or all of the dust removal function. Other scrubbers, particularly the mobile bed type, are also effective but have not been used as much or tested as thoroughly on a full scale.

The spray scrubbers (Pullman Kellogg, Peabody, CE, Chemico) are the simplest and normally the least expensive to build and operate. Pressure drop is especially low, an important factor in operating cost. However, the relatively low mass transfer makes it necessary to pump a large amount of slurry around the loop, and other adjustments (see later) may be warranted to increase removal efficiency. Gas distribution may also be poor and mist generation heavier than with other types; further data are needed in this area.

Other scrubber types are somewhat intermediate, including the B and W sieve tray, the Riley rod packing, and the TVA screen-type packing (also used in Japan).

Scrubber type is one of the trade-offs that must be considered along with several other factors in optimizing design. It is not yet clear which type is the best.

2. Lime vs. limestone. Lime is much more reactive than limestone; under the same conditions it will give better SO_2 removal and is easier to manage in regard to scaling. The choice is complicated by several factors, however, including the higher cost of lime and the energy consumed in producing it from limestone. The system suppliers normally will design for either type depending on local cost and customer preference; others prefer one or the other but do not seem adamant on the matter.

3. Use of Mg^{++}. Carrying magnesium compounds in the loop helps removal efficiency as well as preventing scaling. Recent tests at Phillips gave over 80% removal with MgO added to the lime as compared to about 50% without it. At Mansfield, it will apparently be possible to meet the plus 90% removal requirement with the dolomitic lime, whereas this would almost certainly not be feasible with ordinary lime in the relatively inefficient venturi absorbers. Kawasaki in Japan reports greatly increased efficiency from the Mg^{++} addition.

   It seems likely that the better removal will be a significant advantage for dolomitic lime. However, limestone systems can be helped in the same way by adding Mg^{++}. It is not yet clear what the trade-offs will be, especially in areas where limestone has a major cost advantage.

4. Absorbent utilization. Designers normally plan on use of some excess absorbent (say 110 to 130% of the stoichiometric amount) to improve SO2 removal. This not only increases absorbent cost but also produces more sludge, disposal of which is an even more important cost consideration, and makes mist eliminator operation more difficult (see later). It seems likely that the trend will be to designing for higher utilization by some combination of better mass transfer, use of magnesium, plug flow in the hold tank, and two-stage absorption.

   The problem is easier with lime because of its greater reactivity. For limestone, two-stage absorption, such as tested by EPA and used commercially in Japan, may be the best trade-off. By using the two stages, the first one can be operated at a low enough pH to get all the limestone into solution and available for reaction. For such a system, forced oxidation is essential in order to provide enough sulfate crystals for preventing scaling; the oxidation is highly desirable anyway in order to improve sludge dewatering properties (see later). If the latter is planned for a lime system, which may be a good practice, the two-stage operation is also needed to get the pH low enough for high sulfite solubility, since the sulfite must be dissolved before it can be oxidized.

   The two-stage operation fits in with the two-scrubber systems such as Chemico, B and W, Research-Cottrell, CE, and others. Major test programs on forced oxidation are being planned at operating installations over the country.

Mist Elimination

The problem of entrainment separation has resisted solution longer than any of the other troubles encountered in developing lime-limestone scrubbing. The horizontally-mounted chevron units used in most scrubbers are relatively inexpensive but tend to hold collected mist and act like scrubbers themselves, abosrbing SO_2 and reacting it with absorbent carried up with the mist to form calcium sulfate with resulting scaling and plugging.

Most of the existing scrubbers were plagued with mist eliminator plugging in early operation and some still are. At Shawnee, much of the four years of test operation has been concerned with trying to make the mist eliminators work. The operation is now reasonably acceptable but there is no assurance that the experience can be duplicated elsewhere.

The inter-relationships of the many factors affecting entrainment separator design and operation are extremely complicated, so much so that they cannot be covered adequately in this paper. Instead, the various design approaches will be summarized and evaluated.

1. TVA; Riley. The chevron is placed outside of the scrubber in the exit duct, thus allowing it to be mounted vertically rather than horizontally. In this position it drains better and the collected liquid is removed before much SO_2 absorption can occur.

2. CE. Much the same effect is obtained, even though the unit is located in the upper part of the scrubber, by mounting it in a slanted position. CE has also stressed intermittent, heavy washing to keep the unit clean without exceeding the amount of fresh wash water available; use of recycled supernatant liquor, which does not affect the water balance, usually gives trouble because it already contains dissolved calcium sulfate.

3. B and W; UOP. The upward gas velocity is reduced by using a larger cross sectional area where the mist eliminator is located than in the main scrubber. At the lower velocity, the liquor can drain back with less resistance by the up-flowing gas.

4. Pullman Kellogg. Since the scrubber is horizontal, the mist eliminator logically is mounted in a vertical position. The trailing edges of the baffles are formed in such a way as to make pockets to catch liquid "crawling" along the surface, thus allowing higher gas velocity and better mist removal efficiency.

5. Peabody. A centrifugal type is used rather than a chevron. Efficiency is probably not as good but the open unit is less subject to plugging.

6. Chemico; CEA. Venturis have the advantage that the down-flowing gas impinges on the pool in the bottom of the scrubber before reversing direction and flowing up through the mist eliminator. Thus much of the mist is removed before the gas reaches the mist separation unit.

In Japan, MHI uses both the downflow principle and the location in the exit duct to obtain superior mist eliminator performance.

Factors that may improve mist eliminator performance in the future include better absorbent utilization and use of magnesium. Tests at Shawnee have shown that high utilization reduces scaling in the mist separator, presumably because there is less lime or limestone carried up to react with SO_2 absorbed in the unit. Magnesium is helpful because it reduces scaling from the liquor collected in the unit and also because recycled liquor can be used for washing (thus allowing more wash) with less danger of scaling.

Although these approaches are promising, they have not figured much yet in system planning.

In addition to plugging, mist eliminators can also give trouble by inadequate mist removal. This is not a usual situation, because scrubber droplets are large and relatively easy to remove from the gas, but inadequate removal has occurred; for example, a major current problem at Mansfield is mist carrythrough to ducts, reheater and stack. The causes are not clear but, as a general rule, chevrons mounted vertically and with pockets on the trailing edges are much more efficient than the flat, unpocketed-slat type.

Reheat

The problem of gas reheat is a difficult one because reheat is very expensive and there may not be enough extraction steam available if steam reheat, usually the most economical type, is used. On the other hand, if the gas is not reheated (or not reheated enough), then downstream equipment is subject to corrosion and "acid rain" may be a problem.

This is also a very complicated problem and it is not yet clear what the best solution may be. Since reheat does not enter into scrubber design, the choice is usually left up to the utility no matter which scrubber system is selected. The majority of the systems being planned currently will have reheat but a significant number will not.

Reheater design is difficult because heat exchangers are subject to corrosion and direct-fired heaters not only use expensive fuel but also may not work well. For example, the oil-fired heaters installed at three stations (Phillips, Elrama, Mansfield) are currently inoperable because of vibration.

Heat-exchanger types (usually with steam as the heating medium) are sensitive to the amount of mist impinging on them, which is another reason for emphasizing high mist removal efficiency. There is no good solution yet to the materials problem; ordinary steel has lasted well in some situations but stainless steel or even more resistant materials have failed rapidly in others.

Another of the several inter-relationships is between reheat and fan location. If the fan is located after the reheater, the mist eliminator and reheater must perform well in order to avoid mist impingement on the fan, which is very likely to cause corrosion and imbalance. Since fan damage has occurred in some installations, utilities now generally favor location of the fan before the scrubber. This, however, precludes the scrubber from accepting much of the dust. Another approach, favored by Chemico, is to place the fan between the scrubber and the reheater, thus making it a wet fan and requiring an efficient fan wash system. At Phillips, an effective wash arrangement has been developed but wet fans are not generally popular with utilities.

Sludge Disposal

Disposal of the waste solids has received more publicity than any other scrubber problem, to the extent that the situation may have gotten somewhat out of perspective. The problems are the same as in ash disposal, a large volume of solids, limited space around power plants for discarding them, and the hazard of water pollution by leaching. Scrubber solids tend to have larger settled volume in waste ponds (for high-sulfur coal) but the ash appears to be more of a pollution problem.

The usual course is to put both ash and sludge in waste ponds (lined or unlined), sometimes together and sometimes separately. So far this has been acceptable, although EPA and some local agencies have taken the position that the sludge should be made nonleaching and load-bearing by stabilization. In some areas, stabilization is required by state or local agencies and there is some indication that such regulations will spread.

The principal stabilization methods are those developed by Dravo and IUCS. Both are just coming into full-scale use, Dravo at Mansfield and IUCS at Elrama. Each will be used in systems now in the planning stage.

The main difference is in the first step of the operation, which is "wet" for Dravo and "dry" for IUCS. Dravo mixes the stabilizer additive (proprietary) with thickener underflow (usually about 35% solids) and then pumps the mix either to an intermediate holding pond or direct to the final pond. The supernatant is pumped back to the scrubbers and the settled solids in the pond gradually harden to the final stabilized state. When placed in intermediate ponds, the hardened solids are excavated and used as landfill.

In contrast, IUCS further dewaters thickener underflow by filtration and then adds lime and flyash to give a "dry" solid that can be used immediately as land fill. (The filtration can be omitted if enough ash is available to dry up the mix.)

Thus in a typical operation, IUCS has the extra step of filtration but does not require any ponds, either intermediate or final. Since ponds are quite expensive, this gives IUCS a considerable advantage. The trade-offs are not well worked out, however, and no doubt will be affected greatly by site specific considerations.

The current interest in forced oxidation may also have a significant effect. Gypsum dewaters much better than sulfite sludge and thus should make simple ponding less expensive because of the lower pond volume required. On the other hand, it should also help stabilization (either Dravo or IUCS) by making dewatering easier.

Summary and Conclusions

It is evident from the foregoing that much progress has been made in flue gas desulfurization and that a conventional lime-limestone scrubbing technology is emerging. The main remaining problem is entrainment separation; although acceptable operation has been attained in some systems, this is still a troublesome area in which further development is needed.

Much more work is also needed both in process optimization and in developing a design base that will make it possible to design with confidence for site-specific factors. Although reliability has been improved, lack of confidence by designers still makes a spare scrubber the usual choice for assuring noninterference with boiler operation. Hopefully, scrubbers can eventually be made as reliable as precipitators but this is likely some time off.

In regard to optimization, there is even further to go. There are numerous design questions for which there is as yet no answer. Some of the more difficult ones are as follows.

1. Should the scrubber be preceded by dry dust collection equipment and if so, how efficient should it be?

2. Does a prescrubber give enough advantage to justify the cost?

3. Is is better to use a simple and low-cost scrubber with relatively low mass transfer capability or one with better transfer but with more complicated internals?

4. Is cocurrent operation better than countercurrent?

5. Under what conditions is lime a better choice than limestone?

6. Although Mg^{++} has well-proven advantages, how much extra cost can be justified for using it?

7. What is the best combination of procedures for eliminating scaling?

8. Although a chevron entrainment separator located in the upper part of the scrubber probably is the simplest and most economical arrangement, would it be better to use another approach to get more dependable operation?

9. Of all the possible combinations, where is the best place to put the fan?

10. Is reheat necessary or desirable, and if so, what is the best way to go about it?

11. Is simple ponding of sludge acceptable, and if not, what is the best way to comply with environmental constraints?

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CURRENT STATUS OF THE BI-GAS PROCESS

By

R. K. Young

Presented at

Third Annual International Conference on

Coal Gasification and Liquefaction

University of Pittsburgh

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CURRENT STATUS OF THE BI-GAS PROCESS

By

R. K. Young*

It is a pleasure for BCR to take part in the Third Annual International Conference on Coal Gasification and Liquefaction. Since we have not had the privilege of discussing the BI-GAS coal gasification process with you, perhaps an overview covering our experience would be helpful.

The BCR coal gasification research program was initiated in late 1963 under contract with the Office of Coal Research (now U.S. Energy Research and Development Administration). More recently, as the program moved into the more costly pilot plant stage, the American Gas Association became a substantial sponsor.

Our original scope of work covered two phases, with Phase I being a review of the current state-of-the-art of coal gasification.(1)** The purpose of this review was to determine what processes, if any, were reasonable candidates for further development. From this comprehensive evaluation of processes for production of synthesis gas, it was concluded that the concept of two-stage, entrained, high-pressure, slagging gasification offered considerable promise for production of pipeline gas from coal.

Phase II of our initial program recommended process research and development in the laboratory to provide information required for design of pilot-scale equipment.(2)

SLIDE 1

In the BI-GAS process (Figure 1) coal is introduced into the upper section, or Stage 2, of the gasifier at pressures of 1000 to 1500 psi. The coal comes in contact with a stream of hot synthesis gas produced in the lower section, or Stage 1, of the gasifier and is partially converted into methane and additional synthesis gas.

* Manager, Project Control, Bituminous Coal Research, Inc.

** Numbers in parenthesis refer to Literature References given at end of paper.

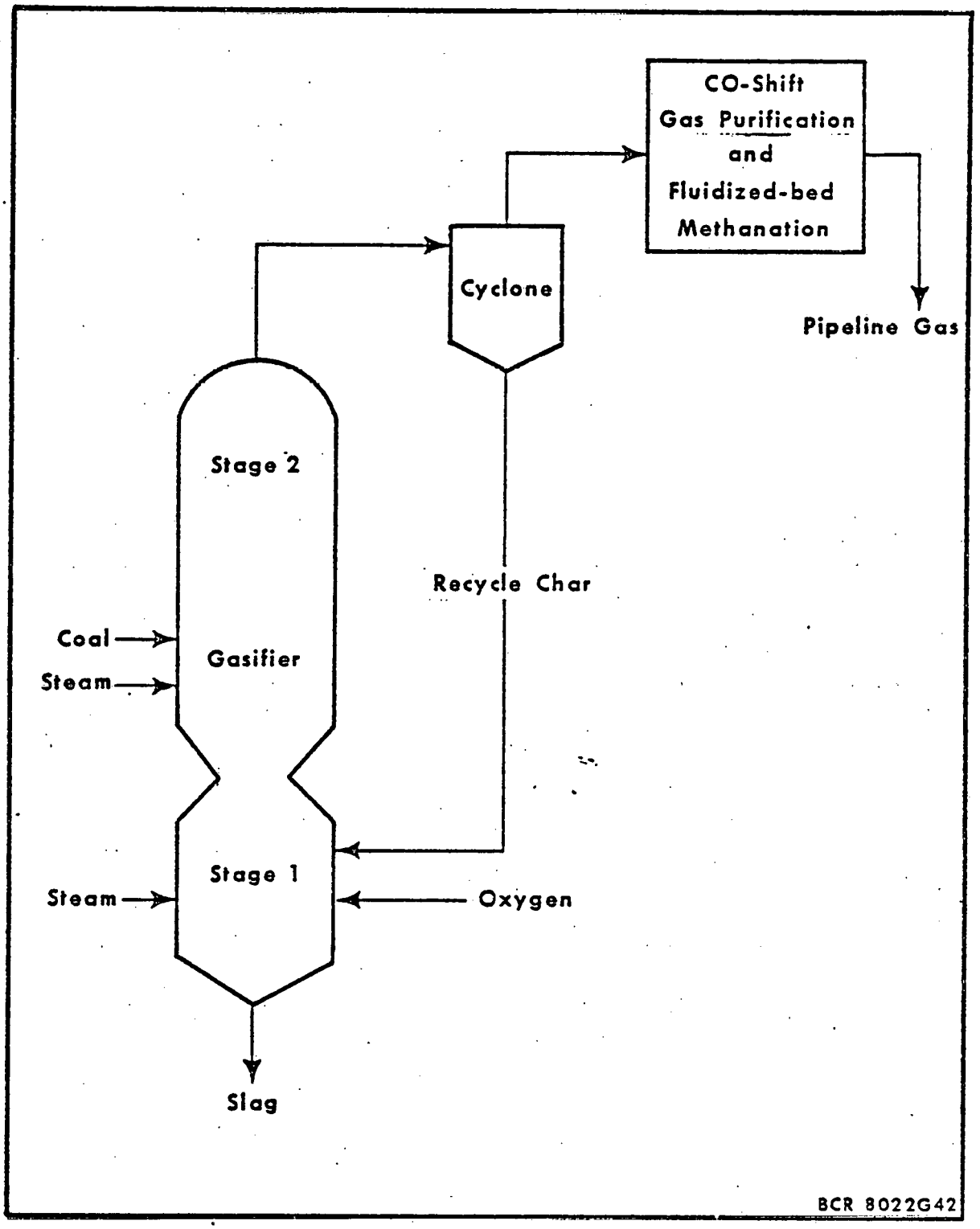


Figure 1. Simplified Flow Diagram
BI-GAS Process

Approximately 50 percent of the methane in the final product gas results from this direct coal methanation.

Residual char in Stage 2 is swept from the vessel along with the gas, is separated by cyclone from the gas stream and returned to Stage 1. The char is then completely gasified under slagging conditions by reaction with oxygen and steam, producing synthesis gas and heat required for Stage 2.

BCR proceeded during the late '60's with laboratory-scale research on Stage 2 gasification reactions which verified the assumptions concerning direct coal methanation. However, products from gasification of coal also require down-stream methanation to produce substitute natural gas. To accommodate BI-GAS operating conditions and product gas compositions, BCR, in a parallel research effort, developed a new concept of fluidized-bed methanation.

I would now like to review briefly some experimental work which BCR has conducted to prove the concept of reactions in Stage 2 of the BI-GAS gasifier.

The initial research involved injection of pulverized coal-water slurry into hot, high pressure autoclaves. As anticipated, high yields of methane were obtained with several different coals.

The next phase of research on Stage 2 involved construction of an electrically-heated, 1-inch I.D. continuous flow reactor which was operated at a coal feed rate of 5 lb/hr, at 1000 psi and 1700 F. Coal, steam, and simulated Stage 1 synthesis gas were fed into this unit. The results showed that direct methanation of coal is a rapid reaction at 1600-1700 F, and with a residence time of several seconds, leads to methane yields corresponding to roughly 20 percent of the coal carbon. (3)

Moseley and Paterson in 1965 gasified coal in pure hydrogen at temperatures of 1750 F and pressures as high as 500 atmospheres.[4] Figure 2 presents their data along with typical data from BCR experiments in the 5 lb/hr continuous flow unit. A single correlating curve passes through the data, verifying the assumption that a high yield of methane can be obtained directly from coal in an entrained system.

SLIDE 2

Although our original assumptions were confirmed, it was deemed necessary to design and operate an intermediate-scale unit prior to design of a pilot-scale gasifier. A 100 lb/hr process equipment development unit (PEDU) was then designed for operation at various pressures, temperatures, and flow rates.[5] In the PEDU, coal and steam were introduced into a hot stream of synthesis gas in order to convert the volatile portion of the coal to methane. In contrast to the 5 lb/hr unit, the synthesis gas and heat for this system were obtained by the partial combustion of ash-free fuels such as liquid hydrocarbons. Figure 3 shows the basic flow diagram for the Stage 2 PEDU.

SLIDE 3

This unit was operated for a two-year period until late 1971, during which time experiments were conducted with lignites, subbituminous, and bituminous coals.[6] Table 1 shows typical results obtained with the Stage 2 PEDU.

SLIDE 4

The following conclusions were evident:

1. Methane yield depends upon coal rank and processing conditions such as temperature and partial pressure of hydrogen.

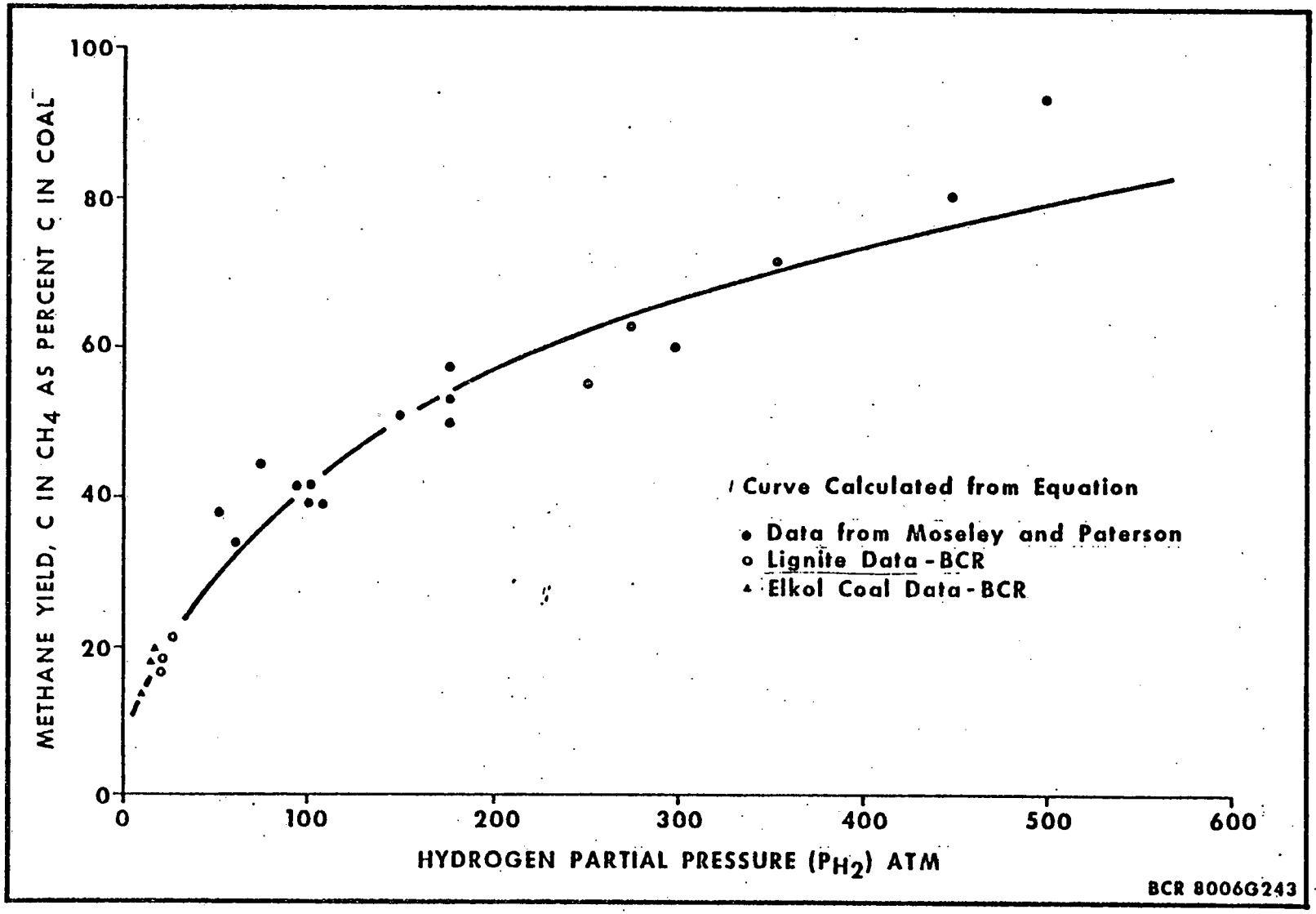


Figure 2. Correlation of Methane Yield with Hydrogen Partial Pressure at 1750 F

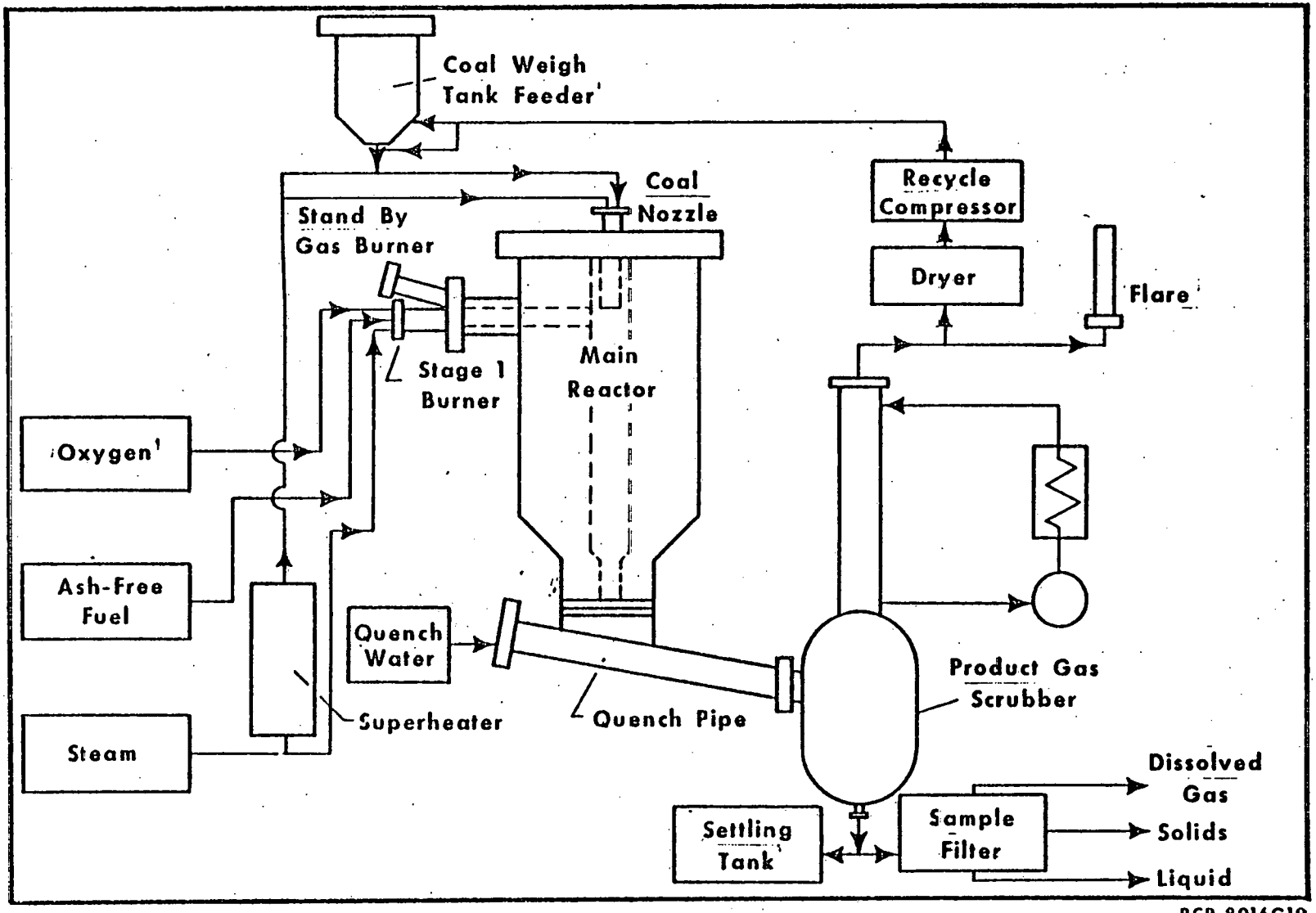


Figure 3. Basic Flow Diagram of 100 lb/hr Stage 2 Process and Equipment Development Unit (PEDU)

TABLE 1. RANGE OF RESULTS FOR 100 LB/HR PEDU TESTS

Coal Tested	Methane Yield, Volume Percent
Lignite	12-20
Elkol	18-23
Pittsburgh	16-26
Lower Kittanning	14-17

2. The direct methanation process is applicable to several different types of coal.

3. Yield of methane expressed as percent conversion of carbon in coal increased with coal rank.

In September 1971, BCR and the sponsor agreed that adequate information for design of the Stage 2 section of a pilot-scale gasifier had been developed from the PEDU program. Realizing that the facilities necessary for pilot testing of the BI-GAS process would be related to other gasification processes under development, OCR suggested planning a multi-purpose research facility. Subsequently, BCR was instructed to plan for design and construction of an oxygen-blown, BI-GAS gasification system as the initial unit.(7)

After review of competitive engineering proposals, OCR directed BCR to retain Koppers Company to assist in developing a process design and bid package for a 5 ton/hr pilot plant. Concurrently, BCR acquired a free plant site of 36 acres at Homer City, Pennsylvania, and arranged for transfer of title to the government.

During the same time period, Air Products and Chemicals, Inc., under a separate OCR contract, performed a detailed review of the BI-GAS process.(8) Among the main conclusions by Air Products was that the BI-GAS concept offered genuine practical advantages in coal gasification, and that the projected economics were competitive with other processes under investigation.

After evaluating several bids for construction and operation of the BI-GAS pilot plant, Stearns-Roger Inc., was selected in mid-1972 as the engineering, construction, and operations subcontractor. Babcock & Wilcox was in turn selected to establish final design and to fabricate the gasifier.

BCR, meanwhile, continued to provide information to assist in design of the gasifier. Recognizing that physical effects in Stage 2 of the unit can influence methane yield from coal, we decided to investigate flow and mixing patterns via a cold flow model program. A series of tests was carried out in a Plexiglass model of the 5 ton/hr unit. This model was full-scale with respect to diameter and length of Stage 1. (See Figure 4.) Many facets of gasifier design were tested with this model; however, major emphasis was placed on determining effects of feed nozzle position and orientation on slag deposition in Stage 1, and on mixing patterns in Stage 2. As a result of this program, several important gasifier design decisions were made.

SLIDE 5

The pilot plant gasifier is a cylinder with a shell diameter of 5 feet and a length of approximately 54 feet (see Figure 5). The main sections of the gasifier are the bottom slag quench zone, the Stage 1 char slagging section, and Stage 2 where coal is injected. The throat area between Stages 1 and 2 aids in separation of molten slag and produces gas velocity great enough to prevent direct return of coal or char particles from Stage 2.

SLIDE 6

Provisions have been made at the pilot plant for a high-pressure coal-water slurry feed system. The slurry is pumped to pressure and conveyed to a spray dryer, after which the coal is separated in a cyclone. Coal and steam are injected into Stage 2 of the gasifier to contact the hot synthesis gas from Stage 1, converting the coal to additional synthesis gas, methane, and char. As mentioned previously, the gas and char are quenched; the raw gas undergoing further processing while the char is recycled to Stage 1.

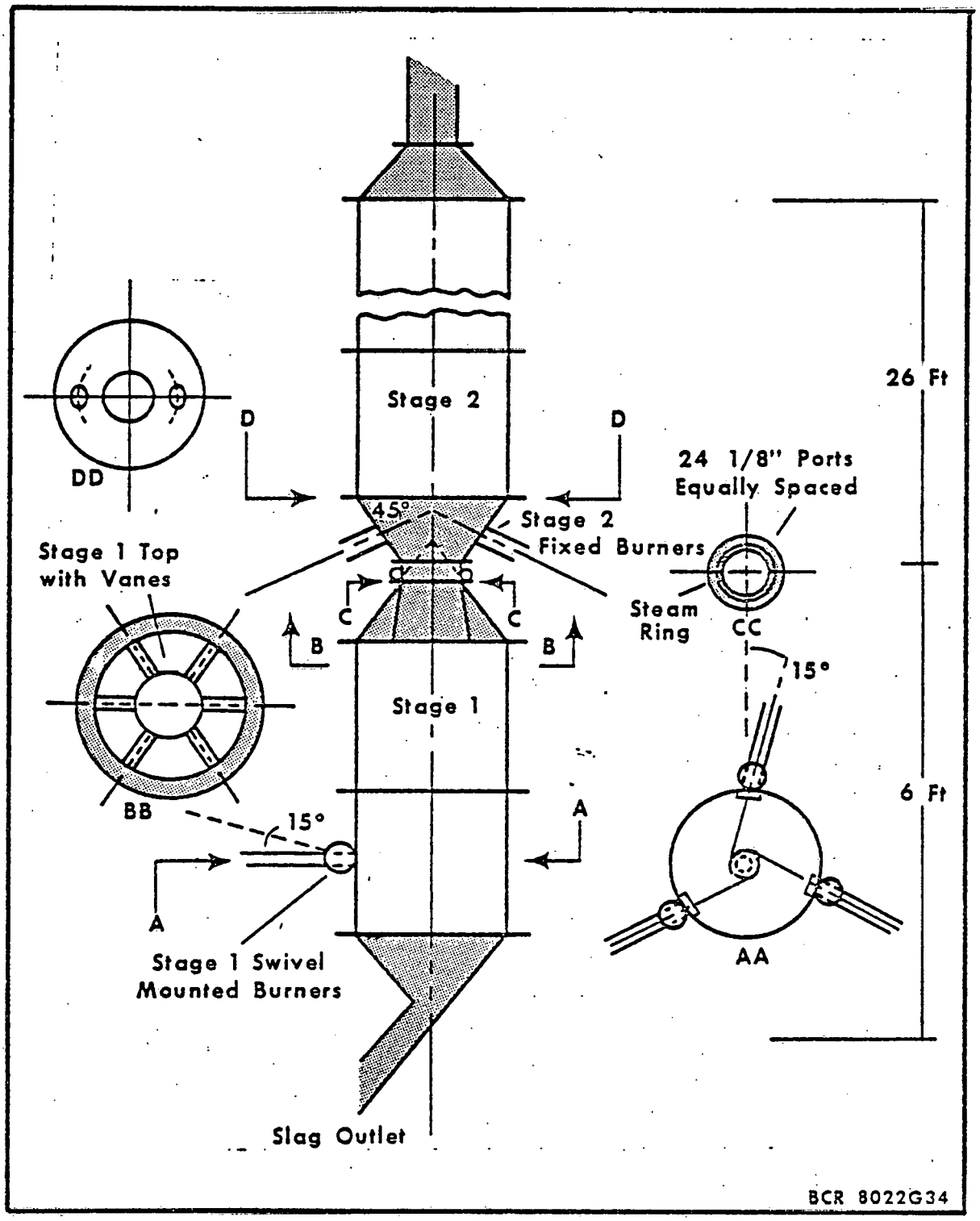


Figure 4. Diagram of Cold Flow Model

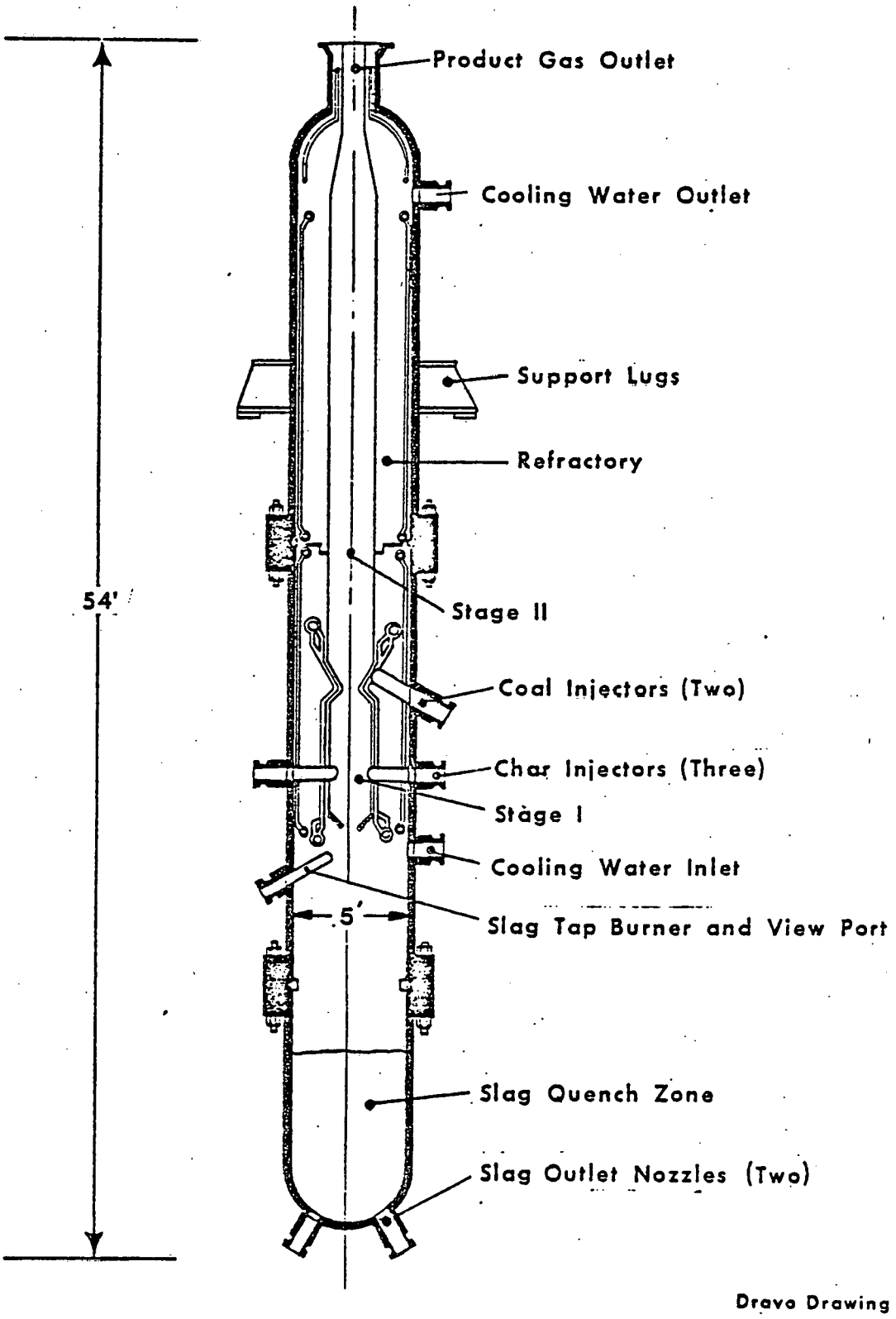


Figure 5. BI-GAS Gasifier

Steam, oxygen, and char are admitted to Stage 1 via three injectors arranged to fire cyclonically which promote slag separation as well as efficient gasification. Molten slag then flows down the Stage 1 walls into the quench zone where it is removed by sequencing two slag lock hoppers.

Operating conditions for the pilot plant gasifier are as follows: temperature in Stage 1 can range from 2600 to 3000 F with estimated mixing temperature in Stage 2 of 2200 F; temperature range of gas exiting Stage 2 is 1500-1700 F and operating pressure may be varied from 500 to 1500 psi.

Table 2 indicates product gas composition expected from gasification of a Montana subbituminous coal which has been selected for startup of the pilot plant. Operation of the pilot plant gasifier will involve the following three phases: SLIDE 7

Phase I - Natural gas firing in Stage 1 and steam cooling in Stage 2.

Phase II - Gas firing in Stage 1 and coal feed to Stage 2.

Phase III - Integrated operation with char firing in Stage 1 and coal feed to Stage 2.

BCR is also successfully conducting research on fluidized-bed methanation systems. While there is proven technology for methanation of gases containing low concentrations of carbon monoxide (1 to 3 percent), and fixed-bed methanation has been demonstrated at the Westfield plant in Scotland, more work must be undertaken on methanation of coal derived product gas because of the substantially higher carbon monoxide content (>20 percent).

TABLE 2. GASIFIER PRODUCT GAS COMPOSITION, (VOLUME PERCENT), MONTANA SUBBITUMINOUS COAL

Carbon Dixoide	18.0
Carbon Monoxide	29.0
Hydrogen	23.0
Methane	12.0
Nitrogen	0.4
Hydrogen Sulfide	0.2
Steam	17.4
Total	100.0

In early 1970, BCR initiated a program to develop kinetic data on methanation catalysts that might be used for developmental gasification processes. Bench-scale experimental units were designed to study conversion of synthesis gas to methane at 1000-1500 psi and at temperatures between 600 and 1000 F. By mid-1970, plans had been made for design of a 5 ton/hr pilot plant; therefore, it was necessary to select a methanation scheme, and based upon limited experience, the decision was made to proceed with the dry fluidized-bed concept. The fluidized-bed provided several advantages over other systems, namely, minimization of temperature gradients, higher heat transfer coefficients compared to fixed beds, and the ability to continually add or remove catalyst.

Experimental work progressed quite satisfactorily; however, it became apparent that the fluidized-bed methanation system should also be investigated on a larger scale in order to provide more meaningful data for pilot plant operations. Design work on a methanation PEDU began in early 1972 with the assistance of Koppers Company. This resulted in construction and installation of the unit shown in Figure 6.(9)

SLIDE 8

Emphasis was placed on providing operational flexibility to permit study of key variables in the production of pipeline-quality gas. An on-line data acquisition and processing computer was coupled with the PEDU to allow operating engineers to obtain complete reports including material and energy balances upon request.

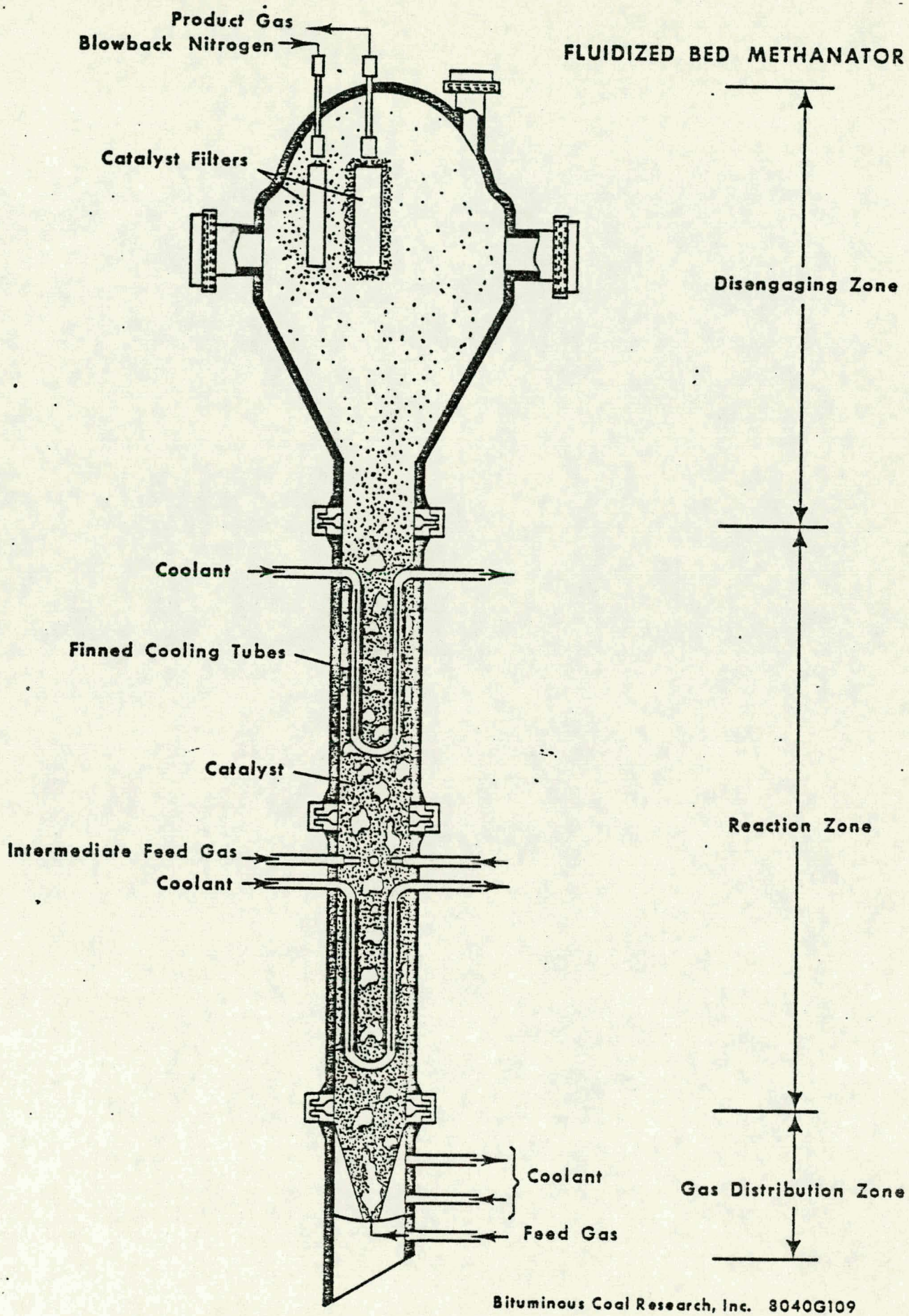


Figure 6. Fluidized-bed Methanator

PEDU tests began in early 1974 and have continued until the present time. The main objective has been to obtain operating data on selected catalysts, exploring such variables as feed gas composition, pressure, bed temperature, and space velocity.(10) Recent PEDU data have been particularly encouraging. For example, at design operating conditions conversions of carbon monoxide and hydrogen to methane of 90+ percent have been attained regularly with no recycle required.

The current experimental program is directed toward optimization of operating conditions for the pilot plant methanator. As a matter of fact, the initial catalyst charge recommended by BCR has been delivered to the pilot plant. Bench-scale and PEDU evaluations are being conducted at present time.

While BCR's main efforts have been in the areas of entrained gasification, and development of the fluidized-bed methanation concept, the pilot plant contains all equipment necessary for complete handling and processing of coal and product gases. The block flow diagram (Figure 7) shows that these facilities include coal storage and preparation, a slurry preparation and drying system, a two-stage slagging gasifier, CO shift and acid gas removal systems, a Claus plant for removing sulfur, and a fluidized-bed methanation system for upgrading the calorific value of product gas. SLIDE 9

Construction of the pilot plant is essentially complete and several plant systems up stream of the gasification area have been checked out and operated satisfactorily. Current planning envisions initial operation of the gasifier during early fall, 1976.

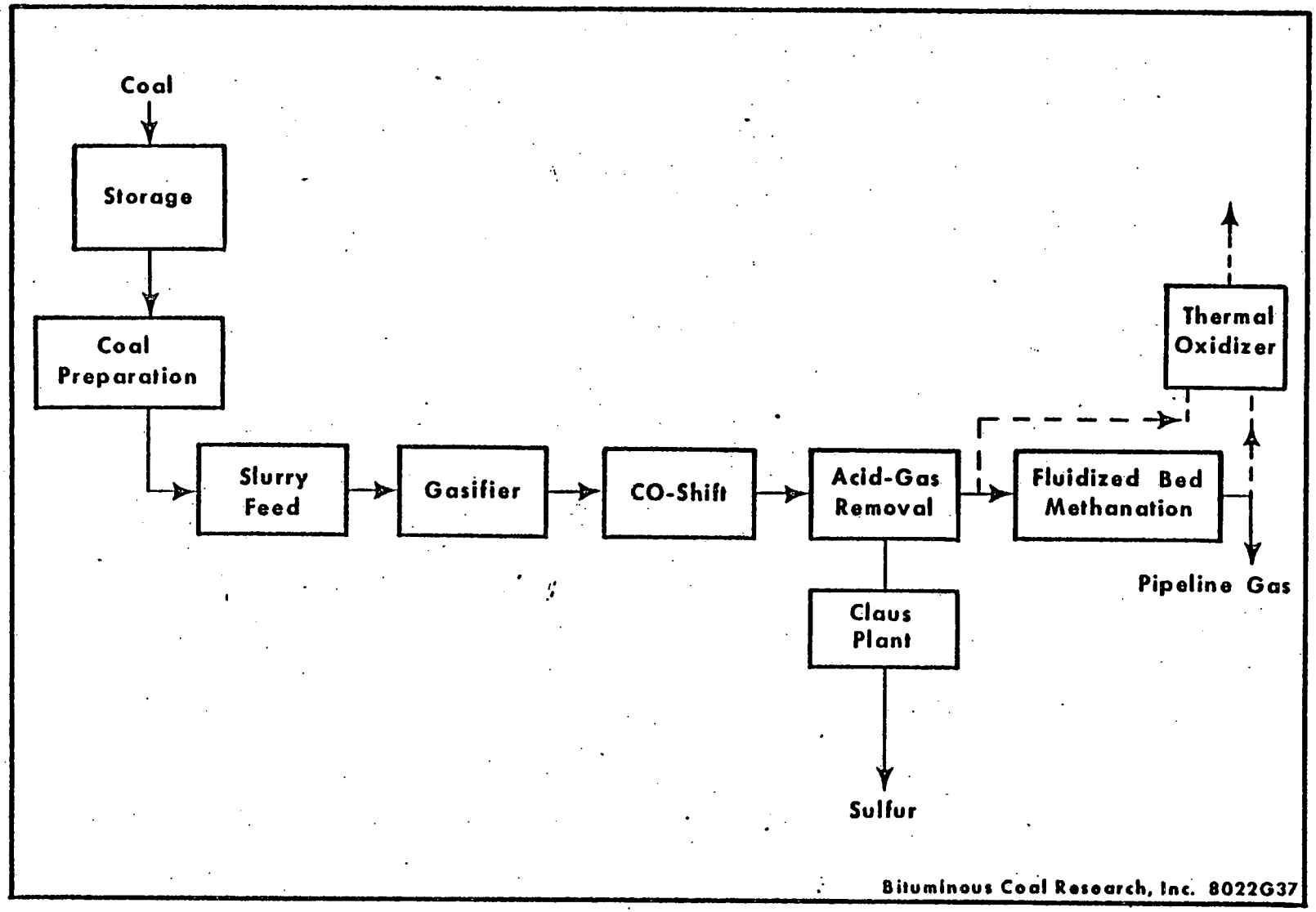


Figure 7. BI-GAS 5 Ton Per Hour Pilot Plant Flow Diagram

The successful operation of the BI-GAS pilot plant at Homer City will involve the coordinated efforts and cooperation of three different organizations. Phillips Petroleum Company has been responsible for insuring that the pilot plant was constructed according to design specifications. Phillips is also responsible for overall program management and planning, plus process evaluation and development. Stearns-Roger Inc., is concerned mainly with pilot plant operation and maintenance of facilities. BCR, as prime contractor, works closely with both Phillips and Stearns-Roger in planning for operations at the pilot plant. Principal contributions by BCR include assistance in planning gasifier and methanator operations, development and installation of data acquisition systems, plus consultation and backup in the areas of analytical and environmental services.

Activities currently in progress at the pilot plant, or at BCR in support of the pilot plant program, include process studies in the coal preparation area, particularly on coal and char flow measuring systems; preparation for phased operations of the gasifier; installation of a complete on-line data acquisition system; experimental work in slag viscometry for the purpose of studying various coal and coal flux mixtures; modifications of our computer simulator to provide specific data for operations with start-up coal; and finally preparations for an environmental sampling program to be conducted by a subcontractor under the guidance of Phillips and BCR.

SERIES OF PILOT PLANT SLIDES

ACKNOWLEDGMENT

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THE CONSOL FGD PROCESS

by

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ABSTRACT

This paper describes a new regenerative process for the removal of 95% or more of sulfur dioxide from power plant stack gases. Scrubbing is conducted by circulation of a concentrated aqueous solution of potassium thiosulfate through a pump-around loop containing a packed bed scrubber for SO_2 removal and an external reaction drum.

A solution containing a mixture of hydrosulfides, polysulfides, carbonates and bicarbonates is injected into the circulating salt solution just ahead of the reaction drum. The hydrosulfides and polysulfides are quantitatively removed from the circulating salt solution before entering the scrubber by reaction with sulfites formed in the scrubber to produce thiosulfate.

The carbonates in the feed solution pass through the reaction drum unconverted, but react in the scrubber with SO_2 to produce the aforementioned sulfites. The net product from the scrubbing process is thus potassium thiosulfate.

The scrubber feed solution is produced by noncatalytic reduction of the thiosulfate with a CO-rich gas. H_2S is simultaneously produced which is converted to elemental sulfur in a Claus plant.

A very low by-product sulfate yield of about 1% of the SO_2 absorbed and very high SO_2 removal efficiency as high as 99% in a simple single-stage scrubber were achieved. The scrubbing solution is a clear salt solution which provides for excellent operability in the scrubber which is free of problems due to scaling and deposition of solids.

The scrubbing process was demonstrated on both a 1000 ACFM and a 30,000 ACFM pilot plant. The latter plant operated with flue gas drawn from a coal-fired power plant of Philadelphia Electric Company. Regeneration of the spent reagent was demonstrated in a separate pilot operation. Finally, closed-loop operation was demonstrated with continuous circulation of reagent between the regeneration and the 1000 ACFM scrubbing pilot plants.

An economic analysis showed the process to be competitive with other regenerable processes.

THE CONSOL FGD PROCESS

INTRODUCTION

The original aim of this project was further development of the aqueous potassium formate process reported by Yavorsky, et al in 1970[1]. The process chemistry, as a result of concurrent laboratory research, was modified and simplified early in the course of the study. The new scrubbing process was demonstrated both in a 1000 ACFM pilot plant at the Research Division of Consolidation Coal Company in Library, Pennsylvania, and in a 30,000 ACFM pilot plant located at the Cromby Power Plant of Philadelphia Electric Company. A regeneration pilot plant was also built and operated at Library. The two units at Library were operated together in closed-loop, integrated fashion in the final phase of the program.

PROCESS DESCRIPTION AND DEVELOPMENT HISTORY

Figure 1 is a schematic flow sheet of the scrubbing system and corresponds to both the Library and the Cromby pilot plants. A venturi scrubber was used in each unit to quench and humidify the hot flue gas. It served also to remove the SO_3, HCl and any residual particulate matter from the flue gas.

The scrubber itself is a packed tower with a pump-around circuit through a reaction drum. Fresh feed, spent scrubber product and make-up potash are fed and withdrawn from the scrubber system as shown. The solid lines indicate the feed and withdrawal points as practiced in the pilot plant operations, the dotted lines indicate the preferred points for a commercial system.

The key reaction in the Consol process is the one that takes place in the reaction drum. Experiments confirmed previous literature data by Hansen and Werres in 1933[2] that sulfite reduction to thiosulfate may be effected by use of KHS as a reducing agent, i.e., by the reaction:

$$KHS + 2\ KHSO_3 = 3/2\ K_2S_2O_3 + 3/2\ H_2O \qquad (1)$$

Similar reactions occur with polysulfides and sulfite.

The feed to the scrubber system is an aqueous solution containing KHS, K_2S_x, $KHCO_3$, K_2CO_3 and KOOCH produced in the regenerator by reduction of the thiosulfate scrub product with CO. The polysulfides and KOOCH are relatively minor components. The polysulfides are consumed in the reaction drum while the KOOCH is substantially inert under the conditions of the process. The feed enters the reaction drum where it is mixed and reacts with the circulating scrub solution. Conditions are imposed such that the KHS and K_2S_x are completely consumed in the reaction drum via reaction (1) and the corresponding reaction for K_2S_x. A residence time of less than 1.5 min is sufficient at a pH of 7. Complete consumption of KHS is desirable since, if it is allowed to enter the scrubber, some evolution of H_2S into the stack gas occurs.

The actual removal of SC_2 in the scrubber occurs by virtue of reactions with carbonates in the feed solution as, for example,

$$K_2CO_3 + H_2O + 2\ SO_2 = 2\ KHSO_3 + CO_2 \qquad (2)$$

The overall reaction in the scrubbing system is now the sum of reactions (1) and (2), and may be written as reaction (3) for the KHS and K_2CO_3.

$$KHS + K_2CO_3 + 2\ SO_2 = 3/2\ K_2S_2O_3 + 1/2\ H_2O + CO_2 \qquad (3)$$

Similar equations apply for polysulfides and bicarbonate.

Control of the scrubbing process is effected in two ways. The first is by controlling the rate of the fresh scrubber feed to maintain the pH in the scrubber effluent at about 7. This insures nearly complete SO_2 removal as well as quantitative removal of the KHS and K_2S_x from the feed reaction drum via reaction (1). The second is to control the composition of the scrubber feed to maintain the proper proportions of KHS and K_2S_x with reference to the content of K_2CO_3 and $KHCO_3$. The stoichiometric relationships may be expressed by the equation:

$$R = \frac{2\ S^{o} + 3\ S^{=}}{\Sigma\ K} \qquad (4)$$

where S^{o}, $S^{=}$ and K refer to the mols of zero-valent sulfur, sulfur with a valance of -2, and total mols of K in the feed solution associated with the compounds KHS, K_2S_x, $KHCO_3$ and K_2CO_3.

The above ratio R is termed the "acceptability ratio." When R equals 1, reaction (3) would proceed quantitatively to produce pure thiosulfate. When R is greater than 1, a surplus of sulfur is present in the feed solution and H_2S may be evolved into the stack. In practice R should be close to but below 1 to maintain an excess of sulfite to help drive reaction (1) to completion.

Lower values of R are undesirable as this permits the concentrations of sulfite and bisulfite ion in the circulating scrub solution to increase. The result is a decrease in the absorption of SO_2 and an increase in sulfate formation as a result of the oxidation reaction:

$$2\ KHSO_3 + O_2 = K_2SO_4 + H_2SO_4 \qquad (5)$$

The sulfate yield in the process is, however, quite low and is easily kept to about 1 mol % or less of the SO_2 absorbed. This is a consequence of the low sulfite concentration maintained in the process, and the fact that concentrated solutions of thiosulfate strongly inhibit the oxidation reaction (5). The low sulfate yield in the revised process is one of the principal advantages. The K_2SO_4 also is readily saleable.

Because of the high solubility of $K_2S_2O_3$ in water, it is possible to operate the scrubber system with a highly concentrated salt

solution. Clear solutions are handled throughout the process except at the point of sulfate removal. Particularly important is the fact that no deposition of solids or scaling takes place in the scrubber.

The regeneration system in the Consol process is illustrated schematically in Figure 2. The basic reaction is the noncatalytic reduction of thiosulfate with a carbon monoxide-rich gas. The reaction may be written as follows:

$$3\ K_2S_2O_3 + 12\ CO + 7\ H_2O = 2\ KHS + 4\ KHCO_3 + 4\ H_2S + 8\ CO_2 \qquad (6)$$

The optimum temperature for the reaction from the practical point of view is in the neighborhood of 450°F. Total pressures in the range of 500-1000 psig are suitable. Typical operating conditions for the various process blocks are shown in Figure 2.

Hydrogen, in the absence of a catalyst, is inert under these conditions. Hydrogen may be utilized, if desired, by introduction of a catalyst containing metal sulfides of Group VI and/or Group VIII supported on active carbon as reported in 1973 by Urban and Alpert[3] and by Urban[4]. The use of a catalyst was tested, but was rejected to maintain simplicity of the system.

Some polysulfides are formed during regeneration. They probably result from reduction of $K_2S_2O_3$ with H_2S:

$$K_2S_2O_3 + 3\ H_2S = K_2S_5 + 3\ H_2O \qquad (7)$$

The ratio of KHS to $KHCO_3$ in the reduction product is dictated by establishment of equilibrium in the reaction:

$$KHS + CO_2 + H_2O = KHCO_3 + H_2S \qquad (8)$$

The acceptability ratio thus may be controlled by the amount of CO_2 in the reducing gas. Operation of the pilot regenerator was normally carried out so that the acceptability ratio of the reactor product was above 1.0. The final adjustment of acceptability ratio to the desired value of about 0.95 was achieved in the flash drum (cf. Figure 2). Here the acceptability ratio is reduced by the decomposition reaction,

$$KHS + KHCO_3 = K_2CO_3 + H_2S \qquad (9)$$

Simultaneously CO_2 is released by the decomposition reaction,

$$2\ KHCO_3 = K_2CO_3 + CO_2 + H_2O \qquad (10)$$

The release of CO_2 does not affect the acceptability ratio, but it does convert slightly soluble $KHCO_3$ into more soluble K_2CO_3.

The off-gas from the regenerator contains unreacted CO, H_2, CO_2 and H_2S. The H_2S is selectively removed, after cooling and condensation of the steam, by scrubbing with a polar organic solvent. Several proprietary processes are available for this service. The absorbed

H_2S-CO_2 gases are released from the solvent in its regeneration. These are combined with the H_2S-CO_2 gas released in the flash drum and sent to a Claus plant where the H_2S is converted to sulfur. The Claus tail gases are incinerated with a portion of the scrubbed regenerator offgas to provide hot flue gases for direct reheating of the scrubbed stack gas.

The major portion of the cleaned regenerator offgas is recycled through an oxygen-fired, partial-oxidation unit. No steam is added to the unit and hydrogen in the recycle gas is converted to CO by reverse water-gas shift:

$$H_2 + CO_2 = CO + H_2O \qquad (11)$$

EXPERIMENTAL

The entire Library scrubbing unit, i.e., venturi, scrubber and reaction drum, was constructed of type 316 stainless steel. The bottom section of the scrubbing tower served as a cyclone separator to remove the venturi water. A chevron-type mist eliminator was used above the separator to prevent water droplets from entering the packed section.

The scrubbing tower was 20 inches ID and contained a 42 inch depth of No. 1 Intalox packing in the form of polypropylene saddles. A distributor was used to supply an even flow of the recycle liquid over the top of the packing. A York mesh-type demister was used above the packing.

Hot flue gas was supplied to the Library unit from a packaged boiler fired with No. 2 fuel oil. A forced-draft fan was used to drive the flue gas through the scrubbing unit. Sulfur dioxide was metered into the gas to achieve concentrations similar to that in the coal-fired units. Provisions were also made to feed flue dust into the gas prior to the venturi.

The Cromby pilot plant was constructed in a similar manner. The flue gases were taken from the output of a 160 MW, coal-fired boiler as a side stream and a separate forced-draft fan was used to drive the gases through the unit. The venturi was constructed of type 316 stainless steel.

The scrubbing tower was 8 ft ID and, in contrast to the Library unit, was constructed of rubber-lined carbon steel. The lower section again was used as a cyclonic separator above which was a chevron-type mist eliminator. The packed section was 5 ft high. The packing initially used was No. 3 Intalox polypropylene saddles. A weir-trough liquid distributor was used, but distribution between the two weirs proved to be unequal. After Run 2, the packing was changed to No. 2 Intalox plastic saddles, and an improved liquid distributor was installed to ensure equal flow to the weirs. The initial configuration also used a York mesh, style 931, teflon mist eliminator at the top of the tower. The reaction drum and all the piping for the pump-around circuits were constructed of rubber-lined carbon steel.

The regeneration pilot unit is shown schematically in Figure 3. The regeneration vessel itself was a mechanically-agitated, gas sparged unit equipped with a Hastelloy liner for protection against stress corrosion cracking due to chloride ion contaminants in the feed. Hot liquid piping also consisted of Hastelloy tubing. All other tubing and equipment such as the flash drum were constructed of 316 stainless steel.

The agitation was effected by a six-bladed turbine impeller, 8.5 inches OD, driven through a mechanical seal by a variable-speed drive, generally at 330 rpm. The rate of absorption of CO was found to vary with the 1.8 power of the rpm, which is similar to the value of 2.0 found by Yoshida et al in 1960[5].

The liquid pool in the regenerator was 15 inches ID x 15 inches high. Level was controlled by an overflow weir through which the regeneration product flowed into the surge pot below the regenerator vessel.

Integrated operation with the scrubber and regenerator was carried out by continuous circulation of salt between the two units. The regenerator and scrubber feed tanks provided surge capacity. The closed-loop operation was conducted continuously for a period of ten days. This corresponds to 12 turnovers of the inventory.

DISCUSSION OF RESULTS

Scrubbing

At Cromby, a series of tests with water scrubbing was run initially to check the mechanical components and the operation of the venturi. The venturi pressure drop was 150% of the design value even with the variable plug fully extended. All tests were therefore run with higher than desired pressure drop. With a flue gas input of 16,000 SCFM, the venturi operated with a pressure drop of 11 to 13 inches of water. The inlet dust loading was 0.22 grains/SCF and the outlet 0.01 grains/SCF for 95% dust removal. Eighty percent of the removed particulates were within the range of 3 to 10 microns. No measurable particulates were removed in the scrubbing loop.

The results of the SO_2 scrubbing tests are given in Table 1. Tables 2 and 3 give the composition of the feed and product solutions and gas, respectively.

The first test, CS-2, was carried out with the larger size, No. 3 saddles. It was a continuous ten-day run. No increase in pressure drop over the packing was observed during this run. The York demister showed a very slight increase in pressure drop for the first seven days and thereafter it went up sharply due to collection of fly ash. It was necessary after this to periodically back-flush the demister with water every 22 hours for a period of 45 to 60 seconds. The back flushing temporarily restored the pressure drop to its original value.

Tests showed that at the 20,000 ACFM gas rate used in this run, less than 0.01% of the potassium in the fresh feed solution would exit the stack without a mist eliminator.

Average absorption of SO_2 was relatively low in this run, i.e., 88%. It was found, however, that proper control could be maintained, and the percent removal of SO_2 held constant during wide swings of as much as a factor of 4 in the SO_2 concentration of the entering flue gas. Inspection of the packing after the run showed it to be clear with only an insignificant film of salt on the packing.

For subsequent runs the No. 3 Intalox packing was replaced by No. 2 packing. A new liquid distributor and pumps of higher capacity also were installed. The mist eliminator at the top of the scrubber was replaced by the Heil blade mist eliminator from the top of the cyclone separator. Runs CS-3A, 3B and 4 were carried out in sequence for a period of continuous operation of 6.5 days. During this period, no increase in pressure drop over the packing was observed. The mist eliminator remained clean and no appreciable foaming or carryover of scrubbing solution was noted. The only difference between Runs CS-3A and 3B was the higher liquid circulation rate employed in the latter run.

The use of the improved distributor and the smaller packing increased SO_2 absorption from 88.0 to 96.2%. Increasing the liquid circulation rate in Run CS-3B raised the absorption of SO_2 to 98%. The use of the higher gas rate increased the SO_2 absorption to 99%, but only at the expense of a considerable increase in pressure drop.

The sulfate formation was small throughout all of these runs and averaged about 1% of the SO_2 absorbed. No appreciable NO_x absorption could be detected within the precision of measurement. Mass transfer coefficients for SO_2 absorption were in the same range reported by Johnstone and Singh in 1937[6].

The rubber lining in all the vessels was in good shape after this operation. Corrosion information was also collected via strips and spools in the system. Based on these short-term results (518 hours), the choice of materials would be among rubber-lined carbon steel, type 316 stainless steel, and fiberglass reinforced pipe.

Closed-Loop Operations

The closed-loop run was conducted over a steady operating period of ten days. During this period, the pressure drop through scrubber and demister remained quite constant. A mesh-type demister was used in the Library scrubber and, in contrast to the experience at Cromby, no plugging problem was encountered. It is concluded from this and other evidence that the plugging of the Cromby demister was due primarily to the entrapment of fly ash.

The sulfate yield averaged slightly less than 1 mol % of the SO_2 absorbed. Only a fraction of the sulfate was removed by the line filters

in the scrubber product line (cf. Figure 1) or the regenerated feed system (cf. Figure 2). The major portion of the sulfate settled out in the regenerated product tank. It is at this point in the system that the sulfate is least soluble. Recent work has demonstrated continuous rapid removal of this sulfate after the flash drum via low-pressure filtration.

The only materials added or removed during the run were water and a small amount of K_2CO_3 added to make up for the K_2SO_4 removed. In a commercial process, no dilution water need be added except to humidify the gas, since the condensate liquor would be returned to the product salts (cf. Figure 2).

