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HIGH-Btu COAL GASIFICATION PROCESSES

by

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FOREWORD

The Community Systems Program of the Division of Buildings and Community Systems, Office of Energy Conservation, of the United States Department of Energy (DOE), is concerned with conserving energy and scarce fuels through new methods of satisfying the energy needs of American Communities. These programs are designed to develop innovative ways of combining current, emerging, and advanced technologies into Integrated Community Energy Systems (ICES) that could furnish any, or all, of the energy using services of a community. The key goals of the Community System Program then, are to identify, evaluate, develop, demonstrate, and deploy energy systems and community designs that will optimally meet the needs of various communities.

The overall Community Systems effort is divided into three main areas: (a) Integrated Systems, (b) Community Planning & Design, and (c) Implementation Mechanisms. The *Integrated Systems* work is intended to develop the technology component and subsystem data base, system analysis methodology, and evaluations of various system conceptual designs which will help those interested in applying integrated systems to communities. Also included in this program is an active participation in demonstrations of ICES. The *Community Planning & Design* effort is designed to develop concepts, tools, and methodologies that relate urban form and energy utilization. This may then be used to optimize the design and operation of community energy systems. *Implementation Mechanisms* activities will provide data and develop strategies to accelerate the acceptance and implementation of community energy systems and energy-conserving community designs.

This report is part of a series of Technology Evaluations of the performance and costs of components and subsystems which may be included in community energy systems and is part of the Integrated Systems effort. The reports are intended to provide sufficient data on current, emerging and advanced technologies so that they may be used by consulting engineers, architect/engineers, planners, developers, and others in the development of conceptual designs for community energy systems. Furthermore, sufficient detail is provided so that calculational models of each component may be devised for use in computer codes for the design of Integrated Systems. Another task of the Technology Evaluation activity is to devise calculational models which will provide part-load performance and costs of components suitable for use as subroutines in the computer codes being developed to analyze community energy systems. These will be published as supplements to the main Technology Evaluation reports.

It should be noted that an extensive data base already exists in technology evaluation studies completed by Oak Ridge National Laboratory (ORNL) for the Modular Integrated Utility System (MIUS) Program sponsored by the Department of Housing and Urban Development (HUD). These studies, however, were limited in that they were: (a) designed to characterize mainly off-the-shelf technologies up to 1973, (b) size limited to meet community limitations, (c) not designed to augment the development of computer subroutines, (d) intended for use as general information for city officials and keyed to residential communities, and (e) designed specifically for HUD-MIUS needs. The present documents are founded on the ORNL data base but are more technically oriented and are designed to be upgraded periodically to reflect changes in current, emerging, and advanced technologies. Furthermore, they will address the complete range of component sizes and their application to residential, commercial, light industrial, and institutional communities. The overall intent of these documents, however, is not to be a complete documentation of a given technology but will provide sufficient data for conceptual design application by a technically knowledgeable individual.

Data presentation is essentially in two forms. The main report includes a detailed description of the part-load performance, capital, operating and maintenance cost, availability, sizes, environmental effects, material and energy balances, and reliability of each component along with appropriate reference material for further study. Also included are concise data sheets which may be removed for filing in a notebook which will be supplied to interested individuals and organizations. The data sheets are colored and are perforated for ease of removal. Thus, the data sheets can be upgraded periodically while the report itself will be updated much less frequently.

Each document was reviewed by several inividuals from industry, research and development, utility, and consulting engineering organizations and the resulting reports will, hopefully, be of use to those individuals involved in community energy systems.

ABSTRACT

This evaluation provides estimates of performance and cost data for advanced technology, high-Btu, coal gasification facilities. The six processes discussed reflect the current state-of-the-art development. Because no large commercial gasification plants have yet been built in the United States, the information presented here is based only on pilot-plant experience. Performance characteristics that were investigated include unit efficiencies, product output, and pollution aspects. Total installed plant costs and operating costs are tabulated for the various processes.

The information supplied here will assist in selecting energy conversion units for an Integrated Community Energy System (ICES).

TECHNOLOGY EVALUATION SUMMARY SHEET OF

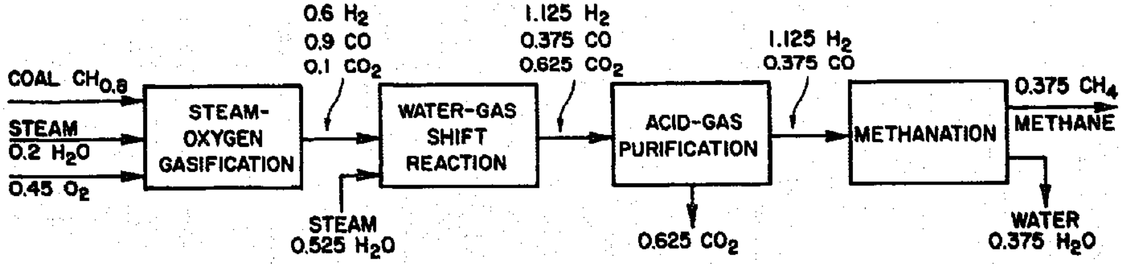
HIGH-Btu COAL GASIFICATION PROCESSESS

By: Christopher F. Blazek, et al, IGT January, 1979

ICES
INTEGRATED COMMUNITY ENERGY SYSTEMS

The conversion of coal to substitute natural gas (SNG) is a promising candidate in the current effort to supplement, and potentially replace, the world's rapidly diminishing supply of natural gas. In addition to SNG, many of the coal gasification processes also can produce synthetic liquid fuels, chemical feedstocks, or electricity.

Two methods are available to produce SNG from coal. The first utilizes a steam-oxygen gasifier to produce a gas mixture of H_2, CO, and CO_2. The hydrogen-to-carbon monoxide ratio then is adjusted to approximately 3 by the water-gas shift reaction. After removal of carbon dioxide in an acid-gas purification unit, the carbon monoxide and hydrogen gas are catalytically reacted to produce a methane-rich gas. The stoichiometry and general process steps of this synthesis-gas methanation technique are shown in Fig. DS-1.



OVERALL REACTION

$$CH_{0.8} + 0.35\ H_2O + 0.45\ O_2 \rightarrow 0.375\ CH_4 + 0.625\ CO_2$$

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Fig. DS-1. Pipeline Gas Production by Synthesis Gas Methanation

In contrast to the system just described, the second method, used by most modern gasification technologies, employs the concept known as hydrogasification. In this system, the incoming coal is reacted initially in a reactor with a hydrogen-rich gas to form substantial amounts of methane. The hydrogen-rich gas for hydrogasification is manufactured from steam utilizing the char leaving the hydrogasifier reactor. For example, hydrogen may be produced in a steam-oxygen gasifier, as illustrated in the stoichiometric and block flow diagram of Fig. DS-2.

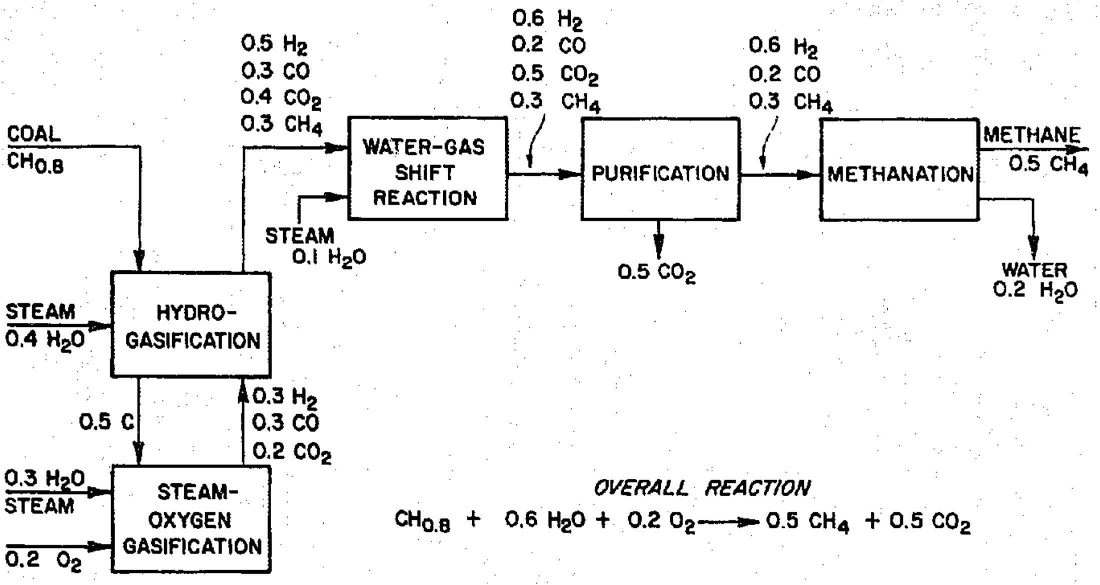


Fig. DS-2. Pipeline Gas Production Using Hydrogasification Plus Methanation

This evaluation analyzes six of the major high-Btu coal gasification processes. Although these processes currently are in the pilot-plant stage of development, conceptual designs of these six processes have been published. In general, these conceptual designs are for commercial-size plants producing 250 billion Btu of SNG per day. To make a valid comparison of these processes, the same (or nearly the same) coal feedstock was chosen.

In all but the COGAS process, a Montana subbituminous coal was assumed. Table DS-1 compares the six processes examined here.

The primary products of these processes form a wide spectrum, from substitute natural gas to highly refined liquids, such as light fuel oils or ammonia. Several of the processes also produce solid products, such as sulfur or char. The overall thermal efficiencies of these systems (i.e., energy out/energy in) are in the range of 56.4 to 77.6%. When SNG is considered as the sole primary energy product of these facilities, the cold gas efficiency (i.e., energy out/energy in) ranges from 51 to 69.7%.

Because no large, commercial gasification plants have yet been built in the United States, little is known about the true operating requirements, i.e., about the operating ranges and procedures or the

Table DS-1. Comparison of High-Btu Gasification Processes

Process Name	Process Type	Developer	Overall Thermal Efficiency (%)	Cold Gas Efficiency (%)	Products	Plant Capital Costs, 250 x 10^9 Btu/day SNG, $\$10^6$ (mid-1978 $)
Lurgi	Fixed Bed	American Lurgi Corp (U.S. Licenser)	71.7	59.0	SNG Oils Ammonia Sulfur Coal Fires	1516.3
Bi-Gas	Entrained Flow	Bituminous Coal Research	63.6	62.9	SNG Ammonia Sulfur	1451.6
Synthane	Fluidized Bed	Bureau of Mines (PERC)	56.4	51.0	SNG Oil Ammonia Sulfur Char	1619.0
COGAS	Fluidized Bed	FMC Corp.	67.6	Not Available	SNG Syncrude Light Oils Ammonia Sulfur Sodium Sulfate	Not Available
HYGAS	Fluidized Bed	Institute of Gas Technology	77.6	69.7	SNG Oils Ammonia Sulfur	1224.3
HYGAS Steam-Iron	Fluidized Bed	Institute of Gas Technology	70	62.2	SNG Oils Ammonia Sulfur	1799.0

safety requirements. Factors, such as turndown capacity, will be influenced greatly by the number of parallel process trains required to attain the desired output. Conversely, however, other pieces of process equipment may have economic constraints that dictate that it is less expensive per unit of material processed to have one large unit, as opposed to having several smaller ones. Thus, a gasification plant may have parallel trains in some areas of the process, and one single train in others. How these two opposing factors interact with each other will depend on the size of the plant. Because of this size variable, "accross-the-board" estimates of turndown capacity are difficult to assess.

Coal gasification plants will require full-time operating maintenance and administrative staffs. These staffs usually are made up of small groups, solely responsible for the operation of the major pieces of equipment used for gasification, pyrolysis, etc. Staff requirements are a function of the plant size. Based on the Lurgi process, it has been estimated that, for a process with a capacity of 250 billion Btu/day SNG, 150 operators, 161 service and administration employees, and 580 maintenance employees would be required. The scaling factor to be used here for estimating staff personnel as a function of plant size may be seen in the following relation:

$$\frac{\text{Desired Staff Size}}{\text{Known Staff Size}} = \left(\frac{\text{Desired Plant Size}}{\text{Known Plant Size}}\right)^{0.40}$$

Annual operating costs for the various gasification process plants shown in Table DS-2, are made up of four costs:

(1) material and supplies including the cost of coal and iron ore (where required);

(2) maintenance materials;

(3) contracted labor; and

(4) catalyst, chemicals, and operating supply costs.

Raw coal was assumed to cost \$20/ton delivered to the plant site. The cost of iron ore for the Steam-Iron HYGAS Process was set at \$23/ton.

Table DS-2. Annual Operating Costs for the Various 250 x 10^9 Btu/day Coal Gasification Facilities[1]

Component	Lurgi	BI-GAS	Synthane	HYGAS Steam-Oxygen	HYGAS Steam-Iron
	$$10^6$ (a)				
Material and Supplies					
Coal, at $20/ton	195.49	148.96	183.07	134.48	149.97
Iron Ore, at $23/ton	-	-	-	-	0.24
Purchased Water, at $0.46/1000 gal	0.53	0.47	0.47	0.41	0.75
Maintenance Materials(b)	14.51	13.89	14.35	11.71	21.16
Catalysts and Chemicals	10.41	5.32	3.66	4.23	5.75
Operating Materials(c)	1.15	0.95	0.95	0.95	0.95
TOTAL MATERIALS AND SUPPLIES	222.09	169.59	202.50	151.78	178.82
Labor					
Operating and Supervision(d)	8.96	8.00	8.14	7.35	10.18
Maintenance(e)	21.77	20.83	21.53	17.55	31.73
Plant Overhead/Administration(f)	18.43	17.30	17.80	14.94	25.14
TOTAL LABOR	49.16	46.13	47.47	39.84	67.05
Property Taxes and Insurance(g)	32.91	31.98	35.71	27.01	39.74
TOTAL OPERATING COSTS	304.16	247.70	285.68	218.63	285.61

(a) Mid-1978 $

(b) 40% of Total Maintenance

(c) 30% of Process Operating Labor

(d) $7.70/h for Operating Labor

(e) 60% of Total maintenance ($10.93/h wage rate)

(f) 60% of Labor

(g) 2.7% of Total Plant Investment

Maintenance material, contract labor, catalyst, chemicals, and operating materials can be scaled by the use of the 0.75 power factor relation. Labor requirements should be scaled by a 0.4 power factor relation.[1]

Operating and maintenance labor and their supervision were estimated by using the personnel requirement estimates shown in Sect. 7. Operating labor and supervision costs were set by applying an hourly average wage of $7.70/h. Maintenance labor and supervision wages were estimated at $10.93/h. The total maintenance labor cost was taken as 60% of the total maintenance and the total supervision cost was assumed to be 20% of the operating and maintenance labor. Wage rates were calculated by averaging the salary estimates used in the process design studies[6] and inflating them to mid-1978 dollars. Plant overhead and administration costs were set at 60% of labor.

Maintenance materials were assumed to be 40% of total maintenance costs, and operating materials were assumed to be 30% of process operating labor costs. Property taxes and insurance were fixed at 2.7% of the fixed capital investment for all plant sizes.

TECHNOLOGY EVALUATION OF
HIGH-Btu COAL GASIFICATION PROCESSES

Prepared by Christopher F. Blazek, et al., IGT
Date January, 1979

1 INTRODUCTION

Conversion of coal to substitute natural gas (SNG) is a promising technology in the current effort to supplement (and potentially replace) rapidly diminishing supplies of natural gas. The primary product of a coal gasification facility is substitute natural gas (SNG), which is composed of methane, hydrogen, carbon dioxide, nitrogen, and higher hydrocarbons. Secondary--but hardly incidental--products from a gasification facility can be synthetic liquid fuels, chemical feedstocks, or electricity.

General methods of coal gasification have been well-stated as follows:[1]

> "Since gaseous products contain a higher H/C ratio than coal, to convert coal to gas one must either add a lot of hydrogen or reject a lot of carbon. The more efficient way is to add as much hydrogen as possible and to reject as little carbon as possible.
>
> "For the production of the various gaseous products, the coal conversion is carried out in steps. In one approach, coal is first reacted with steam and oxygen at a relatively high temperature (1900° to 2500°F) to produce hydrogen and carbon oxides. The reaction of carbon with steam producing carbon monoxide and hydrogen, is highly endothermic:
>
> $$C + H_2O \rightarrow CO + H_2 \tag{1.1}$$
>
> The reaction (1.1) does not occur unless the required heat of reaction is supplied to the system. Usually, this heat is supplied by burning some of the coal:
>
> $$C + O_2 \rightarrow CO_2 \tag{1.2}$$
>
> In this way medium- or low-Btu gas is produced depending on whether pure oxygen or air is used for the combustion step (1.2). The product gas can be cleaned and burned for steam or power generation. Alternatively, for production of high-Btu gas...from medium-Btu gas, the hydrogen-to-carbon

monoxide ratio is adjusted...to 3...by the water-gas shift reaction:

$$CO + H_2O \rightarrow CO_2 + H_2 \qquad (1.3)$$

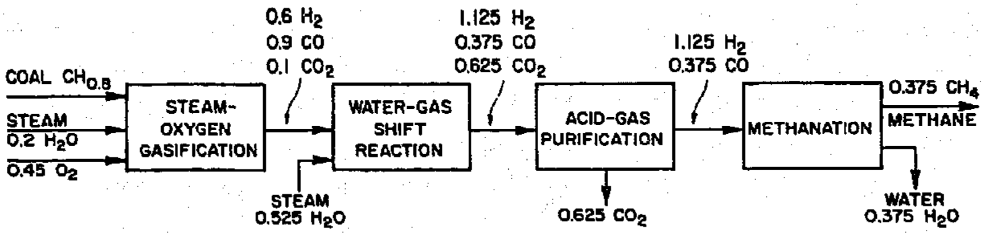
Following water-gas shift, both the sulfur pollutants and carbon dioxide produced during Reactions 1.2 and 1.3 are removed by acid-gas cleaning. The gas then contains only carbon monoxide and hydrogen, in proper proportion for synthesis of high-Btu gas by methanation....

