Gasification of COED Chars in a Koppers–Totzek Gasifier

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Prepared by
Koppers Company, Inc.
Pittsburgh, Pennsylvania

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Gasification of COED Chars in a Koppers-Totzek Gasifier

AF-615
Research Project 264-1

Final Report, July 1978

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ABSTRACT

In December 1974, EPRI entered into a contract with the FMC Corporation to demonstrate COED char gasification in a commercial Kopper-Totzek gasifier. (The Char-Oil-Energy Development (COED) process, which was developed by FMC Corporation under funding from the Office of Coal Research between 1962 and 1975, is a fluidized bed coal pyrolysis process which produces gas, oil and char from the feed coal.) The Energy Research and Development Administration (ERDA), the successor government agency to the Office of Coal Research (OCR), agreed to supply about 900 tons of char derived from test runs with West Kentucky and Pittsburgh coals on the 36 ton per day COED pilot plant at Princeton, New Jersey. A subcontract for the test program was made between FMC and the Koppers Company of Pittsburgh. The Koppers Company contracted with Krupp-Koppers of Essen, Germany, for the actual operation of the tests. Krupp-Koppers entered into agreement with ENFERSA, the Spanish company, for the use of the plant facility selected for the test.

The chars were shipped to Spain in early 1975 and the gasification tests conducted at the ENFERSA plant in Puentes de Garcia Rodriguez, Spain in August 1975. The results of these tests on the two chars demonstrated that COED char could be gasified in the Koppers-Totzek gasifiers. The useful gas yield was about 45 MSCF of carbon monoxide plus hydrogen gas per ton of char. The carbon conversion of the char to gas was 85 to 90 per cent. Some problems were encountered with the refractory lining on the plant; however, technology is claimed to be available to enable proper refractory selection for commercial life. On the basis of these results, confidence exists for the design of larger (30 tons per hour), more modern Koppers-Totzek gasifiers based on the gasification of COED char.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

This final report describes the gasification tests of COED char in a commercial oxygen blown Koppers Totzek gasifier. The Char-Oil-Energy Development (COED) process which was sponsored by the Office of Coal Research (OCR) between 1962 and 1975 is a fluid bed coal pyrolysis process which produces gas, oil and char from the feed coal. It was originally thought that the char would be a suitable power plant boiler fuel; however, the char was found to have about the same sulfur content as the parent coal, and a power plant designed for char from a high sulfur coal would, therefore, require flue gas desulfurization equipment.

An alternative way of utilizing the char is by gasification to a fuel gas which when desulfurized would be a suitable fuel for power plant boilers or combined cycles. In order to provide information which might assist in the rapid commercialization of the COED technology, it was considered desirable to obtain data on the gasification of char in an existing commercial gasifier.

At the time of initiation of this work (1974), there were two proposals before the Office of Coal Research (OCR) in response to their Request for Proposals (RFP) for a Clean Boiler Fuels Demonstration plant. One of these was based on the COED process and another on Union Carbide's COALCON technology. Both were fluid pyrolysis (or carbonization) processes and required gasification of the char to complete conversion to clean fuels.

PROJECT OBJECTIVES

The purpose of this study was to determine whether under gasification conditions the COED chars were sufficiently reactive to give high carbon conversions at reasonable steam and oxygen consumptions in a commercial gasifier.

A related objective of the tests was to provide data for the scale up and design of larger Koppers-Totzek gasifiers in the event that it was decided to proceed with a demonstration or commercial plant based on the COED process.

CONCLUSIONS AND RECOMMENDATIONS

These tests demonstrated that carbon conversions of 85 to 90 percent could be achieved. No large differences in performance between the West Kentucky and Pittsburgh chars were noted. The data should be sufficient to enable the design of larger Koppers Totzek gasifiers based on this COED char feedstock.

Prior to these tests, (and to other tests on the gasification of Coal Liquefaction Residues reported in EPRI Report AF-233) concern was expressed in several quarters that the residues from partial coal conversion processes, such as pyrolysis or coal liquefaction, might be too inactive to enable their conversion to gas in gasification processes. The results of the tests reported here and in AF-233 show that such residues can be converted to synthesis gas at reasonable oxygen consumptions and carbon conversions.

Although the development of pyrolysis processes is still being pursued, they do not appear to be as attractive for most potential applications in the power industry as complete conversion processes. The main reasons are:

- added complexity of processing
- simultaneous marketing of a variety of fuels is required
- pyrolysis liquids need intensive secondary hydrotreatment for storage stability, desulfurization and denitrogenation
- even with gasification of the char, the cost of gas is likely to be greater than direct coal gasification and the processes will be more difficult to integrate with combined cycles.

FOREWORD

This report on the gasification of Char Oil Energy Development (COED) char in a Koppers-Totzek gasifier is issued as partial fulfillment of the FMC/Electric Power Research Institute (EPRI) Agreement RP264-1. The gasification of char was conducted in a Koppers-Totzek unit located at a fertilizer plant in Puentes de Garcia Rodriquez, Spain. This report consists of a summary report of activities and results of the tests, an outline summary of the detailed report of the tests, and the detailed report containing the technical information from the tests prepared by Koppers Company, Inc. U.S.A. and Krupp-Koppers, GMBH, Germany.

The FMC Corporation thanks EPRI for sponsoring the tests, and the Energy Research and Development Administration for sponsoring the COED pilot plant program needed to generate the char for the test. The cooperation of the Koppers Company, Inc., Krupp-Koppers, and Empresa Nacional Fertilizantes S.A. (ENFERSA), the Spanish fertilizer company, made this test a successful demonstration of using COED char in a Koppers-Totzek gasifier.

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PREFACE

The commercial gasification of COED char was sponsored by the Electric Power Research Institute under contract with the FMC Corporation. The char for the demonstration was provided through the Energy Research and Development Administration from the FMC Pilot Plant at Princeton, New Jersey. Koppers Company, Inc., of Pittsburgh organized the char transfer and, in cooperation with Krupp-Koppers of Germany, arranged with Empresa Nacional de Fertilizantes of Spain for the demonstration at the commercial gasification plant of Puentes de Garcia Rodriguez. Following final agreements between all companies, modifications for the tests commenced in early 1975 and the actual gasification demonstration was completed in August 1975.

The following report covers the analytical demonstration data and computer prediction for larger commercial gasifiers using COED char as a feedstock. In conjunction with the char, a demonstration was undertaken for the gasification of petroleum cokes for eight major oil companies.

Acknowledgment is made to EPRI, ENFERSA, ERDA, FMC Corporation, KRUPP-KOPPERS, the oil companies, and the numerous people involved who contributed to the successful completion of the demonstration.

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Prepared by

FMC CORPORATION

SUMMARY

INTRODUCTION

There is a need today for more clean energy than ever before. Importation of oil now accounts for over 40 percent of consumption in the United States. Natural gas is becoming scarce and the price increasing. Coal is available throughout the United States and can be a source of the clean fuel required to again make the country self-sufficient in oil and gas.

A major impact can be made on oil and gas consumption by substituting clean fuel derived from coal in power plants. This can be done by converting high-sulfur coals into fuel gas and/or oil of low sulfur content to meet environmental requirements. The Char-Oil-Energy Development (COED) process is one method of accomplishing this conversion of coal to clean fuels.

The COED process was operated at 36 tons per day of coal in a pilot plant for a period of about four years. During that time, over 20 thousand tons of coal were processed in continuous runs, some of which were 30 days or more, to produce a low sulfur, petroleum-type oil, a fuel gas with a heating value of 500 Btu's per cubic foot, and a solid product called char. Uses of the products were demonstrated by testing the oil as a fuel to a Navy destroyer and in other types of engines. It was found acceptable in most tests, or could be made acceptable through additional hydrogen treating.

The char was tested in a commercial size power plant boiler and found to be suitable fuel. However, the char does have about the same sulfur content as the parent coal. A power plant designed for char from a high sulfur coal would also need environmental equipment for removing sulfur from the stack gas.

An alternate way of utilizing the char is to convert it into a fuel gas for subsequent mixing with the 500 Btu per cubic foot gas produced from the COED process. This combined gas stream would have a heating value of over 300

Btu's per cubic foot and be cleaned of sulfur for use as a fuel gas to a power plant.

There are many second generation gasification processes under development for converting coal or chars into a fuel gas which can subsequently be converted into a high Btu gas. It was thought, however, that if the COED process were to proceed rapidly into commercialization, then the gasification of char should be demonstrated in existing commercial gasifiers. With information from a commercial size gasification test with char, the COED process would be ready for scale-up to a production size plant with a proven option to generate clean fuel gas from the char.

A review of various gasifiers that had been used at the commercial scale concluded that the Koppers-Totzek gasifier would be suitable for gasifying char. In addition, Koppers-Totzek gasifiers were located in several places in Europe and other parts of the world and one could be available for conducting a commercial scale test. Contacts with Koppers Company, Inc. in Pittsburgh confirmed this.

The Electric Power Research Institute (EPRI) accepted and sponsored a proposal by the FMC Corporation to demonstrate char gasification in a commercial Koppers-Totzek gasifier. The Energy Research and Development Administration (ERDA) agreed to make 900 tons of char available for the test through the operation of the COED pilot plant. A subcontract for the test program was agreed to between FMC Corporation and the Koppers Company in November, 1974. The Koppers Company had an agreement with Krupp-Koppers of Essen, Germany, for the actual operation of the tests. Krupp-Koppers had an agreement with ENFERSA, the Spanish company managing the production facility selected for the test, for use of the plant.

COED Process

A schematic diagram of the basic COED process is shown in Figure S-1. In the COED process, coal is crushed, dried and heated to successively higher temperatures in a series of fluidized-bed reactors. In each fluidized bed, a fraction of the volatile matter of the coal is released. The temperature of each bed is selected to be just short of the maximum temperature to which the coal can be heated without agglomerating and defluidizing the bed. Typically, four stages operating at 500°, 800°, 1000°, and 1500°F are used. The number of stages and operating temperatures vary with the agglomerating properties of the coal. Heat for the

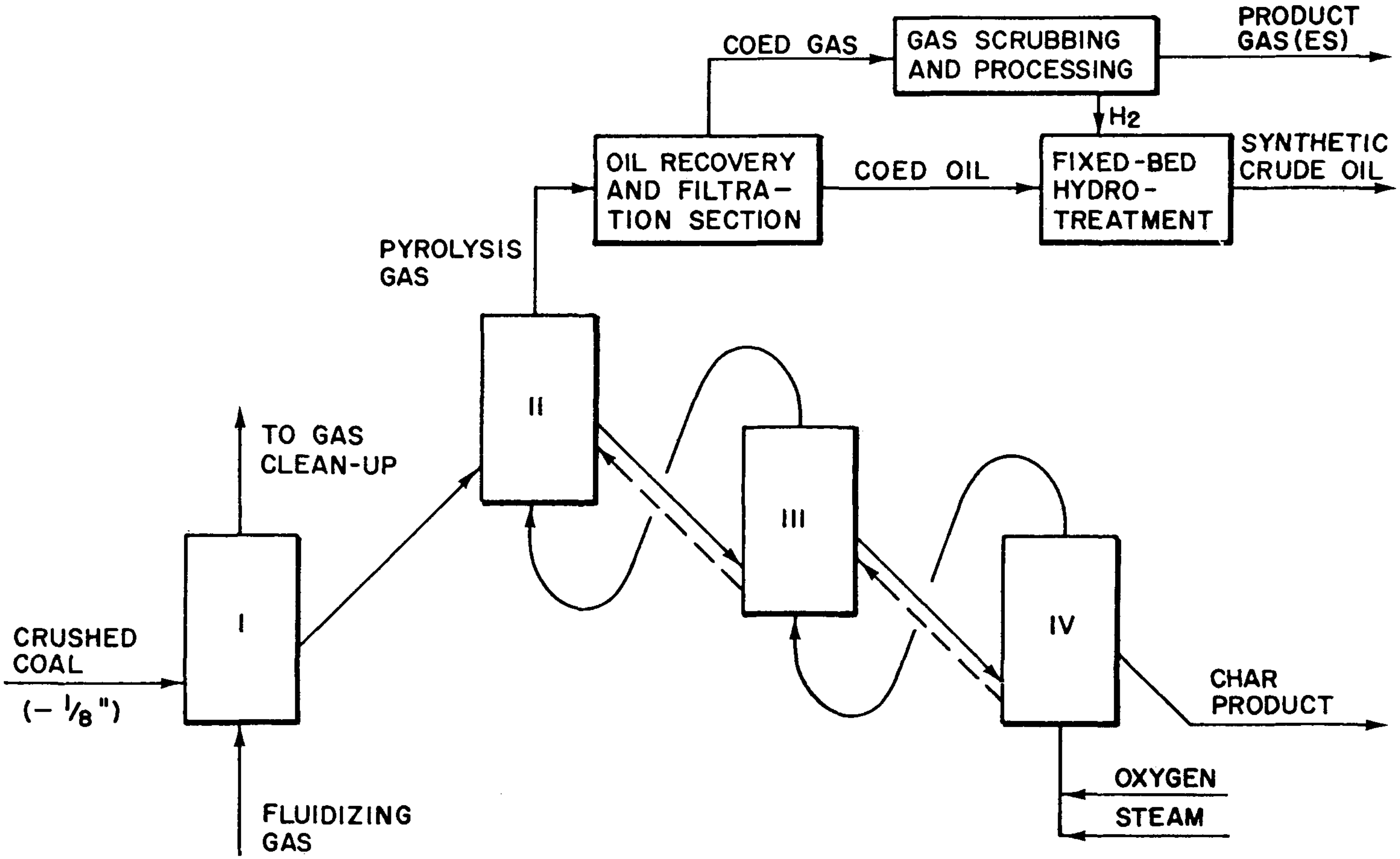

Figure S-1. Schematic Diagram of COED Coal Pyrolysis process is generated by burning char in the fourth stage and then using hot gases and the hot char from the fourth stage to heat the other vessels through recycle.

The volatile products released from the coal in the fluidized-bed reactors are passed to a product-recovery system for recovery of the oil and cooling of the gases. The condensation of the coal oil is accomplished by direct contact with a water-rich stream. The condensed oil-water mixture is decanted and the oil phase dehydrated in a steam-heated separator. The dry oil from pyrolysis is filtered to remove the solids which escape recovery by the cyclones of the fluidized-bed reactors. The solids-free oil is pumped up to pressure and mixed with hydrogen for hydrotreating in a fixed-bed catalytic reactor operating at about 750°F and 2500 psi. Hydrotreating removes sulfur, nitrogen and oxygen from the oil and produces a 25° API synthetic crude oil. The gas from the process can be sold as fuel gas or converted by application of additional technology to pipeline gas or hydrogen. The residual char can be used as fuel to a power plant or gasified to produce a fuel gas, as discussed previously. Selection of the final gas product and end-use of the char depend on the coal used in the process and the marketability of the products in the plant site area. A final report on the COED process has been published.*

Koppers-Totzek Process

The Koppers-Totzek process lends itself to the gasification of most solid and liquid fuels. Since oxygen is used to gasify the input feed, high temperatures, above 2700°F, are reached in the burning chamber. This ensures almost complete combustion of the incoming feed to a gaseous product. When the Koppers-Totzek gasifier operates with a solid feed material, the feed is crushed to about 70 percent minus 90 microns. The gasifier is a horizontal cylindrical type refractory lined vessel with two or more feed heads in each end of the cylinder. The pulverized feed is fed into the gasifier through the feed heads with the proper amount of oxygen and, at times, steam. The control of the oxygen, steam, and carbonaceous feed will optimize the quantity and quality of the gas produced.

The raw gas exits from the top of the gasifier from which heat is removed through a waste heat boiler and thence to a cleaning system for removal of solids and sulfur-containing gases. The clean gas is then suitable for direct use as a fuel gas for energy production, and with CO_2 removal can be used as a synthesis gas for ammonia production or conversion to a high Btu gas.

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* Char Oil Energy Development, Final Report for the Period August 18, 1971 - June 30, 1975. FMC Corporation, Princeton, New Jersey.

Slag formed from the ash is removed from the bottom of the Koppers-Totzek unit through a water cooled chamber for disposal.

Further details of the Koppers-Totzek process are presented in the Appendix which contains the full report of the char gasification tests. These tests were conducted with chars derived from two different coals, a Western Kentucky coal, and a Pittsburgh seam coal. The tests were performed at Puentes, Spain, during the summer of 1975.

PREPARATION OF COED CHAR

Special runs were conducted in the COED pilot plant to produce char from a Western Kentucky coal for the gasification tests. Char was also produced from a Pittsburgh seam coal for the gasification tests as part of the normal evaluation of a new coal in the COED process.

Western Kentucky coal is a high volatile B bituminous coal. The sample (600 tons) was from the Colonial Mine of the Pittsburgh and Midway (P&M) Coal Company. The Pittsburgh No. 8 seam coal is a high volatile A bituminous coal. About 800 tons were supplied by the American Electric Power Corporation from the Ireland Mine of the Consolidation Coal Company.

The analyses of the coals and the chars produced from these coals is shown in Table S-1. The char is essentially devolatilized coal since it has been heated to 1500°F. Char contains more carbon and ash, about the same percentage of nitrogen and sulfur, and less hydrogen and oxygen than the parent coal. Heating values for chars are usually about 5 to 10 percent less than the dry coal because of the higher ash content.

The yield of char was 0.65 and 0.63 ton per ton of dry coal, respectively, for the Western Kentucky and Pittsburgh seam coals. Additional details on runs producing the char for test purposes are in the COED Final Report.*

The char was packaged in half-ton boxes with plastic liners. The boxes were delivered by truck to the Port of Newark for loading on shipment for delivery to the Port of El Ferrol, Spain. The boxes were stored in a warehouse at El Ferrol, and delivered by truck to the plant at Puentes, Spain.

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* Char Oil Energy Development, Final Report for the Period August 18, 1971 - June 30, 1975. FMC Corporation, Princeton, New Jersey.

Table S-1

SUMMARY OF ANALYSIS OF CHARS GASIFIED AND PARENT COALS

	Western Kentucky		Pittsburgh Seam	
	Coal	Char	Coal	Char
Moisture, wt. %	7.8	--	2.5	—
Proximate Analysis, wt. %				
Dry Basis				
Volatile Matter	35.2	4.0	38.3	7.0
Fixed Carbon	55.1	80.5	52.7	79.8
Ash	9.7	15.5	9.0	13.2
Ultimate Analysis, wt. %				
Carbon	69.0	75.6	73.0	76.4
Hydrogen	5.1	1.7	5.2	1.7
Nitrogen	1.5	1.6	1.3	1.4
Sulfur	3.3	3.2	4.3	4.0
Oxygen	11.4	2.4	7.2	3.3
Ash	9.7	15.5	9.0	13.2
Higher Heating Value				
Btu/lb. Dry Basis	12,700	11,900	13,500	12,200
Bulk Density, lb./ft.^{3}	46	38	49	38

TEST OBJECTIVES

The objectives of the Koppers-Totzek gasification test were to determine the qualities and quantities of gas that could be generated from char in the Koppers-Totzek gasifier and to determine if there were any mechanical, process, or other unusual properties in handling char as a feed to this system. To accomplish this, several runs were planned as process variable studies within the restriction placed by the production gasifiers. A long run was also made to look for any unusual occurrences while using char as the feed.

SELECTION OF SITE AT PUENTES, SPAIN

There are no Koppers-Totzek gasifiers in the United States. Therefore, selection of the site had to be overseas. The closest Koppers-Totzek gasifier was in Spain and, fortunately, was of a small enough size that the quantities of char available for test purposes could be run with significant data acquisition for evaluation. The Spanish company, ENFERSA, produces fertilizer at the plant and agreed to make a gasifier available for the test at Puentes.