The initial inventory was turned over twelve times in the course of this run. No foreign materials accumulated in the product. The compositions of all the feed streams, their rates, temperatures and pressures are given in Table 4. The stream numbers correspond to those given in Figures 1 and 3. The figures given in Table 4 are those midway through the operation, i.e., after five days. Sulfur absorption was high throughout this run and averaged about 99%.

Potassium formate is one of the components of the product but is not in reality a foreign material. The amount present is controlled, other conditions being constant, by the percent conversion of the thiosulfate. The higher the conversion, the higher is the formate content of the product. The formate content of the regenerator product shown in Table 4 was, for example, 4 wt. % and the thiosulfate conversion was 95 wt. %. Five days later, at the end of the run, the formate concentration was reduced to 1.7 wt. % when the thiosulfate conversion was decreased to 88 wt. %.

PROCESS ECONOMICS

A cost estimate based on an engineering study for a specific site is shown in Table 5. The scrubbing section is designed to handle the full output from a 200 MW unit with a heat rate of 9,000 Btu/KWH and 35% excess air in the stack gas to be scrubbed. The regeneration section provides the capacity to handle coal with up to 3.5 wt. % sulfur (at 12,000 Btu/lb) when the heat rate is 9,000 Btu/KWH and the average load factor of the utility is 80%. The overall sulfur removal is 93% after the incinerated Claus tail gas has been added to the scrubbed stack gas.

The costs do not include charges for venturis and fly ash removal. Retrofit costs are included. The costs were developed for 1977 startup at a time when the rate of inflation was quite high. The investment estimates correspond essentially to a Chemical Engineering Plant Cost Index of 200. The operating costs also reflect estimates for 1977 startup. Sufficient details have been given in Table 5 to adjust the investment and operating costs to other bases as desired.

The engineering study also may be used to estimate the cost at a new 1000 MW coal-fired power station. On the same basis, i.e., 1977 startup and firing 3.5% sulfur coal, the estimated investment and operating costs are 70 $/KW and 4.1 mills/KWH, respectively.

ACKNOWLEDGMENT

Appreciation is expressed to the utility companies who provided a large part of the funding for this work. These are: Allegheny Power, Consumers Power, Detroit Edison, Niagara-Mohawk, Ontario Hydro and Philadelphia Electric. Appreciation is also expressed to C. Itoh & Company, Inc., for their financial contribution to the work reported here.

Appreciation is also expressed to the Consolidation Coal Company Management for their support of this work and their permission to publish the results.

A large number of engineers and chemists participated in this project and the success of the project was largely due to their efforts. Specifically noteworthy were the contributions by M. D. Kulik, E. B. Klunder, P. J. Dudt, W. E. McKinstry, and N. J. Mazzocco. The cooperative spirit of George Kotnick, J. A. Gille, E. G. Boyer, Jr., and M. W. Moore of the Philadelphia Electric Company during the work at Cromby Station was deeply appreciated.

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Table 1, SUMMARY OF CONDITIONS AND RESULTS
CROMBY SCRUBBING TESTS

Run Number	CS-2	CS-3A	CS-3B	CS-4
Packing Size	No. 3	←	No. 2	→
Flue gas entering venturi, ACFM	20,900	21,100	21,300	31,900
Flue gas leaving scrubber, ACFM	18,200	18,200	18,400	28,500
Liquid recirculation rate, gpm	205	190	250	255
Length of run, hours	240	36	24	94
Superficial gas velocity in scrubber, ft/sec.	6.0	6.0	6.1	9.5
Temperature, °F				
Flue gas entering	300	310	310	317
Gas from venturi	128	136	136	132
Gas from stack	130	138	138	140
Liquid to packing	146	150	151	157
Liquid from packing	142	148	149	140
Pressure Drop, Inches of Water				
Over venturi	11	11	12	25
Over packing	0.5	1.2	1.3	5.0
Total	22	15	16	38
pH of scrubber effluent	7.5	7.4	7.6	6.9
SO_2 absorption, %	88.0	96.2	98.0	99.1
Mol % of SO_2 absorbed producing sulfate	0.9	1.3	1.4	1.0
Calculated value, $K_G a$, lb mol/hr ft^3 atm	21	33	40	74

Table 2, COMPOSITION OF FEED AND PRODUCT SOLUTIONS FROM CROMBY PILOT OPERATION

Run Number	CS-2	CS-3A	CS-3B	CS-4
Average Feed Composition, Wt.%				
KHS	15.5	15.3	15.4	15.4
K_2CO_3	31.9	32.3	32.5	32.3
Acceptability Ratio	0.95	0.94	0.93	0.94
Product Composition, Wt.%				
$K_2S_2O_3$	54.20	54.60	55.60	32.50
$KHSO_3$	1.34	1.20	1.33	1.54
K_2SO_4	0.66	0.92	1.00	0.42
K_2CO_3	0.2	0.2	-	0.1
Fly Ash	-	-	-	0.008
Total salts (by drying), Wt.%	57.6	57.8	59.5	35.3
Distribution of Sulfur, Mol % of absorbed SO_2				
$K_2S_2O_3$	96.2	96.2	95.9	93.7
$KHSO_3$	2.8	2.5	2.7	5.3
K_2SO_4	0.96	1.33	1.40	0.99

Table 3, FLUE GAS ANALYSES CROMBY PILOT OPERATION

Run Number	CS-2		CS-4	
	Intake	Stack	Intake	Stack
Dry Gas Composition, Vol.%				
CO_2	14.1	14.2	14.5	14.5
O_2	4.1	4.2	4.3	4.3
N_2	80.5	80.6	80.2	80.2
Argon	1.0	1.0	1.0	1.0
SO_2, ppm	2975	337	2010	18
Wet Gas				
Vol.% H_2O	-	-	6.7	14.6

Table 4, OPERATING CONDITIONS AND STREAM COMPOSITIONS, CLOSED LOOP OPERATION

	Scrubbing Section		Regeneration Section		
Gas Stream No.	1	3	16	17	18
Flow, SCFH	45,000	47,600	430	704	148
Temperature, °F	350	131	403	220	219
Pressure, psig	0.7	0	595	595	26
Gas velocity, ft/sec	-	7.3	-	-	-
Composition, Vol.%					
CO	0.0	0.0	792	7.4	0.0
H_2	-	-	0.3	0.2	8.2
CO_2	12.0	11.3	19.9	39.3	8.2
O_2	3.3	3.1	0.0	0.0	0.0
H_2S	0.0	0.0	0.0	14.1	8.8
SO_2, ppm	2200	15	0.0	0.0	0.0
N_2 + Argon	74.5	69.1	0.6	4.7	24.2
H_2O	10.0	16.5	0.0	34.3	58.6
% SO_2 Removal	← 99.5 →				

Liquid Stream No.	7	8	9	10	11	13	14	15	19
Flow, gal/hr	648	648	10.6	-	-	-	-	-	-
Flow, lb/hr	-	-	126	-	123	40	105	28	105
Temperature, °F	144	140	149	70	144	70	150	-	-
pH	7.4	6.8	-	-	7.4	-	-	-	-
Composition, Wt.%									
H_2O	49.7	-	37.8	-	49.7	100	51.5	100	57.7
$K_2S_2O_3$	41.5	-	1.8	-	41.5	-	39.7	-	2.2
KOOCH	5.7	-	4.6	-	5.7	-	6.0	-	4.0
$KHSO_3$	0.9	-	0.0	-	0.9	-	0.8	-	0.0
KHS	0.0	-	7.7	-	0.0	-	0.0	-	6.9
K_2SO_4	0.8	-	0.4	-	0.8	-	0.8	-	0.9
K_2CO_3	1.4	-	9.2	47	1.4	-	1.2	-	10.1
$KHCO_3$	0.0	-	15.5	-	0.0	-	0.0	-	14.5
K_2S_2	0.0	-	3.0	-	0.0	-	0.0	-	3.7
Acceptability Ratio	-	-	0.97	-	-	-	-	-	0.97
Absorbed Sulfur Converted to Sulfate, mol %	0.95								

Table 5, ESTIMATED COSTS FOR 200 MW PROTOTYPE, 1977 START-UP

Investment Summary	MM$
Scrubbing and reheat	7.1
K_2SO_4 Recovery	0.5
Regeneration, including oxygen plant, CO generation, acid gas separation and Claus plant	14.3
Support facilities, including storage and boiler feed water, cooling water, process water, and waste water systems	2.8
Total	24.7

Operating Costs	Annual Quantity	Rate	Annual Costs, MM$/Yr	Mills KWH
Operating Labor, incl. Benefits	45,990 hrs	$10.13/hr	0.466	
Supervision, incl. Benefits	8,760 hrs	$10.93/hr	0.096	
Ind. O.H. & Admin., Incl. Benefits	(55% of above)		0.309	
Electricity	22,118 MWH	25$/MWH	0.553	
High-Sulfur Fuel Oil	721,000 MM Btu	3$/MM Btu	2.163	
KOH	898 Tons	355.60$/ton	0.319	
Claus Catalyst	13.9 Tons	200$/ton	0.003	
Misc. other Chemicals	-	-	0.025	
Maint. & Misc. Supplies	(5% of $24.7 MM)		1.235	
Insurance	(0.175% of $24.7 MM)		0.043	
Int. on Working Capital	(10% of $1.43 MM)		0.143	
Sub-Total			5.355	3.82
Capital Charges	(16% of $24.7 MM)		3.952	2.82
Sub-Total			9.307	6.64
Credits:				
Sulfur	16,013 LT	25$/LT	-0.400	
K_2SO_4	982 Tons	42.50$/ton	-0.042	
Sub-Total, Credits			-0.442	-0.32
Net Total			8.865	6.32

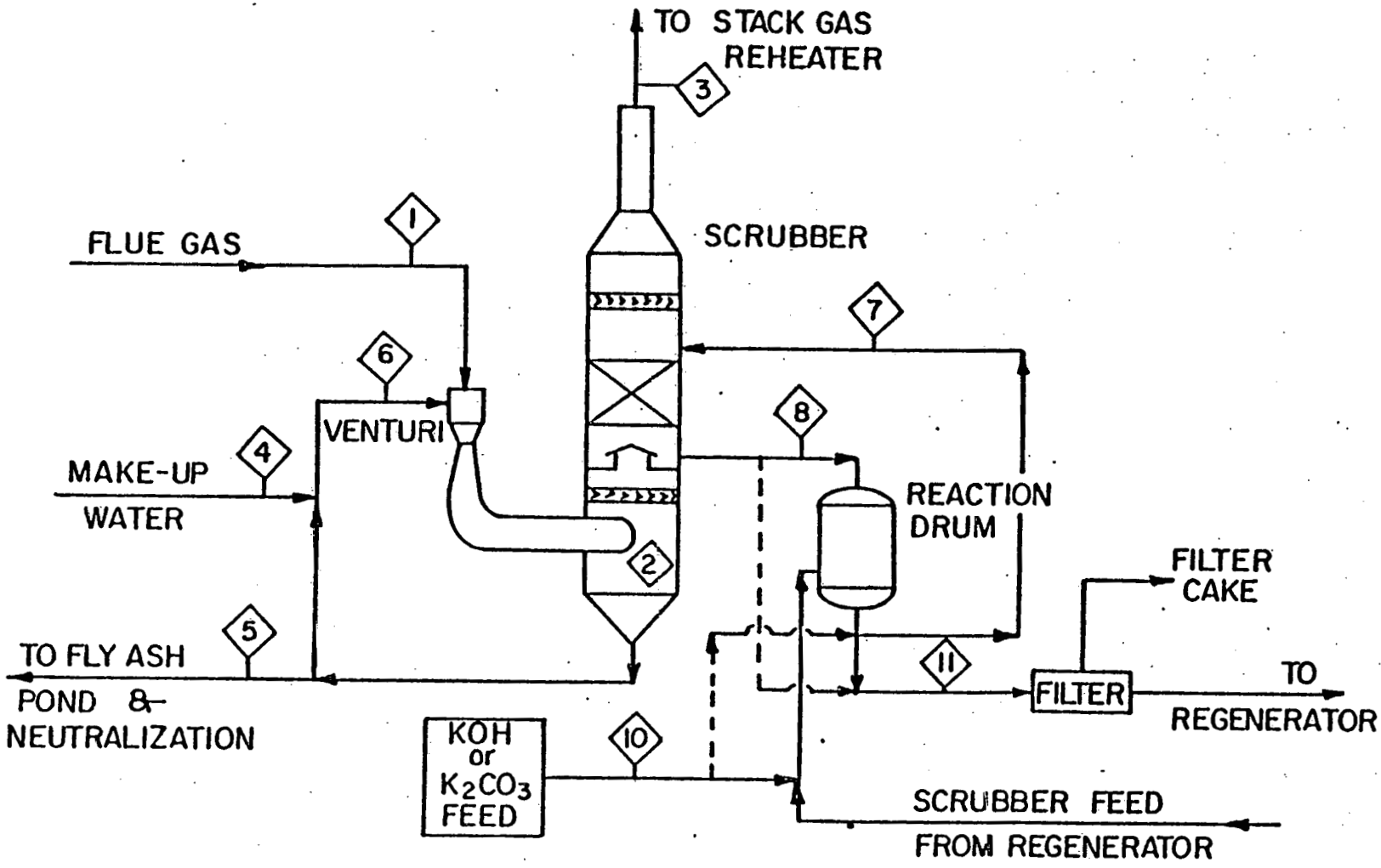


Figure 1 Schematic flow diagram, scrubbing section.

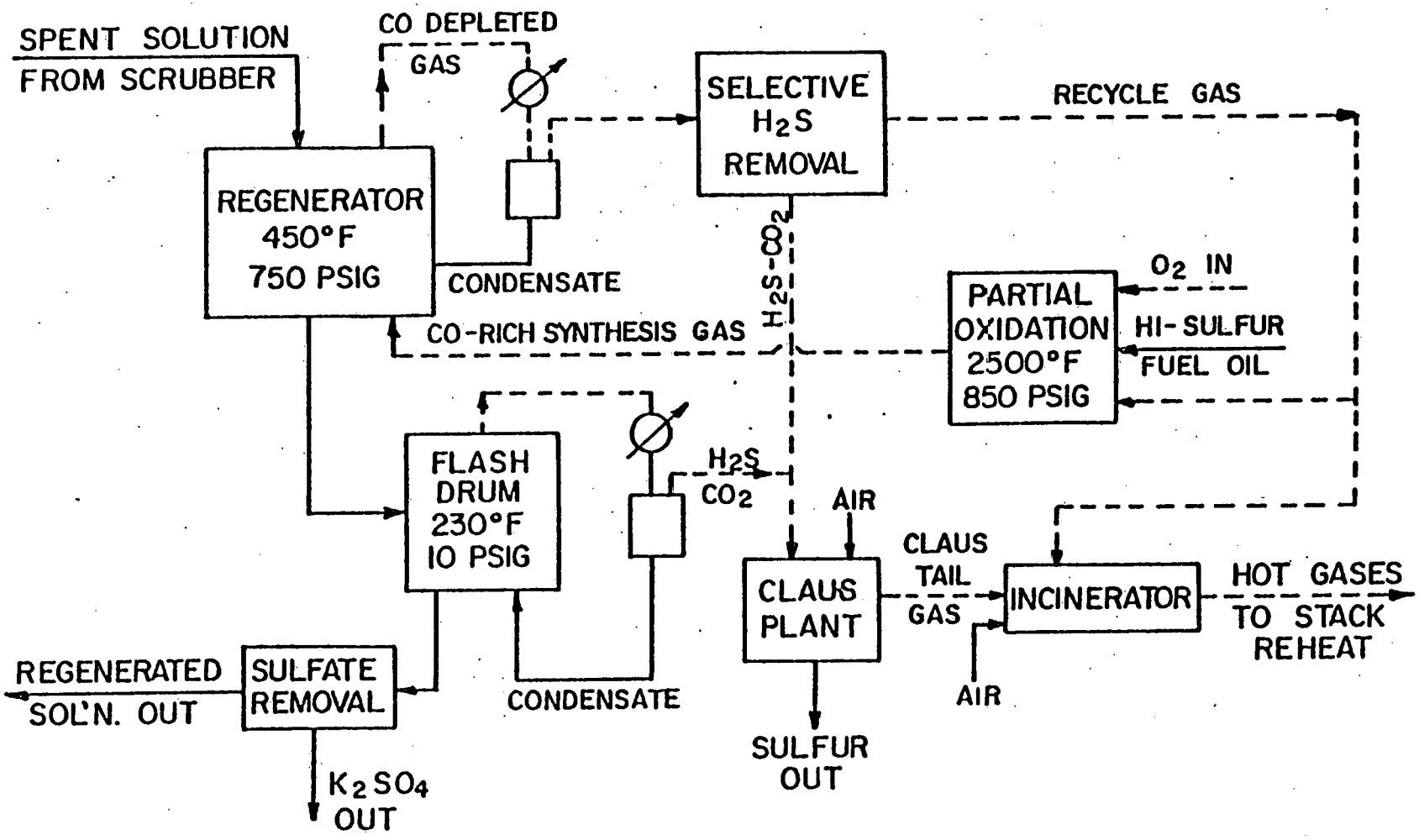


Figure 2 Block diagram, commercial regeneration process.

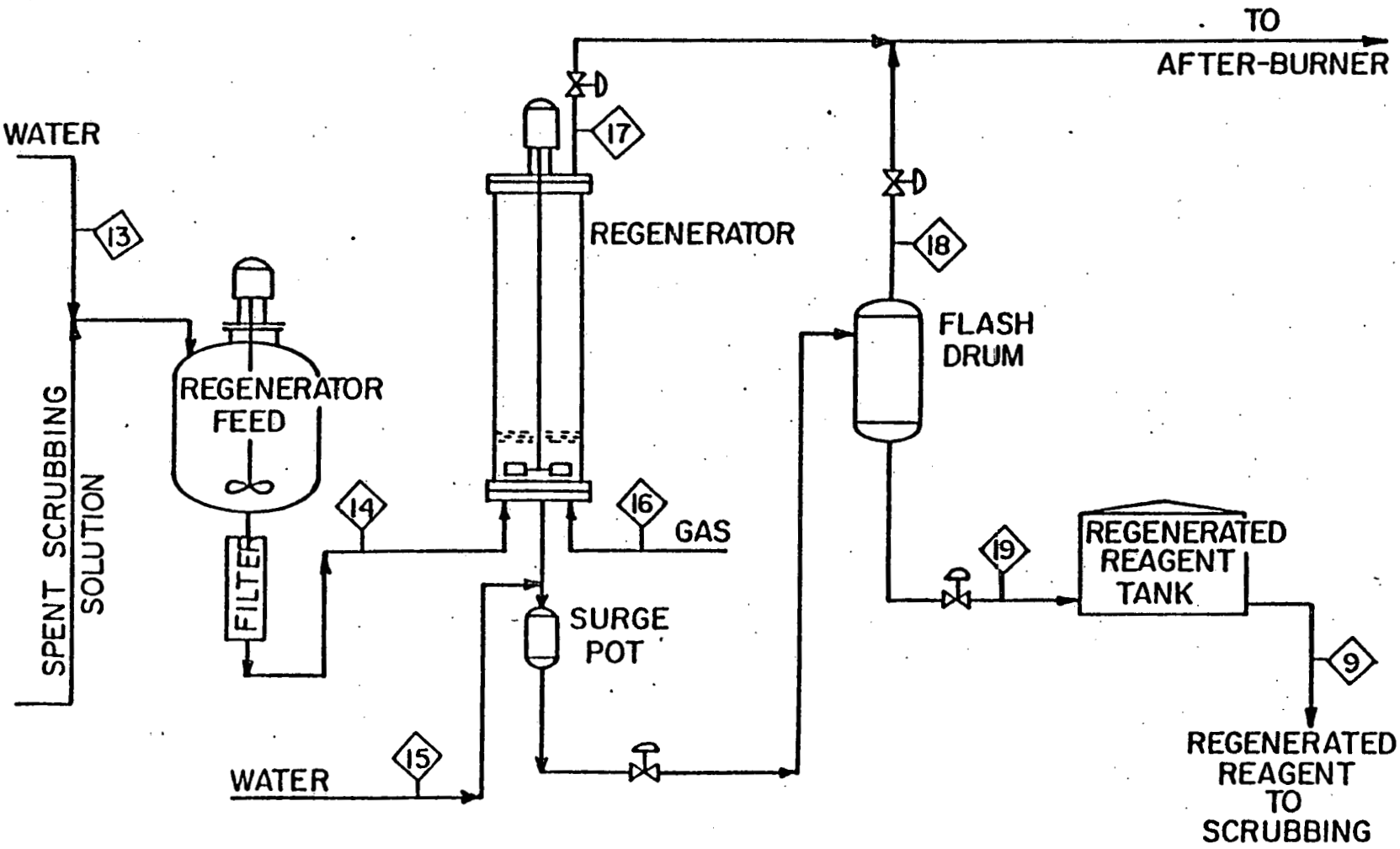


Figure 3 Schematic flow diagram, pilot regeneration.

MOBIL PROCESS FOR THE CONVERSION OF METHANOL TO GASOLINE

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Presented at the Third Annual International Conference on Coal Gasification and Liquefaction, University of Pittsburgh, Pittsburgh, Pennsylvania, August 3-5, 1976.

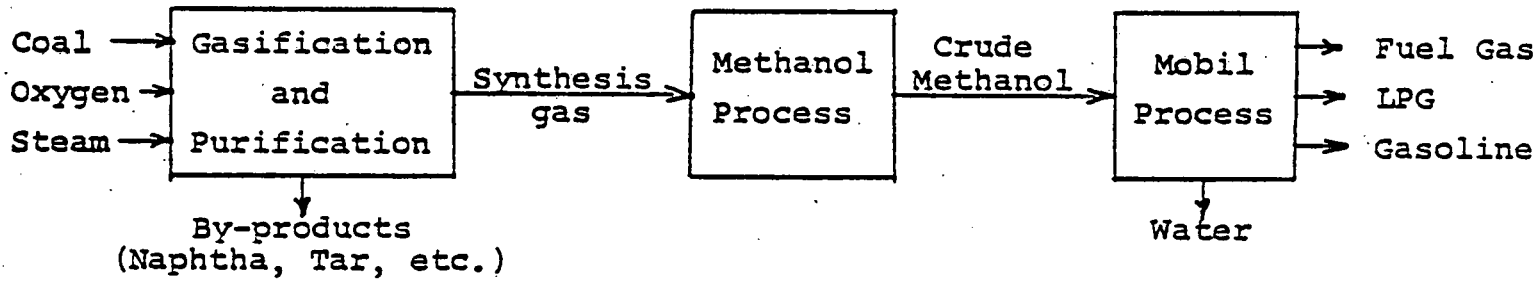
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INTRODUCTION

Mobil Research and Development Corporation is developing a process for converting methanol into a high octane, high quality gasoline.[1] Conversion of coal to methanol is well established commercial technology.[2] Thus, coupling the Mobil Process with coal to methanol technology provides a new route for the conversion of coal into gasoline. As shown in Figure 1, coal can be gasified to form synthesis gas. After purification, the synthesis gas can be converted to methanol. Crude methanol can be then converted to gasoline and water. It is not necessary to remove the water or the small amounts of other oxygenates which are present in the crude methanol stream coming from the methanol synthesis reactor. Small amounts of LPG and high BTU fuel gas are made as by-products. The Mobil Process is a relatively small part of the overall process for converting coal to gasoline via methanol.

Figure 1

Manufacture of Gasoline from Coal



Methanol can be produced from coal and blended directly into gasoline or used by itself as an automotive fuel. However,

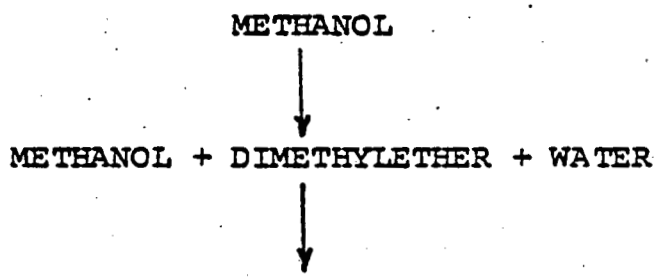
methanol's affinity for water, its corrosiveness, toxicity, and low volumetric energy content present formidable obstacles to its use either alone or as component of gasoline.[3] Large investments would be required to modify engines, storage and distribution facilities, and fuel and lube additive formulations. We calculate that it would be much cheaper to convert methanol into gasoline than to use it in a blend with gasoline.

To further the development of this process, Mobil entered into a cost-sharing contract with the Office of Coal Research and later the Energy Research and Development Administration in January, 1975, to scale-up this process, to study product quality, and to design a 100 B/D pilot plant.

THE MOBIL PROCESS

The conversion of methanol to hydrocarbons and water is highly exothermic. The heat of reaction is 740 BTU/lb. The principle problem in designing a reactor system is heat removal. To minimize the problem, the reaction can be divided into two steps, as shown in Figure 2. In the first step methanol is partially dehydrated to form an equilibrium mixture of methanol, dimethylether, and water. About 20% of the heat of reaction is released in this first step. In the second step this equilibrium mixture is converted to hydrocarbons and water.

Figure 2



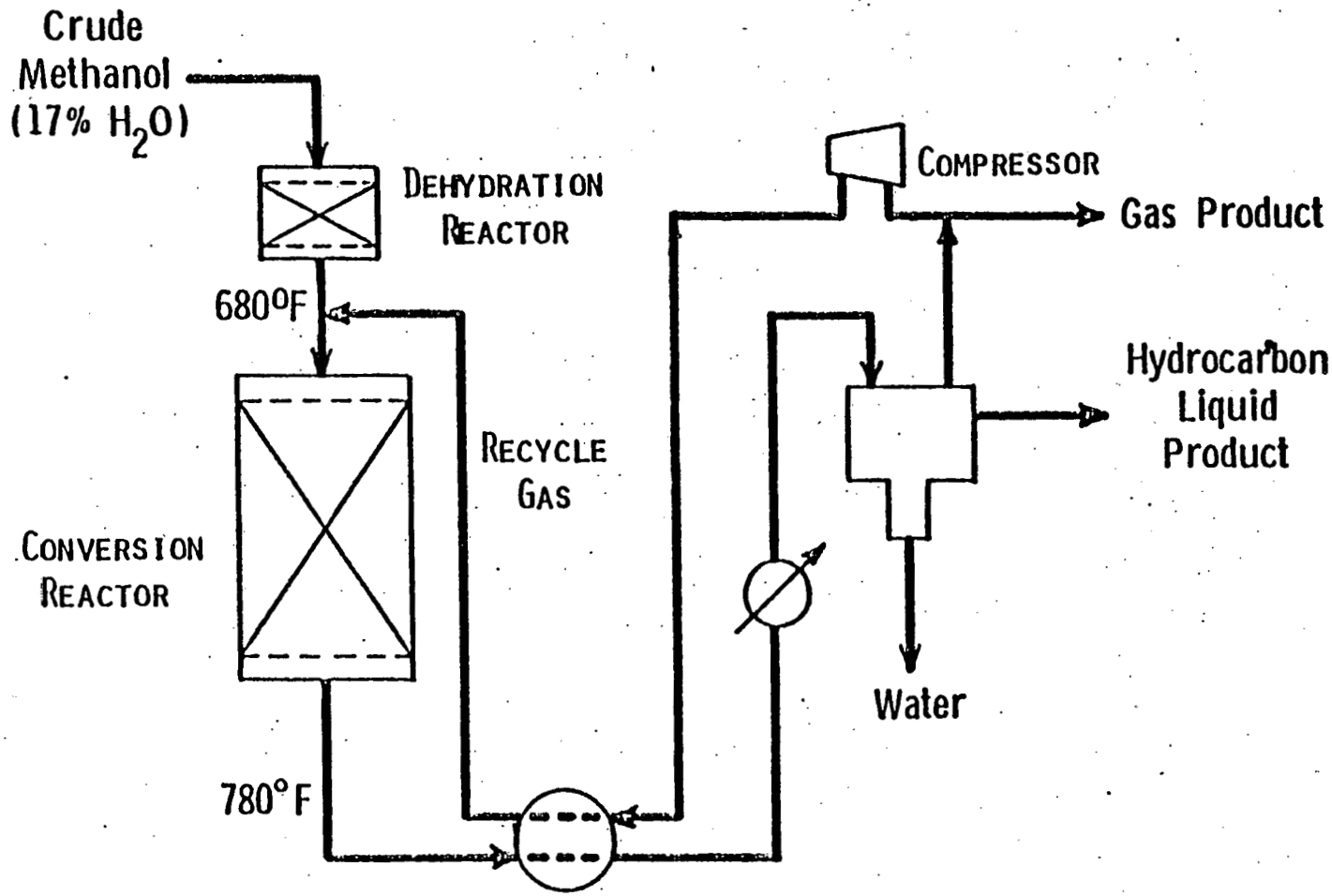
Heat of reaction = 740 BTU/lb

The equipment used in the Mobil Methanol-to-Gasoline Process is similar to a petroleum hydrotreater, as shown in Figure 3. Crude methanol is first passed through the dehydration reactor to form the mixture of methanol, dimethylether, and water. The effluent is diluted with recycle gas and passed through the conversion reactor. The recycle gas provides a heat sink to pick up the heat of reaction and limit the temperature rise. The reactor inlet temperature is typically 680°F. With a 9/1 recycle ratio, the adiabatic temperature rise is about 100°F.

The reactor effluent is condensed, the water and liquid hydrocarbon phases are separated, and the gas is recycled. The liquid and gaseous hydrocarbons are processed through a conventional gas plant. To make additional gasoline, the propylene, butenes, and isobutane are separated and alkylated using conventional petroleum technology.

FIGURE 3

MOBIL METHANOL TO GASOLINE PROCESS



Typical process conditions and yields are shown in Table 1. Methanol is converted to hydrocarbons (44% wt) and water (56% wt) in the expected stoichiometric amounts. Small amounts of CO, CO_2, and coke are formed as by-products. After several weeks on stream, the coke accumulates on the catalyst to the extent that regeneration is required. Activity is restored by burning the coke off the catalyst.

The hydrocarbon product is primarily gasoline. After the butenes and propylene are alkylated and the gasoline is pressured to 9 pounds Reid vapor pressure with the butanes, the finished gasoline has an unleaded Research octane of 93. The yield on the basis of the total hydrocarbons produced is 85.0% wt. The amount of alkylate is limited by the yield of propylene and butenes. Excess isobutane is put into LPG. Yield of valuable LPG is 13.6% wt. The remaining 1.4% of light gases could be used as high BTU fuel gas or blended into a SNG stream.

TABLE 1

Yields From Crude Methanol

(Charge is methanol with 17% wt H_2O)

Temperature, °F		
Inlet	680	
Outlet	780	
Pressure, psig	300	
Recycle Ratio (Mole)	9.0	
Space Velocity (WHSV)	2.0	
Yields, Wt % of Charge		
Methanol + Ether	0.0	
Hydrocarbons	36.2	
Water	63.3	
CO, CO_2	0.3	
Coke, Other	0.2	
	100.0	
Hydrocarbon Product, Wt %		
Light Gas	1.4	
Propane	5.5	
Propene	0.2	
i-Butane	8.6	
n-Butane	3.3	
Butenes	1.1	
C_5^+ Gasoline	79.9	
	100.0	
Gasoline (including alkylate)	85.0	(93 R+O, 9 RVP)
LPG	13.6	
Fuel Gas	1.4	
	100.0	

PRODUCT QUALITY

The gasoline produced in the Mobil Process is very high quality. As shown in Table 2, octanes exceed the current requirements for both unleaded regular or leaded premium. The composition of the gasoline is similar to conventional high quality gasoline. It consists of highly branched paraffins (51%), highly branched olefins (13%), naphthenes (8%), and aromatics (28%). As expected, the gasoline contains no sulfur or nitrogen, since there is none in the methanol feed. Also, the gasoline contains no oxygenates. Its boiling range is typical of premium gasolines.

There is one unusual factor which presented a potential quality problem. This is the presence of durene in the gasoline. Durene is the C_{10} alkylbenzene shown in Table 3. It is the largest molecule produced in any significant quantity when methanol is converted to gasoline in the Mobil Process. It has a high octane (estimated at over 100 R+0), boils in the gasoline range (386°F), but has a freeze point of 175°F. Durene is present in conventional gasoline in very small amounts. In the gasoline synthesized from methanol it can be present in relatively large amounts (3-6% wt of the gasoline), depending upon the process conditions used.

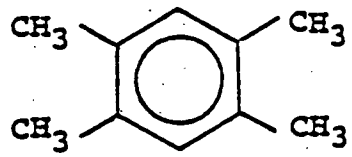
TABLE 2

Typical Properties of Finished Gasoline

Components, % Wt	
Butanes	2.7
Alkylate	3.2
C_5^+ Synthesized Gasoline	94.1
	100.0
Composition, % Vol	
Paraffins	51
Olefins	13
Naphthenes	8
Aromatics	28
	100
Research Octane	
Clear	93
Leaded, 3 cc TEL/USG	100
Reid Vapor Pressure, psig	9.0
Specific Gravity	0.730
Sulfur, % Wt	Nil
Nitrogen, % Wt	Nil
Corrosion, Copper Strip	1A
ASTM Distillation, °F	
10%	114
30%	145
50%	198
90%	330

TABLE 3

Durene



Boiling Point, 386°F

Freezing Point, 175°F

Because of this high freeze point, we were concerned that it might cause problems by crystallizing out in the carburetor of the engine. Indeed, we found that durene levels of 5% wt in gasoline did cause unsatisfactory engine operation under certain test conditions in some cars. However, at a 4% level, durene effects were minimal. Durene levels can be maintained at acceptable levels by proper control of the process parameters. Alternatively, gasoline containing larger than the desired durene levels could be blended with conventional gasoline.

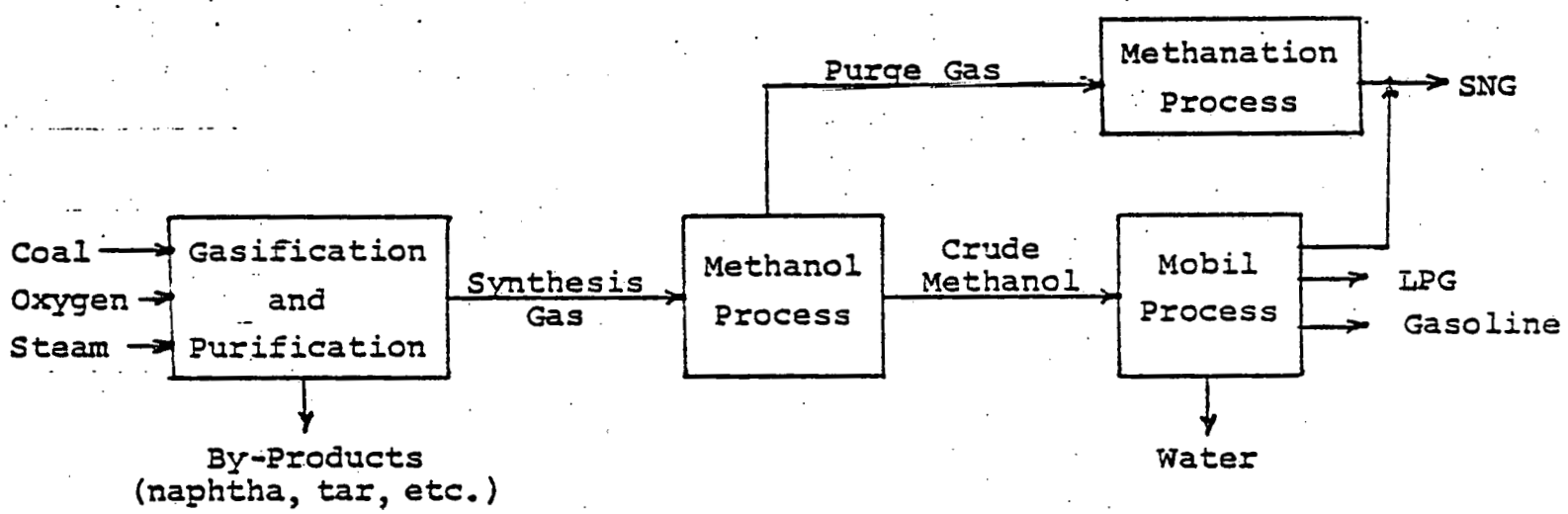
COPRODUCTION OF GASOLINE AND SNG

The coproduction of methanol and SNG would be cheaper and more thermally efficient than production of methanol alone.[4,5] The advantage of coproduction stems from the retention of the methane produced in the gasification step as SNG rather than converting it to additional synthesis gas. By adding the Mobil

Process to convert the methanol to gasoline, an efficient process for the production of both SNG and gasoline results. As shown in Figure 4, synthesis gas containing methane would be used as feed to the methanol process. The purge gas from the methanol unit would be methanated to convert the unreacted CO and H_2 in the gas to additional methane. The Mobil Process would be used to convert the crude methanol to gasoline, LPG and a small amount of additional SNG.

FIGURE 4

Coproduction of Gasoline and SNG



This combination process has high thermal efficiency, as shown in Table 4. It is interesting to note that gasoline and SNG can be produced at almost the same efficiency as SNG alone. Lurgi has reported[5] that the thermal efficiency for converting

coal to pipeline gas plus by-products (naphtha, tar, etc.) by their process is 68.8%. To produce methanol only, it would be 56.2%.

TABLE 4

Thermal Efficiency

$$\eta = \frac{\Sigma \text{ Products (HHV)}^*}{\text{Coal (Process + Utility) (HHV)}}$$

Product	η
Pipeline Gas + By-products	68.8 (5)
Methanol Only	56.2 (5)
Methanol + SNG + By-products	68.7 (5)
Gasoline + LPG + SNG + By-products	65.5

*HHV = Higher Heating Value

However, to produce both methanol and SNG, plus by-products (naphtha, tar, etc.) it would be 68.7%. In other words, the efficiencies of producing pipeline gas or methanol plus pipeline gas are essentially equal. If the methanol produced in the co-process, as described by Lurgi, were to be converted into gasoline by the Mobil Process, we estimate that the efficiency would be 65.5%. This is based on a thermal efficiency of 88% (higher heating value) for the Mobil Process. The efficiency for the Mobil Process includes the associated gas plant and alkylation facilities, as well as all utilities. Any improvement in the efficiency of methanol production as the result of improvements

in gasification or methanol technology would, of course, increase the overall efficiency of making gasoline from coal by the Mobil Process.

The product distribution on a BTU basis from this co-production case is shown in Table 5.

TABLE 5

	Product Distribution, % (BTU Basis)
SNG	46
LPG	5
Gasoline	31
By-products	18
	100

SNG, including the small amount of SNG produced in the Mobil Process, represents 46% of the energy content of the products. Gasoline is 31%, LPG 5%, and the by-products (naphtha, tar, etc.) make up the remaining 18%. In this case, the ratio of SNG to gasoline is 1.5. Currently, the ratio of natural gas to gasoline used in the United States is about 1.9[6], but is expected to decline. The ratio of SNG to gasoline can be varied to some degree by appropriate changes in the methanol process design.

We believe the Mobil Process with coproduction of SNG compares quite favorably with other means of producing gasoline from coal. This is especially true when you recognize that most

direct liquefaction schemes produce a synthetic crude which requires substantial further refining to make a high quality gasoline. In summary, the Mobil Process:

- produces primarily premium products,
- can give high thermal efficiency,
- has relatively few process steps,
- can use proven technology for most of the plant,
- could be scaled up sooner and easier than the more complex and unproven direct liquefaction processes,
- would benefit from any future improvements in gasification or methanol technology.

ACKNOWLEDGEMENT

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THE SYNTHANE PROCESS: CURRENT STATUS

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INTRODUCTION

The SYNTHANE process is one of four integrated coal gasification processes for the production of high-Btu gas that are now being tested in large pilot plants. It was developed at the Pittsburgh Energy Research Center (PERC) of the Energy Research and Development Administration (ERDA), which was previously part of the Bureau of Mines. The decision to build the pilot plant was made about eight years ago and the first coal was gasified on July 7 of this year.

The pilot plant was designed by the Lummus Company, constructed by the Rust Engineering Company, and is operated by Lummus. The construction took two years and was completed in March of 1975. The design and construction cost was about $17 million. A 38-month contract was awarded to the Lummus Company for operation of the plant until October 1978 at a cost of $8.2 million per year. The plant now employs 200 people but this number will be reduced somewhat when some plant construction items are completed and initial startup problems are corrected.

Since the construction completion in March of 1975, many commissioning tests have been conducted. Numerous problems were encountered and corrected before the first light-off of the gasifier. "Light-off" refers to ignition of the coal in the gasifier. This report will describe the process and the pilot plant, the commissioning tests, some of the problems encountered and their solutions, and the experience gained in the first gasification test. The future plans for the plant will also be presented.

PROCESS DESCRIPTION

The SYNTHANE process is designed to operate at pressures up to 1000 psi with any of the coals or lignites available in the U.S. The overall thermal efficiency of the process has been calculated to be 65-70%. It is a fairly simple gasification process with no major recycle streams and therefore has the potential of operating with a high on-stream factor.

An overall flowsheet of the 72 ton per day pilot plant is shown in figure 1.

The first step in the process is coal preparation. Coal is ground in a hammer mill that is swept with hot flue gas to flash dry the coal. The coal is normally ground to pass completely through a 20 mesh screen with less than 30% through 200 mesh. Larger sizes can be used with the western coals which do not require pretreatment. Moisture content has been in the range of 7-8% for the Montana Rosebud coal being used in the pilot plant.

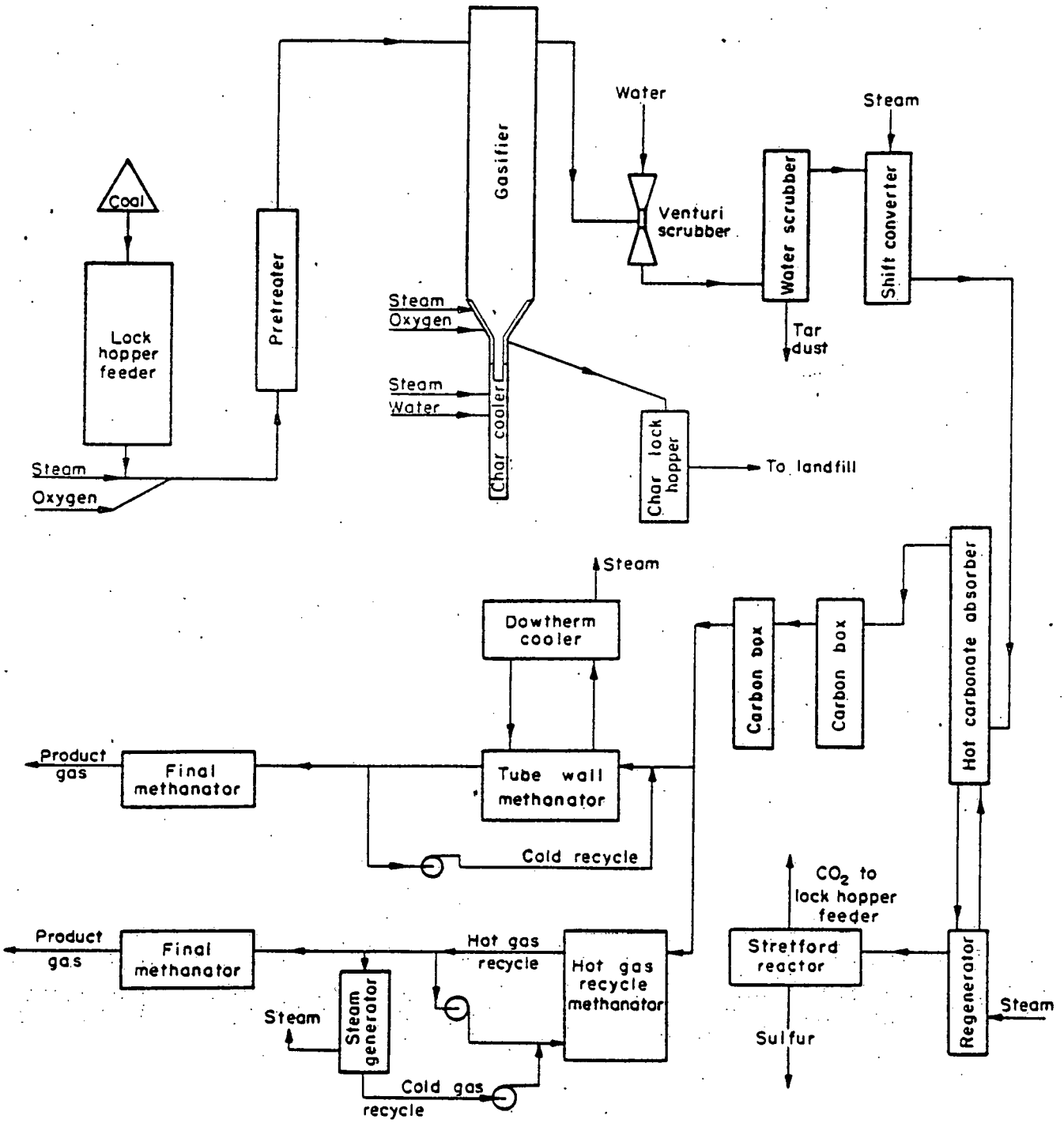


Figure 1. — SYNTHANE pilot plant flowsheet.

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Coal is then brought to operating pressure through a lock hopper system. The pilot plant has a Petrocarb system featuring a weigh hopper, twin lock hoppers with a one ton capacity each, and an injection hopper that serves as a high pressure feed hopper. Cross venting of the lock hoppers is used to conserve energy. The gas used is CO_2 recovered from the process. For a commercial design, a slurry feed system could also be used.

The pretreatment step must be used with the highly caking eastern coals to render them non-caking so that the gasifier can be operated with no agglomeration. This is a mild oxidation step using oxygen at about 430° C in a fluidized bed reactor with an approximate residence time of three minutes. The pretreater operates under pressure and the offgas from the pretreater mixes with the product gas from the gasifier. Level control is accomplished with a nozzle in the side of the pretreater which allows the coal to spill over into the gasifier.

Major features of the 72 ton per day SYNTHANE gasifier are shown in figure 2.

In the gasifier the coal is fluidized and gasified with steam and oxygen at temperatures up to 1800° F. The gasifier is about 100 ft high, the upper 69 ft of which is the gasifier proper. The bed height is normally 20 ft with a maximum of 29 ft. The coal is introduced 20 feet above the bed level and falls through this zone and undergoes partial carbonization as it is heated by the hot gas leaving the fluidized bed. The gasifier is five feet in diameter and is lined with a six-inch layer of light insulating castable refractory and three inches of a dense errosion resistant castable refractory. An Incolloy 800 shroud further restricts the inside diameter to 37 inches. The gases leave the gasifier through an internal cyclone located 43 feet above the bed. The cyclone height is necessitated by the high pressure drop in high pressure cyclones, resulting in a long dip-leg that is needed to return the low density char to the bed.

Char flows from the gasifier to a char cooler where it is fluidized with steam and cooled by spraying atomized water into the bed. The char is cooled to 600° F so it can be discharged into unlined lock hoppers pressurized with steam. The alternate char letdown system is now being used which slurries the char from the char cooler in water and throttles the slurry to atmospheric pressure through a Willis choke valve. The char is filtered and disposed of as a wet filter cake.

In a commercial plant the char produced will be burned to produce steam for use in the plant. With a carbon conversion of about 67% in the gasifier, the char produced is just sufficient to supply the energy needs of the plant. This technique of operating a "balanced" plant results in maximum utilization of the hydrogen content of the raw coal and reduces the size of the gasifier by saving the most unreactive part of the coal for simple high temperature combustion.

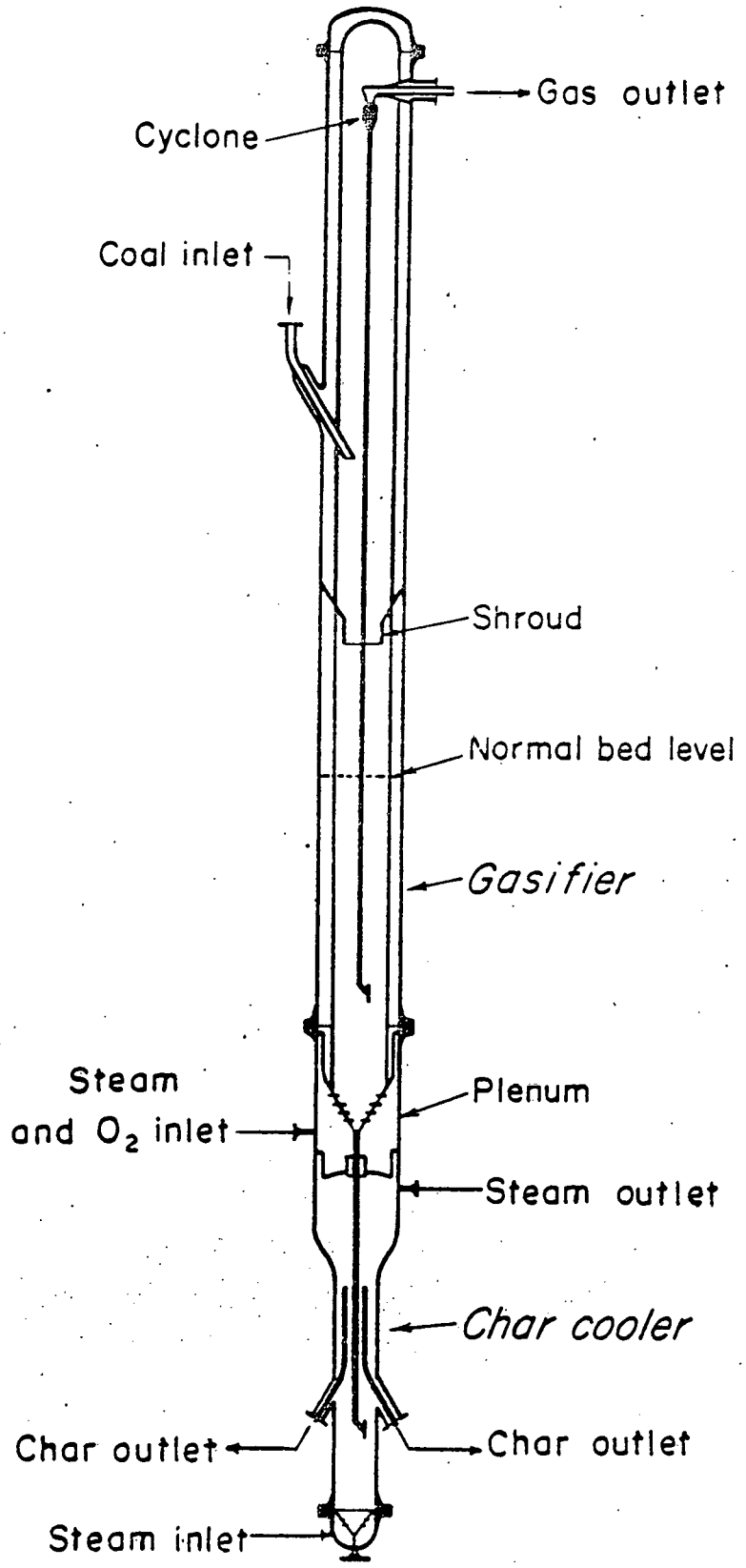


Figure 2. — SYNTHANE gasifier

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The gas leaving the gasifier is quenched and scrubbed with water in a "venturi scrubber" that more closely resembles a spray chamber. It is further scrubbed with water over baffle-trays and packing and then scrubbed with oil over baffle-trays. The oil wash is not being used initially.

The cleaned gas is then shifted from a H_2/CO ratio of about 2 to 1 up to a ratio of slightly over 3 to 1 necessary for methanation. A cobalt-molybdenum shift catalyst made by Catalysts and Chemicals Incorporated will be used initially.

The gas is then scrubbed of CO_2 and sulfur compounds in a Benfield hot potassium carbonate scrubber. The CO_2 - H_2S offgas from the regenerator is sent to a Stretford plant where H_2S is converted to elemental sulfur. A portion of the cleaned CO_2 is compressed and used in the lock hopper system to pressurize the coal.

A portion of the trace sulfur compounds, particularly thiophenes, will not be removed in the Benfield system so two activated carbon traps in series are used to reduce sulfur to below 0.1 ppm. The activated carbon to be tested initially (Katalco) contains copper and chromium impregnants.

Two methanation systems will be tested in the pilot plant. Each system can methanate about 1/8 of the purified gas. Both systems were developed at PERC and use a Raney nickel catalyst. They were described in detail at last years symposium.(1) One system operates isothermally with no recycle in a Tube Wall Reactor (TWR) in which the inside wall of the tube is coated with the catalyst and the heat of reaction is transferred to boiling Dowtherm on the outside of the tube wall. The other system is a Hot Gas Recycle (HGR) reactor which operates adiabatically with heat removed in the recycle stream. High pressure drop is avoided by coating parallel plates with catalyst - a low pressure drop configuration. A third reactor system has also been developed which will be tested in the pilot plant if PDU tests prove satisfactory. It is called a "Hybrid" reactor and some operating results will be presented shortly.(2)

The low level of CO that leaves the methanator as the catalyst deactivates will be converted in an adiabatic final methanator filled with pelletted nickel methanation catalyst. Initially, Girdler[3] modified G-87 RS will be used. The final gas will be about 950 Btu/ft^3 and contain less than 0.1% CO.

PILOT PLANT EXPERIENCE

Efforts to date have been directed toward a gasifier light-off with Montana Rosebud (subbituminous) coal at 600 psig. Downstream equipment such as the Benfield, Stretford, shift and methanation systems will be brought on line in subsequent tests. Prior to the light-off, cold fluidization and hot steam tests were made. Many of problems

were encountered which caused a considerable delay in the start-up. Solutions to nearly all of these have been implemented. A good many of these problems were mechanical, had nothing to do with coal gasification and could be encountered in any large plant. A few of the more important problems will be mentioned here.

Coal feeding and grinding proved to be less than straightforward. Numerous changes had to be made to feed Rosebud coal which is fairly wet (30% moisture), fairly fine and friable, and tends to combust spontaneously. The wet nature of this coal caused plugging in the inlet duct to the bucket elevator, the bottom boot of the elevator, the discharge chute from the top of the bucket elevator to the raw coal storage line and the chute from the weigh hopper to the impact mill. Alteration of chute angles and sizes, installation of breaker bars, and removal of the star feeder on the impact mill has allowed over 5 tons/hr to be fed and ground. A small fire occurred in the baghouse when an explosion door was left open during repairs and the residual dry reactive coal contacted the air, causing the bags to melt. To prevent freezing in the winter in the raw coal storage bin, steam coils have been added. The pulverized coal bin was equipped with steam coils which apparently caused water to vaporize from the coal near the steam coils. This steam then condensed on cold surfaces at the top of the bin and refluxed through the coal, causing "mud" to form at the base of the bin when it had not been used for some time. Shutting off the steam coils eliminated this problem and no freezing of the dry coal has occurred.

Established equipment which was supposedly standard too often failed. The bucket elevator chain failed shortly after it was first used and had to be replaced with a much heavier design. Numerous new instruments proved faulty when tested and had to be replaced by the manufacturer. Such problems are being encountered more and more frequently in American industry.

Several problems were discovered in the coal feeding system. When first tested, 16 of the ball valves leaked. They were all repaired and have operated satisfactorily since then. Level probes had to be redesigned and replaced. Several problems were found in the weigh hopper that prevented obtaining a reliable coal feed weight and these were corrected. Also numerous tramp parts, from welding rods to wire brushes, worked their way out of the system.

The transport line from the Petrocarb Primary Injector was designed to operate with steam as the transport medium. To prevent condensation when the steam contacts the cold coal, oxygen was to be added to cause ignition. Tests at PERC in a small unit showed that ignition of the coal probably would not occur, so a redesign of the system was necessary. Presently, CO_2 is being used as the transport medium. To reduce the

high use of CO_2 which could also possibly overload the Benfield system, the velocity in this line had to be reduced from over 50 ft/sec to below 18 ft/sec. Control of coal flow with the lower pressure drop in the lift line has been difficult, and flows much below 2 ton/hour are not considered controllable. Modifications will be made to introduce separately the coal and the steam - O_2 mixture into the pretreater. Since we are operating with Rosebud coal which does not need pretreatment, the pretreater currently is bypassed.

When the gasifier was first pressure tested with gas, a large leak in the welded flange gasket was found. The refractory was removed and the flange halves separated to make the repair. The entire system is now tight.

Modification of the venturi scrubber was necessary when an analysis showed that unacceptable stresses would occur when the top of the scrubber was contacted by the hot product gas. A new scrubber with a refractory lining was built and installed. When tested with no gas flow, it was discovered that the water could back up and run into the gasifier. This was corrected by extending the scrubber body by 3 feet to develop sufficient head so the water would flow out of the scrubber.

Numerous problems were encountered in the utilities and general service systems, such as the cooling water system, boiler, thermal oxidizer, compressors, and inert gas system. The inert gas system was replaced with a CO_2 storage and vaporizing system when extensive corrosion was found in the inert gas lines. This corrosion was presumably caused by acidic compounds and moisture from combustion of the No. 2 fuel oil (0.3% sulfur content) used to generate the inert gas. A propane fueled system with dryers is now being considered to reduce the cost of purchased CO_2.

Pipe rust was carried into the CO_2 compressors. It mixed with the oil to form an abrasive slurry and caused piston wear and valve failure. After installation of three suction filters, the operation of the compressors has been quite satisfactory. We have recently found that the 50 SCFM purge gas compressors also require suction filters and they are presently on order.

Another major problem was the discovery of off-specification piping in the small-bore piping systems. An extensive testing program was developed to test all the piping for schedule and materials. A sizeable portion of the piping was found to be both below and above specification. All unsuitable piping was cut out, replaced, and then hydrostatically tested.

In the heat-up tests it was found that the required starting temperatures could not be achieved with the inline burner, so the piping to the gasifier was significantly shortened, the pipe diameter was increased, and more insulation was added. Performance is now satisfactory.

During cold fluidization tests with CO_2, an instrument air failure caused the valve on the CO_2 feed line to fly open and result in over-pressuring the gasified plenum zone below the distributor cone. The internal head, which separates the plenum zone from the char cooler and was designed to withstand a pressure difference of 43 psi, was deformed from concave-up to concave-down. In addition, the expansion joint was stretched beyond its limits and failed. The deformed head could not take a high enough ΔP so it was strengthened by adding a series of knee braces which should allow it to take a 25 psi differential in either direction. Two rupture disks were then added to protect the head and the gasifier cone. The normal pressure difference is 3-5 psi with the char cooler higher in pressure. Control of this small ΔP is expected to be difficult.

Cold Fluidization Tests

Prior to the light-off, a series of cold fluidization tests were performed using ground Rosebud coal and CO_2 as a fluidization gas. Due to the limitation on the capacity of the CO_2 compressors, the tests were done at 75 psig. The tests showed that both beds were satisfactorily fluidized, levels could be controlled, and instruments worked as designed. Coal was ground and fed continuously through the Petrocarb system. It was fed through both beds continuously and taken from the char cooler to the high pressure slurry tank, then throttled and filtered with good success on the rotary disk filters. The slurry discharge system is being used because some of the valves in the char lock hopper system have been returned to the manufacturer for repair because of leaks. Operation of the filters with char is expected to improve over operation with raw coal because of the reduced solids flow (0.3 lb. char/lb coal) and the lighter nature of the char. During the cold fluidization tests, correlations of coal flow vs. pressure drop in the lift line were developed to serve as a guide to operation at higher pressure. Measurement of the particle size of the coal in the scrubber water showed that the internal cyclone was removing all particles above 20 microns.

Test Number 1

After the cold fluidization tests, a 600 psig pressurizing test was conducted with steam, and then test number 1 was started. The objective of the first operating campaign is to gasify Montana Rosebud subbituminous coal (20 x 0 mesh size) at 600 psig and 67 percent carbon conversion for a steady state period of 12 to 14 days. Initial gasifier conditions selected were: bed height, 10 ft; coal rate, 2 tons/ hr; fluidizing velocity, .3 ft/sec; and bed temperature, 1550° F. After the carbon conversion is determined for these conditions, gasifier conditions will be adjusted to obtain 67 percent carbon conversion. In this test, the gasifier product gas is water washed and then flared. Subsequent processing steps in the plant are omitted. Figure 3 is a flowsheet of the plant configuration for test number 1.

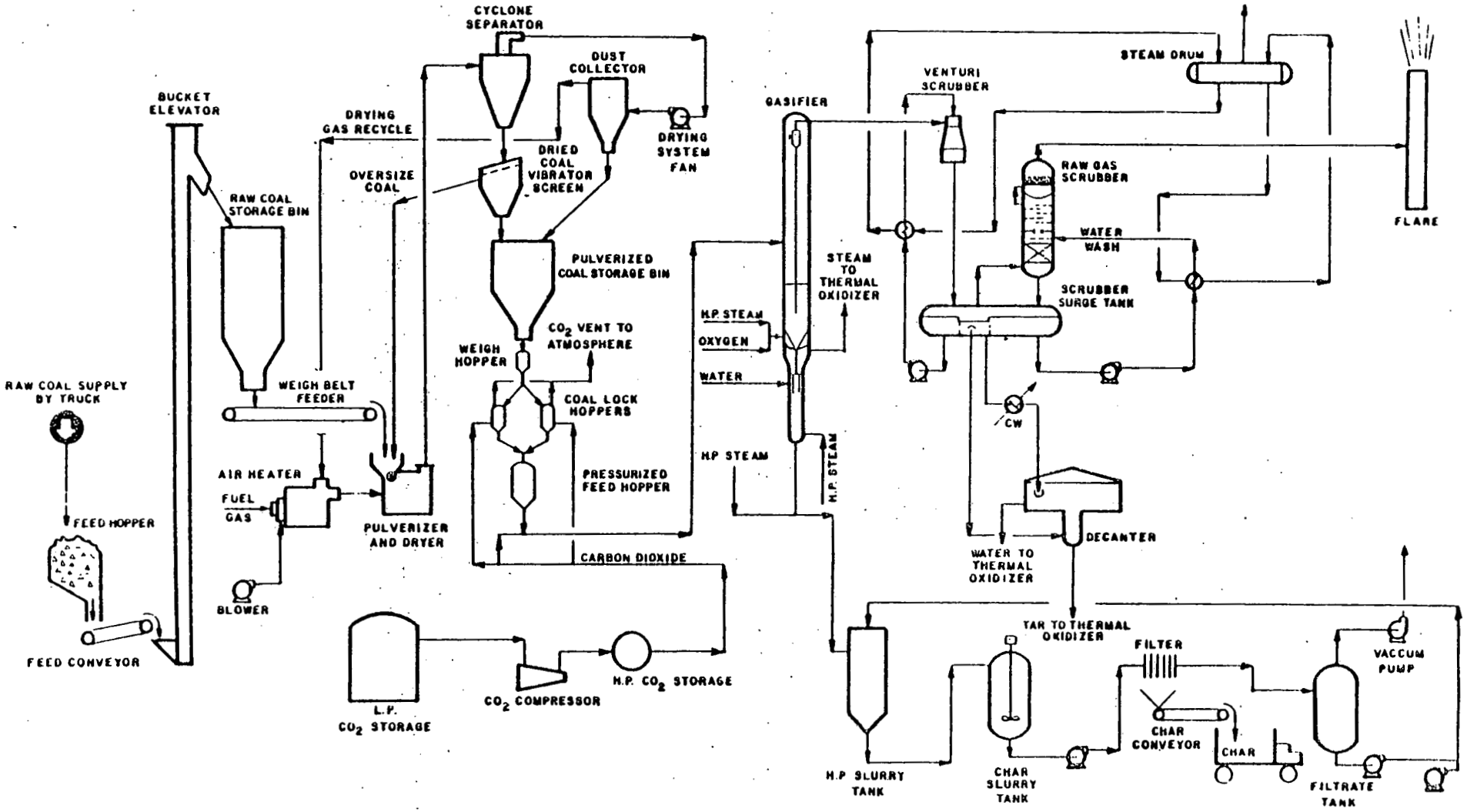


FIGURE 3.—SYNTHANE START-UP OPERATION

On July 7, 1976 the first coal was introduced. When the bed level in the gasifier reached 6 feet, oxygen was added and the temperature was increased to 1200° F. Due to an instrument problem, pressure control became erratic as char was dropped into the char cooler, causing the oxygen valve to trip and a rupture disk to blow. Steam was left in the gasifier, the disk was replaced and the instrument problem corrected.

Coal and oxygen were reintroduced on July 8 and the plant was operated for 7 hours before a lack of CO_2 purge gas and a plug in the char cooler outlet piping caused a shutdown. A bed temperature of 1440° F was reached and a few char and gas samples were obtained. The gasifier was allowed to cool while the line was unplugged and excessive flows of steam and CO_2 bleed streams were reduced. It was later found that coal and water had accumulated in the plenum zone causing a high pressure drop across the gasifier distributor cone. The gasifier was then cooled and opened to correct this problem.

It was found during this run period of test number 1 that the system was responsive and easy to control as far as oxygen addition was concerned. The gas scrubbing system worked well. The high pressure char letdown system through the Willis choke valves performed satisfactorily. The Petrocarb system worked quite well and later pressure tests revealed a leak in only one small vent valve. The one remaining problem was removal of char from the char cooler. At normal fluidizing velocities the char did not discharge satisfactorily and a low fluidizing velocity had to be used. Inspection of this section showed no apparent mechanical problem, so another attempt to discharge char was made when the run was restarted. Gasification was restarted on July 19 at a bed temperature of 1550° F. The system operated quite satisfactorily for an 11 hour period. Char removal from the char cooler worked quite well this time with the normal fluidizing velocity of 0.3 ft/sec in the char cooler. A problem was encountered, however, with an oversized differential pressure control valve which eventually caused a plug in the high pressure char slurry line and aborted the test.

The high pressure slurry line was unplugged, the control valve was changed, and the test was restarted on July 20. The gasifier temperature was then brought to 1650° F to increase the carbon conversion; however, the char did not discharge from the gasifier, necessitating termination of the run period. Visual examination of the resulting char or ash suggests that carbon conversion was excessively high and had significantly changed the flow characteristics of the gasifier bed.

A planned power outage on July 25 has necessitated shut-down of the plant. Upon resumption of operation, gasifier bed temperature will start at 1550° F and be increased incrementally to achieve 67 percent conversion.

The ground Rosebud coal from the pilot plant has been run in the PDU gasifier at PERC. The PDU gasifier is a 4" x 6' high fluidized bed that is operated at 40 atm. Results, which should approximate those expected from the 72 ton-per-day gasifier, are shown in Table 1.