"For high-Btu gas production, a mole of carbon monoxide reacts with three moles of hydrogen and produces methane.

$$CO + 3H_2 \rightarrow H_2O + CH_4 \qquad (1.4)$$

Water is removed from the gas by dehydration, and the high-Btu, methane-rich product is directly substitutable for natural gas.

"This sequence of reactions is illustrated schematically in Fig. 1.1. In the case of high-Btu gas a primary inefficiency in the system results from the methanation reaction (Reaction 1.4) being highly exothermic. Since all the product methane is produced by catalytic methanation, significant heat is released from the process at this point. Since this heat is released at a relatively low temperature it is of little value for the rest of the process. It cannot, for example, be recovered for use in support of Reaction 1.1 because Reaction 1.1 occurs at a much higher temperature. Although some of this methanation heat can be used to raise steam, much of it is discarded, constituting a process inefficiency.



OVERALL REACTION

$$CH_{0.8} + 0.35\ H_2O + 0.45\ O_2 \rightarrow 0.375\ CH_4 + 0.625\ CO_2$$

Fig. 1.1. Pipeline Gas Production by Synthesis Gas Methanation

"In contrast to the system just described, most modern gasification technologies employ the concept known as hydrogasification. In this system, the incoming coal is initially reacted in a reactor with a hydrogen-rich gas to form substantial amounts of methane directly:

$$2CH_{0.8} + 1.2\ H_2 \rightarrow CH_4 + C \tag{1.5}$$

"The hydrogen-rich gas for hydrogasification is manufactured from steam utilizing the char leaving the hydrogasifier reactor. For example, the hydrogen may be produced in a steam-oxygen gasifier, as illustrated in Fig. 1.2.

"The block flow diagrams of Figs. 1.1 and 1.2 have the same overall reaction for the formation of methane from coal. The difference in the two simplified flow sheets is the location of the methane-producing steps. The key to the increased efficiency of the modern coal-to-high-Btu gas processes (Fig. 1.2) is hydrogasification (Reaction 1.5), in which appreciable quantities of methane are formed directly in the primary gasifier. The heat released by methane formation is at a high enough temperature level to be used in the steam-carbon reaction to produce hydrogen. Consequently, less oxygen is used to produce heat for the steam-carbon reaction, and less heat is lost in the low-temperature methanation step. These factors lead to a higher overall process efficiency (65 to 75% compared with 50 to 55% by synthesis-gas plus methanation)."

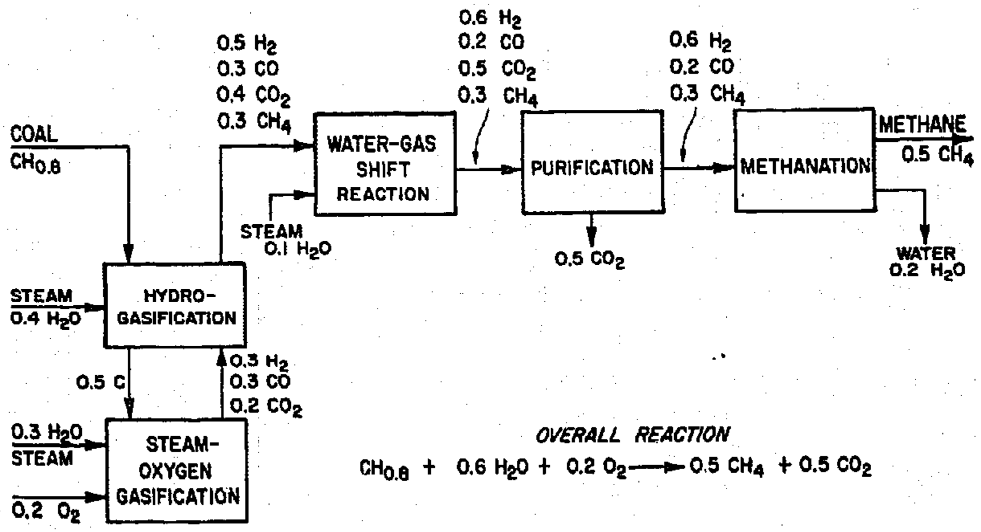


Fig. 1.2. Pipeline Gas Production Using Hydro-Gasification Plus Methanation

2 STANDARD PRACTICE

2.1 APPLICABLE CODES AND STANDARDS

In general, no standards or codes have been issued specifically for coal gasification plants. Components of such a facility would, where applicable, follow ASME codes for boilers and pressure vessels. Because of the similarity between a gasification facility and a petroleum refinery it is possible that existing new-source federal standards for control of gaseous and aqueous contaminants would apply. Noise generation and occupational safety are regulated inside the facility by OSHA; whereas, local noise codes may be applied at the plant boundaries.

2.2 STANDARD RATINGS

It is common practice to classify high-Btu coal gasification facilities by the number of standard cubic feet of SNG produced per day. When a significant portion of the energy produced is in another form (e.g., char, synthetic liquid fuels, or electricity), the facility may be "rated" on either an energy basis (Btu/day) or on a coal consumption rate (tons/ day).

2.3 DERATING FACTORS

One of the main factors that influence the heating value of the raw gas produced is the particular composition of the feed coal. The role of these characteristics can vary from significant in a particular gasifier bed type to insignificant in another. These effects may be generalized according to moisture, volatile matter, fixed carbon, sulfur, and ash contents.

2.3.1 Moisture

Coal moisture content will affect the heating value of the raw gas in a proportional manner. This is especially true in fixed-bed processes because this moisture is removed by the hot gases rising through the drying and devolatilization zones; thus the product gas contains more water vapor. In fluidized-bed units, an increase in the water content tends to increase the production of carbon dioxide. Entrained beds are particularly sensitive to coal moisture content because any moisture inhibits the overall gasification reaction, which must take place quickly in this type of bed.

2.3.2 Volatile Matter

Fluidized and entrained beds, compared with fixed beds, are less sensitive to the volatile matter content of coal because these compounds are gasified very quickly. In fixed-bed units, increases in volatile matter content cause an increase in the heating value of the gases because these are driven off in the devolatilization zone. In single-stage units, this volatile matter will be cracked and polymerized to form heavy tars and pitch that must be removed if the gas is not used directly.

2.3.3 Fixed-Carbon Content

Coals with high fixed-carbon content require more oxygen and steam per pound than do those coals with lower carbon content. This leads to an increase in the percentage of carbon monoxide and hydrogen produced (decreased CO_2 content), and thus increases the heating value of the gas. However, higher fixed-carbon content leads to lower volatile matter contents.

2.3.4 Sulfur Content

Sulfur in the coal will be in three different forms: pyrites (FeS_2), organic, or sulfates. During gasification, the organic sulfur and some of the pyritic sulfur will react with hydrogen to form hydrogen sulfide and with carbon monoxide to form carbonyl sulfide, thereby lowering the heating value of the gas.

2.3.5 Ash

Ash is the remaining inorganic material left after coal is subjected to complete combustion. The composition of this ash will determine the temperature at which its melting will occur. Because all commercially available fixed-bed gasifiers remove the ash in a solid, dry form, the maximum temperatures (and thus the product gas composition) allowed will be governed by this ash-softening temperature. In a fluidized-bed (HYGAS) gasifier, any softening of the ash will cause the bed particles to stick and result in a loss in fluidization. Conversely, in the entrained-bed unit (BI-GAS being the only unit available), the ash is liquefied and removed as a run-off. Therefore, for proper operation and maximum heating value of product gases, high ash-softening temperatures are preferred for fixed and fluidized beds; whereas, low values are preferred for entrained-bed units.

Coal, even taken from the same mine, exhibits wide variations in composition and water content. For example, Table 2.1 shows the range of constituents of a high volatile carbon bituminous Illinois #6 type coal, taken from Orient Mine #5.

This wide range of coal composition over the lifetime of the coal gasification facility (usually taken as 20 yr) can be compensated for by two different design methods. The first method of providing a nearly uniform composition coal would be to blend incoming coal with previously received coal in storage. Continuous sampling and analysis of different coal storage piles, with blending of coal when required, can provide control of coal feed composition and result in smaller fluctuations in coal compositions and rates.

Table 2.1. Range of Materials taken from Orient #5 Mine, Dry Basis[2]

Material	Range, wt %
Carbon	65.0 - 75.0
Sulfur	2.0 - 4.8
Hydrogen	4.3 - 5.4
Oxygen	6.5 - 11.0
Nitrogen	1.3 - 1.5
Ash	11.0 - 22.0

The second design method would provide equipment within the facility with operating flexibility to process variable-composition coal at a fixed feed rate. This design would result in higher capital costs because of the greater design capabilities of the process equipment.

Because moisture content of the raw feed coal can vary by ±5% by weight, water must be driven off at some point in the process. This heat of vaporization does not contribute to the overall thermal efficiency of the design. As a rough estimate, each 1.0 weight percent of water contained in the coal reduces the thermal efficiency by about 0.1 percentage point.

Thus, assigning the equivalent of a "derating" factor to a gasification facility to account for coal feed variation is a tenuous supposition and may be possible only after large commercial plants have provided considerable operating data.

3 PROCESS DESCRIPTION

Most high-Btu coal gasification systems can be categorized into three groups by the same technique used for contacting the solid stream (e.g., coal) with the gaseous streams (e.g., oxygen, steam, hydrogen). These three groups are:

(1) fixed- or moving-bed reactors (e.g., Lurgi);
(2) suspension or entrained reactors (e.g., BI-GAS); and
(3) fluidized-bed reactors (e.g., HYGAS and Synthane).

A description of each group follows.

3.1 FIXED- OR MOVING-BED REACTORS

3.1.1 General Discussion

Because the flow of coal and residue (ash) is countercurrent to the gasifying agents and products (principally carbon monoxide and hydrogen), fixed beds exhibit excellent thermal efficiencies. Outgoing ash heats the incoming gases, and the outgoing products heat the incoming coal. Moreover, the long residence times (typically 1 to 3 hr) involved for a coal particle allow high carbon conversion effeciencies.

Within a fixed bed are the following zones of progressively higher temperatures to which the incoming coal is subjected:

- *Drying Zone*. Raw coal (sized 1/4 to 1-1/2 in.) that is fed to the reactor comes in contact with the hot product gases; moisture in the coal is driven off.
- *Devolatilization Zone*. As the coal is heated further, occluded carbon dioxide and methane are driven off at temperatures less than 400°F. Organic sulfur in the coal is decomposed in the range of 400° to 900°F and is converted to hydrogen sulfide and other compounds. Nitrogen compounds in the coal decompose to release nitrogen and ammonia. Above 550°F, oils and tars are distilled from the coal.
- *Gasification Zone*. Char (the now-devolatilized coal) comes in contact with steam and the hot combustion products from the zone directly below. The chief reactions here are those of carbon monoxide and hydrogen being formed from the combination of carbon with water and carbon dioxide. These

reactions are endothermic, and production of carbon monoxide and hydrogen are favored at high temperatures; whereas, the production of carbon dioxide and hydrogen would be favored at lower temperatures.

- *Combustion Zone*. This zone supplies both the heat and carbon dioxide for the gasification zone. It consists of a layer of ash supporting the combusting, by now gasified, char. The key reaction in this zone is that of carbon with oxygen, producing heat and carbon dioxide. The ash bed acts as a distributor for the oxygen, or air, and steam and, more importantly, provides heat to incoming feeds.

Fixed-bed gasifiers can be classified further into single-stage and two-stage units. Both types will contain the zones described above; they differ in the location of gas removal and in the temperature ranges within the devolatilization and drying zones. A single-stage gasifier has only one product gas offtake, located at the top of the coal bed above the drying zone. Typical temperatures of gas leaving the unit are in the range of 700° to 1100°F. Thus, incoming coal is heated very rapidly, causing the oils and tars from the coal to crack and polymerize into heavy viscous tar and pitch. This violent distillation reaction also causes the coal to decrepitate and gives rise to coal dust, which is carried out with the product gas.

Two-stage producers have one gas offtake above the drying zone and one just at the top of the gasification zone. About half the gas produced by gasification is removed at this point; the remainder flows upward through the devolatilization and drying zones. Because temperatures attained in these two zones are considerably lower than those seen in single-stage units, the incoming coal is heated less violently, and the oils and tars are distilled in a much slower manner. Thus the problems in handling heavy tars, pitch, and soot are avoided.

3.1.2 The Lurgi Process[3,4,5,6]

This process has been commercially proven since 1936 when the first full-scale plant was constructed at Hirschfelde, Germany. Since then, 18 commercial plants have been constructed. Table 3.1 shows the location, size, and feed requirements of these 18 installations.

The Lurgi pressure gasifier, shown in Fig. 3.1 is basically a semi-continuous, slowly moving bed type gasifier of vertical, cylindrical construction. Crushed and screened coal, sized 1/8-in. to 1-1/2-in. is fed into the unit by a lock hopper mounted atop the gasifier. The gasifier, which operates at 350 to 450 psig, is surrounded by a water jacket that utilizes boiler feedwater to recover heat escaping from the gasifier shell. A motorized grate at the bottom of the gasifier withdraws any ash formed in the ash lock hopper.

Table 3.1. List of Lurgi Gasifiers[7,8,9,10]

Plant No.	Location	Year	Type of Coal	Gasifier I.D.	Capacity, 10^6 SCF/day	No. of Gasifiers
1	Hirschfelde, Central Germany	1936	Lignite	3 ft 9 in.	1.1	2
2	Bohlen, Central Germany	1940	Lignite	8 ft 6 in.	9.0	5
3	Bohlen, Central Germany	1943	Lignite	8 ft 6 in.	10.0	5
4	Most, CSSR	1944	Lignite	8 ft 6 in.	7.5	3
5	Zalzui-Most, CSSR	1949	lignite	8 ft 6 in.	9.0	3
6	Sasolburg, South Africa	1954	Subbituminous with 30% ash and more	12 ft 1 in.	150.0	9
7	Dorsten, West Germany	1955	Caking Subbituminous with high chlorine content	8 ft 9 in.	55.0	6
8	Morwell, Australia	1956	Lignite	8 ft 9 in.	22.0	6
9	Daud Khel, Pakistan	1957	High volatile coal with high sulfur content	8 ft 9 in.	5.0	2
10	Sasolburg, South Africa	1958 1958	Subbituminous with 30% ash and more	12 ft 1 in.	19.0	1
11	Westfield, Great Britain	1960	Weakly caking subbituminous	8 ft 9 in.	28.0	3
12	Jealgora, India	1961	Different grades	N/A*	0.9	1
13	Westfield, Great Britain	1962	Weakly caking subbituminous	8 ft 9 in.	9.0	1
14	Coleshill, Great Britain	1063	Caking subbituminous with high chlorine content	8 ft 9 in.	46.0	5
15	Naju, Korea	1963	Graphitic anthracite with high ash content	10 ft 5 in.	15.0	3
16	Sasolburg, South Africa	1066	Subbituminous with 30% ash and more	12 ft 1 in.	75.0	3
17	Luenen, GFR	1970	Subbituminous	11 ft 4 in.	1440 10^6 Btu/h	5
18	Sasolburg, South Africa	1973	Subbituminous with 30% ash and more	12 ft 4 in.	190.0	3

*N/A = not available.

The steam and oxygen that are introduced at the bottom of the Lurgi gasifier, shown in Fig. 3.1., pass upward through four distinct zones. The topmost portion of the gasifier acts as a preheat and coal drying zone where the coal comes into contact with the hot crude gases leaving the reactor. As the coal travels down the reactor, devolatilization starts where the bed temperature ranges from 1150° to 1600°F. After devolatilization, the resulting char undergoes gasification.

The combustion zone is the fourth and last zone located at the bottom of the gasifier. Heat for the endothermic devolatilization and gasification reaction is produced in this zone by the combustion of carbon and oxygen, which produces mostly carbon dioxide. Approximately 14% of the carbon in the coal that enters the top of the reactor reaches the bottom combustion zone. A relatively small amount of unburned carbon remains in the ash discharge.

Crude gases that exit the gasifier at the top are scrubbed and cooled to remove the entrained tar and oils. These gases are cooled further to approximately 360°F by a waste-heat recovery boiler. The resulting wash water from the scrubbing unit is mixed with the condensate from the waste-heat boiler. Tar in the wash water also contains a mixture of coal and ash dust which is returned to the gasifier for cracking and gasification after processing by a tar-gas-liquor separator.