The Puentes plant has adequate facilities for handling and storing another feed raw material to the plant. Also, Puentes was only 25 miles from the Port of El Ferrol, which made for reasonable storage, handling and transport of char.

CONCLUSIONS

The large-scale tests carried out at the plant at Puentes, Spain, demonstrated that COED char could be gasified with good results in the Koppers-Totzek gasifiers. The gas produced during these experiments was fed directly to the gas main at the Puentes plant and used to produce ammonia for their fertilizer operation. The useful gas yield was about 45 MSCF of CO plus H_2 gas per ton of char. The carbon conversion of the char to gas was 85 to 90 percent. Confidence exists for scale-up to the larger (30 tons per hour), more modern gasifiers.

The char is somewhat more abrasive than lignite or coal. The grinding of this char to 70 percent less than 90 microns was of some initial concern. However, as shown in the grinding operation at the Puentes plant, the char could be ground in their existing ball mill to meet the grind size without any operational or mechanical problems. Some of the auxiliary equipment, such as cyclones and bag houses, were not as efficient with char, since the lighter density material tended to overload the system. The grinding system operated at a reduced capacity when grinding char compared with that for lignite.

The refractory lining used in the gasifier was especially selected for operation with petroleum cokes which were also being tested by the Koppers Company. The lining chosen was not completely successful when operating with the COED chars since the ash content in the char partially eroded the lining. The proprietary technology, however, is available for the selection and installation of the proper lining for a gasifier that would use char as its feed.

TEST OPERATIONS

Modifications to Puentes Plant

The Puentes Plant is a production plant that was scheduled to shut down in 1975. With the advent of the energy crisis and higher prices for oil, the plant again became economic in using lignite as the feed to generate the synthesis gas for ammonia production. The plant needed additional test equipment to acquire the necessary data for evaluating results from a program using a new feed. Some of the more important modifications made to the plant before starting the tests are listed below:

- Increased nitrogen capacity was added to the plant for pneumatic conveying of solids.
- A by-pass line and stack were installed so that the raw gas from gasification could be flared during start-up.
- Much instrumentation was added to ensure that all data for test evaluation was collected.
- Various samples were added to ensure that quality samples were taken for analysis on which to base heat and material balances and calculate gasification performance. The sampling points and devices are discussed in detail in the attached report by Krupp-Koppers and the Koppers Company.
- The gasifier was relined with new refractory for the specific tests.
- Miscellaneous equipment such as oxygen/steam mixers and an oxygen preheater were added.

It took approximately six months to make the plant ready for the first start-up test. The ENFERSA plant is a production facility, but the plant operating personnel were very cooperative and assisted as much as possible in making the demonstration runs on the COED chars.

Test Program

The test program plan was as follows:

- Operate the gasifier with two chars, one produced from a Western Kentucky bituminous coal, and the other from a Pittsburgh seam coal.
- Explore process variables such as oxygen feed rate, steam feed rate, and temperature, and their effect on production rate and quality of the product gas.
- Operate the gasifier at its maximum char feed rate for a sufficient length of time to ensure that it reached a steady state and operated without problems.
- Utilize the product gas in the ENFERSA plant system for the production of ammonia.

Prestart-up tests such as determining how well the char grinding system worked and calibrating various instruments and the char feeding system to the gasifier were performed. The best conditions were found for grinding char to meet the size consist of about 70 percent less than 90 microns. Because of later problems in feeding char to the gasifier, some water was sprayed onto the char prior to pulverization to ensure proper feeding to the gasifier.

When cold tests were performed with the screw conveyors used to feed char to the gasifiers, it became evident that char fed differently than lignite mainly because of its dryness and the lighter density of the char. Thus, feed rates were not directly measurable, but were later calculated from carbon balances around the total system.

During any test run, the most readily-available parameters to monitor were the gas temperature and the CO_2 level of the gas. Operations at predetermined temperature and feed rates of oxygen and steam were maintained to give a steady state gas production rate and gas quality.

IMPORTANT RESULTS

Results on the gasification of chars from Western Kentucky and Pittsburgh seam coals were similar since there were no significant differences in the chemical and physical properties of the chars. Five test runs were performed with each char. Data were collected when the gasifier reached steady state conditions as defined by keeping a constant outlet gas temperature and constant CO_2 content in the product gas. The data collection period was 1.5 to 2.5 hours of each test run, although the run itself lasted 3 to 10 hours. One long-term test of over 30 hours was carried out with char from Western Kentucky coal. Longer tests were not possible because of the complexities of scheduling the feed-grinding facilities within a production plant to stockpile the pulverized char for a longer test. Approximately 200 tons of each char was gasified during the five test runs, or a total of 400 tons of char gasified during the demonstration.

The processing parameters studied for the demonstration tests were:

Chars - Two chars, one from Western Kentucky coal, and one from a Pittsburgh seam coal

Oxygen Input - 0.8 to 1.1 lb. O_2 per lb. of char

Steam Input - 0.03 to 0.3 lb. steam per lb. of char

Char Feed Rate - Normally kept constant at highest attainable value, but did vary some because of the wetness of the char; feed rate varied from 2.7 to 3.6 short tons of char per hour

Pressure - Close to atmospheric pressure for all runs

Under these ranges of operating parameters, the test results of importance were:

Gasifier Temperature - 2260 to 2900°F

Useful Gas ($CO+H_2$) Production - 40 to 45 MSCF per short ton of char

Carbon Gasification Efficiency - 80 to 95%

Thermal Gasification Efficiency - 55 to 65%

The material balances for the test runs closed between 96 and 100 percent. Heat loss from the gasifier itself normally was between 1 to 3 percent. This is expected to be lower for the larger Koppers-Totzek gasifiers being built today. The largest thermal efficiency loss is to the generation of CO_2.

Some other operating characteristics were:

- Lower throughputs were experienced with char than with the normal lignite feed. Major reason for this was the lower density of the char and its affect on volume feeders and solids collection equipment such as cyclones. Normal rates with lignite were 4 to 5 tons per hour compared with 2.7 to 3.6 tons per hour with char. However, the higher carbon content of the char and, thus, higher gas yield per ton would have limited its throughput to less than that for the high ash (at times 45 percent) lignite because of limitations of the gas-handling equipment.
- The plant oxygen supply varied somewhat in its purity, 89 to 95 percent oxygen, but normally was about 94 percent oxygen and 6 percent nitrogen.

- The CO content of the raw gas varied between 56 and 71 percent, H_2 14 to 18 percent, and CO_2 5 to 22 percent, depending on the oxygen and steam inputs to each run. In general:

  --Carbon dioxide percentage increased with increases in oxygen input with a corresponding decrease in CO. Hydrogen content remained constant.

  --Proper combination of input ratios of oxygen to char and steam to char can be selected to optimize the quantity of useful gas ($CO+H_2$) produced.

- About 80 percent of the ash contained in the char left the gasifier in the raw gas stream and was subsequently scrubbed out. The remaining 20 percent left as slag through the ash-quenching system.

- The ball mill designed to grind lignite also performed well in grinding char to 70 percent less than 90 microns. The char was a dry material and required some spraying with water, so that it would compress and form a seal in the screw feeders to the gasifier. The screw feeding system, however, can be designed to operate with the dry char in new commercially sized gasifiers.

- The refractory lining used in the gasifier was selected to operate with petroleum cokes which were also tested as part of another program sponsored by eight oil companies. The refractory was eroded in places by the slag from the char ash. However, proprietary knowledge exists for selecting the proper refractory material suitable for operation with char.

The demonstration proved that char can be used as a feed to a Koppers-Totzek gasifier. It was done under adverse conditions in a plant not designed for handling char, but which was readily adapted with some limitations. Design knowledge exists in the Krupp-Koppers Company for using the results from this demonstration to design a larger, commercially sized plant specifically using char as fuel.

An impressive feature of the Koppers-Totzek gasifiers witnessed during the tests was the rapidity with which the units could be started up and brought to steady state operating conditions, and the quickness with which they could be shut down and then restarted. This flexibility could be an important factor for situations where a varying load of fuel gas is needed.

During all tests with char, the product gas was fed into the synthesis gas mains of ENFERSA plant and then used for ammonia production. Similarly, the gas could be used in future plants for conversion to a high Btu gas or burned directly as a fuel gas.

COMMERCIAL PLANT PROJECTIONS

The commercialization of the COED process could require plants with the capacity of 10,000 to 20,000 tons of coal per day (6,000 to 12,000 tons of char per day). To anticipate performance of larger Koppers-Totzek gasifiers, calculations were made using data from the char tests, together with some modifications to reflect operation of the more modern gasifiers. A comparison of results is shown below:

	Puentes Char Test	Predicted for Large Commercial Gasifier
Operating Temperature, °F	2890	2730
Carbon Conversion, %	93.7	92
Oxygen Purity, %	93.4	98
Char Feed, tons/hr.	2.7	32.4
Useful Gas ($CO+H_2$) Make, SCF/ton of char	45,000	47,800
Oxygen, ton/ton of char	0.99	0.9
Steam, ton/ton of char	0.09	0.3
Gas Analysis, Vo. %		
CO_2	12.1	11.7
CO	66.7	62.9
H_2	13.8	22.7
H_2S and COS	1.0	1.2
N_2	6.4	1.5
Gas/Char Thermal Efficiency, %	63	65

The H_2S, COS, and CO_2 are all readily removable to yield a gas for use as clean energy fuel or for conversion to high Btu gas. Prior to commercialization, no further design information would be needed. Chars from other coals are similar enough to those tested to forecast steam and oxygen requirements and gas composition. Only the slag characteristics of the ash in a particular char need to be determined to guide the proper selection of the refractory lining for the gasifier.

Preliminary design and capital cost estimates of a Koppers-Totzek plant to gasify 6,000 to 12,000 tons per day of char from a COED plant were part of the original scope of work. However, this was later deleted from the program because of project cost considerations. Previous work* on economic studies of gasifying char as part of an overall COED processing plant complex showed that the selling price of fuel gas from such a facility was $1.50 to $2.00 per MMBtu based on January 1975 dollars for capital and operating costs. The range of selling price covers

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* Char Oil Energy Development, Final Report for the Period August 18, 1971 - June 30, 1975. FMC Corporation, Princeton, New Jersey.

sensitivities to the price of coal, yield and price of oil from the COED process and capital costs. In December, 1976 dollars, this range of selling price is probably $2.50 to $3.00 per MMBtu. Sufficient data are now available from the final COED report* and the attached report on char gasification to make a more definitive capital estimate and economic evaluation for clean fuel from the combination of the COED and Koppers-Totzek processes.

OUTLINE SUMMARY OF THE REPORT ON THE COMMERCIAL GASIFICATION OF COED CHAR IN THE KOPPERS-TOTZEK GASIFIER AT PUENTES, SPAIN, DURING THE SUMMER OF 1975 BY KRUPP-KOPPERS AND THE KOPPERS COMPANY, INC.

This synopsis summarizes each section of the attached report by the Koppers companies on the gasification test of COED char. The reader may refer to the full report for the details of items of interest.

Introduction (Section 1)

The Koppers-Totzek gasifier is an established commercial process for the gasification of solid and liquid fuels. Seventeen plants have been built operating on feed stocks such as lignite, coal and petroleum residues for generating a clean synthesis gas suitable for ammonia production or for use as fuel. Gasification plants using Koppers-Totzek gasifiers are located in Europe, Middle East, Africa, India and Japan. The largest two-headed gasifier (two feed heads) has a capacity of about 13,000 standard cubic feet per minute of product gas. Four-headed gasifiers (four feed heads) are being built in India with a capacity of about 20,000 standard cubic feet per minute each. Larger gasifiers are projected as the need arises, particularly for converting solid fuels such as coal into a clean fuel gas for power generation or for conversion into a high Btu gas.

The COED process was successfully demonstrated in a 36 ton per day pilot plant by the FMC Corporation under sponsorship of the Energy Research and Development Administration. Many tests were performed on the products from this plant. The hydrotreated liquid oil product was found acceptable as a feed to a petroleum refinery or through some further distillation treatment as a fuel to marine power plants on board ships. About 60 percent of the coal used as feed is converted to a finely divided char with low volatile content and a high heating value. This char has been burned in a commercial power plant boiler. However, this char contains approximately the same percentage of sulfur as the feed coal. For high sulfur coals, it could not be used as a fuel to a power plant without having the same gas clean-up systems as required for burning the parent coal. An alternate

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* Char Oil Energy Development, Final Report for the Period August 18, 1971 - June 30, 1975. FMC Corporation, Princeton, New Jersey.

use of the char is to convert it to a fuel gas in a gasifier, such as the Koppers-Totzek, and remove the sulfur from the gas to make it an acceptable fuel.

With the support of the major U. S. utility companies as represented by the Electric Power Research Institute (EPRI), a program was developed with the Koppers Company of Pittsburgh to perform the gasification demonstration at a commercial plant. Success of such a test would be the quickest way of making the technology of the COED process supplemented with the Koppers-Totzek gasifier system available for future use to meet our clean energy requirements.

The char was prepared at the COED pilot plant at Princeton, New Jersey, and shipped to the site of the Koppers-Totzek gasification plant in Puentes de Garcia Rodriquez, Spain. Gasifiers at this plant are used to convert the local lignite feed to a synthesis gas suitable for production of ammonia. Agreements between EPRI, the FMC Corporation, Koppers Company, Inc. U.S.A., Krupp-Koppers of Germany, and the Spanish operating company, ENFERSA, were negotiated and agreed upon in November of 1974. About 900 tons of char were shipped to Spain in December, 1974, and arrived there in January, 1975. Tests were conducted in Spain as there are no Koppers-Totzek gasifiers in the U. S. The gasifiers at the Puentes plant were the closest available and of a small enough size for meaningful test runs for the quantities of char generated from the COED pilot plant.

This project involves multiple companies and organizations and a gasification plant that had been scheduled to shut down until the oil crisis occurred. Many problems in scheduling, organizing, and modifying the plant occurred. These were all resolved through cooperation with the American, German and Spanish companies. The tests started in July of 1975. Some key areas solved prior to the tests were:

- Scheduling of the test runs within the framework of a production plant.
- Adapting the lignite crushing equipment to grind the char to 70 percent minus 90 microns. The density of the char is lighter than the normal lignite feed, and this meant that the pulverized feed recovery system was less efficient for the lighter char. Also, the char is dry and this gave some feed problems to the gasifier later on. These problems were solved by operating the grinding equipment at a lower capacity so that the fines recovery system was not overloaded, and by spraying water on the char feed to the pulverizer at a controlled rate.
- Special modifications were made to the plant, such as instrumentation and sample collection, to ensure that all the data necessary for evaluating the test would be obtained, by-pass lines, and added nitrogen capacity for pneumatic transport of char.

Description of the COED Chars. Two chars were used for the test runs in the Koppers-Totzek gasifiers. These chars were derived from bituminous coals from Western Kentucky and from the Pittsburgh seams. The chars were similar in composition, and their analyses are presented in this section.

Description of the Koppers-Totzek Gasification Plant at Puentes, Spain (Section 2)

The Koppers-Totzek gasification plant at Puentes was designed for lignite feed and was built in 1959 by Heinrich Koppers Company (now known as Krupp-Koppers). The plant contains three gasifiers, each capable of producing 212,000 SCF (6,000 Nm^3) per hour of raw product gas. A description of the plant with flow sheets is contained in this section.

Char Pulverization. The char was pulverized in the existing ball mill equipment. Since the char was dry, it was conveyed throughout the system with nitrogen rather than with hot combustion gases. The pulverized char from the ball mill was conveyed into bunkers for storage and subsequent feeding to the gasifiers. There was a recycle system for the oversized material, and a system for spraying water onto the char feed to pulverization to give it a moisture content of about one percent. The addition of water to the char was necessary so that the char would compact in the screw feeders used to feed the gasifiers. This slight compaction in the screw feeders was required to achieve a seal between the gasifier and the char feed bins.

The existing crushing system used for lignite was acceptable for char, except that feed rates had to be reduced by about one-half. The reason for this was the lower density of char compared with lignite. The cyclonic and other separation equipment used to recover the pulverized product were less efficient because of this lower density. Operating at the lower feed rates caused no difficulty. Proper recovery equipment could easily be designed for the pulverization of char. Another nitrogen compressor was added for transporting char since the system at Puentes did not have sufficient capacity to transport both the lignite and char at the same time.

Gasification Plant. The Koppers-Totzek gasifiers at Puentes consist of a horizontal cylindrical, water-cooled shell, vessel. On each tapered cylindrical end, burner heads are mounted with two burners for each head. Through these burner heads, an accurately proportioned, homogeneous blend of pulverized char, oxygen, and steam is introduced into the chamber of the gasifier, where ignition under partial oxidation of the feed stock takes place at high temperatures and at about atmospheric pressure. The high reaction temperatures produce a rapid reaction between the char, oxygen and steam. The temperature is sufficiently high to melt any ash contained in the feed. Ash is deposited on the walls of the gasifier and in fluid form flows down through a water seal into a water quench tank. Since some fly ash also leaves with the raw gas, the raw gas from the gasifier is steam quenched to below 1800°F to solidify the ash, and thus to prevent deposit on the waste heat boiler tubes. The raw gas from the water heat boiler is scrubbed to remove the fly ash, and then scrubbed again to remove sulfur containing compounds before storage in the gas holder. At the Puentes plant, the synthesis gas is used for ammonia production.

Plant Modifications. Some modifications made to the existing Puentes plant for the demonstration tests were:

- Oxygen/steam mixture was added, along with an oxygen heater to ensure proper addition of these raw materials to the gasifier.
- A flare stack was added for start-up before feeding the usable gas into the plant's gas main.
- Various instrumentation and sampling systems were added to acquire all the information for evaluation of the tests.
- A new gasifier lining was added since in addition to the test runs with COED chars, test runs were planned with petroleum coke feeds for a series of participating oil companies.

Data Collection and Sampling (Section 3)

This section of the report lists, describes, and illustrates the location of all of the instrumentation and sampling points added to the Puentes plant for the purpose of the test runs with COED chars. Many temperature, pressure and flow measuring instruments were added to assure collection of all information needed to evaluate the effectiveness of the Koppers-Totzek gasifier in gasifying char.

Sampling stations were added for all solid streams including feed, slag, and miscellaneous other solids contained in the gas or as residues from the gasifier. The raw and product gases were sampled at points before and after scrubbing. The water streams from the gas-cleaning equipment and that used to quench and cool the slags were sampled.

A schedule was established prior to each test for the frequency of reading the instrumentation and taking the various samples.

Method of Testing (Section 4)

Prior to starting the gasifiers, the following test work was done:

- The pulverization equipment was adjusted to operate with char and started up to generate sufficient char for the gasification test.
- Preliminary cold test calibration was done with the screw feeders used to feed the char into the gasifiers. This gave an estimate of the feed rate to the gasifier. The actual rate had to be determined later through calculation of a carbon balance.

Main Test Procedures. The most readily-available test parameters were the exit gas temperature and the CO_2 content of the gas. Operation was set at a predetermined exit temperature and associated CO_2 content. The char feed rate, oxygen and steam inputs were determined for steady state conditions. All necessary data were recorded and samples taken according to a predetermined schedule and run plan.

A CO_2 range of 5 to 20 percent of the raw gas was selected for the variable test study. The actual CO_2 percentage varies according to the addition of oxygen and steam to the gasifier.