Table 1. - Gasification of Montana Rosebud Coal - 40 Atm.

Max. Temp: 1663° F
Avg. Temp: 1583° F
Gas Composition:

H_2	36.9%
CO	18.2%
CH_4	13.0%
CO_2	30.5%
C_2H_6	1.1%
H_2S	0.4%

Tar Yield: 1.8% of coal weight
Steam Decomposition: 30.1%

Carbon Conversion: 67.7%

Fluidization Velocity: 0.32 ft/sec

Coal Throughput: 91. 4 lb/hr ft^3

	Coal	Char	Tar
Moisture	6.9	0.3	-
Volatile Matter	35.1	3.6	-
Fixed C	47.4	63.7	-
Ash	10.6	32.4	-
H	5.0	0.9	7.0
O	19.7	1.4	7.0
C	62.8	64.7	83.0
N	0.8	0.3	0.0
S	1.1	0.3	2.7

Future Plans

A series of tests has been planned which will subsequently bring all downstream equipment on line. Tests at 600 and 1000 psig will be made for periods of 7 weeks or less. Four coals will be tested: Montana Rosebud (subbituminous), a mildly caking, high sulfur Illinois No. 6, a highly caking, high sulfur Pittsburgh coal, and last a western lignite.

Three fairly large contracts for design and construction of equipment for testing in the pilot plant have been awarded. The A. G. McKee Company is designing a water treatment PDU. Presently, all the water is incinerated in a thermal oxidizer. This is impractical commercially because of the large energy requirement. The PDU will be designed to handle 1/4 of the gasifier condensate (or all of the condensate if the plant operates 25% of the time). Large tanks will store water for operation when the plant is down. Treatment steps to be included are: oil and tar separation, solvent extraction, free and fixed NH_3 stripping, biological oxidation, adsorption with gasifier char, and reuse of treated water in a cooling tower. The PDU will allow the best sequence of treatment steps and the treatment efficiencies of all steps to be determined. This data is essential to the design of a commercial treatment plant.

The Lummus Company of Bloomfield N.J. was awarded a contract to design and build an improved char letdown system. The system is designed to discharge char directly from the gasifier at 1800° F. This will eliminate the char cooler with its problem of control of small pressure differences, and it will also eliminate the char lock hoppers and the ball valves which must seal at 600° F and 1000 psia with char present. The char will be withdrawn from the gasifier and throttled through a series of specially designed orifices to atmospheric pressure. Char flow and bed level will be regulated by controlled steam addition. This system should use less steam and conserve the sensible heat in the char, all resulting in improved thermal efficiency.

The third contract is with Mechanical Technology Incorporated (MTI) for a centrifugal methanation recycle compressor that can operate at 565° F and 940 psig with a ΔP of 10 psi. This is not a commercially available item. The recycle compressor will replace the existing eductor used for hot gas recycle. The eductor can only operate at one fixed recycle ratio at a given temperature and pressure--a design that is not suitable for a pilot plant. Moreover, it is a low efficiency device. The compressor has several unique features. It has a variable speed drive that operates from 10,000 to 18,000 RPM, allowing the gas flow to be varied from 94,000 to 174,000 SCFH. The compressor is contained in a hermetic pressure case and is lubricated by gas film bearings. MTI is presently conducting acceptance testing of this machine and will deliver the equipment in August of this year.

Lock hopper valve testing is an additional aspect of the experimental program. Two new valve designs will be evaluated. One type will be used in the coal feeding system--these valves were designed and manufactured by Rockwell International. They have the advantage of being able to operate for many thousands of cycles with little or no wear because there is no relative motion of metal to metal surface. In addition, we are expecting delivery from Consolidated Controls Corporation in August or September of this year of two refractory lined, high-pressure, high temperature valves for handling hot char. Both sets of valves will be installed in the pilot plant and tested under actual operating conditions.

CONCLUSION

Shakedown and orientation tests have been completed on the gasifier section of the 72 ton per day SYNTHANE pilot plant. Initial gasification tests with Rosebud subbituminous coal have been started. Thus far, no major process problems have been encountered. Mechanical malfunctions, however, involving equipment and instrumentation, have resulted in operating periods of short duration. Improved equipment and instrumentation more suitable for coal gasification systems should be developed.

During the initial operating periods, the gasifier has been responsive and easy to control. To demonstrate the potential of the SYNTHANE gasifier to achieve a high on-stream factor because of its simplicity, the development test program calls for relatively long operating campaigns of up to 4 weeks duration.

Significant support projects are in progress and include the fields of water treatment, valve development, and hot char discharge. The results from these projects should be of great value to developers of other coal conversion processes.

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KOPPERS-TOTZEK

ECONOMICS AND INFLATION

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SUMMARY

During the last several years, the economics of various synthetic fuel processes have been greatly affected by complex market forces as well as by inflation. This makes the economic analysis of coal gasification difficult to realistically present and interpret. In evaluating the economics of coal gasification, it is important to consider th influence of rising prices, which will almost certainly continue to characterize world economies in the years to come. This paper presents cost data for a large "grass-roots" Koppers-Totzek (K-T) plant which is designed to gasify Eastern bituminous coal to produce 140 billion Btu per day of 300 Btu per cubic foot fuel gas at sufficient pressure (170 psig) to supply industrial customers within an approximate 100 mile radius. The effects of coal price, return on equity, and capital structure on the cost of K-T gas are examined. In addition, however, the effects of general price inflation (4-10% per year) on the future costs of K-T gas and on an alternate fuel such as oil are presented. Although the initial cost of K-T gas may be higher than the cost of an alternate available fuel, K-T gas exhibits a lower cost increase over the life of the project and as a result a point in time is reached where the cost of the alternate fuel is considerably higher. This point is directly related to inflation and occurs sooner as the inflation rate increases or when market forces escalate the price of alternate fuels more rapidly than the price of coal and/or more rapidly than other gasification plant operating costs.

The large "grass-roots" K-T plant considered in this paper requires a mid 1976 plant investment of 367 million dollars, with total capital requirements estimated to be 454 million dollars. The calculated selling price of K-T gas ranges from $3.00 to $3.70 per million Btu for coal costing $15 to $25 per ton whenever cash flows are discounted at a 12% return on equity, with 9% annual rate on debt, and using a 60/40 debt to equity capital structure. In contrast the current selling price of No. 2 fuel oil averages about $2.40 per million Btu. However, as coal and oil prices continue to climb, a K-T plant built today is expected to produce a gas which is ultimately lower in cost than an alternate fuel somewhere during the economic life of the plant. For a modest inflation rate of 4% per year with coal initially available at $20 per ton, this point is reached in the fifteenth year; however at an inflation rate of 10% per year this point is reached in only five years. The analysis assumes for simplicity of presentation that the cited inflation rates are constant and equally applicable to the alternate fuel as well as to those portions of the K-T gas cost which are subject to escalation.

It is emphasized of course, that the rate of inflation or escalation, as it is considered for purposes of this paper, applies to fuel prices and gasification plant operating costs and may or may not be the same as the inflation rate of the economy as a whole. Conceivably, fuel prices may escalate more than the general inflation rate. We do not wish to attempt to preduct future costs and prices particularly since our most noted economists and business and government organizations cannot agree on these matters. We have instead presented a range of what we feel are reasonable rates of inflation and escalation for the future to indicate their differing effects on the price of K-T gas vs. the price of alternate fuel. However, the

hard fact remains that oil and gas are becoming increasingly scarce relative to demand, and the long discussed "Energy Crisis" could be upon us in 3-7 years despite the fact that it presently appears to many people to have disappeared. The United States must continue to sponsor research and development of coal gasification, and must also begin to employ it commercially as well. Coal gasification plants do require a great deal of capital investment, but so does the exploration and development of new sources of oil, natural gas, and other energy forms. Furthermore, location of coal resources are well established and the development of a coal gasification industry can be done completely with American resources and labor.

INTRODUCTION

The Koppers-Totzek gasification process has been commercially proven since 1952, at which time three 2-headed gasifiers were installed at Oulu, Finland. Since then, an additional 13 plants (39 gasifiers) have been installed in the Eastern Hemisphere by Krupp-Koppers, GmbH, of Essen, Germany. The latest plant was placed on stream in 1974 for African Explosives and Chemical Industries (AE&CI) in Modderfontein, South Africa. At this plant, 6 two-headed gasifiers are used to make a synthesis gas for the production of 1000 metric tons per day of anhydrous ammonia. Recently, a portion of the synthesis gas at this plant has been used for the successful production of 65 tons per day of methanol, using the Imperial Chemical Industries (ICI) high-pressure methanol synthesis process. A portion of this methanol is used by the gasification plant as make-up to the Rectisol acid gas removal system. This marks the world's first use of coal gasification for the production of methanol.

Three additional K-T coal gasification plants are under construction in India and each of these employs three 4-headed gasifiers to produce 900 metric tons of ammonia per year. Figures 1, 2, and 3 are recent photographs of the plant under construction in Talcher, India.

In 1975, a 2-headed gasifier at an existing plant in Puentes de Garcia Rodriquez, Spain, was instrumented to permit the test gasification of fluid petroleum coke, delayed petroleum coke, and two varieties of char from the FMC, OCR sponsored, COED process. All materials were successfully gasified. Data on the petroleum coke tests are proprietary since the tests were privately sponsored by eight major oil companies. Data on char gasification is currently being submitted to the Electric Power Research Institute (EPRI).

Features of the K-T process include:

1) The ability to handle a variety of feedstocks, including all ranks of coal, char, petroleum coke, tars, heavy residuals, and light to heavy oils. This advantage is tremendously important when it comes to securing and contracting for an economical fuel supply or in switching from one fuel to another during the life of the plant.

2) No tars, phenols, or other condensible hydrocarbons are produced. Aside from the obvious environmental advantages, the complete gasification of coal results in less maintenance and in the production of a synthesis gas which is readily suitable for chemical processing. Only slag and elemental sulfur are by-products of the K-T process.

3) Simplicity of construction and lack of mechanical complexity lead to a high degree of reliability.

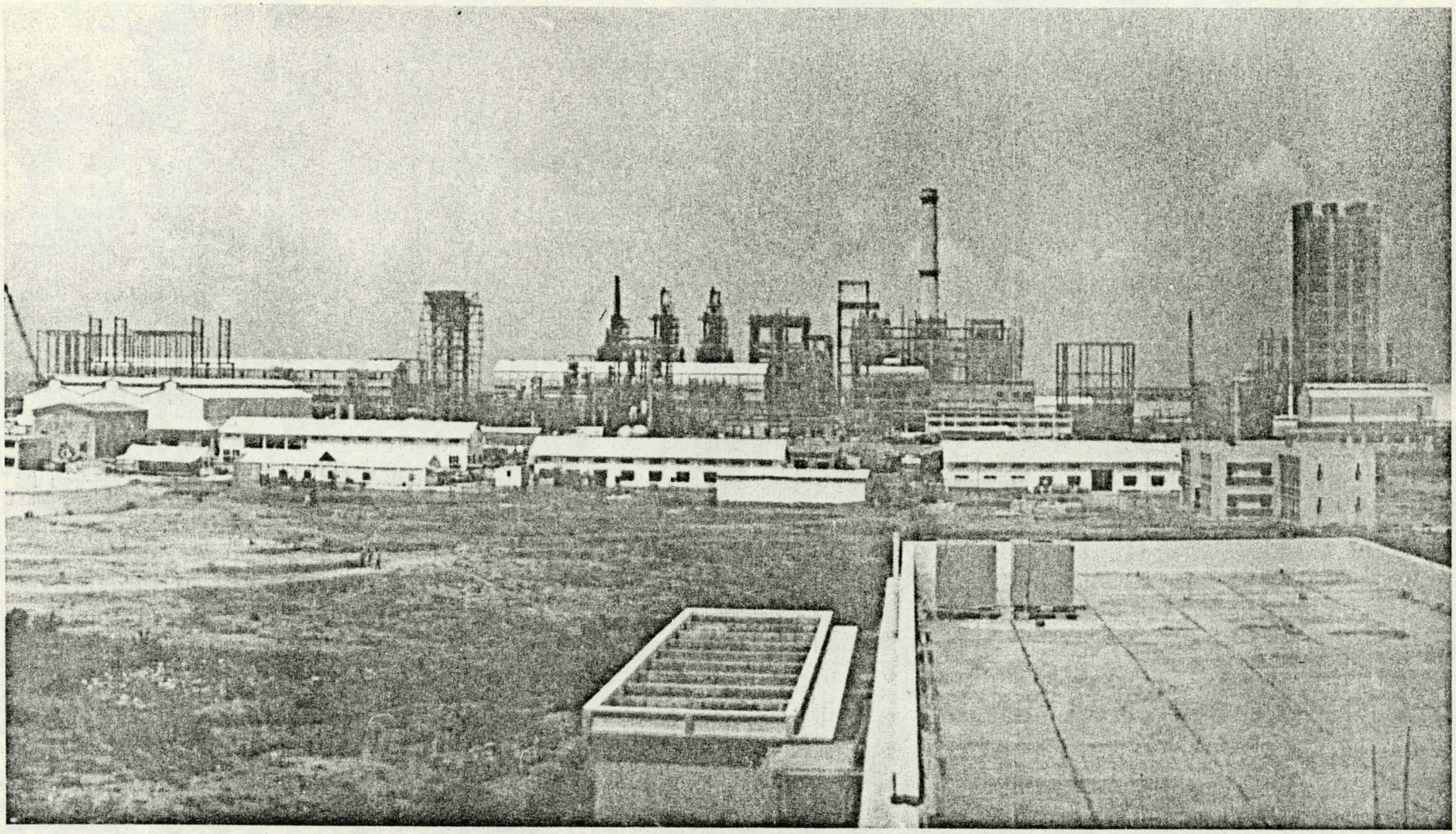


Figure 1. Koppers-Totzek Based Ammonia Plant - Talcher, India

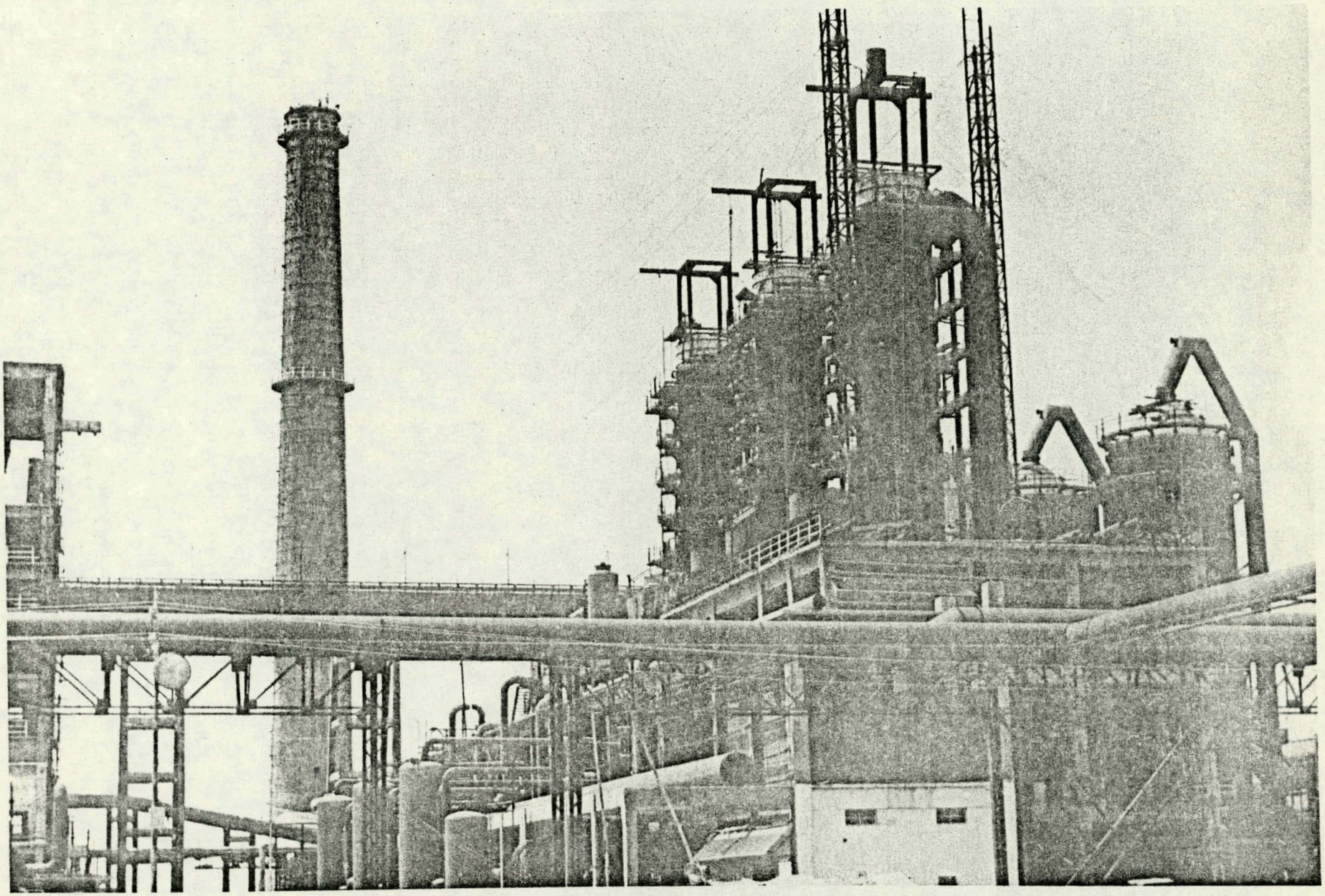


Figure 2. Koppers-Totzek Based Ammonia Plant - Talcher, India



Figure 3. Koppers-Totzek Based Ammonia Plant - Talcher, India

We in Koppers believe, as discussed at last year's Pitt Symposium, that concentrating on the industrial and electric utility segment of the energy market is the most expeditious and economical approach to the implementation of coal gasification. This group currently consumes 12 trillion cubic feet, or 60 percent, of the natural gas used in the United States. Their needs, with few exceptions, do not require a more costly high Btu synthetic natural gas. Converting industry to intermediate Btu (300 Btu per cubic foot) gas in a logical manner conserves natural gas for residential and other uses where the conversion to alternate fuels is too costly or impractical. Thus, this paper concentrates on the economics of a relatively large K-T fuel gas palnt. Aside from fuel gas, however, the K-T process may be employed for other industrial applications such as ammonia, hydrogen and methanol production or for direct ore reduction. These applications have been discussed in other technical papers prepared by Koppers Company, Inc.

PROCESS DESCRIPTION

The Koppers-Totzek entrained slagging gasification process uses the rapid partial oxidation of pulverized coal in suspension with oxygen and steam. Reaction temperature at the burners is about 3500°F, and as a result, no tars, oils, or condensable hydrocarbons are produced. Due to endothermic reactions and radiation, the gas exits the gasifier at approximately 2700°F. Gasifier pressure is slightly above atmospheric.

The gasifier is a steam-jacketed, refractory lined carbon steel shell. A four-headed gasifier (fig. 4), which is capable of processing up to 850 tons per day of coal, employs burners spaced 90° apart. The particular four-headed gasifier shown in the figure is one of three currently under construction at a synthetic ammonia plant in Talcher, India. A two-headed gasifier (fig. 5), utilizes burners spaced 180° apart and has about one-half the capacity of the four-headed version. Figure 6 is a photograph of a 2-headed gasifier at the ammonia plant commissioned late in 1974 at Modderfontein, South Africa.

A wide variety of feedstocks including tars, heavy oils, petroleum coke, char, and all ranks of coal can be gasified. The gas, rich in carbon monoxide and hydrogen, has a gross heating value of 300 Btu per cubic foot, dry basis. Characteristically, coal yields a gas of 50-55% carbon monoxide and 30-35% hydrogen, both on a dry basis. Carbon dioxide, nitrogen and sulfur compounds make up the balance. With liquid feedstocks, the carbon monoxide and hydrogen are produced in an approximately equal volume of 45% each. Table 1 illustrates the performance of a four-headed gasifier using as fuel the bituminous coal considered in the economic analysis presented in this paper. Table 2 is shown to illustrate the heat balance on the gasifier.

Figure 7 is a simplified flow diagram of the K-T gasification system. Cost data in the paper assumes that coal is delivered by rail to the plant site on a 5 day per week basis. Included in the cost of the plant are the necessary spur lines, marshalling yard, switching locomotive, car positioner and dumper, thaw shed, reclaiming system, and all necessary conveyors and mobile equipment. Reinforced concrete bins are included to provide sufficient coal storage for daily plant operation. In addition, a coal yard storage pile of sound environmental design is included to provide at least nine days of coal inventory for emergency use.

Coal is fed to the pulverization system (fig. 8), where the coal is reduced to a size consistency of 70-90 percent minus 200 mesh and is simultaneously dried to a moisture level of approximately 2 percent. In general, however, the degree of drying depends on coal rank. Lignites, for instance, are dried to 8 percent. Wind swept roller mills are used with heat demands for drying met by burning a portion of the desulfurized fuel gas. Combustion gas used for drying is tempered to 800-900°F to keep coal particulate temperature at 180°F to prevent devolatilization. Pulverized coal is transported

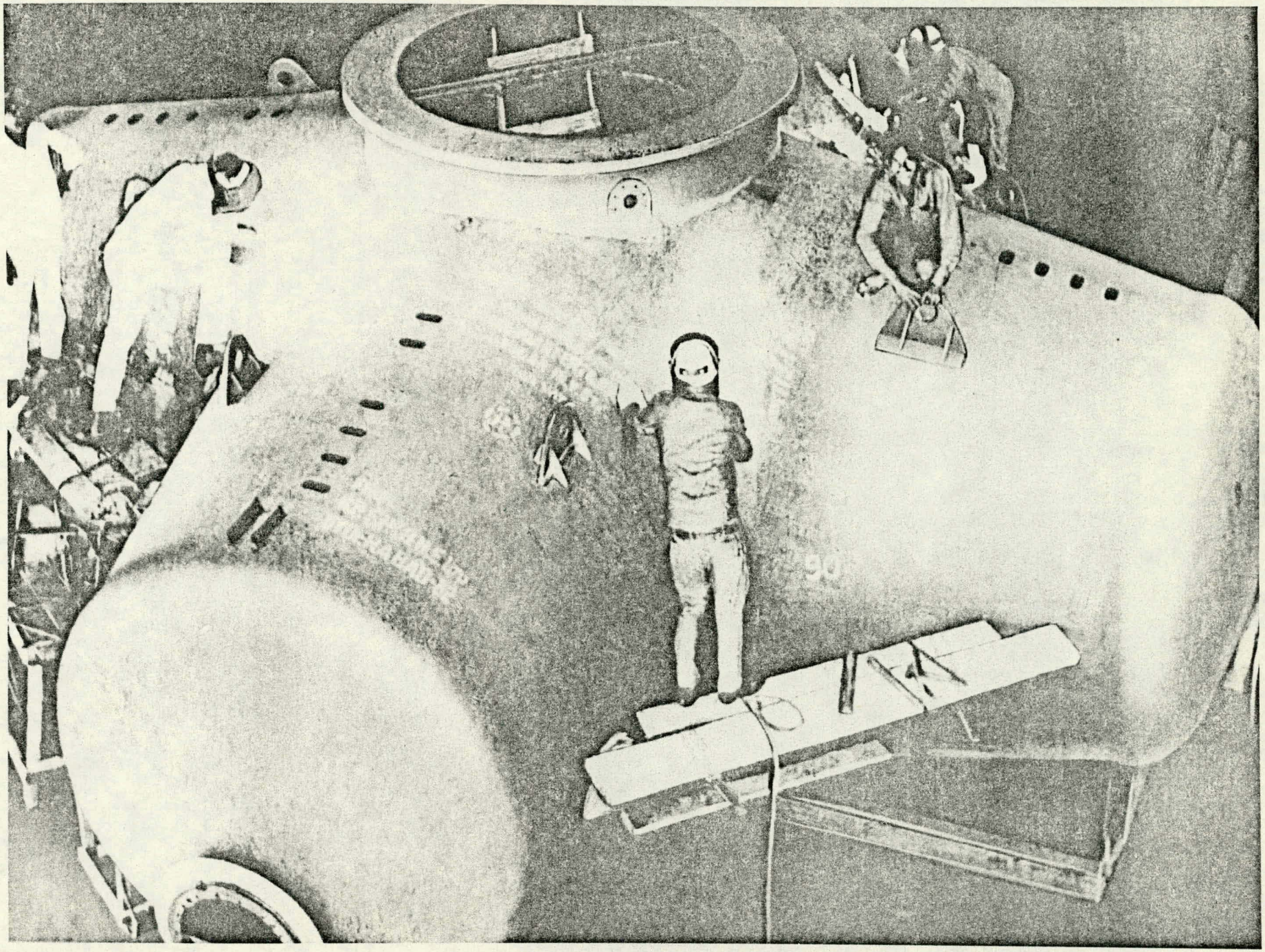


Figure 4. Four Headed K-T Gasifier Under Construction - Talcher, India

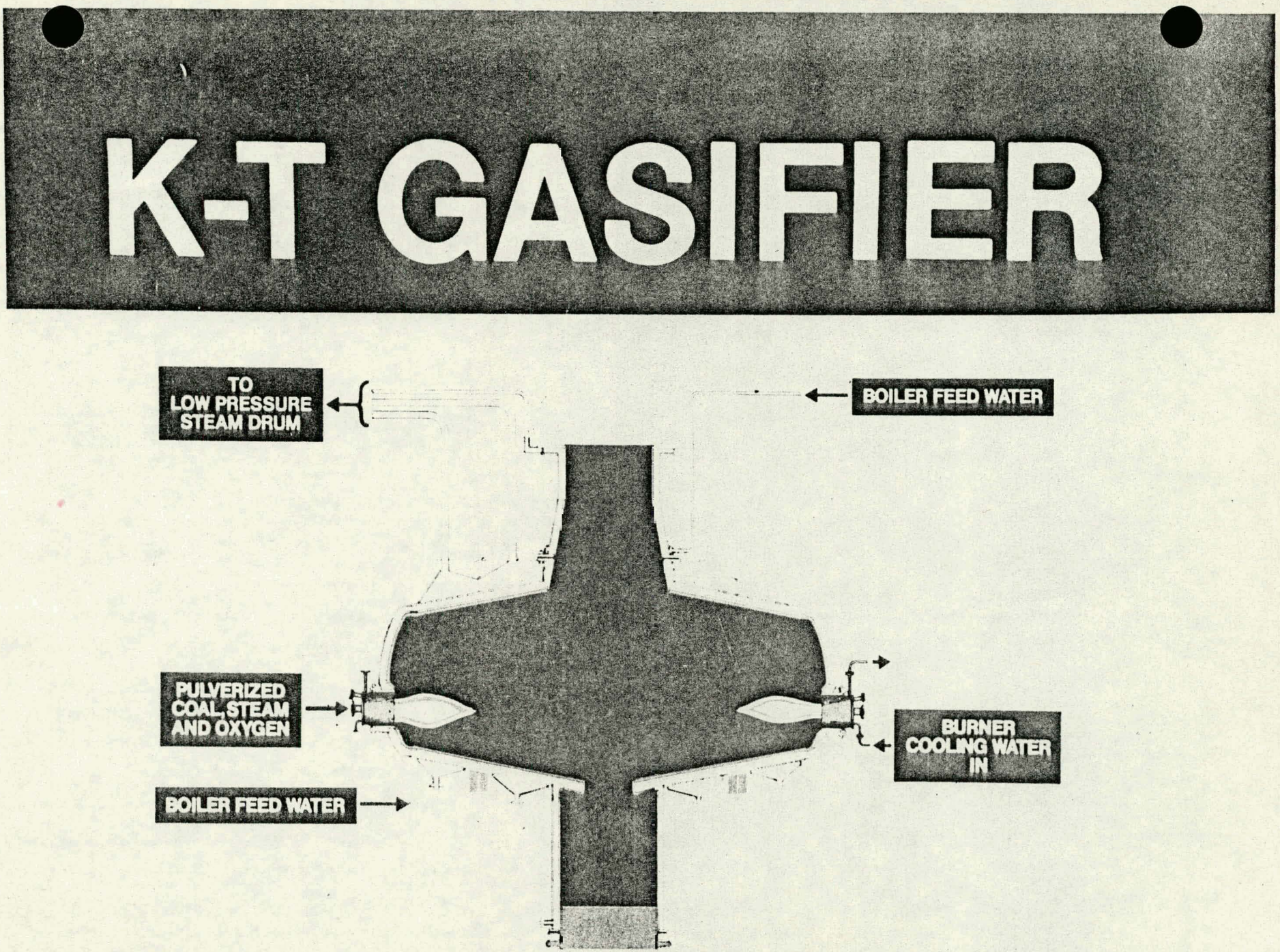


Figure 5. Two Headed K-T Gasifier - Artist's Conception

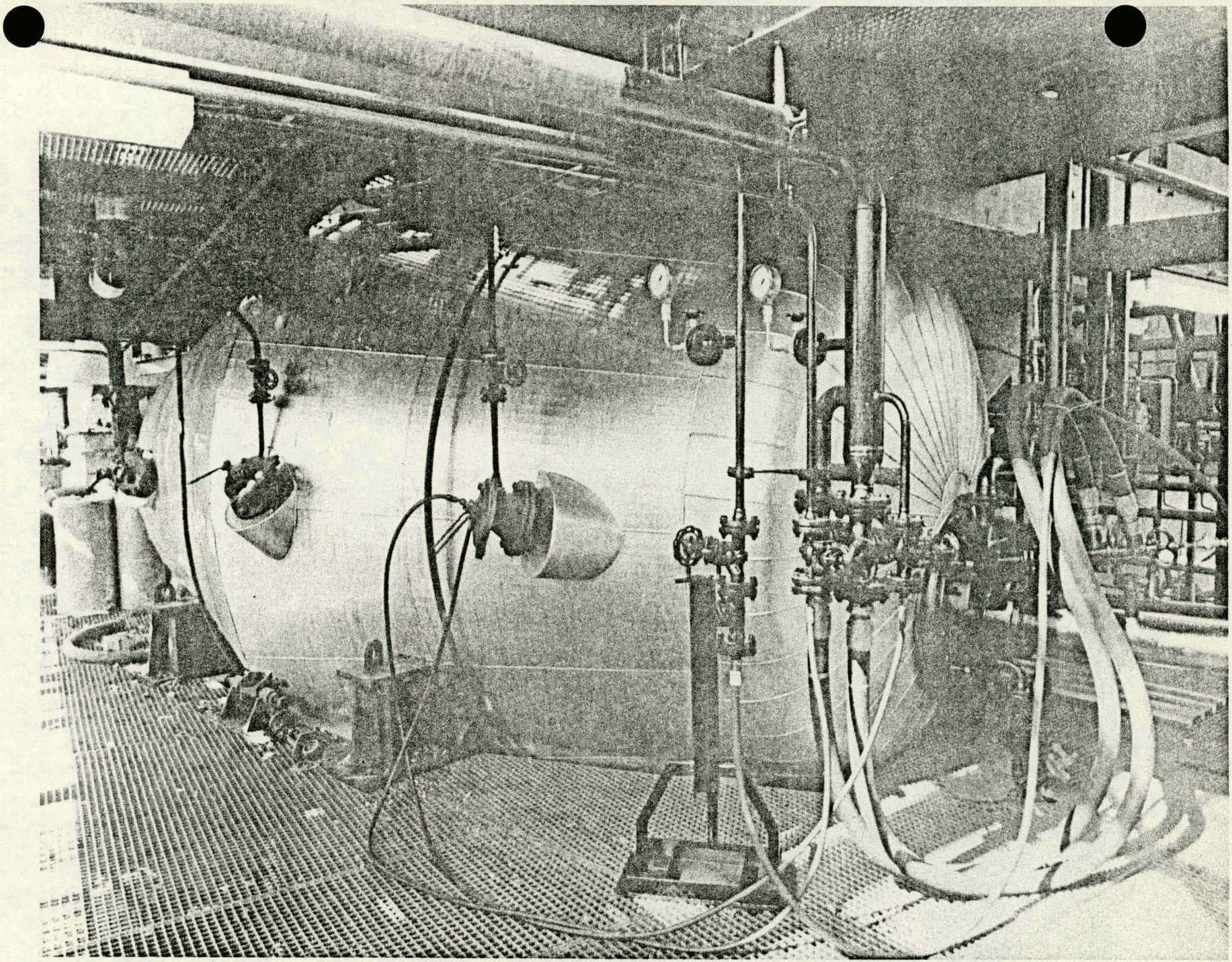


Figure 6. Two Headed K-T Gasifier - Modderfontein, South Africa

TABLE 1

Performance of K-T 4 Headed Gasifier with Bituminous Coal

		Single 4 headed	Plant Basis: 14-4 headed (Operating)
As received coal, tons/day @ 6% moisture		693	9,702
Gasifier Feeds			
Pulverized coal feed tons/day @ 2% moisture		665	9,310
Coal feed analysis, wt. %:			
	C	66.88	
	H	4.92	
	N	1.16	
	S	4.08	
	O	7.00	
	Ash	13.87	
	H_2O	2.00	
		100.00	
Gross heating value of coal feed, btu/lb		12,286	
Oxygen @ 98 vol. % purity	tons/days	561	7,854
	tons/ton coal feed	0.843	
Process Steam @ 250°F	lb/hr	18,200	254,800
	lb/ton coal feed	657	
Gasifier Products			
Jacket steam @ 30 psig			
	lb/hr	25,570	357,980
	lb/ton coal feed	922	
High Pressure steam from waste heat boiler @ 1700 psig, saturated	lb/hr	89,860	1,258,040
	lb/ton coal feed	3,243	
Raw Gas Analysis, vol. %, dry basis			
	CO	53.18	
	CO_2	10.02	
	H_2	34.47	
	H_2S	1.07	
	COS	0.07	
	N_2, Argon, HCN, NH_3	1.19	
		100.00	
Dry gas gross heating value, btu/scf, dry		290	
Dry gas make			
	scfm	29,270	409,780
	scf/ton coal feed	63,380	
% Heating value of gas/heating value coal feed (gross basis)		74.8	

TABLE 2

K-T GASIFIER HEAT BALANCE

BITUMINOUS COAL (2% MOISTURE)

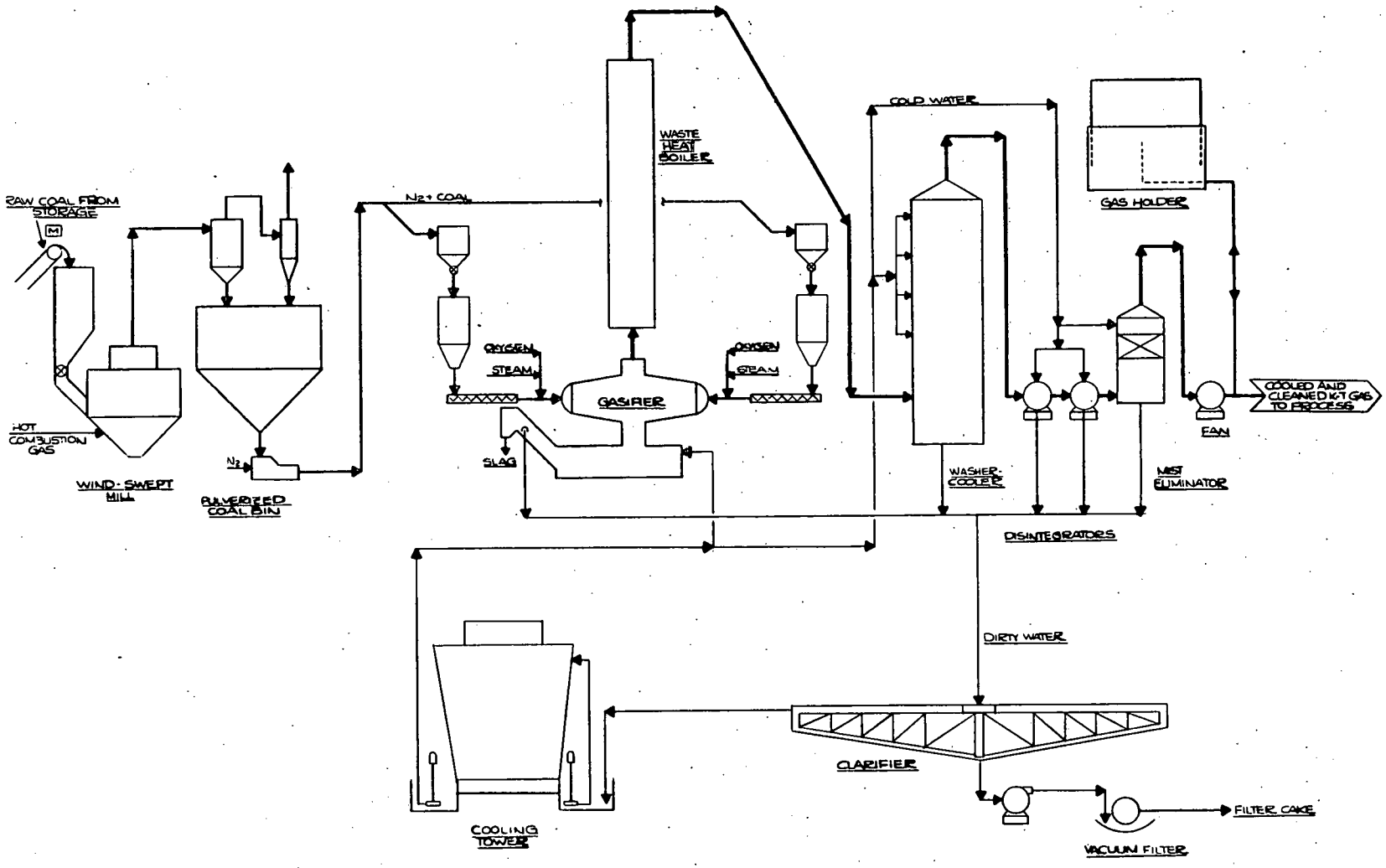
Gas Exit Temperature=2730°F

Reference State = Liquid Water

HEAT INPUT - ABOVE 60°F	Btu/Lb of Coal Feed	%
Calorific Value of Coal	12,286	96.7
Sensible Heat in Coal @ 160°F	30	0.2
Sensible Heat in Oxygen @ 220°F	27	0.2
Total Heat in Steam @ 250°F	367	2.9
Total	12,710	100.0

HEAT OUTPUT - ABOUT 60°F		
Heat in Molten Slag and Ash	104	0.8
Sensible Heat in Unburned Carbon	34	0.3
Calorific Value of Unburned Carbon	470	3.7
Calorific Value of Gas	9,185	72.3*
Sensible Heat Dry Gas	1,784	14.0
Total Heat - Water Vapor in Gas	629	5.0
Total Heat - Gasifier Jacket Steam	448	3.5
Gasifier Heat Loss	56	0.4
Total	12,710	100.0

* Constitutes 74.8% of Gross Combustion Value of Coal



K-T GASIFICATION SYSTEM - SIMPLIFIED FLOW DIAGRAM

Figure 7

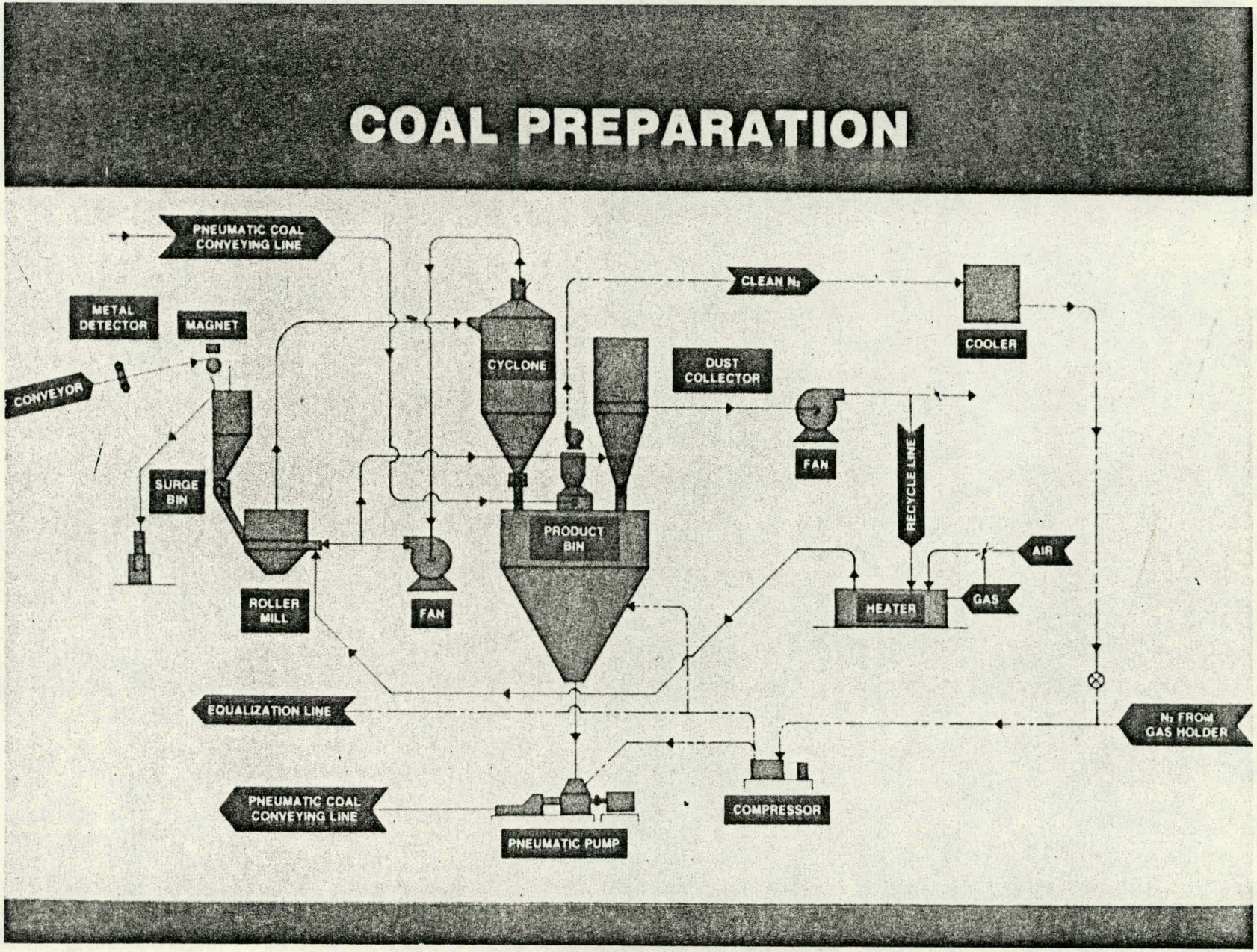

COAL PREPARATION
PNEUMATIC COAL CONVEYING LINE
METAL DETECTOR
MAGNET
CONVEYOR
SURGE BIN
ROLLER MILL
FAN
CYCLONE
PRODUCT BIN
CLEAN N₂
COOLER
DUST COLLECTOR
FAN
RECYCLE LINE
AIR
GAS
HEATER
EQUALIZATION LINE
N₂ FROM GAS HOLDER
PNEUMATIC COAL CONVEYING LINE
PNEUMATIC PUMP
COMPRESSOR

Figure 8.

pneumatically with nitrogen and stored under an inert atmosphere. All vent lines lead to bag filters to prevent dust emissions. The pulverized coal is conveyed from storage to the service bins (two at each 4 headed gasifier), each of which in turn feeds a pair of feed bins. From each feed bin the coal is metered by two feed screws into a mixing zone where it is entrained in a pre-mixed stream of steam/oxygen and fed into the gasifier.

About 50% of the coal ash leaves the gasifier through the bottom in molten form and then is solidified by contact with water in a quench tank situated beneath the gasifier. The ash is removed from the quench tank by means of a scraper-conveyor assembly and is then conveyed to a slag storage bin which permits loading of the slag into railroad cars for disposal at the mine site.

The remaining ash along with any ungasified carbon is entrained with the gas leaving the top of the gasifier, and is sprayed with water, if necessary, to freeze any slag droplets prior to entry into the waste heat boiler. High pressure saturated steam is generated in the waste heat boiler, and gas leaving the economizer section is directed to a washer-cooler wherein the gas is cooled by spray water, typically to about 100^{0} F , while about 90% of the entrained particulates are simultaneously removed. Gas then enters Theisen type irrigated disintegrators to reduce dust loading of the gas to about 0.002 grains per standard cubic foot. After passing through a mist eliminator, the gas is moderately boosted in pressure by a fan to overcome downstream pressure losses. Water used in the cleaning system is recirculated cooling tower water and the solids are recovered in a clarifier in combination with a filtration system. Filter cake is conveyed to a bin for loading into railroad cars.

Gas leaving the fan at each gasifier is manifolded into the suction of three gas compressors which deliver the gas at 180 psig to the acid gas removal system. A short residence time gas holder is provided at the compressors' suction header for control purposes.

The acid gas removal facilities are comprised of three parallel strings, each of which consists of a COS hydrolysis column followed by an absorber. Within the COS hydrolysis column the gas is contacted with hot circulating methyldiethanolamine (MDEA) solution to promote hydrolysis of COS to H_2S in order to facilitate a high degree of sulfur removal within the absorber. Acid gases are stripped from the MDEA absorbent and sent to two multi-stage Claus units, each complete with a tail gas unit, for recovery of elemental sulfur. Recovered sulfur is stored in molten form and loaded into railroad tank cars. Costs presented herein are based on an overall sulfur recovery designed to meet Federal standards as well as local and State regulations typical of the Eastern United States. Gas leaving each H_2S absorber contains approximately 115 ppm of H_2S plus COS, or 0.076 pounds of SO_2 equivalent per million btu of gross calorific heat content of the fuel gas.

Gas from each H_2S absorber is sent to one of three drying columns for removal

of water to meet a -18°F gas dew point so as to permit long distance transmission. Drying is effected using recirculating triethyleneglycol (TEG). Water is removed from the TEG by distillation using iso-octane as an azeotroping agent. Product gas, having a gross heating value 301 btu per standard cubic foot, leaves each drying column at approximately 170 psig and then enters the gas distribution system.

Prior to the gas drying columns, a portion of the desulfurized gas is piped through the gasification plant for use in coal drying, sulfur recovery, and for firing of two auxiliary boilers. The auxiliary boilers are provided to superheat steam from the gasifier waste heat boilers and to also evaporate and superheat additional feedwater to thereby provide sufficient steam to meet demands of the gasification plant. All major compressors, except for the three oxygen compressors and one of the four (4) air compressors at the four air separation plants, are equipped with steam turbine drives to thereby keep the amount of purchased electrical power to a minimum while still allowing a practical means of starting the plant. Overall raw material requirements, product yields, and ultility requirements are indicated in Appendix B.

Blowdown from all cooling towers along with blowdown from the steam distribution system are treated to meet acceptable stream standards. Principally, the treatment involves destruction of HCN and NH_3 which are present in small quantities in aqueous blowdown from the cleaning system by chlorination, dechlorination, and pH adjustment.

Costs presented in the report are based on a grass-roots plant, and hence, general plant facilities are included such as fire protection, raw water treatment, roadways, fences, buildings, storm sewers, sanitary waste treatment, communication systems, maintenance tools, and vehicles.

EFFECTS OF FINANCIAL VARIABLES ON GAS COST

The sensitivities of the calculated cost of K-T gas to changes in parameters such as the cost of coal, rate of return, capital structure, and plant investment are shown on figures 9-12. All costs were calculated for the "grass-roots" fifteen gasifier K-T plant* producing 140 billion Btus/day that was referred to earlier in the paper. The installed cost of the plant was estimated to be $367 MM in terms of mid-1976 dollars. The total capital requirement assuming the coal will be available at $20/ton was estimated to be about $454 MM (see Appendix A for a sample calculation of the total capital requirement). Annual operating costs with coal feedstock at $20/ton are approximately $94.7 MM and are itemized in a tabulation that appears in Appendix B.

The effect of changes in the price of coal feedstock on the cost of K-T gas is shown on figure 9 for a fifteen gasifier plant financed by 60% debt and 40% equity. Interest on debt was 9% per year and the return on equity capital was 12% as computed by a discounted cash flow rate of return method (DCFRR). (A sample cash flow statement that is indicative of the methods used to compute gas costs for the present paper is shown in Appendix C). Figure 9 indicates that a $1 per ton change in coal cost changes the cost of gas by only $0.07 per millions Btus.

Figure 10 indicates the effect of a varying gas cost on the discounted cash flow rate of return on equity capital for constant coal costs of $10, $15, and $20/ton. It should be noted that a rather small increase in the cost of gas from about $3.20 to $3.35 results in an after tax increase in the discounted cash flow rate of return from 8% to 12%. This is due primarily to the fact that project financing was done with only 40% equity. As the amount of equity in the project is reduced from this level, the increase in the rate of return is even more dramatic. Conversely, increases in the amount of equity used to finance the project above 40% yield increasingly smaller gains in the rate of return on equity capital for given upward movements in the cost of K-T gas.

An additional relationship between capital structure (i.e. the proportions of debt and equity capital used for project financing) and the cost of gas is shown on fig. 11. At a constant DCFRR of 12% and a coal cost of $20/ton, the cost of K-T gas decreases from $3.85/MM Btu to $3.15/MM Btu as the percentage debt increases from 20% to 80%. This can be attributed to the fact that the cost of debt capital is 9% compared to 12% for equity capital in this case and to the fact that the cost of debt capital (interest on debt) is tax deductible, whereas the return on equity capital is not.

The final figure in this section (fig. 12) indicates the impact of changes in the expected plant investment on the cost of gas. Horizontal axes show investment level in terms of dollars as well as in percentage changes from the expected plant investment of $367 MM. The sensitivity of gas cost to changes in items such as equipment costs or construction costs that impact plant investment is seen to be only about 25¢/MM Btu for a change of 20% in the plant investment.

* Fourteen operating gasifiers and one spare.

Figure 9

EFFECT OF COAL COST ON GAS COST

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
Debt/Equity: 60/40
DCFRR = 12%

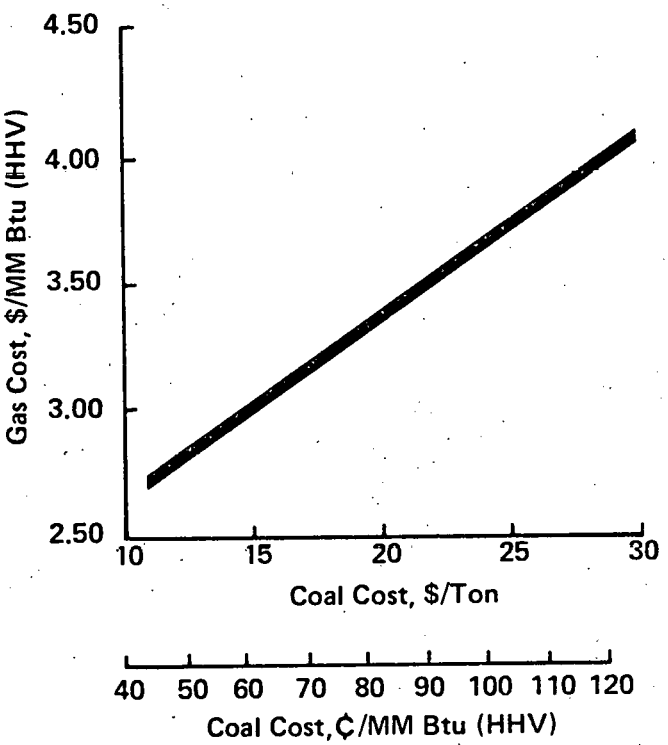


Figure 10

EFFECT OF RATE OF RETURN ON GAS COST

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
Debt/Equity: 60/40

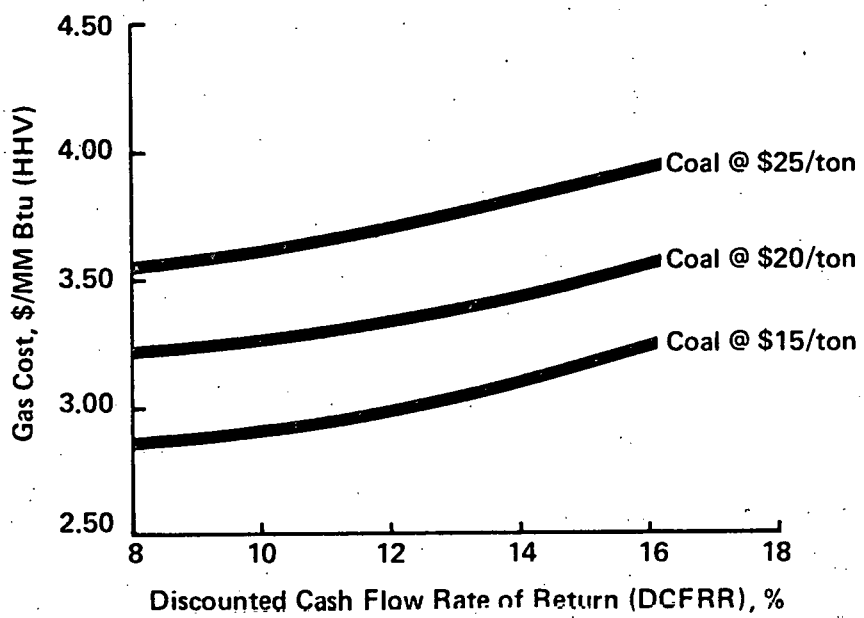


Figure 11

EFFECT OF CAPITAL STRUCTURE ON GAS COST

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
DCFRR = 12%
Coal Cost = $20/ton

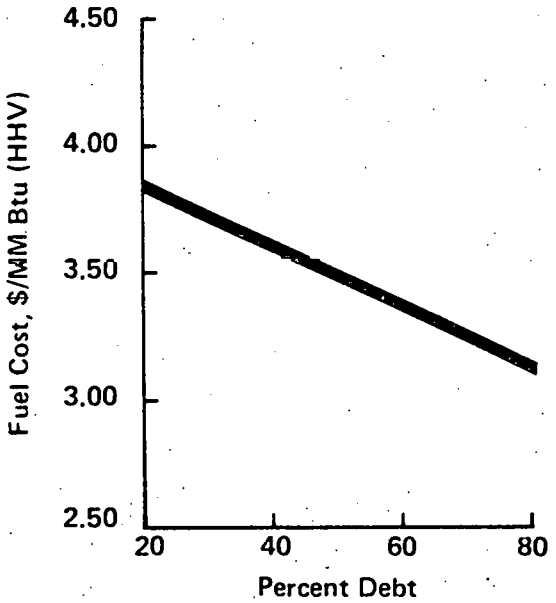
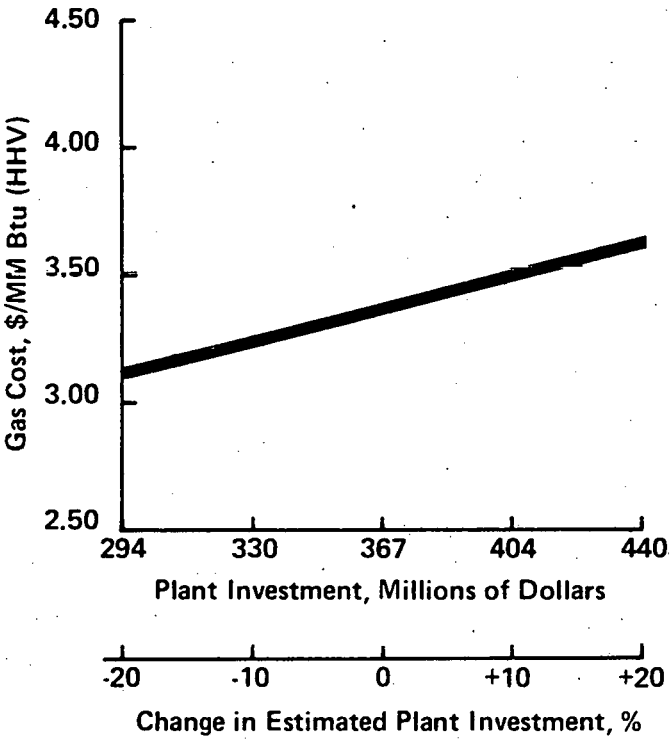


Figure 12

EFFECT OF PLANT INVESTMENT ON GAS COST

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
Debt/Equity: 60/40
DCFRR = 12%
Coal Cost = $20/ton



EFFECTS OF PRICE INFLATION AND ESCALATION ON FUEL COSTS

The gas costs that have been presented in Figures 9 to 12 were based on constant annual operating costs for the twenty year project period. The cost of K-T gas was calculated such that the revenues received from the sale of gas at a constant price over the period would be sufficient to yield a given discounted cash flow rate of return on equity capital (see Appendix C for a sample 20-year cash flow statement). It is, of course, unreasonable to make the assumption that the operating costs will remain constant over the twenty year period. However, it is extremely difficult and controversial to predict the behavior of costs for future periods, even as short as a few months. Predictions of the cost of coal or labor ten years or twenty years into the future can be little more than speculation.

It is, however, reasonable to suspect that inflation will continue to be a characteristic of U.S. and world economies and it is important to temper business decisions based on the possible impact that inflation will have on proposed projects. In this regard, it is helpful to examine typical "make or buy" decisions concerning future supplies of energy. The point of view in the following material was chosen as that of the large industrial consumer or group of consumers who are evaluating the costs of supplying energy to their plants in the form of either gaseous or liquid fuels for purposes such as heating (furnaces or reactors), steam production, or electrical energy production. Their decision (as postulated here) is basically whether to invest in K-T coal gasification facilities to supply future energy needs or to rely on alternate fuels that can be purchased in the market place.

It should be emphasized, before proceeding, that this kind of decision obviously must take into account factors that may be extraneous to those concerning economics. We do not wish to convey the impression that the proper choice hinges on the results of an approach to these problems that would only consider the economics of energy costs. However, many arguments that have been presented in the past and are still being presented dismiss coal gasification as a means of supplying energy because the cost of gas energy from coal is currently more expensive than alternate methods of supplying energy. It is argued below that in many cases this conclusion is not a valid one when the effects of time and price escalation are taken into consideration. The analysis that follows is not intended to be complete, in that the other factors cited above are not fully discussed. These factors vary from case to case and cannot be comprehensively considered in a paper of this type. One of the more obvious of these factors is the availability of capital. If a group of industrial consumers feels that K-T gasification represents a least cost solution to their energy supply problems, but they do not have the capital or access to the capital that is required to build a plant, the purchase of alternate fuels in the market may be the only option available to them. The material below is intended to show that an economic analysis comparing various alternatives of supplying energy often points to the K-T Process as the economic choice when the costs of future energy supplies are investigated.

The effects of inflation on the cost of Koppers-Totzek gas and on the cost of an available alternate fuel are shown in figures 13 to 16. It was assumed that the first year cost for the alternate fuel was $2.40 per MM btu. The alternate fuel was assumed to be No. 2 fuel oil, with an approximate cost today of around $2.40 per MM btus (Ref.1). The future costs of the alternate energy source were determined by annually escalating the initial cost of $2.40 per MM btu at the particular rate of inflation shown on each of the figures (13 through 16). The future costs of Koppers-Totzek gas were determined by escalating the gross operating costs (the portion of the K-T total cost that is subject to inflation) at the appropriate rate of inflation shown on figures 13 to 16*. The costs shown for K-T gas will, of course, still yield a 12% DCFRR to the holders of equity capital over the life of the project. The costs of K-T gas are shown as starting at approximately $3.36 per MM btu and then increasing throughout the life of the project. The entire cost of coal at the beginning of the project ($20/ton) was not fully escalated at the various rates of inflation shown on figures 13 to 16. It was felt that a long term coal contract could be negotiated that limited the price increases of coal such that they would be just sufficient to cover increases in coal mine operating costs such as labor, material, power, and transportation. The percentage of the initial coal cost of $20/ton that is associated with these operating costs was estimated to be 50%.

The above projections indicate that for rates of inflation between 4% and 10% per annum (inclusive) the cost of the alternate fuel at some point in the life of the project becomes equal to and then surpasses the cost of K-T gas at various times according to the rate of inflation. This occurs even though the initial cost of K-T gas is greater than that of the alternate.

It was also determined that over the twenty year period the average cost of K-T gas was less than the average cost of the alternate for rates of inflation of approximately 5 1/4% per year and greater. The relationship between the average cost of K-T gas and the average cost of alternate fuel as a function of the rate of inflation is shown on figure 17.

A key question facing the decision maker (s) then is what the average rate of inflation or price escalation will be for future time periods. If he expects inflation to be approximately 5% or less per year and if the decision maker bases his decision on average prices, or in other words, minimizing his total expenditures for energy for the next twenty years, he will choose to purchase the alternate. If, however, he expects the rate of inflation to be greater than 5%, K-T gasification will be his choice.

An alternative approach to the above would be to discount future energy costs at a rate that is applicable to the decision maker and to compare the costs of K-T gasification and the purchase of alternate fuels on the basis of present dollars. This type of technique will take the time value of money into account and will in effect place greater value on early cost savings and less value on savings that occur later in the project. Each rate of inflation and each particular discount rate that is applied will yield a unique value for the present value of the costs of energy. Figure 18 indicates the relationship between the rate of inflation and the discount rate that is used to determine the present values of both the K-T gas costs and the alternate fuel costs.

*The costs curves are shown as being smooth and continuous to simulate the effects of inflation. Costs were calculated, however, yearly for the twenty year period.