Approximately 50% of the crude gas, which is saturated with steam, then is sent to a shift conversion unit where the following reaction takes place:

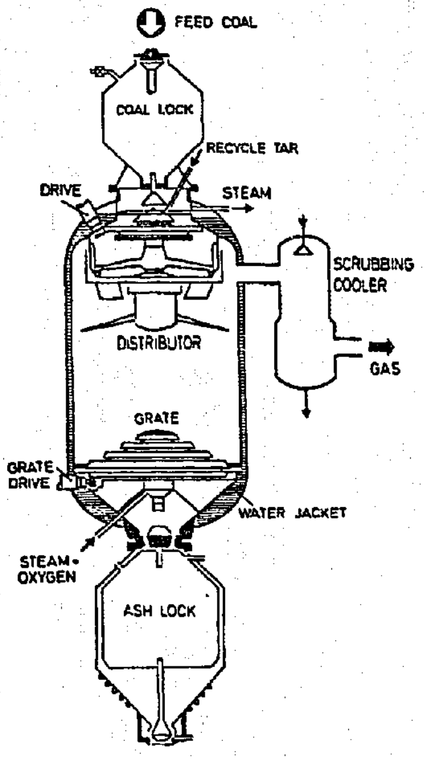


Fig. 3.1. Lurgi Pressure Gasifier

$$CO + H_2O \rightarrow H_2 + CO_2 \qquad (3.1)$$

This adjusts the H_2-to-CO ratio of the gas mixture to an optimum of 3.6 for the methanation reaction. Before the methanation reaction, removal of H_2S, CO_2, and COS is accomplished in a Rectisol unit. Because the physical absorption process uses low-temperature methanol (down to 50°F), all hydrocarbons having two or more carbon molecules are removed.

The resulting CO and H_2 are combined catalytically in a series of fixed-bed reactors, using a pelleted nickel catalyst, to produce CH_4 and H_2O. Heat, generated by this exothermic reaction, is used to generate most of the process steam for the gasification reaction. Approximately 50% of the methane in the product gas is made in the methanation section.

Crude tar and oils as well as anhydrous ammonia and crude phenols are produced in this process. Byproduct naphtha also is recovered in the cool-down step before the shift conversion, acid gas removal, and methanation. The block diagram for this process is shown in Fig. 3.2.

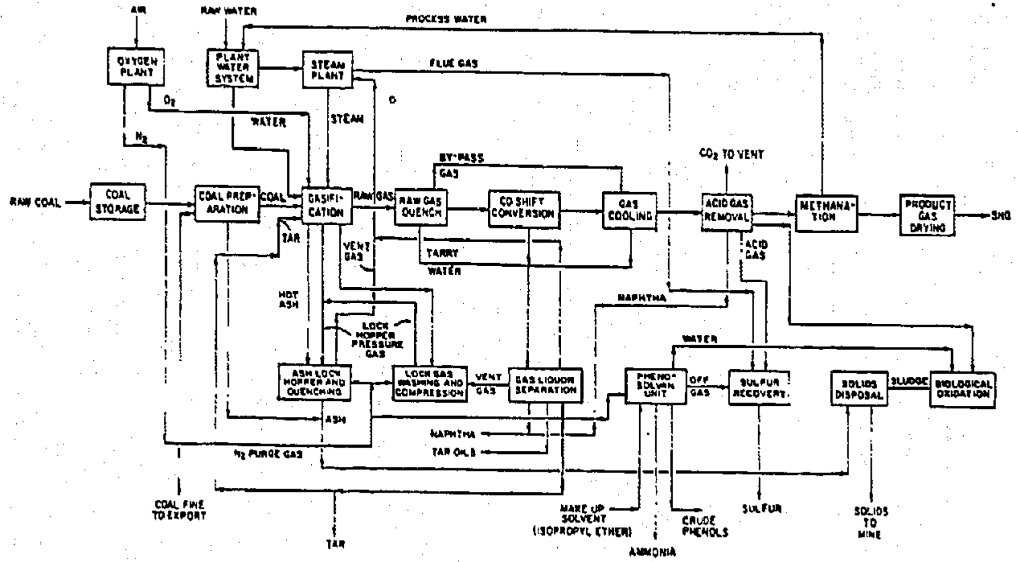


Fig. 3.2. Lurgi Coal Gasification Process

3.2 ENTRAINED FLOW REACTORS

3.2.1 General Discussion

In the conventional, entrained flow reactor, solid and gaseous streams are contacted concurrently, and high temperatures are required to achieve the complete reaction of coal and gases in relatively short periods of time. To achieve high temperature in the reactor requires large amounts of oxygen. Oxygen is introduced through nozzles located near the bottom of the reactor vessel. Ash is removed from the gasifier as slag. Operations of this type are not generally sensitive to the type of coal, i.e., non-caking or caking.

3.2.2 The BI-GAS Process[6,11,12]

Work on the BI-GAS process, initiated in December, 1963, by Bituminous Coal Research, confirmed the belief that coal and steam could be reacted at high pressures and temperatures to produce a high methane yield. Recent development work has produced a fully-integrated 120-ton/day pilot plant at Homer City, Pennsylvania. The pilot plant, in operation since early 1977, has provided design data for the possible construction of a commercial plant.

To react pulverized coal and steam, the BI-GAS process uses an entrained flow, two-stage oxygen-blown gasifier shown in Fig. 3.3. Raw coal is pulverized,

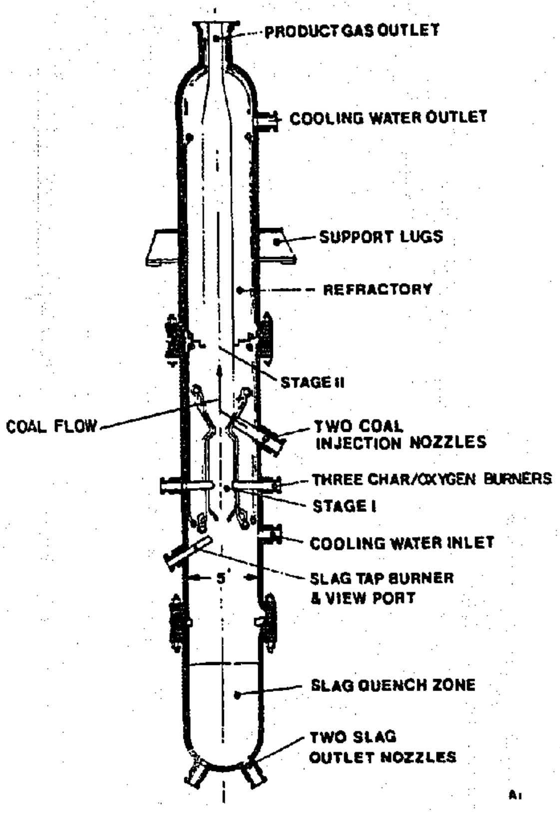


Fig. 3.3. The BI-GAS Gasifier

70% through 200 mesh, and processed to form a water slurry containing as much as 60% solids. A high-pressure slurry pump transports the coal slurry to a spray dryer where recycle gas and hot slurry are contacted for vaporization of the coal surface moisture. The pulverized coal is entrained in the gases and transported to a cyclone at the top of the gasifier where the coal flows by gravity into the bottom of the upper (Stage II) stage.

In Stage II, the entering coal is combined with steam and hot synthesis gas, coming from the bottom (Stage I) stage, at 2200°F to attain rapid conversion of coal to methane, synthesis gas, and char. The residual char is separated from the exiting gas stream and recycles to Stage I of the gasifier. In Stage I, the char is completely gasified under slagging conditions with oxygen and steam, producing the heat required for the endothermic reactions in Stage II.

The raw product gas from the upper stage is first shifted and then stripped of CO_2 and H_2S. The final step of the process is catalytic methanation to upgrade the gas to pipeline quality. Figure 3.4 shows a block diagram of the BI-GAS process.

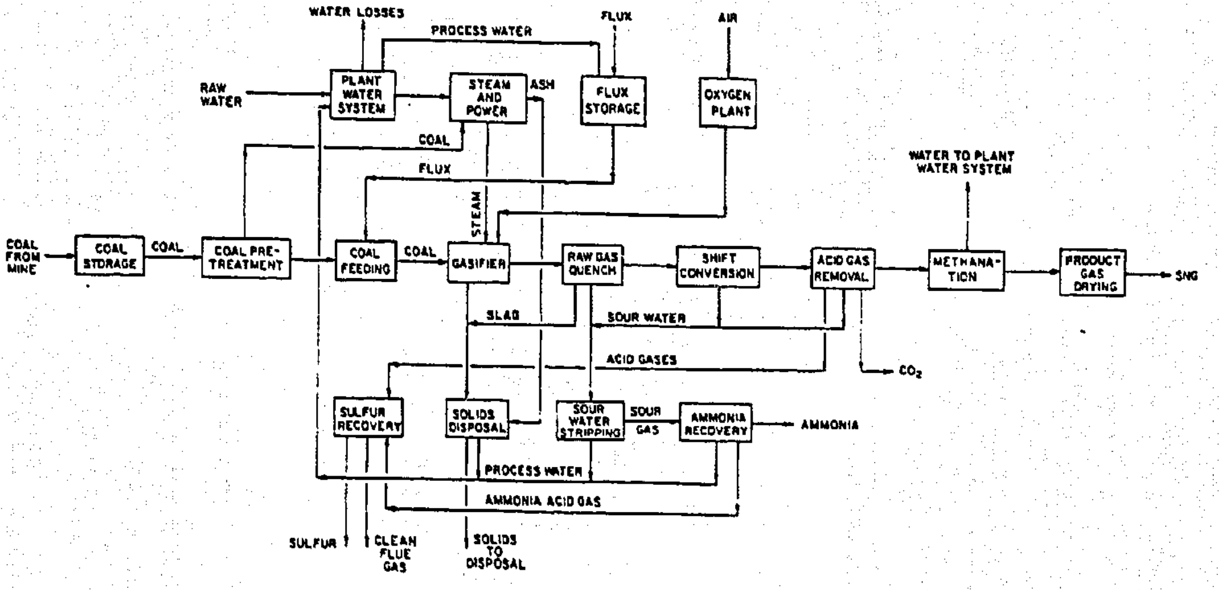


Fig. 3.4. Preliminary Block Flow Diagram for the BCR BI-GAS Process

3.3 FLUIDIZED-BED REACTORS

3.3.1 General Discussion

The fluidized-bed type gasifier allows intimate mixing and contacting of gas and solids and provides a relatively long residence time compared with entrained-bed gasifiers. The gasifying medium (e.g., oxygen, steam, hydrogen) is fed through a bottom distributor plate and acts as the fluidizing medium. Dry ash may be removed continuously from the fluid bed, or the gasifier can be operated at a higher temperature than the ash-softening temperature. This operation will allow removal of ash as agglomerates that are too heavy to fluidize, thereby improving the carbon conversion and reducing the residence time to some extent.

3.3.2 The Synthane Process[6,13,14]

The Synthane gasifier concept was initiated in 1961 by the Pittsburgh Energy Research Center. This process is an off-shoot of research in methods of pretreating caking coals in a fluidized-bed reactor. Early work on the Synthane reactor combined the original two fluidized-bed concepts into a single vessel construction wherein coal pretreatment, carbonization, and gasification occur. Construction of the Synthane 72-ton/day pilot plant was completed in the spring of 1975 and testing currently is under way. Figure 3.5 illustrates the main steps of the process. In this

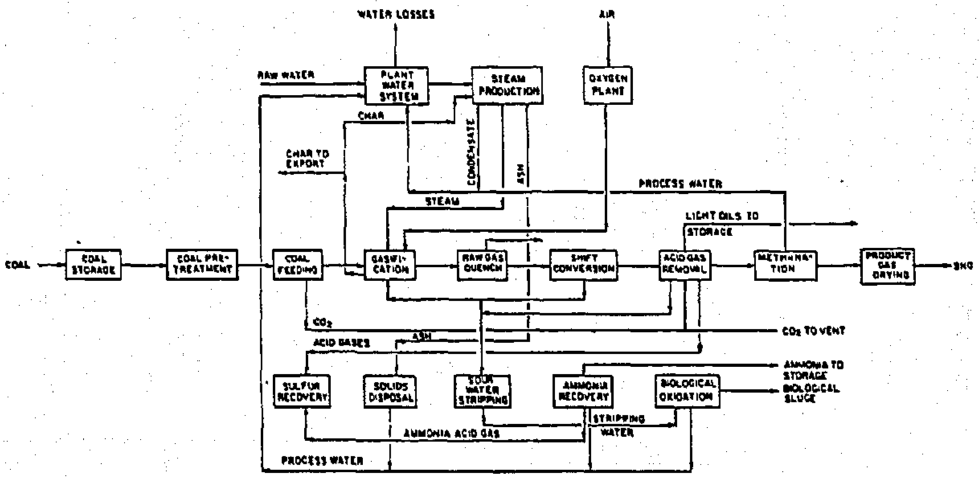


Fig. 3.5. Synthane Coal Gasification Process

process, coal is first crushed to a 20 x 10 mesh size, pressurized to 1000 psi in lock hoppers, and then delivered to the coal pretreater. Here the caking quality of the coal is reduced to a free-swelling index of 1, or less, in the pretreatment fluidized-bed reactor. The pretreator accomplishes this task with oxygen and steam (approximately 0.08 lb of O_2 and 0.27 lb of steam/lb of coal) treatment at 800°F. The mixture of decaked, and partially devolatilized coal, along with any excess steam and volatile matter separated from the coal, then is fed into the top of the fluidized-bed gasifier. In the gasifier section shown in Fig. 3.6, the free-falling coal particles are in countercurrent contact with the raw synthesis gases that rise from the gasifier bed. Devolatilized coal that reaches the bottom of the reactor is gasified with 0.2 lb of oxygen and 0.1 lb of steam/lb of coal at 1800°F. The remaining char, which contains 30% of the original carbon in the coal, is discharged through a lock hopper system to atmospheric pressure. Char, tar, and dust recovered from a lock hopper and raw synthesis gas are burned to generate the steam and power needed in the plant.

The raw gas leaves the gasifier at the top byway of an internal cyclone, and is water-washed to remove residual dust, tar, and excess water vapor. Resulting cleaned gas then is processed in a shift reactor to produce a 3:1 H_2/CO ratio. After this step, hot carbonate scrubbing removes the CO_2 and H_2S, and an activated carbon treatment removes sulfur to < 0.1 ppm before the final catalytic methanation reaction is accomplished.

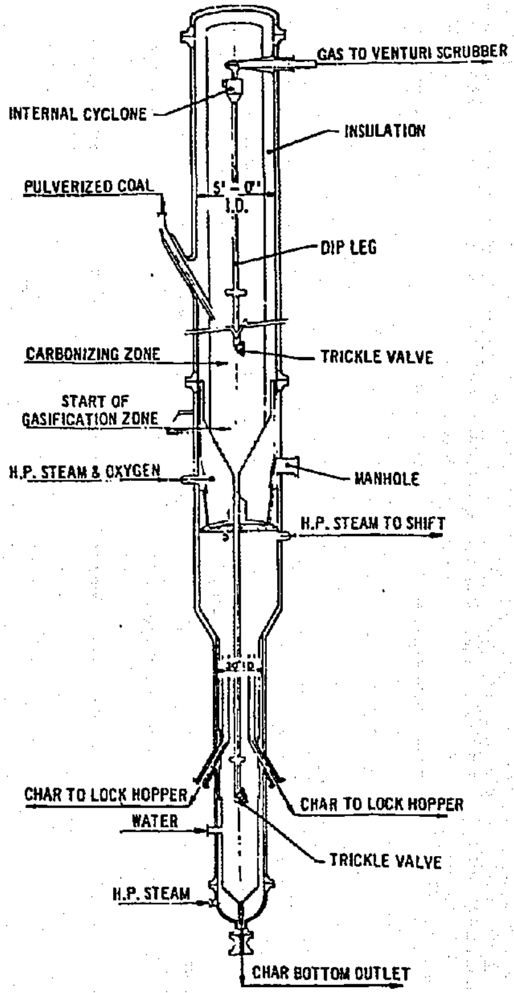


Fig. 3.6. Synthane Gasifier

3.3.3 The COGAS Process[4,15,16,17]

The COGAS Process was developed in the mid-1970s with private funding by the project partners: FMC Corp.; Consolidated Gas Supply Corp.; Tennessee Gas Pipeline Co.; and Panhandle Eastern Pipeline Co. This process is an off-shoot of work performed by the FMC Corp. to study pyrolysis in the Char Oil Energy Development (COED) Project, conducted in the late 1960's. Pilot-plant tests in Leatherhead, England, have shown that substantial quantities of SNG and oil can be produced using this process. A recent contract with the Illinois Coal Gasification Group (ICGG) calls for design, construction, and operation of a demonstration plant capable of processing 2200 tons/day of high-sulfur coal in southern Illinois.

In the COGAS process, a series of fluidized-bed reactors as shown in Fig. 3.7, is used to process crushed coal into gas, synthetic crude oil, and char. Four fluidized beds are operated at 600°, 850°, 1000°, and 1500°F. These temperatures are selected to keep the coal from agglomerating and defluidizing the reactor bed. For nonagglomerating coals, only three stages are required.