The object of each run was to produce the maximum amount of raw synthesis gas under the test conditions specified. The limiting factor was the screw conveyors used to feed the char to the gasifier. These conveyors were designed for the physical properties of lignite and not for the lower density of char. Because of this, the screw feeders fed less weight of char to the gasifier than they did lignite. After reaching a steady gasifier operation, adjustments to the entire system were avoided except for small variations in feed rates or changes in oxygen input.

An average series of three complete collections of data and samples were made during each of the test runs. After a test run, the gasification plant was checked and inspected. The gasifier was opened and slag and carbon deposits were removed for analysis and determining the mass balances. All other samples were put in containers and sent for analysis, either at the plant in Spain, or back at the Krupp-Koppers laboratories in Germany.

Analysis of Results (Section 5)

This section presents the following information:

- Analytical procedures used on all samples. Most of these procedures are according to German standards, but similar to U.S. analytical procedures for similar compounds and elements. Of particular interest in the analytical section is the method and mode of calculation of the dust concentration in the product gas leaving the system.

- A sample calculation is shown in depth for one of the ten gasification runs. Run 27 selected for the sample calculation used a char from the Pittsburgh seam coal. The elements in performing the sample calculation for a run included the following:

  --Organizing and averaging all of the collected data on the run sheets from the instrumentation into a usable form.

  --Completing all of the analytical work for the various samples taken.

  --Computing the volume and mass flow rates of all solid, liquid and gaseous streams.

  -- Computing the error terms of measurements and calculations for possible corrections to balances.

  --Calculating mass flow rates of the elements C, H, N, S, and O.

  --Calculating the carbon balance to determine the char feed rate.

  --Calculating material and heat balances.

  --Calculating and tabulating the characteristic performance data for evaluating the efficiency of the gasification test runs.

- Performance data for all runs. Tables at the end of this section contain the important summary data for all test runs.

In computing the mass balances, the exact feed rate of raw char to the system could not be measured since its flow characteristics through the feed screw conveyors were different from those of lignite. Also, operating the screws on a cold feed test gave different results than when actually feeding char to the gasifier under the hot operating conditions. The exact char feed rate was calculated from the carbon balance of the raw gas, slag, and other outlet streams which contained minor quantities of carbon. With this calculated input feed rate, the mass balance for other elements and for the system was then computed. This form of calculation was used for all the test results and details are shown in this section.

Calculation of the heat balance was straightforward, using the heat inputs from the char, water, steam, and oxygen streams and the heat output associated with the product gas, slag, cooling water, and general heat loss from the system.

The quantities calculated for evaluating a gasification run and as the characteristic performance data were:

- Percentage of carbon gasified.
- Fraction of steam decomposed.
- Quantity of usable gas ($CO+H_2$) produced.
- Thermal efficiency of gasification.
- Heat loss from gasifier.

A list of these results and other important factors in evaluating the gasification runs are listed in Table 5-10 through 5-21. These are the basic results from the char gasification tests that were discussed previously in the Summary Report.

Discussion of Test Results (Section 6)

The two COED chars, Western Kentucky and Pittsburgh seam, have similar elemental analyses, and thus yielded similar results during gasification. Five experiments were carried out with each of the chars. The processing variables studied during the test runs with each of these chars were as follows:

- Ratio of oxygen to char.
- Ratio of steam to char.
- Temperature of gasification.

The test results in varying these parameters are shown graphically in this section of the report in Figures 6-1 through 6-8. Conclusions reached from these tests with respect to these operating parameters are:

- The percentage of CO_2 in the raw gas increases with increasing oxygen to char ratio, whereas the CO percentage decreases. The hydrogen content remains almost constant. Overall, there would be less usable fuel gas at high oxygen to char ratios.
- With the same oxygen to char ratio, an increase in the amount of steam lowers the temperature of the gasifier. The temperature will rise with greater oxygen supplies at a constant steam to char ratio.
- The mass flow of total raw gas and useful gas ($CO+H_2$) produced increases with increasing oxygen to char ratio. In the steam to char ratio range considered, no decisive effect of steam addition on the rate of raw gas was shown. The mass flow of useful gas produced ($CO+H_2$) has a maximum and is influenced by the temperature in the gasifier.

- The efficiency of gasifying carbon increases with increasing oxygen supply, but CO_2 is also formed, so that there is a range in which the oxygen and steam should be varied to maximize the amount of useful gas ($CO+H_2$).

  The overall conclusion was that COED char can be gasified with good results in the Koppers-Totzek gasifiers. All the gas produced during char gasification was fed into the synthetic gas pipeline at Puentes and, subsequently, used to produce ammonia. On the basis of the results from this test, a useful gas, suitable for power generation, was produced at 22.4 SCF of $CO + H_2$ per pound of char.

Another important aspect learned from the test was that COED chars could be ground in existing commercial equipment. The ball mill did a good job in producing the degree of fineness required for the gasification test without showing any signs of wear. The recovery equipment collecting the pulverized feed should have to be sized and designed for the properties of the char.

The lining that was used in the gasifiers at Puentes, Spain, only partially held up to the slag characteristics of the char. The reasons for this are known, and Krupp-Koppers claims there is proprietary information to design refractory linings for gasifiers which will hold up to the char slag.

Prediction of Large Scale Commercial Operation Through Computer Simulation (Section 7)

The commercialization of the COED - Koppers-Totzek combination process within the United States could require gasifiers with a capacity of 6 to 12 thousand tons of char throughput per day. To handle these quantities of char, gasifiers even larger than the four-headed units designed for India would be desirable to reduce the number of parallel units.

To anticipate the performance of larger gasifiers, the data collected during the char gasification tests were used to estimate what the performance may be of these larger commercially-sized gasifiers. The results of this simulation are shown in this section. Essentially, commercial char gasifiers would operate at about 2700°F with carbon conversions of about 92 percent. The carbon conversion is related to the reactivity of the coal. As the coal ranks decrease, the reactivity increases. Operation of the K-T gasifier on lignites and young subbituminous coals will give carbon conversion of nearly 100 percent. The carbon conversion would decrease to 95 - 97 percent for a high volatile A bituminous coal. The thermal efficiencies for such a char gasification plant with a gas make of about 58,000 SCF of char feed would be as follows:

Gas/Char Efficiency = 63 percent $= \frac{\text{(Product Gas HHV)}}{\text{(Feed Char HHV)}}$

Gross Process Efficiency = 80 percent

This includes the energy available in the process steam generated along with the energy in the product gas.

Net Process Efficiency = 72 percent

The net process efficiency considers the energy required for power needs and oxygen production. Steam generated in the process could be used for the power needs and oxygen production.

With the larger commercial gasifiers, there will be an increased carbon efficiency because of the higher temperatures at which they can operate and an increased gas make rate. One of the primary reasons for this is the reduction in the heat loss per unit throughput with the large gasifiers.

The Puentes tests adequately demonstrated the commercial use of gasifying COED char in the Koppers-Totzek gasifiers. The char handled easily, particularly compared with many other solids fed through the Koppers-Totzek gasifier. Gas produced from the char was used throughout the operation in the Puentes plant to produce ammonia. The sulfur in the form of hydrogen sulfide or carbonyl sulfide was readily removed and a gas suitable for a fuel gas to a power plant was produced. No problems are anticipated in designing a large-scale unit for gasifying the COED char.

The COED Koppers-Totzek process has been demonstrated in a pilot plant on a level of 36 tons per day of coal feed to produce oil and char, and in a commercial-size gasifier operating with char at a rate of about 75 tons per day to produce a clean fuel gas. The process now could be scaled up to a reasonably-sized plant to help meet future clean energy needs.

Conversion Factors (Section 8)

Factors for conversion from metric to English system are included at the end of the detailed report by the Koppers Company, Inc. and Krupp-Koppers which follows.

GASIFICATION OF COED CHARS IN A KOPPERS-TOTZEK GASIFIER

Prepared by

KOPPERS COMPANY, INC.

Section 1

INTRODUCTION

The Koppers-Totzek process is an established commercial process for the gasification of solid and liquid fuels. Seventeen plants with a total of forty gasifiers are now in operation. These plants gasify feedstocks as diverse as lignite, coal and petroleum residues to a clean synthesis gas suitable for ammonia production. Four more plants with ten gasifiers are under construction. The establishment of the gasification capacity is shown in Figure 1-1.

The Koppers-Totzek process is used in countries where supplies of natural gas and oil are limited, but where domestic coal and lignite can be economically used for ammonia production. As environmental restrictions have accelerated the demand for clean natural gas and oil, the price of these fuels has greatly increased. Therefore, alternate fuels are being investigated to determine whether these can be converted to clean fuels that can supplement our diminishing natural resources.

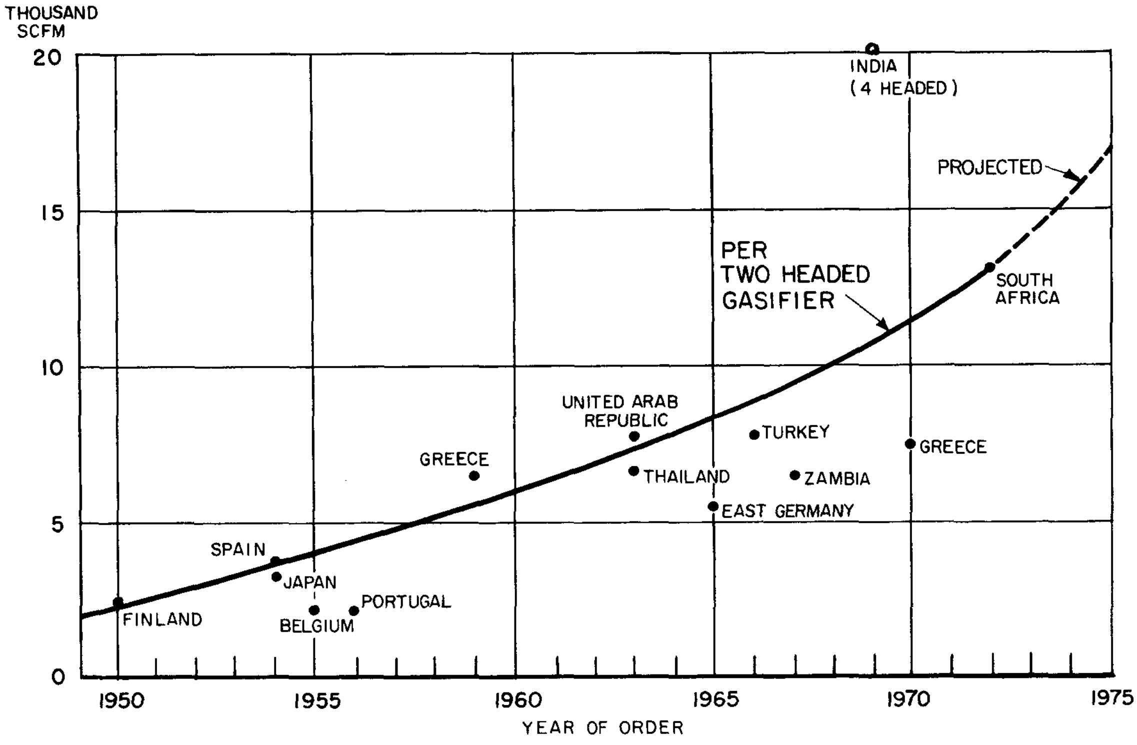


Figure 1-1. Koppers Coal Gasification Process Dry Gas Production Capacity (Chronological Order)

The following report describes the gasification of an alternate fuel, COED char, using the Koppers-Totzek gasification process.

THE COED PROCESS

The Char Oil Energy Development (COED) process subjects finely ground coal to dry distillation (pyrolysis) at successively higher temperatures in a series of fluidized-bed reactors. The volatile constituents of the charged-in coal are almost completely distilled. After being treated with hydrogen, the liquid products are extracted as a synthetic crude oil. The crude oil yield can be as high as 22 percent of the weight of the coal. In addition to the incompletely burnt carbonization gas (4700 to 6100 Kcal/Nm^3), about 60 weight-percent of the coal is converted to a finely divided, dry, low-temperature carbonization char with a high calorific value.

If the char has a high sulfur content its use in power-generating stations is limited. We therefore sought to demonstrate, using a large commercial K-T gasification plant, the economic feasibility of gasifying low-temperature carbonization char. The clean gas produced could then be used either for its heating value or as a chemical synthesis gas. Use of the char in this manner would be environmentally acceptable.

The tests were performed in the facilities of Empresa Nacional de Fertilizantes (ENFERSA) in Puentes de Garcia Rodriguez, Spain. This fertilizer plant, designed for lignite feedstock, was built in 1959. Henrich Koppers G.M.B.H., Essen, the predecessor of Krupp Koppers G.M.B.H., built the gasification plant with three entrained gasifiers, each capable of producing 6000 Nm^3/hr raw gas.

The gasification tests were conducted from June to August 1975. Representatives of FMC, ERDA and EPRI were present during the demonstration as observers.

THE PUENTES TEST SITE

The ENFERSA gasification plant in Puentes was selected as the test site for two reasons: first, there is no Koppers-Totzek gasifier in the United States, and second, the ENFERSA plant has several advantages including adequate raw bunker capacity, two separate material-handling systems in the grinding plants, and space for the char dust with intermediate bunkers that could be separated from each other. Another important feature was the position of the test gasifier in the plant, which enabled the test gas to take a different route from the raw gas in the collecting systems. In addition, the plant had two separate final scrubbing systems with Theisen washer and final cooler, and the gasifier was a suitable size for the available quantities of test char. Finally, the Puentes plant was only 30 km of highway transport from the El Ferrol port, which had good unloading conditions for oceangoing vessels and storage space at the unloading pier. The arrival of char at the El Ferrol port is depicted in Figure 1-2.

Aside from these technical and economic advantages, the good relations between ENFERSA and Krupp-Koppers G.M.B.H. and ENFERSA's years of experience with the gasification of solid fuels made Puentes a good location for the tests.

PROBLEMS

Both during preparations for the tests and during the tests themselves a number of scheduling, operational, and measuring difficulties occurred.

Figure 1-2. Arrival of Char at El Ferrol, Spain, January 29, 1975

Scheduling Problems

The tests were to take place when ENFERSA was out of production for plant repairs, and when, presumably, there would be enough time to make the necessary modifications in test gasifier I and to perform the tests themselves. Because these repairs were completed considerably ahead of schedule, however, the demonstration had to be run during normal plant production.

During normal plant operation, only two gasifiers can operate at the same time. In the preliminary stage of raw gas washing and cooling, each gasifier has its own washer cooler and associated quick-closing valve. In the second state of washing and cooling, however, only one Theisen washer and final cooler is necessary, and the raw gas streams combine.

Because it was necessary to separate the route of the test gas from that of the normal gas produced by gasifiers II and III (Figure 1-3), a connecting pipe was installed between the washer-cooler of test gasifier I and the standby Theisen washer 2, and a flare with a seal pot was built behind final cooler 2. The flare and seal pot were designed so that good-quality raw gas produced in the test could be fed into the normal raw gas main. As it turned out, all of the raw gas produced from the char demonstration test was delivered to the raw gas system and used for NH_3 production. During breakdowns of lignite gasifiers II and III, which normally produced the NH_3, production was sustained by gasification of the COED char only.

Operational Problems

To maintain ammonia production in the plant, it was necessary to operate one of the lignite gasifiers during the char demonstration. Because the lignite

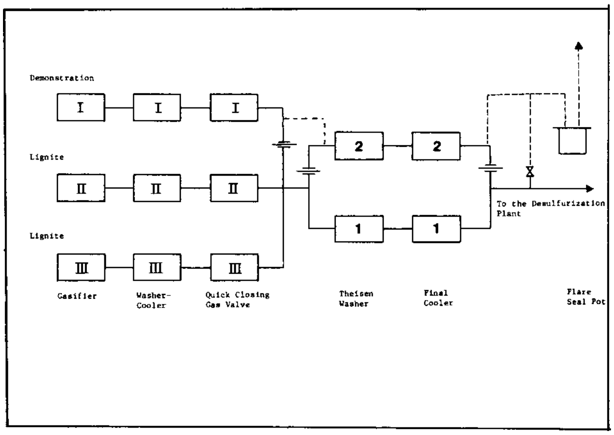


Figure 1-3. Schematic Set-Up of the Gasification Plant Showing Modifications That Were Made to Run the Tests

supplied to the ENFERSA plant during the test period was of variable quality, at times containing more than 40 percent ash, variations occurred in the gasifier's oxygen demand and gas supply. These variations created substantial pressure fluctuations within the plant, which hindered measurements at the demonstration gasifier. This problem resolved itself, however, when early in the test period repair work in an ammonia line reduced one gasifier's demand for raw gas.

ENFERSA therefore agreed to avoid lignite gasification during the periods of char gasification, or until each test was completed. Thus during the demonstration period accurate measurements were made under steady-state conditions.

Milling Plant. The plant is equipped with two ball mills that grind the normal lignite supply down to 72-74 percent below 90 microns. (See Figure 1-4 for a picture of the plant's coal-grinding units.) Each mill grinds approximately 7 tons of lignite (with an ash content of 35 percent) per hour.

The separation of the ground materials to the classifier, above the lignite ball mills, depends essentially on the specific gravity of the ground dust, but the velocity of the entraining gases must also be considered.

A comparison of the specific gravity of COED chars and lignite determined by the Pyknometer method produced the following figures:

Lignite	1.8 kg/dm^3
Western Kentucky Char	1.59 kg/dm^3
Pittsburgh Char	1.57 kg/dm^3

Since the geometric proportions of the classifier could not be modified, it was necessary to reduce the quantity of entraining gas. By partially closing the

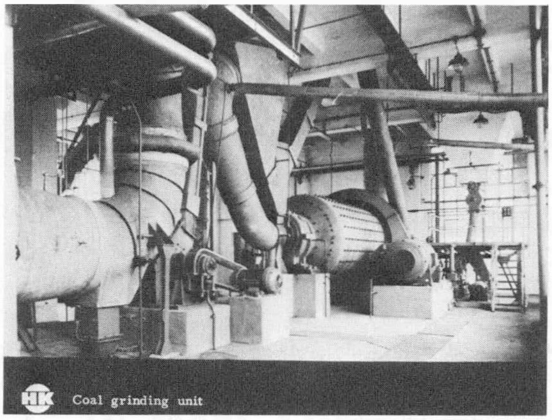


Figure 1-4. Puentes Plant Drying and Grinding Unit

regulating valve to the ball mill fan, the entry flow rate into the classifier was lowered in proportion to the lower specific gravity of the test materials.

However, since a lower flow rate of entraining gases is common to the entire grinding circuit, the reduction in the flow velocity affected the feed distribution between the ball mill and the breaker mill (see flow diagram in Figure 2-1, page 2-3) increasing the amount of coarse material sent to the hammer mill. This increase created an overload, which shut down the entire grinding operation. However, by various endeavors (see Pretest Grinding in Section 4, page 4-1), the entraining gas dust load was reduced until the breaker mill operated without overloading.

To improve the classifier screening, the inlet vanes to the upper section of the outer separator were closed by some 75 percent. This adjustment increased both the flow velocity to the inner section and the cyclone action within the classifier, and made it possible to grind the char down sufficiently for good gasification.

Unfortunately, this procedure reduced char grinding rates to about 2.2 tons/hr and thus shortened test runs, which required up to 4.2 tons/hr of char feed material. However, by utilizing the intermediate storage of some 18 to 24 tons of feedstock, we normally achieved test runs up to about 12 hours and once achieved a test run of over 30 hours.

