2. COAL GASIFICATION PLANTS

Preliminary designs have been made for the gasification processes listed in Table 1. The design for the U-Gas process was for a low Btu product, that for the Winkler and Koppers-Totzek processes were for intermediate Btu products and the rest were designed to produce high Btu gas. In this section of the final report, a summary is given of the results of the studies for the necessary steps in the gasification processes. This summary will include unit descriptions, effluents to the air, solid and liquid effluents, and process alternatives. It cannot be emphasized too strongly that, although tables may be given with results for each process, extreme care should be taken in making comparisons because of the different coal feeds, product slates, furnace feeds, etc. used in the various designs. The Lurgi, Koppers-Totzek and Winkler gasification processes are commercial while the rest of the processes considered are in various stages of development and the designs are conceptual.

Overall environmental considerations of coal gasification have been reported (11,12,13).

2.1 General Gasification Description

Figure 2 is a typical flow plan for coal gasification. Not all the units are the same for different processes and for some, additional units are required.

Coal arrives in the plant and is stored or used directly. Coal preparation may consist of physical cleaning to remove refuse (in many of the designs this step is assumed to be carried out at the mine), crushing and drying. In some cases a slurry preparation step is necessary.

In the gasifier, the coal is reacted with steam and oxygen (pure or as air for low Btu gas) at elevated temperatures and, usually, at elevated pressure. The major reactions in the gasifier are shown in Table 2. The oxygen is necessary to burn part of the coal to supply the heat required for the endothermic reaction of steam with the coal. The products are related to the temperature of the reaction; less methane and carbon dioxide are produced at higher temperatures. Also, by-products such as tar and phenols are reduced at elevated temperatures. Higher pressures tend to increase the formation of methane which is desirable if high Btu gas (substitute natural gas, SNG) is the end product. The quantity of methane is relatively immaterial if fuel gas is desired and may be detrimental if synthesis gas is to be the product. The hot, raw product is normally scrubbed with product liquor or tar to cool it to the point where higher boiling components such as tar and phenols can be removed and to remove particulates.

If SNG or synthesis gas is desired, a shift reactor is normally included to produce more hydrogen by the following reaction:

$$CO + H_2O = CO_2 + H_2 + 17{,}770 \text{ Btu/lb-mole}$$

The hydrogen to carbon monoxide ratio should have a value of approximately 3/1 for the methanation step.

Table 1

Gasification Processes for Which Designs Were Made

(Numbers in the parentheses are references to the Bibliography)

Koppers-Totzek (3)

Synthane (4)

Lurgi (5)

CO_2 Acceptor (6)

BI-GAS (7)

HYGAS (8)

U-Gas (9)

Winkler (10)

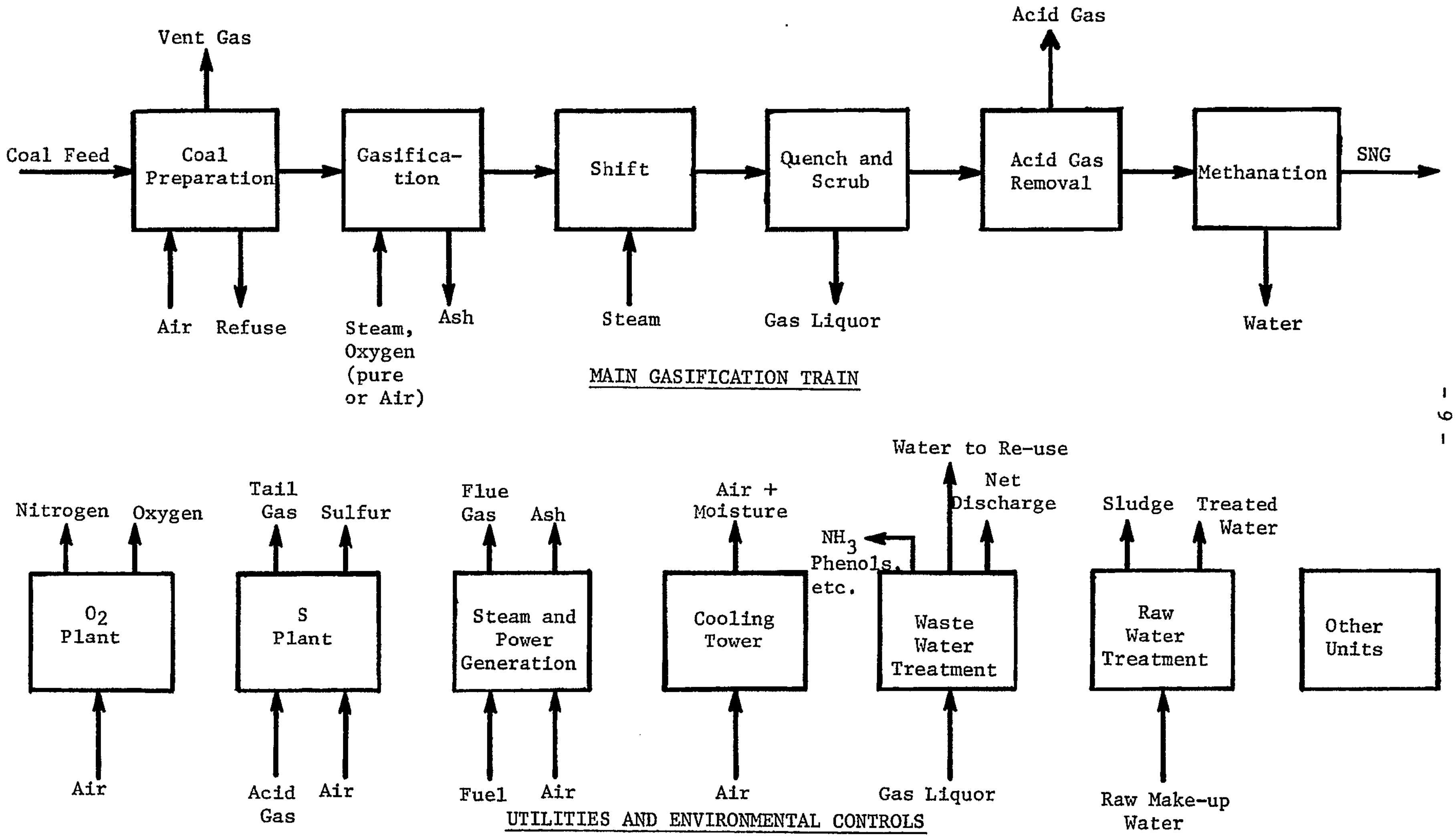

Figure 2

Flow Plan for Coal Gasification

Table 2

Reactions in Gasifier

Devolatilization and Drying

$$\text{Coal} + \text{Heat} \longrightarrow CH_4 + H_2O + \text{Organics}$$

Gasification

$$C + H_2O + 56{,}400 \text{ Btu/lb-mole} \longrightarrow CO + H_2$$

$$C + CO_2 + 74{,}200 \text{ Btu/lb-mole} \longrightarrow 2\ CO$$

$$CO + H_2O \longrightarrow CO_2 + H_2 + 17{,}700 \text{ Btu/lb-mole}$$

$$C + 2H_2 \longrightarrow CH_4 + 32{,}300 \text{ Btu/lb-mole}$$

Combustion

$$C + 1/2\ O_2 \longrightarrow CO + 47{,}550 \text{ Btu/lb-mole}$$

$$C + O_2 \longrightarrow CO_2 + 169{,}200 \text{ Btu/lb-mole}$$

An acid gas removal unit is the next step in the reaction sequence. Hydrogen sulfide must almost always be removed. Carbon dioxide is usually also removed except for fuel gas applications. A number of techniques are available for acid gas removal including hot carbonate solutions, amine solutions, physical absorbtion in cold methanol or other solvents, and, in some cases, chemical reaction of H_2S with appropriate reagents. The reactions occurring are usually reversible so that the materials used are regenerated. This is the last step in fuel gas production.

The final step in the sequence for producing SNG or hydrogen is methanation by the following reactions:

$$CO + 3H_2 = CH_4 + H_2O + 87{,}000 \text{ Btu/lb-mole}$$

$$CO_2 + 4H_2 = CH_4 + 2H_2O + 71{,}000 \text{ Btu/lb-mole}$$

This is a major step in SNG production but is relatively minor when hydrogen is being produced because most of the CO has been removed in the shift section of the plant. SNG is compressed to pipeline pressure and dried.

A number of auxiliary facilities is required for many plants. If oxygen is used in gasification, then an oxygen plant is required. The sulfur compounds from acid gas removal are converted to sulfur in a separate plant if the conversion is not effected in the removal step. For most plants, steam and power must be generated by combustion of an appropriate fuel. Cooling towers, waste water treatment and fresh water treatment are required in all cases. In certain instances, other facilities are required. For example, in the CO_2 Acceptor process, an acceptor regenerator is necessary.

Each of the steps in the overall gasification scheme are discussed in the following sections for the different processes.

2.2 Coal Storage and Preparation

2.2.1 Description of Coal Storage and Preparation

Table 3 gives a summary of the coal preparation and storage assumptions used in the designs. A more detailed description of the individual coal preparation sections is given in Appendix A. A variety of coals were selected by the developers for the processes studied and, thus, comparisons of the processes are difficult. A summary of feed coals with analyses is shown in Table 4. About 30 days storage was assumed for most processes. The size of the coal feed is dictated by the nature of the process and varies from 70% less than 200 mesh up to 1-3/4 inches.

All the coals are dried except for those used in the Lurgi and Synthane processes. In some cases, especially when the moisture content is very high, it is necessary to dry the coal for smooth operation of the process. In others the coal is dried to reduce the heat load in the gasifier, lessening the oxygen requirements. As can be seen from Table 3, a variety of fuels can be used for drying the coal. As indicated below, the purpose of using clean fuel gas in drying is to reduce stack emissions.