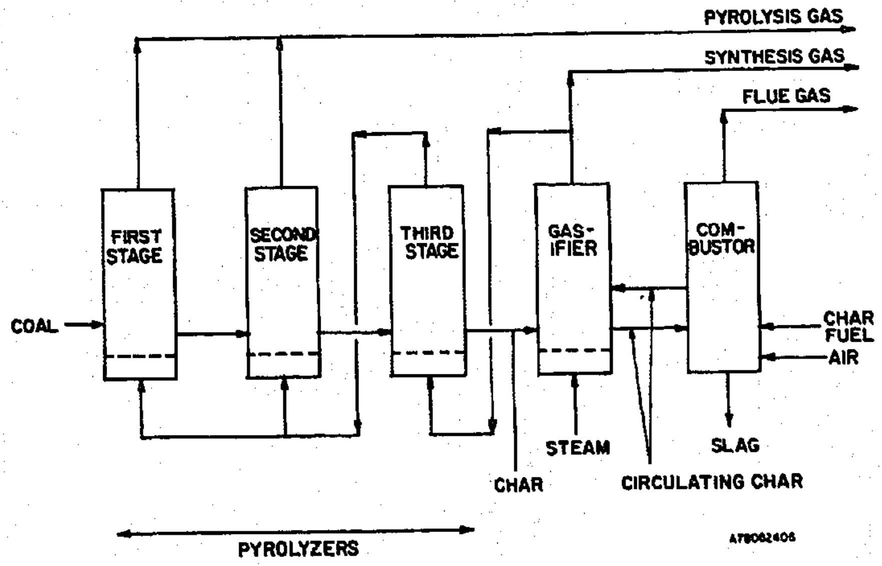


Fig. 3.7. COGAS Pyrolysis and Gasification Reactors

In the first stage of the COGAS Process, coal is fluidized and pyrolyzed by a portion of the gases from the second and third stages. Char from the first fluidized bed is fed into the second fluidized bed and further pyrolyzed and fluidized by gases from the third fluidized bed. This step is repeated by sending the char from the second to the third bed. In the third fluidized bed reactor, the char is pyrolyzed by hot synthesis gases produced by a gasifier. This gasifier converts the remaining char from the third fluidized bed into synthesis gas by steam injection at the bottom of the fluidized-bed gasifier. Heat for the endothermic synthesis reaction is obtained by circulating a hot char stream from combustor into the gasifier. Char for the combustor is recycled from the gasifier.

Use of pyrolysis promotes oil formation by eliminating the coal preoxidation step. The fluidized-bed reactors operate between 6 and 10 psig, depending on the type of coal used. Typically, 1 barrel of oil and 8000 SCF of gas are produced per ton of coal. Figure 3.8 shows a block flow diagram of the COGAS Process.

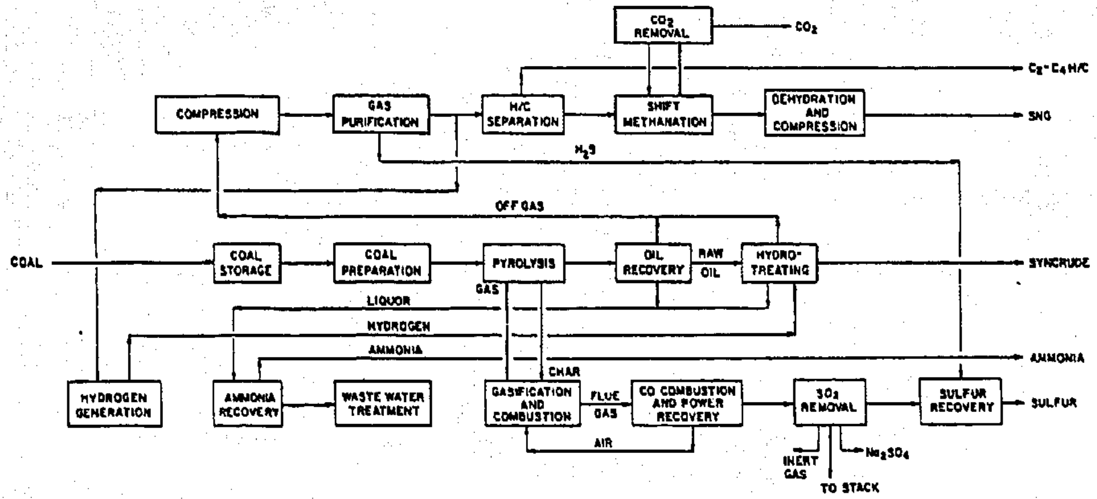


Fig. 3.8. ICGG COGAS Coal Gasification Process

3.3.4 The Steam-Oxygen HYGAS Process[4,6,18,19]

The HYGAS Process has been under investigation by the Institute of Gas Technology (IGT) since 1945. In 1964, further research was sponsored jointly by the American Gas Association, Inc. (A.G.A.) and the Office of Coal Research (OCR), U.S. Department of the Interior. This process is currently in the pilot-plant stage.

The HYGAS process uses the fluidized-bed hydrogasification reactor, shown in Fig. 3.9, to accomplish gasification. The hydrogasifier reactor vessel has four internally connected reaction stages and is 35 ft high with a 5.5-ft i.d. The top stage has a diameter of 2 ft and a fluidized bed 10 ft high that receives an oil coal slurry as a spray. Sensible heat in

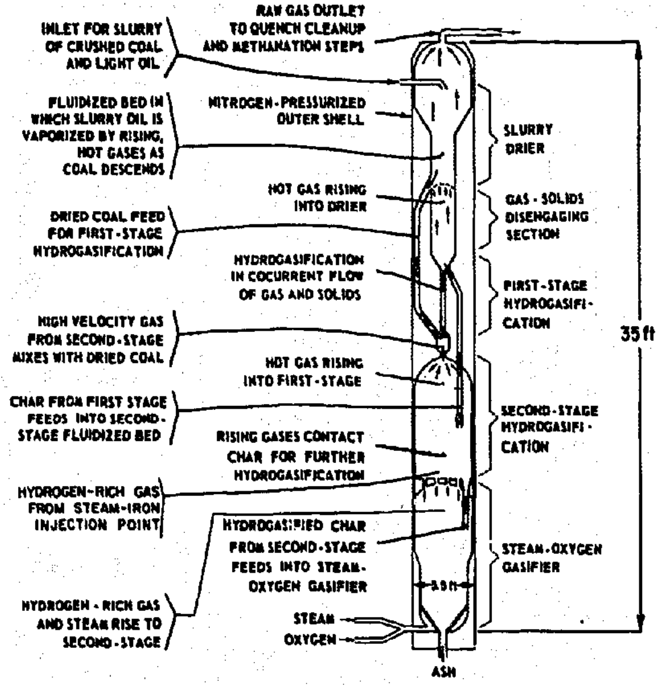


Fig. 3.9. The Steam-Oxygen HYGAS Gasifier

the gaseous reaction products is transferred very efficiently to the slurry. This step drives off the oil as a vapor and leaves dry coal to be fed to the second stage. At this point, the dry coal is at a temperature of about 570°F and at 1000 psi pressure.

From a point near the bottom of the drying bed, the solids flow down through a 3-in.-diameter pipe into a small hopper (lift pot) from which the solids are lifted 20 ft through a 3-in.-diameter pipe into a disengaging conical vessel, 18 in. in diameter. The hot gases (1700°F) from the reaction section below provide:

(1) the lifting force,
(2) the heat to raise the solids temperature to 1200°F, and
(3) the hydrogen to react about 20% of the coal to methane.

This is the first stage of hydrogasification. The first 10 to 20% of the coal can be gasified in the first 3 to 10 sec of Stage I. Although the highest temperature in this stage is only 1200°F, the primary product is methane. The reactions occur in the dilute phase in the lift line. Two 3-in.-diameter pipes are connected near the bottom of the conical disengaging section. One of these can carry the partially reacted coal back to the lift pot where the partially reacted char can be mixed with the incoming fresh coal. The rapid flowrate of coal in the dilute phase, admixed with reacted char, eliminates the tendency for agglomeration and sticking.

The second pipe from the disengaging cone conveys the partially reacted coal down to a high-temperature bed for further gasification. This bed is heated to about 1700°F, fluidized, and gasified by the hydrogen-rich gas rising from the heat-exchanger bed below. This is referred to as the second-stage hydrogasifier. The reactor, which is 30 in. in diameter and 26 ft high, is internally insulated.

At 1700°F the hydrogen reacts in Stage II with the char to produce methane and heat. Steam reacts with the char to produce carbon monoxide and hydrogen and uses up heat. If the temperature rises, the steam-char reaction rate increases, uses more heat, and prevents the temperature from rising any further. If the temperature drops, the steam-char reaction slows down, and the temperature drop is arrested. The steam not only provides this automatic temperature control but also provides additional hydrogen through its reaction. The heat of methane formation produces hydrogen which would otherwise be more difficult to recover. An additional 25% of the coal is converted in this reaction stage, making the total conversion about 45%

From this reaction stage, the char descends through a 3-in.-diameter pipe into a final fluidized-bed section that acts to preheat the incoming hydrogen-carrying gas as it cools the char. This stage is similar in configuration to the second stage. The diameter decreases to 2 ft at the very bottom. Figure 3.10 shows the HYGAS Process.

3.3.5 The IGT Steam-Iron HYGAS Process[6,20,21]

The Steam-Iron Process is used in conjunction with the HYGAS process to supply a mixture of hydrogen and steam to the HYGAS reactor to enhance the hydrogasification process.

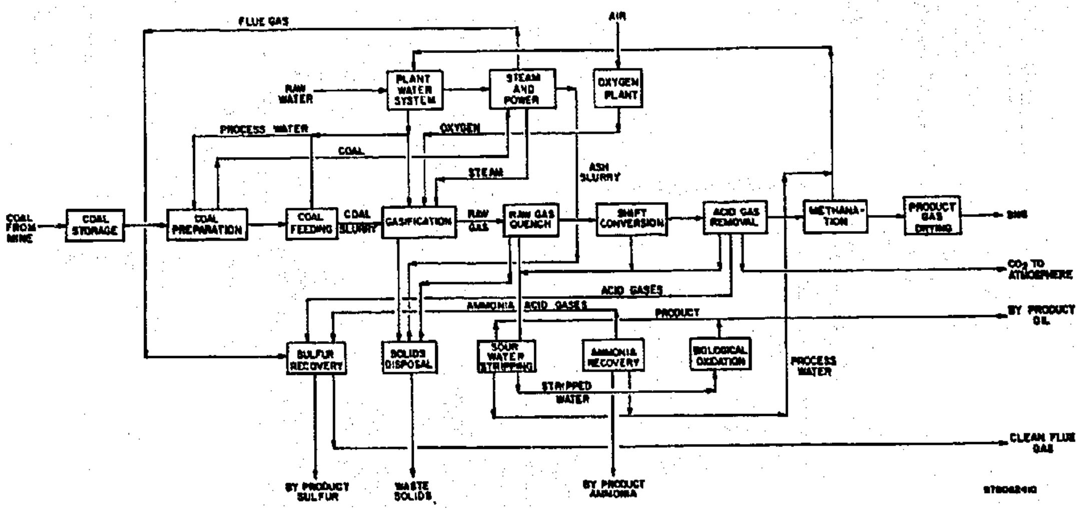


Fig. 3.10. IGT Steam-Oxygen HYGAS Coal Gasification Process

Production of hydrogen from coal by the old cyclic Steam-Iron Process once was a major source of hydrogen in the United States. In this process, a packed bed of iron ore was alternately reduced with producer gas containing carbon monoxide and hydrogen and oxidized with steam to intermittently produce hydrogen. Because of its cyclic nature, the process was limited to low-pressure operation. With the advent many years ago of cheap natural gas in the United States, the cyclic Steam-Iron Process was abandoned in favor of steam reforming of natural gas.

In the 1950s and 1960s, IGT and the U.S. Bureau of Mines revived interest in the Steam-Iron Process for producing hydrogen from coal gasification. Its process concept incorporated solids recirculation through the reduction and oxidation zones, thereby eliminating the need for cyclic operation, which would not be feasible at the high pressures required for coal hydrogasification. In 1976, construction of a 1.1 million SCF/day of hydrogen pilot plant was completed at IGT and has demonstrated the feasibility of the Steam-Iron Process.

Figure 3.11 shows a block flow diagram of the HYGAS Process incorporating the Steam-Iron Process. In this concept, the residual char from the

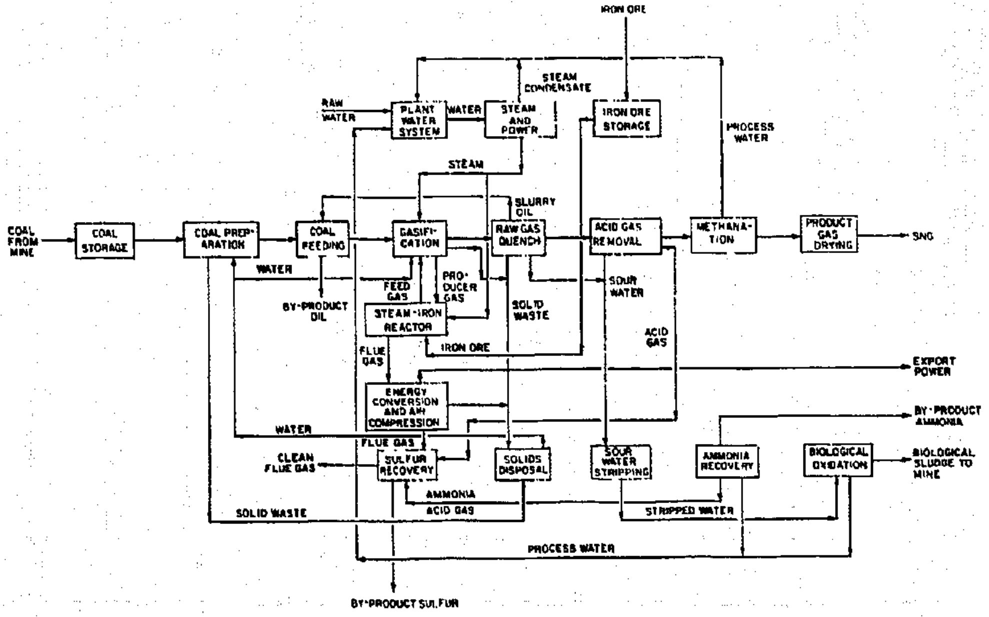


Fig. 3.11. IGT Steam-Iron HYGAS Coal Gasification Process

last hydrogasification step of the HYGAS Process is reacted with steam and air to generate a producer (reducing) gas for the Steam-Iron reactor (shown in Fig. 3.12). The Steam-Iron reactor, in turn, delivers a hydrogen-steam mixture back to the hydrogasifier from the oxidizer. Raw gases from the hydrogasifier are upgraded to synthetic gas in the usual downstream purification and final methanation stages. The spent producer gas from the Steam-Iron reactor is expanded to recover energy for inlet air compression, other plant utilities, and byproduct electric power.

In the producer reactor, the hydrogasified char is reacted with steam and air in a nonagglomerating fludized bed at temperatures in the range of 1900° to 2000°F, depending on the type of coal. Simplified forms of the primary chemical reactions in the producer reactor are:

$$C + H_2O \rightarrow CO + H_2 \qquad (3.2)$$

$$C + O_2 \rightarrow CO_2 \qquad (3.3)$$

Reaction 3.3 is highly exothermic and supplies the heat required for Reaction 3.2, the steam-char reaction. The producer gas generated in this reactor consists primarily of H_2, CO, CO_2, and N_2. It is desirable that the reducing ability of the producer gas be as high as possible to maximize the $(CO + H_2)/(CO_2 + H_2O)$ ratio. In the example discussed above, feed to the producer reactor was hydrogasified. However, operation of the producer reactor is not limited only to hydrogasified char All types of coals and carbonaceous feeds can

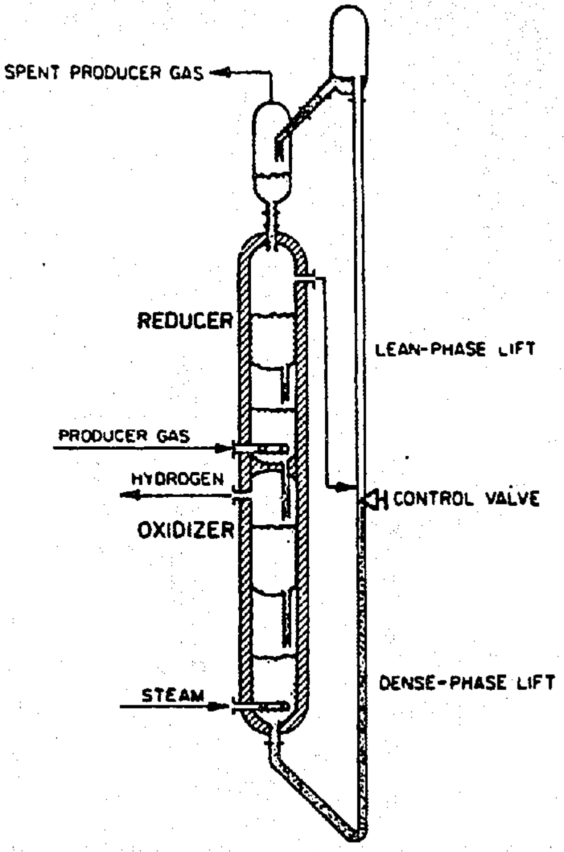


Fig. 3.12. Steam-Iron Reactor

be gasified to make producer gas. Caking coals require pretreatment before feeding into the producer reactor.