In the latter stages of the demonstration a major motor burned out, rendering inoperable an entire train of grinding equipment normally used for lignite. The remaining train of equipment was then used alternatively for lignite and test material grinding. A rest period followed each lignite grinding to allow the mill to cool down.

Pneumatic Conveying. The nitrogen pressure was below the initial design requirements. Although adequate for the pneumatic transfer of lignite alone, the pressure was insufficient for the additional transfer of the chars and caused blockages in the pipelines. The problem was solved by the installation of two additional diesel compressors, which boosted the normal plant nitrogen pressure.

Technical Problems in Measurement

The Char Feed Rate. The feed screws to the gasifier were calibrated by removing the transfer pipes and determining the feed rate against the rotary speed of the screw. When the transfer pipes were replaced, however, feed errors were induced that caused unacceptable variations in the screw feed performance. Thus, as an alternative method, flow quantities yielded from gasification were determined and then an elementary carbon balance was used to determine the feed rate of char to the gasifier.

The Raw Gas Flow Rate. The raw gas flow rate from the gasifier is in the order of magnitude of 4000 to 5000 Nm^3/hr. Because of the heavy dust load and the temperature of the gas leaving the waste heat boiler, an orifice plate inserted in this area would have rapidly eroded and become inaccurate. Consequently, the orifice was positioned in the raw gas line after the final cooler. The appropriate allowances were made to calculate the gas flow at the gasifier outlet (see Raw Gas Measurement in Section 2, page 2-14).

The Quantity of Steam Generated. The steam generated in the Puentes plant from the radiation and tubular boilers was only 5 atm steam which is of only marginal use for a modern K-T plant. Because of the expense of separating the steam system and boiler feed water supply from those of the entire plant, steam quantity was not determined. Heat balance was measured only around the gasifier.

THE FEEDSTOCK

Western Kentucky Char

Production site:	36 tons/day-pilot plant of the FMC Corporation in Princeton, New Jersey, USA
Process:	COED. Low-temperature carbonization of bituminous coal in suspended (fluidized) bed.
Initial raw material:	High-volatile bituminous gas coal from the No. 9 seam of the Colonial open pit mine, Kentucky, USA. Sulfur content: 3.3 weight-percent, dry. Brief analysis of the coal: 33.9% volatiles, 54.3% fixed carbon, 11.8% ash, 13% mine moisture.
Appearance of char:	Dark gray, gleaming homogeneous mixture with a predominating fraction of fine-grained material in the particle size range of 0.5 to 2 mm and with a spherical or cylindrical shape; plate-shaped particles in the fine-particle 0.5 mm range; highly flowable. The first char gasified is shown in Figure 1-5.

The tables and diagrams on the succeeding pages contain the following data:

- Physical properties of raw and ground char
- Particle size distributions of ground char
- Reactivities of the raw samples
- Ash analysis of the raw and ground char
- Ash-fusion characteristics of the raw char

Table 1-1

PHYSICAL PROPERTIES OF RAW AND GROUND FEEDSTOCK

Feedstock: Western Kentucky Char			Raw Sample	Ground Sample		
				019	020 021	022 028
Water Content		Wt. %	5.0	1.63	2.33	2.25
Element Analysis	C	Wt. % Dry	76.0	75.0	77.1	75.6
	H	"	1.8	1.7	1.6	1.7
	O	"	2.9	2.3	1.9	2.4
	N	"	1.6	1.6	1.7	1.6
	S total	"	3.1	3.0	3.0	3.2
	Ash	"	14.6	16.4	14.7	15.5
Calorific Value HHV		$\frac{\text{kcal}}{\text{kg dry}}$	6761	6600	6771	6662
Calorific Value LHV		"	6666	6510	6686	6572
Particle Size Sieve Analysis	> 2 mm	Wt. %	1.2			
	1 mm	"	11.7			
	0.5 mm	"	23.2			
	0.25 mm	"	21.7			
	0.125mm	"	18.5	1.4	3.4	4.4
	0.063mm	"	12.3	11.6	18.4	18.0
	< 0.063mm	"	11.4	87.0	78.2	77.6
Particle Size Microns	> 125 μ	Wt. %		1.4	3.4	
	90	"		1.5	3.1	
	63	"		10.1	15.3	
	50	"		1.0	1.2	
	40	"		11.0	14.0	
	30	"		25.0	30.0	
	20	"		41.0	15.7	
	10	"		5.5	9.8	
	< 10	"		3.5	7.5	
Particle Size Fraction	< 90 μ	Wt. %		97.1	93.5	
Density		g/cm^3		1.588	1.590	
Piled Density		kg/m^3	452			605
Rammed Density		kg/m^3	490			845
Grindability		°H	110			

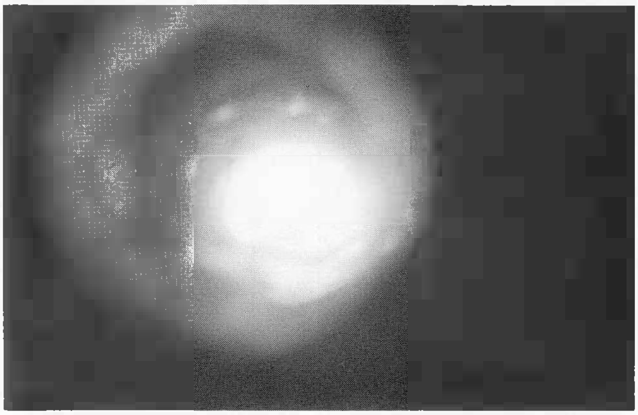


Figure 1-5. Peep Sight Into the Gasifier **Showing Char** Gasified at a Temperature in Excess of 1650°C.

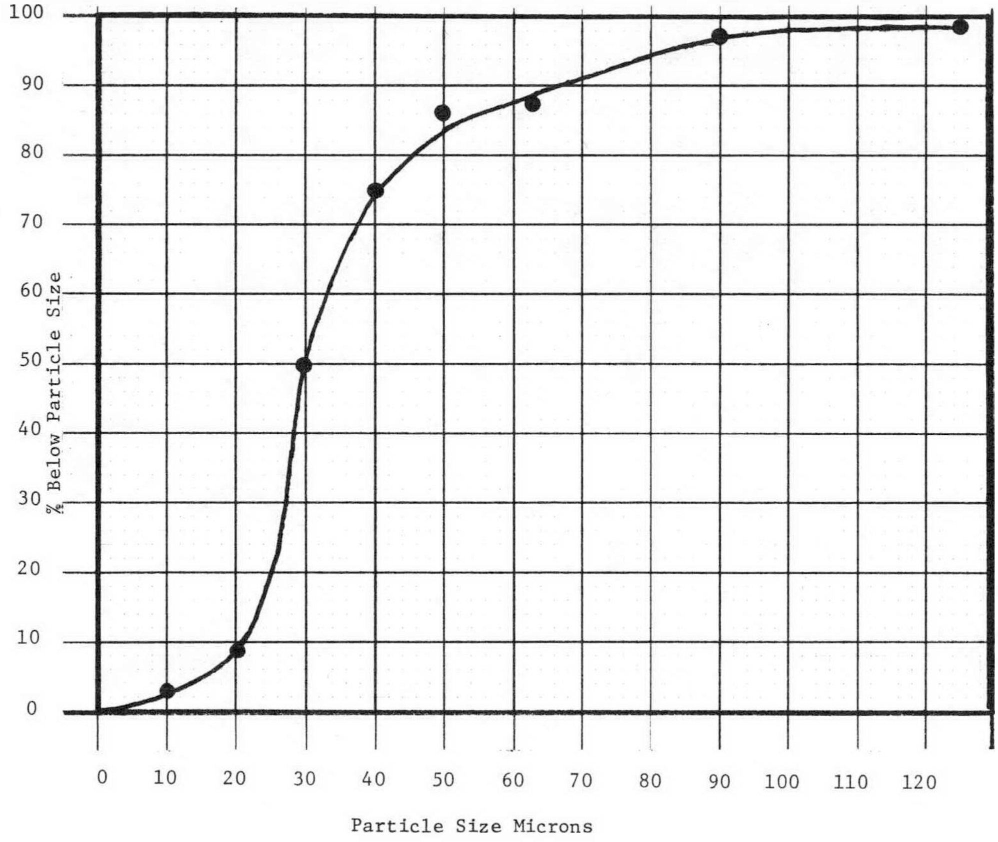


Figure 1-6. Particle Size Distribution of the Western Kentucky Ground Char Sample 19

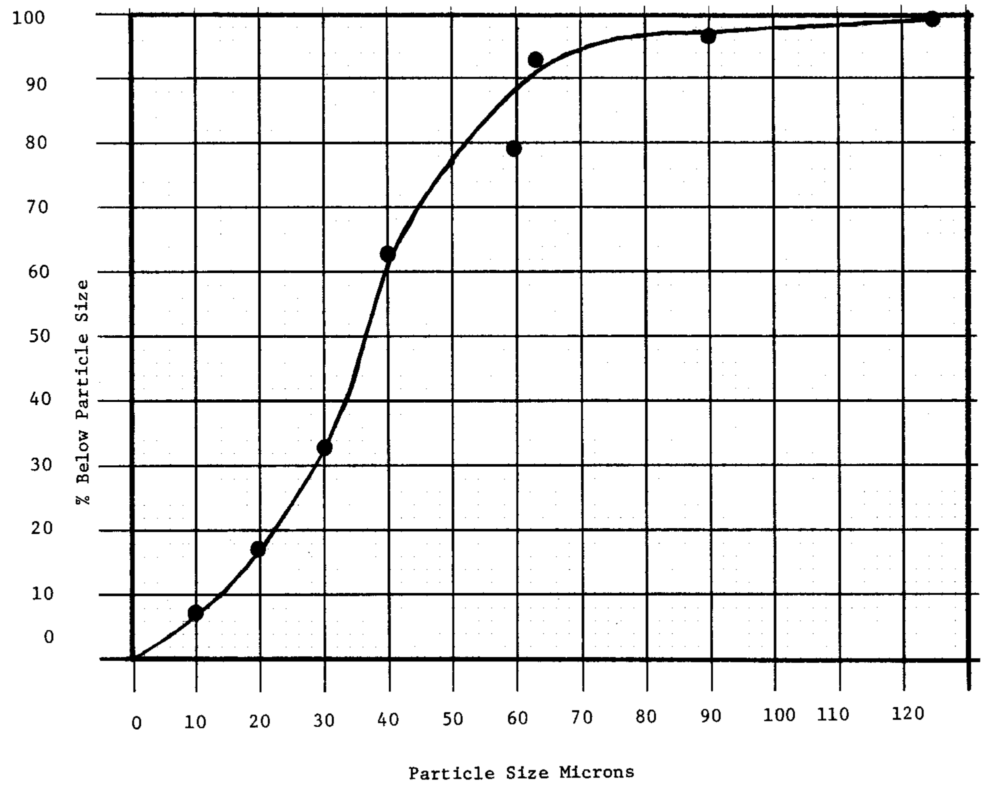


Figure 1-7. Particle Size Distribution of the Western Kentucky Ground Char Sample 020 & 021

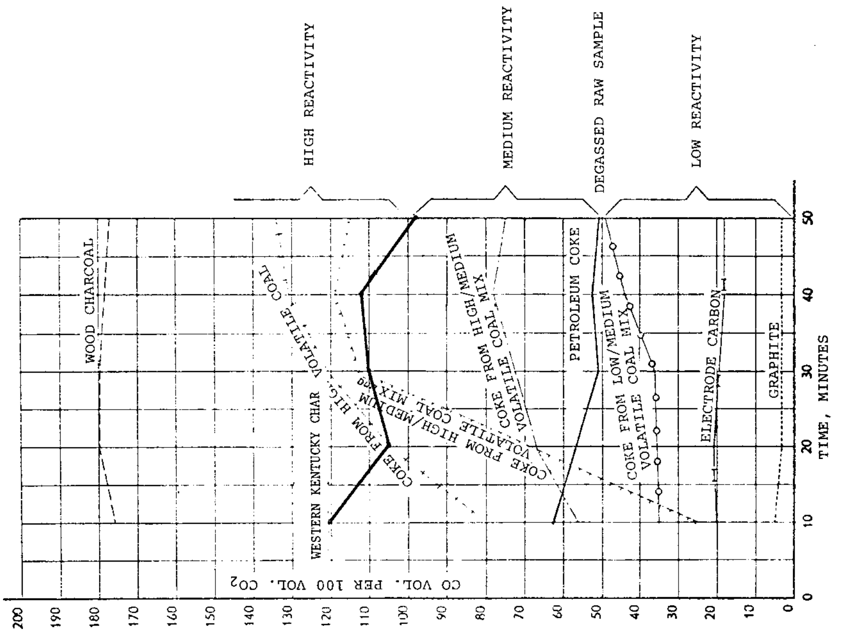


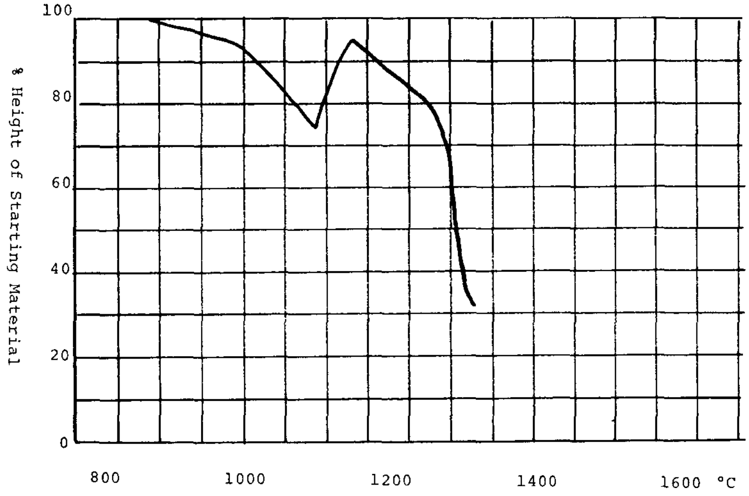
Figure 1-8. Reactivity of Western Kentucky Ground Char at 950°C Compared with Various Fuels

Table 1-2

ASH ANALYSES OF RAW AND GROUND FEEDSTOCK WESTERN KENTUCKY CHAR

Sample	ZnO	CuO	NiO	Fe_2O_3	MnO	V_2O_5	K_2O	CaO	SO_3*	TiO_2	SiO_2	Al_2O_3
Western Kentucky Raw Sample	<0.1	<0.1	<0.1	26.2	<0.1	0.2	2.1	2.5	2.3	1.0	42.4	21.0
019	<0.1	<0.1	<0.1	25.8	0.1	0.1	2.0	3.0	2.6	0.9	42.5	20.5
020												
021	<0.1	<0.1	<0.1	25.5	<0.1	0.1	2.0	2.7	2.3	1.0	42.8	21.0
022												
028	<0.1	<0.1	<0.1	25.9	0.1	0.1	2.1	3.0	2.3	0.9	42.7	20.9

* The SO_3 values are to be considered solely as reference values.



The following characteristic data of the ash-melting process were obtained from observations of shape and state changes in the ash sample at successively higher temperatures.

Softening point:	1090°C
Melting point:	1250°C
Flow point:	1280°C

Figure 1-9. Ash-Fusion Characteristics of the Western Kentucky Char Raw Sample, Ashed at 800°C

Pittsburgh Char

Production site:	36 tons/day-pilot plant of the FMC Corporation in Princeton, New Jersey, USA
Process:	COED. Low-temperature carbonization of bituminous coal in suspended (fluidized) bed.

Initial raw material:	High-volatile bituminous gas coal from the No. 8 seam of the Ireland underground mine in West Virginia, USA. Sulfur content: 4.0 weight-percent, dry. Brief analysis of the coal: 40.2% volatiles, 51.4% fixed carbon, 8.4% ash, 2.3% mine moisture.
Appearance of char:	Dark gray, homogeneous mixture with preponderance of fine-particle fraction; spherical or cylindrical particles in the particle size range from 0.5 to 2 mm; plate shaped in the fine-particle range <0.5 mm, highly flowable.

The tables and diagrams on the succeeding pages contain the following data:

- Physical properties of raw and ground char
- Particle size distribution of ground char
- Reactivities of the raw samples
- Ash analyses of raw and ground char
- Ash-fusion characteristics of the raw samples

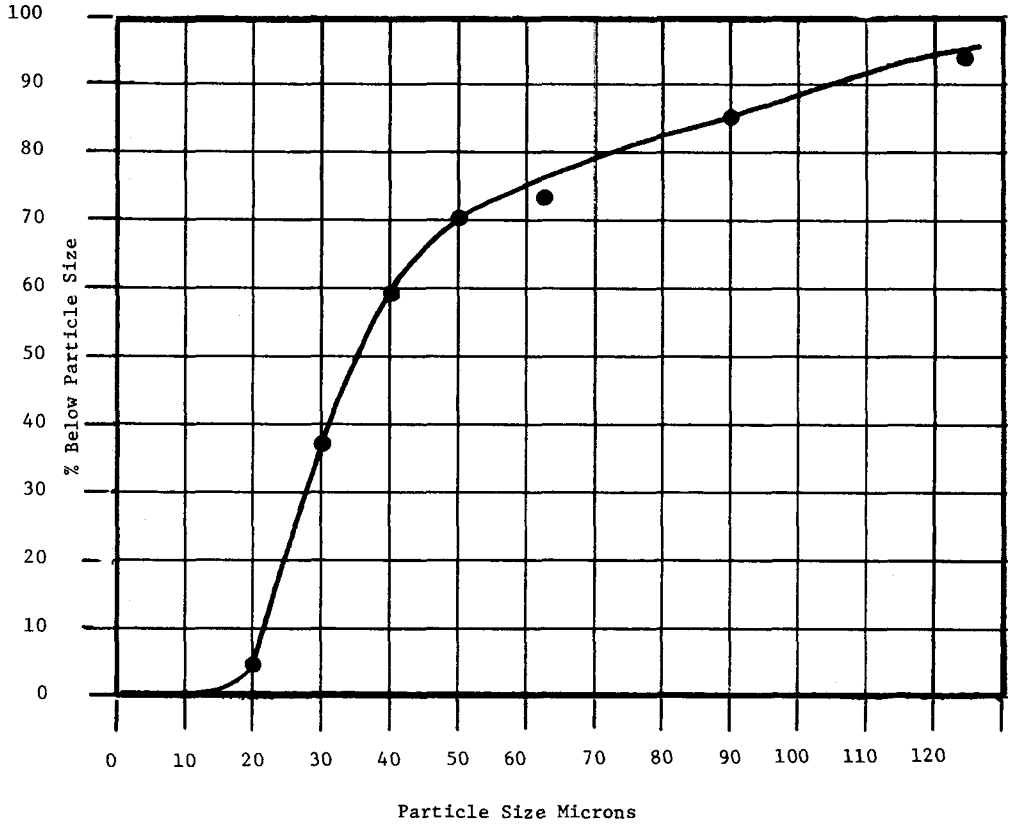


Figure 1-10. Particle Size Distribution of Pittsburgh Ground Char 023-025

Table 1-3

PHYSICAL PROPERTIES OF RAW AND GROUND FEEDSTOCK

Feedstock: Pittsburgh Char			Raw Sample	Ground Sample	
				023 024 025	026 027
Water Content		Wt. %	3.1	1.80	1.63
Element Analysis:	C	Wt. % dry	75.2	75.0	76.4
	H	"	1.7	1.9	1.7
	O	"	2.8	2.3	3.3
	N	"	1.4	1.4	1.4
	S total	"	4.0	4.1	4.0
	Ash	"	14.9	15.3	13.2
Calorific Value HHV		$\frac{kcal}{dry}$	6732	6723	6738
Calorific Value LHV		"	6642	6623	6648
Particle Size	> 2 mm	Wt. %	0.6		
	1 mm	"	11.7		
Sieve analysis	0.5 mm	"	22.5		
	0.25 mm	"	21.2		
	0.125mm	"	18.4	6.0	4.6
	0.063mm	"	13.3	20.4	19.4
	< 0.063mm	"	12.3	73.6	76.0
Particle Size	> 125 μ	Wt. %		6.0	
Microns	90	"		5.0	
	63	"		15.4	
	50	"		3.6	
	40	"		11.0	
	30	"		22.0	
	20	"		32.6	
	10	"		3.6	
	< 10	"		0.8	
Particle Size Fraction	< 90 μ	Wt. %		89.0	
Density		g/cm^3		1.565	
Piled Density		kg/m^3	505	615	
Rammed Density		kg/m^3	575	835	
Grindability		°H	112.8		