Table 3

Coal Preparation and Storage Operations-Gasification

Process	Coal Type	Quantity Stored, tons	Size of Coal Feed	Fuel for Coal Drying	Other Operations
Koppers-Totzek	Navajo Sub-bituminous	200,000	70% < 200 mesh	Coal/Product Gas	---
Synthane	Pittsburgh Seam	400,000	70% < 200 mesh	N.S.	---
Lurgi	Navajo Sub-bituminous	720,000*	1-3/4" x 5/8" & 3/8" x 3/16"	None	---
Co$_2$ Acceptor	Lignite	800,000	8 x 100 mesh	Lignite/Product Gas Fired Furnace	---
BI-GAS	Western Kentucky No. 11	700,000**	70% < 200 mesh	Coal	---
HYGAS	Illinois No. 6	530,000	< 8 mesh	Fuel Gas	Slurry Formation
U-Gas	Pittsburgh Seam	220,000	< 1/4"	Product Gas	---
Winkler	Lignite	600,000	< 4 mesh	Product Gas	---

---

* Plus 650,000 for emergencies.
** Contains 21% refuse.

N.S. = Not specified

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Table 4

Coal Analyses - Gasification

		Proximate, %				Ultimate (MAF), %						
Process	Coal Type	Fixed Carbon	Volatiles	Ash	Moisture	C	H	N	S	O	Other	Higher Heating Value, Btu/lb
Koppers-Totzek	Navajo Sub-bituminous	35.0	31.2	17.3	16.5	76.72	5.71	1.37	0.95	15.21	0.04	8,830
Synthane	Pittsburgh Seam	--	--	7.4	2.5	81.9	5.8	1.7	1.8	8.9	--	13,700
Lurgi	Navajo Sub-bituminous	--	--	17.3	16.5	76.72	5.71	1.37	0.95	15.21	0.04	8,872
CO_2 Acceptor	Lignite	--	--	7.47	33.67	70.53	4.71	1.17	1.00	22.59	--	7,376
BI-GAS	Western Kentucky No. 11	45.4	39.5	6.7	8.4	80.20	5.50	1.62	4.10	8.58	--	12,330
HYGAS	Illinois No. 6	--	--	10.79	6.48	78.45	5.43	1.53	4.75	9.85	--	12,600
U-Gas	Pittsburgh Seam	--	--	10.7	6.0	80.70	5.64	1.35	4.97	7.34	--	12,387
Winkler	Lignite Type	--	--	14.5	13.3	71.2	5.4	0.8	4.3	18.3	--	8,910

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Other operations in coal storage and preparation include slurry feed formation in the HYGAS Process. The use of a slurry feed obviates the problems of lock hoppers operating at high pressures (∿1000 psi) and the consequent handling of lock hopper gas. Evaporation of the slurry liquid is required, however, and pumping slurried coal represents heavy duty on pumps.

2.2.2 Effluents to Air from Coal Storage and Preparation

The coal storage piles represent a potential source of air pollution from dusting and possible fines. In all cases, the storage piles are large and have a large surface area, thus winds can remove significant quantities of dust. Spontaneous combustion could produce obnoxious fumes and proper compaction of the coal piles is necessary (Ref. 14, p. 296-306). Lignites are especially prone to catch fire, but in all cases, proper monitoring of temperatures should be carried out and means should be available for extinguishing fires if they occur.

All coal handling steps are potential sources of dust. Covered conveyors should be used and spills should be recovered promptly or at least maintained in a wet state until recovery is possible.

Crushing and grinding operations can be dusty and should probably be carried out in enclosed spaces provided with sub-ambient pressure control and bag filters. The enclosures would also reduce noise although personnel within the buildings should be properly protected. Environmentally sound disposition of the collected coal dust must be provided. For processes using fine coal, this should present no problem. If fine coal cannot be used in the conversion process, it may be necessary to burn the coal dust for steam generation; in which case, adequate control of stack gas emissions must be provided.

Drying operations present a source of potential pollution. If clean gas is used for drying, one source of pollution is reduced. In all cases, particulate control is necessary since the coal is contacted with a large volume of hot gas. For example, to meet the particulates standard of 0.1 lbs per MM Btu (the level required of stationary boilers) the lignite loss in the CO_2 Acceptor process would have to be less than 0.01 weight percent.

Control of NO_x formation may be desirable. Flame temperature should be kept low and excess oxygen content should be limited to about 10%. This can be accomplished by recycling vent gas. Inert gas (nitrogen from the oxygen plant or carbon dioxide from acid gas removal) can be added to reduce flame temperature and moisture content of the dryer gas. Each process must be considered individually in order to minimize pollution and costs.

2.2.3 Liquids and Solids Effluents from Coal Storage and Preparation

The first problem is due to rain. The storage pile has a very large surface area and the residence time is long so that rain has a chance to react and form acids or extract organics, sulfur, and soluble metals, and in any event contribute suspended matter to the rain runoff. Therefore, it is necessary to collect water from this area as well as from the process area, and send it to a separate retention pond. This pond should have a long enough residence time for solids to settle out; also, there will be a certain amount of biological action which will be effective in reducing contaminants. Limestone can be added in this circuit if needed to correct acidity. The problem may bear some resemblance to acid mine water and should be reviewed from that standpoint (15,16). Run off from the dolomite storage area should also be treated.

In some comparable situations, seepage down through a process area can be a problem in addition to the runoff. Even though storm sewers collect the runoff in a chemical plant or refinery, leaks and oil spills can release enough material that it actually seeps down into the ground water supply. If the ground contains a lot of clay this will not normally be a problem - in fact, the clay can absorb large quantities of metallic ions. In sandy soil it may be necessary to provide a barrier layer underneath the coal storage piles. This could be concrete, plastic or possibly a clay layer.

Water from the coal drainage retention pond will be relatively clean and low in dissolved solids and is therefore a good makeup water for the cooling tower circuit and for preparation of boiler feed water. Normally all of the runoff water can be used in this way so that it will not constitute an effluent from the plant.

No specific solid or liquid effluents are expected from the coal or dolomite grinding, drying, and preheating sections. Coal dryer vent gas will be passed through bag filters to recover the dust. It can be combined with the ash slurry and returned to the mine. Electrostatic precipitators or scrubbers may be used instead of bag filters.

In the BIGAS design, considerable refuse is removed at the plant. It is probable that the refuse will be returned to the mine. Wash water should be sent to a settling pond and recycled.

2.2.4 Process Alternatives

An alternative to minimize dusting and drainage from coal piles is to use the piles only as "dead storage" (17). This stored coal would be used only in emergencies. A much smaller quantity of coal could be stored in silos for day-to-day use. The emergency storage can be covered with a a coating of polymer or asphalt. This reduces the drainage and dusting problems. A further advantage would be loss of coal value due to slow reaction with air. This reaction should decrease with time and coal value losses will be minimized. The use of a cap is, however, contrary to previous recommendations (Ref. 14, p. 298) and should be used with care.

A number of options exist for minimizing air pollution in coal preparation. To minimize coal dusting, for those plants using fine coal feed, the coal can be dried in a relatively coarse form with subsequent grinding. Drying offers a number of other alternatives for optimization with respect to cost and pollution.

One major area for optimization is trade-off between heat load in the gasifier and dryer. This should especially be considered in low Btu processes using air for gasification. Some of the heat of drying in the gasifier can be recovered in subsequent steps in the process. Smaller dryer gas volumes can be used if the moisture content of the coal feed is allowed to increase.

Another major area to be considered is the use of clean gas for dryer fuel vs the use of coal with stack gas scrubbing. The latter alternative should be effective in removing particulates and sulfur.

Nitrogen or carbon dioxide from the process can be used to reduce the oxygen content of the dryer gas. This increases drying capacity of the gas over that obtained by gas recycle.

In those areas where water is a premium, much of the moisture from the coal dryer gas could be recovered using air fin condensers. This might be very useful for Western coals and lignite where the moisture content of the coal is high and fresh water is scarce.

2.3 Gasification and Quench Sections

2.3.1 Gasifiers and Operating Conditions

The gasifiers examined in this study include several types. These range from a counter-current, slowly moving bed to fluidized beds to entrained flow. Temperatures vary considerably, often in the same bed, and range from 600°F in the dryer of the HYGAS process to 3000°F in the bottom of the BIGAS process. Pressures vary widely, from essentially atmospheric pressure in the Koppers-Totzek process to 1200 psia in the HYGAS process. Both air and oxygen gasifiers were examined. The products from the processes include low, medium and high Btu gas. (The processes producing high Btu gas necessarily produce a medium Btu gas before methanation.) A summary of the various process gasifiers is shown in Table 5 together with operating conditions and type of final product gas. A more detailed description of the gasifiers is given in Appendix A and in references 3-10.

The inputs to the gasifiers are given in Table 6. The quantities of coal/lignite feed shown for the processes is actual feed dried to the moisture contents given as footnotes. The CO_2 acceptor process is different in that air is fed to the acceptor regenerator rather than to the reactor. Except for the Koppers-Totzek and U-Gas processes, the Btu contents of the final product gases are roughly the same (231-250.3 X 10^9 Btu/day). The Koppers-Totzek design produces 88.7 X 10^9 Btu/day while the U-Gas design produces 124 X 10^9 Btu/day. These products are discussed in a later section.