Figure 13

EFFECT OF PRICE INFLATION ON FUEL COSTS: 4% Inflation

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
DCFRR = 12%
Debt/Equity: 60/40
First Year Coal Cost: $20/ton
First Year Alternate Fuel Cost: $2.40/MM Btu

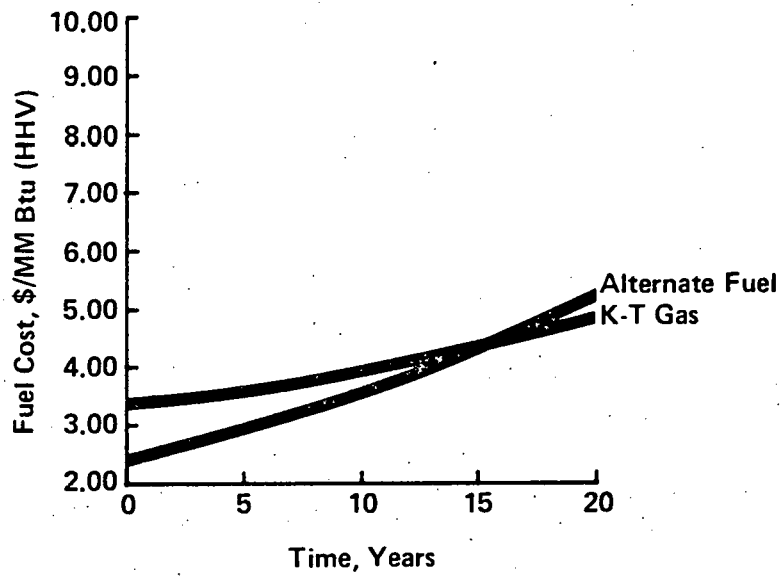


Figure 14

EFFECT OF PRICE INFLATION ON FUEL COSTS: 6% Inflation

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
DCFRR = 12%
Debt/Equity: 60/40
First Year Coal Cost = $20/ton
First Year Alternate Fuel Cost = $2.40/MM Btu

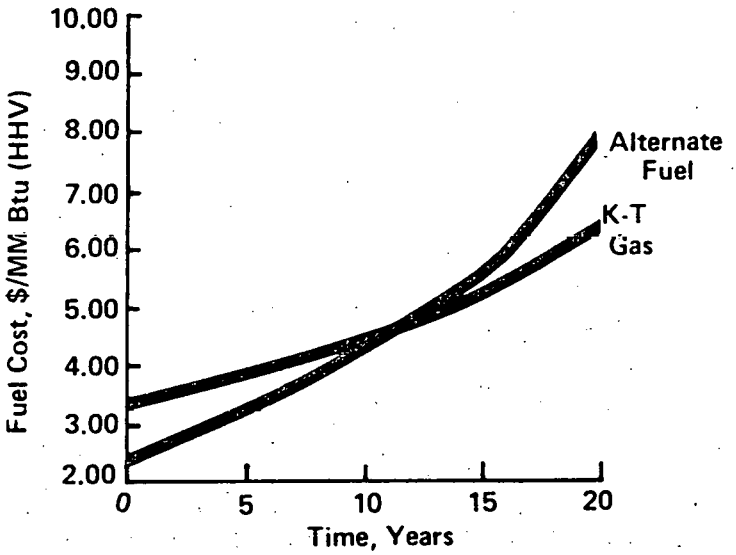


Figure 15

EFFECT OF PRICE INFLATION ON FUEL COSTS: 8% Inflation

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
DCFRR = 12%
Debt/Equity: 60/40
First Year Coal Cost = $20/ton
First Year Alternate Fuel Cost = $2.40/MM Btu

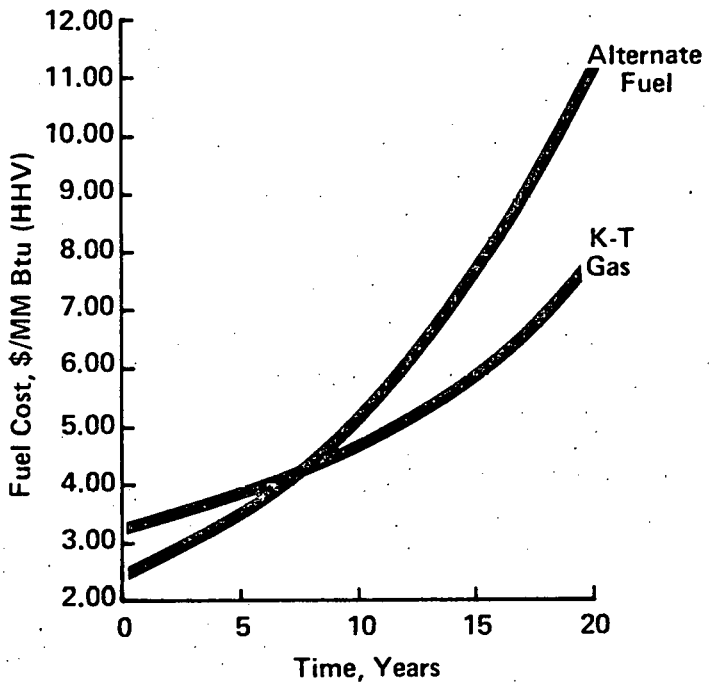


Figure 16

EFFECT OF PRICE INFLATION ON FUEL COSTS: 10% Inflation

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
DCFRR = 12%
Debt/Equity: 60/40
First Year Coal Cost = $20/ton
First Year Alternate Fuel Cost = $2.40/MM Btu

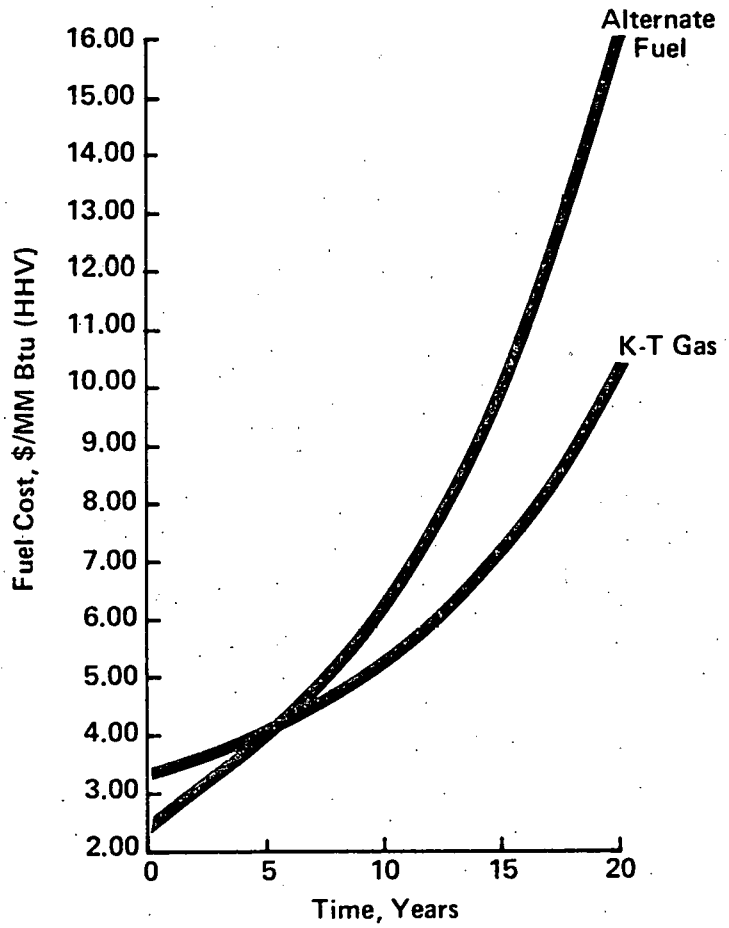


Figure 17

EFFECT OF INFLATION ON 20 YEAR AVERAGE FUEL COSTS

Bases:
15 4-Headed Gasifier Plant
140 Billion Btu's/day @ 170 psig
DCFRR = 12%
Debt/Equity: 60/40
First Year Coal Cost = $20/ton
First Year Alternate Fuel Cost = $2.40/MM Btu

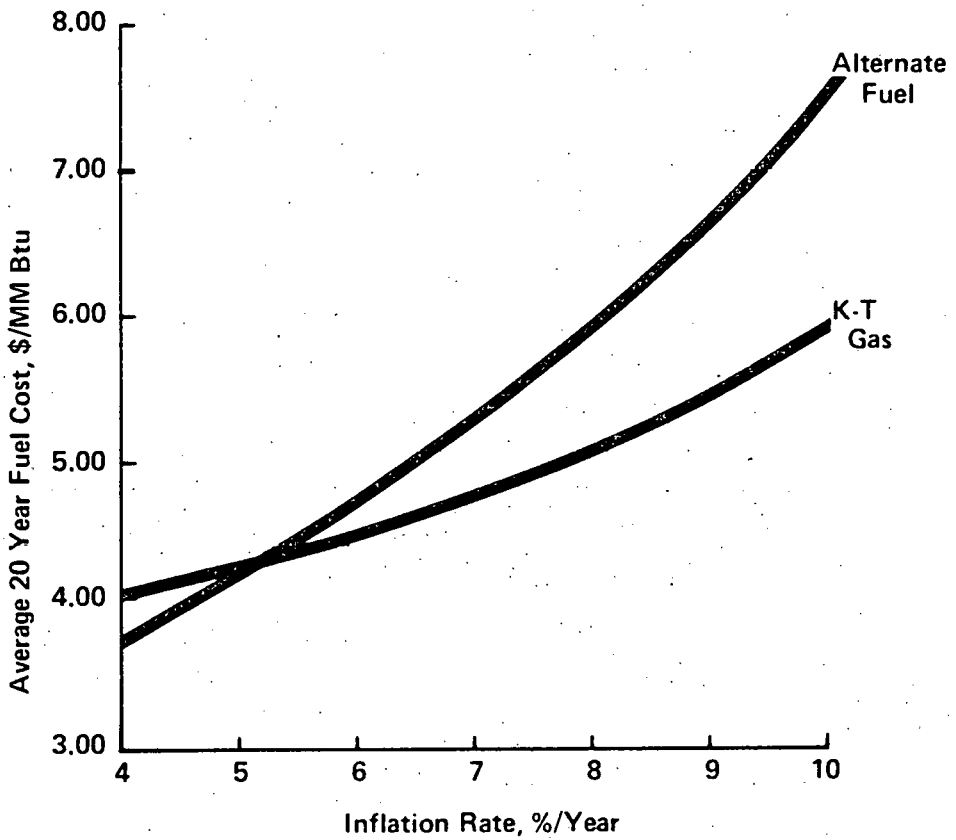
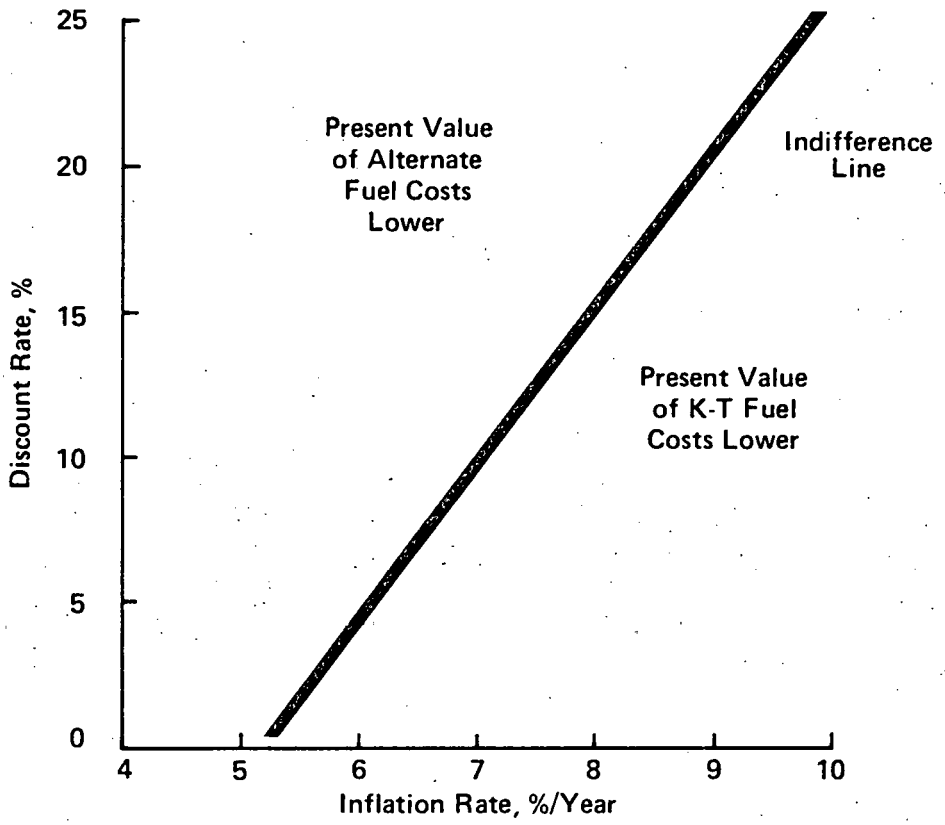


Figure 18

RESULTS OF THE DISCOUNTING APPROACH TO FUEL COSTS



The selection of a particular rate of inflation and a particular rate of return on money (discount rate) results in one of the following sets of circumstances:

1) The present value of the K-T gas costs is lower than that of the alternate and is therefore the economic choice.

2) The present value of the alternate fuel costs is lower than that of the K-T costs and is therefore the economic choice.

3) The present value of each set of costs is the same and the decision maker (s) would be indifferent between the choice of one or the other in an economic sense.

   (These points lie along the line marked "Indifference Line")

For example, if the decision maker (s) minimum acceptable rate of return is 10% and he expects inflation to be about 7% or greater, his economic choice would be for the costs associated with K-T.

The implication of the above is that although the present cost of gas produced via K-T gasification may be more expensive than other fuels that are available today, the effects of inflation and increased prices of natural gas and oil due to diminishing supplies and higher exploratory and development costs, can be such that over a relatively short period of time, the cost of energy from existing K-T gasification facilities will be less than that of alternate sources. It is important to realize that the portion of the cost of synthetic fuel that is associated with the return of interest and principal to the creditors of the project is not generally subject to escalation. Operating costs are subject to inflation but will not increase the cost of gas by nearly the amount of their increase. Therefore, a prudent manager and/or decision maker will apply the above types of analyses using his particular values for parameters such as the cost of capital to arrive at a decision whether K-T gasification or the purchase of an alternate fuel such as No. 2 fuel oil is the economic choice for the future.

REFERENCE

1.) Wholesale Prices and Price Indexes; Data for March 1976; U. S. Department of Labor, Bureau of Labor Statistics p. 79 (Middle Atlantic Region, No. 2 Fuel Oil)
(Average price was 31.3¢/gallon or $2.40/MM Btu)

APPENDIX A

Total Capital Requirement

Bases: 15 gasifier plant (14 operating)
Bituminous Coal @ $20/Ton
"Grass-Roots" Plant

	Million $
Plant Investment	$ 367.00
Interest During Construction (9%/year x $367 MM x 1.875 yrs.)	61.93
Startup Costs (2% of Plant Investment)	7.34
Working Capital (60 day cash supply = $\frac{60}{330}$ x Gross Operating Cost)	17.80
Total Capital Requirement	$ 454.07

APPENDIX B

Gross and Net Operating Costs

Bituminous Coal

Fourteen Operating Gasifiers plus One Spare

330 Days per Year

ITEM	USAGE	UNIT COST	$M/Year	$/MM Btu
Coal	9702 tons/day	$20/ton	$64,033	$ 1.38
Catalysts and Chemicals	-	-	1,884	0.04
Raw Water (from river; treatment included in other costs)	10,500 gpm	no charge	0	0
Electricity, Incoming @ 138 kv	2.26MM kw-hr/day	15 mil/kw-hr	11,190	0.24
Process Operating Labor	96 people	$7.36/hr* avg.	1,470	0.03
Supervision and Management	40 people	$9.50/hr* avg.	790	0.02
Maintenance Labor	74 people	$7.80/hr* avg.	1,200	0.03
Administrative Labor	30 people	$6.33/hr* avg.	395	0.01
Contracted Maintenance Labor	1% of investment		3,670	0.08
Maintenance Materials	1.5% of investment		5,505	0.12
Operating Supplies	30% of process operating labor		440	0.01
Insurance and Local Taxes	2% of investment		7,340	0.16
Slag Disposal at Mine Site	2360 tons/day	$0.70/ton	545	0.01
Gross Operating Costs	-	-	$98,462	$ 2.13
Sulfur Credit	282 LT/day	$40.60/LT	(3,779)	(0.08)
Net Operating Cost			$94,683	$ 2.05

*Includes fringe benefits plus payroll extras

APPENDIX C

Sample Cash Flow Statement

A sample year by year listing of items such as gas revenue, depreciation, cash flow, and discounted cash flow is given on the following page to illustrate accounting methods that were used to arrive at the gas costs presented in the paper. The sample is for a "grass-roots" fifteen gasifier plant using bituminous coal and financed by 60% debt and 40% equity capital.

It was decided for all sections of the paper (except for the section on inflation) to calculate a constant twenty year gas cost such that if this cost were the sales price, sufficient revenues would be generated to satisfy all demands for cash and to provide a given discounted cash flow rate of return on equity capital. It was felt for these cases that a constant cost would be easier to interpret than an average cost calculated from either declining or increasing costs over the twenty year project life. However, when the K-T operating costs were escalated by various rates over the life of the project to evaluate the effect of inflation, it became difficult to adhere to the constant gas cost concept due to the fact that even though a given constant gas cost yielded the specified discounted cash flow rate of return, cash flows in many of the later years of the project became negative. This was due to the higher operating costs that were occurring in these later years. Increasing the constant gas cost by an amount to prevent negative cash flows yielded an after tax cash flow pattern that returned cash at a much higher rate of return than the specified 12%. Since it was felt that the occurrence of negative cash flows was not representative of realistic situations, it was decided to allow the gas cost to increase over the life of the project to prevent negative cash flows and yet yield the specified 12% rate of return on equity capital. Therefore, the cost of K-T gas is not constant but varies over the project life for those cases in which the effects of inflation were considered.

The repayment of debt capital was set up in the manner of a sinking fund, i.e. constant annual payments consisting of interest and principal were made such that by the end of the twenty year period the entire principal had been retired. Other cost items were handled as follows:

1. Working capital was added back to the project's cash flow in the twentieth year of the project.

2. Accelerated depreciation (sum of the years digits method) was used for tax purposes.

3. Startup costs and interest during the construction period were amortized over the project life and were included in Column 5 - Depreciation and Amortization. Interest during the construction period was financed by debt.

SAMPLE CASH FLOW STATEMENT:
DISCOUNTED CASH FLOW RATE
OF RETURN METHOD

BASES: FIFTEEN GASIFIER PLANT
60% DEBT/40% EQUITY
16% DCFRR
INTEREST RATE ON DEBT = 9%
COAL COST $20/TON
PLANT INVESTMENT = $367,000,000
DEBT CAPITAL EMPLOYED = $272,442,000
EQUITY CAPITAL EMPLOYED = $181,628,000
TOTAL CAPITAL REQUIREMENT = $454,070,000

1 Year n	2 Gas Cost $/MMBtu	3 Gas Revenue	4 Operating Costs	5 Depreciation & Amortization	6 Constant Payment to Retire Debt	6A Interest Portion of Col. 6	6B Principal Portion of Col. 6	7 Taxable Income (Cols. 3-4-5-6A)	8 Income Taxes (.50 x Col. 7)	9 Cash Flow (Cols. 3-4-6-8)	10 16% Discount Factor $(1+i)^n$	11 Discounted Cash Flow (Col. 9 x Col. 10)
1	$3533	$163,808,906	$94,683,000	$41,549,643	$29,845,061	$24,519,780	$5,325,281	$3,036,482	$1,528,241	$37,752,603	0.86207	$32,545,348
2				39,472,161		24040,505	5,804,556	5,619,840	2,806,620	36,474,224	0.74316	27,106,291
3				37,394,678		23,518,095	6,826,966	8,213,138	4,106,566	35,174,276	0.64065	22,534,671
4				35,317,196		22,948,668	6,896,393	10,860,048	5,430,021	33.850,823	0.55229	18,695,509
5				33,239,714		22,927,992	7,513,069	13,558,199	6,779,099	32,501,745	0.47611	15,474,504
6				31,162,232		21,631,466	8,193,605	16,318,217	8,156,108	31,124,736	0.41044	12,774,907
7				29,084,750		20,914,082	8,931,029	19,127,124	9,563,562	29,717,282	0.35383	10,514,852
8				27,007,268		20,110,258	9,734,821	22,008,398	11,004,199	28,276,645	0.30502	8,625,097
9				24,929,786		19,234,105	10,610,956	24,968,018	12,481,007	26,799,857	0.26295	7,047,097
10				22,852,304		18,279,119	11,565,942	27,994,483	13,997,241	25,283,603	0.22668	5,731,378
11				20,774,821		17,238,184	12,606,877	31,112,900	15,556,450	23,724,594	0.19541	4,636,148
12				18,697,339		16,103,566	13,741,495	34,325,001	17,162,500	22,118,944	0.16846	3,726,119
13				16,619,857		14,866,853	14,978,250	37,439,217	18,819,608	20,461,236	0.14522	2,971,516
14				14,541,127		13,518,796	16,326,245	41,064,740	20,532,370	18,748,474	0.12519	2,347,222
15				12,464,895		12,049,426	17,795,635	44,611,587	22,305,793	16,975,081	0.10793	1,832,067
16				10,387,411		10,447,816	19,397,242	48,290,676	24,145,338	15,136,506	0.09304	1,408,215
17				8,309,928		8,702,067	21,142,994	52,113,910	26,056,945	13,273,889	0.08021	1,064,653
18				6,232,446		6,799,198	23,045,863	56,094,261	28,047,130	11,233,714	0.06914	776,747
19				4,154,964		4,725,078	25,119,991	60,245,871	30,122,935	9,137,909	0.05961	545,877
20				2,077,482		2,464,271	27,380,790	64,584,152	32,292,076	24,786,768 ②	0.05138	1,273,783
TOTALS				$436,270,000 ①			$272,442,000					$181,628,000 ③

NOTES:

① ANNUAL DEPRECIATION AND AMORTIZATION WERE CALCULATED BY THE SUM OF THE YEARS DIGITS METHOD AND THE TOTAL SUM CONSISTS OF:

PLANT INVESTMENT	= $367,000,000
START-UP COSTS	= 7,340,000
INTEREST DURING CONSTRUCTION	= 61,930,000
TOTAL	= $436,270,000

② $17.8 MILLION IN WORKING CAPITAL WAS RETURNED TO PROJECT AND WAS ADDED TO CASH FLOW IN YEAR 20.

③ THE SUM OF THE DISCOUNTED CASH FLOWS OVER THE TWENTY YEAR PERIOD IS SEEN TO BE EQUAL TO THE INITIAL EQUITY IN THE PROJECT AT A 16% DISCOUNT RATE.

THE RILEY-MORGAN GASIFIER

by

THOMAS F. WALSH, Manager, Steam Generator Sales
RILEY STOKER CORPORATION
WORCESTER, MASSACHUSETTS

Presented at the
THIRD ANNUAL INTERNATIONAL CONFERENCE
ON COAL GASIFICATION AND LIQUIDATION
UNIVERSITY OF PITTSBURGH
SCHOOL OF ENGINEERING
Pittsburgh, Pennsylvania
AUGUST 3-5, 1976



THE RILEY-MORGAN GASIFIER

by

THOMAS F. WALSH, *Manager, Steam Generator Sales*
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During this conference, as at so many others in recent years, the theme of the great energy crisis and the need for a synthetic fuels industry to help overcome the problem has often been heard. Considering the progress we have made to date on commercial applications of synthetic fuels, you have to wonder if anyone has been listening. Perhaps this is an unfair conclusion, more likely the great majority of people in industry have been listening but are as confused as myself by the many contradictory statements being issued by an ever increasing number of sources. Here are a few statements that have caused confusion to our marketing program:

The Federal Energy Administration recently issued the 1976 Executive Summary on the National Energy Outlook. It stated, "Over the next 10 years, the nation can greatly expand its domestic energy production and cut the rate growth in energy demand and still meet its economic objectives. The post-1985 prospects for maintaining independence are less certain, unless technological and economic breakthroughs occur." In determining how the nation will meet its growing energy demands by 1985, the FEA makes the following statement: "Oil production could reach 13.9 MMB/D (including natural gas liquids), if Outer Continental Shelf leasing is strongly pursued and market prices prevail."

Utility Spotlight issue of June 17, 1976 states, "Interior Secretary Kleppe delays two oil and gas leases off Atlantic Coast until next year."

Business Week issue of July 5, 1976 indicates that the bottom has fallen out of the offshore drilling business with 50 rigs laid up compared to only five rigs 18 months ago.

Meanwhile, the American Petroleum Institute reported as of December 31, 1975 that our proved reserves of crude oil peaked in 1970 at 39 billion barrels and has declined to 32.7 billion at the end of 1975. Production from old wells continues to decline and now provides only 60% of our present requirements. The cost of importing the other 40% is approaching $30 billion a year.

What hope does the FEA hold out for us on natural gas? They state, "Natural gas production could reach 22.3 trillion cubic feet (Tcf) if new gas prices are deregulated, but could be 17.9 Tcf under current regulations."

The American Gas Association reports as of December 31, 1975 that our proved reserves peaked in 1967 at 290 Tcf. With the exception of 1970 when we added the Alaska reserves, the trend has been steadily downward. At the end of 1975, The AGA reports the proven reserves at 228 Tcf including 32 Tcf in Alaska. Production is down from a high of 23 Tcf in 1973 to under 20 Tcf in 1975.

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The government is presently considering three routes for the natural gas pipeline from Alaska. It has been reported recently that FEA wants a decision on the Alaska route by October 1, 1977. Assuming the same rate of progress on the gas pipeline that we have seen on the oil pipeline, it is hard to believe that we will get any benefit from the source before 1985.

Much has been said in recent years about high Btu gasification projects offering relief from the natural gas shortage. Numerous announcements have been made of projects to become operational in the 1980's. The announced projects would provide less than 5% of our current gas usage. An article in the May 17 issue of Business Week indicates that all of these projects are held up because of the high cost and lack of financing. Without loan guarantees the projects are probably dead.

The Energy Research Digest letter of June 21, 1976 reports that the synthetic fuels loan guarantee bill is stalled again in the House Commerce Committee. The subcommittee draft eliminates loan guarantees for all coal gasification projects.

The more you read about the problem, the more confusing it gets. We talk and talk about the energy shortage and Energy Research and Development Association spends billions on research and development projects to solve our post-1985 problems, but the real world of industry is today. What will happen to industry and the American economy if our weather pattern changes and the next two or three winters are extremely cold? Our demand for gas will far exceed our production rate and industry will suffer because of its low gas priority.

Most processes presently firing natural gas cannot be converted to direct coal firing. Some cannot even be converted to oil and may have to shut down unless they receive a higher priority from the FEA. Those that can convert their processes to oil will be more fortunate but will face an ever increasing cost for oil. If the present fighting in Lebanon continues to escalate, the Near East oil supply could be disrupted and a sufficient supply may not be available at any price. At the present time converting to an all electric economy is no answer, since recent cancellations of fossil and nuclear power plants could lead to a power shortage by 1980.

It would be nice to think that the Riley-Morgan Gas Producer is the answer to all of the energy problems. Like most big problems, however, there is no *one* right solution but many solutions to small parts of the problem. How a part of this problem can be solved economically by drawing on an historical method of gas production is Riley's approach to the solution.

Most industrial firms have long since abandoned the idea of generating their own electricity or manufacturing their own gas, but it should be remembered that both were a standard practice in the early part of this century. With the current gas shortage and the demand of consumer groups for flat electric rates, we could very well see a return to both practices. Figures 1 through 3 show a typical gas producer installation in a steel plant during the 1920's. The units shown in these photographs are the old Morgan gas producers of which over 9,000 were sold through the 1940's. Traditionally, these units were air blown and operated on anthracite, low-caking bituminous coal or coke. If industries located near a source of anthracite or low sulfur bituminous coal were to adopt this old reliable process, they could assure themselves of a supply of gas at approximately the present price of imported oil.

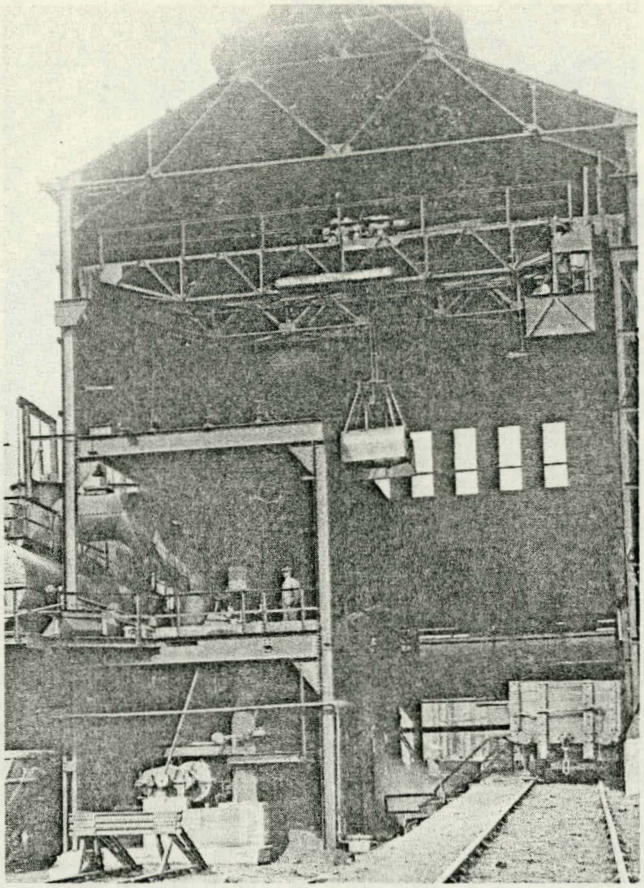


Figure 1

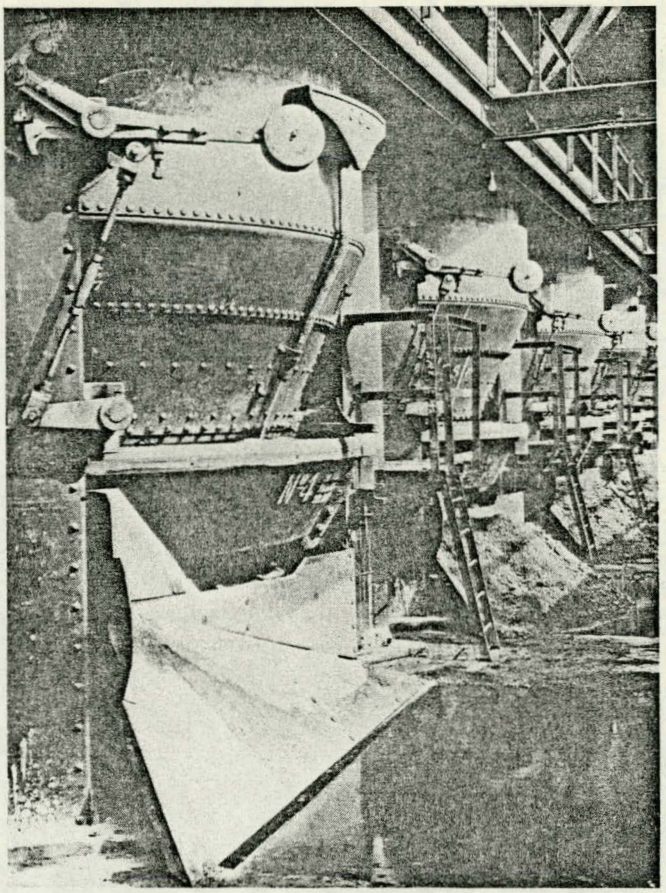


Figure 2

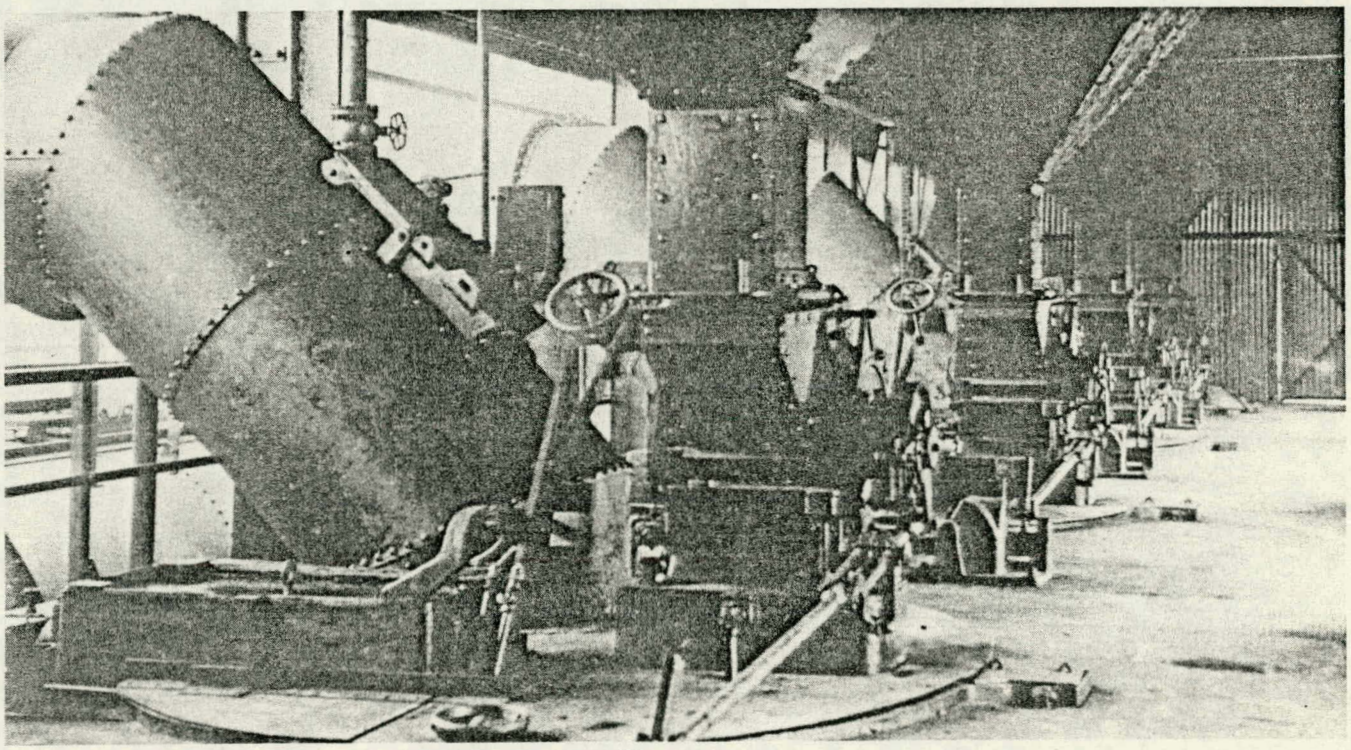


Figure 3

Typical Gas Producer Installation in a Steel Plant in the 1920's

This process can be adapted to other fuels such as high-caking or high-sulfur bituminous coals and can be operated on oxygen to provide an intermediate Btu gas with an average heating value of 300 Btu/cu. ft. . At the present time, however, the added cost of tar and sulfur cleanup and oxygen equipment require relatively large installations with high load factors to be economical.

Riley Stoker Corporation obtained the rights to this fixed-bed gas producer from the Morgan Construction Company of Worcester in late 1973. Figure 4 shows the design of the last Morgan producer, complete with riveted construction, large castings, and numerous potential OSHA problems.

Our first project was to redesign this unit for modern manufacturing practices. Castings were replaced with weldments. An outer gas and dust tight casing was provided for personnel protection. Water seals were added for higher operating pressure. The volume of the producer was increased to handle the swelling tendency of certain coals. A water-cooled agitator was added to handle caking coals. Finally, automated controls were provided to reduce the number of operating personnel. The new Riley-Morgan design is shown in Figure 5.

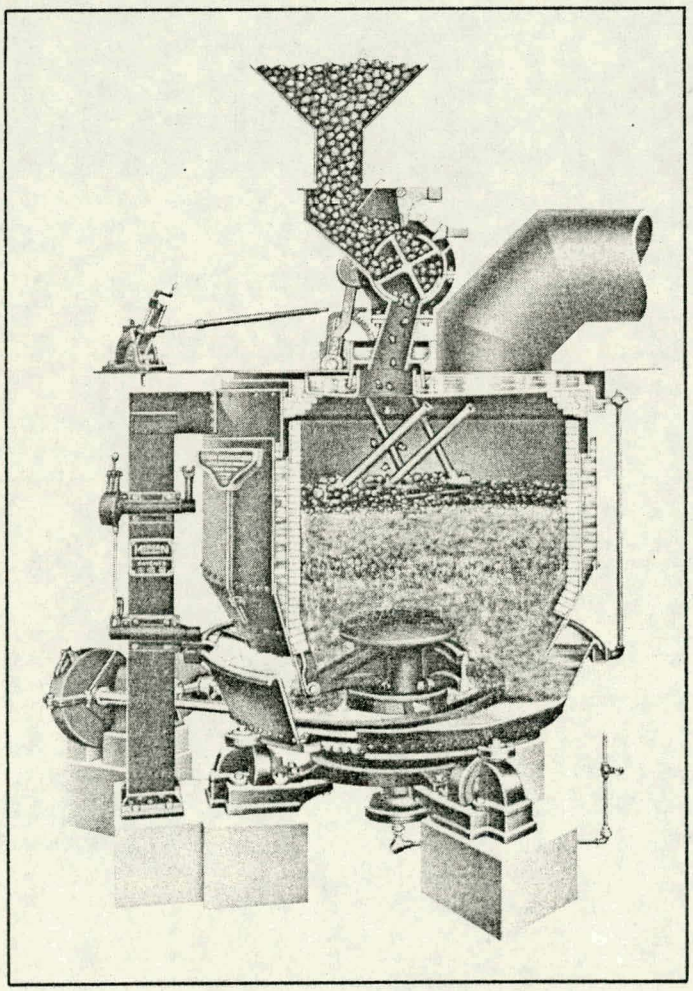


Figure 4 Morgan Gas Producer

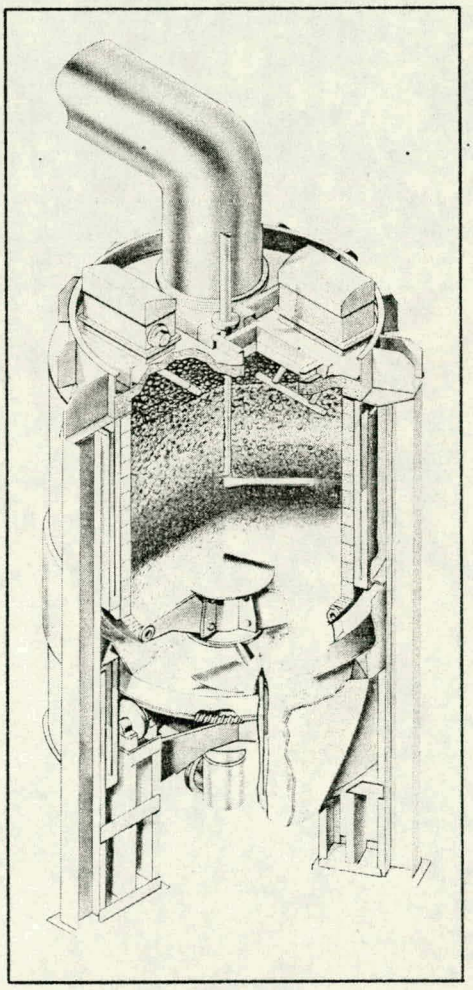


Figure 5 New Riley-Morgan Gas Producer

Merely updating a design was not enough to prove to ourselves or potential customers that the equipment would work under present day restrictions. Riley then began two parallel programs to develop operating data and techniques. In the early part of 1974, Riley moved into its new Research and Development Center in Worcester which is shown in Figure 6. In the high bay portion of this facility, a small pilot plant was installed as shown in Figure 7. A schematic diagram of this pilot plant is shown in Figure 8. The pilot plant was built to provide operating experience and to explore problems associated with tar formation from bituminous coals. This unit was operated for over a year on low ash fusion and highly caking varieties of eastern bituminous coals, using both air and oxygen.

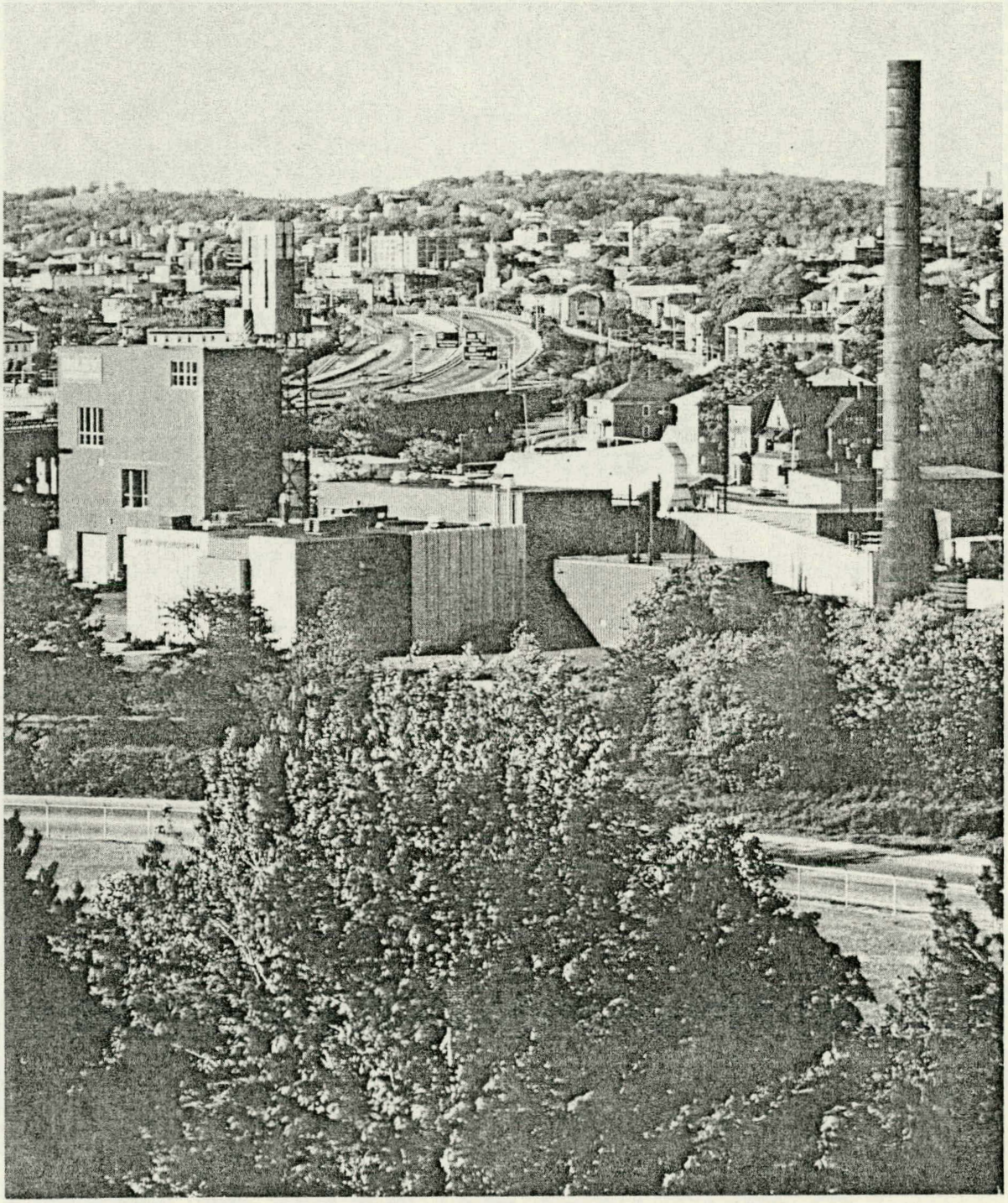


Figure 6 Research and Development Center

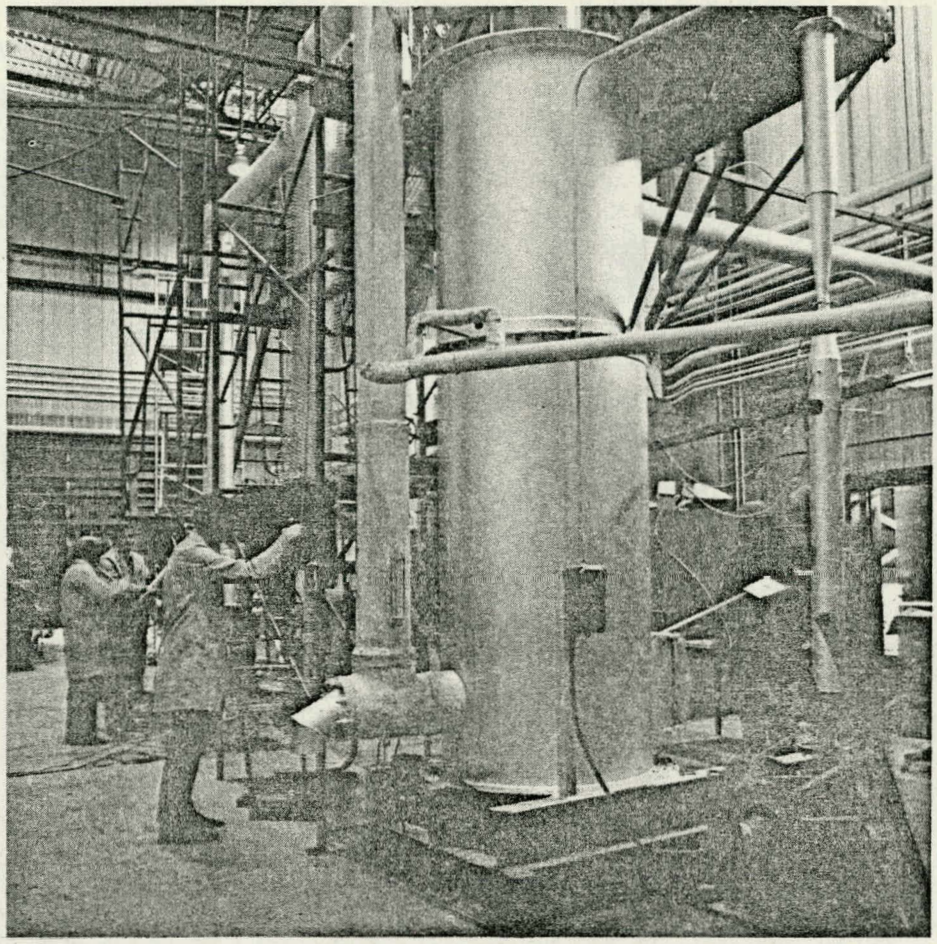


Figure 7 Pilot Plant

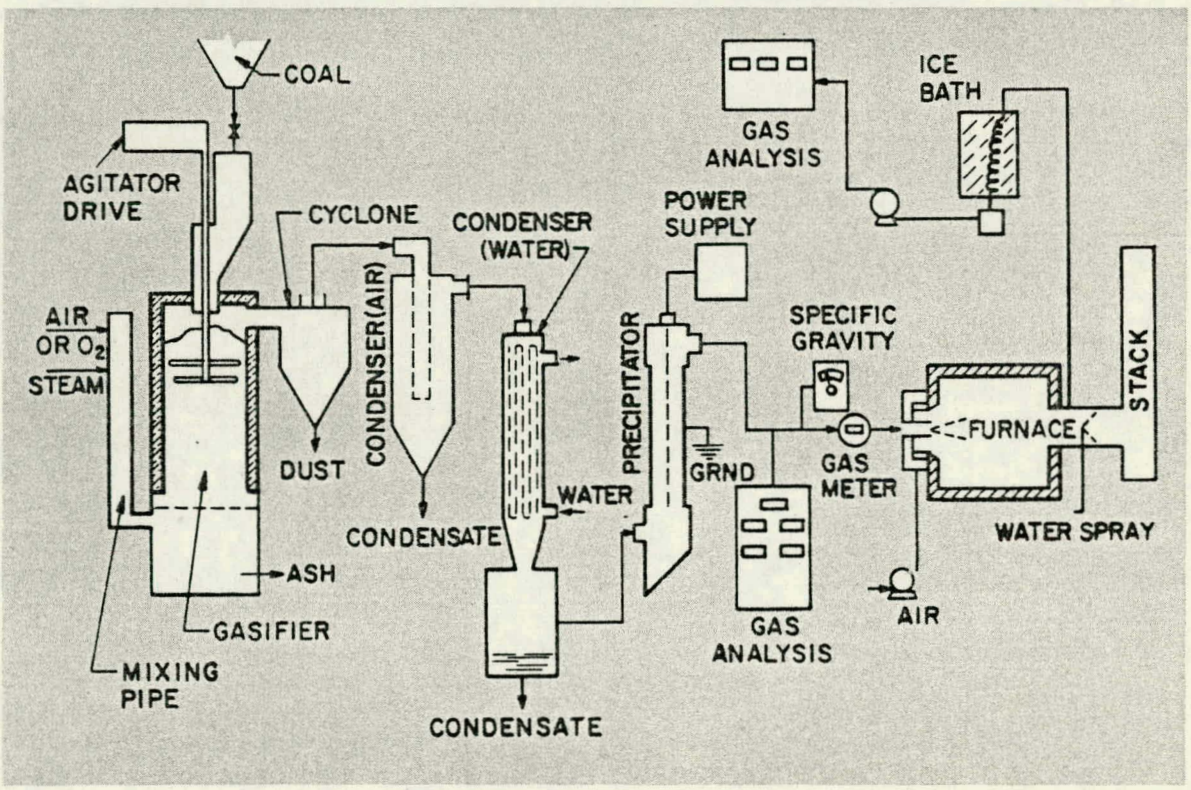


Figure 8 Pilot Plant Schematic

During June of 1975, experience gained from the pilot plant was transferred to the full-scale commercial size installation housed in the building shown in Figure 9. For the past year we have been testing various coals, and Riley is presently completing a series of tests on an Illinois No. 6 coal using air, enriched air, and pure oxygen. Testing of western sub-bituminous coals and lignites is planned if the market outlook warrants. The actual gas producer is shown in Figure 10. The control room of this research and development installation is shown in Figure 11.

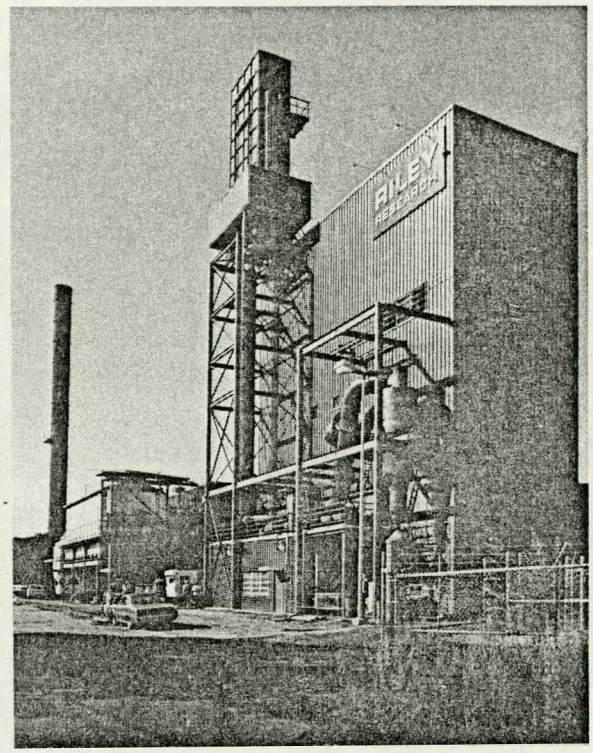


Figure 9 Commercial-Size Gas Producer Installation

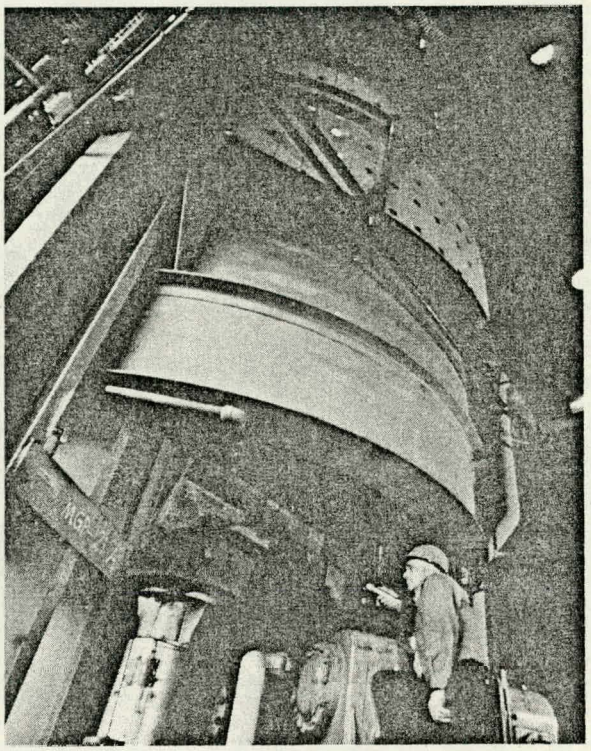


Figure 10 Completed Gas Producer

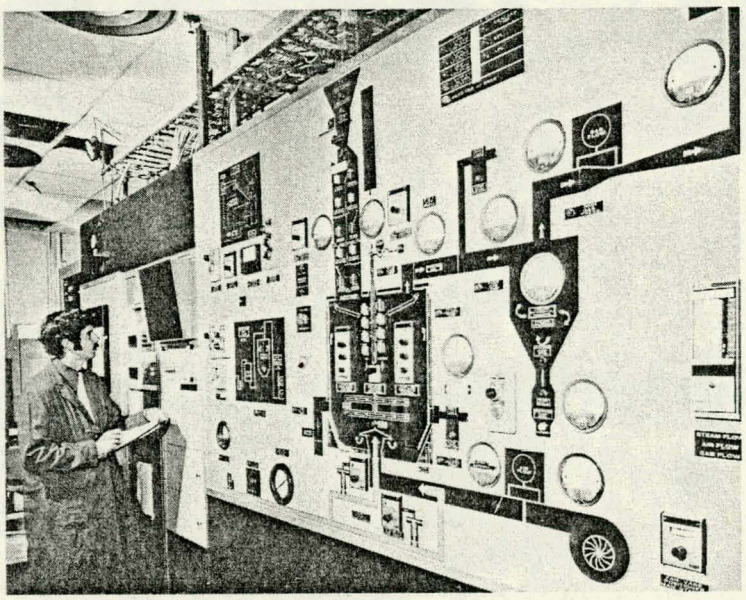


Figure 11 Gas Producer Control Room

Typical mechanical problems associated with a project of this type have been successfully solved. Riley's philosophy has been to solve potential problems at the Research and Development Center rather than on a customer's installation. The biggest problem the company faces is the cost of operating the plant which is entirely financed by Riley. To date, close to $3 million has been spent, which includes both installation and operating costs. Riley has paid as high as $93 per ton for sized anthracite coal. Over $46 per ton was paid for transportation charges on western sub-bituminous coal contributed by the Peabody Coal Company. Oxygen costs $50 per ton plus $1,100 per month for rental of storage equipment. Considering that the gas produced is flared off, you can realize the cost of acquiring data and why so few organizations have undertaken projects of this type.

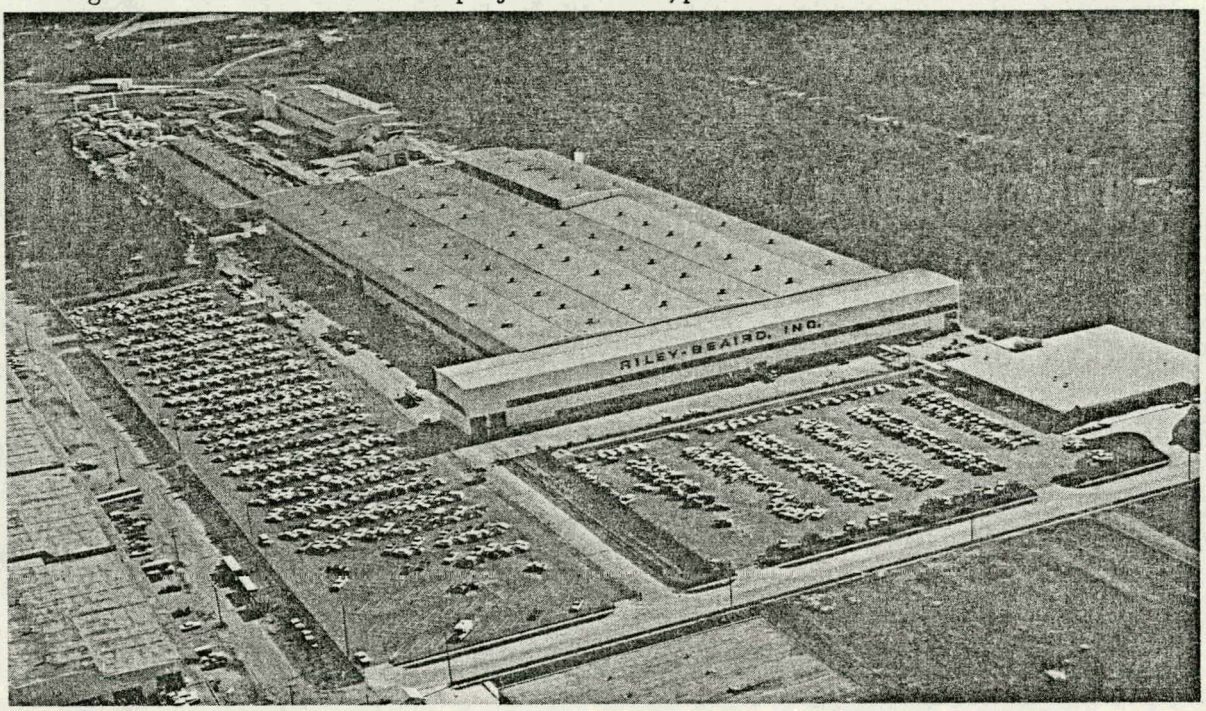


Figure 12 Riley-Beaird, Inc., Shreveport, Louisiana

Because of Riley's elaborate test program, the uncomplicated design of the gas producer, and our large fabricating facilities, as shown in Figure 12, we feel that we can react rapidly to the need of those industries which require gas for their process. While this is not a solution for all energy problems, it is a solution that may keep many companies from shutting down during a real crisis. To meet this crisis, however, the planning and commitment will have to be made soon.

The situation reminds me of a current problem in New England. Everyone is demanding more oil at lower prices and are at the same time fighting off-shore drilling and the locating of oil refineries in their towns. I was therefore amused at seeing a bumper sticker in one of the gulf states that read, "Let the – – – – – freeze in the dark." Let's hope this country does not have to start freezing in the dark before industry takes some action.

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HYGAS® PROCESS UPDATE

by

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for presentation at the

Third Annual International Conference on
Coal Gasification and Liquefaction
University of Pittsburgh
August 3-5, 1976

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HYGAS® PROCESS UPDATE

by

Karl S. Vorres

INTRODUCTION AND HISTORICAL BACKGROUND

The HYGAS® Process involves hydrogen gasification of coal and derives its name from that reaction. The main product of the process is substitute natural gas. The HYGAS Process is a second-generation high-pressure process designed to handle all ranks of coal. It is a method which can produce approximately two-thirds of the product methane in the hydrogasifier by direct hydrogenation. The process is carried out at a pressure of about 1000 psi and temperatures from 600 °F in a slurry drier to 1850 °F in the steam-oxygen gasification zone. It is currently being studied in a pilot plant designed to produce 1-1/2 million SCF/day of methane from 80 tons of coal. This presentation will give some background on the process and an indication of our progress to date.

Early studies of hydrogen gasification of coal at the Institute of Gas Technology (IGT) go back more than two decades. This led to contracts with the U.S. Department of the Interior, Office of Coal Research, and support by the American Gas Association (A.G.A.) starting in 1964. The objectives of the HYGAS program are to verify the technical and economic feasibility of the process and to develop the HYGAS concept as a candidate commercial process by the 1980's, as well as to develop an environmentally acceptable coal conversion process that can utilize all

types of coal. It could supplement natural gas supplies and reserves with gasified coal. This program is one of several pilot plant programs currently being supported by the U.S. Energy Research and Development Administration (ERDA), and in some cases also by the A.G.A. These processes include the Synthane Process, the CO_2 Acceptor, BI-GAS, and the Agglomerating Ash Burner.

The HYGAS Process is based on two processes for converting coal to synthetic pipeline gas. One of these involves the gasification of coal with oxygen and steam to produce a mixture of carbon monoxide and hydrogen which is suitable for methanation. The other involves direct hydrogenation of coal at high temperatures (1200°-1800°F) and high pressures (1000 psi) at moderate residence times. The present HYGAS Process incorporates principles developed in both of these concepts.

The resulting HYGAS Process gives a high yield (65-70%) of the product methane formed in the hydrogasifier by direct hydrogenation of coal. The use of hydrogen and steam in hydrogasification gives an opportunity to balance exothermic reactions with endothermic reactions in order to provide built-in temperature controls as well as internal hydrogen generation.

The HYGAS Process as presently operated is illustrated in Figure 1. Raw coal is crushed to pass a 1/8-inch screen and, if it possesses agglomerating properties, is sent to a pretreater. Otherwise, this step is omitted. The pretreater is a fluidized bed which permits a surface oxidation of the coal at atmospheric pressure and temperatures

in the vicinity of 800 °F with air. Some fuel gas is released in this step. The pretreated coal goes to slurry preparation which is not shown. This slurry is pumped to the top of the hydrogasifier. In the uppermost stage the light oil is removed and the dry solids contact a hot rising stream of hydrogen and other gases. The partially reacted coal falls to a lower portion of the reactor to have a further opportunity to react at a higher temperature with more hydrogen.

The residual char then passes to a steam-oxygen gasifier where the temperature is increased and oxygen reacts with carbon to form carbon monoxide which in turn may react with steam to form hydrogen. The spent material is discharged to the ash receiver. The product mixture goes to a gas purification train for quenching, removal of the light oil and carbon dioxide as well as sulfur compounds and ammonia. The remaining gas mixture, essentially carbon monoxide and hydrogen besides the methane, is in a desirable proportion for converting to methane in the methanator. The product gas may then be dried and placed in the pipeline.

Figure 2 indicates a cross section of the HYGAS pilot plant reactor. The structure is about 132 feet tall. It has been constructed as a balanced pressure reactor with a miaximum 30"-ID for the lower stages. The upper part of the reactor is pressurized between the shell and components with nitrogen. Pressures of 1000 to 1100 psi are maintained in the reactor.

Coal enters at the top as a slurry of up to 50-50 weight percent mixture with a light oil produced in the process. The upward flow of hot

product bases creates a drying zone in which the light oil is removed. The dried coal passes down to a mixing chamber to then undergo first-stage hydrogasification at about 1250 °F. Residence times vary up to 10 seconds before entering the gas-solids disengaging zone. Partially reacted char ghen goes from the first-stage hydrogasification to the second stage. This is a large fluidized bed where the temperature is increased to about 1750 °F depending on coal characteristics. A residence time of about 30 minutes is provided here. Char from second-stage hydrogasification is discharged to the steam-oxygen gasifier in the lowest part for steam and oxygen to produce synthesis gas for the upper stages. The gas flow proceeds vertically upward through the reactor. The raw gas contains a mixture of methane, carbon monoxide, carbon dioxide, hydrogen, steam, ammonia, sulfur bearing gases, and light oils as well as coal fines which did not react.

The first step in treatment of the raw gas is to quench it with a water spray. This gives a three-phase mixture of light oil, water and coal fines in addition to the product gas. The liquid-solid mixture is separated in a horizontal separator. Light oil is recycled to the slurry mixer. Coal fines are separated, filtered, and removed. The water is processed and recycled.

The product gas then goes to a diglycolamine scrubber to remove hydrogen sulfide and carbon dioxide. The amine goes to a steam stripper for removal of hydrogen sulfide which goes to a Claus plant for conversion to elemental sulfur. The lean amine solution is pumped back to the process gas scrubber.

The processed gas then goes to a two-stage methanation unit for conversion of residual hydrogen and carbon monoxide to methane.

The history of the development of the HYGAS Process goes back to the period 1944 to 1964 in which IGT carried out in-house and A.G.A.-funded work on the fundamental investigation of concept parameters. A contract signed in 1964 with OCR and A.G.A. provided joint funding to carry out process development unit studies to form the basis for a pilot plant design. In 1967, the design of the pilot plant was started. This was followed by construction and shakedown, as well as the development of an electrothermal gasifier. In 1971, the pilot plant operations were initiated with hydrogen produced from a reformer and an electrothermal gasifier was constructed. Also a steam-oxygen system for gasifying char was developed. In 1972 through 1974 the pilot plant systems were demonstrated in integrated operation. In 1974 the steam-oxygen gasification system was designed and installed. A light oil recovery unit was installed. Operations with lignite and bituminous coals have been carried out.

In the period 1944 to 1964, during the pregovernment sponsorship era the A.G.A. provided $3 million to fund efforts in a fundamental investigation of the concept parameters. During this time, synthesis gas was generated by steam-oxygen gasification of coal at rates up to 1500 lb/hr. Total methanation of synthesis gas was accomplished on nickel catalysts. The concept of high pressure (1000 to 1500 psi) and high temperature (1200° to 1800°F) hydrogasification of coal was established with hydrogen-steam mixtures. Hydrogasification tests were conducted in batch and continuous flow units in the range of 5 to 10 lb/hr of coal.

In the period of 1964 to 1967 contract efforts under partial OCR sponsorship led to process development unit (PDU) studies to form the basis for a pilot plant design. This work established the fluidized bed rather than a moving bed as a mode of gas-solids contact for ease of scale-up. Continuous PDU gasification tests were conducted at coal feed rates up to 100 lb/hr in a 4-inch-diameter reactor. The two-stage hydrogasification concept was verified for maximum direct formation of methane from coal over a temperature range up to 2000°F and pressures up to 2000 psi. Thirteen coals were tested ranging in rank from lignite through semi-anthracite. Pretreatment conditions were defined in a 10-inch-diameter PDU. Methanation catalysts were screened and a cold gas recycle system was developed for temperature control in methanators.

The period 1967 through 1971 extended the development, initiated the design, and carried out the construction of the pilot plant. IGT, together with Bechtel, developed a preliminary design of the pilot plant. This included the gasifier reactor design which involves multiple reaction stages in one vessel. The single-vessel concept eliminates thermal expansion problems. The gasifier internal diameter was set at 30 inches to reduce reactor heat loss and provide optimum intermediate scale-up data. Slurry feeding of coal was selected. The unreliable life of lock hopper valves led to design of a slurry system for feeding coal into and discharging residue from the gasifier. The detailed design, procurement, and construction of the pilot plant was carried out by Procon Incorporated and was followed by the shakedown of the pilot plant. The shakedown of the pilot plant was carried out from May 1971 to October 1971. The electrothermal

gasifier was also developed. The plant was designed to convert 80 tons of coal/day to 1.5 million cubic feet of pipeline-quality gas. It was also designed to handle lignite, bituminous, and subbituminous coals. Construction was carried out under a guaranteed maximum price (GMP) arrangement for a $7 million total. An electrothermal method of gasification for generating hydrogen from char in a 6-inch-diameter PDU reactor with a power input of 300 kW was tested at 1000 psi pressure and 1700° to 1900°F.

The period 1971 to 1972 included plant operation, construction of the electrothermal gasifier and development of the steam-oxygen gasification method for hydrogen production. In this time, seven initial tests were conducted using hydrogen from the reforming of natural gas. The electrothermal gasification pilot plant unit was designed and constructed by Procon to use up to 2 MW of ac power using on-line computer control for power input. It was also designed for integration with the HYGAS reactor for char feed and constructed under a GMP arrangement for $2 million. Studies of the steam-oxygen gasification method were completed for hydrogen generation from char in the 6-inch-diameter PDU reactor at pressures up to 1000 psi and temperatures between 1600° and 1900°F.

Demonstration of pilot plant systems took place during 1972 through 1974. The use of coal-oil slurry system was demonstrated for pressurizing coal feed, as well as the operability of IGT's cold gas recycle system for methanation. A refractory grid was developed for gas distribution and fluid bed support. Twenty tests were conducted with Montana

lignite culminating in Test No. 27 in March 1974. The test program was completed on batch gasification of char in the electrothermal gasifier. Operability was demonstrated (10 tests conducted). Further development of the unit was not pursued due to the high cost of conventional power generation.

The period of 1974 and 1975 involved design and installation of additional pilot plant equipment as well as operations with lignite and bituminous coal. The steam-oxygen gasification unit was designed, constructed, and integrated into the bottom of the HYGAS reactor. A light oil recovery unit was designed, constructed, and integrated into the pilot plant. Ten tests were conducted using lignite with steam-oxygen gasification. Test No. 37, completed in July 1975, was the most successful test in the pilot plant lignite program. The results of this test were self-sustained operation for 15 days. Over 1000 tons of lignite were processed. During the 15-day period the entire plant was onstream for 92% of the time or 330 hours. Steady-state conditions were achieved for 160 hours. Solids flow through the reactor was smooth. Amine acid-gas removal and methanation sections worked well. Full conversion of carbon monoxide was maintained. The light oil stripping section performed satisfactorily for the first time, and net production of oil was demonstrated. Overall material balances for three major sections, that is gasification, purification, and methanation, were essentially closed. The slurry feeding system operated at feed rates up to 2-1/2 tons/hour. The plant was in excellent condition on post-test examination. The test results were reviewed by the ERDA/A.G.A. Operating Committee and judged to have met the requirements of technical feasibility.

Pilot plant operations with Illinois No. 6 bituminous coal occurred from 1975 to the present. Pilot plant operations started in September 1975 with the bituminous coal. Test No. 54 achieved self-sustained operations for nine days of which seven were at 2-1/2 tons/hour and two were at 3 tons/hour of pretreated char to the reactor. This test gasified 640 tons of high sulfur caking coal. Self-sustained operation means that the plant produced gas from the chemical reaction of water, oxygen, and coal without supplemental sources of hydrogen and heat.

In this run, the coal was pretreated to 800°F prior to being fed into the reactor vessel. Good heat and material balances were recorded during the 214-hour run. The run was terminated voluntarily when the coal supply was depleted.

In summary, the overall accomplishments of the HYGAS program to date are that it has demonstrated the key process concepts and integrated unit operations. It has also demonstrated several methods of hydrogen generation, including catalytic steam reforming of natural gas, electrothermal gasification, and also steam-oxygen gasification. A total of 37 tests were conducted with lignite, including a total of 5500 tons of lignite processed. During this time, it demonstrated the technical feasibility of a process using lignite. A total of 17 tests were conducted involving a total of 3100 tons of bituminous coal. Some specific objectives of the HYGAS program for fiscal 1977 include tests to be conducted with subbituminous coal. Data will be collected for use in the design of an effluent treatment and water reuse cycles in a commercial plant. New methanation catalysts will be tested. Materials testing will continue.

Refractory and metals coupons prepared by separate contractors under sponsorship of the Metal Properties Council will be tested. Data will be gathered on performance of piping and vessels in corrosive or erosive environment.