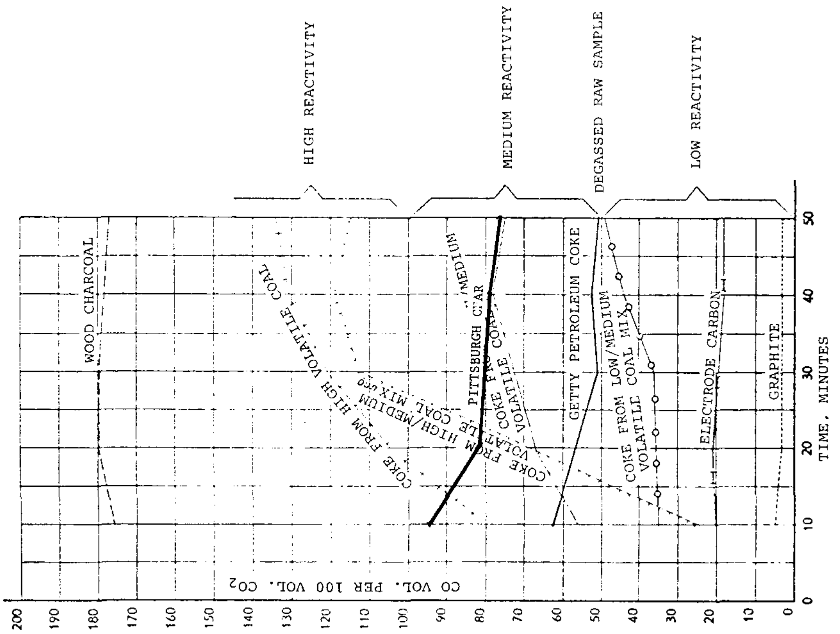


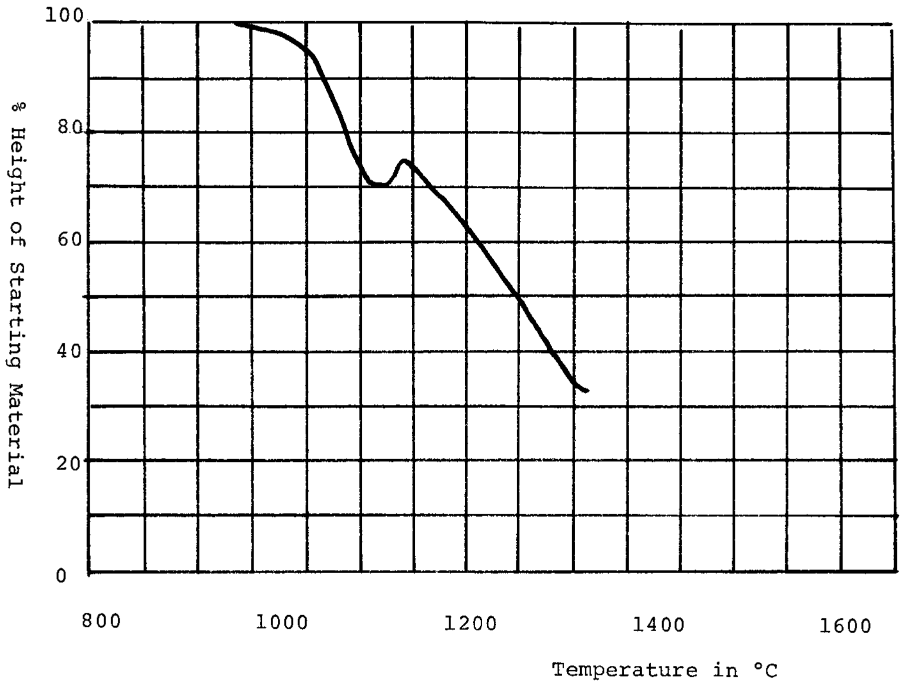
Figure 1-11. Reactivity of Pittsburgh Char at 950°C Compared With Various Fuels

Table 1-4

ASH ANALYSIS OF THE RAW AND GROUND FEEDSTOCK PITTSBURGH CHAR

Sample	ZnO	CuO	NiO	Fe_2O_3	MnO	V_2O_5	K_2O	CaO	SO_3*	TiO_2	SiO_2	Al_2O_3
Pittsburgh Raw Sample	<0.1	<0.1	<0.1	29.6	<0.1	0.1	1.6	1.9	1.0	0.9	41.7	20.2
023 024 025	<0.1	<0.1	<0.1	30.3	<0.1	<0.1	1.6	2.1	1.0	0.9	41.1	19.6
026 027	<0.1	<0.1	<0.1	31.1	<0.1	<0.1	1.5	2.2	1.2	0.9	39.7	19.6

*The SO_3 values are to be considered solely as reference values.



The following characteristic data of the ash-melting process were obtained from observations of shape and state changes in the ash sample at successively higher temperatures.

Softening point: 1060°C
Melting point: 1170°C
Flow point: 1260°C

Figure 1-12. Ash-Fusion Characteristics of the Pittsburgh Char Raw Sample, Ashed at 800°C

Section 2

DESCRIPTION OF THE PUENTES PLANT

COAL PREPARATION

Description

The Puentes plant contains two separate but complete systems for drying and grinding lignite (see Figure 2-1 for a diagram of these systems). One line of equipment, after the rotary predryer through the pulverized fuel storage bunkers was used for the demonstration. A temporary feed system consisting of a hopper and short conveyor feed belt was added to the system to deliver the char into the plant. The feed hopper was installed outside the predrying plant. The conveyor belt fed directly to the main conveyor system, which transferred raw material into the raw fuel storage bunkers.

Bunkers number 3 and 4 were made available for selective storage of the chars. The bunkers had originally included nitrogen injection and temperature sensors to insure safe lignite storage, but this equipment was no longer operative and was not reactivated, since the inert characteristics of the char make it safe to store. The bunker compartments were closed off, however, to prevent excessive ingress of air.

Enclosed drag conveyors with scraper bars transferred the char feed stock from the bunker through a rotary valve and screw feed into the rising inert hot gas main. Since COED char is dry in comparison to lignite (lignite contains about 35 percent moisture after the predryer), the normal inert hot gas, generated from the combustion of the lignite, was replaced by nitrogen.

Entering the rising gas main, the finer particles of the feed material were entrained and carried upwards with the gas flow. The coarser particles of the char feedstock fell through the rising gases and were deflected into the hammer mill to be broken up prior to their entrainment.

The entrained feedstock passed through the overhead feeder and then down into the ball mill, where the heavy steel balls pulverized the feedstock. After leaving the ball mill, the pulverized feedstock was entrained and conveyed to the classifier, which separated the oversized particles from the entrained gas.

The pulverized feedstock was then transferred to the final separator and dust-catcher and fed into the primary pulverized-char bunker. The entraining gas was returned to the ball mill fan which maintained the total flow. At the fan outlet some of the entraining gas returned to the inert gas producer (normally this portion maintained flow and temperature control within the system), another portion entered the transfer system at the ball mill outlet, and the excess progressed to an injection blower that blew it into the atmosphere. (A modern grinding complex would include electrostatic precipitators to extract the total dust content and clean the gases prior to atmospheric release.)

Modifications

In order to grind the different kinds of feedstock, the following modifications were made in the existing coal-preparation plant.

Loading of the conveyor belt behind the drying drum was accomplished by an additional conveyor belt with feed hopper. The feedstock was dumped from trucks in front of the building and loaded into the hopper with a bucket (or front end) loader.

The swirl vanes of the classifier were no longer in operating condition and had to be restored so that the flow velocities in the classifier could be regulated.

In order to achieve the required degree of grinding, the associated problems of overloading the breaker mill and separation of the dust within the classifier required that a substantial reduction be made in the feedstock throughput. This reduction was achieved by installing an additional scraper at the exit slide valve of the char bunker and thus decreasing the feed height on the drag conveyor to a minimum of 65 mm. Based on a width of 750 mm for the chain conveyor, a charging layer-height of 65 mm, and a chain conveyor velocity of 2.4 cm/s, the following milling rates were obtained: Western Kentucky, 2.1 tons/hr.; Pittsburgh, 2.35 tons/hr.

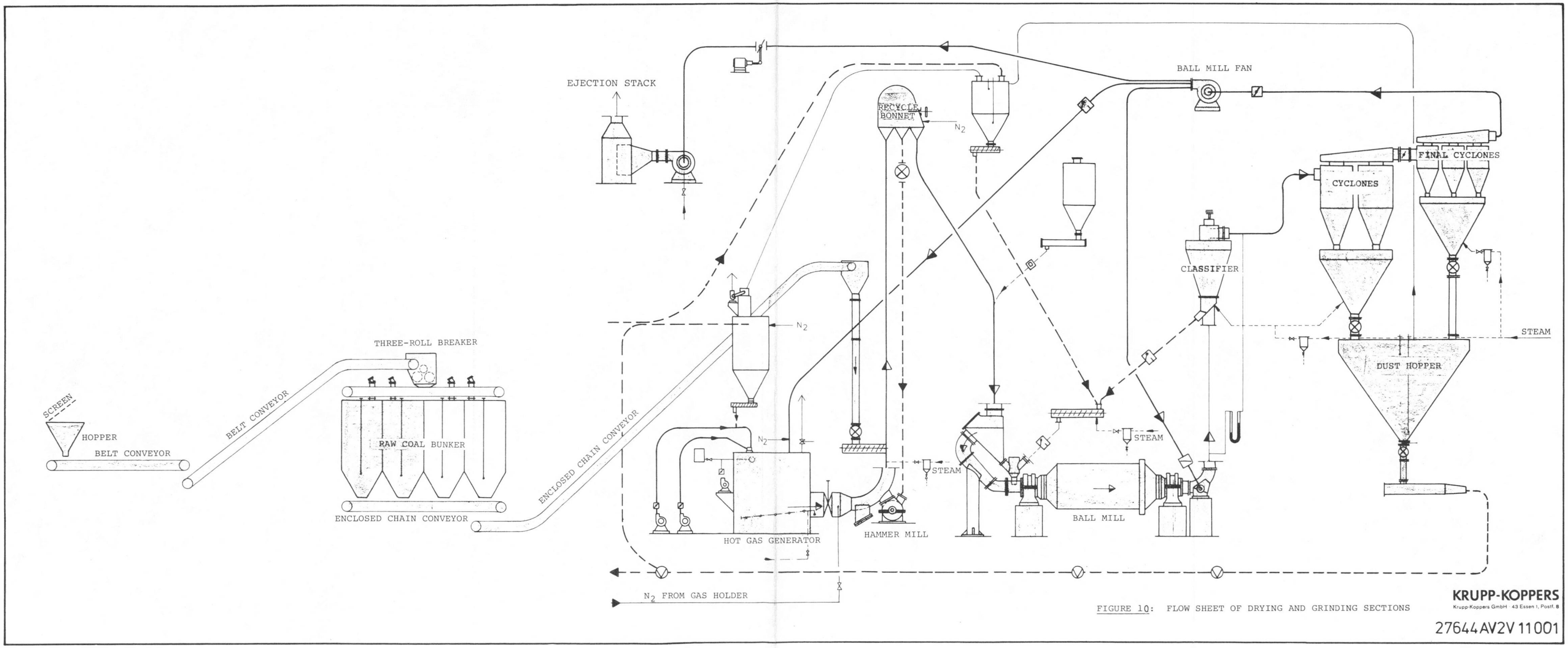


FIGURE 10: FLOW SHEET OF DRYING AND GRINDING SECTIONS

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Figure 2-1. Flow Sheet of Drying and Grinding Sections

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A nitrogen line was installed so that impure nitrogen (containing no more than 9 percent oxygen) would replace the hot flue gas normally used in the grinding system for entraining and drying lignite.

Initially, the pulverized char feed was too dry for smooth control when it reached the gasifier screws. Therefore, water spray equipment was installed over the chain conveyor to moisten the feedstock before it arrived at the ball mill. The spray system was capable of delivering about two liters of water per minute.

PNEUMATIC CONVEYING

Description

The Puentes plant has two separate pneumatic systems completely interchangeable. The pneumatic dust conveying system is diagrammed in Figure 2-2.

Each system is split into two main sections. The first section extends from the pulverized storage bunker through a Fuller pump and pneumatic conveying pipes to an intermediate storage bunker. The second section comprises intermediate storage bunkers, a Fuller pump, and six service bunkers that feed the plant's three gasifiers. From the service bunkers the pneumatic transfer line returns to the intermediate bunker.

The controls for the first section of the pneumatic transfer system are mounted in the grinding plant. The controls for the second section of the system are mounted under the intermediate bunker. The coal level within the intermediate bunker is indicated on an exterior dial.

Modifications

The pneumatic system was designed for operation at 1.5 atm. However, due to wear over the years, the nitrogen rotary compressors provided a final pressure of only 1.1 to 1.2 atm with a nominal line pressure of 0.95 atm. Pressure fluctuations created when a second Fuller pump was activated dropped the line pressure as low as 0.65 atm.

During preparation for the demonstration, the lignite feed system and the char feed system were often operated simultaneously. The line pressure at 0.65 atm was insufficient to support the pneumatic transfer of the char, and consequently, ascending pipes became clogged.

After extensive effort, two diesel compressors were located in Spain, modified in the manufacturer's plant at Victoria, and transferred to the site in time for the demonstration. One of the compressors was connected to the nitrogen lines from the nitrogen holder to the grinding mill, and the other was connected to the nitrogen line from the holder to the blower house. Manually operated slide valves were installed at each connection and operated in conjunction with the compressors. The safety valves on the compressors were adjusted to 2.2 atm, and at a line pressure of 1.5 atm the petroleum coke and char feedstocks were conveyed without difficulty.

During lignite gasification nitrogen used to convey lignite to the service bunkers is continuously recycled to the intermediate bunker. During the demonstration the danger existed that lignite dust could accidently be returned to intermediate bunker 5 and thus contaminate the demonstration feedstock. To avoid this all pneumatic conveying lines used for char were sealed using blind flanges at all times except when char was being transferred.

GASIFICATION PLANT

The Puentes gasifier (BFD-1001), which is illustrated in Figure 2-3, consists of a horizontal, cylindrical, water-cooled steel vessel. (See the flow diagram of the gasification plant in Figure 2-4.) Burner heads are mounted on either tapered, cylindrical end of the gasifier. Each head is equipped with two burners. Through these burner heads an accurate proportioned, homogenous blend of pulverized feed, oxygen, and steam is introduced into the gasifier, where ignition of the feedstock under conditions of partial oxidation takes place at high temperatures.

The high reaction temperatures crack the higher condensible hydrocarbons in the feedstock. The raw synthesis gas leaving the gasifier predominantly consists of CO, CO_2, H_2, and water. The composition of this gas corresponds approximately to the water-gas equilibrium at the exit temperature. The high gas temperatures liquify and slag the ash contained in the feedstock; some of this slag is deposited on the walls of the gasifier and, in fluid form, flows downward through the seal leg into a water quench tank. The remainder of the slag leaves the gasifier with the raw gas, in the form of fly ash.

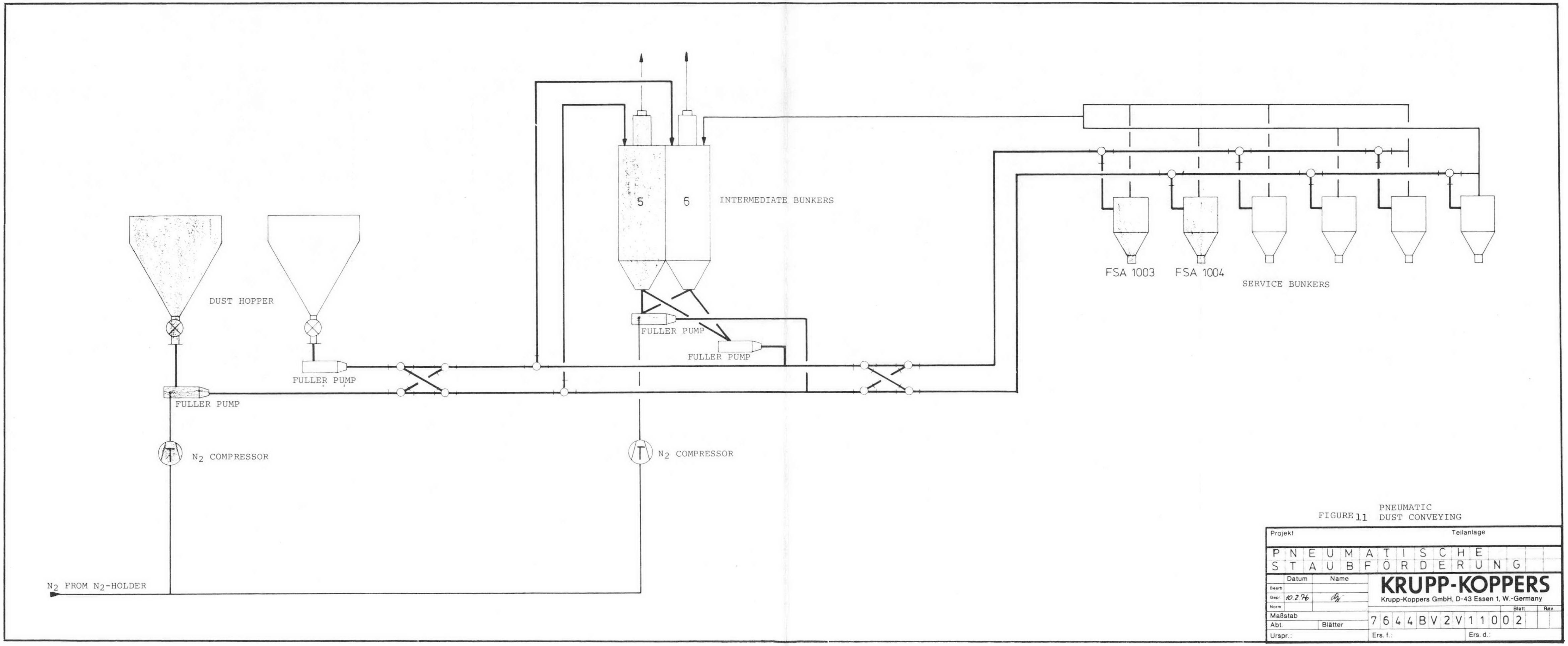

Figure 2-2. Pneumatic Dust Conveying

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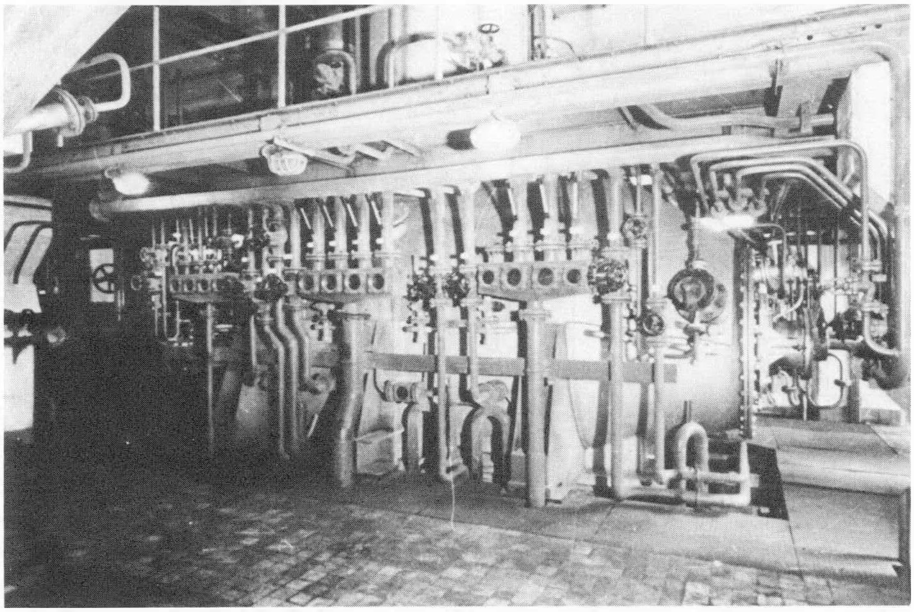


Figure 2-3. Puentes Gasifier Capacity 6000 Nm^3/hr Raw Gas.
Put on Line 1958 for Gasification of Lignite.
1966—Trial Gasification Naptha.
1975—Demonstration Gasification Delayed & Fluid Coke, Western Kentucky and Pittsburgh Char

Ground Feedstock

The ground feedstock prepared in the grinding plant is pneumatically transferred from the finished dust bunkers through the intermediate bunker by Fuller pumps, using pressurized nitrogen, into the gasification plant bin system. Each side of the gasifier has its own bin system consisting of service bin FSA-1004 and feed bin FSA-1005 or FSA-1006. An electronic maximum-level indicator monitors the level of each service bin. Between the two bins is a Bailey distributor RNA-1006, which is controlled by an electronically driven rotating sensor mounted at the top of the feed bin.