Table 5

Gasifier Descriptions and Operating Conditions

Process	Type	Oxidant Supplied	Temperature, °F	Pressure, psia	Product Gas
Koppers-Totzek	Entrained Slagging	oxygen	2700	15	Medium Btu
Synthane	Fluid bed	oxygen	Top - 800 Bottom - 1700	1000	High Btu
Lurgi	Counter-current bed	oxygen	Top - 1100 - 1400 Bottom - ∿1700	420	High Btu
CO_2 Acceptor	Fluid bed	air*	1500	150	High Btu
BI-GAS	Top zone - entrained bottom zone - slagging	oxygen	Top zone - 1700 Bottom Zone - 3000	1200	High Btu
HYGAS	Fluid bed 4 sections	oxygen	Top - 600 2nd Sect. - 1250 3rd Sect. - 1750 Bottom - 1900	1200	High Btu
U-Gas	Fluid bed	air	1900	350	Low Btu
Winkler	Fluid bed	oxygen	1700	30	Medium Btu

---

*To Acceptor regenerator

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

The products in the raw gas from the gasifier/quench section of the plants are tabulated in Table 7, based on dry gas. Other materials leaving the gasifiers in various streams are given in Table 8.

In Table 8, some of the solid effluents from the gasifiers are classified as ash while others are shown as char. The distinction is subjective as some carbon remains in the ash and some chars have relatively low carbon. The analyses of those solids listed as chars are shown in Table 9. Steam is also produced in all cases from the gasifier jackets or in waste heat boilers.

2.3.2 Gasifier Effluents to Air

No major gaseous effluent streams are expected from the gasifier/ quench sections of the plants. It is expected that inert gas or steam used in pressurizing lock hoppers will be returned to the main gas stream. Care must be taken that sources of dust from dry ash or char does not enter the atmosphere. Quench systems for ash or char should be designed to prevent effluent odors, if present. For more details of the containment of gaseous effluents from the gasifier sections of the plants, the individual process reports (3-10) should be consulted.

2.3.3 Liquid and Solid Effluents

The largest liquid and solid effluent streams from the gasifier section of the plants are the ash or char streams. For those processes utilizing char as fuels, these streams are not effluents at this point.

The ash is usually recovered as a slurry and may pass to settling ponds, be returned to the mine, or may have a use such as land fill. In all cases, there exists the possibility of leaching of inorganic materials into general water systems. This can be prevented by using linings for ponds where the soil is sandy. Linings will not be necessary if the soil has a large adsorptive capacity for the soluble ions.

Dirty water streams from the quench sections of each process are sent to some form of waste water treatment. This treatment is reported to consist only of settling ponds for some streams such as ash slurries, but the treatment may be extensive for those streams containing phenols, ammonia, etc. The waste water treatment systems will be discussed later.

There is a purge stream of slurry oil from the HYGAS process that may require treatment. It may contain organic materials as well as trace elements. The disposition of this stream will depend on further definition of its analysis.

2.3.4 Process Alternatives

No major process alternatives exist for the gasifiers since each is defined by the developer. Minor alternatives such as lock hoppers vs slurry feeding, method of pressurizing feed hoppers, methods of ash removal and techniques for quenching the various raw gas streams are discussed for each process in the process reports (Ref. 3-10). A good discussion of these alternatives is presented in the Synthane report (Ref. 4).

Table 6

Inputs to Gasifiers

(lb/hr except as noted)

Process	Coal or Lignite	Higher Heating Value, Btu/lb coal*	Steam	Oxygen	Air
Koppers-Totzek	479,300 (1)	10,327	84,700	326,900	--
Synthane	1,187,500 (2)	13,700	1,169,700	304,000	--
Lurgi	1,722,200 (3)	8,872	1,762,200	468,500	--
CO_2 Acceptor	1,413,400 (4)**	10,945	1,653,700	--	3,373,400***
BI-GAS	946,300 (5)	13,285	409,700	497,600	--
HYGAS	1,057,900 (6)	12,600	981,700	270,300	--
U-Gas	575,400 (7)	13,178	371,750	--	1,849,000
Winkler	1,675,000 (8)	9,320	820,800	961,300	--

Notes:
* With moisture as shown in notes 1-8.
** 7,164,000 lb/hr hot acceptor also enters from regenerator
*** Air used in regeneration of acceptor

(1) 2% Moisture
(2) 2.5% Moisture
(3) 16.5% Moisture
(4) 0% Moisture
(5) 1.3% Moisture
(6) 0% Moisture
(7) 0% Moisture
(8) 8.7% Moisture

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Table 7

Raw, Dry Gas from Gasifiers and Quench

(lb/hr)

Process	CO	H_2	CO_2	CH_4	H_2S	COS	N_2	Higher Hydrocarbons
Koppers-Totzek	575,300	22,200	88,900	600	3,400	700	11,000	0
Synthane	320,000	38,200	871,000	268,000	12,200	N.R.	16,000	15,000
Lurgi	535,500	76,500	1,243,800	174,000	10,700	N.R.	8,800	28,900
CO_2 Acceptor*	431,600	145,000	308,500	98,900	1,142	N.R.	6,200	N.R.
BI-GAS**	1,024,300	40,900	512,300	207,300	40,600	N.R.	15,300	N.R.
HYGAS	650,100	48,300	763,800	244,200	43,300	700	1,700	15,100
U-Gas	520,800	25,600	422,900	72,400	25,600	1,400	1,407,900	N.R.
Winkler	1,094,800	85,700	1,066,500	32,000	51,250	10,000	34,000	N.R.

---

* Does not include gas from acceptor regenerator

** Output includes 104,100 lb/hr (dry) recycled product gas

N.R. = Not reported

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Table 8

Other By-products from Gasifier and Quench

(lb/hr)

Process	Ash	Char	Tar & Oil	Phenols	NH_3	Hydrocarbon liquids
Koppers-Totzek	111,500	-	negligible	negligible	negligible	negligible
Synthane	-	362,200	43,200	N.R.	13,200	7,400
Lurgi	314,000	-	126,400	10,100	16,900	18,400
CO_2 Acceptor	*	496,800**	N.R.	N.R.	N.R.	N.R.
BI-GAS	68,400	-	N.R.	N.R.	7,700	N.R.
HYGAS	-	138,900	N.R.	1,300	11,300	39,800
U-Gas	-	86,400	N.R.	N.R.	N.R.	N.R.
Winkler	-	372,500	N.R.	N.R.	N.R.	N.R.

---

* See regenerator section in section 2.16.1.

** Char passes to regenerator. 7,977,000 lb/hr of acceptor passes to regenerator section.

N.R. = Not reported

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Table 9

Char Analysis

Process	Char Analysis, wt. %						HHV, Btu/lb
	C	H	O	S	N	Ash	
Synthane	71.4	0.9	1.8	1.5	0.5	23.9	11,000
CO_2 Acceptor*	63.41	0.54	2.26	0.97	0.25	32.57	9,450
HYGAS	10.3	N. A.	N. A.	N. A.	N. A.	N. A.	1,488
U-Gas	20.33	1.43	-	0.58	1.78	75.88	3,877
Winkler	31.7**	N. A.	N. A.	N. A.	N. A.	N. A.	4,810

---

* Char is burned in acceptor regenerator

** Average of two streams

N. A. = not available

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

2.4 Shift Conversion and Cooling

Shift conversion, that is, the reaction of carbon monoxide with steam to produce hydrogen and carbon dioxide, is only used in those processes where SNG is the final product and only then when there is not sufficient hydrogen in the raw gas to effect the methanation step. (Processes designed for maximum hydrogen production would of course also use the shift reaction.) Of the processes studies in this work, only the Synthane, Lurgi, BIGAS and HYGAS processes use shift conversion.

2.4.1 Description of Shift Conversion

When SNG is to be the final product of the process, it is usually necessary to convert carbon oxides to methane by hydrogenation. Since the principal reaction is carbon monoxide with hydrogen, a ratio of hydrogen to carbon monoxide of about 3:1 is required prior to methanation. This ratio is obtained in the shift reactor section of the plant by reacting carbon monoxide with steam in the following reaction:

$$CO + H_2O = CO_2 + H_2 + 17{,}770 \text{ Btu/lb-mole.}$$

(The CO_2 Acceptor process is an exception since sufficient hydrogen is present in the raw gas.) Usually, only a fraction of the total raw gas stream passes through the converter system since only part of the carbon monoxide is reacted. Before entering the converter reactors, the gas is usually washed to remove most of the tars, dust, etc., to prevent bed plugging.

Although low temperature catalysts (ca. 450°F) are available for carbon monoxide conversion, these catalysts are deactivated by sulfur compounds. In the designs for SNG production, it has been assumed that acid gas removal is most economically carried out after the shift reaction so that carbon dioxide, formed during shift conversion, is also removed. Thus, high temperature (ca. 700°F) catalyst are used that are not grossly affected by sulfur compounds. These usually consist of chromia promoted iron oxide and have a life of up to three years. The exothermic heat of the shift reaction is removed by intercooling and preheating the cool raw gas.

After the shift reaction the shifted gas and bypassed gas are cooled and remixed. During cooling, as much useful heat is recovered as possible. Also during cooling, organic compounds may be removed and sent to storage or to other units in the plant. Large quantities of water are condensed and must be treated prior to reuse or discharge. (In some cases, at least part of the dirty water can be used for quench.) The cooled gas is then sent to the acid gas removal section for further purification.

A more detailed description of the individual shift converter sections may be found in Appendix A.

2.4.2 Effluents to Air from Shift Conversion and Cooling

There are normally no effluents to the air from the shift conversion and cooling section of the plant; any vent gases are collected, recompressed and returned to the system.

2.4.3 Liquids and Solid Effluents from Shift Conversion and Cooling

The only solid effluent from the shift conversion section is the periodic catalyst removal required after about three years operation. The relatively small quantities involved should present no disposal problems. A note of caution is warranted, however. It is possible that there would be a buildup of trace metals on the catalyst that could present environmental problems. The spent catalyst should be examined carefully before disposal to assure that the disposal method used will be environmentally sound.