In the Steam-Iron reactor, the major reactions are based on reversible reduction and oxidation of iron oxides. Three iron phases enter into these reactions — magnetite (Fe_3O_4), wustite (FeO), and metallic iron (Fe). The highest oxidation form, hematite (Fe_2O_3), is not produced in the system. The following reduction and oxidation reactions occur with the iron and iron oxides:

Reduction

$$Fe_3O_4 + H_2 \rightarrow 3\ FeO + H_2O \tag{3.4}$$
$$Fe_3O_4 + CO \rightarrow 3\ FeO + CO_2 \tag{3.5}$$
$$FeO + H_2 \rightarrow Fe + H_2O \tag{3.6}$$
$$FeO + CO \rightarrow Fe + CO_2 \tag{3.7}$$

Oxidation

$$Fe + H_2O \rightarrow FeO + H_2 \tag{3.8}$$
$$3\ FeO + H_2O \rightarrow Fe_3O_4 + H_2 \tag{3.9}$$

Because of the cyclic nature of the reactions, the recirculating solids stream tends, after several cycles, toward a high FeO concentration. When steady-state chemistry has been achieved, all Fe and Fe_3O_4 produced are subsequently consumed in their respective reverse reactions. Furthermore, process kinetics and thermal requirements dictate that the amounts of Fe and Fe_3O_4 produced be very small relative to the FeO content. Approximately 10% of the total iron content is converted to FeO in actual cyclic operation of iron ore where high conversions to Fe are necessary, the Steam-Iron Process encounters only low concentrations of Fe. This low concentration of Fe and high concentration of nonsintering-type solids in the iron ore prevent sintering problems in the Steam-Iron reactor.

As shown in Fig. 3.12, four fluidized beds are used to achieve full countercurrent operation. The producer gas, at a temperature ranging from 1900° to 2000°F, enters the steam-iron reactor as feed to the lower reducer stage where, because of its high reducibility, FeO is converted to Fe.

Because of its high temperature, the producer gas also serves as a heat source for the endothermic reduction reactions. The producer gas reaches a conversion level of about 20% in the lower stage before entering the upper stage, where the bulk of the final conversion level of the producer gas (65%) is attained by the conversion of Fe_3O_4 to FeO. Thus, high reducing-gas conversion is attained, and sufficient metallic iron is produced to ensure the necessary hydrogen production.

Steam fed to the lower oxidizer stage converts FeO to Fe_3O_4. Steam conversion in this bed is limited to about 30% by equilibrium. Further conversion (above 30%) in the upper oxidizer stage must be effected by conversion of Fe to FeO. Although the equilibrium limit of about 66% steam conversion is possible at 1089°K (1500°F), conversion is limited by the amount of Fe formed in the lower reducer stage. For use as feed in coal gasification, the production of Fe is controlled to produce a final steam conversion of about 45%.

4 MATERIAL BALANCE

4.1 PRIMARY MATERIAL INPUT/OUTPUT

Typical primary material balances for various gasification processes, shown in Figs. 4.1 through 4.6, are based on coal input of 100 units, by weight. The coal is as-mined and thus contains — in addition to coal — water, minerals, refuse, etc. This is the type of coal encountered by the first piece of coal preparation equipment, be it for drying, grinding, or cleaning. Water used for the various processes, is raw river water; all necessary separation treatment equipment are integral to the plant. Only the exported, saleable products are shown as outputs and are discussed in detail under Sect. 4.3, *Product Material Specifications*. Solid, liquid, and gaseous effluents to the environment are discussed under Sect. 6.0, *Environmental Effects*.

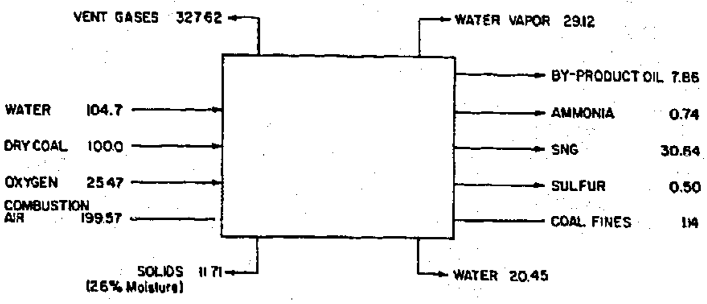


Fig. 4.1. Single-Feed Lurgi Material Balance[6]
(Basis: Coal input = 100, by weight)

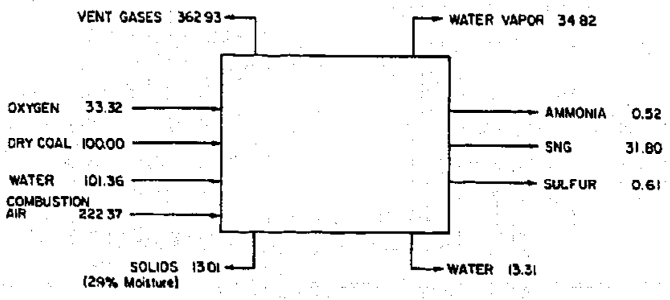


Fig. 4.2. BCR BI-GAS Material Balance[6]
(Basis: Coal input = 100, by weight)

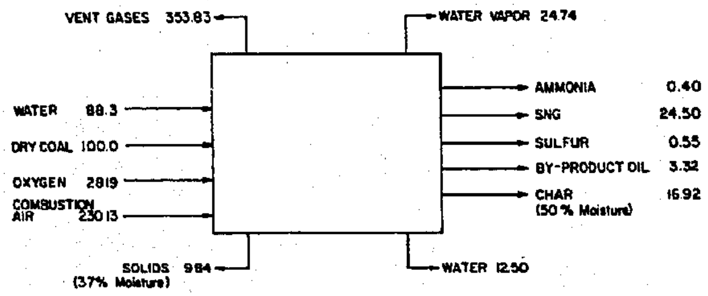


Fig. 4.3. PERC Synthane Process (Slurry Feed, Export Char) Material Balance[6]
(Basis: Coal input = 100, by weight)

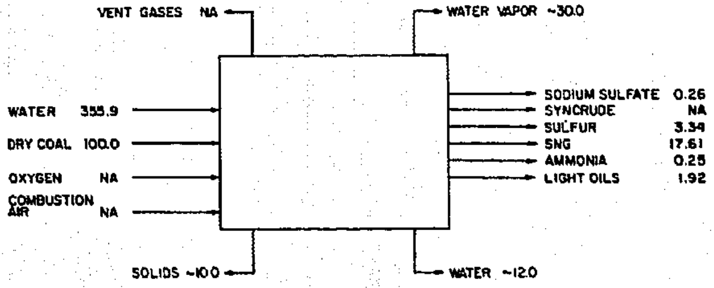


Fig. 4.4. COGAS Process Material Balance[17]
(Basis: Coal input = 100, by weight)

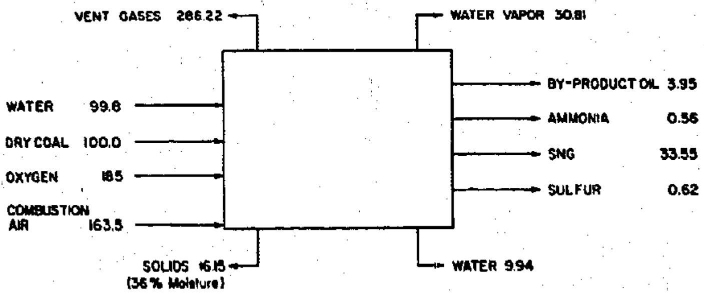


Fig. 4.5. IGT Steam-Oxygen HYGAS Material Balance[6]
(Basis: Coal input = 100, by weight)

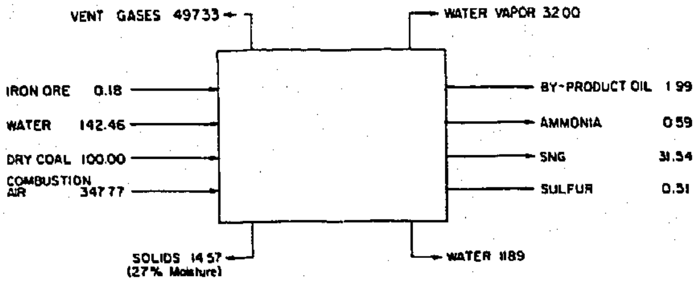


Fig. 4.6. IGT Steam-Iron HYGAS Material Balance[6]
(Basis: Coal input = 100, by weight)

4.2 MATERIAL SPECIFICATIONS

Approximately one-fourth of the world's known coal reserves is located in the United States. As shown in Fig. 4.7, about 14% of the land in the Continental States is underlaid with coal-bearing deposits. This amounts to an estimated 3.968 billion short tons of coal, nearly half of which is ranked as bituminous. In general, coal has a carbon content ranging from 60 to 85 weight percent.

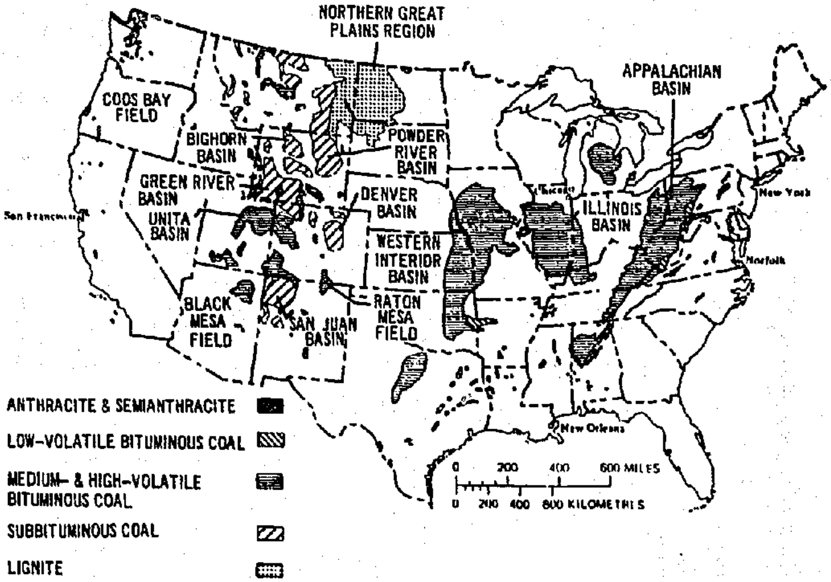


Fig. 4.7. Coal Fields of the Continental United States[22]

Coal in these formations generally lies from 0 to 3000 ft below the surface. In some places, such as in central and northern Wyoming, the coal-bearing rock is located as deep as 20,000 ft below the surface. Approximately 91% of coal formation is located in less than 1000 ft of overburden.

In general, a plant that converts a substance to another form is designed for a specific feed composition, which sets both the oprating condition and yields of the process.

Because of their vast quantities of coal, fields west of the Mississippi River have received considerable attention for energy conversion processes. As seen in Fig. 4.8, five of the top six coal-bearing states in the United States lie west of the Mississippi. For this reason alone, many coal gasification processes have been as thoroughly evaluated on Western coals as on Eastern coals. Of the processes evaluated here, only the COGAS Process characterization is based on an Illinois bituminous coal. The remainder of the processes are characterized for a Western subbituminous coal from Montana. In contrast to the Illinois coal, Western coals are typically low in sulfur and therefore do not present as much of a pollution problem. One additional point that may influence the gasification plants' location, and thus the coal source, is the availability of water. Unlike water-rich Eastern states, the coals of the Western states are located in a water-poor region. For this region, special

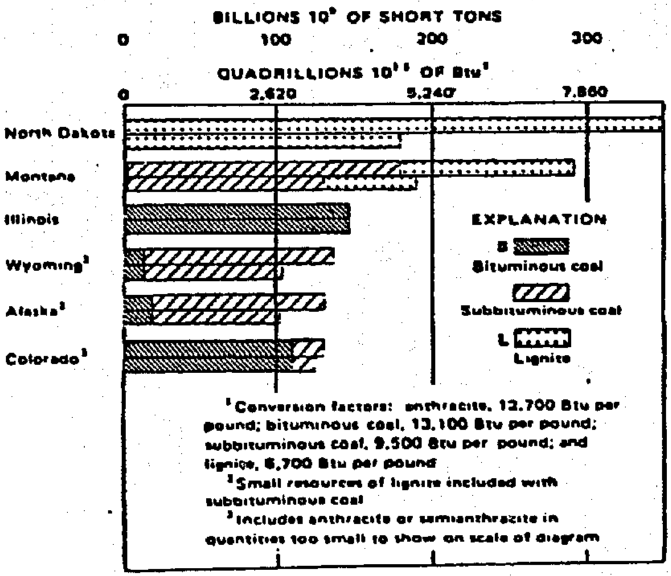


Fig. 4.8. Coal Resources of the Top U.S. States[22] (Top bar indicates tonnage and bottom bar indicates heating value)

consideration must be given to site selection for coal gasification facilities because of their vast water requirements.

Several proposed coal gasification facility designs have been based on a captive coal mine incorporated within the overall complex. Because of a lack of consistency and the variables involved in incorporating a captive coal mine into the designs that do not have this feature, it was assumed, for this evaluation, that the coal gasification facilities are near the coal mine. In either case, the mined coal composition can be stated with reasonable accuracy, and a process design based on this feed composition can be made.

Coal may contain water in two forms: inherent and surface moisture. Surface moisture may be reduced considerably by simply crushing the coal; whereas, inherent moisture requires the application of heat. Moisture content of the coal affects the overall thermal efficiency of the plant, because both crushing and drying operations, which are included in the process designs, consume energy. Therefore, in the following tables, the coal proximate analysis will be stated before any water is removed.

Table 4.1. Montana Subbituminous Coal Properties

Proximate Analysis as Received, wt %	
Moisture	22.0
Volatile Matter	29.4
Fixed Carbon	42.6
Ash	6.0
	100.0
Ultimate Analysis (dry), wt %	
Carbon	67.70
Hydrogen	4.61
Nitrogen	0.85
Oxygen	18.46
Sulfur	0.66
Ash	7.72
	100.00
Heating Value of Dry Coal	
Btu/lb (HHV)	11,290
Heating Value of Coal as Received	
Btu/lb (HHV)	8,800
Hardgrove Grindability Index	52
Fusibility of Ash °F	
In Reducing Atmosphere	
Initial Deformation	2,226
Softening Temperature	2,240
Hemispherical Temperature	2,254
Fluid Temperature	2,271
In Oxidizing Atmosphere	
Initial Deformation	2,284
Softening Temperature	2,299
Hemispherical Temperature	2,320
Fluid Temperature	2,388
Chlorine Content, wt %	0.0

Ash Analysis, wt%	
SiO_2	41.0
Al_2O_3	15.8
TiO_2	0.6
Fe_2O_3	6.2
CaO	14.4
MgO	4.6
Na_2O	1.1
K_2O	0.5
P_2O_5	0.4
SO_3	14.7
Form of Sulfur as % of Total Sulfur	
Pyritic	41
Sulfate	5
Organic	54
	100
Size Consist, % Below	
1-1/4 in.	100.0
1	96.7
3/4	87.1
5/8	76.5
1/2	65.5
3/8	56.4
1/4	36.6
1/8	-
6 mesh	26.9
8	21.6
10	19.1
18	11.3
28	-
100	2.3
200	1.3

Coal specifications are shown in Tables 4.1 and 4.2. Table 4.1 illustrates the Montana coal characteristics for all of the process designs except

Table 4.2. Coal Property Range of Typical Illinois #5 and Illinois #6 Coals Utilized in the COGAS Process Design[23] (Proximate Analysis, wt %)

Harrisburg-Springfield (Illinois #5) Coal (as received basis)	
Moisture, %	8-13
Volatile Matter, %	35-38
Fixed Carbon, %	40-44
Ash, %	9-12
Sulfur, %	4-5
Calorific Value, Btu/lb*	11,000-11,400
Rank Index	124-135
Ash Fusion, °F	N/A
Herrin (Illinois #6) Coal (as received basis)	
Moisture, %	8-12
Volatile Matter, %	35-39
Fixed Carbon, %	38-44
Ash, %	9-13
Sulfur, %	1-4
Calorific Value, Btu/lb*	10,800-11,300
Rank Index	124-133
Ash Fusion, °F	1,920-2,610

*Calorific value of moist coal on a mineral-matter-free basis.

the COGAS Process. Table 4.2 represents the analysis of typical Illinois No. 5 and Illinois No. 6 coals that would be used in the COGAS Process. The IGT Steam-Iron HYGAS Process requires, in addition to coal, iron ore for the production of hydrogen used in the hydrogasification reaction.

In general, the water required by the gasification plants has no specifications as to quality. However, for all but the COGAS Process, the coal gasification facilities were designed to utilize fresh water with the qualities listed in Table 4.3. This raw water is treated within the plant before utilization in the various components and is assumed to be withdrawn from a nearby river or lake.