At the bottom of the feed bin the ground feedstock is distributed through an inverted Y pipe to the two screw conveyor units. Emergency sensors installed above the screw feed boxes in the Y pipe operate according to the same principle as the rotating sensors, but they automatically shut down the gasifier when no feed is available. The feed bin is under 200 to 400 mm w.c. nitrogen pressure. Surplus nitrogen is released through an adjustable water seal.

Oxygen

A rotary compressor draws in oxygen from the oxygen holder and compresses it to approximately 4000 mm w.c. The quantity of oxygen flowing to each mixer head is regulated by manually operated valves located downstream from the automatic high-speed closing valves. Each oxygen supply is preheated by steam in one of two heat exchangers, EA-1001 or EA-1002, to approximately 110°C. This temperature is required to prevent condensation of steam when oxygen and steam flow together in blender UAD-1001 or UAD-1004 on the way to the mixing heads. In the mixing heads, the O_2/steam mixture is thoroughly blended with the ground feedstock and sent through the transfer blower pipes. The blend of ground feedstock/oxygen/steam is fed to the gasifier through the burners.

Supplementary Steam

The steam required for the gasification process is taken from the plant steam system, at a pressure of 5.7 kg/cm^2, and supplied through orifices to oxygen/steam blenders UAD-1001 to UAD-1004.

Raw Synthesis Gas

The raw synthesis gas produced in gasifier BFD-1001 flows to the gasifier outlet and through a quenching zone to the boiler system. The boiler system consists of a radiation boiler BCC-1002 and tubular boiler BCD-1003, in which the sensible heat of the gas is made available for generating a 5 atm saturated steam. During the starting operation the raw synthesis gas is carried through seal pot FKA-1007 into the flare stack. This heat recovery process is illustrated in Figure 2-5.

After leaving the BCD-1003 tubular boiler, the gas arrives in the washer-cooler DBC-1001, where a large percentage of the fly ash is removed and the gas is cooled to approximately 25°C. To clean the gas sufficiently for compression and subsequent synthesis, the gas is then subjected to thorough scrubbing in a Theisen washer DBF-1003 and finally passed through a mist eliminator.

Slag

In the gasifier, a portion of the fluid slag flows downward into the seal leg and drops into the water quench tank of the slag removal system RKK-1005, where it is granulated. A scraper conveyor then removes the granulated slag from the bottom of the quench tank.

Flue Dust

Part of the flue dust entrained in the synthesis gas is removed by a dustcatcher on the tubular boiler. From there it is drawn off through seal lines and carried

GASIFIER	BFD-1001
RADIATION BOILER	BCC-1002
TUBULAR BOILER	BCD-1003
WASHER COOLER	DBG-1001
FINAL COOLER	DBG-1002
THEISEN WASHER	DBF-1003
OXYGEN HEATER	EA-1001/2
COOLING WATER ELEVATED STORAGE TANK	FBB-1001
COOLING WATER COLLECTING TANK	FBB-1002
SERVICE BUNKERS	FSA-1003/4
FEED BUNKERS	FSA-1005/6
START UP FLARE SEAL POT	FKA-1007
FLARE SEAL POT	FKA-1008
NITROGEN SEAL POT	FKA-1009
SEAL POT	FKA-1010/11
QUICK GAS SHUT-OFF	FKA-1012
SCREW FEEDER	RKG-1001÷1004
ASH EXTRACTOR	RKK-1005
BAILEY FEEDER	RNA-1006/7
OXYGEN/STEAM MIXER	UAD-1001÷1004

FIGURE 16: INSTRUMENTATION AND SAMPLING

Projekt
Teilanlage
MEßSTELLEN UND PROBENNAHME
Datum Name
Bearb.
Gepr. 10.2.76
Norm
Maßstab
Abt. Blätter
Urspr.:
KRUPP-KOPPERS
Krupp-Koppers GmbH, D-43 Essen 1, W.-Germany
Blatt Rev.
7644BV2V11006
Ers. f.:
Ers. d.:

Figure 2-4. Flow Sheet Gasification Section

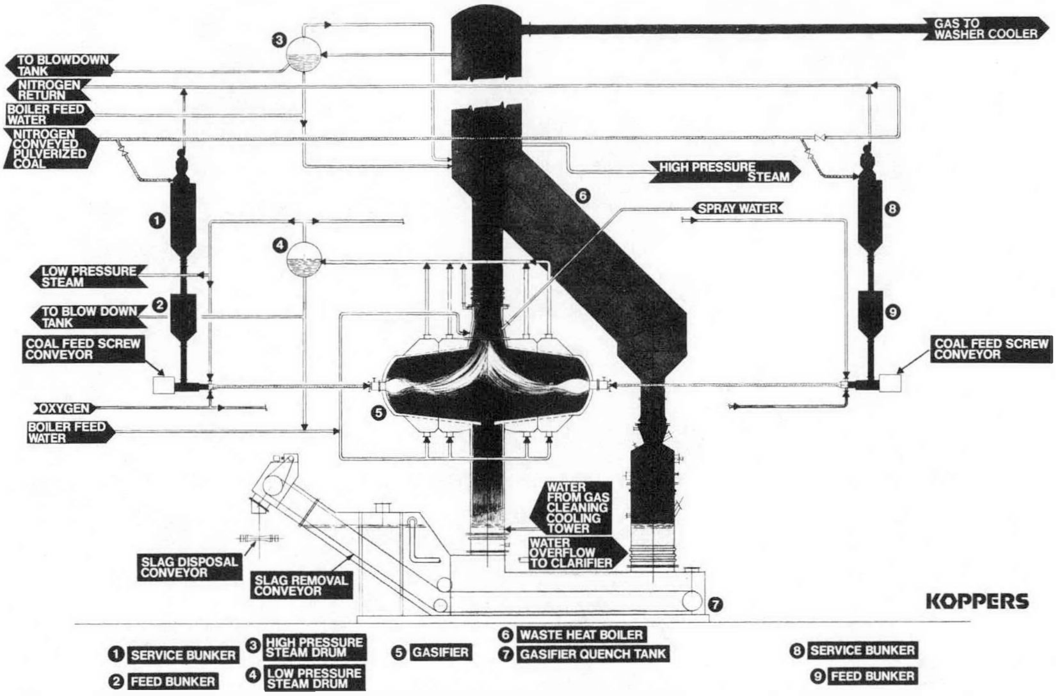


Figure 2-5. K-T Gasification and Heat Recovery

away in the wash water. The remainder of the flue dust entrained in the raw gas is scrubbed out in the washer-cooler and Theisen washer.

Nitrogen

In the event of a safety failure, nitrogen is used to flush and deactivate the gasifier. In this process the gas main is purged and deactivated up to the main synthesis gas flare.

Nitrogen is also used in the service and feed bins as described earlier in this section, under Ground Feedstock, page 2-9, in the gasifier sight glasses, and as purging gas for the test lines.

The nitrogen which is at least 95 percent pure, is drawn from the gas holder by a two-stage rotary compressor, compressed to about 1.1 atm, and stored in tanks.

Cooling Water

Cooling water is supplied from the plant system. Recycled cooling water is taken to storage tanks and mixed with fresh cold water, while the corresponding quantity of heated recycled water is drawn off from the circuit. The cooling water is then pumped into a gravity tank and from there flows to the cooling stations.

Wash Water

For the wash water required for the washer-cooler, final cooler, Theisen washer, quenching zone, and various seals, fresh treated water from the plant water supply is used at a pressure of 4.0 to 5.50 atm. Pipes carry the wash water back to the river.

Modifications

Oxygen/Steam Mixer. In order to ascertain the effect of steam on the gasification of the feedstock, oxygen/steam mixers were installed in each of the four supply lines conveying oxygen to the gasifier heads. A supply of saturated steam was available for addition.

Oxygen Heater. In order to mix the specified quantity of steam with the oxygen, the temperature of the oxygen had to be raised to 110°C. Saturated steam was available for heating purposes.

Since the heating of the oxygen was to be confined to the test gasifier, new heaters had to be installed that conformed to the existing oxygen pipelines to the two gasifier heads.

Test Flare. A test flare stack was installed to separate the demonstration raw gas from the main flow of gas from lignite. This flare stack is depicted in Figure 2-6. First the raw gas line was separated from the gas main at the high-speed gas seal after washer-cooler 1. A new gas line was then installed, connecting the gas seal directly to Theisen washer 2, right before final cooler 2 (see Figure 2-1). The flare stack was placed after final cooler 2, and the lines were run so that the raw gas produced in the demonstration could be either flared or delivered to the raw gas main and used for ammonia production. A seal pot insured that air would not be drawn through the flare.

Raw Gas Measurement. Because the orifices in the raw gas main measured the total gas flow from all gasifiers, they could not be used to measure the raw gas produced only from the demonstration. Thus, a new orifice was fitted across the pipeline between final cooler 2 and the flare seal pot, where there is negligible dust in the raw gas and thus little danger from erosion of the orifice. This modification is depicted in Figure 2-6.

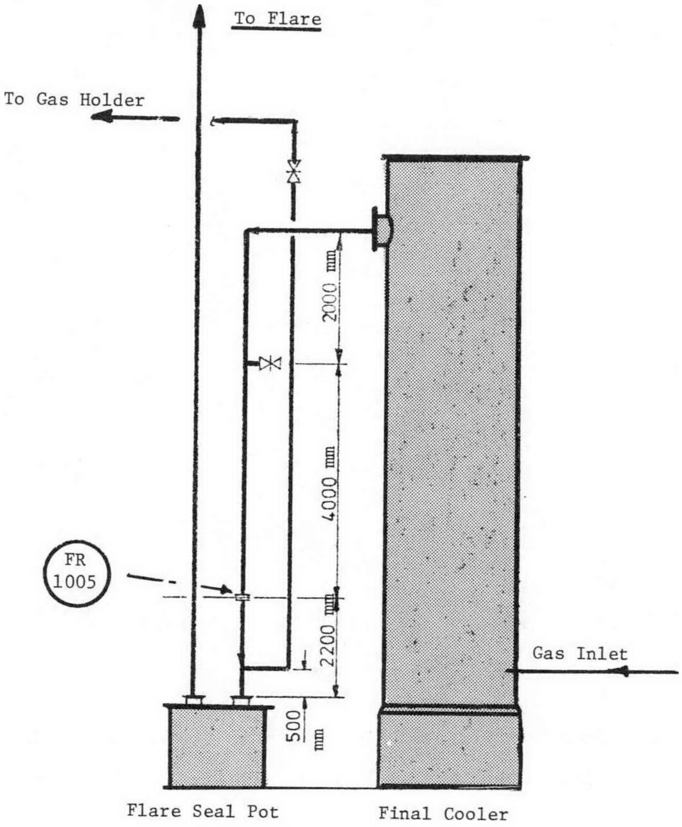


Figure 2-6. Flare Stack

Temperature Measurement in Gasifier Inner Chamber. In order to measure the temperature inside the gasifier two Pt-PtRh thermocouples were inserted (at TI 1006A and TI 1006B) on the top side of the gasifier between the gasifier burners and the gasifier outlet (see Figure 2-7).

Welded supports were used to anchor the thermocouples, which were also protected by a 24 mm double-walled sillimanite tube. The thermocouples were positioned so that the junction extended inside the chamber 200 mm beyond the gasifier lining. To prevent errors in measurements due to slag deposits, the thermocouples were replaced after each demonstration run. A test rod checked the lining thickness at regular intervals so that the depth of the thermocouples could be accurately determined.

Gasifier Lining. The K-T gasifier has an inner rammed lining of refractory material. For lignite, a chromium ore lining material with a high percentage of Cr_2O_3 is normally used.

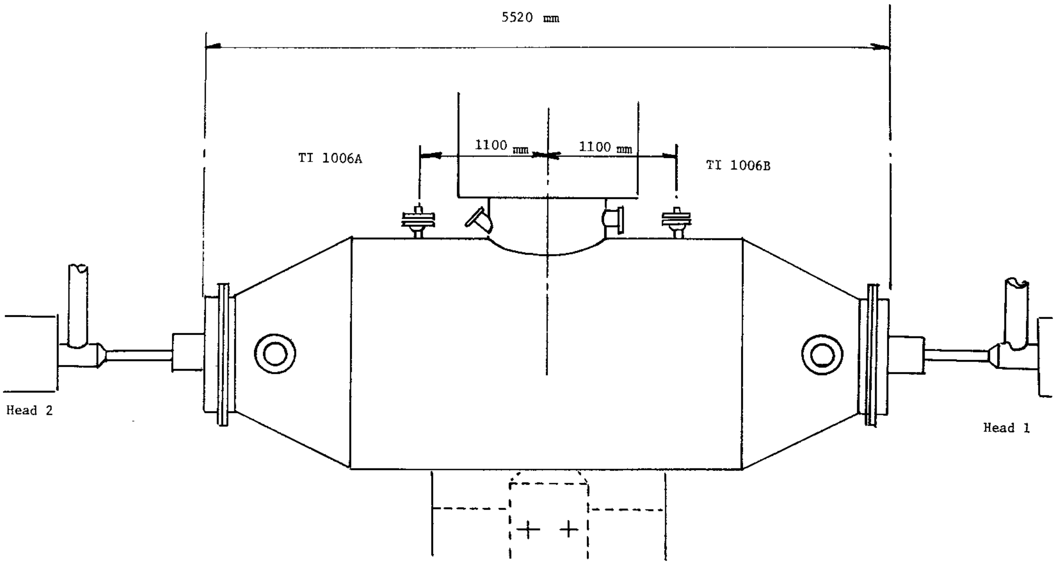


Figure 2-7. Arrangement of Temperature-Measuring Supports on the Gasifier

A major purpose of the demonstration was to determine how operating conditions affect the gasification of COED chars. Therefore the normal gasifier lining used for lignite was replaced with a corundum refractory lining, which permitted operation at a higher temperature (1700°C).

The corundum lining used was Norton CA 70. Its chemical composition is as follows:

Al_2O_3	95.98%
SiO_2	0.05%
Fe_2O_3	0.11%
CaO	3.62%
Na_2O	0.16%
MgO	0.08%

The material was mixed with about 10% water in a concrete mixer. After the inside walls of the gasifier were cleaned, sandblasted, and studded, the lining was applied by hand. Hot water was circulated in the gasifier jackets to dry in the lining, and gasifier oil burners were used for final burning.

Section 3

DATA AND SAMPLE SELECTION

MEASURE (CONTROL) POINTS

The measurement and sampling points are depicted in Figure 3-1. Basically, measuring and instrumentation layout had two functions:

- To insure and maintain steady and safe operations in the overall gasification process.
- To collect the necessary demonstration data for evaluation.

Tables 3-1 to 3-4 identify new and existing instruments and control points. In all the tables KS refers to the No. 1 head of the gasifier on the washer-cooler side and MS refers to the No. 2 head of the gasifier on the control station side.

Temperature Measurement

For most of the measuring points, especially in the gasifier area, the control data were checked according to the first point, above.

The position of temperature points TI 1020 to TI 1041 for the gasifier cooling system is shown in Figure 3-2. The temperature points inside the gasifier TI 1006A and TI 1006B were read directly in degrees Celsius and corrected according to the cold junction TI 1042A and TI 1042B.

Temperature-measuing points are designated by the index range TI 1001 to TI 1043 (index 1019 was not used).

Pressure Measurement

Because measuring station PR 1006 was constantly in error, it was replaced by local measuring station PI 1018 during the tests. The measuring stations were designated by index PI-1001 to PI 1021 and are described in Tables 3-5 and 3-6.