Liquid streams leaving the shift/cooling section may include oil products to storage or other use and contaminated water. The latter must be treated and will be discussed later in Section 2.11. The quantities of dirty water leaving the shift conversion and cooling areas are shown in Table 10. The water from the cooling areas is also included for those processes without a shift reactor.

2.4.4 Process Alternatives in Shift Conversion and Cooling

Few process alternatives exist in the area of shift conversion. The technique is quite old and most variables have been optimized.

One alternative that might offer advantages in some cases is the use, as much as possible, of the dirty water before treatment. This is done in the BIGAS process. Use of the water in place of steam would offer credits for steam production as well as decrease the load on waste water treatment. It should be noted, however, that water in the Koppers-Totzek process is relatively clean and requires only a settling pond for treatment for reuse.

Cooling of the gas stream prior to acid gas removal should be carried out so as to conserve as much heat as possible for subsequent use. The level at which this heat is recovered will be determined by its subsequent utilization. Air fin coolers can be used as far as possible in the final cooling to conserve cooling water.

2.5 Acid Gas Removal

The acid gas removal section of the plant has the duty of removing sulfur compounds, carbon dioxide and any other materials that would interfere with subsequent methanation. There is a large number of options for this section and no attempt will be made to describe them all. Brief descriptions will be given of those chosen for the processes in the present study together with the effluents from each as far as information is available.

2.5.1 Description of Acid Gas Removal

The procedures chosen for acid gas removal generally involve chemical or physical absorption of the acidic materials in a suitable liquid with subsequent desorption of the acid gases at a lower pressure

Table 10

Sour Water from Shift Conversion, Cooling and Scrubbing

Process	Water, lbs/hr	Disposition
Synthane	1,110,000	To waste water treatment
Lurgi	1,277,500 (1)	To waste water treatment
BI-GAS	866,600	To quench (2)
HYGAS	806,500	To waste water treatment
Koppers-Totzek (3)	7,142,800 (4)	To clairifier
CO_2 Acceptor (3)	612,000	To waste water treatment
U-Gas (3)	230,800	To waste water treatment
Winkler (3)	928,300	To waste water treatment

---

(1) Contains sour water from initial cooling

(2) Perhaps 86,000 lb/hr must be treated to prevent build up of trace contaminants

(3) No shift conversion. Sour water from quench and cooling

(4) This water is reported not to contain sour components; the large quantity is needed for solids removal.

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

(and in some cases at higher temperatures) to regenerate the absorbent. Table 11 lists the acid gas removal techniques suggested for the processes studied in this work. It should be noted that acid gas removal is a major consumer of utilities.

The hot carbonate process has been described in a number of publications (e.g., Ref. 18 and 19). Basically, the process involves absorption of acid gases in a solution of potassium carbonate (and additives) at about 230°F. The acid gases are desorbed in a regeneration tower at lower pressure with steam stripping. Variations (such as operation at a lower temperature in the absorber) have been suggested to increase absorption. Other sulfur compounds, such as COS, CS_2 and mercaptans can be removed to a certain extent depending on conditions. Thiophenes should not react with the carbonate solution but partial removal has been reported (18). Cyanides and sulfur dioxide may react irreversibly with the solution. By modification of the design two acid gas streams can be obtained: one high in sulfur content (suitable for a Claus plant) and the other high in carbon dioxide. The latter stream will still contain significant sulfur as hydrogen sulfide that will have to be incinerated or removed (see the description of the BIGAS process (7)).

The Rectisol cold methanol process operates by absorption of acids in methanol at reduced temperatures (ca. -50°F) and has been described in the literature (e.g., Ref. 20). This process is capable of removing all the types of sulfur compounds but can also remove significant quantities of combustibles. One design (21), after reducing combustibles to a minimum, incinerates the acid gas after sulfur removal. Although it is possible to obtain a relatively pure CO_2 stream and a high concentration of H_2S in a separate stream, the relationship between the loss of product gas and the concentration of H_2S in such an arrangement is not clear. (The unit described in Reference 21 produces only one stream with a low H_2S content unsuitable for a Claus plant.) The Rectisol process may use stripping gas (N_2) in some cases and can be integrated with the final product gas compression step to remove water from the final product gas.

The Koppers-Totzek process makes use of an amine system (methyl diethanolamine) for acid gas removal at about 120°F. This system is capable of producing a high concentration of H_2S in the sour gas stream which can be sent to a Claus plant. Several hundred parts per million of sulfur compounds and most of the CO_2 remain in the product gas, but for fuel use this is acceptable. If it were necessary to methanate the product gas, further treatment would be necessary.

Selexol acid gas removal, indicated for use in the U-Gas process, absorbs acid gases in dimethoxy tetraethylene glycol. (See References 22 and 23 for a description of the Selexol process.) A high concentration of H_2S in the product gas stream can be obtained by this process, but no information is available as to the concentration of product gas in the acid gas stream.

Table 11

Acid Gas Removal

Process	Type of Acid Gas Removal	Volume of Acid Gas, MM scfh (1)	Acid Gas Analysis (1)		HHV of Acid Gas, Btu/scf (1)	Type of S Guard
			V % H_2S	V % Total S Compounds		
Koppers-Totzek	Methyl diethanolamine	0.138	23.1	24	N. S.	N. N.
Synthane	Hot carbonate (Benfield)	9.2	1.5	N. S.	35 (2)	Iron oxide/ char (or activated carbon)
Lurgi	Cold methanol (Rectisol)	13.5	1.1	N. S.	38	Zinc oxide
CO_2 Acceptor	N. S.	0.22 (3)(4)	5.9 (4)	N. S.	N. S.	Zinc oxide
BI-GAS	Hot carbonate (Benfield)	2.92 (5)	14.6	N. S.	N. S.	Zinc oxide
HYGAS	Cold methanol (Rectisol)	1.71 (6)	29.8	N. S.	N. S.	Zinc oxide
U-Gas	Dimethoxy tetraethylene glycol (Selexol)	1.58	17.9	18.2	N. S.	N. N.
Winkler	Hot carbonate	4.04	15.0	15.0	N. S.	N. N.

N. N. = Not needed
N. S. = Not specified

(1) Dry Gas
(2) Estimated
(3) Does not include gas from regenerator
(4) N. S. if wet or dry gas
(5) Does not include all CO_2 - 9.91 MM scfh vented separately
(6) Does not include all CO_2 - 9.88 MM scfh vented separately

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

The sulfur content of the gas from the acid gas absorption system is usually decreased further by reaction with iron oxide, or zinc oxide or by adsorption. This step is frequently necessary to protect the methanation catalyst which is highly sensitive to the presence of sulfur compounds. The clean gas, if SNG is to be the final product, then passes to the methanation section. The acid gas stream containing the H_2S passes to a sulfur recovery plant.

2.5.2 Effluents to Air From Acid Gas Removal

The only atmospheric emission from the acid gas removal section is, in some cases, a carbon dioxide stream containing sulfur compounds and combustible materials (H_2, CO, CH_4, etc.). The quantities that could be emitted depend on the type of system used and the specific design of the system. The sulfur compounds, such as H_2S, COS, thiophenes, etc., can be removed by various treating processes, such as adsorption by molecular sieves. Alternatively, the combustible materials can be converted to carbon dioxide and the sulfur can be emitted as sulfur oxides by inceration. Unless the HHV of the carbon dioxide stream is sufficiently high, the cost of incineration can be expensive due to the large quantities of carbon dioxide that must be vented.

When guard boxes are to be regenerated (usually by air blowing at elevated temperatures), appropriate disposition of the exit gases must be available. These gases normally will contain sulfur oxides. The effluent may be directed to a furnace stack if the SO_x concentration is not too high. Otherwise some sort of scrubbing will be necessary.

2.5.3 Liquids and Solid Effluents from Acid Gas Removal

Condensate streams are formed in the acid gas removal sections of the plants. These streams are normally sent to a waste water treatment section. Build up of impurities in the absorption medium requires purging of the absorbent and the disposition of these purges requires individual examination. One frequently suggested technique of disposal is by incineration followed by mine burial of the solid residues.

When guard boxes are necessary prior to the methanation step, it is necessary to dispose of the spent solids from time to time. One suggested method is mine burial. A determination of leachability of the solids will be necessary to assure that contamination of ground water is not a problem. The solids should especially be examined to assure that potentially hazardous trace elements have not accumulated which could present an environmental problem. If such is the case, techniques will have to be devised to assure the environmental soundness of the ultimate fate of the solids.

2.5.4 Process Alternatives in Acid Gas Removal

Besides those discussed in 2.5.1, other alternatives exist for removal of acid gases from the main gas streams (24, 25, 26). In particular, it should be mentioned that aqueous solutions of monoethanolamine (MEA) and diethanolamine (DEA) have been used for removal of acid gases from gas streams (see, for example, Reference 27). If COS is present, MEA reacts with it irreversibly, while the COS passes through DEA. The MEA and DEA are not particularly selective for H_2S removal vs CO_2 removal. However, triethanolamine (TEA) preferentially removes hydrogen sulfide and a combination of TEA and a CO_2 removal system could be used to obtain a highly concentrated hydrogen sulfide stream for a Claus plant.

Alternatives for trace sulfur removal should also include the use of molecular sieves alone or in conjunction with methods discussed above.

Any type of acid gas removal unit chosen can be varied extensively. The choice of configuration will be dictated by such restrictions as gas composition, temperature and pressure, type of sulfur recovery facility, availability of excess steam, economics of the final trace sulfur clean-up system, and others. Each case must be examined individually, not only to choose the best type of acid gas removal system for the particular application, but also as to what modification to choose for the best type. It should be kept in mind, however, that ultimate disposition of effuents can be a major factor in the final choice for acid gas removal.