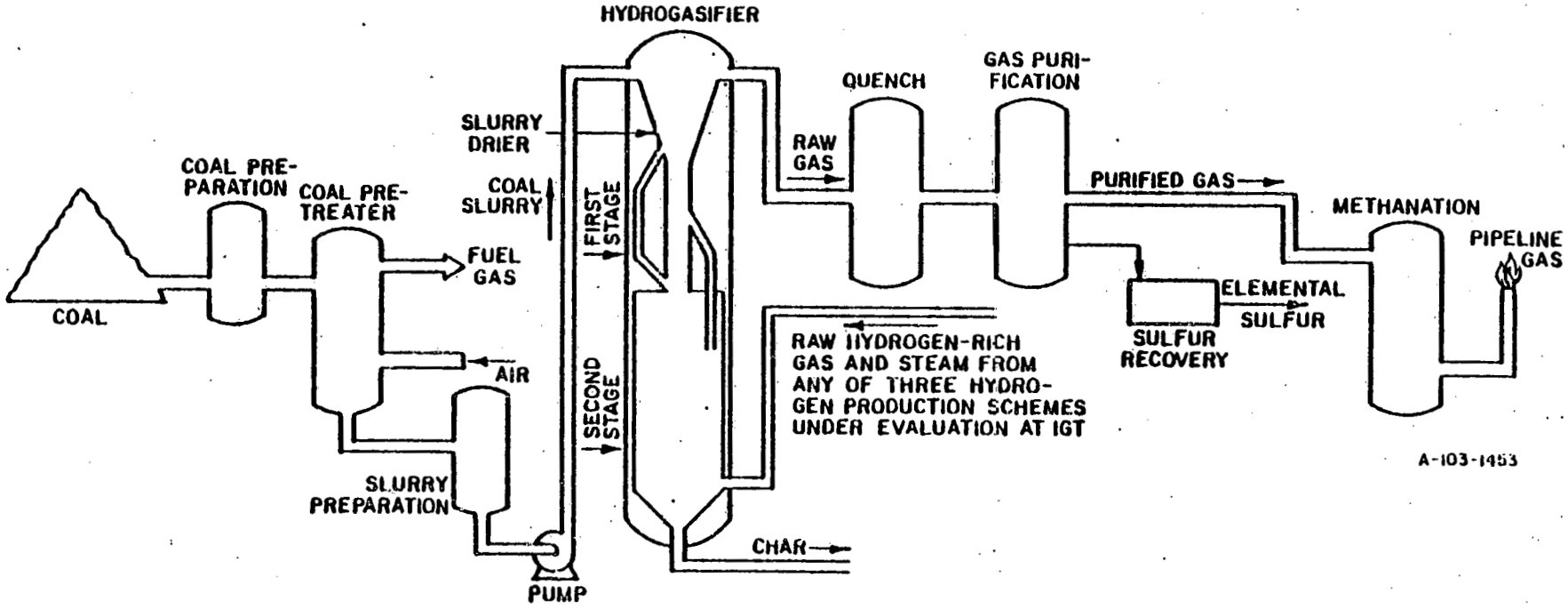


Figure 1. SIMPLIFIED FLOW DIAGRAM OF THE IGT HYGAS PILOT PLANT

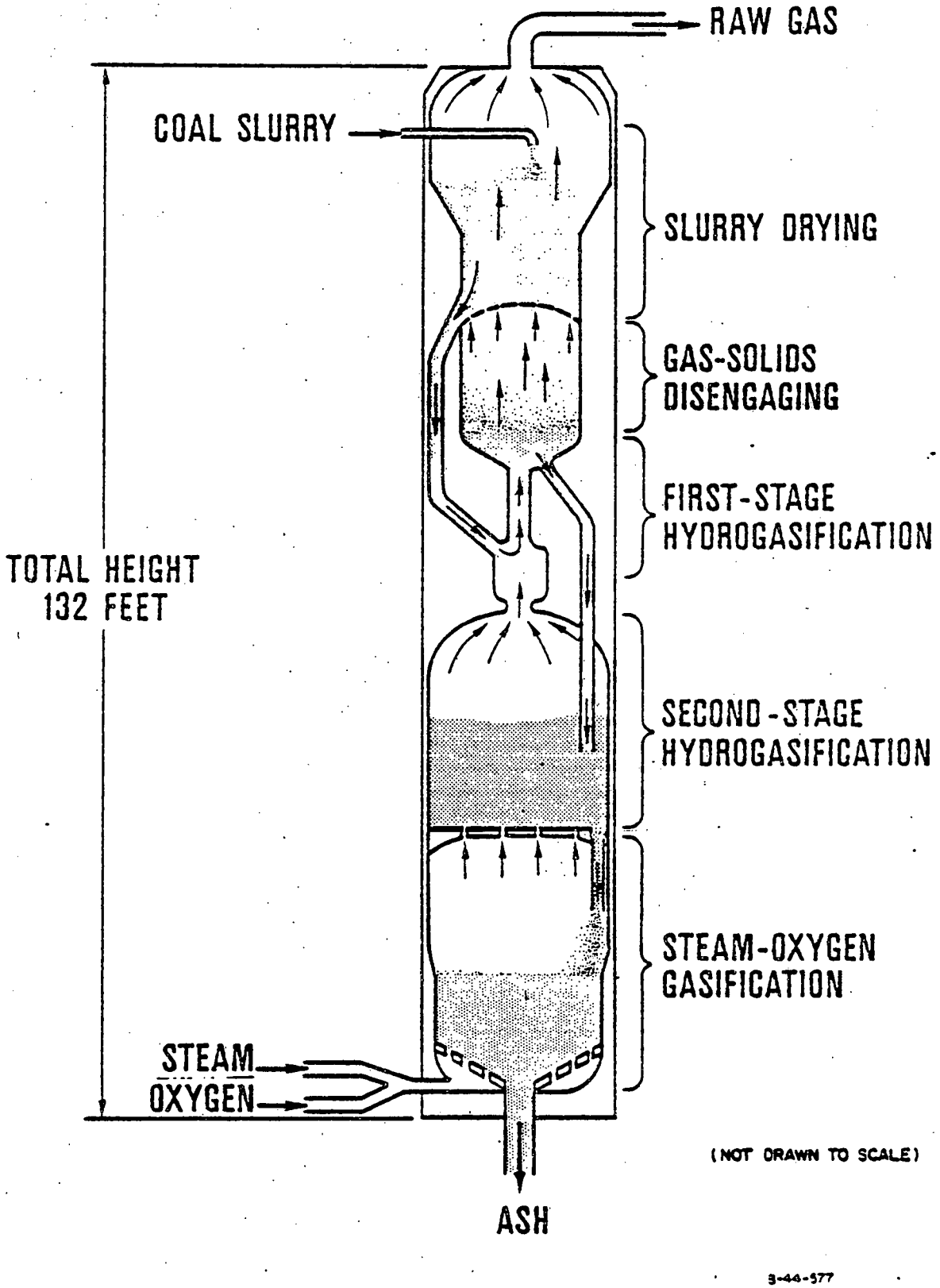


Figure 2. HYGAS PILOT-PLANT REACTOR

FLUE GAS DESULFURIZATION PROCESS AT THE D. BRUCE MANSFIELD POWER PLANT

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THE FLUE GAS DESULFURIZATION PROCESS AT THE D. BRUCE MANSFIELD PLANT

The Bruce Mansfield Plant is located in the southwestern part of Pennsylvania on the Ohio River in Shippingport, Pennsylvania. The plant is being constructed and will be operated by Pennsylvania Power Company, a subsidiary of Ohio Edison Company. It is owned by the Central Area Power Coordination Group (CAPCO) made up of Ohio Edison Company, Pennsylvania Power Company, the Cleveland Electric Illuminating Company, Duquesne Light Company, and the Toledo Edison Company. The original concept included the installation of two 880,000 (net) KW coal-fired generating units of conventional design with steam conditions of 3500 psi, 1000 DegF main steam and 1000 DegF reheat. The air quality control system envisioned at that time incorporated high efficiency, electrostatic precipitators, and a tall chimney. Engineering for the project started late in 1969 and construction started in June, 1971.

At the start of engineering the Commonwealth of Pennsylvania did not have a state-wide standard applicable to sulfur emissions. Engineering proceeded on the basis of Pennsylvania Regulation V (since superseded by Chapter 123.11) which established a very stringent particulate emission limitation. The Company was aware that SO_2 emission regulations might be promulgated sometime in the future and, therefore, engineering effort was directed toward the installation of an air quality control system satisfying the particulate emission regulation that could be later coupled to an SO_2 removal device of some kind. This prompted initiating an investigation during early 1970 of the state-of-the-art and the expected commercial availability of SO_2 removal systems.

When engineering design had proceeded to the point at which the Company could notify the Pennsylvania Department of Environmental Resources (DER) of the design concept and sizing of major equipment, a partially completed form for Application for Construction Approval was submitted. In November of 1970 the Pennsylvania DER advised the Company that they had reviewed the preliminary Application and had concluded that since the design did not include a full-scale SO_2 removal system it was unlikely that approval of the Application would be recommended, even though Pennsylvania apparently had no applicable SO_2 emission limitation at that time. Nevertheless, this notification accelerated and intensified a series of studies then in progress in connection with SO_2 removal systems.

The studies concluded that the technology associated with SO_2 removal systems was then in an early state of development and totally unproven. For this reason, the Company proposed discussions with the Pennsylvania DER and the Federal Environmental Protection Agency (EPA) to build a module of 20% to 25% size to serve #1 unit. This module would provide the vehicle to solve operating, chemical, and disposal problems. It would also prevent

severe operating problems and outages of equipment required to made modifications on the full number of modules in a system installed to treat all of the boiler flue gas and would allow the remaining portions of the SO_2 removal modules not initially installed to be successfully operated at an earlier date. In the interim period a tall chimney (800' to 1000') would prevent ground level SO_2 concentrations from exceeding the ambient air quality standards. (A 950' chimney height was later recommended pursuant to a study performed by Battelle Columbus Laboratories.) It was further pointed out that the single module approach would avoid the cost of duplicating modifications, which no doubt operating experience would prove necessary.

As a part of this study a world-wide search was conducted for vendors capable of supplying a full-scale SO_2 removal system of the size required on the Mansfield units with a performance warranty that would satisfy the requirements of the Pennsylvania SO_2 emission regulations. Only one vendor was willing to undertake this full-scale project, and that vendor limited the performance warranty to a very modest amount. The DER and EPA did not approve the application for a 20% to 25% module for Unit #1 and, therefore, a full-scale SO_2 removal system was constructed.

The study concluded that the only arrangement showing any promise of achieving the level of SO_2 removal efficiency required by the Regulations, about 92%, was the installation of several trains of two-stage wet lime scrubbing. Six trains per unit were chosen in order to minimize the magnitude of scale-up in size and to provide a spare train for each unit.

Pennsylvania Power Company also commissioned a parallel study with a company to determine and evaluate problems involved in the disposal of wastes produced by the lime throw-away system. After very detailed studies of several methods of disposal it was concluded that the only one available that provided reasonable assurance of continued operation of the plant was the treatment of the sludge with a compound then under development that would tend to cause it to solidify over some period of time after pumping it to an impounded disposal area. The large quantities of sludge generated, expected to be upwards of 2,000,000 tons per year from the two units, require a very substantial land area for disposal. To satisfy this need, some 1400 acres of property were purchased (an entire valley) about six miles from the plant site. A 420' high hydraulic dam is being constructed to contain the sludge in this area.

The two-stage venturi scrubber for Mansfield Units 1 and 2 is designed by Chemico. As indicated above, six trains are installed, each train consisting of a scrubber vessel, induced draft fan, and an absorber vessel. There are two oil-fired reheaters. Three scrubber trains discharge into

a single reheater. The scrubber and absorber vessels are about 35' in diameter and 50' high. The vessels and duct work are lined with polyester flakeglass material. The induced draft fan housing is lined with rubber, and the induced draft fan rotors are made of Inconel 625.

The flue gas enters the top of the scrubber vessel, passes down and around an adjustable plumbob through the venturi throat, turns 180 deg., passes up through a mist eliminator out of the absorber vessel and into the suction of the induced draft fan. The adjustable plumbob makes it possible to control the velocity of the gas at the throat of the absorber. Pressure drop across the throat is maintained at approximately 25" water. Scrubber liquor is circulated with two recycle pumps taking their suction from the base of the scrubber vessel, and then pumping the liquor to the top of the vessel where the venturi throat and plumbob surfaces are wetted. The pumps also supply liquor to secondary sprays located in the gas stream ahead of the mist eliminator. An intimate mixing of the gas and liquor in the venturi throat and with the secondary sprays is designed to remove practically all of the particulate matter and about 70% of the sulfur dioxide.

From the induced draft fan the gas enters the top of the absorber vessel, passes down through a fixed throat venturi, turns 180 deg. and passes through a mist eliminator, leaves the absorber vessel, and enters the reheater. Absorber liquor is circulated with two recycle pumps in a manner similar to the scrubber. The absorber is designed to remove the remaining particulate matter and enough sulfur dioxide to meet the Commonwealth of Pennsylvania emission standards. The water-saturated gas enters the reheater at about 120 DegF and is designed to leave the reheater at about 160 DegF, thus evaporating any moisture carryover, adding buoyancy to the gas and preventing precipitation from the stack plume in the vicinity of the plant. The pH of the scrubber and absorber liquor is maintained at 7 by the addition of lime to these vessels.

The liquor in the scrubber and absorber contains calcium sulfite and calcium sulfate, lime and fly ash. Enough of the liquor in the absorber is bled from the discharge of the recycle pumps to the scrubber vessel to maintain 8-10% solids in the absorber liquor. The scrubber vessel liquor is bled in the same manner but to a 200' diameter thickener located on the plant site. Thickener overflow water is used to maintain a liquid level in the scrubber and absorber vessels. Thickener underflow is controlled to approximately 30% solids and pumped to the 1400 acre impoundment site with large reciprocating sludge pumps.

The pumping station is located on the plant site and pumps the sludge over six miles to the impoundment area without booster pumps. The Dravo Corporation designed and built the sludge handling system including the hydraulic dam for the impoundment area. Dravo will also supply the lime and the sludge hardening agent called calcilox for this project. Calcilox is added to the sludge on the plant site at the sludge pump suction. This material is designed to solidify the sludge to the extent that it will support about four and one-half tons per square foot or equal to solid earth. Operating costs are high. This system almost doubles the station power requirements, and along with the circulating water cooling tower, reduces the net output from an original design of 880 MW to 825 MW.

The unit was placed on the line December 11, 1975, and was placed in full commercial operation June 1, 1976. The overall unit operation has been satisfactory. Operating experience with the scrubbers has exposed certain problems requiring modifications to the system. The cavitation of the recycle pumps caused pulsing of the liquor flow and, therefore, an unstable gas flow through the scrubber vessel. Baffle plates were installed over the pump suction openings in the scrubber vessel which solved the problem but required several modifications of the baffles before one was found that could withstand the forces exerted upon it and the corrosive atmosphere that it was exposed to. One of the scrubber mist eliminators was plugged up very early in the operation of the system probably because of the gas flow through the scrubber being much higher than the design flow.

The material plugging the mist eliminator hardened to a point where it could not be removed and the mist eliminator had to be replaced. The mist eliminators are normally cleaned by intermittent spraying with recycle water. After the problem of the mist eliminator pluggage occurred, a system was installed that permitted the mist eliminator to be flooded with large volumes of water in case the pressure drop across the device became excessive. The mist eliminators, even at design flow, are not meeting the design criteria of one grain per cubic foot carryover. Tests have indicated the water carryover in the mist eliminator is on the order of two to three grains per cubic foot.

To solve this problem a section of vertical mist eliminator is planned for the second stage outlet. The carryover problem has caused a calcium sulfate scale to build up in the chimney flue. This scale flakes off when the outside air temperature fluctuates, is carried out of the chimney and drops to the ground in the vicinity of the stack. The gas velocity in the chimney at full load is approximately 90' per second.

An attempt was made to place the oil-fired reheaters in service in February, 1976. Several problems were encountered but the major stumbling

block was a severe vibration within the burner. This problem has yet to be solved, and the unit is operating without reheaters as of this date. With the exception of the one incident with the mist eliminators, only a little plugging or scaling has been encountered.

Performance testing of the scrubbers has not been done because the reheaters do not work, and in order to determine if the system meets the State standards, it is necessary to do the testing with the reheaters in service.

With the limited experience acquired to date at the Mansfield Plant, it is impossible to make predictions about the future of this particular desulfurization system. However, major modifications are necessary, the cost of which will be measured in millions of dollars. A wiser course of action would have been to experiment with one scrubber train to develop the process and then install the full-size system.

ENVIRONMENTAL ASPECTS OF COAL CONVERSION PLANT SITING AND COST OF POLLUTION CONTROL

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V. COST OF POLLUTION CONTROL

I. INTRODUCTION

I-1. GENERAL

The often complicated though important government regulations, requiring the construction of a coal conversion facility in an environmentally acceptable manner, can impose severe restrictions on the location of such facilities. This is especially true in geographical areas where a delicate balance between various natural environmental systems exists. It is difficult to precisely define this environmental balance in a general paper like this since it is unique in each of the geographical areas that are potential sites for the location of coal conversion facilities. However, it can be explained, in terms of adverse and irreversible damage to the environment, as follows:

> if a certain component of the environment is damaged by the intentional or unintentional introduction of new materials it would eventually destroy all the environmental systems.

In an effort to prevent damage of environmental components like the flora and fauna, certain pollution control regulations applicable to coal conversion plants have been established. These regulations, on the Federal, state and local levels, in most cases require the use of best available control technology. They may in certain cases also require the development of new technology or modification of the process itself.

Environmental analysis of coal conversion facilities have so far only considered individual components mostly limiting it to the facility itself. Comprehensive environmental impact statements involving air, land, water and noise pollution, and the entire sequence of activities starting from the mining of coal to the distribution of products have not been developed. However, when one considers the size of these facilities, the overall pollution problem is greatly simplified by having the pollutants generated at a single processing site rather than at multiple points of coal utilization. This does not reduce the complexity of the technical problems associated with the technology involved in controlling pollution from such facilities but it greatly reduces the overall economic impact.

I-2. COAL CONVERSION

Coal conversion normally refers to processes for converting all types of coal, i.e., anthracite, bituminous, sub-bituminous and lignite, to a product generally more acceptable from an environmental point of view, of a higher heating value, more expensive in terms of cost per unit of energy (¢ per million Btu or KCAL), and to various other organic and inorganic chemicals. The first known coal conversion process was developed in the 17th century and is referred to now as carbonization or destructive distillation of coal for the production of coke. Coke production, solvent refining or chemical cleaning of coal, production of low, intermediate and high heat content gas, manufacture of synthesis gas, manufacture of methanol, production of liquid feed stocks, etc. fall into the category of coal conversion technology. Coke production, by virtue of its association with the iron and steel industry can be treated as a separate entity and, therefore, I will exclude it from this paper. However, it should be recognized that coke production is one of the dirtiest operations and has been quite difficult

to clean up.

Coal at present forms the world's largest reserve of concentrated organic raw materials and for many essential industries from dyes, medicines, pesticides, and elastomers to modern plastics shares to a lesser extent with petrochemicals the furnishing of the basic raw materials. However, it is our fundamental fuel and will remain so for a long time to come.

Coal is a hetrogeneous substance, both physically and chemically, and this hetrogenicity can be attributed to the fact that the parent material itself was highly hetrogeneous, originating from different plant, and to a smaller extent animal, forms mixed with inorganic sediments. Its structure consists of aromatic ring clusters generally no larger than 6 or 7 rings tied together by bridge linkages such as methylene chains of 1 to 4 carbon atoms, ethers, sulfides, disulfides, and biphenyl groups (1)*. A representation of the bituminous coal structure is shown in Figure 1 (2). The solvent extraction, volatilization, oxidation and hydrogeneration of coal with a complicated structure such as this is difficult because the reactive linkages are accessible only with difficulty. Inquisitive scientists have been attempting to understand the reactions of coal for centuries and in recent years, interest has been renewed because additional sources of energy and chemicals are needed. The conversion of coal to heat, and liquid or gaseous fuels is extremely complicated, and for simplicity the reaction of carbon (i.e., β-graphite) as shown in Figure 2 can be used to gain a better understanding of the process involved (3). Reactions 1 to 6 involve the direct reaction of coal, 7 to 9 involve the products of coal and are useful in converting coal to gaseous fuels, and reactions 10 and 11 are combinations of other simple reactions.

Figure 3 shows some of the products that could be derived from coal involving the destructive distillation, pyrolysis, oxidation, gasification or hydrogeneration and recovery of by-products. These products have generally been derived so far from the distillation of coal tar and light oils obtained as a by-product of the coking operation (4). However, it should be recognized that hydrogeneration and oxidation of coal also would produce similar chemicals and in some cases produce char as a by-product.

Environmental analysis of coal conversion facilities have so far only considered individual components mostly limiting it to the facility itself. Comprehensive environmental impact statements involving air, land, water and noise pollution, and the entire sequence of activities starting from the mining of coal to the distribution of products have not been developed. However, when one considers the size of these facilities the overall pollution problem is greatly simplified by having the pollutants generated at a single processing site rather than at multiple points of coal utilization. This does not reduce the complexity of the technical problems associated with the technology involved in controlling emissions from such facilities but it greatly reduces the overall economic impact.

In the following chapters, I will describe the status of coal conversion technology and potential health effects due to exposure to products from coal conversion plants very briefly. I will then describe the environmental rules and regulations applicable to coal conversion facilities in 14 states that have potential sites for the location of such facilities and costs of compliance with these regulations.

*Numbers in parentheses refer to bibliography at the end of this report.

II. STATUS OF COAL CONVERSION TECHNOLOGY

II-1: PRESENT ENERGY SITUATION:

Energy demand in the United States has been growing rapidly since the turn of the century - keeping pace with a growing population, increasing industrialization and greater affluence. For the past 10 years, the United States demand for energy has grown at an annual rate of between 4 and 5 percent. Today, the per capita energy consumption here is 8 times the average of the rest of the world. Through 1950, the United States was totally self-sufficient, meeting its energy demand with cheap and abundant coal and what appeared to be abundant and cheap domestic oil and natural gas. But through the 60's and 70's imports of crude oil and other petroleum products increased to an extent whereby 1973 imported petroleum jumped to 35% of the total domestic use. The United States now relies on oil for 46% of its energy and on natural gas for 32%, while coal, which provides only 17% of our consumption, represents about 3/4 of the total domestic resources.

As contrasted with the more than 800 years of coal supply available at the current rate of consumption, we have only about 10 years of proven oil and gas reserves in the 48 contiguous states. In 1974, we consumed 35% of the world's total production of fossil fuels, including 25% of the total coal production, 40% of the total liquid petroleum production, and 70% of the total natural gas production. At this rate of use of natural gas, the projected deficit between demand and supply is expected to be 34%, 60% and 90% by 1980, 1985 and 1990 respectively and production is projected to steadily decrease. Some experts have doubts if the United States will be able to attain energy self-sufficiency under the project independence program by 1985. Some argue that this may not be in the best interest of the consumers since there will have to be a net expenditure of non-revenue producing money to attain self-sufficiency. Nevertheless, President Ford has outlined a comprehensive energy program for achieving energy independence by 1985. One component of this proposed program is a national synthetic fuels commercialization program with an initial goal of producing the equivalent of 350,000 barrels of oil a day and the final goal of 1 million barrels of oil per day by 1985. The Synfuels Interagency Task Force report(5) projects that 25 plants with the following fuel mix would be required to meet this goal: Shale Oil (50,000 CBE*) 6; Syncrude (50,000 CBE) 2; high Btu gas (40,000 CBE) 7; and low Btu gas (25,000 CBE) 10.

The time required to build and operate a coal conversion facility is shown in Figure 4. According to this time schedule if clean fuels from coal are expected to make an impact on the availability of clean energy by 1985 then firm commitments to build such plants would have to be made in 1978 or 1979. Hopefully, by then the methodology for analyzing total environmental impact of these plants would be developed. This would be important to prevent any delays in the construction and operation of these plants if the energy independence goal has to be achieved.

*CBE - Crude Barrel Equivalent.

II-2: COAL CONVERSION TECHNOLOGY:

While processes and technology for converting coal to higher value products have been known for many years, commercial scale application in the United States has not come about mainly due to the economic competition from petroleum and natural gas. However, in countries with abundant coal supplies and restricted availability of petroleum and natural gas, coal conversion on a commercial scale has been utilized for the production of gasoline to nitrogenous fertilizers.

II-2.1; COMMERCIAL SCALE PROCESSES:

II-2.1.1; THE LURGI GASIFIER:

The best known commercially proven process that has been opted for by companies planning to build large scale coal gasification processes in the United States is the Lurgi processes. The gasifier used in the process was developed by the Lurgi Mineralotechnik GMBH of Frankfurt, West Germany. Sixteen commercial plants using such gasifiers and variety of processes for cleaning and upgrading the gas have been built worldwide during the last 30 years. This gasifier, shown in Figure 5, is a high-pressure (operating between 300-500 psig), fixed-bed, non-slagging, steam oxygen system producing a synthesis gas from coal. Its main drawback has been the agglomeration of caking coals within the gasifier thereby rendering it inoperable. This problem may be overcome by either pretreating the coal or converting a conventional non-slagging gasifier to a slagging gasifier that could operate at higher temperatures. So far, only lignite, sub-bituminous, and non-caking and weakly caking bituminous coals have been used wherever Lurgi gasifiers are a part of the coal conversion facility.

The raw gas from the gasifiers normally have a heat content of about 450 Btu/scf and this has to be upgraded through methanation if gas with a heat content of around 1,000 Btu/scf is desired. A simplified process flow diagram of a coal gasification facility producing pipeline quality gas using a Lurgi pressure gasifier is shown in Figure 6.

The best known coal conversion facility using 13 Lurgi gasifiers is located in Sasolburg, South Africa and has been successfully converting coal into a variety of products since 1954. The process flow diagram and the products produced at this plant are shown in Figure 7. The confidence shown by the operators of the Sasol plant was recently reflected in annoucing a 1.3 billion dollar expansion program projected for completion in 1981.

A detailed process flow diagram of a coal conversion facility, reproduced from reference 6, for the production of high heat content gas from low sulfur western coal is shown in Figure 8. The bold line indicates that flow of primary gasification process and the encircled number refers to the stream compositions shown in Table 1. The overall energy balance is shown in the table on the right hand side of Figure 8 and note that the product gas retains only about 56.2% of the total energy supplied to the facility. Rhombic shaped units represent intermediate products such as boiler feed water, uses such as electric power, and sources such as return water, for which distribution is not shown. Non-integral pollutant clean-up process are indicated by sloping rectangles. Inverted trapezoids at the bottom denote residual storage. Also, the composition of the streams

before and after treatment can be obtained from Table 1.

It should be noted that this process flow diagram represents the application of the best available technology and would in all cases meet the present pollution control requirements even if Illinois coal with 3.5-4% sulfur were used. As far as air pollution is concerned, I am only referring to specified pollutants and non-significant air quality deterioration provisions. The concentrations of other substances, i.e., non-specified pollutants, has not been mentioned due to the fact that theoretical calculations of the amount of such substances present in the gas streams is difficult, if not impossible.

II-2.1.2: THE KOPPERS-TOTZEK GASIFIER:

The Koppers-Totzek gasifiers, shown schematically in Figure 9, is an oxygen blown, atmospheric pressure, entrained suspension system used extensively for hydrogen production in the ammonia synthesis process (7). The gasifier can be operated on all types of coal and so far about 50 of them have been built throughout the world. Also, change over from solid to liquid feeds such as tars and heavy residual involves changing of the burners and a few other minor modifications. Other advantages claimed for the gasifier include its simple construction, low maintenance, high capacity and safe operation. The process can be instantly shut down with full power resumed in 1/2 hour.

Use of the Koppers-Totzek gasifier for pipeline gas production is limited by the fact that all of the methane must be manufactured in the methanator. Application of the gasifier for producing hydrogen from residual chars in other pipeline gasification processes can be considered.

The process flow diagram of a coal conversion facility using the Kopper-Totzek gasifier is shown in Figure 10 (6). The stream compositions if the process feed was Illinois coal with about 4.87% sulfur (2% moisture coal basis) are shown in Table 2. The same symbols as described for the process flow diagram using the Lurgi gasifier apply here and compliance with all the applicable regulations can be expected.

II-2.1.3: THE WINKLER GASIFIER:

The Winkler gasifier shown in Figure 11 is a commercially proven fluidized bed, atmospheric pressure system for the production of synthesis or low heat content gas from a variety of coals generally not including caking coals. The synthesis gas can be used for the production of methanol, ammonia, as well as liquid hydrocarbons, via the Fischer-Tropsch synthesis, or can be shifted and methanated to produce pipeline quality gas.

Let me now describe some of the processes under development here in the United States - those that have operated at the pilot plant scale.

IV-2.2: PROCESSES UNDER DEVELOPMENT FOR COMMERCIAL SCALE INSTALLATIONS:

IV-2.2.1: IGT'S HYGAS PROCESS:

The Hygas process, like most advanced coal gasification processes utilizes the hydrogasification reaction to produce methane. Two-thirds of this product gas is produced directly in the prime reaction. Maximizing direct methane formation improves the process thermal efficiency and decreases the extent

of catalytic methanation. The gasification reactor operates at 1000° to 1400°F, gasifies the most reactive portions of the feed and forms large amounts of methane directly; the second stage, operating at up to 1850°F, gasifies the less reactive portions. Residual carbon from this stage is withdrawn and used to generate the hydrogen needed for hydrogasification. A mixture of steam, hydrogen, carbon monoxide, methane, and other reaction products serve as fluidizing gases. The steam also aids in controlling gasifier reaction temperatures.

Three alternatives have been studied on ways to generate hydrogen from the carbon residue. These include:

Electrothermal gasification - Heat required for the steam-carbon reaction is supplied by passing an electric current through the bed of fluidized solids.

Steam-oxygen gasification - Partial combustion of the residual char with oxygen supplies the heat required for the steam carbon reaction.

Steam-iron reaction - Hot reduced iron reacts with steam to form a hydrogen-rich gas. The iron is reduced in a separate vessel by gas generated from char by partial combustion with air and steam.

A schematic flow diagram for a commercial scale hygas process producing 250 million cubic feet of pipeline quality gas per day using no. 6 Illinois coal as feed is shown in Figure 12. Energy balances are shown on the right hand side of Figure 12 and Table 3 gives all the stream compositions. All the possible atmospheric, water and land pollution streams are also shown.

A pilot plant for the demonstration of the hygas process is currently being operated by IGT in Chicago. They report that several continuous runs of 100 hours of more duration have been made. Procon, the builder of the pilot plan for IGT has completed a preliminary design of a demonstration plant to convert 5000 tons per day of bituminous coal to 50 million cubic feet per day of pipeline quality gas. A full scale commercial plant producing 250 million cubic feet per day of pipeline quality gas may be built by 1981.

II-2.2.2: THE SYNTHANE PROCESS:

The Bureau of Mines synthane processes has been under development since 1963 at the Pittsburgh Energy Research Center in Bruceton, Pennsylvania. This research facility has now been taken over by the Energy Research and Development Administration (ERDA). A simple process flow diagram is shown in Figure 13. Dried ground coal is fed into a fluidized pre-treatment reactor by use of pressurized lock hoppers and appropriate feeders. In the pre-treatment reactor the caking properties of the coal are destroyed by treating with steam and oxygen. From the pre-treater the coal, along with any volatile matter and excess unreacted steam, is fed to the top of the fluid bed gasifier. Here the coal is reacted with steam and oxygen at about 1700°F to form methane, carbon monoxide, hydrogen and carbon dioxide. However, not all the carbon reacts and about 30% of the heat content in the coal is removed from the bottom of the gasifier in the form of char. This char could be used as boiler fuel or further gasified if the high sulfur content does not allow it to be used as boiler fuel. The gas leaving the top of the gasifier contains entrained dust plus tar formed in the low temperature pre-treatment reactor. These impurities have to be removed prior to upgrading the gas to pipeline quality.

Again, it is expected that a commercial scale facility would use best available technology to control all the pollutants.

At this point, let me just mention a few other processes that have the potential of commercial scale operation for the production of high heat content gas and those that treat coal prior to burning or pre-treatment processes.

A) CO_2 acceptor process developed by the Consolidation Coal Company.
B) Bi-gas process developed by Bituminous Coal Research, Inc.
C) The coed processes or the char-oil-energy-development process developed by FMC Corporation under sponsorship of the Office of Coal Research.
D) Solvent refined coal or SRC process originally developed by Spencer Chemical Company. A 50 ton/day pilot plant using this process was built by Pittsburgh and Midway Coal Mining Company, the design for which was provided by Stearn-Roger Corporation, in Tacoma, Washington. In this process a highly aromatic solvent, a by-product of the process itself, dissolves coal at 850°F and 1000 psi. The solution is filtered to remove coal ash and unreacted carbon, and the solvent removed by distillation. The solid coal product is very low in ash and sulfur, high in heat content and is normally referred to as synthetic coal.

About 25 more different processes are being developed at present and I will not mention them here. For a listing of various processes under development, see references 8, 9 and 10.

III. POTENTIAL HEALTH EFFECTS DUE TO EXPOSURE TO PRODUCTS FROM COAL CONVERSION FACILITIES:

III-1: GENERAL:

Table 4 shows the principal components, disposition and treatment of waste streams that could be expected to be generated by a coal conversion facility. It should be noted that almost all of these waste streams are subjected to treatment prior to the final disposition. Some of the process streams identified in Figures 8, 10, and 12 and Tables 1, 2 and 3 can be picked out from this table for a better understanding of the type of treatment and final disposition that may be necessary.

III-2: HEALTH EFFECTS OF ENVIRONMENTAL CONTAMINATION:

The potential adverse health effects of exposure to both the waste materials and products generated from a coal conversion facility have not been thoroughly investigated in the United States probably due to the non-existence of commercial scale facilities. Such facilities are in operation elsewhere in the world but very little, if not none, data is available to show the nature of the health consequences. Such studies if conducted at any of the numerous pilot plants presently in operation have not yet been reported in the general literature.

The literature in the field of experimental carcinogenesis of chemicals derived from coal and of synthetic products similar in structure is voluminous (11). The health consequences of by-products of coal combustion, such as sulfur dioxide, have been well established (12) and (13).

Because of the large quantities of raw materials that any economically viable coal conversion facility consumes, in the order of 20,000 to 30,000

tons per day of coal, the potential for environmental discharge of considerable quantities of hazardous materials exists. These materials include the trace elements shown in Table 5 and various quantities of hydrogen cyanide, phenols, cresols, carbonyl and hydrogen sulfides, ammonia, mercaptans, thiocyanides, aniline, arsenic and various other polycylic hydrocarbons (14) and (15). Of the trace elements shown in Table 5, arsenic, beryllium, mercury, lead and vanadium have all been identified as materials with properties toxic to man.

Heavy oil and tars constitute a large quantity of the total amount of products generated in a coal hydrogenation plant. Product liquids derived from these may have higher aromatic contents than naturally occurring hydrocarbons with the same boiling point characteristics. Since many of these aromatic or cyclic hydrocarbons have been identified as being hazardous to human health, the potential of these liquids for causing adverse effects on the central nervous system, liver, kidney, and skin, as well as their possible carcinogenicity, has to be determined.

Another undefined area is the effect exerted by the large amounts of carbon dioxide (around 16,000 tons per day from a commercial facility) emitted from gasification facilities (14).

Wastewaters from coal conversion process originate from (16):

A) moisture in the coal,
B) water of constitution or decomposition,
C) water added for stoichriometric process requirements, and
D) water introduced for by-product recovery or gas scrubbing.

Since these process waters come into direct contact with the various contaminants in coal, they may be a source of pollution. These types of water pollution sources could be greatly minimized since the potential exists for recycling them to extinction because coal conversion plants are net consumers of water. Table 6 from reference 16 compares the concentrations of various undesirable contaminants in waste ammonia liquor from a by-product coke plant and by-product waters from the Bureau of Mines synthane process. The presence of cyanides, phenols and other cyclic compounds in waste waters could pose a health hazard if the receiving bodies constitute drinking water supplies for the surrounding communities.

III-3: OCCUPATIONAL HEALTH IMPLICATIONS:

The occupational health implications of coal conversion facilities may pose a few problems. At the coal hydrogenation plant in West Virginia that was started in 1952 by Union Carbide pamphlets were issued and lectures conducted by the medical department to explain the hazards of skin cancer and preventive measures to reduce the incidence of prolonged exposure and/or contact of the skin with the chemicals produced in the hydrogenation process (17). It was eventually reported that out of group of 350 workers at this plant, over a five year period, ten men developed skin cancer. One reason may be the high level of an identified carcinogen namely, benzo(a) pyrene, which was measured in concentrations often in excess of 50 micrograms per 1000 cubic meters, with a peak value of 18,000 (18). The typical urban levels of this substance has been found to be about 3-6 micrograms per 1000 cubic meters (14).

Health statistics on occupational groups in other coal conversion operations,

such as coke ovens and coal-tar processing, have shown significantly higher lung cancer rates relative to groups without such occupational exposure (19). However, as mentioned earlier, coke production is one of the dirtiest operations and no comparison can be made with the well controlled modern coal conversion facility.

These facilities are expected to use the best available technology, and in some cases, innovative technology, to assure that environmental and health impacts are reduced to a minimum. Such is the case with a few plants that have gone past the design stage.

IV. ENVIRONMENTAL RULES AND REGULATIONS APPLICABLE TO COAL CONVERSION FACILITIES:

IV-1: GENERAL:

The regulatory framework established under Federal, state and local pollution control laws in general have set up agencies for implementing the specific requirements contained in such laws. At the Federal level, the Clean Air Act, as amended, 42 U.S.C. §1857 et seq (1970), the Federal Water Pollution Control Act, as amended, 33 U.S.C. §1151 et seq (1970), the Solid Waste Disposal Act, as amended, 42 U.S.C. §3251 et seq (1970), the Noise Pollution and Abatement Act of 1970, 42 U.S.C. §1858 et seq (1970), the Energy Supply and Environmental Coordination Act of 1974, 15 U.S.C. §791 et seq (1974), and the National Environmental Policy Act of 1969, 42 U.S.C. §4321 et seq along with the guidelines of the Council on Environmental Quality, dated April 23, 1971 (36 FR 7724) established the regulatory framework, as it exists today.[1] The Federal Water Pollution Control Act Amendments of 1972 requires each state to establish and obtain approval of receiving water quality standards that protect fish life and recreational use of its water ways. All the fifty states and most of the territories have at present some type of legislative requirements that grants an appropriate agency the authority to establish and enforce environmental regulations. There are literally hundreds of local agencies and departments that also have local jurisdiction for establishing and enforcing environmental regulations. In general, regulations establishing specific pollution control requirements at the state levels are required to be as strict or stricter than similar Federal requirements, and local requirements are to be as strict or stricter than state requirements for state and local agencies to obtain approval to enforce such requirements.

Instead of summarizing regulations established by all the fifty states the following section contains the present environmental regulations and siting requirements as applicable to coal conversion facilities in fourteen different states. Also, included under Section IV-2 are the regulations established at the Federal level that would be quite uniform throughout these states. Section IV-3 deals with the environmental impact statement which may, one way or the other, be required prior to the construction of a coal conversion plant. Sections IV-4 to IV-17 contain specific requirements established by

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[1] Regulations developed in compliance with the Occupational Safety and Health Act of 1970, 29 U.S.C. 655 et seq will not be discussed in this paper since they deal with in-plant atmosphere. The Occupational Safety and Health Administration under the Department of Labor administers these regulations at the Federal level.

the 14 states mentioned previously. These are the states that have large coal reserves and potential sites for the location of coal conversion facilities.[1] It should be pointed out that Federal requirements, e.g., the development of implementation plans to implement that national ambient air quality standards required to be promulgated under Section 109 of the Clean Air Act, are uniform for all the fifty states. The states may have chosen different ways of compliance but the goals established at the Federal level are quite similar. Appendix B contains tables listing specific pollution control regulations promulgated by the 14 states.

IV-2: FEDERAL REGULATIONS:

IV-2.1: NATIONAL EMISSION STANDARDS FOR HAZARDOUS AIR POLLUTANTS (NESHAPS):

Under Section 112 of the U.S. Clean Air Act, as amended, the Administrator of the U.S. EPA has to promulgate emission standards for hazardous air pollutants applicable to specific sources of such pollutants. So far, no emission standards of hazardous pollutants specifically applicable to coal conversion facilities have been promulgated. However, if this does become necessary, e.g., for organic carcinogens, etc., the U.S. EPA may promulgate such emission standards. Emission standards for asbestos, beryllium and mercury were promulgated by the U.S. EPA on April 6, 1973, but these are presently not applicable to coal conversion facilities (38 FR 88200; 39 FR 15396 and 23543; and 40 FR 48292). In addition to these three pollutants a NESHAP for vinyl chloride was proposed on December 24, 1975 and the regulations have not yet been finalized (40 FR 59477 and 59532).

IV-2.2: NEW SOURCE PERFORMANCE STANDARDS (NSPS):

Apart from the new source performance standards promulgated by the U.S. EPA for fuel combustion emission sources and incorporated into various state's rules and regulations, if a coal preparation plant is a part of a coal conversion facility, new source performance standards applicable such a plant were promulgated on January 15, 1976 (41 FR 2232). All new plants constructed or modified as of October 24, 1974 are required to comply with particulate emission standards applicable to: a) thermal dryers - 0.031 gr/dscf (grains per dry standard cubic feet) and 20% opacity; b) pneumatic coal-cleaning equipment - 0.018 gr/dscf and 10% opacity; c) coal processing and conveying equipment - 20% opacity; d) coal storage systems - 20% opacity; and, e) transfer and loading systems - 20% opacity. In cases where state's emission standards are stricter than the ones mentioned above, then these plants would have to comply with such state's standards. The NSPS applicable to coal conversion facilities and incorporated in state's regulations will be discussed under individual state's regulations in the following sections.

IV-2.3: PREVENTION OF SIGNIFICANT AIR QUALITY DETERIORATION:

In response to a preliminary injunction issued by the district court for the District of Columbia Circuit, the Administrator of U.S. EPA disapproved all state implementation plans insofar as they failed to provide for the prevention of significant deterioration of existing air quality. This injunction also required the Administrator to promulgate regulations which would prevent significant deterioration of air quality. Subsequently, regulations were promulgated on December 5, 1974 (39 FR 42510) and Section 52.21(d) of these standards requires the Administrator of U.S. EPA to review

[1]See Keystone's Map of the Coal Fields of the United States; page 66.

19 new sources and to determine the effect of these sources on existing air quality. These 19 new sources include coal cleaning plants and fuel conversion plants (nos. (ii) and (XViii) respectively). Under Section 52.21(c), allowable increases in concentration over the base line air quality are specified for particulates and sulfur dioxide for areas designated Class I, II, and III. These concentrations are shown in Table 7. States were required, under this same section, to notify the Administrator of the U.S. EPA by June 1, 1975 if any areas which are presently above the national ambient air quality standards, like the St. Louis (Illinois) major metropolitan area, should be exempted from being designated. However, subsequent amendments (40 FR 25006) allow the states to notify the Administrator at any time if any "counties or functionally equivalent areas that pervasively exceeded any national ambient air quality standard during 1974" for sulfur dioxide and total suspended particulates. Also, the incremental deterioration provisions are not applicable to attainment (AQMA) areas, since progress in reducing emissions is needed.

It should be noted that the entire significant air quality deterioration provisions are currently under debate in the U.S. Congress as it prepares its amendments to the Clean Air Act. Therefore, no definitive conclusions can be made of the impact of these regulations, which may be extensively amended in the near future, on the location of coal conversion facilities. However, the site selection procedure should include the measurement of background concentrations of these two pollutants to determine the allowable increment.

Simple calculations made for a full scale coal conversion facility at the New Athens, Illinois site[1], after taking into account Illinois Power Company's 1905 Mw Baldwin power plant located about 13 miles south of this site, had shown that the increments allowed for sulfur dioxide in Class II areas would not be exceeded. Implementation of these provisions through preconstruction review of stationary sources would require more sophisticated modeling and air quality analysis. It is, therefore, suggested that the selection of a specific site be made with emphasis placed on the present or any future significant air quality deterioration provisions. It should be noted that due to the lack of precise particulate and sulfur dioxide emission rates from the coal conversion plant to be located at New Athens the emission rates were assumed to be similar to the WESCO facility (20). Total sulfur dioxide emissions from this plant were assumed to be 7.46 tons per day compared to the Baldwin power plant's average daily emissions of 900 tons with a maximum of 1200 tons/day of sulfur dioxide.

IV-2.4: REVIEW OF INDIRECT SOURCES:

Regulations establishing a permit issuing system for complex or indirect sources of air pollution had been published by the U.S. EPA and proposed by the Illinois EPA. The Federal EPA complex (indirect) source regulations appear in the Federal Register Vol. 39, No. 138, July 9, 1974 and they were to be applicable to those projects beginning construction or modification following January 1, 1976.

The intent of these regulations were to assure that developments leading to

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[1]This is site in Illinois where the first coal conversion demonstration facility funded by ERDA is going to be built by COALCON.

large increases in traffic generation would not create sufficiently high concentrations of traffic-related pollutants to cause a violation of the national ambient air quality standards. The indirect source regulations were to affect projects such as streets and highways; parking facilities; retail, commercial and industrial facilities, recreation, amusement, sports and entertainment facilities, airports; apartment and condominum buildings; office and government buildings; and education facilities.

At present, the proposed Federal regulations have been indefinitely suspended but may in the near future be revived again. Some of the states do have indirect source regulations that require the design and operation of sources so as not to cause violations of the air quality standards. Due consideration should be given to such sources, if they are a part of a coal conversion facility, and their impact on air quality prior to selecting a specific site for the location of a coal conversion facility.

IV-2.5: AIR QUALITY MAINTENANCE AREAS (AQMAs):

On January 31, 1973, the U.S. Court of Appeals for the District of Columbia Circuit decided the case of "Natural Resources Defense Council, Inc., et al vs. Environmental Protection Agency" (Civil Action No. 72-1522) and seven other related cases. The Court's order required the Administrator of the U.S. EPA to review within 30 days the air quality maintenance provisions of all state implementation plans. The Administrator was directed to disapprove plans which did not include measures necessary to insure the maintenance of the primary national ambient air quality standards after May 31, 1975. Completion of such a review was reported on March 8, 1973 (38 FR 6279) and no state plan was found to contain adequate growth projections and air quality maintenance measures. Accordingly, all state plans were disapproved and maintenance plans were required to be submitted by the states by April 15, 1973. However, the final regulations pertaining to the analysis and development of plans for AQMAs were promulgated on May 3, 1976 (41 FR 18382). The AQMA names, pollutants and geographical compositions for all the states were published on three different dates as follows: April 29, 1975 (40 FR 18726); June 2, 1975 (40 FR 23746); and, September 9, 1975 (40 FR 41942). AQMA designations for various areas in the 14 states that have potential sites for the location of coal conversion facilities are shown in Table 8 (21).

Volume 1 of the AQMA guidelines (22) specifies the procedure to be used in designating AQMAs. All Standard Metropolitan Statistical Areas (SMSAs) were investigated along with national resource and industrial development areas in the 14 states listed in Table 8. Natural resource development areas are non-SMSAs which have the potential to exceed any national ambient air quality standards due to existing or planned development of natural resources. Some areas containing coal deposits and potential coal conversion plants are included under natural resource development areas.

Selection of sites for the location of coal conversion facilities should take into consideration the AQMA status of various areas and plans, either developed or in the process of being developed by U.S. EPA or the individual states, for air quality maintenance. Some areas may have to restrict the location of large individual complexes and sources of specified pollutants if such sources would not allow for the maintenance of the air quality standards. Both direct and indirect sources that may be associated with industrial complexes, such as coal conversion facilities, would have to be taken into consideration.

Guidelines for the review of stationary sources have been published by the U.S. EPA (23). If a coal conversion plant is to be located in an AQMA these guidelines should be consulted along with the state's procedures, to determine the sources impact on air quality. The purpose of the review would be to: a) provide assurance that the facility will not cause ambient air quality standards to be violated; b) to bring to light procedures, source designs and siting alternatives which would further reduce the impact of air pollutants, especially on the more susceptible individuals and crops; and, c) to provide sufficient knowledge of potential air pollution problems resulting from the new source's existence or subsequent growth and that such problems can be adequately dealt with.

IV-2.5: STATE AND AREAWIDE WATER QUALITY MANAGEMENT PROGRAMS:

The purpose of the state and areawide water quality management programs is the development and implementation of State Water Quality Management Plans so that the longer range goals of the Federal Water Pollution Control Act Amendments of 1972 (P.L. 92-500) can be met. To achieve these goals, it has been found to be necessary to develop a water quality management process at the state and local level that assures continuous planning for and implementation of pollution control measures. Draft guidelines for state and areawide water quality management program development have been issued by the U.S. EPA (24). The management plans are required to be directed towards meeting the following two principal mandates of the Federal Water Pollution Control Act Amendments of 1972: a) the determination of effluent limitations needed to meet applicable water quality standards including the requirement to at least maintain existing water quality (Section 303); and b) development of state and areawide management programs to implement abatement measures for all pollutant sources (Section 208).

The continuous planning process has to take into consideration point sources such as discharges from municipal treatment plants, combined sewer overflows, and industrial waste effluents. Non-point sources such as accummulated pollutants in the streams, diffuse run-off, seepage and percolation contributing to the degradation of the quality of surface and ground waters also have to be taken into consideration.

Discharges such as cooling tower blowdowns, sewage, etc. from coal conversion plants can be classified as point sources and other related activities such as coal mining and site preparation may generate non-point sources of water pollution. Therefore, the consideration given to these two categories of sources in the individual state's planning program should be evaluated prior to selecting a specific site.

The main objective for including industrial wastewater point sources in the planning process is to provide the most efficient approach for serving the present and future industrial waste water treatment needs of the area. Under the plan, direct discharges of industrial wastes to receiving water would have to comply, at a minimum, with the provisions of the pertinent effluent limitations and new source performance standards. Higher treatment levels or internal waste load reductions may be required where waste load allocations dictate more stringent restrictions. Application of higher treatment levels to meet water quality standards may be mitigated through restricting the location of future industrial development to areas where receiving water can more readily assimilate the treated waste water. Control of industrial

location incorporated into the land use plan with recognition of other constraints such as air quality control may also be included as a part of the planning process.

One area that should be investigated more thoroughly is the effect of withdrawal of large quantities of water from stream with flows barely adequate to sustain present waste water loadings. Such withdrawal may require downstream discharges to comply with more stringent effluent standards. The planning process is required to identify such problems and water bodies that will not accommodate any increases in pollutant discharges.

All the states are presently in the process of developing water quality management plans and by November 1, 1978 are required to submit plans to the U.S. EPA for approval. Also, by that date, governors are required to designate management agencies to implement approved water quality management plans.

IV-3: ENVIRONMENTAL IMPACT STATEMENTS (EIS):

Pursuant to the National Environmental Policy Act (P.L. 91-190, 42 U.S.C. et seq) and executive order 11514 (35 FR 4247) all Federal departments, agencies, and establishments are required to prepare environmental impact statements in connection with their proposals for legislation and other major Federal actions significantly affecting the quality of the human environment. Final guidelines for the preparation of such environmental impact statements were published by the Council on Environmental Quality on August 1, 1973 (38 FR 20550). The general definition of what constitutes "major Federal actions significantly affecting the quality of the human environment" is given in these guidelines under Section 1500.6. A number of Federal courts have interpreted NEPA requirements and have identified the basis of when an environmental impact statement is required, who has to prepare it and the general format and contents.

On November 30, 1971, the Federal Power Commission issued "Statement of General Policy to Implement Procedures for Compliance with the National Environmental Policy Act of 1969" (36 FR 22738). Under this policy FPC will have to prepare detailed environmental impact statements when the regulatory action taken by FPC under the Federal Power Act and Natural Gas Act will have a significant environmental impact.

These laws, guidelines, policy statements and various other rules and regulations issued by Federal agencies requires the preparation of a detailed environmental impact statement prior to the construction of just about any coal conversion facility. The assumption here is that construction of coal conversion facilities would involve some sort of a major action by any one of more of than 38 Federal agencies, commissions, etc. that may have to grant approval for some aspect of the facility.

The basic steps involved in the assessment of environmental impact and the preparation of an EIS are (25): a) the identification of anticipated types of impacts including any adverse environmental effects that cannot be avoided; b) the estimation of the extent and magnitude of anticipated impacts; c) the interpretation of impact significance; d) alternatives to the proposed action; e) relationship between local short-term uses of the environment and long-term productivity; and, f) identification of any irretrievable commitment of resources.

Table 9 is an outline of an EIS for a coal conversion facility. This outline covers essentially all the Federal requirements resulting from NEPA. Please note that this outline only covers the coal conversion facility and does not include the environmental impact of the associated coal mining activity or water withdrawal from either surface or underground water bodies. Preparation of EISs for the coal mining activity or the water withdrawal could be combined with the EIS for the coal conversion facility or done separately. However, a document describing the overall impact may be necessary and more appropriate.

One of the first EISs that was ever put together is included as Appendix A. I thought it would be interesting for the reader to note the basic ecological insight - all things are connected - embodied in this EIS.

IV-4: ILLINOIS:

Illinois, at present, has great potential for attracting commercial scale coal conversion facilities because of the abundant bituminous coal reserves and adequate supplies of water. Research conducted by the Illinois Institute for Environmental Quality has identified numerous sites in Illinois where coal conversions plants can be built (26). The New Athens site, which is about 30-40 miles southeast of St. Louis, has been chosen by ERDA for the construction of the first coal-to-clean-fuels demonstration facility. Two other demonstration coal conversion plants, one in Perry County in Southern Illinois and the other in Pekin, Illinois, are also going to be built. The plants at New Athens and in Perry County are expected to prove technical and economic viability and environmental acceptability of coal conversion technology. These plants are partially funded by ERDA under the synthetic fuels commercialization program.

Due to the fact that Illinois may emerge as a leader as far as the number of coal conversion facilities are concerned, I am going to describe its environmental regulations in a little more detail than other states.

IV-4.1: WITHDRAWAL OF WATER FOR COAL CONVERSION FACILITIES:

Based upon the design data and the type of process involved at present, a coal conversion plant producing up to 250 million cubic feet per day of high heat content gas has a consumptive water usage of about 5,500 gpm or 8 million gpd or 12.4 cfs. Site selection for building such plants should take into consideration this large consumptive water use, historic low flow periods of a given water source and the excess water flows over the life span of the facility.

There are at present no specific regulations in Illinois limiting the withdrawal of surface or ground water. The Illinois State Water Survey has been responsible for the collection and interpretation of information on the water resources of Illinois and has conducted surveys to determine the availability of water resources for use in siting coal conversion plants (27).

Regarding the use of ground water, the Illinois Supreme Court has adopted the English common law rule for cases involving percolating ground water. In essence, the rule states that the owner of the land owns all the percolating water underlying his land. For streams within the state, the Illinois Supreme Court has subscribed to the doctrine of riparian rights

with respect to the reasonable use of water in natural water courses.

The possibility of using effluent water from sewage treatment facilities has also been considered in some cases. However, no definitive information is presently available regarding any problems that may be encountered due to the use of treated waste water. The Illinois State Water Survey has identified 14 municipal sewage treatment plants in the State of Illinois that have a 7-day low-flow adequate enough to meet the water demands, on a year round basis, of a coal conversion plant (27). However, the technical and economic feasibility of using such a source have yet to be established.

IV-4.2: WATER QUALITY AND SEWAGE DISPOSAL STANDARDS:[1]

IV-4.2.1: WATER QUALITY STANDARDS:

Legal limits - water quality standards applicable to coal conversion facilities are established under Chapter 3, Part II, of the Illinois Pollution Control Board's Rules and Regulations. These water quality standards are in addition to the effluent standards defined in Chapter 3, Part IV of the Rules and Regulations and are to be met in the receiving waters outside the mixing zone. Such mixing zones are not to exceed the area of a circle with a radius of 600 feet. The water quality standards take into account additional factors not included in the effluent standards, such as dissolved oxygen level in the receiving stream, unnatural sludge or bottom deposits, and unnatural color of turbidity.

With respect to temperature, there generally cannot be any abnormal temperature changes that may adversely effect aquatic life unless caused by natural conditions. The maximum temperature rise above natural temperatures cannot exceed 5°F. In addition, the temperature outside the mixing zone cannot exceed a maximum of 60°F during December through March and 90°F during April through October more than one percent of the hours in any 12 month period. The water temperature cannot exceed the maximums by more than 3°F at any time.

Any discharge from coal preparation plant is considered to be process water and falls under Chapter 3 of the Pollution Control Board's Rules and Regulations.

Some of the pertinent general standards specified under Rule 203 are as follows:

A. Freedom from unnatural sludge or bottom deposits, floating debris, visible oil, odor, unnatural plant or algal growth, unnatural color or turbidity, or matter in concentrations of combinations toxic or harmful to human, animal, plant or aquatic life of other than natural origin.
B. pH to be within the range of 6.5 to 9.0 except for natural causes.
C. Phosphorus: Phosphorus as P not to exceed 0.05 mg/l in any reservoir or lake, or in any stream at the point where it enters any reservoir or lake.
D. Dissolved oxygen cannot be less than 6.0 mg/l during at least

[1]Illinois Control Board Rules and Regulations, Chapter 3, Water Pollution; Adopted March 7, 1972; Amended June 28, 1973; July 19, 1973, January 31, 1974; March 6, 1975, Effective March 20, 1975; August 14, 1975.

16 hours of any 24-hour period, nor less than 5.0 mg/l at any time.

E. Based on a minimum of five samples taken over not more than a 30-day period, fecal coliforms shall not exceed a geometric mean of 200 per 100 ml.

F. Any substance toxic to aquatic life shall not exceed one-tenth of the 48-mile hour median tolerance limit (48-hour tlm) for native fish or essential fish food organisms.

In addition, if the process water use in the commercial plant concentrates materials contained in the intake water, the two constituents which may present a water quality problem are iron and total dissolved solids. The effluent standard for total dissolved solids would probably be 3,500 mg/l assuming that pollution abatement practices are employed at the plant. However, the water quality standard for total dissolved solids, after mixing with the receiving stream, is 1,000 mg/l. The ability to meet this standard during low river flow conditions has to be considered when siting the plant to use surface waters.

Permits required - The types of permits that may be required, depending again on the type of process used and the site selected, are as follows:

A. A construction permit is required from IEPA prior to the start of construction on any water pollution control, sewage collection, or treatment facility.

B. A NPDES permit is required from the U.S. EPA prior to the start of any new discharge. It is anticipated that the State of Illinois will soon have this permitting authority.

C. A construction permit is required from the U.S. Army Corps of Engineers for any intake or discharge structure that may be built in navigable waters.

D. A permit from the Illinois Department of Transportation, Division of Water Resources, is required for construction of water intake and discharge flumes.

E. A permit is required from the Illinois Department of Mines and Minerals for the drilling of water well.

IV-4.2.2: SEWERAGE DISPOSAL STANDARDS:

Septic systems - Septic systems which provide primary treatment need to be followed by secondary treatment plus chlorination prior to surface discharge. Septic tank subsurface seepage field systems which are designed not to have a surface discharge still require a construction permit from the IEPA. The Department of Public Health has jurisdiction over septic systems serving less than 15 population equivalent (PE) but does not require a permit. The IEPA requires a permit for systems serving 15 PE or greater.

Discharge requirements - Currently, the discharge standards are defined by the U.S. EPA under the NPDES permit program. This would limit the discharge to 30 mg/l or BOD, 30 mg/l of suspended solids, and 200 fecal coliforms per 100 ml. Permitting requirements for sanitary treatment facilities are as follows:

A. A permit to construct the treatment facility is required from the IEPA. Sewage treatment plant design shall be in accordance with the design criteria contained in Water Pollution Control Technical Policy 20-24.

B. An NPDES permit is also required from the U.S. EPA. However, it is anticipated this permit will soon be issued by Illinois.

IV-4.3: AIR POLLUTION CONTROL REQUIREMENTS:

IV-4.3.1: POLLUTION CONTROL BOARD'S RULES AND REGULATIONS GOVERNING THE CONTROL OF AIR POLLUTION:

Under the present rules and regulations adopted by the Illinois Pollution Control Board on April 14, 1972, construction and operating permit requirements for sources of five different air pollutants are specified in Part I. Emission limitations for these five different pollutants are specified in Part II. Parts III, IV, V, VI, VII, and VIII specify rules and regulations dealing with ambient air quality standards, episodes, open burning, asbestos, and mobile sources respectively. Of direct concern to a coal conversion plant to be located in the State of Illinois would be the following rules and regulations. It should be noted that when these rules and regulations were adopted no direct testimony was presented to evaluate the impact on air quality of a large coal conversion facility and also no thought was given to the possibility of establishing emission limitations specifically applicable to such facility. If in the future a need for specific regulations does arise, the presently applicable general emission limitations may be revised to an extent necessary to protect public health and welfare.

A. Illinois Environmental Protection Act:
Illinois Revised Statute, Chapter III 1/2, §1001-1051 (1970 as amended). Section 9(a) of this Act prohibits any person from emitting any contaminant into the environment so as to cause air pollution. Air pollution is defined in Section 3 of the Act as the presence of any contaminant in the atmosphere that would be injurious to human, plant or animal life, to health, or to property, or to unreasonably interfere with the enjoyment of life or property. This Section 9(a), therefore, would not allow emissions of any type of substance into the atmosphere in quantities that would generate complaints from people living around the plant. Other requirements under Section 9 are included, in more detail, in the rules and regulations.

B. Illinois Pollution Control Board's Chapter II, Air Pollution Control Regulations:
   a) Part I: General Provisions
      - Rule 103(a) of Part I requires all new sources or any modified existing source to obtain a construction permit before construction of such a source is commenced.
      - Rule 103(b) requires an operating permit before a source is put into operation unless the initial operation is part of a testing program which was previously approved as a part of the construction permit. Other rules under Part I specify certain conditions and information required for the evaluation of a construction or an operating permit. The construction and operating permits required to be submitted are shown in Table 10.

   b) Part II. Emission Standards and Limitations for Stationary Sources: This part contains a comprehensive set of emission limitations, some of them applicable generally to all non-exempted sources of particulates, sulfur dioxide, hydrocarbons, carbon monoxide and oxides of nitrogen. In certain cases, specific processes and plants are required to comply with emission limitations solely developed for such processes and plants. As mentioned earlier, since coal conversion was not given

any special treatment, the general provisions specified in this part are applicable. However, the power or steam generation plant associated with any coal conversion facility would have to comply with the emission limitations specifically applicable to fuel combustion emission sources. A list of rules and regulations generally applicable to a coal conversion facility is shown in Table 11. It should be noted that these emission limitations are compatible with the U.S. Clean Air Act, as amended, applicable to new fuel combustion emission sources (or any one of the first five categories only). Changes in the present Illinois emission regulations to reflect the emission standards contemplated to be promulgated by the U.S. EPA applicable to new sources, could be forthcoming.

c) Part IV: Episodes:
Sources of greater than 100 tons per year of any specified air contaminant would have to file an episode action plan as specified in this Part IV of the rules and regulations.

IV-414: NOISE POLLUTION CONTROL REQUIREMENTS:

Part II of Chapter 8 of the Illinois Pollution Control Board Rules and Regulations contains limitations on the sounds emitted from the facility, to nearby residential property in terms of allowable octave band sound pressure levels in decibels. The allowable limits are shown in Table 12. The regulations apply to the operation of the facility and do not apply to construction noises.

There are no construction or operating permits required for the noise regulations, and no future permitting requirements are currently being planned or considered for facilities of this type.

IV-4.5: LAND POLLUTION CONTROL REQUIREMENTS:

On July 27, 1973, the Illinois Pollution Control Board adopted solid waste rules and regulations relating to the design, development, and operation of solid waste management sites. Since waste generated from a coal conversion facility can be expected to be non-putrescible in nature, the requirement that land fill waste be covered daily could be waived.

The use of coal mine as a repository for solid waste generated from the facility (ash and sludge) is considered proper and will not be in violation of any existing or proposed solid waste regulations. However, contract waste disposal sites are generally available in the areas which have been identified as potential sites for the location of coal conversion facilities.

Since Appendix B contains all the specific pollution control standards, in the following Sections IV-5 to IV-17, only the general water pollution control requirements in each state will be discussed. It should be noted that air pollution control requirements in some cases may be a more important factor that water pollution control requirements in the selection of a site. Potential sites located in areas where ambient air quality standards attainment problems have been identified may not be suitable from an air pollution point of view. Emission standards in such areas may have to be strengthened, compliance with which may impose additional cost penalties. Comparison of specific emission standards, present ambient air quality, future growth of industrial activity in the area and associated increase in pollutant emissions, and specific air quality goals for each site should be

compared prior to the selection of a specific site.

IV-5: ARIZONA:

IV05.1: WATER POLLUTION CONTROL:

Arizona water quality standards were established pursuant to Section 36-1857 of the Arizona Water Pollution Control Law.[1] Section 2-3 of the State Department of Health contains rules and regulations for sewerage systems and waste treatment works. These rules and regulations do not specify any water quality standards or effluent standards in terms of allowable concentrations or quantitative limitations. Design criteria for community and industrial waste collection and treatment facilities are contained in the Arizona State Department of Health Engineering Bulletin No. 11. Bulletin No. 12 specifies design criteria for septic tanks and other similar sewerage facilities.

IV-6: COLORADO:

IV-6.1: WATER QUALITY CONTROL:

The Colorado Water Quality Control Act[2] created the Colorado Water Quality Control Commission. The Commission is required to develop and maintain a comprehensive and effective program for prevention, control, and abatement of water pollution and for water quality protection throughout the entire state. The Commission has established regulations for the state discharge permit system, water quality standards, waste discharge standards, regulations for effluent limitations, treatment work site approval regulations and storm sewer discharge regulations. These standards and regulations would be of direct concern to the selection of a site for the location of a coal conversion facility. The antidegradation clause contained in Colorado's water quality standards should be carefully evaluated to guage its impact on the location of a plant with large water consumptive use because of the reduction in flow and waste loading capability.

IV-6.1.1: WATER QUALITY STANDARDS:

All the state waters are classified into four different classes depending upon the designated use. Class A_1 or A_2 waters are designated to be or to become suitable for all purposes including primary contact recreation such as swimming or water skiing. Class B_1 or B_2 are designated to be or to become suitable for all purposes except primary contact recreation. Specific characteristics such as bacteriological concentrations, dissolved oxygen, pH, temperature and other wastes are defined for each one of the 4 classes of waters mentioned above. Such characteristics are shown in Appendix B in the table listing the water quality standards.

A comprehensive list of all the main streams in the 11 river basins in Colorado with their classification is included as a part of the water quality

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[1]Arizona Revised Statutes, Title 36, Water Pollution Control, Chapter 16, Article 1; effective March 16, 1967; amended by laws of 1971.
[2]Title 25, Article 8, Colorado Revised Statutes 1973; effected July 1, 1973.

standards. Also, included is a list of tributaries and standing bodies of water on the main stem and tributaries along with their classification.

IV-6.1.2: DISCHARGE STANDARDS:

Standards for the discharge of wastes into any classified waters of the state or any waters of the state which are tributories to such classified waters have been established by the Colorado Water Pollution Control Commission.[1] Specific standards include allowable levels for settleable solids, suspended solids, biological oxygen demand, chemical oxygen demand, pH, fecal coliform, dissolved oxygen, etc. and are included in Appendix B.