Table 3-1

TEMPERATURE-READING POINTS TI 1001 - TI 1012

		Measuring Point				Indication					
Measuring Station Index	Medium	Location	Type of Instrument	Existing	New	Dimension	Local	Control Room	Recorded	Test Data	Control Data
TI 1001	O_2/steam mixture	O_2 line inlet 1	Thermometer		X	°C	X			X	
TI 1002	O_2/steam mixture	O_2 line inlet 2	Thermometer		X	°C	X			X	
TI 1003	O_2/steam mixture	O_2 line inlet 3	Thermometer		X	°C	X			X	
TI 1004	O_2/steam mixture	O_2 line inlet 4	Thermometer		X	°C	X			X	
TI 1005	Raw synthesis gas	Outlet Washer-cooler	Thermometer	X		°C	X				
TI 1006A	Raw synthesis gas	MS Inside gasifier	Thermocouple Pt-PtRh		X	°C		X		X	X
TI 1006B	Raw synthesis gas	KS Inside gasifier	Thermocouple Pt-PtRh		X	°C		X		X	X
TR 1007	Raw synthesis gas	Outlet of Radiation boiler	Thermocouple Pt-PtRh	X		°C		X	X		
TI 1008	Oxygen	Outlet O_2 heater Screw 1 and 2	Thermometer		X	°C	X			X	
TI 1009	Oxygen	Outlet O_2 heater Screw 3 and 4	Thermometer		X	°C	X			X	
TI 1010	Cooling water	Entering Circulation Pump	Thermometer	X		°C	X			X	
TI 1011	Wash water	Entering washer-cooler	Thermometer	X		°C	X			X	
TI 1012	Wash water	Leaving washer-cooler	Thermometer		X	°C	X			X	

GASIFIER	BFD-1001
RADIATION BOILER	BCC-1002
TUBULAR BOILER	BCD-1003
WASHER COOLER	DBG-1001
FINAL COOLER	DBG-1002
THEISEN WASHER	DBF-1003
OXYGEN HEATER	EA-1001/2
COOLING WATER ELEVATED STORAGE TANK	FBB-1001
COOLING WATER COLLECTING TANK	FBB-1002
SERVICE BUNKERS	FSA-1003/4
FEED BUNKERS	FSA-1005/6
START UP FLARE SEAL POT	FKA-1007
FLARE SEAL POT	FKA-1008
NITROGEN SEAL POT	FKA-1009
SEAL POT	FKA-1010/11
QUICK GAS SHUT-OFF	FKA-1012
SCREW FEEDER	RKG-1001÷1004
ASH EXTRACTOR	RKK-1005
BAILEY FEEDER	RNA-1006/7
OXYGEN/STEAM MIXER	UAD-1001÷1004

FIGURE 12 : FLOW SHEET GASIFICATION SECTION

Projekt
Teilanlage
FLIESSBILD DER VERGASUNGSANLAGE

	Datum	Name
Bearb.		
Gepr.	10.2.76	
Norm		

Maßstab
Abt. Blätter
Urspr.:

KRUPP-KOPPERS
Krupp-Koppers GmbH, D-43 Essen 1, W.-Germany

Blatt Rev.
7644BV2V11003
Ers. f.:
Ers. d.:

Figure 3-1. Instrumentation and Sampling

Table 3-2

TEMPERATURE-READING POINTS TI 1013 - TI 1025

		Measuring Point				Indication					
Measuring Station Index	Medium	Location	Type of Instrument	Existing	New	Dimension	Local	Control Room	Recorded	Test Data	Control Data
TI 1013	Cooling Water	Ash extractor outlet	Thermometer		X	°C	X			X	
TR 1014	Raw synthesis gas	Tubular boiler inlet	Thermocouple Pt-PtRh	X		°C		X	X		X
TR 1015	Raw synthesis gas	Tubular boiler outlet	Thermocouple Pt-PtRh	X		°C		X	X		X
TI 1016	Cooling Water	Transfer Pipe outlet	Thermometer	X		°C	X				X
TR 1017	Raw synthesis gas	Outlet of final cooler	Resistance Thermometer		X	°C	X		X	X	
TI 1018	Cooling water	Return to storage	Thermometer		X	°C	X			X	
TI 1019	Not used										
TI 1020	Cooling water	C3 cooling chamber outlet	Thermometer	X		°C	X				X
TI 1021	Cooling water	C4a cooling chamber outlet	Thermometer	X		°C	X				X
TI 1022	Cooling water	C4b cooling chamber outlet	Thermometer	X		°C	X				X
TI 1023	Cooling water	C5 cooling chamber outlet	Thermometer	X		°C	X				X
TI 1024	Cooling water	C2 cooling chamber outlet	Thermometer	X		°C	X				X
TI 1025	Cooling water	C6 cooling chamber outlet	Thermometer	X		°C	X				X

Table 3-3

TEMPERATURE-READING POINTS TI 1026 - TI 1038

		Measuring Point				Indication					
Measuring Station Index	Medium	Location	Type of Instrument	Existing	New	Dimension	Local	Control Room	Recorded	Test Data	Control Data
TI 1026	Cooling water	C1 cooling chamber outlet	Thermometer	X		°C	X				X
TI 1027	Cooling water	C7 cooling chamber outlet	Thermometer	X		°C	X				X
TI 1028	Cooling water	Front cover cooling outlet MS	Thermometer	X		°C	X				X
TI 1029	Cooling water	Jacket steam chamber outlet MS	Thermometer	X		°C	X				X
TI 1030	Cooling water	Inspection hole outlet MS	Thermometer	X		°C	X				X
TI 1031	Cooling water	Burner cooling outlet MS	Thermometer	X		°C	X				X
TI 1032	Cooling water	Inspection hole outlet MS	Thermometer	X		°C	X				X
TI 1033	Cooling water	Front cover cooling outlet KS	Thermometer	X		°C	X				X
TI 1034	Cooling water	Jacket steam chamber outlet KS	Thermometer	X		°C	X				X
TI 1035	Cooling water	Inspection hole outlet KS	Thermometer	X		°C	X				X
TI 1036	Cooling water	Burner cooling outlet KS	Thermometer	X		°C	X				X
TI 1037	Cooling water	Inspection hole outlet KS	Thermometer	X		°C	X				X
TI 1038	Cooling water	Gasifier outlet	Thermometer	X		°C	X				X

Table 3-4

TEMPERATURE-READING POINTS TI 1039 - TI 1043

		Measuring Point				Indication					
Measuring Station Index	Medium	Location	Type of Instrument	Existing	New	Dimension	Local	Control Room	Recorded	Test Data	Control Data
TI 1039	Cooling water	K1 seal leg outlet	Thermometer	X		°C	X				X
TI 1040	Cooling water	K2 seal leg outlet	Thermometer	X		°C	X				X
TI 1041	Cooling water	K3 seal leg outlet	Thermometer	X		°C	X				X
TI 1042A	Ambient air	MS cold junction TI 1006A Thermocouple	Thermometer		X	°C	X			X	
TI 1042B	Ambient air	KS cold junction Thermocouple TI 1006B	Thermometer		X	°C	X			X	
TI 1043	Cooling water	Entering overhead tank	Thermometer	X		°C	X			X	

Table 3-5

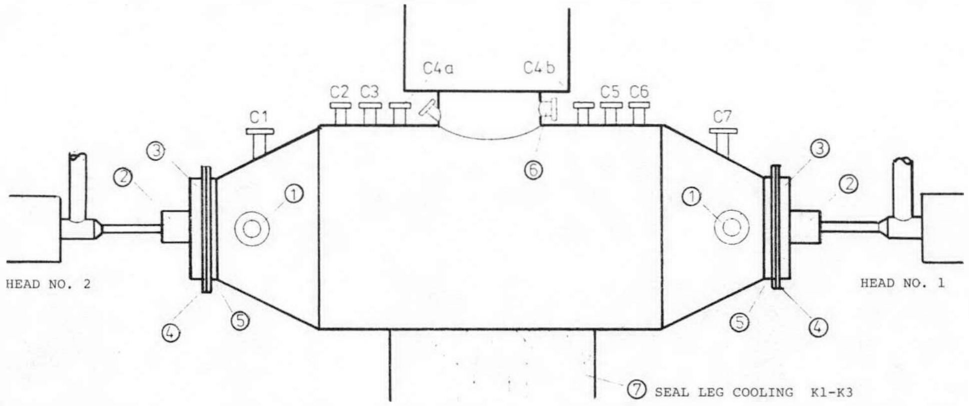
PRESSURE-READING STATIONS PI 1001 - PI 1013

		Measuring Point				Location					
Measuring Station Index	Medium	Location	Type of Instrument	Existing	New	Dimension	Local	Control Room	Recorded	Test Data	Control Data
PI 1001	Oxygen	No.1 orifice inlet	U-tube		X	mmHg		X		X	
PI 1002	Oxygen	No.2 orifice inlet	U-tube		X	mmHg		X		X	
PI 1003	Oxygen	No.3 orifice inlet	U-tube		X	mmHg		X		X	
PI 1004	Oxygen	No.4 orifice inlet	U-tube		X	mmHg		X		X	
PI 1005	Raw synthesis gas	Final cooler outlet	U-tube		X	mmWS	X			X	
PR 1006	Raw synthesis gas	Leaving gasifier	Ring balance	X		mmWS		X	X		X
PI 1007	Raw synthesis gas	Entering tubular boiler	U-tube		X	mmWS	X				X
PI 1008	Raw synthesis gas	Leaving tubular boiler	U-tube		X	mmWS	X				X
PI 1009	Steam	Entering reducing station	Manometer	X		kg/cm^2	X			X	
PI 1010	Raw synthesis gas	Entering washer-cooler	U-tube		X	mmWS	X			X	
PI 1011	Raw synthesis gas	Leaving washer-cooler	U-tube		X	mmWS	X			X	
PI 1012	Raw synthesis gas	Entering Theisen washer	U-tube		X	mmWS	X				X
PI 1013	Raw synthesis gas	Leaving Theisen washer	U-tube		X	mmWS	X				X

Table 3-6

PRESSURE-READING STATIONS PI 1014 - PI 1022

		Measuring Point				Location					
Measuring Station Index	Medium	Location	Type of Instrument	Existing	New	Dimension	Local	Control Room	Recorded	Test Data	Control Data
PI 1014	Coke/O_2/Steam	Blow pipe No.1 burner	Ring balance	X		mmWS		X			X
PI 1015	Coke/O_2/Steam	Blow pipe No.2 burner	Ring balance	X		mmWS		X			X
PI 1016	Coke/O_2/Steam	Blow pipe No.3 burner	Ring balance	X		mmWS		X			X
PI 1017	Coke/O_2/Steam	Blow pipe No.4 burner	Ring balance	X		mmWS		X			X
PI 1018	Raw synthesis gas	Leaving gasifier	U-tube		X	mmWS	X				X
PI 1019	Raw synthesis gas	Leaving orifice after final cooler	U-tube		X	mmWS	X				X
PI 1020	Wash water	Entering washer-cooler	Manometer	X		kg/cm^2	X				X
PI 1021	Steam	Entering O_2/steam blender from orifice	Manometer		X	kg/cm^2	X			X	
PI 1022	Barometer level	-	Barometer	X		mmHg	X			X	



Water Cooled Locations:
1. Sightglass
2. Burner
3. Burner Head
4. Steam Jacket Chamber
5. Head Cover
6. Gasifier Neck
7. Seal Leg
8. Cooling Chambers C1-C7

Figure 3-2. Location of Cooling Water Temperature-Measuring Locations at Gasifier

Flow Measurement

The flow-measuring stations, designated by index FI 1001 to FI 1015, are described in Tables 3-7 and 3-8.

SAMPLING (TESTING) OF SOLIDS

See Figure 2-1, page 2-3 for a diagram of the measuring stations and sampling.

Raw Sample of Char Feed A1001 (not indicated in Figure 2-1)

A sample of 2 kg of each feed material was taken from each truckload. The individual specimens were intermixed to produce an average specimen of 10 kg.

Taken from:	conveyor belt to the raw coal bin
Specimen index:	KLW (truck) specimen

Ground Feedstock

Ground Samples at the Mill A1002 (not indicated in Figure 2-1). The samples served as a control for particle size during the grinding operation and were taken at intervals of 1 to 2 hours.

Taken from: final dust bin in the grinding mill plant

Ground Samples at the Screws A1003. Samples were taken during the entire test period. The average specimen was approximately 2 kg.

Taken from:	dust entering gasifier screw 1 to 4
Specimen index:	01

Table 3-7

FLOW-MEASURING STATIONS FI 1001 - FI 1013

		Measuring Point				Indication					
Measuring Station Index	Medium	Location	Type of Instrument	Existing	New	Dimension	Local	Control Room	Recorded	Test Data	Control Data
FI 1001	Steam	Entering No.1 O_2/ steam blender	Orifice		X	% Flow	X			X	
FI 1002	Steam	Entering No.2 O_2/ steam blender	Orifice		X	% Flow	X			X	
FI 1003	Steam	Entering No.3 O_2/ steam blender	Orifice		X	% Flow	X			X	
FI 1004	Steam	Entering No.4 O_2/ steam blender	Orifice		X	% Flow	X			X	
FR 1005	Raw synthesis gas	Leaving final cooler	Orifice ring balance		X	mmWS	X		X	X	
FR 1006	Oxygen	Entering No.1 O_2/ steam blender	Orifice ring balance	X		% Flow		X	X	X	
FR 1007	Oxygen	Entering No.2 O_2/ steam blender	Orifice ring balance	X		% Flow		X	X	X	
FR 1008	Oxygen	Entering No.3 O_2/ steam blender	Orifice ring balance	X		% Flow		X	X	X	
FR 1009	Oxygen	Entering No.4 O_2/ steam blender	Orifice ring balance	X		% Flow		X	X	X	
FR 1010	Cooling water	Leaving overhead tank	Orifice U-tube		X	mmHg	X			X	
FI 1011	Cooling water	Weir 1 leaving ash extractor	Triangular spillway		X	mm	X			X	
FI 1012	Wash water cooling	Waste water drain	Rectangular spillway		X	mm	X			X	
FI 1013	Quench water	Entering gasifier outlet head 1	Rotameter	X		l/h	X			X	

Table 3-8

FLOW-MEASURING STATIONS FI 1014 - 1015

		Measuring Point				Indication					
Measuring Station Index	Medium	Location	Type of Instrument	Existing	New	Dimension	Local	Control Room	Recorded	Test Data	Control Data
FI 1014	Quench water	Entering gasifier outlet, head 2	Rotameter	X		1/h	X			X	
FI 1015	Oxygen	Entering seal leg	Ring balance	X		Nm^3/h		X		X	X

Slags A1004

An average specimen of about 2 kg was taken from all of the slag collected during a test.

Taken from: Scraper conveyor of slag extractor
Specimen index: 06

Flue Dust Under Tubular Boiler A1005

By plugging the dustcatcher under the tubular boiler the quantity of dust in the dustcatcher was ascertained after a test or series of tests.

Taken from: Dustcatcher under the tubular boiler
Specimen index: 07

Residue in Gasifier A1006

The quantity of residue deposited between burner heads and the center of the gasifier (seal leg) was estimated and a specimen was taken in order to determine C and H content.

Taken from: inside of gasifier
Specimen index: 08

Flue Dust in the Raw Synthesis Gas A1007

Individual specimens were taken during the entire test period by the special method for determining flue dust (see Tables 5-1 through 5-5, pages 5-6 and 5-7).

Taken from: raw gas line before washer-cooler inlet
Specimen index: 02

WATER SAMPLING

Wash Water From Washer-Cooler A1008

Individual samples of 10 liters each were taken during the entire test period.

Taken from: washer-cooler discharge cone outlet
Specimen index for filtered wash water: 03
Specimen index for solids filtered out of wash water: 04

Cooling Water From Slag Extractor A1009

Individual samples of 10 liters each were taken during the entire test period.

Taken from: drain passage from slag extractor
Specimen index for solids filtered from cooling water: 05

GAS SAMPLING

Raw Gas Analysis A1010

Individual specimens were taken during the entire test period for complete Orsat analyses and abbreviated Orsat analyses (CO_2, CO).

Taken from: seal pot after the final cooler

Raw Gas--Full Analysis A1011

A complete analysis (including the determination of H_2O, COS, SO_2, NH_3, HCN) was performed once during each test. An abbreviated form of the analysis (H_2S, COS, SO_2) was made toward the end of each test. The complete analysis was arranged to coincide with dust measurement (see Flue Dust in the Raw Synthesis Gas A1007, page 3-13). During the complete analysis an average specimen was taken for a complete Orsat analysis.

Taken from: raw gas line at tubular boiler outlet

Oxygen A1012

Individual specimens were taken during the entire test period.

Taken from: O_2 line entering gasification plant

Specimen Analyses

All gas analyses were immediately analyzed in the laboratory. The water samples were filtered to eliminate reactions observed earlier between solids and wash water.

Any specimens that could not be evaluated at the test site were hermetically sealed and sent to Essen for study. The individual specimens were identified by test number and appropriate specimen index.

FREQUENCY OF DATA TAKING AND SAMPLING

After the gasifier had been started or adjustment had been made for new test conditions, and a steady operating situation had been achieved, the test director set up a schedule for taking samples and data. Each test began with an Orsat analysis. At the same time, or after 15 minutes, test data and Orsat gas analyses were begun and repeated at half-hour intervals. Samples were taken between every two data-taking operations.

The first wash water specimen was taken 15 minutes after the first data were obtained. The first complete gas analysis and the first dust sampling, was well as the second wash water specimen, were performed 1 to 1 1/2 hours later, depending on when the dust measuring equipment was ready (see Flue Dust Determination, page 5-8). At the end of each test (approximately 2 to 2 1/2 hours after the first data were obtained), the third wash water specimen and the last dust specimen were taken, and the partial gas analysis (H_2S, COS, SO_2) was performed.

From the first Orsat analysis to the last data taking, the individual tests lasted from 1 1/2 to 3 hours.

Section 4

TEST PROCEDURE

TEST PREPARATION

Pretest Grinding

Western Kentucky Char. The flow conditions and equipment adjustments within the grinding plant, although suitable for the normal plant operation with lignite, were not acceptable for the COED chars. Preliminary char-grinding runs gave an unsatisfactory ground particle size distribution.

Tests were made with various adjustments of the classifier helical vanes. The best values were obtained when the vanes beneath the screen were 25 percent open.

Nevertheless, adjustment of the screen alone was not sufficient to obtain the desired ground size analysis. A reduction in the feed rate was essential. Therefore, the height of the feed layer below the feed bunkers was reduced to 65 mm and the drag link conveyor was reduced to its minimum speed, 2.4 cm/sec.

Because the ground char was too dry to form plugs in the screw conveyor to the gasifier, a waterspray was installed over the chain conveyor for controlled spraying of the raw char feed to the ball mill.

As a result of these adjustments, the ground char achieved a suitable particle size of 95 percent <90 micron and a moisture content of 1.6 to 2.3 percent*

Before each grinding operation, the inert gas stream was preheated in the fuel-gas producer to about 60 to 80°C. A temperature of ~60 C set in during the grinding operation.

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*This moisture content was suitable for formation of a satisfactory plug in the feedscrew conveyors.

Pittsburgh Char. The milling operation took place with the lowest entrained gas velocity. During the operation the helical vane in the separator was 25 percent open, the minimum charging height was 65 mm and the minimum chain conveyor velocity was 2.4 cm/sec. The raw char feed was water-sprayed. Under these conditions the ground char achieved a particle size of about 89 percent < 90 micron and went from a raw-material moisture content of about 3.0 percent to a final moisture content of 1.6 to 1.8 percent.

Pretest Gasification

Cold tests with the screw feeders were conducted prior to gasification in order to obtain settings and adjustment for each feed material. The tests showed that the feed materials, which differed considerably from lignite dust in composition, ash content, moisture content, and bulk density, could not be immediately handled by the screw feeders, which had been designed specifically for lignite dust. Good feed conditions in the screw feeders depend on the formation of a sufficiently resistive coal plug (char plug). The plug is formed in the space between the last flight of the screw feeder and the peripheral coal slot. The design of the coal slot and the plug space greatly influence plug formation.

In order to attain a good material plug, the coal slot and plug space were altered as necessary within the limits of the original design. Spacer rings were installed between the screw feeder and the metering head in order to increase the material volume.

The moisture content of the feed material also influences feed control. When raw char was conditioned with water, a suitable moisture content was obtained and feed control was greatly improved.

During the start-up preparations, the oxygen slots were also adjusted to the necessary dimension.

The screw feeders were calibrated, in an attempt to obtain a relationship between feed rate and r.p.m. The actual feed rates, however, were sometimes quite a bit lower than those measured during calibration. This variation was due in part to the pressure increase at the mixing head during operation, but the screws themselves, which had been designed specifically for lignite and had been in operation for over eighteen years, also contributed to the reduced feed rate. A gas flow rate of 4800 Nm^3/hr of dry gas 20 percent below the gasifier design rate was the maximum achievable with char.