As in other cooling operations, air fin cooling rather than the use of cooling water can be advantageous in areas where water is scarce. Where the absorber and regenerator operate at different temperatures, heat exchange can be used to reduce the heat load. Another possible alternative to be considered is the use of heat pumps to minimize energy consumption. Still further energy conservation can be had by the use of liquid turbines in the depressurization of the absorber solution. These options must, of course, be considered from the standpoint of cost, availability and environmental effect.

2.6 Methanation Section

2.6.1 Description of the Methanation Section

When SNG is the desired final product, a methanation step is required. The reactions involved in methanation are

$$CO + 3H_2 = CH_4 + H_2O + 87{,}700\ \text{Btu/lb-mole}$$

$$CO_2 + 4H_2 = CH_4 + 2H_2O + 71{,}000\ \text{Btu/lb-mole}$$

It is usually desirable to reduce the need for the last reaction to conserve hydrogen requirements. Fortunately, the reaction of CO_2 is slow in the presence of CO. The above reactions are generally carried out over a nickel catalyst that is easily deactivated by sulfur compounds, hence the need for very clean feed gas.

Methanation has been used for years in, for example, ammonia plants where the levels of carbon monoxide to be removed has been low. In the production of SNG, the concentration of CO is high and special considerations are then necessary (28).

The methanation reactions are highly exothermic and it is necessary to design the unit to keep the temperature within limits dictated by catalyst life. It is also desirable to recover as much as possible of the heat released in the reactions at as high a temperature as possible. Other design considerations involve the possible formation of nickel carbonyl at low temperatures and the reaction of CO to give CO_2 and carbon at high temperatures; methanators usually operate at about 750°F.

Temperature control can generally be effected by large recycle of the cooled effluent gas. This keeps the carbon monoxide low and hence the temperature rise is minimized. The U.S. Bureau of Mines (now PERC of ERDA) has proposed the use of a heat exchanger in which a nickel catalyst is sprayed onto the exchanger tube walls (4). Heat can then be transferred to a suitable liquid.

2.6.2 Effluents to Air from the Methanation Section

During normal operation, there should be no effluents to the air from the methanation section. During start-up, recycle of the process gas is necessary, and during shut-down, facilities are required for flushing the catalyst bed with inert gas and for oxidizing the catalyst with a stream containing low amounts of oxygen. The effluent gases can be incinerated.

There is the possibility of the formation of nickel carbonyl, especially at low temperatures, and care must be taken that this is not released to the atmosphere (or, for that matter, that the final SNG product is not contaminated).

2.6.3 Liquids and Solid Effluents

The only liquid from methanation is a relatively clean condensate that can be sent to raw water treatment. Gases evolved during decompression of the water should be recompressed and returned to the system. No solids leave this section except during catalyst replacement; the catalyst will probably be reworked to recover the nickel content.

2.6.4 Process Alternatives in Methanation

Few alternatives exist for methanation and these generally have to do with methods of heat recovery/temperature control. Internal cooling, as in the Bureau of Mines (now PERC of ERDA) technique, is one possibility. The generally accepted method is recycle of a large stream of cooled gas. Heat is then extracted from the hot gas from the reactor before recycle. However, the recycle compressor can be a large energy user. A desirable alternative, but one that is not available at present, would be a catalyst that was more tolerant of sulfur compounds.

2.7 Compression and Drying

For high Btu gas a compression step may be required to bring the gas to pipeline pressure. (For some other applications, compression of the gas from the atmospheric gasification processes may be required.) This compression does not release atmospheric pollutants but does require considerable energy. The gas is then dried, using, for example, a glycol system. (Systems using a cold methanol carbon dioxide removal step do not require further drying.) The water from the gas stream is sent to raw water treatment. Gases evolved during decompression are recompressed and returned to the gasification system. Thus there are normally no gaseous, liquid or solid effluents from the compression and drying section. The materials used for drying will have to be replaced infrequently.

2.8 Final Product Gas

Table 12 shows the analyses of the product gas produced in each of the processes studied, along with the total volumes, heat contents and pressures.

2.9 Oxygen Plants

All the gasification processes studied in this work except the CO_2 Acceptor and U-Gas processes, require an oxygen plant. Oxygen in some form is required to burn part of the coal to produce the heat required in the gasifier. The U-Gas process uses air to accomplish this. The CO_2 Acceptor process carries out the oxidation in a separate reactor where air can be used without contamination of the product gas with large quantities of nitrogen. Other than for process operability, the use of pure oxygen allows the production of a higher Btu product from the gasifier than could be obtained by the use of air and the consequent introduction of nitrogen into the gas stream. Reference 29 presents a good discussion of oxygen separation from air.

Table 13 lists the oxygen requirements of the processes studied. The oxygen plants are relatively clean; the major effluents to the atmospheres are those that come in with the air. The liquid effluent is the water from the air and this can be directed to boiler feed water treatment. However, oxygen plants consume considerable energy for compression; approximately 0.2 hp-hr is required per pound of oxygen. Supplying this energy represents the major environmental effect of the oxygen plant.

2.10 Sulfur Recovery

2.10.1 Description of Sulfur Recovery

Sulfur recovery is a major concern with respect to its effect on the environment. There are quite a number of alternatives available for sulfur recovery, each with its own problems. Sulfur recovery has long been an active area for research and development and has been discussed extensively in the literature (see, for example, Refs. 24,25,26,30,31).

Basically, sulfur recovery usually depends on the oxidation of sulfur according to the equation

$$H_2S + (O) \longrightarrow H_2O + S$$

Classically, the oxygen came directly from air but newer processes depend on intermediate compounds which oxidize the hydrogen sulfide.

Table 12

Net Dry Product Gas

Process	Volume of Product Gas, MM scfd	HHV of Product Gas, Btu/scf	Pressure of Product Gas, psia	Gas Analysis, Volume %					
				CH_4	H_2	N_2	CO_2	CO	H_2S + COS
Koppers-Totzek	290	303	166	0.1	32.6	1.2	5.2	60.9	0.03
Synthane	250	927	1000	90.5	3.6	2.1	3.7	0.1	----
Lurgi	251	972	915	95.9	0.8	1.2	2.0	0.1	----
CO_2 Acceptor	263	952	1000	93.0	4.8	0.8	1.3	0.1	----
BI-GAS	250	943	1075	91.8	5.1	1.9	1.1	0.1	----
HYGAS	260	1000	958	93.0	6.6	0.2	0.1	0.1	----
U-Gas	784	158	300	4.9	13.8	54.4	6.7	20.2	0.015
Winkler	886	282	Ca 15	2.0	42.7	1.2	15.1	38.9	0.08

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Table 13

Gasification Process Oxygen Requirements

Process	Oxygen Required, lb/hr	Oxygen Required, lb per MM Btu in Gasifier Feed Coal
Koppers-Totzek	326,900	66.04
Synthane	304,000	18.69
Lurgi	468,500	30.67
BI-GAS	497,600	39.58
HYGAS	270,300	20.28
Winkler	961,300	61.58

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

The Claus process, developed about 1890, oxidized H_2S over a bauxite or iron ore catalyst. Later modifications included oxidation of part of the hydrogen sulfide completely to recover heat and the subsequent reaction of the SO_2 with the remaining H_2S to produce sulfur. The latter technique allows operation at lower H_2S concentrations in the feed stream. At very low concentrations of H_2S, fuel must be added to the hydrogen sulfide stream to support combustion. The biggest problems with the Claus plant approach is that high concentrations of H_2S are required and the tail gas from the plant still contains sulfur at such a level as to make its atmospheric release undesirable.

The liquid phase production of sulfur has been used to decrease sulfur content of the exit gases. These processes use an intermediate compound such as vanadates to oxidize the H_2S and can operate with dilute feeds. Commercial examples of these processes are the Stretford (32, 33, 34), the Giammarco Vetrocoke (24) and the Takahax (24) processes. These processes suffer due to problems in removal of other sulfur compounds, such as COS, and difficulty of liquid effluent disposal.

Details on sulfur recovery for the various processes are summarized in Table 14.

2.10.2 Effluents from Sulfur Recovery

The principal effluent from sulfur recovery plants is the tail gas. In the past it has been common practice to incinerate Claus plant effluents. In coal gasification, the large volume of CO_2 in the effluent makes incineration expensive. Furthermore, the SO_2 content of the incinerated gas can be excessive.

A number of processes have been announced for removing most of the sulfur from the Claus tail gas (24). Among these are the Beavon, IFP, SCOT, Sulfreen and the W-L processes. In the Beavon and SCOT processes, the sulfur compounds are converted to H_2S. The H_2S, in the Beavon process, is converted to sulfur in a Stretford unit. In the SCOT process, the H_2S is separated from the CO_2 using a selective alkanolamine absorber and is returned to the Claus plant. In the IFP process, the tail gas is incinerated and scrubbed with aqueous ammonia. The sulfates are reduced to sulfites, SO_2 is generated from the sulfites in solution and is reacted with a slip stream of H_2S to produce sulfur. The Sulfreen process is an extension of the Claus process. The H_2S and SO_2 are reacted catalytically at low temperature to form sulfur. The W-L process produces SO_2 solutions by incineration of the Claus tail gas followed by absorption. The SO_2 is removed from solution and returned as a concentrated stream to the Claus unit.

The use of tail gas clean-up adds to the cost of the gasification plants. Also, those processes utilizing liquids usually have a liquid effluent to dispose of with attendant environmental consequences that must be taken into account. Each case must be investigated individually to determine the environmental effects and at present no firm commitments have been made as to the process to be used.