Table 4.3. Assumed Raw Water Properties[a][6]

Total dissolved solids, ppm	496
Total hardness, ppm $CaCO_3$	232
Conductivity	-
Calcium, ppm $CaCO_3$	150
Magnesium, ppm $CaCO_3$	82
Sodium, ppm $CaCo_3$	148
Potassium, ppm $CaCO_3$	3.7
Iron, ppb Fe	20
Carbonate, ppm $CaCO_3$	0
Bicarbonate, ppm $CaCO_3$	145
Sulfate, ppm $CaCO_3$	224
Chloride, ppm $CaCO_3$	16
Fluoride, ppm F	0.5
Nitrate, ppm $CaCO_3$	1.5
Color, cobalt platinum units	4
Turbidity, Jackson turbidity units	270
Temperature, °F average, (range)	48(35-75)
pH	8.3
Silica, ppm SiO_2	12

(a) These data do not reflect raw water properties for the COGAS Process. Because of a lack of published information, the COGAS raw water requirements have not been characterized.

4.3 PRODUCT MATERIAL SPECIFICATIONS

Products from the various gasification processes range from substitute natural gas to synthetic crude oil. Some processes, e.g., the COGAS Process, yield a wide range of products; others, e.g., the BI-GAS Process, produce a single fuel product. This spectrum of product differentiation may be attributed either to fundamental differences in the gasification approach, or to the addition of equipment that would further refine the product materials. In general, all gasification facilities evaluated here are designed to produce a finished product, i.e., substitute natural gas, that can be sent directly to the distribution or transmission network. Other products, such as syncrude, will require further processing outside of the gasification facility to convert them into a finished marketable form.

Although no commercial coal gasification plants have yet been built in the United States, some of the processes are at a stage of development such that the product characteristics can be projected with some certainty, based on the operation of smaller process units; and their application or use can be defined.

4.3.1 Lurgi Process (Single-Feed Case)[6]

The hydrocarbon products of the Lurgi Process include substitute natural gas, tar, tar oil, and naphtha. In addition to these, ammonia and sulfur also are produced.

- ***Substitute Natural Gas (SNG) (Lurgi Process)***

Substitute Natural Gas (SNG) has produced the composition shown in Table 4.4 and should be easily integrated into the natural gas distribution network. Because of the slightly lower heating value of this SNG, some blending may be required.

Table 4.4. Projected Lurgi SNG Composition

Component	Mol %
Methane	85.99
Hydrogen	9.70
Carbon Monoxide	0.05
Carbon Dioxide	3.18
Nitrogen	1.07
Water	0.01
Higher Heating Value	902.6 Btu/SCF
Available at	1012 psig and 105°F

- *Naphtha (Lurgi Process)*

Naptha which is produced in the Lurgi Process could be used as a reformer feedstock for the refining and petrochemical industries, and would have the following characteristics:

Gross Heating Value, Btu/lb (HHV)	18,500
Sulfur, wt %	0.20
Boiling Range at Atmospheric Pressure, °F	120-320

- *Tar Oil (Lurgi Process)*

Tar oil produced by the Lurgi Process has been characterized as having the following properties:

Gross Heating Value, Btu/lb (HHV)	17,200
Sulfur, wt %	0.30
Boiling Range at Atmospheric Pressure, °F	285-680

- *Tar (Lurgi Process)*

Tar produced in this process will require further processing in a refinery and has the following properties:

Gross Heating Value, Btu/lb (HHV)	16,000
Sulfur, wt %	0.50
Boiling Range at Atmospheric Pressure, °F	> 500

- *Sulfur and Ammonia (Lurgi Process)*

Both sulfur and ammonia products of the Lurgi Process are over 99% pure. Ammonia would be of value in the fertilizer industry; whereas, the sulfur produced would depend on market conditions and could be used in the chemical industry.

4.3.2 BI-GAS Process[6]

- *Substitute Natural Gas (BI-GAS Process)*

The only energy product of this process is SNG having the projected properties shown in Table 4.5. As with most coal gasification processes, some

Table 4.5. Projected BI-GAS SNG Composition

Component	Mol %
Methane	86.68
Hydrogen	9.98
Carbon Monoxide	0.10
Carbon Dioxide	2.64
Nitrogen	0.59
Water	0.01
Higher Heating Value	907.6 Btu/SCF
Available at	1012 psig

blending with a higher energy content natural gas will be required to meet the average heating value of 1000 Btu/SCF for today's natural gas.

- *Sulfur and Ammonia (BI-GAS Process)*

These byproducts of the BI-GAS Process are exported from the plant in the form of elemental sulfur and anhydrous ammonia.

4.3.3 Synthane Process[6]

- *Substitute Natural Gas (Synthane Process)*

The SNG product of the Synthane Process will need blending before it can be directly interchangeable with natural gas. It has a heating value of 902.2 Btu/SCF and has the composition shown in Table 4.6.

- *Tar Oil (Synthane Process)*

The tar oil product of the Synthane Process has not been sufficiently characterized to permit exact specifications; however, it is expected to have an average molecular weight of 352 and to contain 0.05 wt % ash.

- *Ammonia and Sulfur (Synthane Process)*

Ammonia produced in the Synthane Process could be of value in the fertilizer industry; sulfur, depending on market conditions, could be used as a feedstock in the chemical industry.

Table 4.6. Projected Synthane SNG Composition

Component	Mol %
Methane	85.18
Hydrogen	12.96
Carbon Monoxide	0.09
Carbon Dioxide	1.08
Nitrogen	0.68
Water	0.01
Higher Heating Value	902.2 Btu/SCF

- *Char (Synthane Process)*

Char produced in the Synthane Process is low in sulfur and can be burned directly. The composition of this char is shown in Table 4.7.

4.3.4 COGAS Process[16,17]

- *Substitute Natural Gas (COGAS Process)*

The SNG produced in the COGAS Process has a heating value of 950 Btu/SCF and should be of sufficient quality to be introduced into the natural gas transmission network. This gas has the typical gas composition shown in Table 4.8.

- *Syncrude (COGAS Process)*

The syncrude produced by the COGAS Process has been characterized as containing only 0.1% sulfur and as having a 25° API gravity. Its principal value would be as a refinery feed and, as such, could yield LPG, refinery gases (for internal use), gasoline, middle distillates (jet, diesel, and No. 2 fuel oil), and residual fuel such as No. 6 fuel oil.

- *Light Oils (COGAS Process)*

The light oils produced in the COGAS Process are upgraded from the raw oil by hydrotreatment. This oil would have the projected value shown in

Table 4.7. Projected Wet Char Composition of Synthane Process

Component	Mol %
Carbon	84.15
Hydrogen	2.76
Nitrogen	0.61
Sulfur	0.35
Oxygen	12.15

Table 4.8. Projected COGAS SNG Composition

Component	Mol %
Methane	91.5
Carbon Dioxide	0.1
Hydrogen	0.4
Inerts	5.0
C_2-C_4	3.0
Higher Heating Value	950.0 Btu/SCF

Table 4.9 and is predominantly naphthemic with a large aromatic but low paraffin content. By supplying less hydrogen from reforming of the synthesis gas to the hydrotreatment section, a low sulfur #4 or #5 fuel oil can be produced.

- *Ammonia, Sulfur, and Sodium Sulfate (COGAS Process)*

Both the sulfur and ammonia products of the COGAS Process are in relatively pure form. Ammonia would be of value in the fertilizer industry; whereas, the sulfur could be used as a chemical feedstock. The sodium sulfate is of sufficient purity to be sold as a feedstock to the craft paper industry or to any other industry that utilizes it.

4.3.5 Steam-Oxygen Process[6,18]

- *Substitute Natural Gas (Steam-Oxygen HYGAS Process)*

The SNG produced by the Steam-Oxygen HYGAS Process can be blended with existing natural gas supplies to produce a pipeline quality gas. Table 4.10 shows the projected composition of this gas.

- *Byproduct Oil (Steam-Oxygen HYGAS Process)*

The oil is produced in the form of B-T-X (benzene, toluene, and xylene). The B-T-X has been characterized as approximately 80% toluene, with the remainder as benzene and xylene.

Table 4.9. Projected COGAS Hydrotreated Oil Analysis

Component	Weight %
Carbon	87.00
Hydrogen	11.00
Nitrogen	0.10
Sulfur	0.70
Oxygen	0.60
°API	22
Pour Point, °F	30
Viscosity, SUS, 100°F	43

Table 4.10. Projected Steam-Oxygen HYGAS SNG Composition

Component	Mol %
Methane	86.65
Hydrogen	11.75
Carbon Dioxide	0.01
Carbon Monoxide	1.58
Water	0.01
Higher Heating Value	913.9 Btu/SCF

- ***Ammonia and Sulfur (Steam-Oxygen HYGAS Process)***

Ammonia and sulfur from the Steam-oxygen HYGAS Process are exported from the plant in the form of anhydrous ammonia and elemental sulfur.

4.3.6 Steam-Iron HYGAS Process[6,21]

- ***Substitute Natural Gas (Steam-Iron HYGAS Process)***

The SNG produced by the Steam-Iron HYGAS Process will require blending because of its low heating value before being interchanged with existing pipeline quality gas. It has a higher heating value of 904 Btu/SCF, and the projected gas composition is shown in Table 4.11.

- ***Byproduct Oil (Steam-Iron HYGAS Process)***

The oil produced in the Steam-Iron HYGAS Process is similar to B-T-X produced in the Steam-Oxygen HYGAS Process. It is composed primarily of toluene with the remainder being benzene and xylene.

- ***Ammonia and Sulfur (Steam-Iron HYGAS Process)***

Ammonia and sulfur from the Steam-Iron HYGAS Process are produced as anhydrous ammonia and elemental sulfur, both of which can be sold as chemical feedstocks.

Table 4.11. Projected Steam-Iron HYGAS SNG Composition

Component	Mol %
Methane	86.16
Hydrogen	9.76
Carbon Monoxide	0.09
Carbon Dioxide	1.63
Nitrogen	2.35
Water	0.01
Higher Heating Value	904.0 Btu/SCF

5 ENERGY BALANCE

5.1 GENERAL DISCUSSION

In general, coal gasification plants are designed to be both efficient and energy self-sufficient. The only source of energy input is in the form of coal; all electrical, steam, and fuel needs for use inside the plant are generated and integrated within the plant design.

5.2 ENERGY INPUT/OUTPUT

Figures 5.1 through 5.6 show the energy input/output block diagrams. These diagrams are based on an energy input of 100 units, using the higher heating value of the feed coal, and show the amount of energy contained in the products, byproducts, and heat rejected to the environment. Energy values of the sulfur and ammonia are included in the overall thermal efficiency calculation.

All the process energy balances, except that for the COGAS Process, can be compared directly because they use the same Montana subbituminous coal with a higher heating value of 11,290 Btu/lb dry.[6] The COGAS Process is characterized on an Illinois bituminous coal with an assumed higher heating value of 12,530 Btu/lb dry.[17] The data presentation for the COGAS Process operating on Illinois coal was necessitated by the unavailability of published information covering the use of Montana subbituminous coal in the COGAS Process.

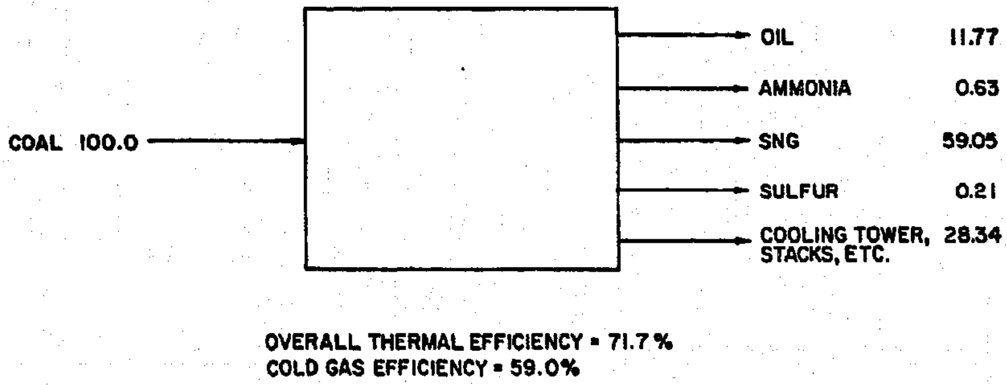


Fig. 5.1. Single-Feed Lurgi Energy Balance[6]
(Basis: Coal input = 100, HHV)

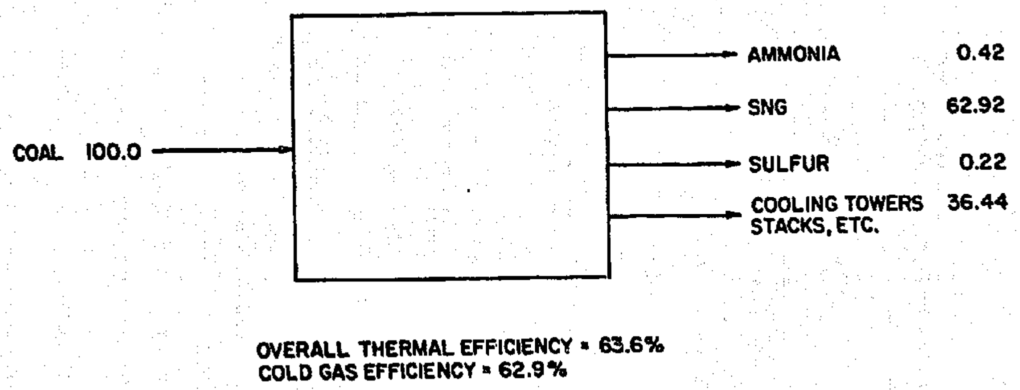


Fig. 5.2. BCR BI-GAS Energy Balance[6]
(Basis: Coal input = 100, HHV)

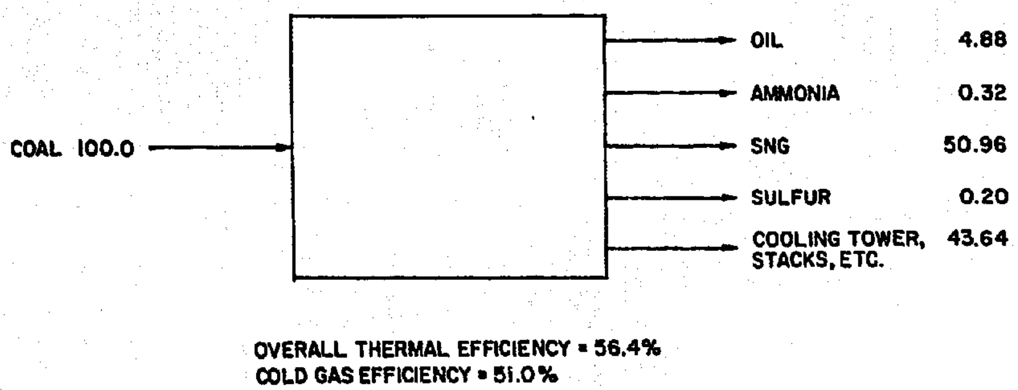


Fig. 5.3. PERC Synthane Energy Balance (Slurry Feed, Export Char)[6]
(Basis: Coal input = 100, HHV)

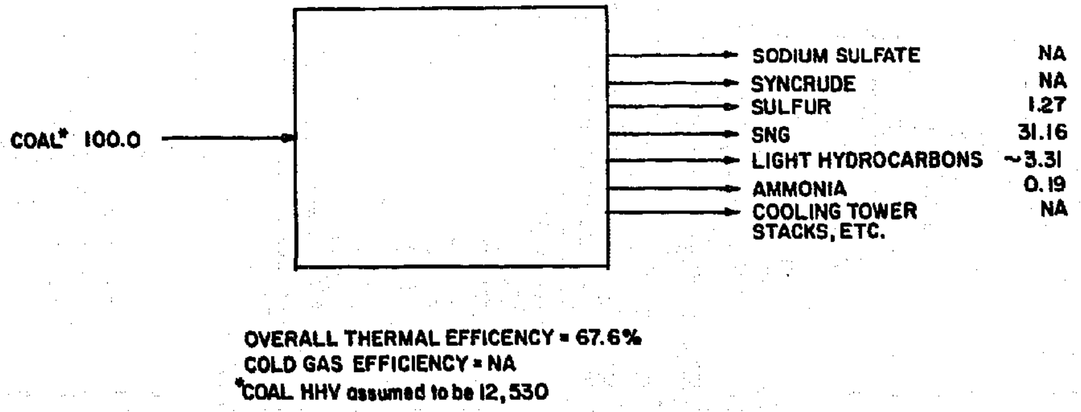


Fig. 5.4. COGAS Process Energy Balance[17]
(Basis: Coal input = 100, HHV)

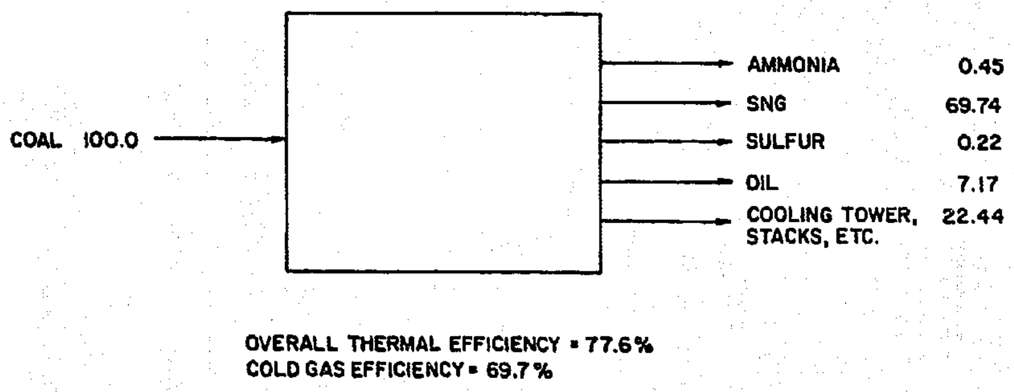


Fig. 5.5. Steam-Oxygen HYGAS Energy Balance[6] (Basis: Coal input = 100, HHV)

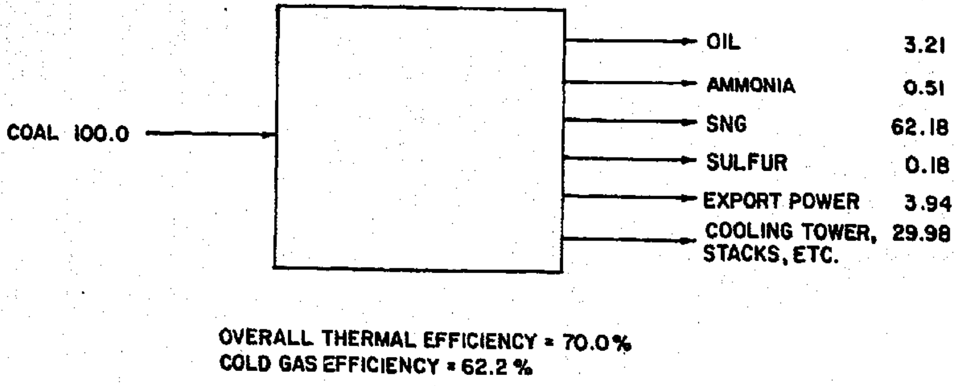


Fig. 5.6. Steam-Iron HYGAS Energy Balance[6] (Basis: Coal input = 100, HHV)

As can be seen from the energy balance figures, the thermal efficiency, as well as the product mix, varies from process to process. Overall thermal efficiencies range from a low of 56.4% for the Synthane Process to a high of 77.6% for the Steam-Oxygen HYGAS Process. The product mix varies from six different products in the COGAS Process to just three in the BI-GAS Process.