IV-7: INDIANA:

IV-7.1: WATER POLLUTION CONTROL:

The Indiana Stream Pollution Control Law[2] created the Stream Pollution Control Board of the State of Indiana and granted it authority to promulgate water quality standards. Subsequently, the Board promulgated both water quality criteria (water quality standards) and waste water treatment requirements. Included under Section 1 of the water quality standards is a non-degradation provision which has to be taken into account if the site selected happens to be on a stream that is relatively clean and the plant has the potential of generating waste waters. Water use designations and the application of such designations to various streams and water bodies are included under Sections 2 and 3. Section 4 contains a definition of mixing zones and Section 5 provides for the applicability of stream quality criteria except when the stream flow is less than "the average minimum seven-consecutive-day low flow which occurs once in ten years."

IV-7.1.1: WATER QUALITY CRITERIA:

The water quality criteria under Section 6 specifies standards for pH, taste and odor, toxic substances, dissolved oxygen temperature, fecal coliform bacteria, dissolved solids, radio active substances, and chemical constituents. Different standards are applicable within and outside the mixing zones and for the months of April through October which are designated as the recreational season. Specific standards including qualitative limits are given in Appendix B.

IV-7.1.2: WASTE WATER TREATMENT REQUIREMENTS:

As specified in Section 7, all municipal and semi-public waste waters are required to be subjected to either secondary or advanced treatment prior to its discharge into the state waters. Phosphorous removal and effluent disinfection is also required for certain types of effluents from waste water treatment plants. Industrial waste water treatment plants are required to treat wastes containing organic material and/or suspended solids in a manner that would produce an effluent equal to or better than that discharged

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[1]Colorado Water Pollution Control Commission Standards for the Discharge of Wastes; adopted November 21, 1972, effective January 15, 1973.

[2]Chapter 214, Acts of 1943; amended by Chapter 132, Acts of 1945; and amended by Chapter 64, Acts of 1957.

from municipal or semipublic treatment facilities. Waste waters containing other types of contaminants are required to use "best practicable degree of treatment or control consistent with technological feasibility, economic reasonableness and sound engineering judgement." Phosphorous removal is also required for industrial waste waters.

IV-7.1.3: NPDES PERMIT REGULATIONS:

Regulations prescribing the policy and procedures to be followed in connection with issuance of construction, operation, and discharge permits under the Environment Management Act and issuance of discharge permits under the National Pollutant Discharge Elimination System (NPDES) program required by the Federal Water Pollution Control Act are contained in the Indiana Stream Pollution Control Board Regulations SPC-15.

IV-8: KENTUCKY:

The Kentucky Environmental Protection Law[1] established an Environmental Quality Commission, and a Department of Environmental Protection. The Department of Environmental Protection has the authority, power and duty to administer and enforce all rules, regulations and orders promulgated by the Commission. They are required to develop comprehensive plans and programs for the management of water, land and air resources. Regulations effecting coal conversion plants have been promulgated. Plans and programs which may in the future effect the location of such plants in various areas are being developed and may be promulgated in the near future.

The Kentucky Environmental Protection Law includes the Ohio River Valley Water Sanitation Compact between the states of Illinois, Indiana, Kentucky, New York, Ohio, Pennsylvania, Tennessee and West Virginia. It also includes the Tennessee River Basin Water Pollution Control Compact and a few other interstate water quality control agreements. These should be evaluated to determine possible impact on the siting of coal conversion facilities in Kentucky as well as other participating states.

IV-8.1: WATER POLLUTION REGULATIONS:

IV-8.1.1: WATER QUALITY STANDARDS:

Section 2 of the water quality standards[2] prescribes minimum standards applicable to all the waters of the state. The standards require waters to be substantially free from the following; substances attributable to municipal, industrial and other discharges; floating debris, oil, scum and other floating materials; materials producing color odor or other conditions causing nuisance; and substances toxic to human, animal, plant or aquatic life. Sections 4 to 9 contain minimum conditions specified for designated use of the water.

Section 4 specifies criteria applicable to surface water at the point at

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[1]Kentucky Revised Statutes, Chapter 224; amended through September 15, 1973; amended by Chapter 355, Laws of 1974. Approved March 29, 1974.

[2]Kentucky Department for Natural Resources and Environmental Protection, Division of Water Quality, 401 KAR 5:025; adopted June 5, 1975; effective July 2, 1975.

which water is withdrawn for use for public water supply or by a food processing industry. Standards are established for bacteria, threshold-odor number, dissolved solids, radio active substances and chemical constituents (arsenic, barium,cadmium, chronium, cyanide, fluoride, lead, selenium and silver). The location of a coal conversion facility upstream from such a withdrawal point may pose problems. Even if no discharges are generated from the coal conversion plant itself, discharges from other plants located upstream or in between the withdrawal point and the conversion plant, may allow for the standard to be exceeded due to withdrawal of large quantities of water by the conversion plant.

Criteria for industrial water supply, aquatic life, put-and-take trout streams, recreational, agricultural, and multiple uses are contained in Sections 5 to 10 respectively.

Specific qualitative and quantitative water quality standards are shown in Appendix B.

IV-8.1.2: STREAM DISCHARGE REGULATIONS:

Section 2 of the Kentucky Stream Discharge Regulations[1] requires the use of best practicable control technology, considering such factors as the total cost of the application of such technology in relation to the effluent reduction benefits to be achieved, etc., for treatment facilities not later than July 1, 1977. Best available control technology economically achievable by July 1, 1983 is required to be applied for treatment of effluent for all point sources. Discharges containing coliform bacteria are required to disinfect effluent to meet standards specified in the section dealing with the water quality standards.

IV-9: MONTANA:

The Montana Environmental Policy Act[2] and Water Pollution Control Law[3] established the framework for the development, promulgation and enforcement of pollution control rules and regulations in the State of Montana. The Environmental Policy Act is administered by the Montana Environmental Quality Council and the Water Pollution Control Law by the Montana Department of Health.

IV-9.1: WATER POLLUTION REGULATIONS:

IV-9.1.1: WATER QUALITY CRITERIA:

The water quality criteria adopted by the Montana Department of Health and Environmental Sciences on July 13, 1973 and subsequent amendments establishes maximum allowable changes in water quality and limits for pollutants which affect prescribed beneficial uses of state waters. Regulations prescribe the

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[1]Ibid., 401KAR 5:035; adopted March 13, 1975; effective July 2, 1975.
[2]Revised Code of Montana, Title 69, Public Health and Safety, Chapter 65, Montana Environmental Policy Act; effective March 9, 1971.
[3]Revised Code of Montana, Title 69, Public Health, Chapter 48 - Water Pollution as amended.

use of best practicable treatment and control of waste. Water-use classifications and descriptions of these classifications with specific water quality criteria for each one of the classifications are given in Sections 4 and 5 respectively. The general water quality criteria is included in Section 6. The water quality criteria under Sections 5 and 6 specify standards for fecal coliform, dissolved oxygen, pH, turbidity, temperature, sediment and settleable solids or residues, toxic substances, radionuclides and metals. These standards are shown in Appendix B.

IV-9.1.2: MONTANA POLLUTANT DISCHARGE ELIMINATION SYSTEM:

The Montana State Department of Health and Environmental Sciences Regulation 16-2.14(10)-S14460 establishes Montana Pollutant Discharge Elimination System under which all point sources are required to obtain a permit. Permits are granted if the point source is shown to have used treatment methods that would allow compliance with all the applicable state and Federal effluent standards and water quality standards. The minimum state applicable effluent standards are those standards adopted by the U.S. Environmental Protection Agency. No specific effluent standards for Montana were available for inclusion in this paper.

IV-10: NEW MEXICO:

IV-10.1: WATER POLLUTION REGULATIONS:

The New Mexico Water Quality Act[1] established the Water Quality Control Commission and granted the Commission to adopt a comprehensive water quality program; adopt water quality standards as a guide to water pollution control, and adopt regulations to prevent or abate water pollution. Subsequently, water quality standards for interstate and intrastate streams were adopted by the Commission on August 22, 1973.

IV-10.1.1: WATER QUALITY STANDARDS:

Section B of the water quality standards does not allow degradation of existing water quality unless it is proved to be economically and socially justifiable. To protect existing water, the Commission may require the highest and best degree of effluent treatment practicable.

Qualitative general standards for stream bottom deposits, floating solids, oil and grease, color, odor and taste , plant nutrients, hazardous substances, radio activity, pathogens, temperature, turbidity, salinity, dissolved gases, and mixing zones and zones of passage are included under Section C. The stream use designations and specific standards applicable to such designations are included under Section G and in this paper in Appendix B. Since these standards vary from stream to stream the selection of a specific site should take into consideration the applicable water quality standards and the effluent treatment requirements along with the effects on water quality of withdrawal of large quantities of water.

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[1]Chapter 190, Laws of 1967 as amended by Chapter 64, Laws of 1970 and Chapter 277, Laws of 1971; 75-39-1 through 75-39-12 NMSA, 1953 compilation.

IV-10.1.2: EFFLUENT STANDARDS:

There are no specific effluent standards applicable to point sources in New Mexico. The application of best practicable control technology is required in most cases along with compliance with the water quality standards.

IV-11: NORTH DAKOTA:

IV-11.1: WATER POLLUTION REGULATIONS:

The North Dakota Water Pollution Control Act[1] authorizes the control, prevention, and abatement of pollution of the surface waters of the state. It established a State Water Pollution Control Board and authorized the State Department of Health to jointly establish rules and regulations designed to prevent water pollution. Subsequently, surface water quality standards were promulgated in October of 1973 for purposes of maintaining and improving the quality of waters in the state.

IV-11.1.1: WATER QUALITY STANDARDS:

The general requirements under Section IV do not allow the discharge of any untreated domestic or industrial waste into the surface waters. The degree of treatment of municipal waste is based on certain qualitative and quantitative standards for BOD, suspended solids, fecal coliform and pH. Industrial wastes are required to receive a minimum of secondary treatment after maximum in-plant control of waste waters. Treatment has to be adequate so as not to cause a violation of the water quality standards and meet all the requirements of the National Pollutant Discharge Elimination System.

Specific standards of quality for designated classes of waters of the state are contained in Section VI of the water quality standards. All the waters are put into four classes and specific standards are established for each of these classes. In cases where waters contain certain natural characteristics or properties exceeding the limits specified under any of the four classes then such natural background level is used as the standard for that particular parameter as a basis for controlling the addition of wastes from controllable discharges. The specific water quality standards are included in Appendix B.

IV-11.1.2: EFFLUENT STANDARDS:

Standards applicable to waste water effluent are included as a part of the water quality standards and were discussed in Section IV-11.1.1

IV-12: OHIO:

The Ohio Environmental Protection Agency is responsible under the Environmental Protection Act[2] for administering the laws and regulations pertaining to the prevention, control, and abatement of air and water pollution, public water supply, comprehensive water resource management planning, and the disposal

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[1]Sections 61-28-01 through 61-28-08, North Dakota Century Code, 1971 Supplement, effective July 1, 1967; amended by Chapter 501, approved March 27, 1973).

[2]Ohio Revised Code, Sections 3745.01 through 3745.09; amended by S.B. 288, effective September 30, 1974.

and treatment of solid wastes, sewage, industrial waste, and other wastes. The Act also established an Environmental Board of Review which is a quasi-judicial authority responsible for the adoption of rules and regulations, conducting public hearings, etc. The Environmental Board of Review has adopted air and water pollution control and other waste disposal regulations that would have to be taken into consideration when selecting a site for the location of a coal conversion facility.

IV-12.1: WATER POLLUTION CONTROL REGULATIONS:

The Ohio Water Pollution Control Act[1] along with the Environmental Protection Act are administered by the Ohio Environmental Protection Agency. These two Acts together provide the frame work for the establishment and enforcement of various water pollution regulations in Ohio.

The Ohio Water Pollution Control Board (not in existence at present) adopted a resolution on February 9, 1971 regarding discharge of 'toxins' into the waters of the state. All sources that produce toxins (i.e., materials containing toxic substances) are required to eliminate any conceivable possibilities of discharges of said toxins into the waters. Also, under this resolution the Ohio Department of Health was to be provided with a list of all chemicals handled or used in industrial, commercial, agricultural and related operations along with their toxicity. This list should be evaluated to determine the degree of control required to reduce the possibility of toxic substances being discharged from a coal conversion facility into the waters.

Ohio EPA's regulations EP-31-01 through EP-31-10 (adopted December 9, 1973; effective December 30, 1973) establishes the Ohio NPDES permit requirements. The issuance of a NPDES permit is based on compliance with the following water quality and effluent standards established by the Ohio EPA.

IV-12.1.1: WATER QUALITY STANDARDS:

Ohio EPA regulations EP-1-01 through EP-1-09 (the last one adopted December 10, 1974) specify water quality standards. Unlike a few other states, under regulation EP-1-01 all surface waters in Ohio are classified as appropriate for warm water fisheries, for primary contact recreation, and for processing by conventional treatment into public, industrial, and agricultural water supplies. The only exceptions are when the stream flow falls below the annual minimum 7 day average flow that has a reoccurrence period of once in ten years, when a portion of a water course is determined to be a low-flow stream, and when chemicals are applied for control of aquatic plants or animals. The general water quality standards are specified under regulations EP-1-02 and includes standards for dissolved solids, phenols, nitrate, dissolved iron, chromium (hexavalent), cyanide, dissolved manganese, dissolved oxygen, pH, fecal coliform, temperature, threshold odor, radio activity and other chemicals. Some of these general standards vary depending upon the location of any public water supply intake or any other water intake. Regulations pertaining to mixing zones are specified under regulation

[1]Ohio Revised Code Sections 6111.01 through 6111.09, 6111.31 through 6111.40, 6111.99, 6121.02 through 6121.05 and 6161.02, and amendments.

EP-1-03. Regulations prohibiting deterioration of waters whose quality was better than the water quality standards, mentioned above, as of July 1, 1973 are specified under regulation EP-1-04. The application of best available control technology is required for new sources that may discharge into high quality waters.

Separate water quality standards are specified under regulation EP-1-06 for the Mahoning River and its tributaries, under regulation EP-1-07 for Lake Erie, under regulation EP-1-08 for cold water fisheries and regulation EP-1-09 for Lower Cuyahoga River.

IV-12.1.2 : EFFLUENT STANDARDS:

The Ohio River Valley Water Sanitation Commission adopted standards on sewage and industrial wastes on November 13, 1970. Pollution control standard no. 1-70, administered by the Commission, specifies effluent standards applicable to all sewage and industrial waste. Such wastes are required to be treated to substantially remove settleable solids, oil, debris, scum and other floating material. Chemical constituents radioactive materials, fecal coliform, pH, BOD and temperature are required to be reduced to specified levels shown in Appendix B of this paper.

Pollution Control Standard No. 2-70 specifies allowable maximum temperature for cooling water discharges into the Ohio River starting from the point of confluence of the Allegheny and Monongahela Rivers at Pittsburgh to mile point 0.0 at Cairo, Illinois located at the confluence of the Ohio and Mississippi Rivers.

All the water quality and effluent standards should be taken into consideration when selecting a site for the location of coal conversion facilities.

IV-13: PENNSYLVANIA:

The Pennsylvania Department of Environmental Resources Law[1] provided for the reorganization of environmental responsibilities assigned to various state agencies and the establishment of the Department of Environmental Resources to assume these responsibilities. The law also provided for the establishment of an Environmental Quality Board with the power to develop a master environmental plan for the state. The Pennsylvania Department of Environmental Resources has the primary responsibility for developing and enforcing environmental regulations.

IV-13.1: WATER POLLUTION CONTROL:

The Pennsylvania Clean Streams Laws[2], along with the other laws regarding environmental protection, provides for the preservation and improvement of the purity of the waters of the Commonwealth of Pennsylvania for the protection of public health, animal and aquatic life, for industrial consumption and recreation. General provisions relating to the administration

[1]Act No. 275, 1970, House Bill No. 2213: effective January 19, 1971.
[2]Act of June 22, 1937, PL1987, as amended May 8, 1945, PL435; August 23, 1965, PL372, and July 31, 1970, PL 2221.

and enforcement of regulations and permitting of sewage treatment facilities were issued by the Department of Environmental Resources (Pennsylvania Code, Title 25, Environmental Resources, Part I, Article 2, Chapter 91, General Provisions, adopted September 2, 1971, amended October 31, 1973).

Pursuant to the requirements of the Pennsylvania Sewage Facilities Act[1], the Department of Environmental Resources of the Commonwealth of Pennsylvania issued regulations (Pennsylvania Code Title 25, Environmental Resources, Part I, Article I, Chapter 71, as amended) dealing with the municipal planning requirements and the administration of permits for individual and community sewage systems. These regulations would not directly effect the location of a coal conversion facility but would require planning and obtaining permits for sewage systems associated with the development of a community at the plant site.

IV-13.1.1: WATER QUALITY CRITERIA:

Pennsylvania water quality criteria (Pennsylvania Code, Title 25, Part I, Environmental Resources, Chapter 93, Water Quality Criteria, adopted September 2, 1971, as amended) are based upon water uses to be protected and are considered in regulating the discharge of waste waters. Section 93.2 and 93.3 gives a description of 'protected water uses' and 'standard water uses' respectively. The general water quality criteria established in Section 93.4 does not allow substances attributable to municipal, industrial or other wastes discharges in concentrations or amounts sufficient to be harmful to human, animal, plant or aquatic life. Substances to be controlled include floating debris, oil, scum and of floating materials, toxic substances and substances which produce color, taste, odors, or settle to form sludge deposits.

Specific water quality criteria are established under Section 95.3 and includes limits on pH, dissolved oxygen, iron, temperature, dissolved solids, bacteria, turbidity, odor, fluoride, cyanide, chlorides, phenol, ammonia-nitrogen, etc. Designated water uses and water quality criteria for such uses is specified under Section 93.6.

Specific water quality criteria for various designated uses is included in this paper in Appendix B and should be compared to determine the most cost effective site for the location of a coal conversion facility.

IV-13.1.2: EFFLUENT STANDARDS:

Pennsylvania NPDES permit regulations (Pennsylvania Code, Title 25, Environmental Resources, Part I, Article 2, Chapter 92, adopted September 28, 1973) requires all point sources discharging waters into surface streams to obtain a NPDES permit. Permits are issued only if the source is shown to be in compliance with all the effluent limitations under the Federal Water Pollution Control Act, standards of performance for new sources where applicable, and any other more stringent limitation applicable to the source.

Waste water treatment regulations are specified under Chapter 95 and

[1]Act of January 24, 1966, PL 1535, as amended, 35 PS §750.1 et seq.

and industrial waste regulations under Chapter 97 (Pennsylvania Code, Title 25, Environmental Resources, Part I, Article 2). Specific effluent standards applicable to coal washeries and mineral preparation, handling, or processing plants are included under Section 97.31. Drainage waters from active mineral refuse piles, mineral stock piles and other related facilities are required to meet the standards specified under Section 97.31. Section 97.66 prohibits the discharge of cyanides or other cyanogen compounds in any amounts into the streams. Section 97.81 specifies general standards related to cooling water discharge temperatures.

The Pennsylvania Mine Drainage Permit requirements (Chapter 99) should be taken into consideration, along with the other water pollution regulations mentioned previously, prior to the selection of a specific site.

IV-14: SOUTH DAKOTA:

IV-14.1: WATER POLLUTION CONTROL:

The South Dakota Environmental Policy Act[1] regulates domestic water quality and requires impact statements for actions having significant environmental effects. The South Dakota Water Pollution Law[2] establishes the frame work for the formulation, adoption and enforcement of water pollution control regulations.

IV-14.1.1: WATER QUALITY STANDARDS:

Water quality standards applicable to surface waters are established under the Rules and Regulations of the Board of Environmental Protection (Article 34:04, Chapters 34:04:02 to 04; effective June 30, 1974). Standards are specified for visible pollutants, toxic materials, radioactive substances, pH, taste and odor, nuisance aquatic life, dissolved gases, and petroleum products. Beneficial uses of public waters are specified in Section 34:04)02:30 and subsequent sections contain criteria these various beneficial uses.

Chapter 34:04:03 specifies surface water quality standards and uses which in turn would establish the specific criteria contained in Chapter 34:04:02 under Sections 34:04:02:33 to 46. The specific standards contained in these sections are included in Appendix B and should be taken into consideration when selecting sites for the location of coal conversion facilities.

IV-14.1.2: EFFLUENT STANDARDS:

Under Section 46-25-53 of the South Dakota Water Pollution Law effluent limitations based upon the application of such levels of treatment, technology, and processes as are required under the Federal Water Pollution Control Act are conditions prior to obtaining a discharge permit. Further permit conditions that would allow for more stringent effluent limitations necessary to meet either Federal or state water quality criteria may be established.

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[1]H.B. 753, 1974 Legislature; approved March 2, 1974.

[2]Chapter 46-25, Water Pollution Control, South Dakota Compiled Laws; approved March 27, 1973, and H.B. 567, Laws of 1974, approved February 15, 1974.

IV-15: UTAH:

IV-15.1: WATER POLLUTION CONTROL:

The Utah Water Pollution Control Act[1] established the State Water Pollution Committee and authorized it to develop and enforce programs for the prevention, control and abatement of new or existing pollution of the state waters. The committee is also required to establish water quality standards and classify waters according to their reasonable uses in the interest of the public under such conditions as the committee may prescribe for the prevention, control and abatement of pollution.

IV-15.1.1: WATER QUALITY STANDARDS:

The Utah State Department of Health water quality standards contains under Section II-3 anti-degradation policy and water classifications under Section II-6 to II-16. Specific water quality standards are applicable to each class of waters and these standards are shown in Appendix B of this paper.

Guidelines for formulating water quality standards for the interstate waters of the Colorado river system are also included as a part of the water quality standards. Under these guidelines minimum water quality criteria applicable to interstate waters at agreed state line sampling points are specified. Such interstate water quality agreements should be given careful consideration when a site located on interstate waters is being considered.

IV-15.1.2: EFFLUENT STANDARDS:

No specific effluent standards have been established for discharges of pollutants into the Utah waters. Construction permits to construct treatment facilities or discharge of waste water are required. Waste water treatment facilities are required to be operated so as not to pollute the receiving waters.

IV-16: WEST VIRGINIA:

IV-16.1: WATER POLLUTION CONTROL:

The West Virginia Water Pollution Control Act[2] authorized the State Water Resources Board to promulgate rules and regulations setting standards of water quality applicable to the waters of the state. The Act also required the establishment of a waste water discharge permit program and specified other administrative procedures for the abatement of water pollution and the maintenance of water quality.

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[1]Utah Code Annotated, 1953, Title 73, Chapter 14, Pollution of Waters, as amended by Chapter 197, Laws of 1969; amended October 24, 1973, effective December 12, 1973.

[2]Chapter 20, Article 5A, West Birginia Code, as amended, 1969.

IV-16.1.1: WATER QUALITY STANDARDS:

West Virginia water quality standards were promulgated by the State Water Resources Board (Chapter 20, Articles 5 and 5A, Code of West Virginia, effective April 15, 1974). Mixing zone dimensions for various designated water uses are given under Section 2.03 of these regulations followed by general water quality conditions under Section 3. Section 5 specifies certain acid mine drainage control measures.

Seven water use categories are specified in Section 6 including water contact recreation, public and industrial water supply, etc. Water quality criteria for these seven categories are specified in Section 7 through 14. The specific standards for the these categories of water are given in Appendix B of this report.

IV-17: WYOMING:

The Wyoming Environmental Quality Act[1] created the State Department of Environmental Quality consisting primarily of air, water and land quality divisions and an Independent Environmental Quality Council. The Council is authorized to promulgate rules and regulations necessary for the administration of the requirements specified in the Act after holding public hearings and receiving recommendations from the Director of the Department, the Administrator of the various divisions and their respective advisory boards. Rules and regulations controlling water, air and land pollution has been adopted by the Council and these one way or the other effect the construction of a coal conversion facility. The general provisions are mentioned in the text of this paper with specific standards included under Appendix B at the end of the paper.

IV-17.1: WATER POLLUTION STANDARDS:

Department of Environmental Quality's Rules and Regulations, Chapter II, deals with the NPDES permit program. These regulations specify that the owner or operator of any point source within the state planning to or presently discharging, or increasing the volume of discharge is required to obtain a permit at least 180 days prior to the commencement of such an activity.

IV-17.1.1: WATER QUALITY STANDARDS:

Water quality standards were adopted by the Department on July 30, 1974 and became effective on August 28, 1974. Section 4 of the standards divide the waters of the state into three classes depending on the use. High quality waters are defined under Section 13 and standards for 15 different parameters are specified in Sections 15 through 29. These parameters include settleable and floating solids, toxic materials, temperature, pH, etc. and the specific standards are included in Appendix B.

---

[1]Enrolled Act No. 107, 42nd Session, effective July 1, 1973, amended effective February 12, 1974, amended March 12, 1975.

V. COST OF POLLUTION CONTROL:

The incremental cost of pollution control for a commercial scale coal conversion facility would depend upon alot of variables including the type of process, quality of the feedstock to be used, and the stringency of the pollution control standards. For processes designed to use best available pollution control technology, and if such technology allows compliance with the most stringent pollution control standards, variation in the incremental cost of pollution control would mainly depend upon the type of process and the quality of the feedstock.

The types of waste streams generated in general from a coal conversion facility along with the specific source and treatment are listed in Table 4. Any or all of these waste streams could be associated with coal conversion facilities using processes presently in commercial scale application or those that are being developed for such application. As can be seen from Table 4 all the waste streams will receive extensive treatment and then disposed off into the environment or reused within the facility itself. In some cases, e.g., sulfur, the product of the waste treatment process will be stored for eventual sale or sold as by-product immediately.

The effect of the quality of the feedstock on the incremental cost of pollution control for a coal fired boiler associated with a coal conversion facility is shown in Figure 14 in terms of sulfur content of coal (28). The increases in levelized unit operating costs, which includes both capital charges and operating costs, shown are for a constant 90% sulfur dioxide removal. Therefore, the sulfur dioxide emission rate would increase proportionately with the sulfur content of coal. However, 90% removal of sulfur dioxide while burning coal with a sulfur content of as high as 5 % would still allow the source to be in compliance with the new source performance standard of 1.2 lb of sulfur dioxide per million Btu heat input (assuming a heat content of 10,000 to 10,500 Btu/lb of coal). If further reductions in sulfur dioxide emission rates are required, e.g., because of more stringent regulations or air quality maintenance plans or significant air quality deterioration provisions, the sulfur dioxide removal rate may have to be increased above 90%. If such cases are encountered, then the unit operating costs shown in Figure 14 would also increase along with the total incremental cost of pollution control.

The same type of argument is valid for most of the waste treatment processes listed in Table 4. The only exceptions are those that would generate saleable by-products in quantities enough to offset the increase in costs. One such example may be the sulfur recovery processes.

The increases in cost of complying with a range of pollution control regulations should be calculated after identification of various alternate sites. Such increases in cost may be traded off by selecting a site close to the raw materials. In cases where pollution control costs could be reduced because of less stringent standards the coal conversion facility may have to be located away from the raw materials and the cost of transportation of such raw materials would have to be taking into consideration. Also, in some cases, the cost of transporting waste materials, e.g., ash, if it was to be disposed off in the coal mine, would have to be taken into consideration.

Just to give you an idea of the incremental cost of pollution control, Table 13 gives the costs of pollution control for various processes listed in reference (6). These cost figures seem to have been calculated assuming use of best available pollution control technology during the later part of 1973 and beginning of 1974. Further reductions in concentrations of pollutants in waste streams, if required because of more restrictive standards today, would further increase this cost. Also, since these figures were calculated in 1974 the incremental cost of pollution control can be expected to be alot higher in terms of the total cost and somewhat higher in terms of percentage of the total cost of the coal conversion facility.

For comparison purposes, the cost of producing synthetic gas is shown in the last column and it is evident that the approximate incremental cost of pollution control in 1974 ranged from a low of about 5% to a high of about 24%. If the relative cost remains the same today, it can be concluded that the cost of pollution control has a substantial impact on the cost of synthetic gas.

Time and monetary constraints have not allowed the calculation of incremental cost of pollution control associated with the pollution control standards for each one of the 14 states listed in Chapter 4 of this paper. The specific standards, as applicable to various geographical areas within these states, are shown in Appendix B and for each set of these standards the incremental cost will be calculated in the near future. This would facilitate the selection of a specific site and estimation of the total cost of the coal conversion facility.

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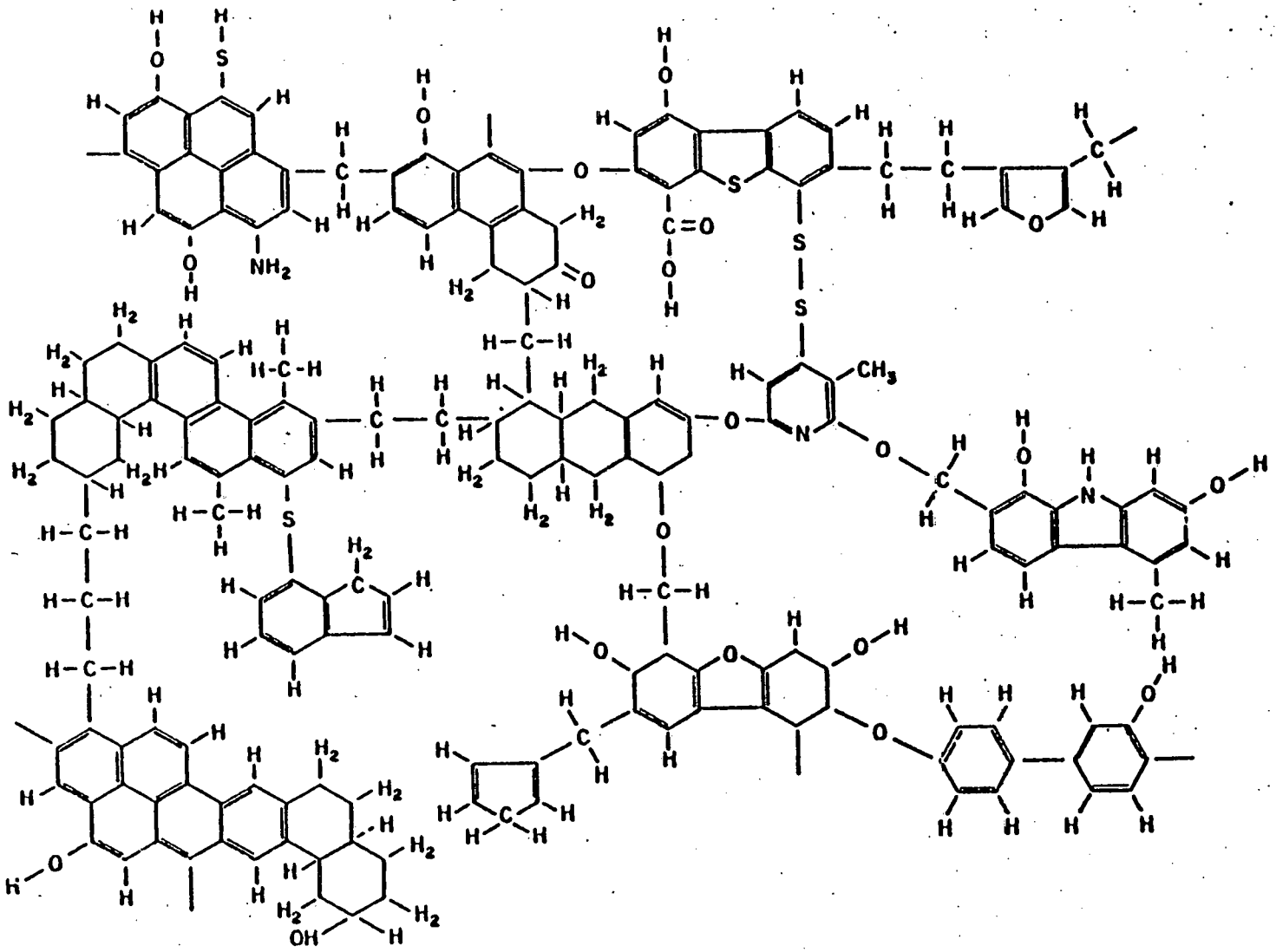
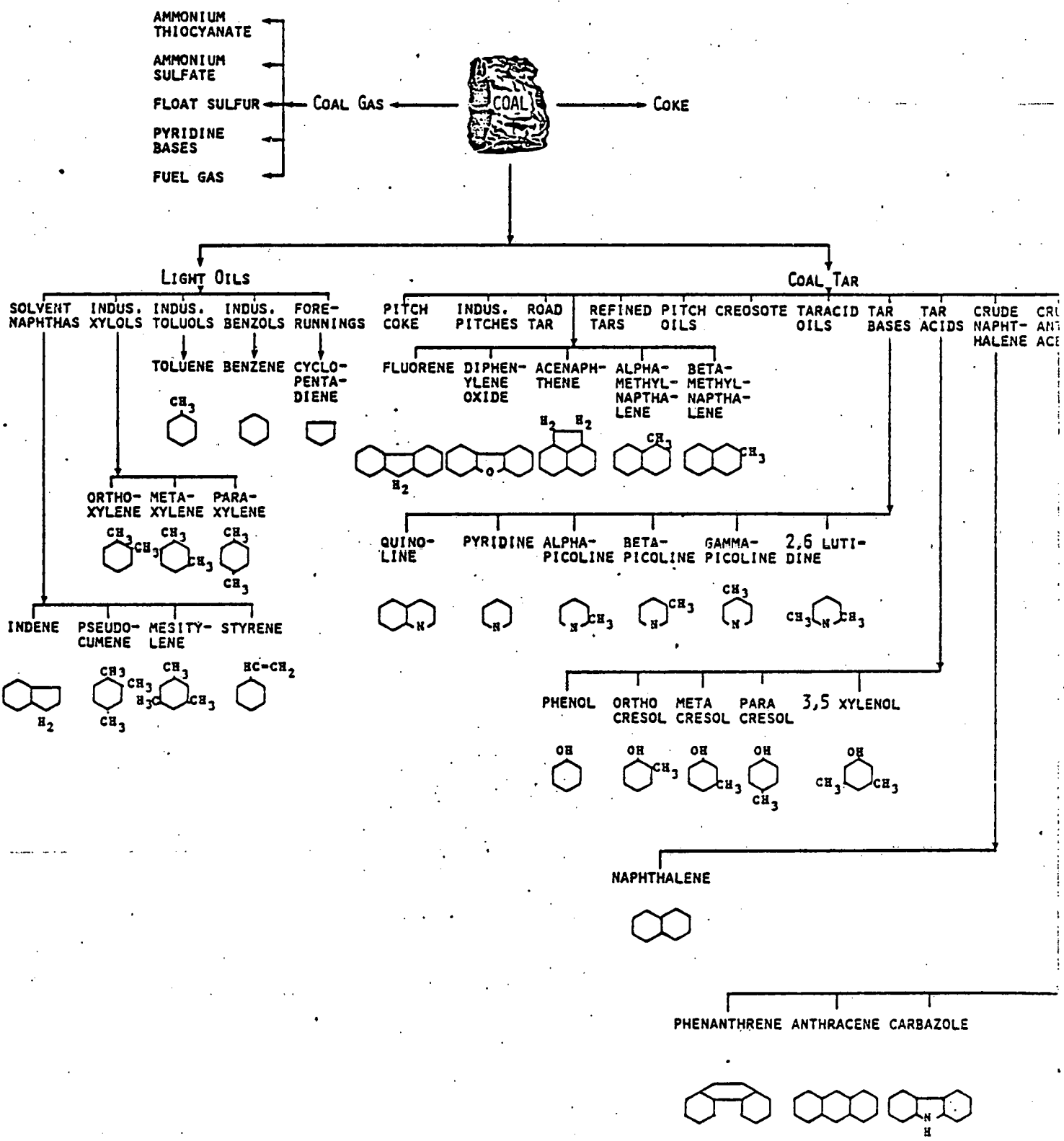


FIGURE 1: A REPRESENTATION OF THE BITUMINOUS COAL STRUCTURE SHOWING AROMATIC RING CLUSTERS AND BRIDGE LINKAGES (1)

Reactions	$\Delta H°_{298}$ Heats of Formati At 25°C in Kcal/mole
(1) $C + 1/2O_2 \rightarrow CO$	-26.42
(2) $C + O_2 \rightarrow CO_2$	-94.05
(3) $C + H_2O \rightarrow CO + H_2$	+31.38
(4) $C + CO_2 \rightarrow 2\ CO$	+41.21
(5) $C + 2H_2 \rightarrow CH_4$	-17.89
(6) Coal + $\Delta \rightarrow$ Char + H_2 + CO + CO_2 + Hydrocarbons + Tars	About 0
(7) $H_2O + CO \rightarrow H_2 + CO_2$	-9.83
(8) $CO + 3H_2 \rightarrow CH_4 + H_2O$	-49.27
(9) $2\ CO + 2H_2 \rightarrow CH_4 + CO_2$	-59.10
(10) $2C + H_2O + H_2 \rightarrow CO + CH_4$	+13.49
(11) $2C + 2H_2O \rightarrow CO_2 + CH_4$	+3.66

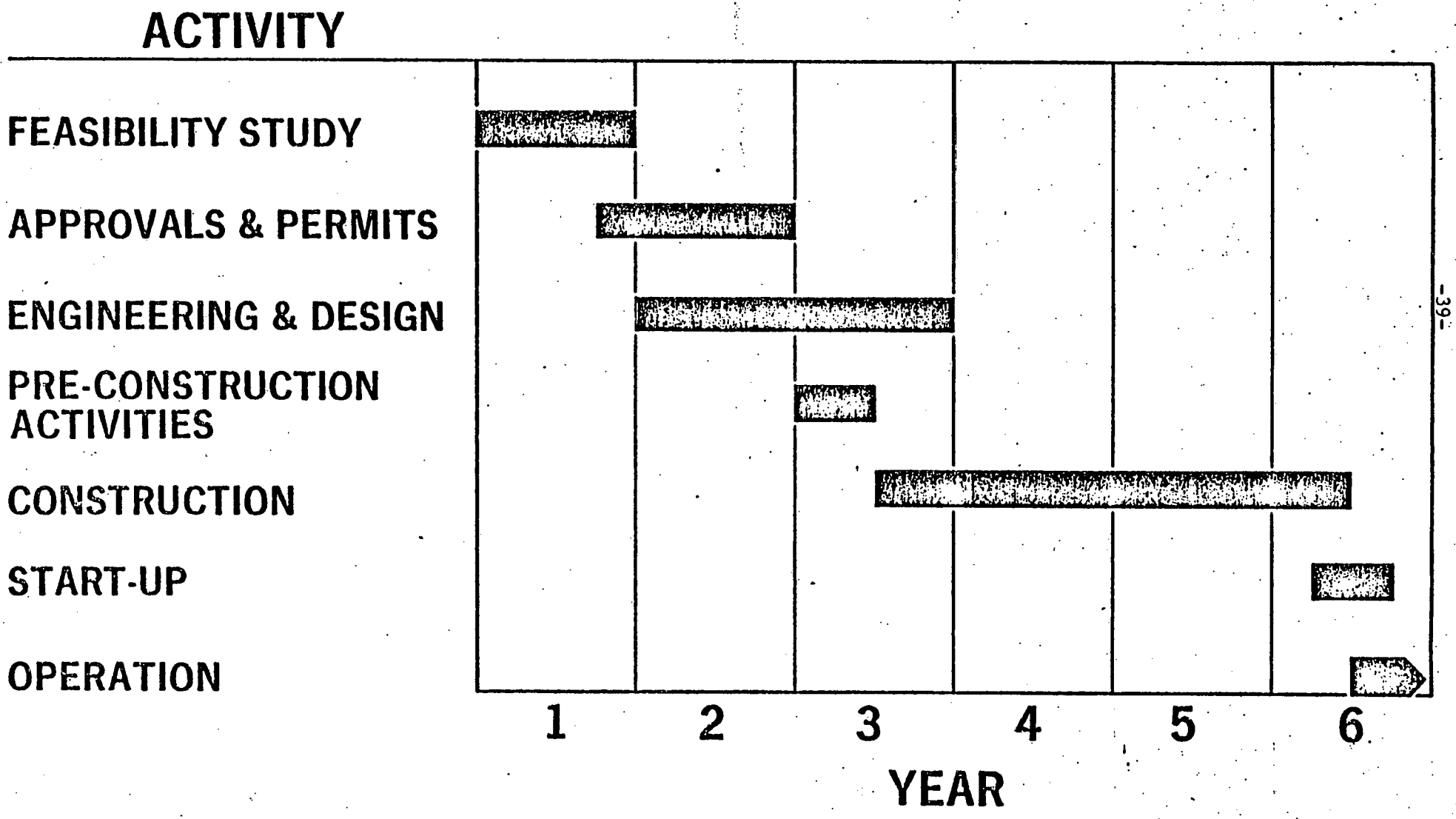
Figure 2: Important Reations of Carbon Associated With the Conversion of Coal to Heat, and Liquid or Gaseous Fuels (2)

FIGURE 3: CHEMICALS FROM COAL



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FIGURE 4: PROJECT SCHEDULE FOR BUILDING COAL CONVERSION FACILITY



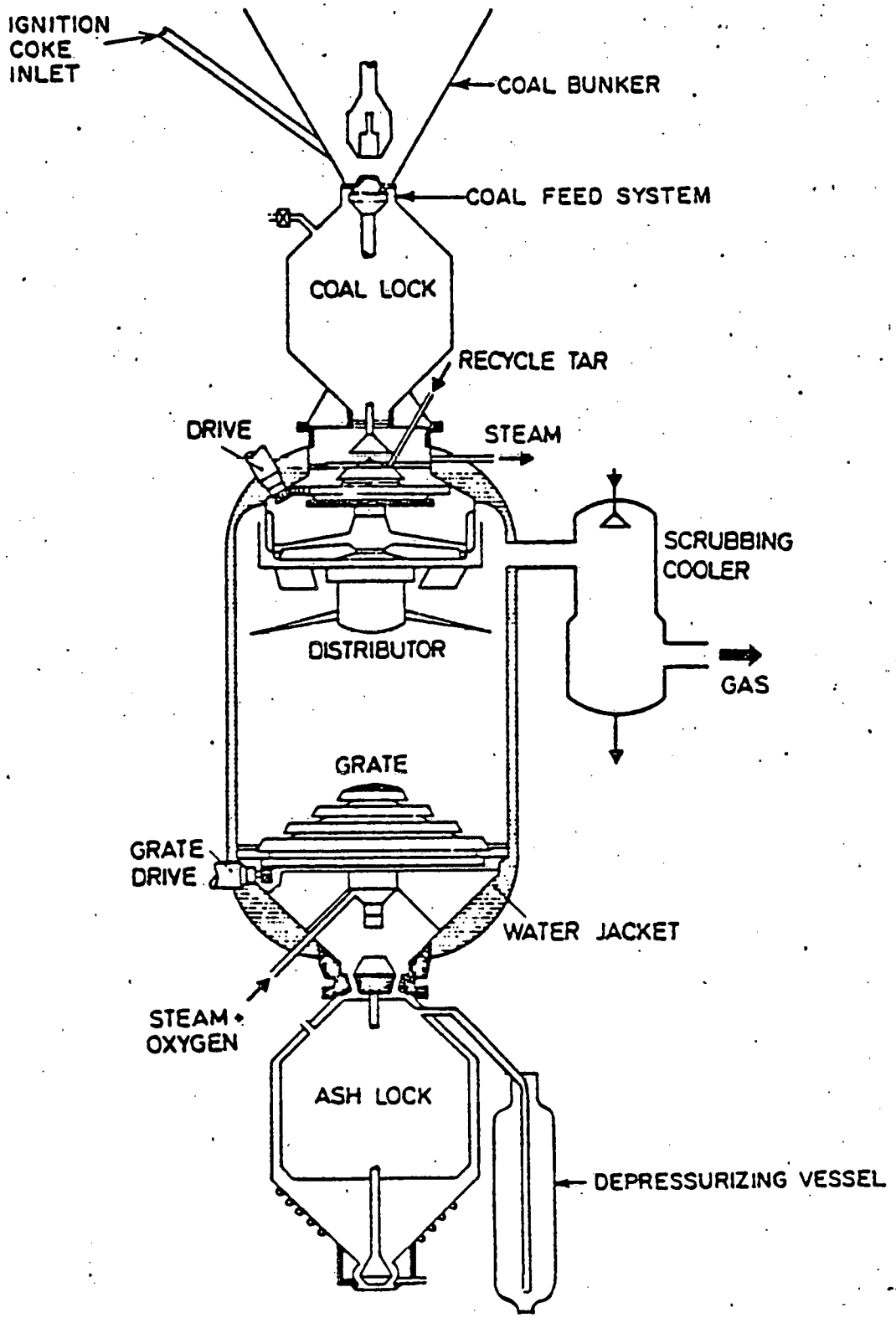


FIGURE 5: THE LURGI GASIFIER

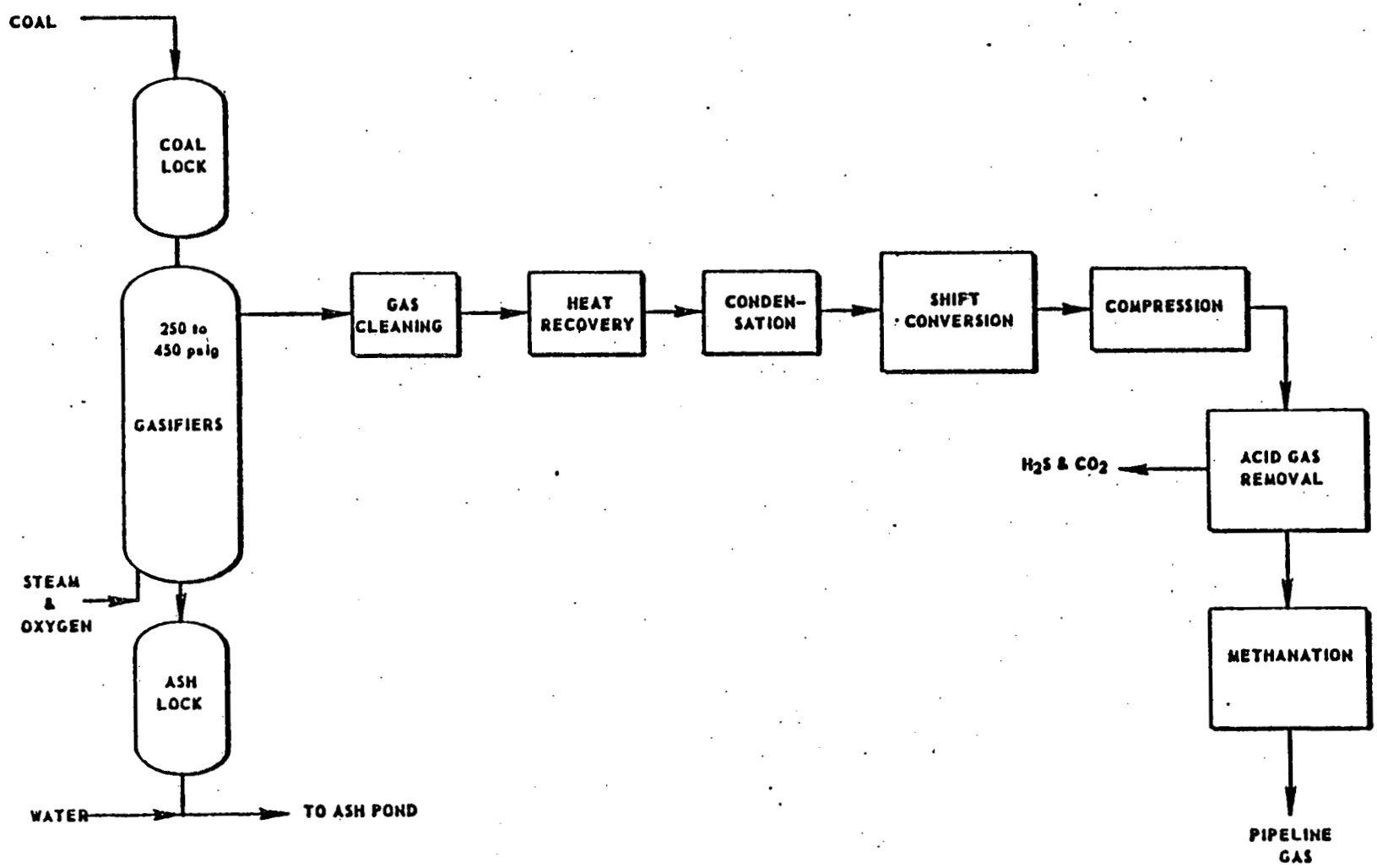


FIGURE 6: SIMPLIFIED PROCESS FLOW DIAGRAM OF A COAL CONVERSION FACILITY USING LURGI PRESSURE GASIFIERS

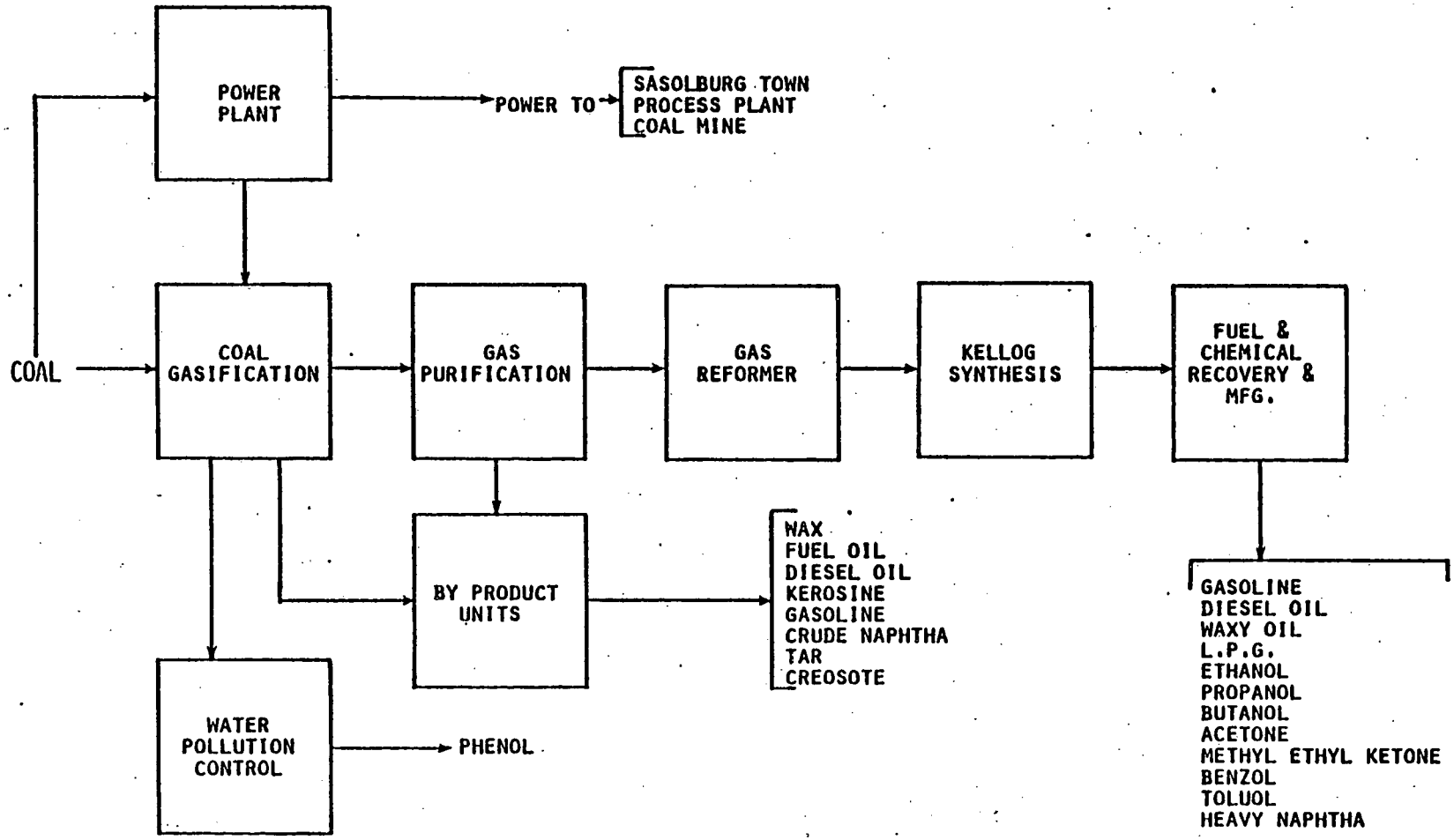
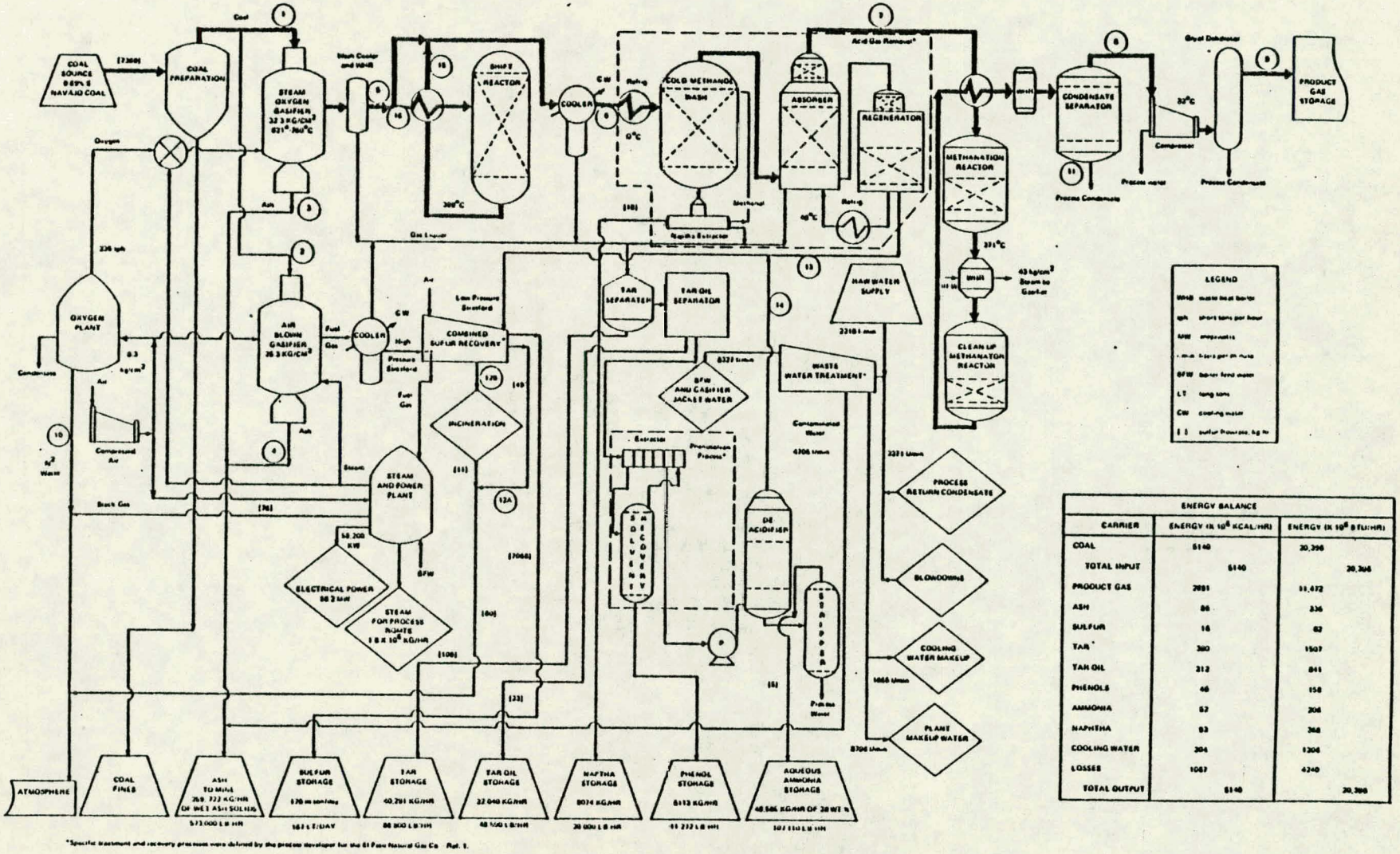


FIGURE 7: PROCESS FLOW DIAGRAM OF THE SASOL COAL CONVERSION FACILITY



ENERGY BALANCE		
CARRIER	ENERGY (X 10^6 KCAL/HR)	ENERGY (X 10^6 BTU/HR)
COAL	5140	20,396
TOTAL INPUT	5140	20,396
PRODUCT GAS	2891	11,472
ASH	86	336
SULFUR	16	62
TAR	380	1507
TAR OIL	212	841
PHENOLS	40	158
AMMONIA	52	206
NAPTHA	83	328
COOLING WATER	304	1206
LOSSES	1067	4240
TOTAL OUTPUT	5140	20,396

*Specific treatment and recovery processes were defined by the process developer for the El Paso Natural Gas Co Ref. 1.

FIGURE 8: PROCESS FLOW DIAGRAM OF A COAL CONVERSION FACILITY USING LURGI PRESSURE GASIFIERS (6).

TABLE 1: COMPOSITION OF STREAMS SHOWN IN FIGURE 8 (6).

STREAM COMPOSITIONS

STREAM NAMES	Oxygen Gasifier Coal Feed	Air Gasifier Coal Feed	Oxygen Gasifier Ash	Air Gasifier Ash	Crude Gas	Gas Purification Feed	Methanation Feed	Methanation Effluent	Pipeline Gas	N_2 Waste to ATM	Methanation Condensate	Off Gas to ATM	Off Gas to Incineration	Combined Acid Gas	H_2S From Waste Liquor	Shift Effluent	Shift Feed
STREAM NUMBERS	1	2	3	4	5	6	7	8	9	10	11	12A	12B	13	14	15	16
Gas Components (Vol. %)																	
O_2	---	---	---	---	---	---	---	---	---	0.86	---	2.29	---	---	---	---	---
CO_2	---	---	---	---	28.03	32.36	3.10	1.81	1.81	---	22.58	85.45	64.79	96.12	95.91	36.95	28.03
H_2S, COS, CS_2	---	---	---	---	0.37	0.34	---	---	---	---	---	80ppm	0.06	1.06	4.09	0.32	0.37
C_2H_4	---	---	---	---	0.40	0.39	0.45	---	---	---	---	0.20	0.11	0.22	---	0.35	0.40
CO	---	---	---	---	20.20	11.70	16.91	.81	0.01	---	---	0.15	7.22	0.41	---	5.03	20.20
H_2	---	---	---	---	38.95	43.63	63.49	4.16	4.16	---	---	0.39	9.63	0.76	---	46.80	38.95
CH_4	---	---	---	---	11.13	10.70	14.92	92.93	92.93	---	77.42	0.47	2.10	0.59	---	9.75	11.13
C_2H_6	---	---	---	---	0.61	0.59	0.69	---	---	---	---	0.27	0.16	0.30	---	0.53	0.61
N_2 + Ar	---	---	---	---	0.31	0.29	0.41	1.09	1.09	---	---	---	---	---	---	0.27	0.31
N_2	---	---	---	---	---	---	---	---	---	99.14	---	10.78	15.93	0.54	---	---	---
FLOW RATES (lb/hr)																	
Total Dry Gas	---	---	---	---	2,280,447	2,450,001	829,704	513,694	513,694	1,577,467	69	1,769,323	46,513	1,670,209	8,853	1,392,164	1,246,945
Water	314,950	67,522	---	---	1,394,960	2,680	---	1,316	66	9,995	314,625	34,569	1,006	10,900	8,870	357,765	702,764
Naphtha	---	---	---	---	20,005	20,005	---	---	---	---	---	---	---	---	---	10,939	10,939
Tar Oil	---	---	---	---	28,007	---	---	---	---	---	---	---	---	---	---	15,314	15,314
Tar	---	---	---	---	7,314	---	---	---	---	---	---	---	---	---	---	3,999	3,999
Crude Phenols	---	---	---	---	9,127	---	---	---	---	---	---	---	---	---	---	4,991	4,991
NH_3	---	---	---	---	17,629	---	---	---	---	---	---	---	---	---	---	9,640	9,640
Coal (MAF)	1,250,310	268,053	19,639	4,209	---	---	---	---	---	---	---	---	---	---	---	---	---
Ash	373,220	80,012	373,220	80,012	---	---	---	---	---	---	---	---	---	---	---	---	---
Total lb/hr	1,938,480	415,587	392,859	84,221	3,757,489	2,472,686	829,704	515,010	513,760	1,587,462	314,694	1,803,892	47,519	1,681,009	17,723	1,794,812	2,054,592
(kg/hr)	(879,530)	(188,560)	(178,250)	(38,210)	(1,704,850)	(1,121,910)	(376,450)	(233,670)	(233,100)	(720,260)	(142,780)	(818,460)	(21,560)	(762,710)	(8,040)	(814,340)	(932,210)
Total Gas Flow Rate x 10^6 ft^3/day	---	---	---	---	982.2	1079.7	734.9	288.6	288.6	511.0	---	386.3	11.9	351.3	1.8	612.9	537.1

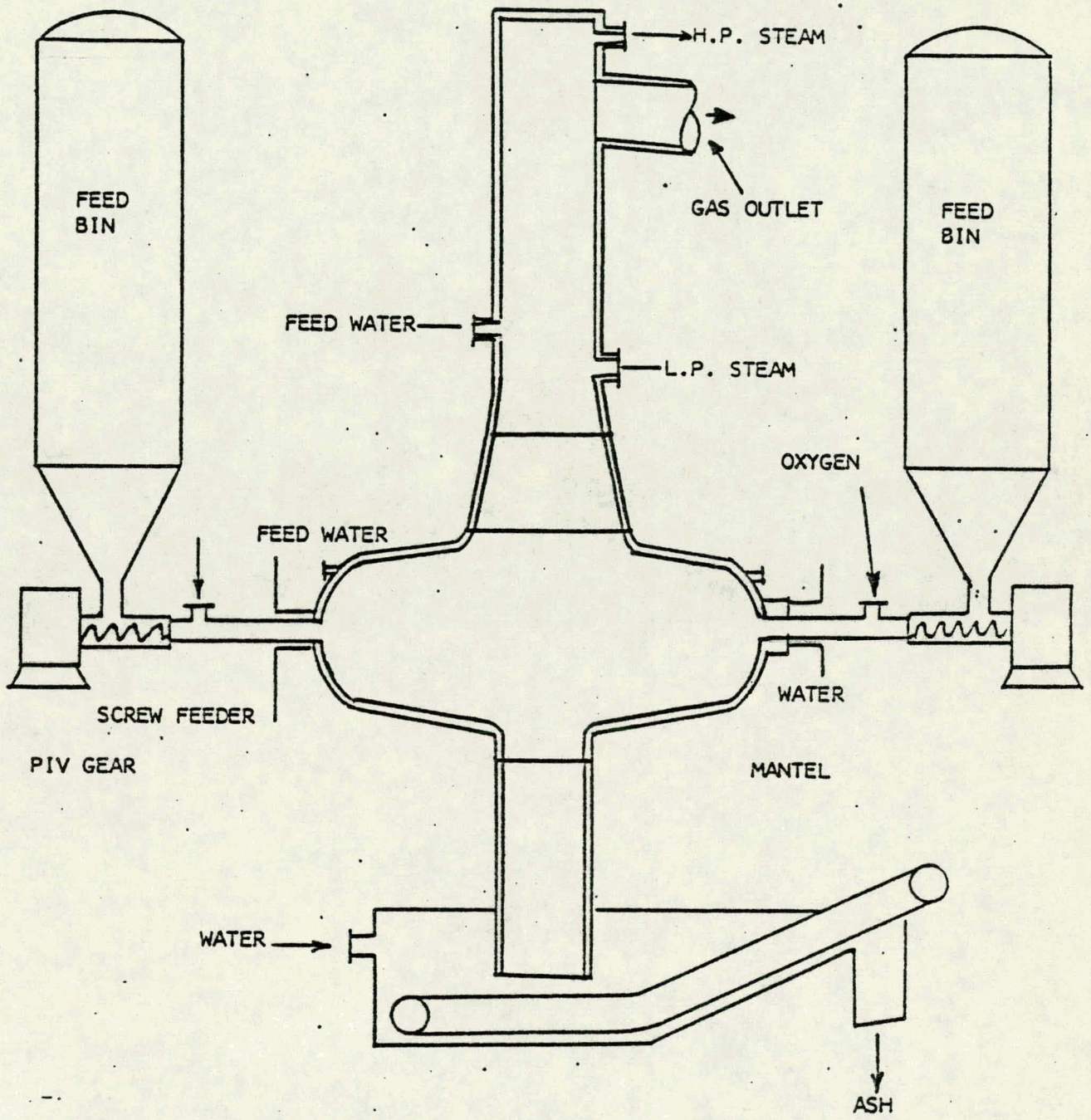
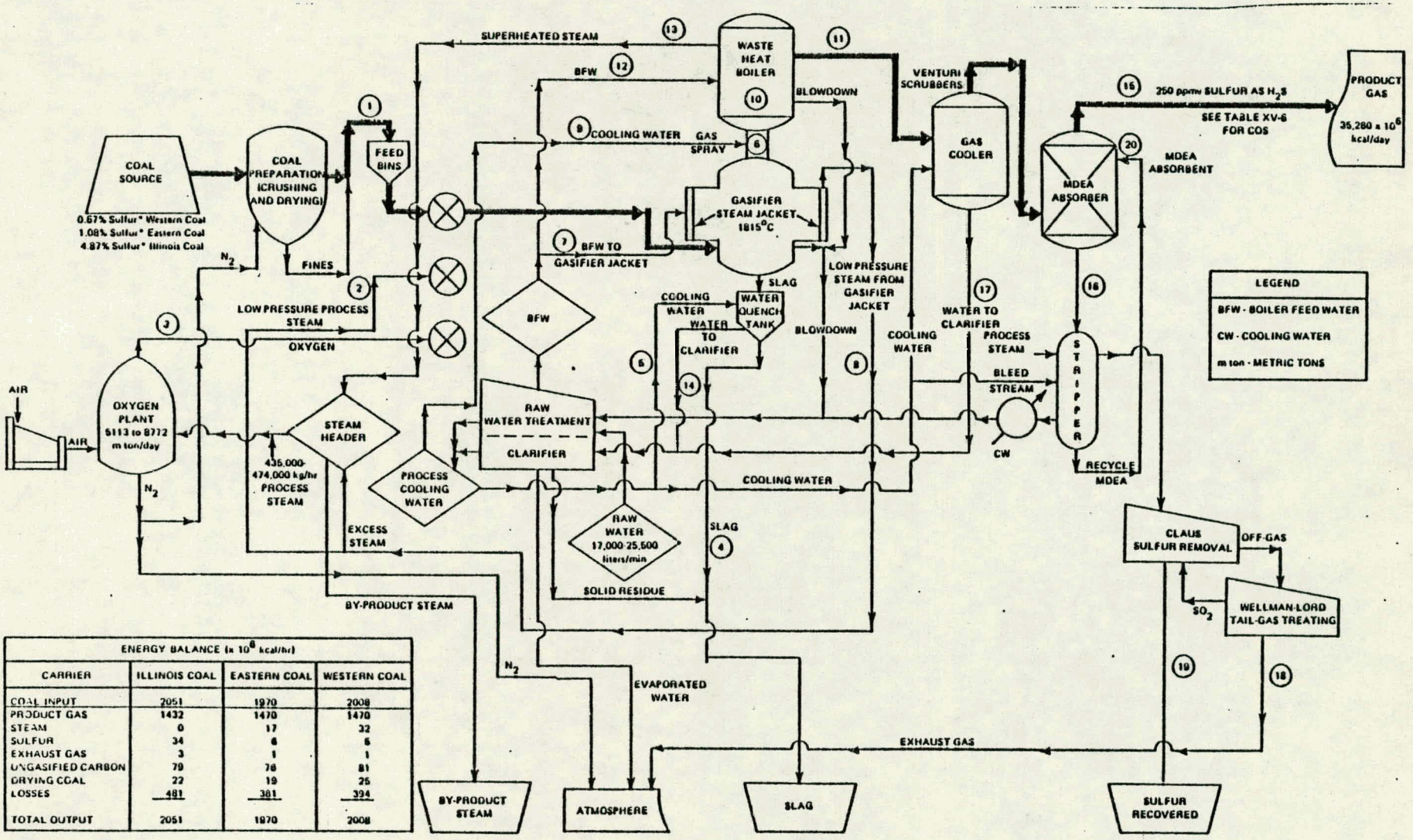


FIGURE 9: SCHEMATIC DIAGRAM OF THE KOPPERS-TOTZEK COAL GASIFIER

FIGURE 10: PROCESS FLOW DIAGRAM OF A COAL CONVERSION FACILITY USING KOPPERS-TOTZEK GASIFIERS (6)

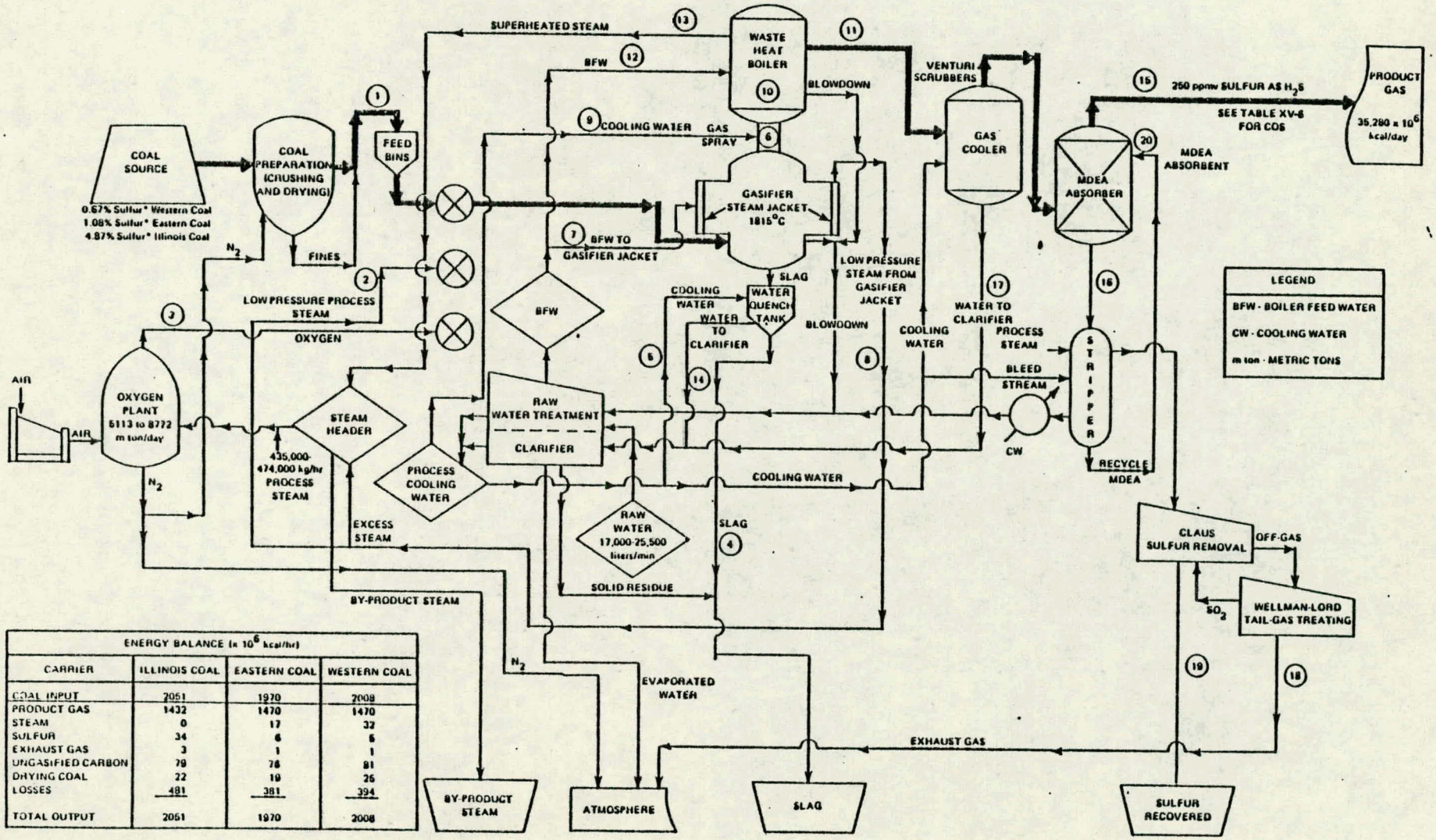


ENERGY BALANCE (x 10^6 kcal/hr)

CARRIER	ILLINOIS COAL	EASTERN COAL	WESTERN COAL
COAL INPUT	2051	1970	2008
PRODUCT GAS	1432	1470	1470
STEAM	0	17	32
SULFUR	34	6	6
EXHAUST GAS	3	1	1
UNGASIFIED CARBON	79	78	81
DRYING COAL	22	19	25
LOSSES	481	381	394
TOTAL OUTPUT	2051	1970	2008

*SULFUR CONTENT WHEN DRIED TO 2% MOISTURE IN EACH CASE.