Table 14

Sulfur Recovery in Gasification Systems

Process	Type of Sulfur Recovery	Acid Gas to Sulfur Plant MM scfd (1)	H_2S Concentration in Acid Gas (1) V%	Total Sulfur Produced, lb/hr	Sulfur in Tail Gas, V ppm	HHV of Tail Gas, Btu/scf	Tail Gas Disposal
Koppers-Totzek	Claus	0.138	23.1	3,330	3,390	N.S.	Clean-up (N.S.)
Synthane	Stretford	9.2	1.5	11,670	5	26(7)	To boiler stack
Lurgi	Stretford	13.5(2)	0.93(2)	12,340	740	29	Incineration
CO_2 Acceptor	N.S.	0.22(3)(4)	5.9(3)	9,920	N.S.	N.S.	Incinerate and clean up with flue gas
BI-GAS	Claus	2.92(5)	14.6	35,130	2,431	N.S.	Clean-up (N.S.)
HYGAS	Claus	1.7(2,6)	29.8(2)	55,500	3,010	N.S.	Clean-up Wellman-Lord
U-Gas	Claus	1.58	17.9	23,580	N.S.	N.S.	Clean-up (N.S.)
Winkler	Claus	4.04	15.0	40,420	N.S.	N.S.	Clean-up (N.S.)

N.S. = Not specified
(1) Dry Gas.
(2) Does not include gas from auxiliary fuel gasification unit.
(3) Does not include gas from regenerator.
(4) N.S. if wet or dry gas.
(5) Does not include all CO_2 - 9.91 MM scfh vented separately.
(6) Does not include all CO_2 - 9.88 MM scfh vented separately.
(7) Estimated

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

If the concentration of H_2S in the acid gas to the sulfur plant is very low, then a liquid phase recovery of sulfur is necessary as indicated above. The available processes have difficulty in removing compounds of sulfur other than H_2S. If the sulfur content is sufficiently high, the gas from these processes must be controlled. One technique could be incineration as in the Lurgi design. This may be necessary in any case since most acid gas removal processes also remove combustibles that must be removed or destroyed before venting. (See the Lurgi and Synthane processes as examples, Table 11.) The liquid processes also have a liquid effluent for which inceration may be required (32,33). The effluent may be quite large, amounting, in some cases, to 0.2 to 0.3 gal/lb sulfur recovered(33). The resulting solids from incineration must be examined to determine if mine disposal is safe as they may contain heavy metals such as vanadium or arsenic as well as soluble salts. It would also be wise to determine that no heavy metals enter the atmosphere during incineration.

2.11 Ash and Solids Disposal

2.11.1 Description of Ash and Solids Disposal

The solids from the gasifiers are removed in different ways depending on the process. The Koppers-Totzek process removes molten slag at low pressure and quenches it with water. The Winkler process, also at low pressure, removes the char via water cooled screw conveyors. The Lurgi process, at intermediate pressures, uses a lock hopper. The remaining processes are conceptual and suggested methods of removal are indicated in Table 15.

It should be pointed out that not all the solids are removed directly from the gasifiers. For example, in the Koppers-Totzek process, only about one-half the solids are removed directly; the remainder exits with the raw gas and is subsequently removed by an elaborate series of washing operations. Smaller amounts of dust are carried overhead in the Lurgi gasifier and are removed in a tar scrubber and a final wash before shifting. A major portion of the solids in the Winkler process is removed from the raw gas by cyclone, water scrubbing and an electrostatic precipitator. In all cases, sufficient care must be taken to assure essentially dust free gas before shifting.

2.11.2 Effluents to Air from Solids Disposal

There should be little air contamination from solids handling and disposal from the gasifiers. Odors may occur when ash or char is quenched, but this must be checked in each case. Care must be taken to prevent dusting; dust can be controlled by keeping the solids moist.

2.11.3 Liquids and Solids Effluents from Solids Disposal

The solids from the gasifiers represent the largest source of solids effluents (directly or later from their fuel use). The water quench streams are also very large.

Table 15

Solid Gasifier Product

Process	Type of Solid	Quantity of Solid, lb/hr	Type of Cooling, Removal	Solids Disposition	Liquid Disposition
Koppers-Totzek	Slag	111,500	Water quench	To mine	Recycle
Synthane	Char	362,200	Dry let-down, fluid bed	To power plant	--
Lurgi	Ash	314,000	Water cool, ash locks	To mine	Used in plant
CO_2 Acceptor	Char/Spent Acceptor	496,800	Char to regenerator Spent acceptor overhead, water cool	N.S.	Recycle
BI-CAS	Slag	68,400	Water quench, lock hoppers	N.S.	Steam to Reactor Water N.S.
HYGAS	Ash/Char	138,900	Water cool, lock hoppers	N.S.	Returned to system
U-Gas	Char	86,400	Water cooling venturi throat	N.S.	Returned to system
Winkler	Char	372,500	Water cooled screw conveyors	To power plant	--

N.S. = Not specified

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Chars that can be utilized as boiler fuel are, of course, directed to the steam plant. Other solids present more of a problem. Permanent mine storage has been suggested in a number of cases but this may not always be an environmentally acceptable solution. Although the metal content of the solids was originally removed from the mine, its physical and chemical nature has been changed so that its return to the mine could present problems. The major potential problem involves leaching of contaminants which may show up in surface water streams or in sub-surface water. If the soil contains sufficient clay, run-off will be the chief problem, but in sandy soils the ions can move considerable distances in the soil. In these cases, consideration must be given to providing an impervious layer of asphalt, concrete or other material to prevent movement of the inorganic materials. If surface movement is the only problem, impoundment would solve the problem.

Liquids that have been in contact with the ash or slag present similar problems as the solids themselves. Recycle of clarified water can be used as much as possible but a purge may be necessary to prevent continued build up of dissolved solids. This purge can be directed to impervious evaporation ponds.

Perhaps solids from other areas of the plant will present as much of a disposal problem as the ash from the gasifiers even though the quantity is lower. Purge streams from various units (e.g. sulfur recovery) may contain hazardous, soluble metallic ions. Incineration and mine disposal may be the answer in most cases, but each process must be examined individually and in detail. Care should be taken in incineration that hazardous metals do not escape into the atmosphere as vapors or entrained liquids or solids.

2.11.4 Process Alternatives in Solids Disposal

Although alternatives are available for solids removal and disposal, in general, when the individual process and operating conditions are considered, very few options exist. Problems connected with solids removal are discussed in reference 4 and in the section on the Synthane process in Appendix A. Consideration should be given in all cases to the possibility of recovering valuable chemicals from the solids before disposal.

2.12 Wastewater Treatment

The handling of the process and cooling water streams can represent one of the major pollution problems in coal conversion plants. These water streams have the potential for both air, water and land pollution if not handled properly as they can give off gaseous, liquid and solid wastes. For economic and other reasons many conversion plants are seriously considering recycling all process water to extinction. The water treatment systems will have to be designed specifically for each plant; no one process will be universally applicable. The variety of coal sources and gasifier operating conditions will differentiate the aqueous wastes in the various processes under development.

Water treatment in coal conversion plants is very much like that used in petroleum refineries and petrochemical plants. Reference 35 discusses in detail the treatment of aqueous wastes from such plants. Reference 36 outlines the design of such treatment facilities.

Water treatment technology for petroleum refineries can be classified as primary, secondary and tertiary. Primary treatment can be described as gross removal of materials, secondary treatment provides for reasonably clean effluents, and tertiary treatment methods are for polishing the effluent or for removal of special materials to acceptable levels. Some of the methods of each classification are listed in Table 16.

Wastewaters are generally segregated in some fashion such as oily water (including oily rainwater run off), high solids clean water, sour water, and very hazardous waters. These streams are handled separately to minimize the size of treatment units.

The treatment system necessary for each process has not been specified. A detailed examination of the individual process would be necessary, including stream analyses, before such a system could be outlined. Table 17 shows the quantities of water treated by the gasification plants studied here. The treatment of rainwater run-off, minor purges, etc. is not indicated. In most cases, secondary and tertiary treatment has not been determined.

High temperature processes such as Koppers-Totzek and BIGAS form negligible quantities of heavy organic materials and there is little sour water to be treated. In the Koppers-Totzek case, it has been suggested that the water from the particulate removal spray can be directed to a clarifier and then recycled. This water has been in contact with hydrogen sulfide and should dissolve a certain quantity of this toxic substance. The H_2S would be removed from the system in the cooling tower. Such practice should be checked to see if it meets reasonable air enviornmental requirements.

Some water may contain such materials as phenols, acids, ammonia and sulfur compounds. In many of the processes considered, the sour water stream is large and requires special treatment. This usually involves phenol removal by extraction with a suitable solvent and stripping to remove ammonia and hydrogen sulfides. The phenols can be sold as such, burned or recycled. The ammonia can be sold or burned while the hydrogen sulfide can be routed through the acid gas removal section. Water leaving the strippers still contains materials that cannot be allowed to enter the environment. This water is sent to secondary treatment.

Suspended matter is usually removed by coagulation or flocculation. The sludge from these operations can be disposed of with other sludges from, for example, biological oxidation. Techniques available are described in Reference 35 and include incineration and burial. Final solids disposal can be handled with the ash from the gasifier.