A direct comparison among the various processes, using thermal efficiency as a criterion, can be misleading. For example, in the Steam-Iron HYGAS Process, a major product is electrical power, generated by steam.

Although electricity is readily useable with a rather defined demand, a reduction in the overall thermal efficiency results from the steam-to-electricity conversion. Another case in point is the production of unfinished products, such as oils and tars, that may require additional processing. This additional processing will require another energy loss before a useable end-use product is produced. The selection of which process to use in any given case can depend on the desirability of a process' product mix.

6 ENVIRONMENTAL EFFECTS

6.1 GENERAL DISCUSSION

Any process, such as coal gasification, obviously will impact on the environment by its production of noise and heat, as well as gaseous, solid, and liquid effluents. Currently, there are no federal standards written specifically for coal gasification processes to regulate these discharges to the environment. In some cases, standards, regulating discharges from other similar processes, may be used to estimate the permissible discharge levels for gasification processes. One example of this would be the use of federal regulations for petroleum refineries with ragard to effluent concentration levels. Conversely, standards regulating the production of noise and the discharge of heat to the environment are essentially independent of plant function. Noise production falls under the jurisdiction of the U.S. Occupational Safety and Health Administration, which controls the noise level to which a worker at a plant may be subjected. Thermal discharges to bodies of water are regulated; those to the air are not.

6.2 THERMAL DISCHARGES

The primary mode of heat rejection from coal gasification processes is through the air; the higher the thermal efficiency of the process, the lower the amount of rejected heat. Essentially, three methods are available for heat rejection: stack losses, air coolers, and wet cooling towers. Stack losses consist of the heat contained in the exhaust gases that are vented to the atmosphere. Air coolers are simply heat exchangers that cool some process gas or liquid by the heating of air. Cooling towers are heat exchangers that reject heat by causing the evaporation of water.

The temperature of the water exiting the cooling towers is approximately 85°F; the entering water temperature usually is between 100° and 120°F.

Water lost by evaporation is made up by fresh water from a river or other source that is treated before use. The major end-use of the water inputs to the plant material balances, shown in Sect. 4, is this cooling tower makeup water.

Amounts of rejected heat by means of cooling towers, relative to the amount rejected by stack losses or air coolers, may vary with climatic conditions, geographical location, or design philosophy. All the processes evaluated here are designed to reject as much heat as is economically feasible through the use of air coolers.

6.3 NOISE ATTENUATION

By law, the amount of noise generated during the construction and operation of a plant is regulated with regard to the effect on workers at the plant. As a result, noise abatement is a constraint in the design of these plants. High-level noise sources, such as turbines, fans, compressors, valves, and pipelines are attenuated by either redesign or the use of sound and vibration absorbing materials or barricades.

Local codes, which take effect at the plant property line, also may set a limit on the generation of noise. However, coal gasification plants most likely would be treated as any industrial plant and not specifically as a gasification plant.

6.4 AESTHETICS

Coal gasification plants should be similar in overall appearance to petroleum refineries. It is not expected that they would be located in, or even near, residential areas.

6.5 GASEOUS EFFLUENTS

Emission standards for coal gasification facilities have not been issued by the federal government. Only one state, New Mexico, has issued regulations directly pertaining to coal gasification. Similarly, Illinois has issued standards for petrochemical process plants that are as strict as, or stricter than, the federal standards and are often used for comparison purposes with gasification plants because of Illinois' position in the Interior coal region. The New Mexico air quality standards, shown in Tables 6.1 and 6.2, are considered far stricter than the National Ambient Air-Quality Standards amd are presented in Table 6.3. Because the New Mexico standards are considered to be stricter than any expected federal standards that may be issued on coal gasification facilities, many recent plant designs have been directed at meeting the New Mexico standards.

Table 6.1. New Mexico Ambient Air-Quality Standards

	Maximum Concentration
μg/m³	
1. PARTICULATES	
24-hr avg.	150
7-day avg.	110
30-day avg.	90
Annual Geometric Mean	60
Beryllium	0.01
Asbestos	0.01
Heavy Metals (Total)	10
ppm	
2. SULFUR DIOXIDE	
24-hr avg.	0.10
Annual Arithmetic Mean	0.02
3. HYDROGEN SULFIDE	
State, except Pecos-Permian Basin, 1-hr avg.	0.003
Pecos-Permian Basin, 1/2-hr avg.	0.10
4. TOTAL REDUCED SULFUR	0.003[a]
5. CARBON MONOXIDE	
8-hr avg.	8.7
1-hr avg.	13.1
6. NITROGEN DIOXIDE	
24-hr avg.	0.10
Annual Arithmetic Mean	0.05
7. PHOTOCHEMICAL OXIDANTS	
1-hr avg.	0.06
8. NON-METHANE HYDROCARBONS	
3-hr avg.	0.19

(a) Applies to the entire state except the Pecos-Permian Basin. Includes H_2S, 1-hr avg.; other standards apply to the Pecos-Permian Basin.

Table 6.2. New Mexico Emission Standards

Source	Allowable Emission[e]
COAL-FIRED BOILER[a]	
Sulfur Dioxide	0.34
Particulates	
Total	0.05
Fine Particulates	0.02
Nitrogen Dioxide	0.45
COAL GASIFICATION PLANT[b],[c]	
$H_2S + COS + CS_2$, ppm	100
H_2S only, ppm	10
HCN, ppm	10
HCl, ppm	5
Particulates, gr/ft³[d]	0.03
NH_3, ppm	25
Total Sulfur	0.008[f]
Gas-Burning Boiler	
Particulates	0.03[g]
SO_2	0.16[g]
Briquette Forming Particulates, gr/SCF	0.03

(a) Applies to new sources with high heating value heat, input > 250 million Btu/h.

(b) Stack heights must be at least 10 diameters in height and have adequate sample ports and platforms.

(c) See the *Environmental Reporter* for a complete tabulation.

(d) Grains/cubic ft of gas at 70°F and 14.7 psia.

(e) All emissions are lb of pollutant/million Btu of heat input (higher heating value) unless stated otherwise.

(f) Based on heat input to gasifier.

(g) Based on lower heating value.

Table 6.3. Summary of National Ambient Air-Quality Standards

Pollutant	Time of Average	Primary Standard (at 25°C and 60 mm of Hg)	Secondary Standard
Particulate Matter (TSP)	Annual Geometric Mean 24 h	75 μg/m^3 260 μg/m^3	60 μg/m^3 150 μg/m^3
Sulfur Dioxide (SO_2)	Annual Arithmetic Mean 24 h 3 h	0.03 ppm (80 μg/m^3) 0.14 ppm (365 μg/m^3) None	None None 0.5 ppm (1300 μg/m^3)
Carbon Monoxide (CO)	8 h 1 h	9 ppm (10 mg/m^3) 35 ppm (40 mg/m^3)	Same as Primary Same as Primary
Photochemical Oxidants(O_3)	1 h	0.08 ppm (160 μg/m^3)	Same as Primary
Non-Methane Hydrocarbons (N-MHC)	3 h (6 to 9 a.m.)	0.24 ppm (160 μg/m^3)	Same as Primary
Nitrogen Dioxide (NO_2)	Annual Arithmetic Mean	0.05 ppm (100 μg/m^3)	Same as Primary

Note: All standards with averaging time of 24 h or less are not to be exceeded more than once per year.

μg/m^3 = micrograms/cubic meter

mg/m^3 = milligrams/cubic meter

ppm = perts per million by volume

Gaseous effluents of coal gasification facilities consist of:

(1) combustion stack gases;

(2) process vent gases from units such as the oxygen production unit or the sulfur recovery unit;

(3) cooling tower evaporation and drift;

(4) fugitive emissions from pump seals, valves, flanges, storage facilities, etc.; and

(5) pond evaporation.

Table 6.4. Stack Gas Emissions from the Lurgi Process[24]

Flue Gas Component	LB/H
H_2O	27,466
CO_2	688,296
N_2	5,124,597
O_2	1,022,215
SO_2	384
NO_x	570
Particulates	Negligible
Total Flue Gas	6,863,528

Gaseous emissions from combustion stack gases for the Lurgi Process are shown in Table 6.4. This gaseous discharge originates from the turbines, boilers, incinerators, steam superheaters, and fuel gas heaters. The discharge consists primarily of water, carbon dioxide, nitrogen, and oxygen; the contaminant discharge of NO_x, SO_2, and particulates is relatively low when compared with the overall gas volume. Vent gases from the various process units contain many different effluents. Tables 6.5 and 6.6 show the contaminants of these streams for the Lurgi and Synthane Processes, respectively. Also shown in Tables 6.5 and 6.6 are the remainder of the gaseous emission characteristics, such as temperature, pressure, stream rate, and contaminants.

Several of the gasification processes have been designed to the point where the gaseous emission levels expected from the plants can be stated. Other processes, still in the development stage or without adequate pilot plant experience, have only stated in their design criteria that these facilities will meet any existing standard that might pertain to coal gasifiction facilities.

Table 6.5. Summary of Gaseous Effluent Streams from the Lurgi Coal Gasification Process(a)[24]

Waste Stream	Temperature °F	Pressure psi	Steam Rate lb/h	Contaminants	Quantity of Contaminants lb/h
Stack Gases	300	15	7.11×10^6	Particulates	Negligible
				NO_x	570
				SO_2	384
Oxygen Production Vent Gases	Ambient	15	1.59×10^6	None	-
Sulfur Recovery Vent Gases	150	15	1.80×10^6	CO	1,780
				COS	172
				CS_2	6
				H_2S	12
				CH_4	3,230
				C_2H_4	2,390
				C_2H_6	3,390
Water Treatment Degasser Vent	Ambient	15	17.5×10^3	None	-
Steam and Power Production Deaerator Vents	200	15	31.9×10^3	Negligible	
Cooling Tower Evaporation	Ambient	15	1.04×10^6	H_2S	-
				NH_3	-
Cooling Tower Drift	Ambient	15	16.5×10^3	Dissolved Solids	-
				Trace Elements	-
				Trace Organics	-
Pond Evaporation	Ambient	15	606×10^3	Negligible	-
Fugitive	-	15	50	H_2S	-
				NH_3	-
				CO	-
				HC	-
				Trace Elements	-

(a) Based on El Paso project design with coal feed rate of 28,200 tons/day of subbituminous coal and 288×10^6 SCF/day of SNG capacity.

Table 6.6. Summary of Gaseous Waste Effluents from the Synthane Coal Gasification Process(a)[24]

Waste Stream	Temperature °F	Pressure psia	Steam Rate lb/h	Contaminants	Quantity of Contaminants lb/h
Combustion Stack Gases	200	15	3,100,778	Particulates	8
				Hydrocarbons	51
				CO	171
				SO_2	3,595
				NO_x	5,153
				Trace Elements	-
				Trace Organics	-
Sulfur Recovery Vent Gases	150	15	2,421,860	H_2S	1 ppm
				CO	8
Oxygen Production Vent Gases	75	15	2,701,492	-	-
Cooling Tower Evaporation	Ambient	15	3,260,000	H_2S	-
				NH_3	-
				Trace Organics	-
				Trace Elements	-
Cooling Tower Drift	Ambient	15	30,000	H_2S	-
				NH	-
				Trace Organics	-
				Trace Elements	-
Fugitive Emissions	-	11	45	H_2S	-
				NH_3	-
				CO	-
				Hydrocarbons	-
				Trace Elements	-
				Trace Organics	-

(a)Based on Laboratory data for a 255×10^6 SCF/day plant using 22,000 tons/day of coal.

6.6 AQUEOUS EFFLUENTS

As was the case with gaseous effluents from coal gasification plants, there are no federal standards directly issued for coal conversion facilities. Federal standards for petroleum refining and the Illinois State standards for all sources should be used as goals for those processes that have not yet characterized their liquid effluents.

Liquid effluents from gasification facilities consist of:

(1) boiler blowdown,
(2) cooling tower blowdown,
(3) process condensates,
(4) sewage treatment sludge,
(5) demineralizer and zeolite softener regeneration wastes, and
(6) quenching water overflows.

Tables 6.7 and 6.8 summarize the projected liquid effluents from the Synthane and Lurgi Processes, respectively. All, or part of, these liquid waste effluent streams can be eliminated through the use of cleanup and recycling equipment. In this manner, it is possible to attain zero liquid effluents by recycling the cleaned liquid back into the process water feed lines. Residue from the liquid cleaning processes is disposed of as a solid waste.

Table 6.7. Summary of Liquid-Phase Effluent Streams from the Synthane Coal Gasification Process(a)[24]

Waste Stream	Temperature °F	Pressure psia	Steam Rate lb/h	Contaminants
Cooling Tower Blowdown	Ambient	15	332,000	H_2S NH_3 Trace Elements Trace Organics
Sodium Zeolite Softener Regeneration Wastes	Ambient	15	13,000-79,500	Na^+ Cl^-
Demineralizer Regeneration Wastes	Ambient	15	500-4,000	$SO_4^=$ Cl^- Na^+

(a) Based on laboratory data for a 255 x 10^6 SCF/day plant using 22,000 tons/day of coal.

Table 6.8. Summary of Liquid-Phase Effluent Streams from the Lurgi Coal Gasification Process(a)[24]

Waste Stream	Temperature °F	Pressure psi	Steam Rate lb/h	Contaminants	Quantity of Contaminants lb/h
Gas Liquors Before Treatment	.150	15	1.44 x 10^6	H_2S	606
				NH_3	21,400
				Tars	88,800
				Tar Oils	48,600
				Phenols	11,300
				CO	64
				CH_4	42
				Trace Organics	-
				Trace Elements	-
Gas Liquors After Treatment	≤150	15	1.36 x 10^6	Trace Elements	-
				Trace Organics	-
				H_2S	-
				NH_3	-
Boiler Blowdown	.200	15	228 x 10^3	Dissolved Solids	-
Process Condensates	.200	15	419 x 10^3	Negligible	-
Cooling Tower Blowdown	Ambient	15	244 x 10^3	Trace Elements	-
				Trace Organics	-
				Dissolved Solids	-
Ash Quencher Water	.150	15	530 x 10^3	Trace Elements	-
				Trace Organics	-
				Dissolved Solids	-
				Mineral Matter	-

(a) Based on El Paso project design for a coal feed rate of 28,200 tons/day of subbituminous coal and 288 x 10^6 SCF/day of SNG capacity.

As in the case of gaseous emissions, several gasification processes have characterized the liquid effluents; whereas, other processes, still in the pilot plant stage, have only stated that the commercial design will meet existing effluent standards.

6.7 SOLID WASTES

Solid wastes from the coal preparation units, ash from gasification or combustion units, and sludges created in the water treatment facilities are either returned to the mine for disposal, or buried in a suitable landfill. The spent catalysts that are not a continuous solid waste problem may be disposed of in the same manner or returned to the manufacturer for regeneration or recycling. As long as these disposal techniques do not allow these solids, or the mineral matter contained in them (the most important ones being selenium, chromium, boron, mercury, and barium) to affect local water supplies, there are no or few applicable federal or state standards.