FIGURE 10: PROCESS FLOW DIAGRAM OF A COAL CONVERSION FACILITY USING KOPPERS-TOTZEK GASIFIERS (6)



ENERGY BALANCE (x 10^6 kcal/hr)			
CARRIER	ILLINOIS COAL	EASTERN COAL	WESTERN COAL
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DRYING COAL	22	19	25
LOSSES	481	381	394
TOTAL OUTPUT	2051	1970	2008

*SULFUR CONTENT WHEN DRIED TO 2% MOISTURE IN EACH CASE.

TABLE 2: COMPOSITION OF STREAMS SHOWN IN FIGURE 10 FOR PROCESS USING ILLINOIS COAL (6)

Stream Names	Dried Coal Feed	Steam	Oxygen	Slag	H_2O From Cooling Tower	Gas Out Gasifier	BFW to Gasifier Jacket	L.P. Steam from Jacket	Spray Cooling Water	Gas to Waste Boiler	Gas Out Waste Boiler	BFW to Boiler	Steam from Boiler	Slag Cool H_2O Return	Gas Out Absorber	Liquid Out Absorber	Condensate Out Cooler	Gas Out Claus	Sulfur Out Claus	MDEA to Absorber
Stream No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
Gas composition (Vol. %)	Wt. %																			
CO	–					51.14				47.38	47.38				61.05					
CO_2	–					5.60				5.18	5.18				–					
H_2	4.35					31.61				29.28	29.28				37.73					
N_2	0.97		2.0			0.98				.91	.91				1.17					
H_2S + COS	–					1.77				1.64	1.64				H_2S .02 COS .01					
HCl	–					0.06				.05	.05				–					
H_2O	2.00					8.84				15.54	15.54				–					
Ash	19.08																			
C	61.83																			
S	4.87																			
Cl_2	0.179																			
O_2	6.72		98.0																	
Flow rate (lb/hr)																				
CO						880,950				880,950	880,950				880,950					
CO_2	–					151,580				151,580	151,580				–	153,824		154,198		
H_2	31,082					38,867				38,867	38,867				38,867					
N_2	6,931		9,956			16,887				16,887	16,887				16,887			59,175		
H_2S + COS	–					38,740				38,740	38,740				H_2S 466 COS 510	H_2S 35,928 COS 510				
HCl	–					1,315				1,315	1,315				–		1,315			
H_2O	14,291				1,521,252	97,887	153,545		87,904	185,792	185,792	992,964		1,521,250	–		185,792	19,020		
Ash	136,333			68,167		68,167				68,167	68,167				–		68,167			
C	441,798					22,090				22,090	22,090				–		22,090			
S	34,797																		34,032	
Cl_2	1,279																			
O_2	48,017		487,830																	
Steam	–	172,338						146,234					945,688							
SO_2	–																			
MDEA	–															1,085,565		109		1,085,565
Total (lb/hr)	714,528	172,338	497,786	68,167	1,521,252	1,316,483	153,545	146,234	87,904	1,404,388	1,404,388	992,964	945,688	1,521,250	937,680	1,275,827	277,364	232,502	34,032	1,085,565
(kg/hr)	324,110	78,173	225,796	30,921	690,040	597,157	69,648	66,332	39,873	637,030	637,030	450,408	428,964	690,039	425,332	578,715	125,812	105,463	15,437	492,412

* BOILER FEED WATER
1 and 8- SERVICE BUNKERS; 2 and 9- FEED BUNKERS;
3 - HIGH PRESSURE STEAM DRUM;
4 - LOW PRESSURE STEAM DRUM;
5 - GASIFIER;
6 - WASTE HEAT BOILER;
7 - GASIFIER QUENCH TANK;

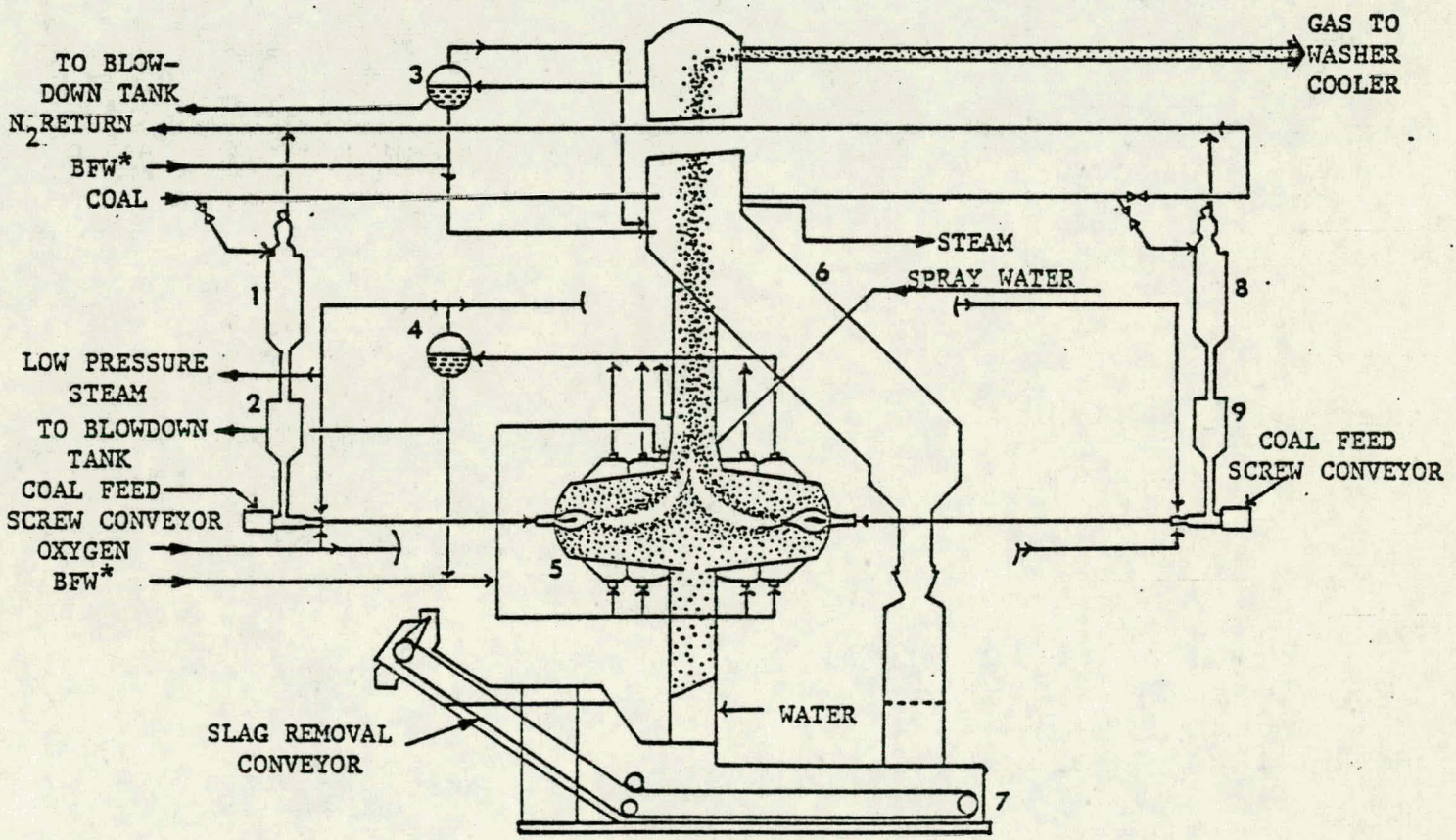
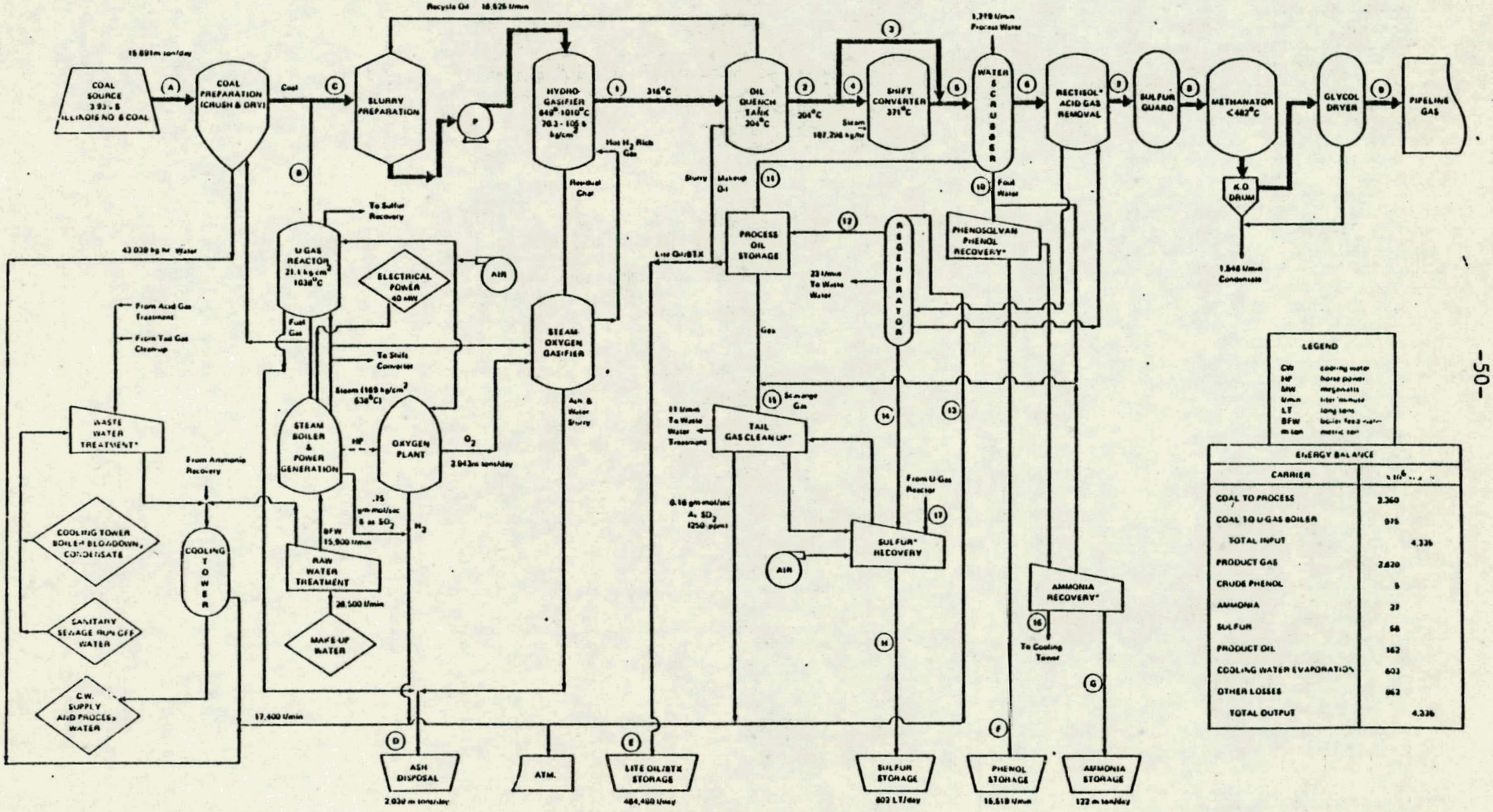


FIGURE 11: SCHEMATIC DIAGRAM OF THE WINKLER GASIFIER

FIGURE 12: PROCESS FLOW DIAGRAM FOR THE HYGAS PROCESS USING ILLINOIS COAL(6)



*Specific Treatment and Recovery Processes were Selected by Senior IGT Advisors for This Analysis.

TABLE 3: COMPOSITION OF STREAMS SHOWN IN FIGURE 12 (6)

Stream name	Raw Gas		Oil Quench Effluent		Shift By-Pass		CO Shift Feed		Combined Shift Effluent		Scrub Effluent		Acid Gas Effluent		Methanation Feed		Product Gas	
Stream number	1		2		3		4		5		6		7		8		9	
Temperature, °C	316		204		--		52		52		52		38		38		38	
Pressure, kg/cm^2	81		81		81		79		79		79		76		76		70	
Component	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %
CO	23,217	19.84	23,217	22.25	7,739	22.25	15,478	16.74	11,304	8.88	11,299	12.04	11,290	18.00	11,290	18.00	28	0.10
CO_2	17,359	14.83	17,359	16.64	5,786	16.64	11,573	12.51	29,294	23.02	28,946	30.85	88	0.14	88	0.14	38	0.13
H_2	24,167	20.66	24,167	23.16	8,056	23.16	16,111	17.42	36,106	28.37	36,090	38.46	35,971	57.33	35,971	57.33	1,866	6.54
H_2O	19,501	16.66	19,501	18.69	6,500	18.69	35,920	38.84	30,459	23.93	165	0.18	--				2	0.01
CH_4	15,260	13.04	15,260	14.63	5,087	14.63	10,173	11.00	15,260	11.99	15,253	16.26	15,212	24.25	15,212	24.25	26,524	93.01
C_2H_6	502	0.43	502	0.48	167	0.48	335	0.36	502	0.39	502	0.54	119	0.19	119	0.19		
C_6H_6	149	0.13	149	0.14	50	0.14	99	0.11	149	0.12	142	0.15						
NH_3	628	0.54	628	0.60	209	0.60	419	0.45	654	0.51	--							
HCN	39	0.03	39	0.04	13	0.04	26	0.03	13	0.01	--							
H_2S	1,361.6	1.16	1,361.6	1.30	453.9	1.30	907.7	0.98	1,357.9	1.07	1,348.4	1.44	trace	0.1 ppmV				
COS	11.8	0.01	11.8	0.01	3.9	0.01	7.9	0.01	15.5	0.01	15.5	0.02						
N_2	59	0.05	59	0.06	20	0.06	39	0.04	59	0.05	59	0.06	59	0.09	59	0.09	59	0.21
Phenol	8	--	8	0.01	3	0.01	5	0.01	8	0.01								
Oil	14,765	12.62	2,079	1.99	693	1.99	1,386	1.50	2,079	1.63	6	0.01						
Total lb-mol/hr	17,027	100.00	104,341	100.00	34,781	100.00	92,480	100.00	127,260	100.00	93,826	100.00	62,739	100.00	62,739	100.00	28,517	100.00
(gm-mol/sec)	(2,150)		(13,160)		(4,390)		(11,660)		(16,050)		(11,830)		(7,910)		(7,910)		(3,600)	
($\times 10^6$ ft^3/hr)	44.3		39.5		13.2		35.1		48.2		35.6		23.8		23.8		10.8	

Note: All flow rates are rounded to the nearest mol/hr except for sulfur bearing compounds. Volume flow rates are calculated at normal conditions of 14.7 psia, 60°F.

TABLE 3: (CONT'D)

Stream name	Foul Water		Oil		BTX- Oil		Acid Gas CO_2 Vent to Stack		H_2S Stream to Claus Unit		Scavenged Gas		Wastewater		SO_2 From U-Gas	
Stream number	10		11		12		13		14		15		16		17	
Component	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %	lb-mol/hr	mol %
CO	5	0.01					9	0.03			5	1.20				
CO_2	348	0.78					25,696	97.87	3,162	70.00	348	86.45				
H_2	16	0.04					119	0.45			16	3.97				
H_2O	43,338	97.61									6	1.49	43,332			
CH_4	7	0.02					41	0.16			7	1.74				
C_2H_6	--						383	1.46			--					
C_6H_6	7	0.02			142	95.50										
NH_3	654	1.47											0.2	5 ppmW		
HCN	13	0.03									12.7	3.16	0.3	12 ppmW		
H_2S	8		1.5	0.07	0.7	0.47	0.3	--	1347.4	29.83	7.8	1.94	0.20	10 ppmW		
COS	0.03	--					7.7	0.03	7.8	0.17	0.03	0.01				
N_2	--														1,475	79.00
Phenol	8	0.02											0.2	20 ppmW		
Oil			2073	99.93	6	4.03										
SO_2															392.1	21.00
Total lb-mol/hr	44,404	100.00	2075	100.00	149	100.00	26,256	100.00	4,517	100.00	403	100.00	43,333		1,867	100.00
(gm-mol/sec)	(5,600)		(260)		(20)		(3,310)		(570)		(50)		(5,470)		(235)	
($\times 10^6$ ft^3/hr)	0.014		4.4×10^{-3}		3.6×10^{-4}		9.9		1.7		0.2		0.014		0.7	

Note: All flow rates are rounded to the nearest mol/hr except for sulfur bearing compunds. Volume flow rates are calculated at normal conditions of 14.7 psia, 60°F.

TABLE 3: (CONT'D)

Stream Name	Combined Coal Feed	Coal to U-Gas System	Coal to Hygas System	Ash	By-Product Oil-BTX	Crude Phenol	Anhydrous Ammonia	By-Product Sulfur
Stream Number	A	B	C	D	E	F	G	H
kg/hr	662,139 (6.5% Moisture	139,347 dry	479,853 dry	84,974	18,047	581	5050	25,505
Component	wt% dry	wt% dry	wt% dry		lb-mol/hr	lb-mol/hr	lb-mol/hr	lb-mol/hr
C	69.40	69.40	69.40	10.3% of Carbon	C_6H_6 142	C_6H_6 7	NH_3 653.8	Sulfur 1753.8
H	4.80	4.80	4.80	0.3% of Sulfur	H_2S 2.2	Phenol 7.8		
O	8.71	8.71	8.71		Oil 212			
N	1.35	1.35	1.35					
S	4.20	4.20	4.20					
Ash	11.54	11.54	11.54					
Total	100.00				356.2	14.8	653.8	1753.8
gm-mol/sec					(45)	(2)	(82)	(221)

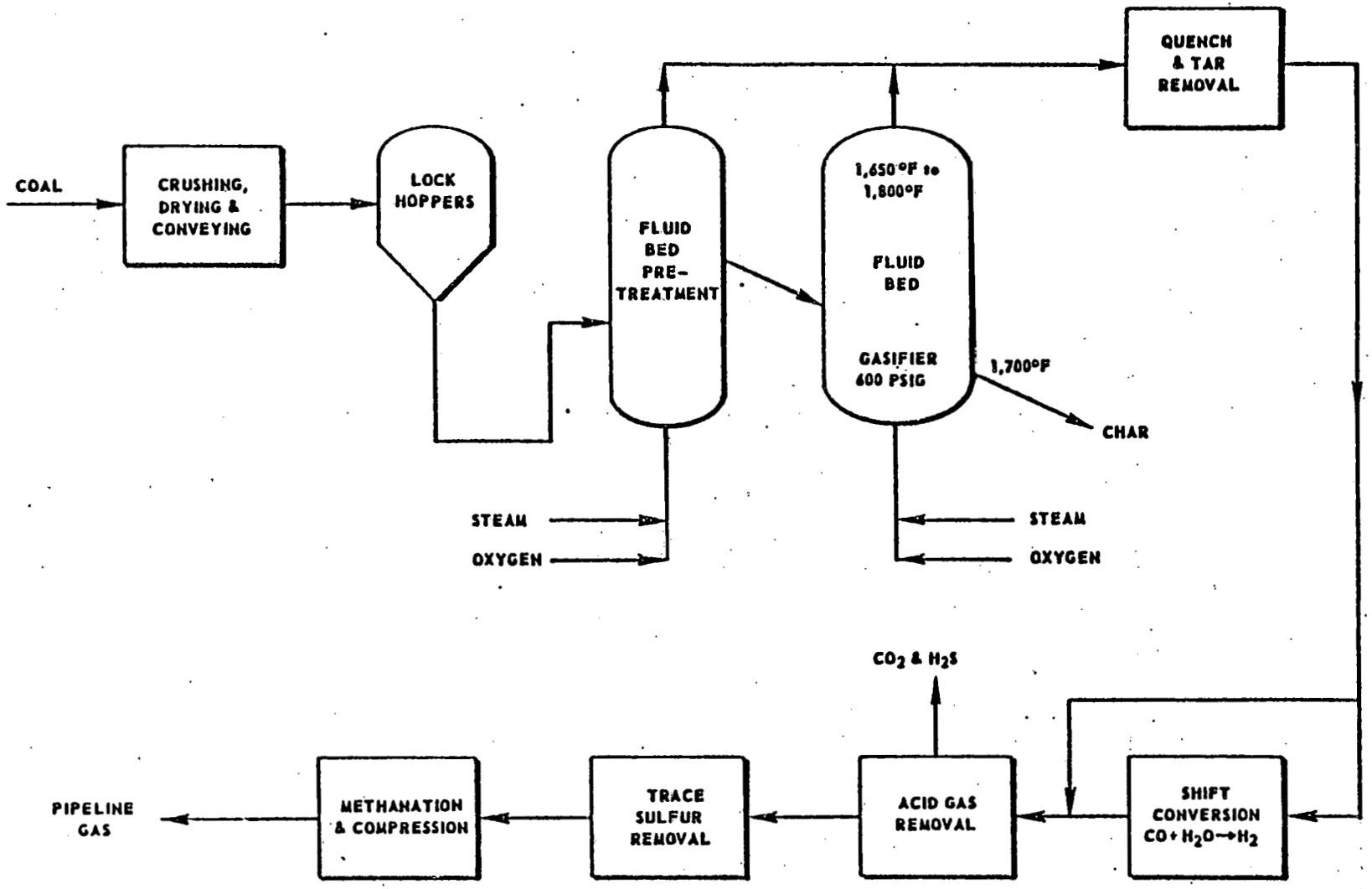


FIGURE 13: SIMPLIFIED PROCESS FLOW DIAGRAM OF THE BUREAU OF MINES SYNTHANE PROCESS

TABLE 4: TYPE, SOURCES, AND TREATMENT OF WASTE STREAMS

SOURCE	WASTE STREAMS	PRINCIPAL COMPONENTS	DISPOSITION	PROCESS/UNIT
FEEDSTOCK STORAGE	LEACHINGS AND RUNOFF	PYRITES	TREATMENT	NEUTRALIZATION WITH LIME
FEEDSTOCK CLEANING	ROCKS AND DEBRIS		DISPOSAL	WATER WASHING, FILTERATION
FEEDSTOCK CRUSHING AND GRINDING	COAL AND SHALE DUST	CARBON PARTICLES	CONTAINMENT AND RECOVERY	CYCLONE SEPARATORS, BAG FILTERS, ENCLOSURE
	COAL FINES	COAL PARTICLES	RECOVERY	
WATER TREATMENT	WASTE WATER	DISSOLVED SOLIDS	TREATMENT/UTILIZATION	LIME SOFTENING, ZEOLITE TREATMENT
STEAM GENERATION	STACK GAS	SO_2,CO,NO_x,PART.	TREATMENT AND DISCHARGE	WELLMAN LORD, LIMESTONE INJECTION, LIME SCRUBBING, CATALYTIC OXIDATION
	ASH	MINERALS, CARBON, S	DISPOSAL/UTILIZATION	LAND FILLING/INCORPORATION IN CONSTRUCTION MATERIALS
STACK GAS CLEANUP	STACK GAS	SO_2,CO,NO_x,PART.	DISCHARGE	
	SULFITE SLUDGE		TREATMENT AND DISPOSAL	
OXYGEN GENERATION	STACK GAS	NITROGEN	DISCHARGE	
GASIFICATION	WASTE WATER	COD,DS,SS,PHENOL, NH_3,CYANIDE,THIO-CN	TREATMENT AND REUSE	CHEVRON, PHENOSOLVAN,PHOSAM,OXIDATION BIOLOGICAL
	TAR		UTILIZATION	CONVERSION TO OIL,DIRECT GASIFICATION DIRECT COMBUSTION,INDIRECT COMBUSTION
	CHAR	MINERALS,CARBON,S	UTILIZATION	GASIFICATION,DESULFURIZATION,DIRECT COMB.
	ASH	MINERALS,CARBON,S	DISPOSAL/UTILIZATION	LAND FILLING/INCORPORATION IN CONSTRUCTION MATERIALS
WASTE WATER TREATMENT	TREATED WATER		REUSE	
	AMMONIA AND PHENOL		RECOVERY	
	HYDROGEN SULFIDE		SULFUR RECOVERY	CLAUS,STRETFORD
	SOLID SLUDGE		TREATMENT AND DISPOSAL	DEWATERING,OXIDATION,LANDFILLING
ASH PONDS	LEACHATE,DRAINAGE	DISSOLVED SALTS	TREATMENT AND REUSE	
SHIFT CONVERSION	ACID GAS	H_2S,COS,CO_2	SULFUR RECOVERY	HOT POTASSIUM CARBONATE,RECTISOL, SULFINOL,MEA,DIPA,ETC.
	WASTE WATER	SULFUR,CO_2	TREATMENT AND REUSE	CHEVERON,PHENOSOLVAN,PHOSAM,OXIDATION, BIOLOGICAL
	HEAT		RECOVERY	
	SPENT CATALYST	COBALT,MOLYBDENUM	TREATMENT AND RECOVERY	OXIDATION
SULFUR RECOVERY	SULFUR		STORAGE	
	TAIL GAS		DISCHARGE OR TREATMENT	INCINERATION,BEAVON,SCOT
	SPENT CATALYST	BAUXITE	DISPOSAL	
METHANATION	WASTEWATER		TREATMENT AND REUSE	CHEVRON,PHENOSOLVAN,PHOSAM,OXIDATION, BIOLOGICAL
	HEAT		RECOVERY	
	SPENT CATALYST	NICKEL	TREATMENT AND DISPOSAL	OXIDATION
FINAL PURIFICATION AND DRYING	VENT GAS	SULFUR DIOXIDE	DISCHARGE	
	WASTEWATER		DISCHARGE,REUSE	
	SPENT IRON OR ZINC OXIDE		DISPOSAL	OXIDATION,IRON AND ZINC OXIDE TREATMENT
BOILERS AND COOLING TOWERS	BLOWDOWN WATER	DISSOLVED SALTS	TREATMENT AND REUSE	
	STEAM		DISCHARGE	
HEAT EXCHANGERS	COOLING WATER	HEAT	UTILIZATION/DISPOSAL	WASTE HEAT RECOVERY
WASHROOMS AND WORK AREAS	SANITARY SEWAGE	ORGANIC/NITROGEN COMPOUNDS	TREATMENT AND DISPOSAL	BIOLOGICAL
LAND SURFACE	SURFACE RUNOFF	SEDIMENT OIL	DISCHARGE	

TABLE 5: SOME TRACE ELEMENTS IN COAL AND RAW GAS FROM GASIFIER (14)

ELEMENT	POUNDS OF TRACE ELEMENT/1,000 TONS OF COAL	
	IN COAL	IN RAW GAS
AS	28.0	27.7
B	204.0	97.3
BE	3.2	0.1
CD	5.0	
CO	19.1	
CR	27.5	
CU	30.3	
F	122.0	121.0
HG	0.4	0.4
MN	98.8	6.8
MO	15.1	13.7
NI	42.1	
PB	69.6	25.2
SB	2.5	1.9
SE	4.2	
SN	9.6	
V	65.4	
ZR	145.0	

TABLE 6: WASTE WATER POLLUTANT COMPOSITIONS (16)

POLLUTANT	AMMONIA LIQUOR MG/LITER	SYNTHANE WASTE WATER MG/LITER
pH	8.3-9.1	7.9-9.3
COD	2,500-10,000	1,700- 43,000
AMMONIA	1,800-4,300	2,500-11,000
CYANIDE	10-37	0.1-0.6
THIOCYANATE	100-1,500	21-200
PHENOLS	410-2,400	200-6,000
SULFIDE	0-50	NOT DETERMINED
ALKALINITY ($CaCO_3$)	1,200-2,700	NOT DETERMINED
SPECIFIC CONDUCTANCE	11,000-32,000	NOT DETERMINED

TABLE 7: INCREASES IN POLLUTANT CONCENTRATIONS ALLOWED OVER BASELINE AIR QUALITY (39 FR 42515)

POLLUTANT	CLASS I, µG/CU.M[1]	CLASS II µG/CU.M	CLASS III[2] µG/CU.M
PARTICULATE MATTER:			
ANNUAL GEOMETRIC MEAN	5	10	60
24-HR MAXIMUM	10	30	150
SULFUR DIOXIDE			
ANNUAL ARITHMETIC MEAN	2	15	60
24-HR MAXIMUM	5	100	260
3-HR MAXIMUM	25	700	1300

[1] µG/CU.M — MICROGRAMS PER CUBIC METER

[2] CONCENTRATIONS OF THESE TWO POLLUTANTS IN AREAS DESIGNATED CLASS III ARE LIMITED TO THE NATIONAL AMBIENT AIR QUALITY SECONDARY STANDARDS (SECTION 52.21(c)(2)(ii) 39 FR 42515).

TABLE 8: AQMA DESIGNATIONS IN 14 STATES

No.	State and AQMA	AQMA Number (By U.S. EPA)	Particulates	Pollutants SO2	CO	Ox	NO2
1	ILLINOIS						
	a) Decatur	075A	X				
	b) Illinois-Indiana-Wisconsin Interstate (Chicago)	067A-I	X	X	X	X	X
	c) Peoria	065A	X	X			
	d) St. Louis Interstate	070A-I	X	X		X	
2	ARIZONA						
	a) Phoenix SMSA	015A	X		X	X	
	b) Tucson SMSA	015B	X			X	
3	COLORADO						
	a) Colorado Springs	038A	X		X		
	b) Colorado-Utah Oil Shale Interstate[1]	035A-I	X	X	X	X	
	c) Metropolitan Denver	036A	X		X	X	X
	d) North Central Colorado	037A	X		X	X	
	e) Pueblo	038B	X		X		
4	INDIANA						
	a) Evansville Interstate	077A-I	X	X			
	b) Illinois-Indiana-Wisconsin Interstate	067B-I	X	X		X	
	c) Indianapolis	080A	X	X		X	
	d) Louisville Interstate	078A-I	X	X			
5	KENTUCKY						
	a) Cincinnati Interstate	079A-I	X			X[2]	
	b) Evansville Interstate	077B-I	X				
	c) Louisville Interstate	078B-I	X	X			
6	MONTANA						
	a) Anaconda-Butte[3]	142A	X	X			
	b) Billings[3]	140A	X	X	X		
	c) Helena[3]	142B		X			
	d) Kalispell[3]	144A	X				
	e) Missoula	144B	X		X		
	f) Southeastern Montana Coal Resource[1]	143A	X	X			
7	NEW MEXICO						
	a) Albuquerque	152A	X		X	X	
	b) Four Corners[1]	014A			X		
	c) Las Cruces	153A	X		X		
	d) Roswell	155A			X		
	e) Santa Fe	157A	X		X		
8	NORTH DAKOTA						
	a) Cass	130A	X				
	b) McLean-Mercer-Oliver[1]	172A	X	X		X	X
9	OHIO						
	a) Akron-Canton	174A	X	X			
	b) Cincinnati Interstate	079B-I	X			X	
	c) Cleveland	174B	X	X			
	d) Columbus	176A	X				
	e) Dayton	173A	X	X			
	f) Mansfield	175A	X				
	g) Steubenville	181A	X	X			
	h) Toledo Interstate	124B-I	X	X			
	i) Youngstown	178A	X				
10	PENNSYLVANIA						
	a) Allegheny County AB[4]	197A	X	X		X	
	b) Allentown-Bethlehem-Easton Interstate	151B-I	X				
	c) Beaver Valley AB	197B	X	X			
	d) Erie AB	178B	X				
	e) Harrisburg AB	196A	X				
	f) Johnstown AB	195A	X				
	g) Lancaster AB	196B	X				
	h) Metropolitan Philadelphia Interstate	045B-I	X	X		X	
	i) Monongahela Valley AB	197C	X	X			
	j) Reading AB	151C	X				
	k) Scranton-Wilkes-Barre AB	151D	X				
	l) York AB	196C	X				
11	SOUTH DAKOTA						
	a) Black Hills	205A	X				
	b) Sioux Falls	087A	X				
12	UTAH						
	a) Colorado-Utah Oil Shale Interstate[1]	035B-I	X	X			
	b) Northcentral Utah	220A	X	X			
	c) Provo	219A	X				
	d) Salt Lake City	220B	X	X			
	e) Southeastern Utah Coal Resource[1]	014B	X	X			
	f) Southwestern Utah Coal Resource[1]	014C	X	X			
	g) Wayne County Coal Resource[1]	014D	X	X			
13	WEST VIRGINIA		NO AQMAS				
14	WYOMING						
	a) Powder River Basin[1]	241A	X			X	
	b) Sweetwater[1]	243A	X	X			

[1]Natural Resource Development Area.
[2]In part of AQMA only.
[3]Industrial Development Area.
[4]Air Basin.

TABLE 9: OUTLINE FOR THE PREPARATION OF AN ENVIRONMENTAL IMPACT STATEMENT

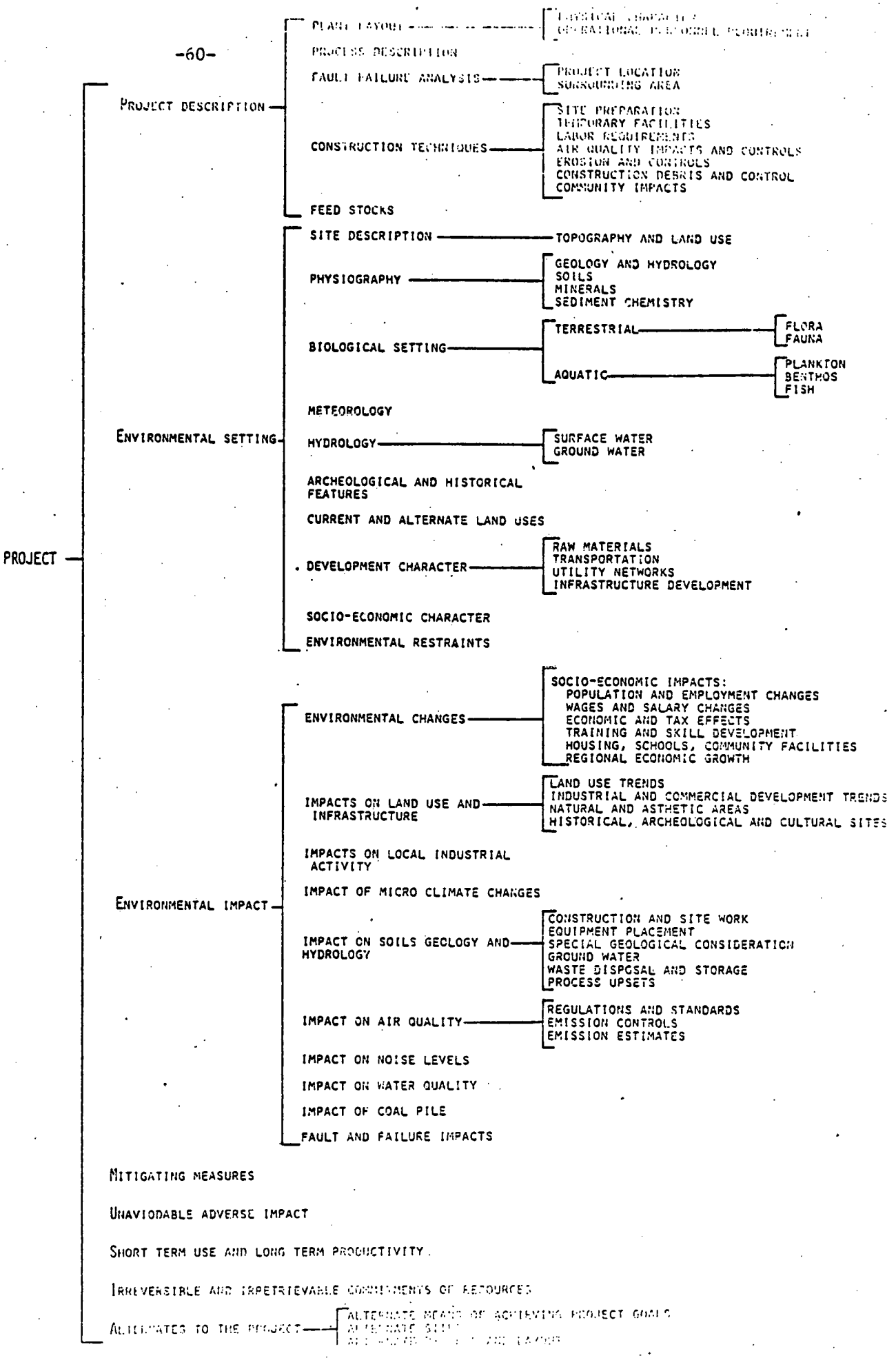


TABLE 10: CONSTRUCTION AND OPERATING PERMITS REQUIRED BY THE AIR POLLUTION CONTROL DIVISION FOR COAL CONVERSION FACILITIES

1. GENERAL OPERATION
   - APC-200 PERMIT TO CONSTRUCT OR OPERATE
   - APC-203 OPERATION DURING STARTUP
   - APC-204 OPERATION DURING MALFUNCTION AND BREAKDOWN
   - APC-100 EPISODE ACTION PLAN

2. GASIFIER
   - COAL CONVEYING AND HANDLING
     - APC-220 PROCESS EMISSION SOURCE
     - APC-260 CONTROL EQUIPMENT
     - APC-103 DRY COLLECTOR WASTE DISPOSAL
   - RESIDUAL LOCK GAS VENT
     - APC-220 PROCESS EMISSION SOURCE
   - WET ASH CHAMBER VENT
     - APC-220 PROCESS EMISSION SOURCE
     - APC-260 CONTROL EQUIPMENT
     - APC-104 WET COLLECTOR WASTE DISPOSAL
   - CRUDE GAS SCRUBBER
     - APC-220 PROCESS EMISSION SOURCE
   - GAS COOLER
     - APC-220 PROCESS EMISSION SOURCE
   - TAR SEPARATOR
     - APC-220 PROCESS EMISSION SOURCE
   - ASH CONVEYING AND HANDLING

3. SHIFT CONVERSION
   - APC-220 PROCESS EMISSION

4. GAS COOLING UNIT
   - APC-220 PROCESS EMISSION SOURCE

5. TAR OIL SEPARATION
   - APC-220 PROCESS EMISSION SOURCE

6. RECTISOL UNIT
   - APC-220 PROCESS EMISSION SOURCE
   - APC-231 REACTOR, DRUM, TOWER OR HEAT EXCHANGER

7. MATHANATION UNIT
   - APC-220 PROCESS EMISSION SOURCE

8. PHENOLSOLVAN UNIT
   - APC-220 PROCESS EMISSION SOURCE
   - APC 231 REACTOR, DRUM, OR HEAT EXCHANGER

9. CLAUS SULFUR PLANT
   - APC-220 PROCESS EMISSION SOURCE
   - APC-231 REACTOR, DRUM, TOWER OR HEAT EXCHANGER
   - APC-260 CONTROL EQUIPMENT (AFTERBURNER, SCRUBBER)
   - APC-104 WET COLLECTOR WASTE DISPOSAL

10. OXYGEN PLANT
    - APC-220 PROCESS EMISSION SOURCE

11. AMMONIA BURNER BOILER
    - APC-240 FUEL COMBUSTION SOURCE

12. COAL-FIRED BOILER
    - APC-240 FUEL COMBUSTION SOURCE
    - APC-260 CONTROL EQUIPMENT

13. OIL-FIRED SUPERHEATER
    - APC-240 FUEL COMBUSTION SOURCE

TABLE 11: ILLINOIS POLLUTION CONTROL BOARD'S RULES AND REGULATIONS GENERALLY APPLICABLE TO COAL CONVERSION FACILITIES

1) RULE 202: VISUAL EMISSION STANDARDS AND LIMITATIONS.
RULE 202(A)(1) WOULD BE APPLICABLE TO FUEL COMBUSTION EMISSION SOURCES WITH HEAT INPUTS OF GREATER THAN 250 MILLION BTU PER HOUR.

2) RULE 203: PARTICULATE EMISSION STANDARDS AND LIMITATIONS.
RULE 203(A) SPECIFIES PARTICULATE EMISSION LIMITATIONS FOR ALL NEW PROCESS EMISSION SOURCES.
RUL3 203(F) IS APPLICABLE TO EMISSIONS OF FUGITIVE PARTICULATE MATTER FROM THE FACILITY.
RULE 203(G) SPECIFIES PARTICULATE EMISSION LIMITATIONS FOR FUEL COMBUSTION EMISSION SOURCES.

3) RULE 204: SULFUR STANDARDS AND LIMITATIONS.
RULE 204(A) SPECIFIES SULFUR DIOXIDE EMISSION LIMITATIONS FOR NEW FUEL INPUTS OF GREATER THAN 250 MILLION BTU PER HOUR. THIS IS THE SAME AS THE U.S. EPA'S NEW SOURCE PERFORMANCE STANDARD FOR SULFUR DIOXIDE APPLICABLE TO POWER PLANTS.
RULE 204(F)(1)(A) SPECIFIES EMISSION LIMITATIONS APPLICABLE TO PROCESS EMISSION SOURCES SUCH AS SULFUR RECOVERY UNITS.

4) RULE 205: ORGANIC MATERIAL EMISSION STANDARDS AND LIMITATIONS.
RULE 205(A) IS APPLICABLE TO STATIONARY TANKS STORING IN EXCESS OF 40,000 GALLONS OF ORGANIC MATERIAL.
RULE 205(B) IS APPLICABLE TO LOADING OF ORGANIC MATERIAL AT A LOADING FACILITY WITH A THROUGHPUT OF GREATER THAN 40,000 GALLONS PER DAY OF ORGANIC MATERIAL.
RULE 205(C) IS APPLICABLE TO ORGANIC MATERIAL - WATER SEPARATION EQUIPMENT LIKE THE ONES THAT COULD BE USED TO SEPARATE THE TARS FROM THE EFFLUENT FROM A TAR SCRUBBER IN A COAL CONVERSION FACILITY.
RULE 205(F) SPECIFIES CERTAIN LIMITATIONS APPLICABLE TO THE USE OF ORGANIC MATERIALS.
RULE 205(F)(2) WOULD BE APPLICABLE TO VAPOR BLOW DOWN SYSTEMS CAPABLE OF RELEASING MERCAPTANS OR HYDROGEN SULFIDE INTO THE ATMOSPHERE (SEE DEFINITION OF "EXCESSIVE RELEASE" SPECIFIED IN RULE 201).
RULE 205(G)(3) IS APPLICABLE TO SETS OF UNREGULATED SAFETY RELIEF VALVES CAPABLE OF CAUSING AN "EXCESSIVE RELEASE".
RULE 205(H) IS APPLICABLE TO EMISSIONS DURING CLEAN-UP OPERATIONS.

5) RULE 206: CARBON MONOXIDE EMISSION STANDARDS AND LIMITATIONS.
RULE 206(A) IS APPLICABLE TO FUEL COMBUSTION SOURCES WITH ACTUAL HEAT INPUTS OF GREATER THAN 10 MILLION BTU PER HOUR.

6) RULE 207: NITROGEN OXIDES EMISSION STANDARDS AND LIMITATIONS.
RULE 207(A) IS APPLICABLE TO NEW FUEL COMBUSTION EMISSION SOURCES WITH HEAT INPUTS OF GREATER THAN OR EQUAL TO 250 MILLION BTU PER HOURS.

TABLE 12: ALLOWABLE SOUND PRESSURE LEVELS

OCTAVE BOND CENTER FREQUENCY (HERTZ)	ALLOWABLE OCTAVE BOND SOUND PRESSURE LEVEL (dB)	
	DAYTIME (7 A.M.-10 P.M.)	NIGHTTIME (10 P.M.-7 A.M.)
31.5	75	69
63	74	67
125	69	62
250	64	54
500	58	47
1000	52	41
2000	47	36
4000	43	32
8000	40	32

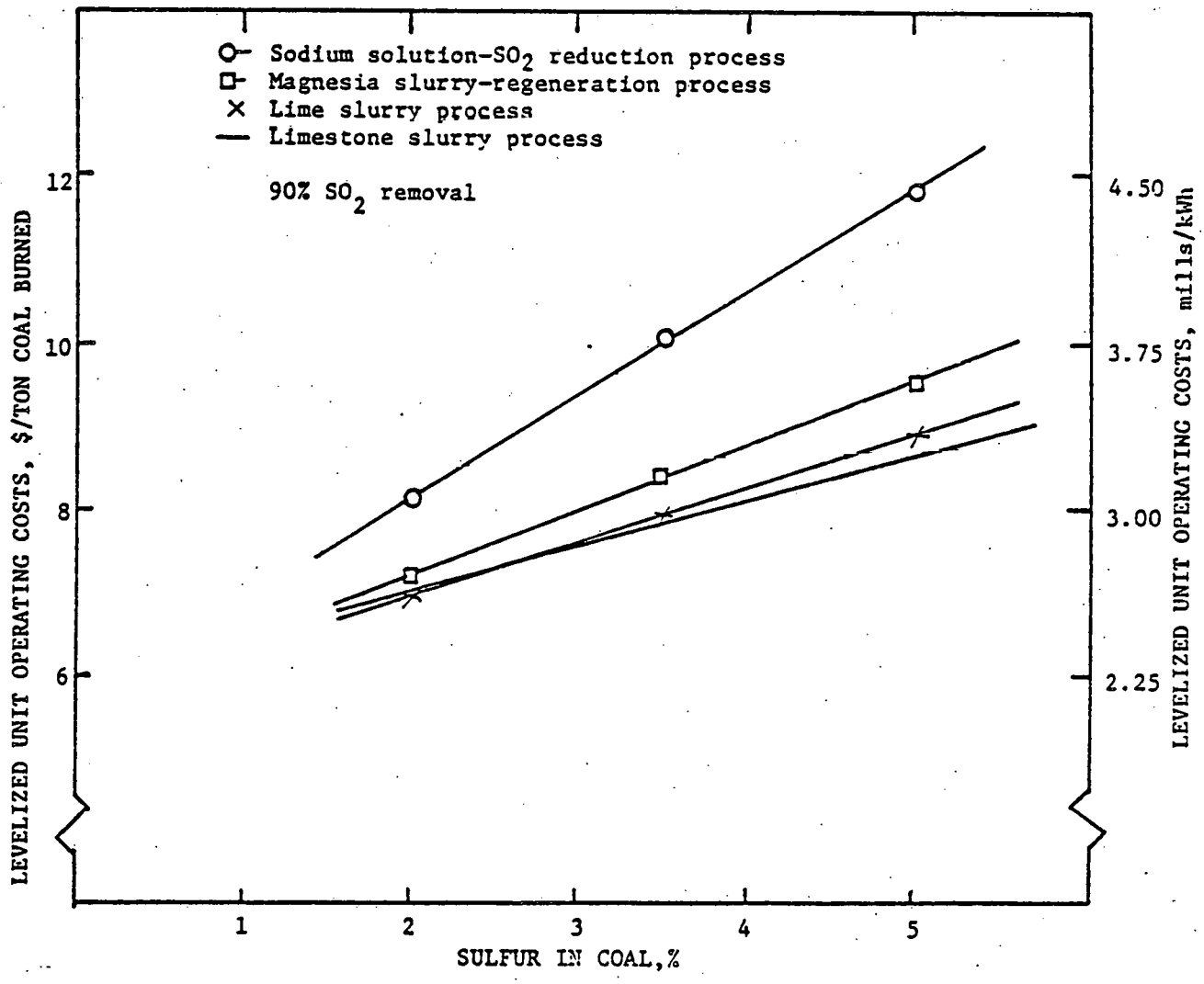


FIGURE 14: EFFECT OF SULFUR CONTENT OF COAL ON LEVELIZED OPERATI"G COSTS FOR A NEW 500-MW BOILER

TABLE 13: INCREMENTAL COST OF POLLUTION CONTROL

PROCESS	FEEDSTOCK QUALITY % S BY WT.	INCREMENTAL ANNUAL COST OF GAS, $/YR. DCF[1]	UTILITY FINANCING	INCREMENTAL COST OF GAS, $/10^6 BTU DCF[1]	UTILITY FINANCING	AVERAGE COST OF SYNTHETIC GAS[7] $/10^9 BTU REFERENCE (29)[8]	REFERENCE (30)
1) SYNTHANE PROCESS							
Direct Char Combustion[2]	1.6	14,444,900	11,135,500	0.19	0.15	1.20	1.23
	2.9	16,633,600	12,968,900	0.22	0.17		
Indirect Char Combustion[3]	1.6	19,356,000	14,517,300	0.25	0.19		
	2.9	22,103,800	16,724,300	0.29	0.22		
2) CO_2 ACCEPTOR PROCESS	0.59	6,151,300	4,260,600	0.075	0.052	1.11	1.42
3) LURGI PROCESS							
Direct Combustion of Coal[4]	0.69	17,077,100	10,807,200	0.19	0.12	1.28	1.77
Indirect Combustion of Coal[5]	0.69	21,971,500	14,086,800	0.24	0.16		
4) HYGAS PROCESS	0.51	16,567,000	11,742,900	0.201	0.142	1.20	1.50
	3.93	16,707,700	11,117,400	0.204	0.136		
5) CCED PROCESS							
Gas Reforming[6]	0.50	3,440,100	2,686,800	0.046	0.036	Not Available	Not Available
	3.70	12,341,500	9,117,200	0.189	0.140		
Char Oxidation[6]	0.50	3,484,000	2,708,000	0.050	0.039		

1 DISCOUNTED CASH FLOW METHOD.

2 CHAR IS BURNED IN A BOILER EQUIPPED WITH FLUE GAS DESULFURIZATION OR PRETREATMENT PROCESSES TO REMOVAL OF SULFUR COMPOUNDS.

3 CHAR IS GASIFIED TO A LOW BTU GAS, AND AFTER RAW GAS TREATMENT, THE GAS IS BURNED IN A BOILER.

4 COAL IS BURNED IN A BOILER EQUIPPED WITH FLUE GAS DESULFURIZATION SYSTEMS.

5 COAL IS GASIFIED AND, AFTER RAW GAS TREATMENT, THE GAS IS BURNED IN A BOILER.

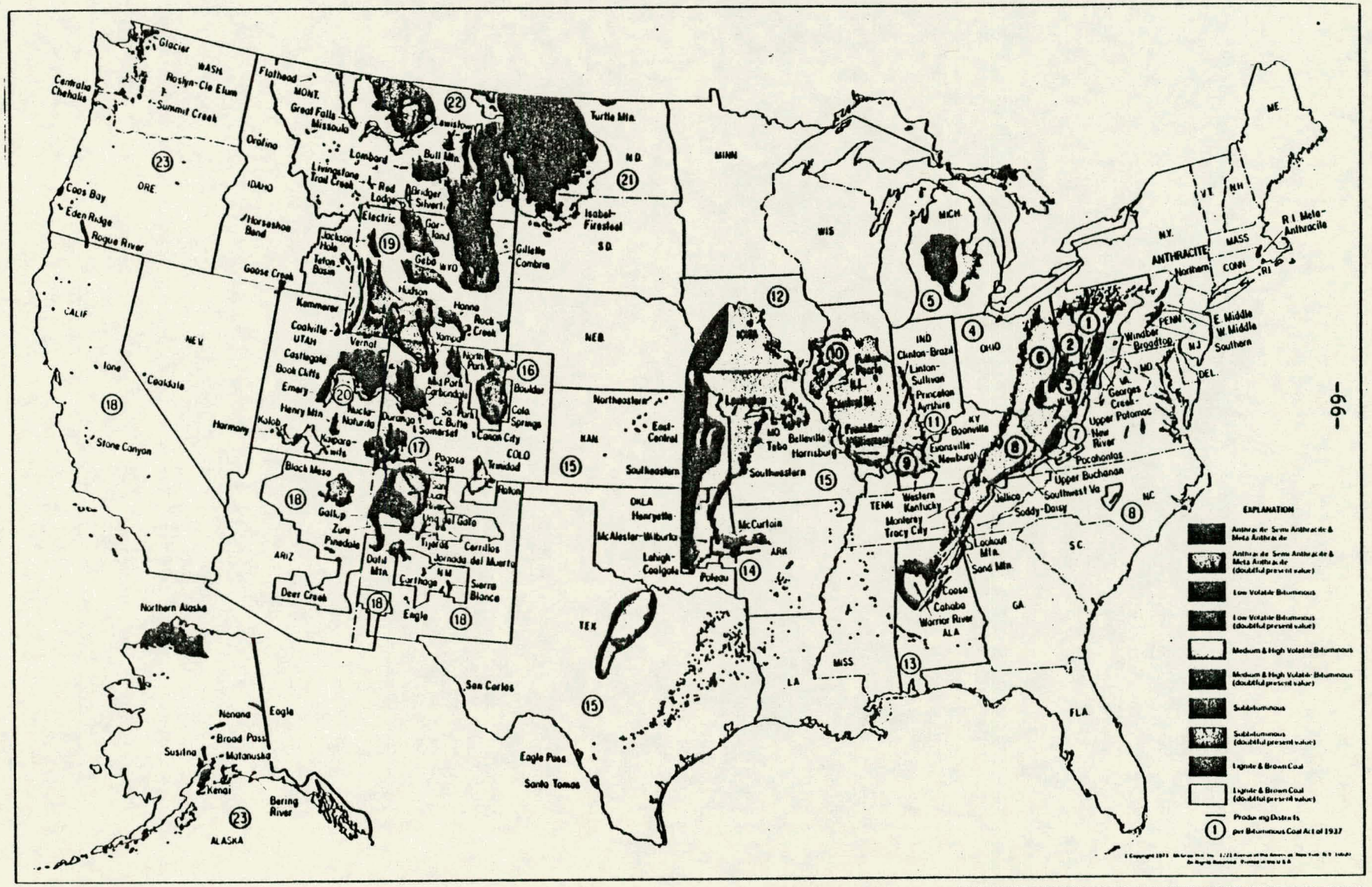
6 METHOD OF HYDROGEN GENERATION.

7 OBTAINED FROM REFERENCES (29) AND (30). 1972 COST DATA GIVEN BY HITTMAN (29) AND REFERENCE (30) WAS MULTIPLIED BY 1.25 ASSUMING A 25% INCREASE IN COST OF SYNTHETIC GAS BY 1974 OVER THE 1972 ESTIMATES.

8 COAL COST AT $6 PER TON.

9 COAL COST AT $8 PER TON.

KEYSTONE'S MAP OF THE COAL FIELDS OF THE UNITED STATES



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Appendix A

THE IMPACT STATEMENT

FOLLOWING ARE EXCERPTS FROM A LETTER WRITTEN TO PRESIDENT FRANKLIN PIERCE IN 1855 BY A CHIEF OF THE DUWAMISH INDIANS NAMED SEALTH. FROM A CORRUPTION OF HIS NAME WE DERIVE SEATTLE, THE CITY BUILT ON THE LAND HIS TRIBE INHABITED.

"We know that the white man does not understand our ways. One portion of the land is the same to him as the next, for he is a stranger who comes in the night and takes from the land whatever he needs. The earth is not his brother, but his enemy, and when he has conquered it, he moves on. He leaves his fathers' graves, and his children's birthright is forgotton. The sight of your cities pains the eyes of the redman. But perhaps it is because the redman is a savage and does not understand.

"There is no quiet place in the white man's cities. No place to hear the leaves of spring or the rustle of insect's wings. But perhaps because I am a savage and do not understand, the clatter only seems to insult the ears. The Indian prefers the soft sound of the wind darting over the face of the pond, and the smell of the wind itself cleansed by a mid-day rain, or scented with a pinon pine. The air is precious to the redman. For all things share the same breath - the beasts, the trees, the man. The white man does not seem to notice the air he breathes. Like a man dying for many days, he is numb to the stench.

"What is man without the beasts? If all the beasts were gone, men would die from great loneliness of spirit, for whatever happens to the beasts also happens to man. All things are connected. Whatever befalls the earth befalls the sons of the earth.

"It matters little where we pass the rest of our days; they are not many. A few more hours, a few more winters, and none of the children of the great tribes that once lived on this earth, or that roamed in small bands in the woods, will be left to mourn the graves of a people once as powerful and hopeful as yours.

"The whites, too, shall pass -- perhaps sooner than other tribes. Continue to contaminate your bed, and you will one night suffocate in your own waste. When the buffalo are all slaughtered, the wild horses all tamed, the secret corners of the forest heavy with the scent of many men, and the view of the ripe hills blotted by talking wires, where is the thicket? Gone. Where is the eagle? Gone. And what is to say goodby to the swift and the hunt, the end of living and the beginning of survival? We might understand if we knew what it was that the white man dreams, what hopes he describes to his children on the long winter nights, what visions he burns into their minds, so they will wish for tomorrow. But we are savages. The white man's dreams are hidden from us."

WORKING IN A MOST UNSCIENTIFIC WAY, WITH NOTHING BUT INTUITION AND LOVE TO GUIDE HIM IN THE COLLECTION AND INTERPRETATION OF HIS RANDOM DATA, CHIEF SEALTH WROTE AN ENVIRONMENTAL IMPACT STATEMENT WHICH -- 120 YEARS AGO -- EMBODIED THE BASIC ECOLOGICAL INSIGHT; ALL THINGS ARE CONNECTED. WHATEVER BEFALLS THE EARTH BEFALLS THE SONS OF THE EARTH.

FROM AN ADDRESS BY
RUSSELL PETERSON, CHAIRMAN
COUNCIL ON ENVIRONMENTAL QUALITY

APPENDIX B

ENVIRONMENTAL ASPECTS OF COAL CONVERSION PLANT SITING AND COST OF POLLUTION CONTROL

By
G. NARENDER REDDY

77-15 Read
0480-15 x 77-13

TABLE B-1: WATER QUALITY STANDARDS IN ILLINOIS

No.	Water Quality Criteria	WATER USE CLASSIFICATION			
		General	Public Or Food Processing Water Supply[1]	Secondary Contact And Indigenous Aquatic Life Standards	Lake Michigan
1	General Quality	No Floating Debris, Etc.	*	*	*
2	pH	6.5-9.0	*	6.0-9.0	7.0-9.0
3	Phosphorus AsP, mg/l.	0.05	*	*	*
4	Dissolved Oxygen, mg/l.	> 6.0 for 16 hrs. or > 5.0 any time	* *	> 3.0 for 16 hrs. or >2.0 any time [2]	> 90% of saturation
5	Radioactivity, Gross Beta pCi/l. Radium 226 Strontium 90	100 1 2	* * *	- - -	* * *
6	Chemical Constituents, mg/l.				
	a) NH_3 - Nitrogen (as N)	1.5	*	-	0.02
	b) Arsenic (Total)	1.0	0.01	-	**
	c) Barium (Total)	5.0	1.0	-	**
	d) Boron (Total)	1.0	*	-	*
	e) Cadmium (Total)	0.05	0.01	-	**
	f) Chloride	500.0	250.0	-	12.0
	g) Chromium (Total Hexa.)	0.05	*	-	*
	h) Chromium (Total Tri)	1.0	*	-	*
	i) Copper (Total)	0.02	*	-	*
	j) Cyanide	0.025	0.01	-	**
	k) Fluoride	1.4 or 5.0	*	-	*
	l) Iron (Total)	1.0	0.3	-	**
	m) Lead (Total)	0.1	0.05	-	**
	n) Manganese (Total)	1.0	0.05	-	**
	o) Mercury (Total)	0.0005	*	-	*
	p) Nickel (Total)	1.0	*	-	*
	q) Phenols	0.1	0.001	-	**
	r) Selenium (Total)	1.0	0.01	-	**
	s) Silver (Total)	0.005	*	-	*
	t) Sulfate	500.0	250.0	-	24.0
	u) Total Dissolved Solids	1000.0	500.0	-	180.0
	v) Zinc	1.0	*	-	*
	w) Carbon Chloroform Extract (CCE)	-	0.2	-	**
	x) Methylene Blue Active Substance (MDAS)	-	0.5	-	**
	y) Nitrates Plus Nitrites	-	10.0	-	**
	z) Oil (Hexane Soluble or Equivalent)	-	0.1	-	**
	aa) Phosphorous (as P)	-	-	-	0.007
7	Fecal Coliform, geo. mean per 100 ml.	5 Samples in 30 days; < 200.	*	5 Samples in 30 days; < 1000.	5 Samples in 30 days; < 20.
		10% of Samples in 30 days; < 400.	*	10% of Samples in 30 days; < 2000.	*
8	Toxic Substances	< 1/10 of 48-hr. TLm.	*	-	*
9	Temperature, °F.	Variable.	*	< 93°F 95% of the time or Max. 100°F.	Variable.