Following removal of the suspended matter, a biological oxidation unit (biox) may be used to reduce further the contaminant levels of the water. Several techniques are available for biological treating including

TABLE 16

CLASSIFICATION OF WASTEWATER TREATMENT METHODS

Primary Treatment

- Stripping
- Primary Incineration
- Neutralization
- Oil Separation

Secondary Treatment

- Activated Carbon Adsorption
- Chemical Coagulation
- Flocculation
- Air Flotation
- Biological Treatment
  - Aerated Ponds
  - Activated Sludge Processes
  - Trickle Filter Processes
  - Biological Oxidation in Cooling Towers

Tertiary Treatment

- Chlorination
- Activated Carbon Adsorption
- Evaporation
- Ozone Oxidation
- Ion Exchange
- Reverse Osmosis
- Dialysis
- Precipitation

Table 17

Dirty Water Treatment Systems of Gasification Plants

Process	Total Dirty Water Treated, lb/hr (1) (2)	Sour Water, lb/hr	Secondary Treatment Type	Tertiary Treatment
Koppers-Totzek	8,297,800 (3)	0	None	None
Synthane	1,773,900	1,311,100	N. S.	N. S.
Lurgi	1,644,500	1,282,000	Activated sludge	Evaporation ponds
CO_2 Acceptor	612,000 (3)	612,000	N. S.	N. S.
BI-GAS	686,000 (4)	86,000	N. S.	N. S.
HIGAS	809,600 (5)	809,600	N. S.	N. S.
U-Gas	397,400	230,800	N. S.	N. S.
Winkler	928,300 (5)	928,300	N. S.	N. S.

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(1) Does not include rain water, miscellaneous purges, filter backwash, septic sewer, stack gas scrubber.
(2) Does not include "clean" water from condensate in oxygen plant, methanation or compression.
(3) Only clarifier treatment used and water is recycled.
Cooling tower blowdown is disposed of with ash in mine.
(4) Disposition of cooling tower blowdown N. S. S.
(5) Cooling tower blowdown not included; its disposition N. S.

N. S. = Not specified

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

activated sludge treatment, aeration ponds, trickle filters and biological oxidation in cooling towers. Such techniques are very effective in removing phenol and reducing BOD but problems still exist. For example, an activated sludge system has been found inadequate in ammonium ion removal and erratic in its removal of cyanide and thiocyanate (37). It has been pointed out that reduction of BOD does not necessarily mean that all harmful organics have been removed (38). Materials such as polynuclear aromatics may not show up in a BOD test and, likewise, would not be removed by bacterial action in a biox unit. A third problem arises from the possibility of air pollution from biox units. The large liquid surface area necessary for transport of oxygen and carbon dioxide makes biox units ideally suited for stripping of contaminants into the air (39). This area of foul water treatment deserves close attention when used in a coal gasification plant for it would be very easy to convert a water pollution situation into an air pollution problem. A final problem with biological oxidation involves its sensitivity to upset conditions. Fluctuations in the concentration of contaminants in the water influent may cause a decrease in the biological activity. In some cases, the activity can be destroyed by a sudden increase in some component (e.g. cyanide).

Activated carbon adsorption can be used as a final polishing step after a biox unit or may be substituted for such a unit. The performance of carbon adsorption of materials from petroleum refinery effluent has been investigated (40). Advantages of activated carbon are that effluents are concentrated and can be disposed of rather easily and the system is relatively insensitive to fluctuations in contaminant concentrations. A disadvantage is the semi-batch nature of the process with its necessary regeneration step.

The use of API separators for bulk oil removal are usually necessary in gasification plants. Not only is it frequently necessary to treat oily process water, but rain water run off from process areas and tank farms contains oil that must be removed. It may be necessary to follow the separation with flotation units before sufficient oil is removed.

Non-oily rain water run-off can usually be impounded and used, after raw water treatment, as make-up water. Boiler blowdown water can also be used as cooling tower make-up water.

One of the larger streams in gasification plants is the blowdown from the cooling water system. Not only does this water contain large amounts of dissolved solids but may contain other contaminants introduced through leaks in heat exchangers, pump seals, etc. Furthermore, materials, such as chromimum, added to prevent algae formation in the cooling towers, present a special problem in water treatment. Chromimum can be precipitated before other treatment of the blowdown water or ion exchange can be used for metals removal. In the Koppers-Totzek and CO_2 acceptor processes, cooling tower blowdown is disposed of with ash in the mine. In the Lurgi process, final disposal is provided by evaporation ponds. Blowdown water may be eliminated by using softened water for cooling. Drift loss in the cooling towers keeps the solids level sufficiently low in the cooling water circuit. In areas where water is scarce, this total recycle of cooling water might be especially attractive.

In any gasification plant there will be minor streams to be considered. These will include minor purge streams, filter backwash, contaminated water from raw water treatment, etc. These streams must be considered individually and treatment may consist of special techniques such as neutralization, precipitation, incineration, and evaporation ponds.

In conclusion, it should be stated again that careful evaluation of waste water treatment is necessary. Care should be taken to see that contaminants are not transferred from the water to the air and proper management of solids, which are often the product of water treatment processes, is necessary. Further work in the area of water treatment is needed.

2.13 Power and Steam Generation

2.13.1 Alternatives in Power and Steam Generation

A number of alternatives exist as regards the methods and fuels used to generate power and steam and the resulting pollution problems. Our basis for design of all gasification plants has been one wherein the plants were self-sufficient with respect to steam and electrical requirements. Table 18 shows the steam and electricity requirements for each process, together with type and quantity of fuel and whether or not flue gas scrubbing is required.

No separate steam plant is required for the CO_2 Acceptor, U-Gas and Winkler plants. In fact, so much by-product high pressure steam is available in the CO_2 Acceptor process that all electricity needed in the plant and mine could be produced by bleeding the high pressure steam to 165 psi and 377,000 lb/hr of 165 psi steam would be available for sale.

The major area where a number of alternatives are available is that of fuel used to generate steam. Alternatives include the use of coal, clean intermediate product, clean final product, char, and manufactured low Btu gas. In the processes studied here, alternatives chosen are coal, char, and manufactured clean, low Btu gas. The use of manufactured gas or product gas decreases thermal efficiency of the overall process but has the advantage of low sulfur emissions. If coal or char is used as fuel, then stack gas scrubbing is frequently required to remove sulfur and particulates. This too, of course, reduces thermal efficiency over the use of coal or char alone and increases liquid or solid effluents.

In some cases an alternative to reduce sulfur emissions is the use of some clean product gas along with coal for fuel to the boilers. In this way, sulfur can be reduced to acceptable levels and particulates can be removed by electrostatic precipitators. This elminates the large quantities of spent limestone that must be disposed of from scrubbing operations. It was estimated that 150-200 tpd of sulfated lime would result from scrubbing the flue gas from the Synthane process (4).

Table 18

Generation of Steam and Electricity in Gasification Plants

Process	Steam Plant, lb/hr	Boiler Fuel		Flue Gas Scrubbing	Total Steam Generated,* lb/hr		Electrical Generation, kW
		Type	Quantity, lb/hr (MM Btu/hr)		High P	Low P	
Koppers-Totzek	646,000	Coal	113,300 (1,000)	Yes	1,307,800	109,500	19,400
Synthane	2,840,000	Char	362,500 (3,980)	Yes	3,346,800	809,600	6,000
Lurgi	1,488,800 (1)	Low Btu gas	(1,725) (3)	No	3,067,900	1,263,100	58,500 (2)
CO_2 Acceptor	(4)	-	-	-	2,142,000 (4)	253,000	17,500
BI-GAS	1,329,400	Coal	179,600 (2,220)	Yes	2,931,200	670,600	41,900
HYGAS	N. S.	Low Btu gas	(2,923)	No	N. S.	N. S.	57,000
U-Gas	0	-	-	-	984,000	600,000	10,000
Winkler	0	-	-	-	753,800	502,500	20,000

N. S. = Not specified

* Extraction steam not included twice.

(1) Besides steam plant, 41,354,260 lb/hr saturated steam from methanation waste heat boiler is superheated from 562°F to 930°F in superheater using 430 MM Btu/hr of low Btu gas and 443 MM Btu/hr of off-gas from sulfur removal (the heating value of the latter is small)

(2) Includes 1,500 kW produced in oxygen plant

(3) Plus effluent from gas turbine

(4) No steam plant required; after producing electricity, 377,000 lb/hr excess steam available at 165 psig for sales

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

If clean gas is used for fuel, the use of combined cycle operation becomes an alternative. This alternative has been used in the Lurgi, HYGAS and U-Gas designs.

2.13.2 Effluents from Power and Steam Generation

The effluents from the production of steam and electricity are given in Table 19. The major effluents are ash (when coal or char is used as fuel), flue gas, and heated cooling water when condensers on turbines are water cooled. The ash may be handled along with gasifier ash described in Section 2.11. Particulates, sulfur oxides and nitrogen oxides are the main consideration in the flue gas, although plume formation can sometimes be a problem. If coal is used, stack gas scrubbing may be necessary to reduce sulfur emissions. Less information is available on NO_x emissions; this subject was discussed in reference 5.

2.14 Cooling Water System

The cooling water system includes some of the largest streams in the plant and can represent a major source of pollution unless handled adequately. Table 20 summarizes some of the streams associated with the cooling water circuit.

The quantity of recirculated cooling water can be varied by the use of air-cooled heat exchangers where applicable. Cooling water is then used only for trim-cooling and low temperature heat transfer. The Lurgi design (5) is a good example of the use of air-fin cooling. In areas where water is scarce, the use of air cooling may be necessary. This method of cooling is not without debits, however. Added investments are necessary and electrical requirements are increased. Balanced against this is a reduction in water treatment and pumping costs. It has been estimated in one case that the decrease in thermal efficiency attendant to the use of air cooling is 1.5% (4).

The possibilities for air pollution caused by the cooling towers mainly result from leaks in equipment. Especially at high pressures, leaks in heat exchangers can result in contaminants being transferred to cooling water. These contaminants can enter the atmosphere with evaporated water or drift losses. The only technique for preventing such pollution is continuous monitoring of appropriate cooling water streams and provision of facilities for immediate removal of offending equipment from the system. This obviously requires spare equipment to allow for such removal from service.