A summary of the solid effluents from the Lurgi Coal Gasification Process is shown in Table 6.9. Table 6.10 summarizes the solid effluents from the Synthane Process. Most of the remaining processes have

Table 6.9. Summary of Solid-Phase Effluents from the Lurgi Coal Gasification process(a)[24]

Waste Stream	Temperature °F	Pressure psi	Steam Rate lb/h	Contaminants
Wet Gasifier Ash	Ambient	15	572×10^3	Minerals Dissolved Solids Trace Elements Trace Organics
Lime Softener Sludge	Ambient	15		$CaCO_3$ $Ca(OH)_2$
Spent Catalyst	Ambient	15	Intermittent	Cobalt Chromium Molybdenum Vanadium Nickel

(a) Based on El Paso project design for a coal feed rate of 28,200 tons/day of subbituminous coal and 288×10^6 SCF/day of SNG capacity.

Table 6.10. Summary of Solid-Phase Streams Effluents from the Synthane Coal Gasification Process(a)[24]

Waste Stream	Temperature °F	Pressure psia	Steam Rate lb/hr	Contaminants
Utility Boiler Ash	350	15	122,000	Trace Elements Trace Organics Mineral Matter
Limestone Wet Scrubber Sludge	Ambient	15	111,754	Trace Elements
Spent Catalysts	Ambient	15	Intermittent	-

(a) Based on laboratory data for a 255 x 10^6 SCF/day plant using 22,000 tons/day of coal.

characterized the solid wastes only as meeting any solid-waste emission standards that would be applicable at the time of commercial phase construction.

7 OPERATING REQUIREMENTS

7.1 CONSIDERATIONS

Because no large commercial gasification plants have yet been built in the United States, little is known about the true operating requirements, i.e., what the operating ranges and procedures or safety requirements will be. These may be estimated or inferred from detailed designs and/or experience with other, similar processes or units.

With regard to operating ranges, a gasification plant is not designed to follow a fluctuating demand for its products; these demand fluctuations are modulated by storage facilities at the plant site or by underground storage via the distribution system it feeds. Therefore, gasification plants are designed to operate at or near 100% of design capacity. For very large plants, it may be possible to decrease the process plant output by a factor inversely proportional to the number of parallel trains within the process. For example, if there are four parallel trains (i.e., a set of process units, in parallel they may be operated independently of the other three sets), three of these trains could be turned down or off, while the remaining train operates at full capacity. This would be equivalent to operating the plant at 25% of the designed capacity. Because many pieces of equipment in a gasification plant, particularly high-pressure and temperature vessels, have size limitations, as the plant design capacity increases, the number of parallel units would tend to increase.

Conversely, other pieces of process equipment may have economic constraints that dictate that it is less expensive, per unit of material processed, to have one large unit as opposed to several smaller ones. Thus, a gasification plant may have parallel trains in some areas of the process, and one single train in others. Plant size will determine how these two opposing factors interact with each other. Because of this size variable, "across-the-board" estimates of turndown capacity are difficult to determine.

For those operations that are unique to specific gasification plants, operating procedures and safety requirements are being determined at the pilot plants operating for each process. For those operations that are similar to other industrial processes, e.g., coal handling, gas purification, etc., the operating procedures and safety requirements may be considered as standard practice.

7.2 OPERATING PERSONNEL

Coal gasification plants will require a full-time operating staff. Usually this staff will be made up of small groups, solely responsible for the operation of the major pieces of equipment represented by the equipment blocks shown in Sect. 3 (i.e., gasification, pyrolysis, etc.). The number of staff required has been estimated by projecting data from other similar pieces of equipment currently used in industry.

Conceptual designs for commercially sized, high-Btu, coal gasification facilities have been based on a 250 billion Btu/day of SNG product output capacity.[6,17] Only the more detailed of these conceptual designs were used in this evaluation to assess accurately the process characteristics. As a result of using these detailed design studies, the size of the plant, the capital requirements, and the operating costs are clearly fixed. Therefore, "scaling factors" are required to apply the above design study results to other plant sizes. However, caution should be used in applying these factors because fractional sizes may result in less economical process trains.

The scaling factor to be used here for estimating the number of operating personnel as a function of plant size may be seen in the following relation:

$$\frac{\text{Desired Operating Staff Size}}{\text{Known Operating Staff Size}} = \left(\frac{\text{Desired Plant Size}}{\text{Known Plant Size}}\right)^{0.40}$$

The known or desired plant size to be used here is based on an energy output. In this manner, differences in coal properties and product outputs are removed; these constituents do not necessarily influence the design and efficiency of the process, and they have little effect on the staff size.

In TRW's *Project Independence* report to the Federal Energy Administration,[25] a commercial-scale Lurgi gasification plant with a 250-billion-Btu/day SNG capacity requires approximately 311 persons: 150 operators and 161 service and administration employees. In a different paper on the COGAS process,[17] an estimated 1075 people will be required. However, this personnel requirement has not been broken down in terms of operating, maintenance, and supervisory personnel. For the remainder of the processes discussed here, no accurate estimate is available. For this reason, the personnel requirements for the CO_2 Acceptor Process (a high-Btu coal gasification process not discussed in this report but similar in technology and stage of development) were chosen as representative of the personnel requirements of the remainder of the processes discussed. The operator labor requirements were estimated in the TRW report as 150 people/day.[25]

8 MAINTENANCE AND RELIABILITY

8.1 GENERAL

Coal gasification plants will require a full-time maintenance staff for conducting preventive and routine maintenance. During plant shutdowns, major maintenance and refurbishing usually are carried out by contract labor.

8.2 MAINTENANCE REQUIREMENTS

Mainteance requirements for a coal gasification plant consist of two factors: material and labor. Maintenance materials, i.e., those items that are required to keep the units operating as designed, normally comprise about 40% of the total maintenance costs. Maintenance labor comprises the plant staff and contracted labor required to perform this maintenance and constitutes about 60% of the total cost of maintenance.

Because coal gasification still is a developing technology, the cost and number of staff required for maintnance may only be estimated. Although many pieces of equipment are common to the processes evaluated here, many other units are unique to each design and require different maintenance procedures. Thus, although it is common to use a statement that the maintenance costs (and thus staff size) are some percentage of the total plant costs, the use of such a factor for all the processes examined here could be misleading because it would tend to mask these differences.

The number of plant personnel required for maintenance is based on published estimates[6,17,25] for a particular plant size (see Sect. 7.2), and scaled for other plant sizes by use of the following relation:

$$\frac{\text{Desired Number of Maintenance Staff}}{\text{Known Number of Maintenance Staff}} = \left(\frac{\text{Desired Plant Size}}{\text{Known Plant Size}}\right)^{0.40}$$

In TRW's *Project Independence* Report,[25] a commercial Lurgi Process with a 250 billion Btu/day capacity would require approximately 580 maintenance employees. For a process that is nearing commercialization, an estimated 425 maintenance employees would be required.

8.3 PLANT LIFE AND RELIABILITY

The design life of commercial gasification plants is usually taken as 20 yr, with plant startup being year one. The plants are designed to operate 330 days/yr, or 90.4% of the time at full design output. The remaining time consists of scheduled and unscheduled maintenance.

9 COST CONSIDERATIONS

9.1 GENERAL

The gasification processes examined here can be compared on an equivalent basis. All operating and capital costs, as well as equipment requirements and size, have been escalated to represent mid-1978 dollars for a 250-billion-Btu SNG output coal gasification facility.

9.2 CAPITAL COST

The capital cost of the coal gasification process plants, at their designed 250 billion Btu/day SNG capacity, is shown in Table 9.1. The fixed capital investment is the total cost of installing the plant at some location. This includes the amount of capital need to build and start operation of the gasification plant. All costs shown in this table have been adjusted to mid-1978 dollars by use of the *Chemical Engineering* Plant Cost Indexes published by the journal every two weeks.

Table 9.1. Capital Costs of Coal Gasification Plant(a)[6]

	Mid-1978 $$10^6$				
Process	Lurgi	BI-GAS	Synthane	Steam-Oxygen HYGAS	Steam-Iron HYGAS
Installed Cost	956.9	931.2	1038.5	784.0	1155.1
Home Office Cost	106.8	103.5	115.0	86.6	128.3
Project Contingency	155.3	149.5	169.1	130.0	188.6
Paid-Up Royalties	1.1	1.75	1.7	1.8	8.6
FIXED CAPITAL INVESTMENT	1220.1	1186.0	1324.3	1002.4	1480.6
Start-Up Cost	60.8	49.4	57.1	43.7	57.1
Construction Financing	192.2	187.3	208.5	153.4	233.2
TOTAL DEPRECIABLE PLANT	1473.1	1422.7	1590.0	1199.5	1770.9
Initial Catalyst and Chemical Costs	23.9	11.9	9.3	10.0	8.4
Working Capital(b)	19.3	17.0	19.7	14.8	19.7
TOTAL CAPITAL REQUIRED	1516.3	1451.6	1619.0	1224.3	1799.0

(a) Design basis of 250 billion Btu of SNG/day.

(b) Does not include a byproduct credit.

Using Table 9.1 as a base, the capital requirements for the gasification process plants were scaled to four different plant sizes (100, 150, 200, and 250 billion Btu/day of SNG product) by use of the following relation:

$$\frac{\text{Desired Plant Size Capital Required}}{\text{Design Plant Size Capital Required}} = \left(\frac{\text{Desired Plant Size, Billion Btu of SNG/day}}{\text{Design Plant Size, Billion Btu of SNG/day}}\right)^{0.75}$$

Tables 9.2 through 9.6 show the values produced by using this relation. Construction time was not changed from the original estimates (of about four years) published in the design studies.

Table 9.2. Capital Requirements for Lurgi Gasification Process Plant[6]

	SNG Output, Billion Btu/day HHV Mid 1978 $\$10^6$			
	100	150	200	250
Fixed Capital Investment	613.7	831.8	1032.1	1220.1
Total Depreciated Plant	740.9	1004.3	1246.1	1473.1
Total Capital Required	762.7	1033.7	1282.6	1516.3

Table 9.3. Capital Requirements for BI-GAS Gasification Process Plant[6]

	SNG Output, Billion Btu/day HHV Mid 1978 $\$10^6$			
	100	150	200	250
Fixed Capital Investment	596.5	808.5	1003.2	1186.0
Total Depreciated Plant	715.6	969.9	1203.5	1422.7
Total Capital Required	730.1	989.6	1227.9	1451.6

Table 9.4. Capital Requirements for Synthane Gasification Process Plant[6]

	SNG Output, Billion Btu/day HHV Mid 1978 10^6			
	100	150	200	250
Fixed Capital Investment	666.1	902.8	1120.2	1324.3
Total Depreciated Plant	799.7	1084.0	1345.0	1590.0
Total Capital Required	814.3	1103.7	1369.0	1619.0

Table 9.5. Capital Requirements for Steam-Oxygen HYGAS Gasification Process Plant[6]

	SNG Output, Billion Btu/day HHV Mid 1978 10^6			
	100	150	200	250
Fixed Capital Investment	504.2	683.4	847.9	1002.4
Total Depreciated Plant	603.3	817.7	1014.7	1199.5
Total Capital Required	615.8	834.6	1035.6	1224.3

Table 9.6. Capital Requirements for Steam-Iron HYGAS Gasification Process Plant[6]

	SNG Output, Billion Btu/day HHV Mid 1978 10^6			
	100	150	200	250
Fixed Capital Investment	744.7	1009.4	1252.4	1480.6
Total Depreciated Plant	890.7	1207.3	1498.0	1770.9
Total Capital Required	904.8	1226.4	1521.8	1799.0

9.3 ANNUAL OPERATING COST

Annual operating costs for the various gasification process plants shown in Table 9.7, are made up of four costs:

(1) *material and supplies,* including the cost of coal and iron ore (where required);

Table 9.7. Annual Operating Costs(a) for the Various 250 x 10^9 Btu/day Coal Gasification Facilities[6]

	$\$10^6$(a)				
Component	Lurgi	BI-GAS	Synthane	HYGAS Steam-Oxygen	HYGAS Steam-Iron
Material and Supplies					
Coal, at \$20/ton	195.49	148.96	183.07	134.48	149.97
Iron Ore, at \$23/ton	-	-	-	-	0.24
Purchased Water, at \$0.46/1000 gal	0.53	0.47	0.47	0.41	0.75
Maintenance Materials(b)	14.51	13.89	14.35	11.71	21.16
Catalysts and Chemicals	10.41	5.32	3.66	4.23	5.75
Operating Materials(c)	1.15	0.95	0.95	0.95	0.95
TOTAL MATERIALS AND SUPPLIES	222.09	169.59	202.50	151.78	178.82
Labor					
Operating and Supervision(d)	8.96	8.00	8.14	7.35	10.18
Maintenance(e)	21.77	20.83	21.53	17.55	31.73
Plant Overhead/Administration(f)	18.43	17.30	17.80	14.94	25.14
TOTAL LABOR	49.16	46.13	47.47	39.84	67.05
Property Taxes and Insurance(g)	32.91	31.98	35.71	27.01	39.74
TOTAL OPERATING COSTS	304.16	247.70	285.68	218.63	285.61

(a) Mid-1978 \$

(b) 40% of Total Maintenance

(c) 30% of Process Operating Labor

(d) \$7.70/h for Operating Labor

(e) 60% of Total maintenance (\$10.93/h wage rate)

(f) 60% of Labor

(g) 2.7% of Total Plant Investment

(2) *maintenance materials;*
(3) *contracted labor;* and
(4) *catalyst, chemicals, and operating supply costs.*

Raw coal was assumed to cost \$20/ton delivered to the plant site. The cost of iron ore for the Steam-Iron HYGAS Process was set at \$23/ton. Maintenance material, contract labor, catalyst, chemicals, and operating materials can be scaled by the use of 0.75 power factor relation. Labor requirements should be scaled by a 0.4 power factor relation.[6]

Operating and maintenance labor and their supervision were estimated by using the personnel requirement estimates shown in Sect. 7. Operating labor and supervision costs were set by applying an hourly average wage of $7.70/h. Maintenance labor and supervision wages were estimated at $10.93/h. The total maintenance labor cost was taken as 60% of the total maintenance; the total supervision cost was assumed to be 20% of the operating and maintenance labor. Wage rates were determined by averaging the salary estimates used in the process design studies[6] and inflating them to mid-1978 dollars. Plant overhead and administration costs were set at 60% of labor.

Maintenance materials were assumed to be 40% of total maintenance costs; the operating materials were assumed to be 30% of process operating labor costs. Property taxes and insurance were fixed at 2.7% of the fixed capital investment for all plant sizes.

9.4 SALEABLE PRODUCT OUTPUT RATES

The product slates for the six gasification processes, at various plant sizes are shown in Tables 9.8 through 9.13. Product quality, heating value, and other characteristics were discussed in Sect. 4. All product rates are shown on a daily basis. Annual outputs may be calculated by multiplying by 330 days/yr, the estimated operating availability.

Table 9.8. Lurgi Product Slate[6]

	SNG Output, 10^9 Btu/day HHV			
	100	150	200	250
Sulfur, lb/day	76,800	115,200	153,600	192,000
Ammonia, lb/day	112,320	168,480	224,640	280,800
Phenols, gal/day	15,206	22,810	30,412	38,016
Naphtha, gal/day	45,677	68,515	91,354	114,192
Tar, gal/day	35,251	52,877	70,502	88,123
Light Oil, gal/day	61,978	92,966	123,955	154,944
Surplus Coal Fines, tons/day	2,227	3,341	4,454	5,568

Table 9.9. BI-GAS Product Slate[6]

	SNG Output, 10^9 Btu/day HHV			
	100	150	200	250
Sulfur, lb/day	86,400	129,600	172,800	216,000
Ammonia, lb/day	73,920	110,880	147,840	184,800

Table 9.10. Synthane Product Slate[6]
(Slurry Feed, Export Char)

	SNG Output, 10^9 Btu/day HHV			
	100	150	200	250
Sulfur, lb/day	95,040	142,560	190,080	237,600
Ammonia, lb/day	69,120	103,680	138,240	172,800
Light Oil, gal/day	24,192	36,288	48,384	60,480
Tar, gal/day	47,808	71,712	95,616	119,520
Char and Coal Fines, tons/day	736	1,104	1,472	1,840

Table 9.11. COGAS Product Slate[17]

	SNG Output, 10^9 Btu/day HHV			
	100	150	200	250
Sulfur, lb/day	812,000	1,218,000	1,624,000	2,030,000
Light Hydrocarbons, lb/day	467,200	700,800	934,400	1,168,000
Sodium Sulfate, lb/day	63,200	94,800	126,400	158,000
Ammonia, lb/day	60,000	90,000	120,000	150,000
Syncrude, bbl/day	13,912	19,788	26,384	32,980

Table 9.12. HYGAS (Steam-Oxygen) Product Slate[6]

	SNG Output, 10^9 Btu/day HHV			
	100	150	200	250
Sulfur, lb/day	79,680	119,520	159,360	199,200
Ammonia, lb/day	71,040	106,560	142,080	177,600
Light Oils, gal/day	77,184	115,776	154,368	192,960

Table 9.13. HYGAS (Steam-Iron) Product Slate[6]

	SNG Output, 10^9 Btu/day HHV			
	100	150	200	250
Sulfur, lb/day	72,000	108,000	144,000	180,000
Ammonia, lb/day	84,480	126,720	168,960	211,200
Light Oil, gal/day	39,974	59,962	79,949	99,936
Surplus Electric Power, kWh/h	77,304	115,956	154,608	193,260

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