* Same standards as shown under 'General' apply.

** Standards shown for Lake Michigan are in addition to those shown for Public Or Food Processing Water Supply.

1 Waters after treatment have to meet "Public Health Service Drinking Water Standards - 1962".

2 After December 31, 1977; > 4.0 at any time.

TABLE B-7: WATER QUALITY STANDARDS IN NEW MEXICO[1]

No.	Water Quality Criteria	Sections of Rio Grande Basin[2]					San Francisco River Basin		
		1	3	6	10	13	1	2	3
1	Dissolved Oxygen, mg/l.	>5.0	>5.0	>6.0[4]	>6.0[4]	>6.0[4]	>5.0	>6.0	>6.0[4]
2	pH	6.6-8.8	6.6-8.8	6.6-8.8	6.6-8.8	6.6-8.8	6.6-9.0	6.6-8.8	6.6-8.8
3	Temperature, °F.	<93.2	<77.0	<68.0	<68.0	<68.0	<90.0	<68.0	<68.0
4	Fecal Coliform Bacteria Monthly Log Mean/100 ml. 10% of Samples/100 ml.	<1,000 <2,000	<1,000 <2,000	<100 <200	<100 <200	<100 <200	<1,000 <2,000	<1,000 <2,000	<100 <200
5	Total Dissolved Solids[3], mg/l.	<2,000	--	--	--	--	--	--	--
6	Sulfate[3], mg/l.	< 500	--	--	--	--	--	--	--
7	Chlorides[3], mg/l.	< 490	--	--	--	--	--	--	--
8	Total Phosphrous, mg/l.	--	--	<0.1	<0.1	<0.1	--	--	<0.1
9	Organic Carbon, mg/l.	--	--	<7.0	<7.0	<7.0	--	--	<7.0

1 Standards for only two of six basins shown in this table. Also not included are the general standards.

2 Standards for only 5 of 15 section shown here.

3 Monthly average concentration.

4 Or 85% of saturation.

TABLE B-8: WATER QUALITY STANDARDS IN NORTH DAKOTA

No.	Water Quality Criteria	Class I	Class IA	Class II	Class III
1	Chemical Constituents, mg/l.				
	a) Ammonia	1.00	*	1.50	2.00
	b) Arsenic	0.05	*	*	0.10
	c) Barium	1.00	*	*	*
	d) Boron	0.50	*	*	0.75
	e) Cadmium	0.01	*	*	*
	f) Chlorides	100.00	175.00	250.00	*
	g) Chromium (Tri or Hexa)	0.05	*	*	*
	h) Copper	0.05	*	0.10	*
	i) Cyanides	0.01	*	*	0.10
	j) Lead	0.05	*	*	0.10
	k) Nitrates	4.00	*	5.00	*
	l) Phenols	0.01	*	*	*
	m) Phosphates (as P)	0.10	*	0.20	*
	n) Selenium	0.01	*	*	*
	o) Zinc	0.05	*	1.00	*
2	Total Dissolved Solids, mg/l.	500.00	1,000.00	*	1,000.00
3	Fecal Coli, per 100 ml.	200.00[1]	*	*	1,000.00[1]
		400.00[2]	*	*	2,000.00[4]
4	Coliform, per 100 ml.	1,000.00[1]	*	*	*
		2,000.00[3]	*	*	*
5	Turbidity, Jackson Units	10.00	*	25.00	50.00
6	Color	15.00	*	25.00	40.00
7	pH	7.0-8.5	*	6.0-9.0	*
8	Dissolved Oxygen, mg/l.	> 5.0	*	*	*
9	Temperature, Maximum	85.0°F	*	90.0	*
	Maximum Increase	5.0°F	*	5.0	*
10	Radiological Constituents, pCi/l.				
	a) Iodine - 131	5.0	*	*	*
	b) Radium - 226	1.0	*	*	*
	c) Strontium - 89	100.0	*	*	*
	d) Strontium - 90	10.0	*	*	*
	e) Tritium	3,000.0	*	*	*

1 Geometric mean of a Min. of not less than 5 samples obtained during separate 24-hr. periods for any 30-day period.

2 10% of total samples.

3 20% of samples during any 30-day period.

4 10% of samples during any 30-day period.

TABLE B-10: WATER QUALITY STANDARDS IN PENNSYLVANIA

	Water Quality Criteria	1	2	3	4	5	6	7	8	9	10	11
a.	pH	6.0-815	6.5-8.5	7.0-9								
b.	Dissolved MDA[1] Oxygen, mg/l MV[2]	6.0 5.0	5.0 4.0	5.0 6.5[3]	3.5 6.5[3]	5.0 4.0-5.0[3]	- 7.0	5.0[4] 4.0[4]	5.0[5]	7.0 6.0		
c.	Iron total mg/l dissolved	1.5 -	- 0.3									
d.	Temperature °F	5° above 58°, 2° per hr.	5° above 87°, 2° per hr.	-	-	Month by temperature,	5° above 74°, 2° per hr. 5° above 87°, 2° per hr.	5° above specified, max. 86°	5° above ambient until 50° 2° between 50° and 58°	Heat dissipation area is specified	Heat dissipation area is specified	Heat dissipation area is specified
e.	Dissolved solids, mg/l	500[6] 750[7]	- 1,500[7]	1.33 times background or 500	1.33 times background							
f.	Bacteria; fecal coliform density per 100 ml.	5 sample geometric mean <200	5000[6] or as specified	5000[6]	770 max. geometric mean							
g.	Turbidity, units	30-5/30 to 9/15 or 40[6] max. 150	40[6] max. 150	100	40-5/15 to 9/15 100-9/16 to 5/14	10[6] max. 150	20[6] max. 150	30[6] max. 150				
h.	Threshold odor number	NOT MORE THAN 24 AT 60°C										
i.	Alklinity, mg/l	>20	20-120	20-100								
j.	Methylene blue active substance (MBAS), mg/l	<0.5	<1.0									
k.	Total manganese, mg/l	<1.0										
l.	Fluoride, mg/l	<1.0										
m.	Cyanide, mg/l	<0.025										
n.	Sulfate, mg/l	<250[8]										
o.	Chlorides, mg/l	<150	<250	<200	50[9]							
p.	Phosphate (total soluble $AsPO_4$), mg/l	<0.1[8]	<0.3[8]	<0.4[8]	<0.03[8]							
q.	Phenol, mg/l	<0.005	<0.02									
r.	Color, units	<50										
s.	Copper, mg/l	<0.02	<0.10									
t.	Zinc, mg/l	<0.05										
u.	Ammonia-nitrogen, mg/l	<0.5	<1.5									
v.	Radioactivity, alpha pc/l beta	max. 3 max. 1000										
w.	Hardness, mg/l	150[6]	95[6]									

[1]Minimum daily average
[2]Minimum value
[3]Seasonal average or seasonal minimum value
[4]For epilimnion of lakes, ponds and impoundments
[5]For lakes, ponds and impoundments
[6]Monthly average
[7]Maximum allowable value
[8]Or natural level whichever is greater
[9]Maximum 15-day mean

NOTE: For each water body, the specific use and criteria (e.g., a_1, b_4, c_2, etc.) are given

TABLE B-15: AIR POLLUTION EMISSION STANDARDS IN FOUR STATES

NO.	POLLUTANT AND SOURCE	ILLINOIS	PENNSYLVANIA	NEW MEXICO[8]	NORTH DAKOTA
1	VISIBLE EMISSIONS;	-	$<$ 20% Opacity	-	
	a) Fuel Combustion	$\leq$ 20% Opacity	-	-	No. 1 Ringelman
	b) Process	$\leq$ 30% Opacity	-	-	No. 1 Ringelman
2	PARTICULATES;				
	a) Fuel Combustion	0.1 lb/mmBtu[1]	0.1 lb/mmBtu	0.03 lb/mmBtu[9]	$E=0.811(H)^{-0.131}$
	b) Process	$P < 450$[2]	0.2 gr./dry SCF	-	$P \leq 30$[2]
		$E=2.54(P)^{0.534}$	or		$E=4.10(P)^{0.67}$
		$P \geq 450$[2]			$P > 30$[2]
		$E=24.8(P)^{0.16}$	$A=0.76(E)^{0.42}$		$E=55(P)^{0.11}-40$
			(See Footnote 5)		
3	SULFUR DIOXIDE;				
	a) Fuel Combustion, lb/mmBtu	1.2-Solid Fuel	For Heat Inputs	0.16 - Gas	3.0
		0.8-Resid. Oil	$\geq$ 50 BUT $<$ 2,000[6];	0.34 - Coal	
		0.3-Dist. Oil	1.0-0.6	0.34 - Oil	
			For $>$ 2,000; 0.6		
	b) Process	2,000 ppm	500 ppm[7]	-	Variable
4	HYDROCARBONS;				
	a) Storage	$>$40,000 gal. Tanks; Pressure Tank Or Floating Roof Or 85% Control	$>$40,000 gal. Tanks; Pressure Tank Or Floating Roof Or Vapor Recovery	Floating Roof Tank or Vapor Recovery	$>$65,000 gal. Tanks; Floating Roof Or Vapor Recovery
	b) Loading	Throughput $>$40,000 gal/day 8 lb/hr.[3]	Throughput $>$20,000 gal/day Vapor Recovery	Vapor Recovery Or Containment	Submerged Loading Or Vapor Control
	c) Organic Material - Water Separation	$\geq$ 200 gals; 85% Control[3]	$\geq$ 200 gal; Floating Roof	-	-
				-	-
	d) Use of Organic Material	8 lb/hr. or 85% Control[4]	-	-	-
5	CARBON MONOXIDE,				
	a) Fuel Combustion	200 ppm	-	-	-
	b) Incinerators	500 ppm	-	-	-
6	OXIDES OF NITROGEN;				
	a) Fuel Combustion				
	Gas	0.2 lb/mmBtu	-	0.2 lb/mmBtu	-
	Oil	0.3 lb/mmBtu	-	0.3 lb/mmBtu	-
	Dual - Gas & Oil	0.3 lb/mmBtu	-	-	-
	Solid	0.7 lb/mmBtu	-	0.7 lb/mmBtu	-
	b) Process	5.0 lb NO_2 per ton acid used or produced	-	-	-
7	COAL BRIQUETING OPERATIONS				
	a) Particulates	-	-	0.03 gr./SCF	-
8	GASIFICATION PLANT				
	a) Sulfur Compounds Hydrogen Sulfide	-	-	$\leq$ 100 ppm	-
				$\leq$ 10 ppm	-
	b) Hydrogen Cyanide	-	-	$\leq$ 10 ppm	-
	c) Hydrogen Chloride	-	-	$\leq$ 5 ppm	-
	d) Particulates	-	-	0.3 gr/SCF	-
	e) Ammonia	-	-	$\leq$25 ppm	-
	f) Sulfur	-	-	0.008 lb/mmBtu[10]	-

NOTE: Emission standards only for new sources and those fuel combustion sources with heat input of greater than 250 million BTU per hour.

Footnotes continued on next page.

TABLE B-15: (FOOTNOTES)

1 Denotes pounds per million Btus.

2 Process weight rate -- P in tons/hr.; emission rate -- in lb/hr.

3 If no odor nuisance exists - limitation is applicable only to organic material with vapor pressure of $\geq$ 2.5 psia at 70°F.

4 If no odor nuisance exists - limitation is applicable only to photochemically reactive organic materials.

5 A -- allowable emissions in lb/hr; E -- emission index = FxW lb/hr; F -- process factor (given in regulations for 30 different processes) in lb/unit; and W -- charging rate in units/hr.

6 Heat inputs given in million Btu per hour and emissions in lb per million Btu.

7 Separate sulfur dioxide regulation exists for elemental sulfur recovery plants.

8 New Mexico is presently the only state that has emission standard specifically applicable to coal conversion facilities.

9 For gas fired boilers.

10 Elemental sulfur emission standards based on total heat input to gasifier.

to clean up.

Coal at present forms the world's largest reserve of concentrated organic raw materials and for many essential industries from dyes, medicines, pesticides, and elastomers to modern plastics shares to a lesser extent with petrochemicals the furnishing of the basic raw materials. However, it is our fundamental fuel and will remain so for a long time to come.

Coal is a heterogeneous substance, both physically and chemically, and this heterogeneity can be attributed to the fact that the parent material itself was highly heterogeneous , originating from different plant, and to a smaller extent animal, forms mixed with inorganic sediments. Its structure consists of aromatic ring clusters generally no larger than 6 or 7 rings tied together by bridge linkages such as methylene chains of 1 to 4 carbon atoms, ethers, sulfides, disulfides, and biphenyl groups (1)*. A representation of the bituminous coal structure is shown in Figure 1 (2). The solvent extraction, volatilization, oxidation and hydrogeneration of coal with a complicated structure such as this is difficult because the reactive linkages are accessible only with difficulty. Inquisitive scientists have been attempting to understand the reactions of coal for centuries, and in recent years interest has been renewed because additional sources of energy and chemicals are needed. The conversion of coal to heat, and liquid or gaseous fuels is extremely complicated, and for simplicity the reaction of carbon (i.e., β-graphite) as shown in Figure 2 can be used to gain a better understanding of the process involved (3). Reactions 1 to 6 involve the direct reaction of coal, 7 to 9 involve the products of coal and are useful in converting coal to gaseous fuels, and reactions 10 and 11 are combinations of other simple reactions.

Figure 3 shows some of the products that could be derived from coal involving the destructive distillation, pyrolysis, oxidation, gasification or hydrogenation and recovery of by-products. These products have generally been derived so far from the distillation of coal tar and light oils obtained as a by-product of the coking operation (4). However, it should be recognized that hydrogenation and oxidation of coal also would produce similar chemicals and in some cases produce char as a by-product.

In the following chapters, I will describe the status of coal conversion technology and potential health effects due to exposure to products from coal conversion plants very briefly. I will then describe the environmental rules and regulations applicable to coal conversion facilities in 14 states that have potential sites for the location of such facilities and costs of compliance with these regulations.

*Numbers in parentheses refer to bibliography at the end of this report.

THE SRC-II PROCESS

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THE SRC-II PROCESS

SUMMARY

Continuing laboratory, pilot plant and engineering work on the Solvent Refined Coal process, carried out by Gulf Oil Corporation under the sponsorship of the Energy Research and Development Administration, has resulted in the development of an improved version of the process, designated as SRC-II. The primary product of the SRC-II process is a low-sulfur distillate fuel oil containing less than 0.4% sulfur, in contrast to the solid fuel produced in the original version of the process. Additional products of the process are pipeline gas, LPG, a light distillate liquid (C_5-400°F.), phenols and ammonia.

In the SRC-II process, a portion of the product slurry is recycled as a "solvent" for the coal, rather than using an all-distillate liquid for this purpose. The major result of this is to significantly increase the conversion of dissolved coal to liquid and gaseous products. The increased conversion is sufficient to permit recovery of the liquid fuel oil product by vacuum distillation, while the bottoms from the vacuum distillation step is sent to a gasifier for production of hydrogen. The quantity of organic material remaining in the vacuum bottoms is controlled so that it is just sufficient to produce the hydrogen required for the process. In this way, conversion of most of the undissolved coal to liquids makes it possible to avoid complex, unproven process steps such as filtration, mineral residue drying and product solidification, which still require extensive development work.

Upon completion of some additional pilot scale development work, the process would be ready to proceed to the demonstration phase, using commercial size equipment. It is anticipated that the demonstration plant would be designed as one complete module of a future commercial plant, so that the equipment and facilities could eventually be utilized by expansion of the demonstration plant to a full size commercial plant.

THE SRC-II PROCESS

I. Introduction

Solvent Refined Coal technology has been under development by Gulf Oil Corporation, under sponsorship of the Energy Research and Development Administration, for some years. Much of the current work on the project is being carried out by The Pittsburg & Midway Coal Mining Co., a Gulf subsidiary, in a 50 T/D pilot plant at Ft. Lewis, Washington. The overall process design for the pilot plant, plus early start-up experience, was discussed in a paper[1] published about one year ago. Data from subsequent operations were discussed in a presentation at the Second Annual Conference on Gasification and Liquefaction at the University of Pittsburgh.[2] The pilot plant has continued to operate with considerable success since that time. Much of the recent pilot plant work has been directed toward development of a process for production of a solid Solvent Refined Coal product for use in power plants, but experimental work has also been progressing on certain variations of the SRC process. One such variation which appears to be particularly promising has been designated as the SRC-II process.

In the SRC-II process a portion of the product slurry is recycled as a "solvent" for the coal rather than using an all-distillate liquid as in the original version of the process. In addition to dissolving the coal, the process also converts much of the dissolved coal to distillate liquid and gaseous products. The major liquid product is separated in a vacuum distillation step, while the residue from vacuum distillation is fed to a gasifier for hydrogen production. The quantity of organic material remaining in the vacuum residue is controlled so that it is just sufficient to produce the hydrogen required for the process. Thus, conversion of a large part of the dissolved coal to liquids makes it possible to eliminate the filtration step, the mineral residue drying step, and the product solidification step of the SRC-I process. Consequently, the SRC-II

process is less complex and, of even greater importance, has fewer uncertain process steps requiring substantial development work.

II. Process Characteristics

A. Process Chemistry

The major reactions occurring in the Solvent Refined Coal process are:

1. Solution of the coal in the solvent, accompanied by formation of a gel as the coal begins to dissolve.
2. Hydrogenation and depolymerization of the gel and of the coal in solution.
3. Hydrocracking of the dissolved coal to lower molecular weight hydrocarbons ranging from liquid middle distillates to methane.
4. Removal of organic sulfur by hydrogenation to hydrogen sulfide.
5. Conversion of iron pyrites to ferrous sulfide.
6. Hydrogenation of oxygen in the coal to water.

The SRC-II process involves all of the above reactions, but the major difference between it and the original version of the Solvent Refined Coal Process (SRC-I) is that reaction No. 3 is much more extensive in the SRC-II process. For example, in the SRC-I process, typical yields are about 6% light hydrocarbon gases and 15% hydrocarbon liquids boiling below 500°F., with no net yield of 500-800°F. solvent. (The yield of this fraction is merely sufficient to make up for the quantity in this boiling range hydrocracked to gases and lighter liquids.) The yield of solids-free 850°F.+ residue (Solvent Refined Coal) is usually about 60% by weight of dry coal in the SRC-I Process.

In the SRC-II process, however, the yield of solids-free 850°F.+ residue is typically only about 20% by weight of the coal, and this fraction goes to the gasifier with the mineral residue. Except for the undissolved coal solids (mineral residue) all of the coal is converted to liquid and gaseous products. In addition, conversion of sulfur to hydrogen sulfide and conversion of oxygen to water (reactions 4 and 6 above) are somewhat more extensive in the SRC-II process.

The more extensive hydrogenation and hydrocracking reactions are attained in several ways. First, recycle of the product slurry results in return of much of the dissolved coal (solids-free vacuum residue) to the reaction zone, so that its effective residence time is significantly increased even at the same per-pass residence time. In addition, however, the per-pass residence time is higher in the SRC-II process. A third reason for the increased conversion is that the recycle of slurry results in a greater concentration of mineral residue in the dissolver. Since the mineral residue is known to be a catalyst for the SRC reactions, the reaction rates are significantly increased.[3] Thus, the increased conversion is a result of both an increased effective residence time and increased rates of reaction.

While the specific reasons for the catalytic effect of the mineral residue have not yet been clearly defined, it appears that the catalytic activity results from both iron (sulfides or oxides) and the silica-alumina-magnesia, etc., which constitutes most of the mineral matter in the coal.[4] The fact that these minerals are present in very small particles (1 to 20 microns) undoubtedly contributes to their effectiveness.

One notable feature of the process is that the equilibrium activity of the mineral residue "catalyst" is sufficiently great so that the quantity of vacuum residue remaining is no more than that necessary to produce the hydrogen required for the process. This makes it possible to eliminate the solids separation step, as well as mineral residue drying and product solidification, and thereby greatly simplify the process. By effective utilization of the catalytic activity of the coal solids, the required conversion can be attained without the use of a separate catalyst. This is a major advantage of the SRC-II process, since the use of an added catalyst would increase costs, as well as to make the process more complex and add a degree of uncertainty to scale-up problems.

Typical yields from laboratory experimental work on the SRC-II hydrogenation-hydrocracking step are given in Table I. Substantial yields of both gas and liquid hydrocarbons are obtained, while the solids-free vacuum residue (850°F.+) yield is about 20% by weight of dry coal. The hydrogen consumption is 4.4% by weight of dry coal (about 17,000 SCF/ton).

B. Process Description

A schematic flow diagram of the SRC-II process as it is currently visualized for a large scale demonstration or commercial plant is shown in Figure 1. Raw coal is pulverized and dried in the coal preparation area, then mixed with hot recycle slurry "solvent" from the process. The coal-recycle slurry mixture (in the range 1.5-2.5 parts of slurry to one part of coal) is pumped, together with hydrogen, through a fired preheater, to a reactor, or "dissolver." The ratio of hydrogen to coal is usually about 25,000 SCF/Ton.

TABLE I

TYPICAL LABORATORY YIELDS FOR SRC-II PROCESS

Coal Feed: Western Kentucky Bituminous

Yields: % by Wt. of Dry Coal		
C_1-C_4 Gases		16.6
Liquids		43.7
C_5-380°F.	11.4	
380-480°F.	9.5	
480-850°F.	22.8	
850°F.+ Residue (Solids-Free)		20.2
Undissolved Organic Matter		3.7
Ash		9.9
H_2S		2.3
$CO + CO_2$		1.1
H_2O		7.2
Total		104.7
Hydrogen Consumption: % by Wt.		4.7

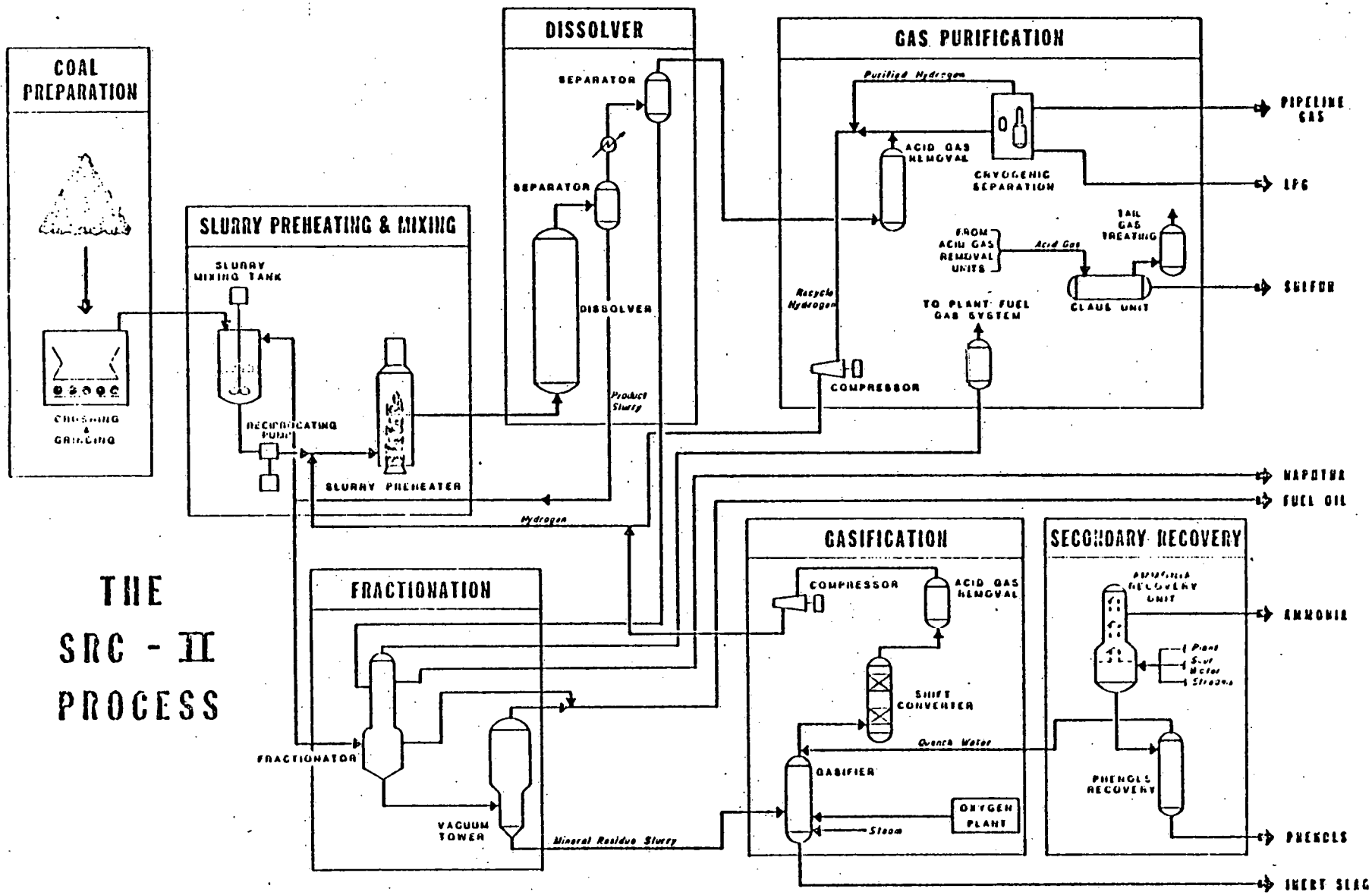

COAL PREPARATION
CRUSHING & GRINDING
SLURRY PREHEATING & MIXING
SLURRY MIXING TANK
RECIPROCATING PUMP
SLURRY PREHEATER
DISSOLVER
SEPARATOR
SEPARATOR
DISSOLVER
Product Slurry
GAS PURIFICATION
Purified Hydrogen
ACID GAS REMOVAL
CRYOGENIC SEPARATION
PIPELINE GAS
LPG
FROM ACID GAS REMOVAL UNITS
Acid Gas
TAIL GAS TREATING
SULFUR
CLAUS UNIT
Recycle Hydrogen
TO PLANT FUEL GAS SYSTEM
COMPRESSOR
Hydrogen
NAPHTHA
FUEL OIL
GASIFICATION
COMPRESSOR
ACID GAS REMOVAL
SHIFT CONVERTER
Quench Water
GASIFIER
OXYGEN PLANT
Steam
SECONDARY RECOVERY
AMMONIA RECOVERY UNIT
PHENOLS RECOVERY
PHENOLS
INERT SLAG
FRACTIONATION
FRACTIONATOR
VACUUM TOWER
Mineral Residue Slurry
THE SRC - II PROCESS

FIGURE I

The temperature of the reactants at the outlet of the preheater is about 700-750°F. At this time the coal is already partially dissolved in the recycle slurry "solvent," and the exothermic reactions of hydrogenation and hydrocracking have just begun. The heat generated by these reactions in the dissolver further raises the temperature of the reactants to the range 820-870°F. Hydrogen quench is injected at various points in the dissolver to control the reaction temperature and alleviate the impact of the exothermic reactions.

The dissolver effluent goes first to a hot high pressure separator. The hot overhead vapor stream from this separator is then cooled in a series of heat exchangers and additional vapor-liquid separation steps. The condensed liquid from these separators goes to the fractionation system. The non-condensed gas consists of unreacted hydrogen, methane and other light hydrocarbons, plus H_2S and CO_2. The gas is first sent to an acid gas treating unit for removal of H_2S and CO_2. A portion of the treated gas is then further processed in a cryogenic unit for removal of much of the methane and other light hydrocarbons. The recovered hydrogen (90% pure) is added to the remaining gas from the acid treating step. The resulting mixture of recycle gas, plus fresh make-up hydrogen from the gasifier train, comprises the hydrogen fed to the process. A mixture of methane and ethane recovered in the cryogenic unit is produced for sale as pipeline gas, while propane and butane are recovered for sale as LPG. The H_2S removed in the acid gas treating unit is converted to elemental sulfur in a standard Claus unit equipped with a tailgas treating unit to insure compliance with environmental standards.

The liquid (slurry) from the hot high pressure separator, after flashing to lower pressure, is split into two major streams. One of these streams comprises the recycle "solvent" for the process, while the other goes to the fractionation system for separation of the major products of the process.

In the fractionation system the slurry product is first distilled at atmospheric pressure to separate a gas stream, a naphtha stream, a middle distillate stream, and a bottoms stream which goes to a vacuum flash system for further separation. The temperature of the feed to the fractionation system is maintained at a high enough level so that no additional preheating is needed, other than for startup operations. The overhead from the vacuum flash tower, together with middle distillate from the fractionator, makes up the major fuel oil product of the process. For operation in the lower range of "solvent" to coal ratios, a portion of this product may be recycled to the feed slurry mixing tank to insure that the solids content of the feed slurry does not exceed the maximum pumpable level.

The bottoms from the vacuum tower, consisting of all of the undissolved mineral residue plus the vacuum residue portion of the dissolved coal, goes to an oxygen-blown gasifier for ultimate conversion to hydrogen. The synthesis gas produced in the gasifier goes through a shift conversion step for conversion of steam and carbon monoxide to hydrogen plus carbon dioxide, then to an acid gas removal step for removal of the carbon dioxide. The hydrogen (94% pure) is then compressed and fed as make-up hydrogen to the preheater-dissolver of the SRC-II process.

All of the waste water streams from the process are collected and sent to a distillation unit for removal of ammonia. After ammonia removal, the water is sent to a phenols recovery unit for separation of

TABLE II

ANALYSIS OF FEED COAL FOR DEMONSTRATION PLANT DESIGN

Illinois No. 6 Bituminous Coal

Analysis: % by Wt.	As Received	Dry Basis
Carbon	61.4	69.8
Hydrogen	4.3	4.9
Sulfur	4.6	5.2
Nitrogen	0.9	1.0
Oxygen	8.4	9.6
Ash	8.4	9.5
Moisture	12.0	--
Total	100.0	100.0

TABLE III

PRODUCT YIELDS FROM SRC-II PLANTS

Coal Feed: Illinois No. 6 Bituminous

Coal Feed Rate (Dry Basis): T/CD	6,000	30,000
Products		
Pipeline Gas: MM SCF/CD	15	34
LPG: B/CD	900	7,440
Light Distillate: B/CD	3,400	15,000
Fuel Oil: B/CD	12,250	61,200
Sulfur: T/CD	250	1,300
Ammonia: T/CD	13	60

TABLE IV

THERMAL EFFICIENCY OF SRC-II PROCESS
(Based on Conceptual Commercial Plant Design)

		MM Btu/D
Input		
Coal		759,700
Electrical Power (41 MW(e))		8,600 (a)
	Total	768,300
Output		
Pipeline Gas		44,800
LPG		29,100
Light Distillate		82,900
Low-Sulfur Fuel Oil		383,100
	Total	539,900
Thermal Efficiency		70.3%
Equivalent Petroleum Refinery Basis (BTU)		100,000 Bbl/D

(a) Based on Power Plant Thermal Efficiency of 39%.

phenols. The ammonia and phenols are recovered for sale as by-products. Even after such purification, however, the waste water is not discharged to the surrounding watershed, but is recycled to the process to avoid any possibility of adverse environmental effects.

C. Product Yields and Properties

A conceptual design has been developed for a 6,000 T/D demonstration plant, based on recent experimental results from laboratory and pilot plant units. The design was based on charging Illinois No. 6 coal with properties as given in Table II. This coal is very similar in properties to the Western Kentucky coal used in the laboratory experiments. The design incorporates the basic process scheme outlined in the previous sections, including a gasifier section for hydrogen production and all supporting facilities. The overall yields from such a plant are given in Table III, together with yields from a 30,000 T/D plant made up of five 6,000 T/D modules.

The overall product yields from the 30,000 T/D plant on a Btu basis are given in Table IV. The yields represent the products remaining for sale after deducting the plant fuel requirements. Fuel for the plant is supplied from the gas produced in the process. The only outside source of energy involves purchase of about 40 megawatts of electrical power. This is necessary since the net power required is too small to justify an in-plant electrical generation system. All of the major compressors and pumps in the plant are steam driven. The overall thermal efficiency of the process, after adjusting for the purchased power, is 70% (high heating value basis).

While the yield of pipeline gas product is relatively low in this case, the quantity of gas could be increased by appropriate changes in operating conditions. The process is flexible enough to produce significant quantities of pipeline gas, in addition to a low-sulfur fuel.

In view of the fact that the SRC-II process tends to produce both gas and liquid fuels, it is more efficient and economical to do this rather than to try to force the production of one or the other of the two products. The economic advantages of such an approach were shown in an earlier economic study for a coal conversion plant producing both gas and liquid products by a somewhat different combination of process steps.(3)

The major product of the process is a low-sulfur fuel oil suitable for use in power plants and industrial installations. In contrast to the solid fuel produced by the SRC-I process, the fuel oil product from SRC-II is an all-distillate liquid containing no ash and less than 0.4% sulfur. The properties of the product fuel oil are given in Table V.

The fuel oil is sufficiently low in sulfur that it will not only meet current EPA guidelines for low-sulfur fuel oil, but will also meet many of the more stringent requirements specified in certain major metropolitan areas. Since the viscosity of the fuel oil is only about 50 SUS at 100°F., it is readily transportable by pipeline. The heating value of the fuel oil is 17,284 Btu per pound, substantially more than that of the solid SRC-I product. The sulfur content of 0.3%, combined with a heating value of greater than 17,000 Btu/lb., represents an effective reduction in potential SO_2 emissions of greater than 90% on a Btu basis. Although the SRC-II product is a better quality fuel, its cost promises to be no higher than that of the SRC-I product, primarily because of elimination of the filters, dryers, and solidification equipment.

TABLE V

TYPICAL PROPERTIES OF SRC-II FUEL OIL

Gravity: °API		5.0
Approximate Boiling Range: °F.		400-900
Molecular Wt.:		230
Flash Point: °F.		168
Viscosity SUS at 100°F.		50
Elemental Analysis: % by Wt.		
Carbon		87.2
Hydrogen		7.9
Sulfur		0.3
Nitrogen		0.9
Oxygen		3.9
	Total	100.0
Heating Value: Btu/lb.		17,284

The process also produces a significant quantity of light distillate consisting of liquids in the boiling range C_5-400°F. This liquid could be burned as a process fuel to make more pipeline gas available for sale, or alternately could be utilized in a nearby petroleum refinery. The naphtha portion of this liquid is high in cyclic hydrocarbons, and with appropriate treatment for removal of oxygen and nitrogen, would make an excellent charge stock for production of high octane gasoline by catalytic reforming. Since this fraction also contains 20-30% phenols, these materials could be recovered for sale as by-products of the process. The properties of the light distillate fraction are given in Table VI.

The pipeline gas product consists of 74% methane and 24% ethane, plus very minor amounts of propane, nitrogen and hydrogen. Its heating value is 1,200 Btu per standard cubic foot. The LPG from the process consists of approximately 84% propane, 13% butane and 3% ethane on a volume basis. The heating value of the LPG fraction is 3,900,000 Btu per barrel.

III. Proposed Demonstration Plant Program

A. Scale-Up Problems

The SRC-II process appears to be one of the most promising variations of the many proposed applications of the Solvent Refined Coal technology, and represents a logical choice for commercialization. Beyond the pilot plant work, we believe that commercialization must be next pioneered through the design, construction and operation of a commercial module which consists of the minimum size processing train that would be represented in a commercial plant. Based on a series of engineering evaluations, a single full size module for the SRC-II process would have a nominal capacity of 6,000 tons per day. A full scale commercial plant

TABLE VI

TYPICAL PROPERTIES OF SRC-II LIGHT DISTILLATE

Gravity: °API		39.0
Approximate Boiling Range: °F		100-400
Elemental Analysis: % by Wt.		
Carbon		84.0
Hydrogen		11.5
Sulfur		0.2
Nitrogen		0.4
Oxygen		3.9
	Total	100.0
Heating Value: Btu/lb.		19,048

would consist of several such modules, with five modules appearing as a logical and economical commercial size. On this basis, a commercial plant would have a capacity of 30,000 tons per calendar day, the output of which would be equivalent to a 100,000 barrel per day petroleum refinery.

While many aspects of the SRC-II process have and are being verified at the Ft. Lewis SRC pilot plant, there are certain features of the process which can be proven only in a full scale unit with commercial size equipment. Thus, further development in anything less than full size equipment would still leave some definite uncertainties to be resolved in the commercial commitment.

A brief summary of the features of the process which must be proven in the commercial module demonstration program is given in the following paragraphs. For certain items, particularly the cryogenic unit and the oxygen compressor, unproven designs would be used primarily because of cost advantages.

1. Slurry Mixing and Pumping

After the hot recycle slurry is mixed with the feed coal, the temperature of the resulting mixture is 450°F. At this temperature the coal-slurry mixture forms a gel rather rapidly, resulting in a sharp increase in slurry viscosity. Since the gel is a non-Newtonian fluid, the effects of the high viscosity can be overcome by strong shear forces generated by mixing and pumping. Such shear effects, however, are strongly dependent upon the size of the equipment. Thus, even though extensive pilot plant work is now being carried out, ultimate demonstration of this step can be accomplished only in full scale equipment

2. Slurry Preparation

Since the presence of a highly viscous gel persists throughout much of the length of the preheater, there is uncertainty regarding the actual effective viscosity of the reactants. Again, the effective viscosity will be greatly dependent upon the size (primarily diameter) of the preheater. Since the effective viscosity has a pronounced influence on pressure drop, heat transfer and erosion, it is obvious that some uncertainty will remain in the design of the preheater until after the step is demonstrated in full size commercial equipment.

3. Dissolver System

While hydrogen quench is important in controlling the reaction temperature in the dissolver, another major factor in dissipating the exothermic heat of reaction is backmixing of the reactants. Good backmixing is essential for distribution of the heat to the maximum extent throughout the reactor. The extent of backmixing is greatly affected by reactor dynamics, which are in turn greatly dependent upon the size and configuration of the dissolver vessel. In addition, dissipation of heat by heat loss to the surroundings is greatly affected by dissolver size. Thus, final optimized dissolver performance can be verified only in full scale commercial equipment.

4. Fractionation

While basic fractionation technology is well established for solids-free systems, the presence of the solids in both the fractionator and the vacuum flash drum of the SRC-II process is unique. For example, the bottoms stream from the vacuum tower may

contain as much as 40% solids. The liquid in the drum is a very viscous material, solid at room temperature, and must be maintained at about 600°F. or higher to keep it fluid enough to flow readily through the lines and pumps.

While it is generally anticipated that most of the problems associated with the presence of solids can be solved in pilot plant work, we will not be certain of the operability of either of these fractionation steps until they are demonstrated in a full size commercial unit.

5. Pumping and Handling of Concentrated Slurries

The SRC-II process involves pumping and handling of a number of fairly concentrated slurries, resulting primarily from the continuing recycle of slurry. The high viscosity and erosiveness of these streams requires a major effort in designing adequate pumping and handling systems. While most of these problems are being worked out at the pilot plant scale, the effects of the high viscosity and erosiveness, as well as the problems of keeping the pumps properly sealed, may be more severe in larger equipment. Thus, full scale tests will be necessary to optimize the slurry pumping systems, even after extensive pilot plant testing.

6. Hydrogen Purification

Recycle hydrogen is purified in a cryogenic system "cold box" prior to recycle to the slurry mixing and preheating system. A cryogenic system has not previously been used for feed gas compositions typical of the SRC-II process operating as high as 2,000 psig, and thus will require demonstration as part of the overall process.

7. Gasifier

Gasification of a feed stock such as the high-ash content vacuum residue slurry from the SRC-II process has never before been accomplished in a full scale commercial plant. It is anticipated that mechanical problems of charging the gasifier feed and withdrawal of inert slag, as well as erosion problems inside the gasifier itself, will be highly dependent upon the scale of operation. Thus, successful demonstration in full size commercial equipment will be necessary to fully prove out this step.

8. Oxygen Compression

The oxygen compression unit, consisting of a three-stage centrifugal compressor and inter-coolers, compresses 99.5% pure oxygen from atmospheric pressure to 1,600 psig. Since an oxygen compressor has not previously been built for this pressure, its operability and reliability must be demonstrated on a commercial scale.

B. Investment Costs

While the major objective of the demonstration plant program is to prove the operability, reliability and performance of the plant on a full size commercial scale, another important consideration in the overall plan for commercialization of the process is the eventual commercial value of the plant as part of a commercial facility after the process has been successfully demonstrated. This is particularly significant in view of the large capital investment required in pioneering efforts at commercial scale. It is generally anticipated that the plant would be expanded to a commercially economical size, consisting of several process trains, probably the 30,000 T/D size indicated earlier. Since the demonstration plant would have been initially conceived and engineered with its utility as one module of a commercial plant in mind,

such expansion could be accomplished making use of much of the equipment and facilities already available in the demonstration plant.

The capital investment cost estimate, based on an engineering design developed in conjunction with the Stearns-Roger Corp. of Denver, Co., for the 6,000 T/D commercial module exceeds $300 million in 1975 dollars, with the full 30,000 T/D commercial facility estimated to cost about $1 billion. Figure 2 shows one possible configuration wherein subsequent commercial expansion of the module facility might be accomplished with minimum disruption. In addition to the engineering estimates, an examination has been made of the potential use of the demonstration plant equipment and facilities in an expanded commercial plant, including an estimate of the economic value of the existing facility as an integral part of the commercial plant. The results of this study are summarized in Table VII.

Pre-commercial modifications to the demonstration facility include modifications as a result of indicated process improvement, replacement of equipment unduly worn as a result of extensive experimental testing, and removal of experimental units.

Yield penalty is a discounted value penalty which assumes that the original module achieves yields of approximately 95% of the added modules for the commercial project lifetime (20 years).

The net value of the demonstration plant is thus represented by the $179 million "displacement" value less the pre-commercial modification expense and the yield penalty, or $142 million. Significantly, the "worth" of the commercial-module demonstration plant for the SRC-II advanced coal liquefaction is almost half of the original investment.

FIGURE II CONCEPTUAL SITE ARRANGEMENT FOR SRC-II FACILITY

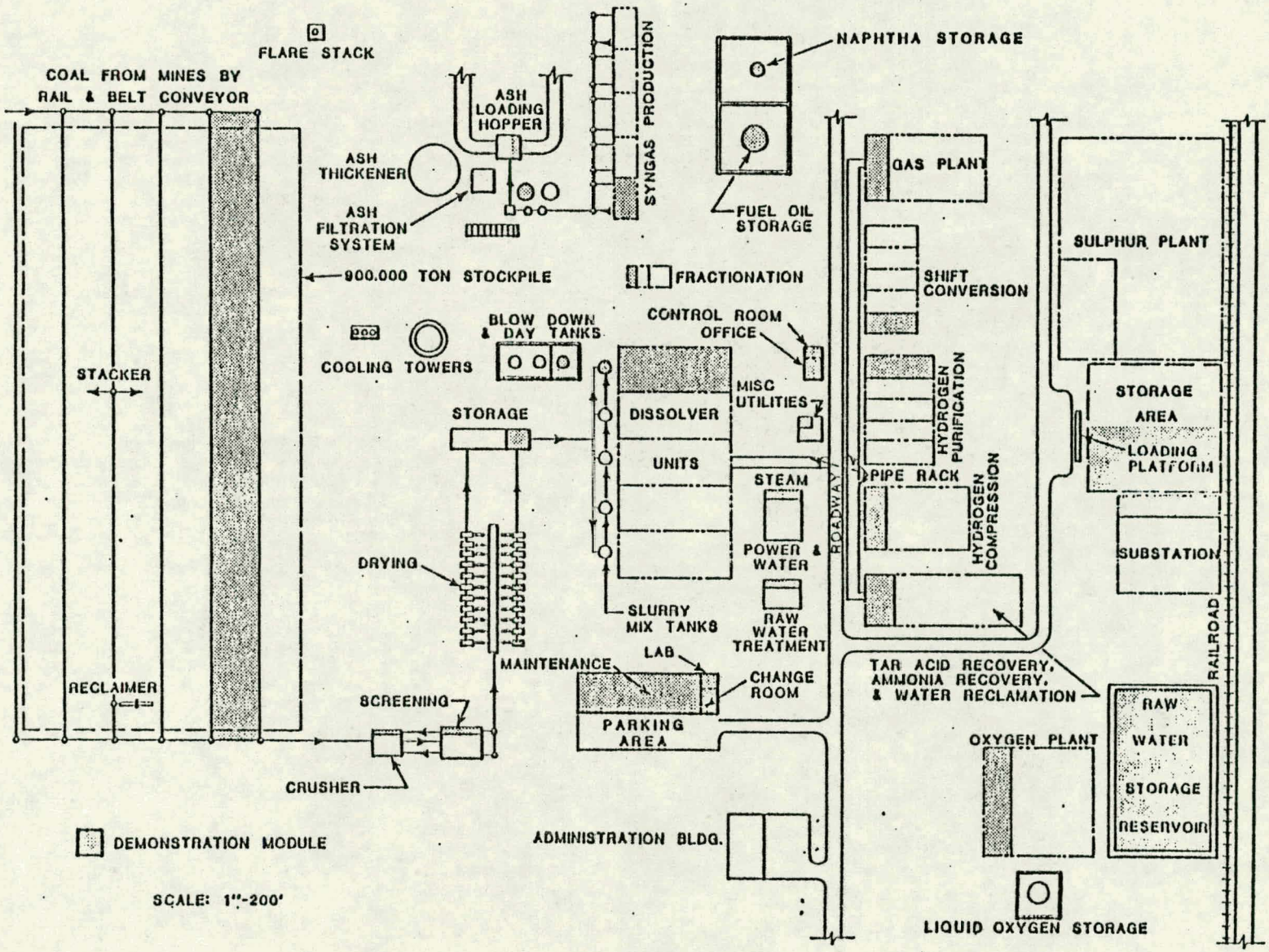


TABLE VII

VALUE OF DEMONSTRATION PLANT IN COMMERCIAL EXPANSION

Millions of Dollars (1975 Basis)

System	Demonstration Plant	Value to Commercial Unit	Additional Costs for Commercial Expansion	"Grass Roots" Commercial Cost
Coal Preparation and Ash Handling	18	15	58	73
Primary Process Plants	95	62	289	351
Hydrogen Production Plants	56	35	132	167
Gas Plants	21	6	86	92
Utility Systems	54	27	165	192
General Facilities	26	24	47	71
Offsites	21	10	5	15
Sub-Total	301	179	800	979
Pre-Commercial Modification		(10)	10	
Yield Penalty	—	(19)	19	—
TOTAL	301	142	837	979

IV. Conclusions

Extensive experience and understanding of Solvent Refined Coal processing has developed from almost two years of operation of the Fort Lewis pilot plant. A very promising version of the SRC process, known as SRC-II, has now evolved which will efficiently produce a distillate hydrocarbon fuel oil of less than 0.4% sulfur content. The SRC-II process has a major advantage in that it does not include filtration, or other steps with difficult process and mechanical problems. We believe that the current phase of pilot scale work at the Fort Lewis facility will confirm that the process is ready for further demonstration in a commercial module facility.

Uncertainties remaining include process scale-up, large scale operational reliability, optimization of process performance, accurate determination of investment and operating costs, as well as development of materials and componentry needed for a coal liquefaction industry.

We have recommended that the SRC-II process be demonstrated in a commercial module facility. The plant would be designed as one complete module of a future commercial plant, using full size equipment. By taking this approach, not only could calendar time be saved in commercialization of the process, but much of the cost of the demonstration plant could be recoverable through utilization of the equipment and facilities in an expanded commercial plant.

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POLITICAL AND ECONOMIC JUSTIFICATION FOR IMMEDIATE REALIZATION OF A SYN FUELS INDUSTRY

BY

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TO BE PRESENTED AT:

THE THIRD ANNUAL INTERNATIONAL CONFERENCE ON COAL GASIFICATION AND LIQUEFACTION
WHAT NEEDS TO BE DONE NOW!

August 3 - 5, 1976

UNIVERSITY OF PITTSBURGH
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INTRODUCTION

During the past few years, since the Arab boycott of late 1973, there has been much published concerning the impact of the embargo on U. S. energy needs and the derived policy of "Project Independence". Much of what has been said regarding curtailments and economic hardship has not come to pass, and as a consequence, our country and the public, in particular, have been lulled into a false sense of security. This winter we will again play Russian roulette.

Recently the Federal Power Commission predicted a further 30% decrease in gas supply this winter, compared to last. As with past due predictions, it is difficult to assess the impact of the imminent shortfall. As FPC Chairman Richard Dunham said, "the effect of this shortfall will depend on the weather, industrial demand and the extent to which users convert to alternative fuels."[1] Time will tell whether this worsening shortfall will catch up with us this winter.

Compounding our bleak domestic gas supply situation is our weak bargaining position with our energy suppliers. I truly believe that, in order for our country to maintain its economic and political independence, we must begin a program now, which will allow the United States, in the minimum, a reasonable negotiating position with our international suppliers who furnish much of our energy needs.

One way to achieve some measure of independence is to implement greater use of our largest energy resource, coal. Current annual consumption is 640,000,000 tons per year. A comparison of the reserves with current consumption clearly indicates that we have more than adequate reserves of coal for centuries to come.

Despite this great abundance of coal in our country, we still import increasingly large amounts of our energy from foreign, sometimes unfriendly and undependable sources. Currently, we import approximately 40% of our oil needs. This is predicted to go up to 50% by 1980. In order to keep our imported energy from increasing even further, we must begin now to increase development of indigenous natural resources.[2]

While the news has not been very encouraging of late, some progress has been made.

As you know, there now is a law which calls for gradual de-control of oil pricing over the next 40 months.

The Federal Power Commission has handed forth a decision on the Wesco project which calls for building an SNG from coal plant in the Four Corners area in New Mexico. This latter ruling appears to be satisfactory to the pipeline company, as it calls for a guaranteed reasonable return

on investment and a pricing mechanism for syn fuels. However, this plant, in 1980 dollars including the mine development, will cost approximately $1 billion. Thus, the remaining all-critical area left to the realization of this project is that there must be an adequate federal loan guarantee which is a prerequisite for the financing of such huge first-of-a-kind projects. Last December, Congress voted against such loan guarantees. During recent months new attempts have been made to introduce legislation in Congress that will allow such guarantees. At this time, it is uncertain as to whether this attempt will be successful.

The ERDA R & D and Demonstration plant programs are taking shape and the foundation is being laid for the second generation coal gasification plants of the late eighties and nineties. The Phase I engineering plan for three large scale low and high BTU demonstration plants has been established and some design work is under way. These plants will be completed in the mid eighties, ready for commercialization in the latter part of that decade. In addition, ERDA has a number of other coal gasification programs ready for imminent contract award.

Now that it appears that U. S. Federal policy is slowly taking shape and plans are now in effect for a coal gasification industry that will reach maturity by around the year 2000, can

we afford to sit back and let this plan slowly take effect? Should we forget the potential for immediate realization of a first generation coal gasification industry which could result in plant operation less than four years from today? The answer is a clear no!

The list of immediate political and economic advantages to SYN FUELS NOW is long and quite impressive. I would like to devote the remainder of this paper to exploring with you some of the more important benefits of an immediate commercialization program.

ADVANTAGES IN THE POLITICAL AREA

As stated at the beginning of this discussion, with a concerted effort in Syn Fuels, U. S. industry would be in a much stronger bargaining position with OPEC. A serious start toward Project Independence would temper the price demands of our international suppliers.

A second important benefit of SYN FUELS NOW would be the impact it would have on international boycotts. Admittedly, Syn Fuel production would initially be quite modest, thus, its influence on near term boycotts would not be significant. However, as the industry grows, its impact will become more pronounced.

It should be noted that some countries take energy independence

quite seriously. One notable country, South Africa, is presently building a two billion dollar Syn Fuels plant from coal. This plant will employ approximately 36 commercial size gasifiers complete with gas treatment and gas catalytic reformation to produce a myriad of liquid fuel products. When one considers the total gross national product of South Africa, which is on the order of $25 billion, and the $2 billion+ capital cost for this new Syn Fuels plant, a comparable figure in the U.S. economy is the equivalent of an expenditure of more than $140 billion. When one measures this tremendous undertaking on the part of the South Africans, one can readily see that some countries take energy independence very seriously. This one project alone will increase the South African indigenous production of liquid hydrocarbons from 10% to 40%.

ECONOMIC BENEFITS

As important as the political benefits of energy independence are, one can made even a stronger case for its economic benefits. These latter benefits are broadly based and include regional, national and international incentives.

REGIONAL AND NATIONAL

Pending severe regional natural gas shortages are faced by certain of our states. Montana is a classic example of this problem. This state obtains sixty-eight percent of its natural gas from

Canada. Based on the total volume allotment given to the Montana Power Company by Canada, and the current import rate, this allowment will be totally depleted by 1984. Montana desperately requires this natural gas to fuel its homes and economy. The Montanans would welcome the opportunity to be assured of a gas supply, which admittedly at this time, because of price controls, is somewhat more expensive than natural gas. However, over the long-term and because of the relatively modest coal escalation cost, the delivered SNG cost is forecasted to be far more stable than imports, which include Canada. The price of imported natural gas from Canada to the States over the last three years has increased over 500%. Can we assume that, even if such gas were to continue to be available, that these increases would diminish? If you draw the cost versus time curves for Canadian natural gas and SNG from coal, we believe these curves will intersect in the not too distant future. One SNG plant will more than supply the entire needs of Montana. Montana would certainly welcome a project to make SNG from coal.

The Californians are also desperate for natural gas. They are now scrambling for LNG from Indonesia. The cost of LNG is also increasing rapidly. Again, a similar analogy to that just mentioned for Canada can be established for California. Today, the cost of imported LNG in the California distribution system will not be much different than the cost of SNG from coal. Even assuming it would be slightly less today, what will it be five years or ten years from now?

The economic stimulization derived from a Syn Fuels industry would have immediate favorable regional impact on technical employment. Consider the situation in New Jersey. Despite the general improvement of the unemployment statistics on the national scene, New Jersey unemployment is still languishing at greater than 10%. One significant area of this state's employment sector is the engineering/construction industry.

The aggregate total technical and other home office staffs of the companies residing in New Jersey is approximately 6,000. Many more thousands of engineers live in New Jersey and work in adjacent states.

The award on one SNG plant to a New Jersey contractor will in effect use up approximately 25% of the entire State's capability, as one SNG plant requires approximately 1500 man years of technical support. This, of course, would have the derived benefit of increasing technical employment for other process plant design activities and metal fabricating shop activity. The infrastructure that supports the towns in which these offices are located will also benefit greatly.

I would like to draw a simple analogy. If one were to build a coal gasification plant, which requires around 600 operating and maintenance personnel, say in the Far West, where an input-output analysis can readily isolate and measure the impact of such a plant, this plant would support a town of 10,000. This has been studied by one of the major pipeline

companies, and I believe can be well documented.[4]

Assuming that a similar input-output analogy would prevail with the employment of 1500 technical personnel, one can readily calculate that this would support far more than a town of 10,000. This increase is just the vanguard of the increase in employed population. Employment in our equipment manufacturing and metal fabricating industries will increase several fold over the engineering gains. Recently it has been estimated that one SNG from coal plant requires approximately 60,000 tons of fabricated metal products.[5] Using industry rules-of-thumb, this translates to about 4,000 man years of shop labor.

There is an even greater generation of jobs for the construction industry. As you know, there is very high unemployment in the construction trades. One SNG plant will require approximately 11,000 man years of construction labor. Can you imagine what this will do for the region where this plant is built, and the incomes that will be received by the craft workers? It is not surprising that the A. F. of L leadership has recently given its endorsement of the Federal Loan Guarantee Act to the House Science and Technology Committee.

The growth impetus from the initial SNG plants would, in turn, spill over into other capital plant construction. A clearer energy policy would give encouragement to private industry to invest more money. We believe now that with corporate cash flow

increasing and given a clear-cut federal energy policy, private industry would gain confidence and would invest in other new capital plant construction. Furthermore, the construction of syn fuels plants will help to relieve the growing feedstock shortage faced by the U. S. chemical producers.

Utilizing the gross data for the manpower requirements required for one SNG from coal plant and the input-output analysis study prepared by one of the pipeline companies, I have summarized, very roughly, the gross impact on employment from the building of one SNG plant.

I apologize to the Economists present in the audience for the primitive approach used, but the numbers should be roughly indicative of the real world.

In viewing the numbers presented in Figure I, please note that the input-output multiplier extrapolated from the SNG plant site study for use with the transitory employment of the technical service industries is only appropriate, if the initial impetus to build the SNG plant carries into other construction activities.

FIGURE I

ECONOMIC IMPACT ON EMPLOYMENT FROM THE BUILDING OF A COMMERCIAL SIZE SNG FROM COAL FACILITY [3]

	Man Years
Engineering, Design and Other Home Office Services	1500
Shop Labor	4000
Field Construction Labor	11000
Plant Operating and Maintenance Labor (600 personnel x 20 year plant life)(1)	12000
	28500

Total infrastructure gain in employment =

$$\frac{10000}{600}\ \frac{\text{(Regional Population Supported)}}{\text{(Employment-Basic Industry)}}^{(2)} \times .3\text{(Adult Employment Factor)} \times 28{,}500 = 142{,}500$$

(1): Assumes 20 year useful plant life

(2): Based on studies by Pipeline Industry, basic industry employment supports a regional population of 10,000 people.

(3): Neglects employment gains in derived industries such as refining plant by-products or follow-up technical service activities furnished by the expanded engineering staff.

International

While I don't have the courage to state variance and confidence levels for the Figure I conclusions, I believe that these data give one sufficient courage to say that it is preferable to spend $1,000,000,000 in coal mine development and SNG capital construction rather than import LNG for foreign and sometimes undependable sources. Furthermore, we would not be shipping any of our wealth outside of the country. Capital formation would occur in our country and in foreign lands. We would be financing fewer large projects in Indonesia and Algeria; we would not be building LNG tankers, which in many instances are not built in our shipyards; we would be putting our engineers, fabricators, and construction workers to work; and we would eliminate a drain on our balance of payments with other nations in the world.

ALTERNATIVES TO THE DEVELOPMENT OF A SYN FUELS INDUSTRY:

Aside from the obvious desirable goals of employing improved conservation techniques and encouraging greater development of conventional indigenous fossil fuel production, we have several other options to the development of a Syn Fuels industry.

a) We can continue to increase importation of crude oil from sometimes unfriendly countries and pay them more than $14.00 per landed barrel, while we control payments to U. S. producers at $7.66/bbl. plus some allowance for escalation;

b) We can continue to import natural gas from Canada, where the cost has increased over 500% in the last three years, or;

c) We can increase our imports of LNG from Algeria and Indonesia.

I believe that we can catagorically state that Syn Fuels are preferable to these alternatives.

IS SYN FUELS TECHNOLOGY READY?

One of the often voiced myths currently in fashion is the belief that coal gasification is not ready for commercialization. There are a number of countries that don't agree, most notable of which is South Africa who, as stated earlier, is building a 2 billion dollar syn fuels facility. Such an undertaking is a prodigious feat for such a small economy.

Are we paying too great a price for building first generation plants when compared to second generation plants, that may be ready for commercialization in about ten years? The answer is a clear no!

To prove my contention...let us explore the sensitivities of hypothetical R & D improvements in coal gasification. As backgound, I would like to present a typical analysis which sets forth those elements that contribute to the capital and operating cost of a coal gasification plant.

As shown in Figure II, while coal gasification is the key process, it is still only one step of the overall scheme and it represents only 20% of the total investment cost. Even if a new process reduces the investment cost of the gasification section by 50%, (nobody claims that this is possible) the total overall capital requirement would be reduced by only 10%.

A simplified approach to calculating the gas production cost for a plant to produce 250 MMM BTU/d (= 10.42 MMM BTU/hr.) is as follows:

FIGURE II

BREAKDOWN OF INVESTMENT COST
1ST QUARTER - 1976
OF AN SNG FROM COAL PLANT[6]

Coal Handling	8%	
Gasification	20%	PROCESS PLANTS = 49%
Shift Conversion	3%	
Waste Liquor Treatment By-Product Recovery	7%	
Gas Purification	10%	
Sulfur Recovery	3%	
Methane Synthesis	6%	
Oxygen Plant	11%	UTILITY SUPPLY = 34%
Steam and Power Supply Distribution	18%	
Water Supply and Treatment Cooling Water	5%	
Site Development Buildings Miscellaneous	9%	
	100%	= ($900 MILLION for a plant to produce 250 MMM BTU/d.)

A more detailed estimate of gas production cost can be seen from the El Paso, Wesco and ANG F.P.C. applications. The purpose of the above simplified scheme is merely to show the influence of capital requirement and cost of coal on the gas production cost.

Using Figure III as a reference, let us further develop some of the critical sensitivity analyses:

a) Further to the very theoretical assumption that a new process is perfected, which reduces the cost of gasification by 50%; the total capital requirement would be reduced by 10%, and the gas cost would be reduced by 0.144 $/MM BTU or only 5%.

b) The thermal efficiency is a key criteria in the evaluation of a process. There is no current known way to improve this efficiency in practice beyond a figure of about 72%. Assuming this figure is reached, the gas production cost would be reduced by 0.026 $/MM BTU or by about 1%.

c) At the first glance, the cost element "operation and maintenance" appears to be high. It should be noted that the expenses for operating personnel are only equivalent to 0.1 $/MM BTU. The balance are expenses for maintenance, catalyst, chemicals, administration and insurance. Thus, only a marginal saving can be realized

FIGURE III

TYPICAL SNG PRODUCTION COST[7]

1)	Net investment	900 MM $	
	interest during construction + start up expenses + working capital	245 MM $	
	Total capital requirement	1145 MM $	
	Interest + depreciation	10.5 %	
	Capital charges		
	$\frac{1145 \cdot 10^6 \$ \cdot 0.105}{8000 \text{ hrs/yr.} \cdot 10420 \text{ MM BTU/hr}}$ =		1.44 $/MM BTU
2)	Taxes		0.39 $/MM BTU
3)	Coal = 0.5 $/MM BTU thermal efficiency = 69% $\frac{0.5 \$/\text{MM BTU}}{0.69}$		0.72 $/MM BTU
4)	Operation and maintenance		0.50 $/MM BTU
	TOTAL GAS PRODUCTION COST		3.05 $/MM BTU

in this cost center.

The conclusions that can be drawn from these analyses are that coal gasification produces a gas which is unavoidably more expensive than indigenous natural gas at the present price level, and that, new processes when they are ready to go commercial will not significantly reduce the cost for gas-from-coal. In fact, the inflation rate will easily eliminate all savings which might result from the successful development of new processes.

First generation technology is not just some old item that is taken off the shelf, dusted and marketed with a new coat of paint. It is a living entity. Typical of this evolution is that experienced with our technology at Lurgi.

The historical development of the Lurgi process started in 1936. However, comparing the 1936 designs with present day is analogous to comparing a 1936 coal fired 100 MW boiler with the 1000 MW highly sophisticated modern plant of today. Each of these developments was made possible only by a stepwise refinement after the first commercial plants were placed into operation.

Since the initial first effort, our commercial experience has helped us to refine the process. Our expanded chemical and mechanical engineering know-how influenced the design of many individual components, all of which culminated into the design

of the present total process plant. A typical example of this evolution is shown in Figure IV, which illustrates the stages of the development of the Lurgi gasifier.

Our current Mark IV design which represents the improvements from the "third generation" has already been demonstrated in a gasifier processing 750 MM BTU coal input/hr. We believe that this design has further potential for an increase in capacity. Conceptual designs for a reactor to process more than 1000 MM BTU/hr. are available and this prototype, Mark V, is now in the final design stage.

There has been tremendous progress in the technology. However, in order to continue to build on this progress and to seek other improvements in a logical way, it is vital that commercial plants be built, based on the status of today's technology.

We would like to point out that the LURGI coal gasification process comprises much more than just the gasifier. It includes all process steps which are necessary to produce gas of the required quality. As with the evolution of today's modern gasifier, the remainder of the process has undergone a major transformation from the initial town gas plants built many

years ago.

8

Some milestones in the development of the overall process are:

- Replacement of a series of gas purification processes by the Rectisol process which makes it possible to clean large quantities of gas to a very high degree of purity.

- Introduction of the hot Crude Gas Shift Conversion with the attendent economic feature of utilizing the steam in the crude gas rather than injection of additional steam.

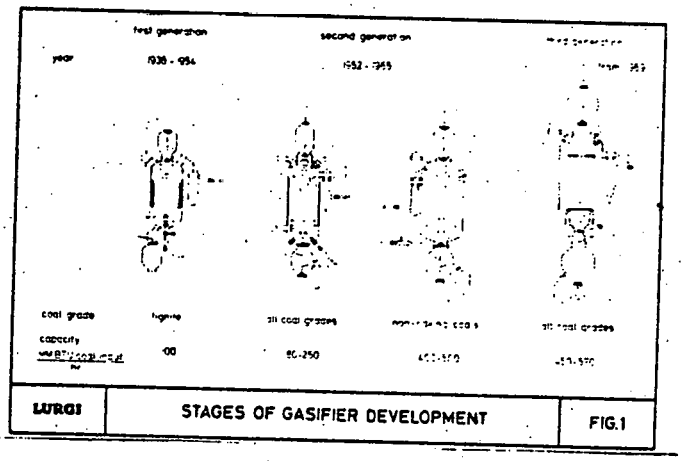
- Introduction of the tar recycle.
- Improvement in the by-product recovery plants.
- Application of modern remote control systems.
- Development of a coal distributor which makes it possible to process highly caking and swelling Eastern U. S. coals.
- Improved overall design backed up by large-scale test results in plants at Sasolburg (South Africa), Westfield (Scotland) and Luenen (Federal Republic of Germany).
- Most recent developments are:

  Operation of a LURGI gasifier as slagger

  Integration of the LURGI gasifier as a front-end process for advanced power cycles

  Re-use of waste liquor

FIGURE IV


first generation
second generation
year
1936 - 1954
1952 - 1965
coal grade
lignite
all coal grades
capacity
100
80-250
LURGI
STAGES OF GASIFIER DEVELOPMENT
FIG.1

It should be noted that the LURGI gasifier in conjunction with the British Gas Corporation slagging bottom, has been selected as the leading second generation technology by ERDA for a large scale high BTU demonstration plant. Such a program would not be possible if industry had written off first generation technology.

Finally, the first generation of commercial plants will serve as a useful benchmark from which to evaluate further R & D efforts.

In conclusion, it is my belief that a strong justification exists for the launching of a "Syn Fuels Industry", not next year, the year after, or in five years, but now. In less than four years, and with government cooperation, four large SNG plants could be built and operating. To reach this level of readiness, U.S. private gas distribution companies have already spent $50,000,000 in front end planning and engineering effort. Each of these plants is past the planning engineering phase; ready for detailed design, procurement and construction, and could by 1980 begin to make a significant contribution to the U. S. energy scene.

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