The cooling water system is also a potential source of water pollution. Chemicals used to treat make-up cooling water may include chromium or zinc compounds, acids, chlorine and others. Some of these materials are toxic. Furthermore, because of evaporation, the concentration of dissolved solids builds up in the cooling water and must be purged. Drift loss acts as a purge and additional purge can, in some cases such as the Koppers-Totzek and CO_2 Acceptor processes, be used for ash quench with subsequent mine disposal. Waste water treatment was discussed in Section 2.12. There it was pointed out that one possibility for reducing water effluent would be to use softened water in the cooling water circuit.

Table 19

Effluents from Steam and Electricity Generation

Process	Boiler Fuel	Ash, lb/hr	Spent Limestone, lb/hr	Flue Gas, MM scfd	SO_x, lb/hr	NO_x, lb/hr	Cooling water, gpm
Koppers-Totzek	Coal	19,600	3,300	320	Less than 1.2 lb/MM Btu	N. S.	16,400
Synthane	Char	87,500	15,000	1,070	Less than 1.2 lb/MM Btu	N. S.	N. S.
Lurgi	Fuel Gas	nil	0	1,440(1)	2,004	676	0(3)
CO_2 Acceptor	(2)	(2)	(2)	(2)	(2)	(2)	N. S.
BI-GAS	Coal	12,000	44,500	625	Less than 1.2 lb/MM Btu	N. S.	N. S.
HYGAS	Low Btu gas	nil	0	905	Low	N. S.	N. S.
U-Gas	---	---	---	---	---	---	N. S.
Winkler	---	---	---	---	---	---	N. S.

N. S. = not specified

(1) Includes flue gas from steam superheater which includes incinerated gas from sulfur plant
(2) Does not include limestone regenerator
(3) Uses air-cooled condenser

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Table 20

Cooling Water Requirements and Effluents in Gasification

Process	Cooling Water Circulated	Cooling Tower Water, lb/hr			Air to Cooling Tower, MM scfd
		Blowdown	Drift Loss	Make-up	
Koppers-Totzek	73,764,000 (1)	302,000 (1)(2)	148,000	1,500,000	48,000
Synthane	50,000,000	250,000	150,000	1,700,000	20,000
Lurgi	65,000,000	105,000	130,000	1,405,000	N. S.
CO_2 Acceptor	21,450,000	N. S.	43,000	N. S.	15,000
BI-GAS	131,285,000	600,000	263,000	3,489,000	85,000
HYGAS	100,000,000	N. S.	200,000	N. S.	74,000
U-Gas	25,700,000	167,000	N. S.	891,000	16,000
Winkler	31,500,000	150,000	63,000	996,000	25,000

N.S. = Not specified

(1) Does not include cooling tower on water scrubber.

(2) Blowdown from utility cooling tower sent to scrubber.

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

2.15 Raw Water Treatment

The gasification processes reviewed require raw water for make-up purposes. The treatment of this water will depend, of course, on the nature of the water as well as its ultimate use. In general, the water is treated with lime and filtered. A sludge stream from filter back wash is an effluent that must be disposed of; this can frequently be concentrated and returned to the mine with ash. Boiler feedwater make-up can be demineralized by ion exchange. The blow-down water from the demineralization must be treated; at least it must be neutrailized. Cooling tower water make-up will usually be treated with chemicals such as chromium to prevent algae formation. Other chemicals that may be included in raw water treatment include alum, chlorine and acids.

Table 21 summarizes the raw water treatment operations in gasification.

2.16 Miscellaneous Plant Sections

This section presents a brief review of other plant installations that are not common to all processes. These are the acceptor regeneration section of the CO_2 Acceptor process and the low Btu fuel gas generation facilities in the Lurgi and HYGAS processes.

2.16.1 CO_2 Acceptor Regeneration

The dolomite Acceptor in the CO_2 Acceptor process removes sulfur and CO_2 in the reaction section. This reacted dolomite is removed from the reactor and passes to a regenerator section. The char from the reactor section is burned with air in the regenerator section to regenerate the acceptor material which is then returned to the reactor. The hot gas produced by the char combustion is used to superheat steam and its CO content is reduced by addition of more air. The hot exhaust gases then pass through an expansion turbine.

Dust, separated by cyclones from the hot gases, passes to ash desulfurization where it is reacted with carbon dioxide and water to form $CaCO_3$ and H_2S. The sulfur-containing gas from the desulfurizing unit passes to the acid gas removal unit and the ash is handled as discussed in Section 2.13. A more detailed description of the regeneration section is given in Appendix A.

The feed and effluents to the regenerator section are tabulated in Table 22.

2.16.2 Low Btu Fuel Gas Production in the Lurgi Process

In the design of the Lurgi process used in this study, fuel needs are supplied by fuel gas with 229.1 Btu/scf higher heating value. This gas is produced in a Lurgi gasifier operating at 285 psig and using air as the oxygen source. The system is very similar to the Lurgi high Btu gas train except for the use of air for gasification, a hot carbonate acid gas removal unit instead of the Rectisol unit used in the main train, and the lack of methanation which is not required. The low and high Btu gas operations cannot be combined because of contamination of the high Btu

Table 21

Raw Water Treatment in Gasification

Process	Raw Water Treated, lb/hr	Chemicals Added, lb/hr	Sludge from Water Treating, lb/hr	Water Treatment Sludge Disposal	Contaminated Water from Water Treatment, lb/hr	Contaminated Water Disposal
Koppers-Totzek	1,575,100	N. S.	N. S.	Concentrated; to mine	N. S.	Neutralized; ash slurry
Synthane	N. S.	N. S.	N. S.	N. S.	N. S.	N. S.
Lurgi	2,531,000	N. S.	90,000	Evaporation Ponds	275,000	Ash quench
CO_2 Acceptor	1,420,000	N. S.	N. S.	Concentrated; to mine	N. S.	Neutralized; ash slurry
BI-GAS	3,489,000	N. S.	N. S.	N. S.	N. S.	N. S.
HYGAS	3,536,000	N. S.	N. S.	Dispose of with char	N. S.	N. S.
U-Gas	1,245,000	N. S.	N. S.	Dispose of with ash	N. S.	N. S.
Winkler	1,197,000	N. S.	N. S.	N. S.	N. S.	N. S.

N.S. = Not specified

Values shown in this table depend on the original bases chosen; plant sizes as well as other factors differ and direct comparison of the values is difficult. The process reports in references 3-10 should be consulted to determine each design basis, information sources, and qualifications (see Section 1.5) if individual numbers are to be utilized.

Table 22

Feed and Effluents of the CO_2 Acceptor Regeneration Section

Material	Quantities
Inputs	
Char, lb/hr	496,810
Reacted acceptor, lb/hr	7,977,000
Air, scfh	44,500,000
Dolomite makeup, lb/hr	254,454
CO_2, scfh	600,000
Water, lb/hr	15,800
Outputs	
Regenerated acceptor, lb/hr	7,164,000
Carbonated ash slurry (50% water), lb/hr	466,000
Acid gas, scfh	450,000
Flue gas, scfh	57,300,000

gas with nitrogen, but waste liquids and liquid by-products are combined. Thus, the coal tar, ash, gas liquor and acid gas produced in the low Btu complex is combined with those materials from the high Btu system for ultimate disposal. A major portion of the product gas is heated, by burning a portion of the product, and expanded through an expander turbine to provide part of the air compression energy requirements of the low Btu complex.

The disposition of the low Btu fuel gas is shown in Table 23. The inputs and outputs of the low Btu plant are given in Table 24.

2.16.3 Low Btu Fuel Gas Production in the HYGAS Process

Fuel for coal drying and for the utility furnace in the HYGAS process is provided as low Btu fuel gas from a U-Gas gasifier. The U-Gas process was the object of a special process report (9) and has been described in that report. Therefore, no special description will be given here. The major inputs and outputs of the production of low Btu gas for use in the HYGAS process is given in Table 25.

Table 23

Disposition of the Low Btu Fuel Gas in the Lurgi Process

Gas Disposition	Volume, MM scfd
Burned internally to heat gas for expander turbine on air compressor	19.5
To gas turbines in oxygen plant	82.1
To steam super heater	44.9
To power boilers	180.2

Table 24

Inputs and Outputs of the Lurgi Low Btu Gasification System

Material	Quantities
Inputs	
Steam (to gasifier), lb/hr	258,060
Boiler feed water, lb/hr	54,440
Air (including 3,679 lb/hr water), MM scfd (dry)	184.0
Coal, lb/hr	440,000
Outputs	
Low Btu product gas (HHV, 229.8 Btu/scf), MM scfd (dry)	307.2
Ash, lb/hr	80,224
Coal tar, lb/hr	21,846
Boiler blowdown, lb/hr	560
Gas liquor, lb/hr	213,165
Acid gas, MM scfd (dry)	40.3
Flue gas (low sulfur)	not specified
Utility Requirements	
Steam, lb/hr	166,600
Electricity, kW	4,230
Cooling water, gpm	2,000

Table 25

Major Inputs and Outputs of the Low Btu Gasification Plant Used in the HYGAS Process

Material	Quantity
Inputs	
Coal, lb/hr	273,800
Air (to gasifier), MM scfd (dry)	309.3
Air (for sulfur acceptor regeneration, MM scfd	17.2
Steam, lb/hr	213,300
Quench water, lb/hr	333,300
Make-up chemicals to remove sulfur	not specified
Outputs	
Fuel gas (33 MM scfd for coal preparation), MM scfd	482
Char slurry, lb/hr (dry)	37,500
Dust	not specified
SO_2 stream (to sulfur recovery), lb/hr	66,400