MAIN INVESTIGATION

Description of Test Procedures

The most readily available test parameters were the exit gas temperature and the CO_2 level of the gas. Operating at a predetermined exit temperature and associated CO_2 level, the char feed rate and oxygen and steam inputs were determined for steady-state conditions. The gas rate and all necessary data were recorded and samples were taken.

As a layer of slag built up on the thermocouple sheaths during long test runs, a stable CO_2 level became necessary for a steady temperature and steady operating conditions. A CO_2 range of 5 to 20 percent in the raw gas was selected for the test series. A CO_2 content of 20 percent corresponded to a steam addition of 0.3 kg of steam/kg char.

To obtain the best possible operating conditions, the maximum production of raw synthesis gas was striven for. Production was limited only by the original design of the lignite screws, which restricted the char feed rate. By adjustments in the screws the char feed rate was increased and the oxygen flow rate regulated to obtain the desired gasifier exit temperature and associated CO_2 content. The limitation on char feed rate, which restricted full gas flow, adversely affected gas quality, primarily because of the associated percentage increase in heat loss.

After steady-state conditions were achieved, adjustments were avoided, but a correcting control was made for small variations in the feed rates or changes in the oxygen purity. On several occasions the tests were discontinued because of mechanical problems or excessive flow variations.

On the average, a series of three tests was made during a test run, following which the gasification plant was checked and inspected. The gasifier was opened and deposits were removed for use in the mass balance. The dust hoppers beneath the boiler and the connecting duct were also checked for dust that was included as appropriate in the mass balance.

Following a series, the tests were quickly evaluated for guidance in the next test sequence.

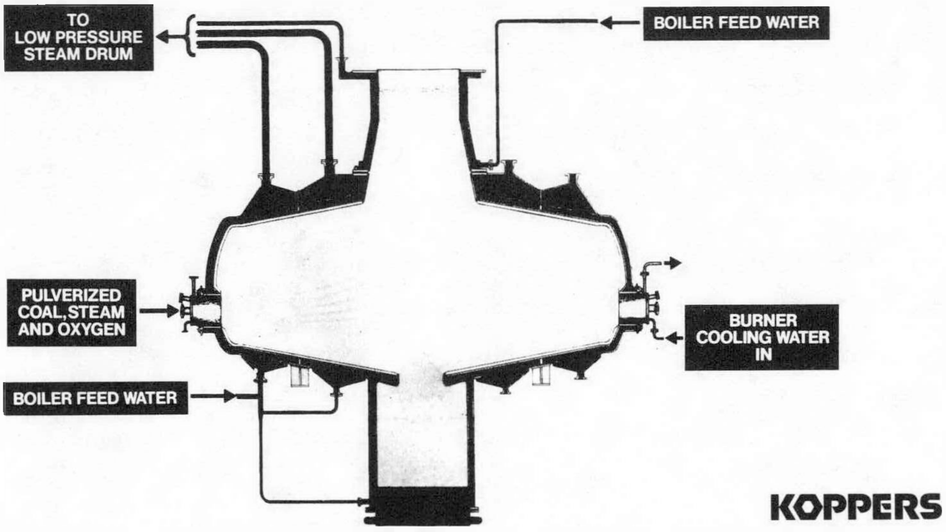

Figure 4-1. Two-Headed K-T Gasifier (Modern Design)

Section 5

EVALUATION

ANALYSIS OF SOLIDS

Raw Material

Water Content: DIN* 51 718 Drying oven process.

The sample was dried at 106 ± 2°C until constant weight.

Accuracy: 0 to 10% water content ± 0.2% absolute

Particle Size Analysis (granulation): DIN 51 704 and DIN 51 033.

The sample was sieved in accordance with DIN 51 704, using a mechanical sieving machine and various sieves sized (DIN 4188) down to 63 microns. The size of particles smaller than 63 microns was determined in accordance with DIN 51 033, using the Andreasen pipette method.

The process was based on the different settling times for single particles of different size in a dispersion medium. The settling time of the particles, which were assumed spherical, was computed on the basis of Stokes's law.

Carbon and Hydrogen: DIN 51 721

The fuel was burned in the air-dried condition in oxygen. The carbon and hydrogen contents were computed according to the weight-determined quantities of the combustion products, carbon dioxide and water.

Accuracy: Carbon ± 0.2%, hydrogen ± 0.1% absolute

Sulfur: DIN 51 724

The total sulfur content was determined by burning the fuel in a stream of oxygen at a temperature of 1300 C, and absorbing the resulting sulfur oxide in a hydrogen peroxide solution, followed by indirect titrimetric determination of the sulfate ions.

Accuracy: ± 0.03% absolute

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*DIN - Deutsche Industrie Norm.

Nitrogen: DIN 51 722

The fuel was decomposed in concentrated sulfuric acid in the presence of a catalytic mixture. The ammonia was expelled from the resulting ammonium sulfate by sodium hydroxide, collected in sulfuric acid, and titrimetrically determined.

Accuracy: 2% relative

Ash Content: DIN 51 719

The sample was heated in an electric muffle furnace to 815°C and then completely ashed at this temperature for at least one hour. The ash content was determined from the combustion residue.

Accuracy: 0.2% absolute

Calorific Value: DIN 51 900

The weighed-out sample was burned in a bomb calorimeter under specified conditions. The resulting rise in the temperature of the calorimeter system was measured. Because of the heat transfer between the calorimeter vessel and the environment, this measured rise in temperature then had to be corrected.

The calorific value of the fuel was computed from the weight of the sample, the corrected temperature rise, and the heat capacity of the calorimeter system.

Accuracy: ± 15 kcal/kg

Reactivity: (Koppers-Jenkner method)

The method is based on the state of equilibrium that sets in between carbon dioxide, carbon monoxide, and carbon under specified working conditions. The sample was degassed at 800°C and then treated at 950°C with carbon dioxide. The reactivity of the resulting gas mixture was computed from the volume percentages of carbon dioxide and carbon monoxide.

Bulk Density

The sample was charged into a measuring receiver from a specified height. The bulk density was then computed from the sample's weight and volume.

Rammed (or tapped) Density: DIN 53 194

One hundred grams of the sample were inserted into the measuring cylinder of the tapping volumeter so that no cavities remained. The measuring cylinder was firmly fixed in the holder of the tapping volumeter and tapped 1250 times. The volume of the sample was read to 1 milliliter.

Specific Gravity, Pycnometer Method

A weighed-out sample was inserted into a pycnometer having a precise volume of 50 cm^3 and then covered with about 30 cm^3 of isopropyl alcohol. The bubbles were removed by careful shaking and boiling. After this initial conditioning operation, the pycnometer was filled to the calibration mark with isopropyl alcohol and weighed.

Ash Investigations

Ash Production. Ashing of the chars was carried out under standard conditions of 800°C in a porcelain or platinum dish.

Ash Analysis. X-ray fluorescence analysis was used to determine ZnO, CuO, NiO, Fe_2O_3, MnO, V_2O_5, CaO, SO_3, TiO_2, SiO_2, and Al_2O_3.

Ash-Fusion Behavior: DIN 51 730

In order to characterize the melting behavior, ash samples of specified dimensions were heated on the high temperature microscope. Observation of form and state changes established the temperatures at which characteristic forms appeared.

Slag Investigations

In the slag investigations, carbon, sulfur, and ash content were determined in the same manner as for the char. Ash analysis was carried out and ash-fusion behavior determined as described under Ash Investigations.

WATER ANALYSIS

Hydrogen-Ion Concentration (pH-value): DEV*C5

The determination of the pH-value was based on the potential difference between electrodes immersed in the water under investigation. One of the electrodes was named the reference electrode, and its potential with respect to the standard hydrogen electrode had a known value. The potential of the measuring electrode was determined in the regular manner, from the pH-value of the water.

Ammonia: DEV E5

Ammonia was analyzed photometrically, as described in the following section on gas analysis.

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*DEV - Deutsche Einheits Vorschrift

Free Cyanide Ion: Published in "Gesundheitsingenieur" 1955, page 374, and 1960, page 248.

Feigl's silver reagent (dimethylaminobenzilidene rhodamine) was used as the indicator for the agentometric cyanide titration. If a silver nitrate solution is added drop by drop to a weakly alkaline cyanide solution, the positive silver ion first bonds with the cyanide ion. When the cyanide is depleted, the excess silver ions react with the silver reagent to form a silver salt with a deep red coloration. The color reaction is very distinct and, as a result, the titration point is easily recognized.

Sensitivity: 0.2 mg CN'.

Total Cyanide

Total cyanide was analyzed by the same method used in the gas analysis.

Hydrogen Sulfide and Sulfur Dioxide: DEV G3 and DEV D15

Again, the same method was used as in the gas investigation.

Range of applicability: concentrations above 2 mg/liter

Total Sulfur: DEV D5

The sample combined with hydrochloric acid was oxidized with bromine water. The sulfates that formed during the reaction were precipitated with barium chloride, filtered off, calcined, and weighed.

Iron: "Photmetrische Metall-Und Wasseranalysen" (Photometric Metal and Water Analyses) Wissenschaftliche Verlagsgesellschaft mbH. Stuttgart.

A red complex was formed by iron ions with thioglycolic acid in ammoniacal solution.

Range of application: 0.03 - 15 mg Fe/l

Nickel: Wasser, Luft und Betrieb 1966 (10)

Red complex salts are formed by combining nickel ions with dimethylglyoxime. The complex can be extracted quantitatively with chloroform. The extinction of the chloroform extract is measured photometrically.

GAS ANALYSIS

Gas Analysis with the Orsat Equipment: Ruhrgas Handbuch

The following gases were determined: CO_2, C_nH_m, O_2, CO, H_2, CH_4, and N_2. A measured volume of gas was brought, in turn, into intimate contact with specific reagents. In each contact, a specific constituent of the gas was removed. The volume difference before and after the reaction yielded the percent fraction of the individual constituents in the total volume.

Accuracy: 0.2 volume % absolute

Organic Sulfur (COS): Ruhrgas Handbuch

Gas from which hydrogen sulfide had been removed was burned in a purified stream of air; thus, any sulfur dioxide that formed originated from organic sulfur. The waste gases were passed through caustic soda solution, as a result of which part of the solution was reacted. The rest was back-titrated with sulfuric acid.

Accuracy: ± 1% relative (experimental value)

Hydrogen Sulfide and Sulfur Dioxide: Ruhrgas Handbuch

The hydrogen sulfide in a measured quantity of gas reacted quantitatively with cadmium acetate to form cadmium sulfide. This was filtered off and reacted with hydrochloric acid, and the hydrogen sulfide was determined iodometrically. The filtrate was treated with acetic acid, and the thiosulfate formed from the reaction between hydrogen sulfide and sulfur dioxide was determined iodometrically. The sulfur dioxide content was calculated from the amount of iodine consumed.

Total Cyanide: Modified Ruhrgas method (Ruhrgas Handbuch)

The gas to be analyzed was passed through an alkali hydroxide, a process in which hydrogen cyanide, hydrogen sulfide, and carbon dioxide were absorbed. When this alkaline absorption solution was acidified with sulfuric acid in the presence of bromine, hydrogen cyanide reacted with bromine to form bromine cyanide. After removing the excess bromine by means of phenol, the bromine cyanide was determined iodometrically.

Accuracy: ± 0.2 mg absolute

Ammonia: DEV E5

The ammonia contained in the gas was collected by reaction with sulfuric acid. The absorbing solution containing the ammonium ions was put in a potassium hydroxide solution with Nesslers reagent (dipotassium-tetraiodomercurate-II), to form a Millon-base salt, the yellow-brown mercury oxide aminoiodide, which was determined photometrically.

Limit of detection: 0.02 mg/liter

Particle Size Distribution and Ash Composition of Flue Dust and Slag

Table 5-1

PARTICLE SIZE ANALYSES FLUE DUST IN RAW GAS BEFORE WASHER-COOLER
WESTERN KENTUCKY CHAR

Feedstock:			Flue Dust Sample From Test			
Western Kentucky Char			019	020	021	022
Particle Size:	> 125 μ	Weight %	0.4	1.2	0.4	2.0
Microns	63	"	6.4	14.8	9.2	15.2
Sieve analysis	< 63	"	93.2	84.0	90.4	82.8
Particle Size:	> 125 μ	"	0.4			2.0
Microns	90	"	2.3			5.0
According to Andreasen	63	"	4.1			10.2
	50	"	1.7			3.8
	40	"	2.5			4.0
	30	"	6.0			23.0
	20	"	43.0			36.0
	10	"	17.0			7.5
	< 10	"	23.0			8.5

Table 5-2

PARTICLE SIZE ANALYSES FLUE DUST IN RAW GAS BEFORE WASHER-COOLER
PITTSBURGH CHAR

Feedstock:			Flue Dust From Test				
Pittsburgh Char			023	024	025	026	027
Particle Size:	> 125	Weight %	2.0	2.4	1.6	2.0	1.2
Microns μ	63	"	17.2	21.2	20.8	20.0	14.0
Sieve analysis	< 63	"	80.0	76.4	77.6	78.0	84.8
Particle Size:	> 125	"	2.0	2.4			1.2
Microns μ	90	"	6.8	8.6			5.8
	63	"	10.4	12.6			8.2
According to Andreasen	50	"	5.0	5.0			6.0
	40	"	8.0	10.0			12.0
	30	"	21.0	23.0			41.8
	20	"	32.3	27.7			14.0
	10	"	8.8	7.1			6.0
	< 10	"	5.7	3.6			5.0

Table 5-3

ASH ANALYSIS IN WEIGHT-PERCENT IN FLUE DUST IN RAW GAS BEFORE WASHER-COOLER WESTERN KENTUCKY CHAR

Sample	ZnO	CuO	NiO	Fe_2O_3	MnO	V_2O_5	K_2O	CaO	SO_3*	TiO_2	SiO_2	Al_2O_3
019/02	0.1	<0.1	<0.1	25.2	0.1	0.2	2.4	2.3	1.6	1.0	45.2	19.5
020/02	0.1	<0.1	<0.1	26.0	0.1	0.2	2.4	2.6	1.5	1.1	43.4	20.9
021/02	0.1	<0.1	<0.1	26.1	0.1	0.2	2.4	2.5	1.5	1.2	43.2	20.2
022/02	0.1	<0.1	<0.1	26.6	0.1	0.1	2.3	2.3	1.6	1.0	44.1	19.8
028/02	0.1	<0.1	<0.1	27.0	0.1	0.1	2.4	2.4	1.7	1.0	42.3	20.0

Table 5-4

ASH ANALYSIS IN WEIGHT-PERCENT IN FLUE DUST IN RAW GAS BEFORE WASHER-COOLER PITTSBURGH CHAR

Sample	ZnO	CuO	NiO	Fe_2O_3	MnO	V_2O_5	K_2O	CaO	SO_3*	TiO_2	SiO_2	Al_2O_3
023/02	<0.1	<0.1	<0.1	30.5	0.1	<0.1	1.8	2.0	0.4	0.9	41.8	19.0
024/02	<0.1	<0.1	<0.1	30.3	0.1	<0.1	1.7	2.0	0.4	0.9	42.2	19.1
025/02	<0.1	<0.1	<0.1	29.1	<0.1	<0.1	1.9	1.7	0.3	0.9	44.2	18.6
026/02	<0.1	<0.1	<0.1	31.0	<0.1	<0.1	1.8	2.0	0.3	1.0	42.4	19.0
027/02	<0.1	<0.1	<0.1	30.6	0.1	<0.1	1.8	2.2	0.5	1.0	41.2	19.4

Table 5-5

ASH ANALYSIS IN WEIGHT-PERCENT IN SOLIDS FROM WASH WATER FROM WASHER-COOLER WESTERN KENTUCKY AND PITTSBURGH CHAR

Sample	ZnO	CuO	NiO	Fe_2O_3	MnO	V_2O_5	K_2O	CaO	SO_3*	TiO_2	SiO_2	Al_2O_3
Western Kentucky												
019/04	0.1	<0.1	<0.1	25.2	0.1	0.1	2.2	2.2	0.7	1.0	45.0	20.8
020/04	0.1	<0.1	<0.1	25.9	0.1	0.1	2.2	2.3	1.0	1.1	43.8	21.4
022/04	<0.1	<0.1	<0.1	26.8	0.1	<0.1	2.1	2.3	1.0	1.0	44.0	20.6
Pittsburgh												
023/04	<0.1	<0.1	<0.1	30.2	<0.1	<0.1	1.6	2.2	0.9	0.9	42.1	19.6
026/04	<0.1	<0.1	<0.1	31.2	0.1	<0.1	1.7	1.9	0.9	0.9	41.3	19.4

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*The SO_3 values are only to be considered as approximate.

Table 5-6

ASH ANALYSIS IN WEIGHT-PERCENT IN SOLIDS FROM WASH WATER FROM SLAG EXTRACTOR
WESTERN KENTUCKY AND PITTSBURGH CHAR

Sample	ZnO	CuO	NiO	Fe_2O_3	MnO	V_2O_5	K_2O	CaO	SO_3*	TiO_2	SiO_2	Al_2O_3
Western Kentucky												
019/05	<0.1	<0.1	<0.1	25.7	0.1	0.2	1.9	2.8	0.3	1.0	43.4	19.1
020/05	<0.1	<0.1	<0.1	24.7	0.1	0.2	2.0	2.5	0.3	1.1	44.4	19.5
021/05	<0.1	<0.1	<0.1	26.1	0.1	0.1	2.0	2.6	0.3	1.2	43.3	22.2
022/05	<0.1	<0.1	<0.1	28.8	0.1	0.1	1.9	2.3	0.3	1.1	43.3	21.4
028/05	<0.1	<0.1	<0.1	26.9	0.1	0.1	1.9	2.4	0.1	1.0	43.7	21.5
Pittsburgh												
023/05	<0.1	<0.1	<0.1	28.4	0.1	0.1	1.6	2.1	0.3	1.0	42.6	21.7
024/05	<0.1	<0.1	<0.1	28.8	<0.1	0.1	1.6	2.0	0.3	1.0	42.2	20.9
025/05	<0.1	<0.1	<0.1	29.7	<0.1	0.1	1.6	2.0	0.2	1.0	41.6	21.1
026/05	<0.1	<0.1	<0.1	29.7	<0.1	0.1	1.6	2.0	0.3	1.0	41.6	21.3
027/05	<0.1	<0.1	<0.1	30.8	<0.1	0.1	1.6	2.1	0.3	1.0	41.0	21.0

Table 5-7

ASH ANALYSIS IN WEIGHT-PERCENT IN SLAG FROM SLAG EXTRACTOR
WESTERN KENTUCKY AND PITTSBURGH CHAR

Sample	ZnO	CuO	NiO	Fe_2O_3	MnO	V_2O_5	K_2O	CaO	SO_3*	TiO_2	SiO_2	Al_2O_3
Western Kentucky												
019/06	<0.1	<0.1	<0.1	12.6	0.1	0.6	0.7	5.0	0.6	0.5	29.0	51.6
020/06	<0.1	<0.1	<0.1	17.2	0.1	0.4	0.9	4.4	0.8	0.6	31.7	43.2
021/06	<0.1	<0.1	<0.1	17.4	0.1	0.3	1.0	4.4	0.8	0.6	33.9	41.6
022/06	<0.1	<0.1	<0.1	23.7	<0.1	0.1	1.3	3.4	1.0	0.8	40.1	28.4
028/06	<0.1	<0.1	<0.1	24.1	0.1	0.1	1.7	3.6	1.0	0.8	39.4	29.1
Pittsburgh												
023/06	<0.1	<0.1	<0.1	20.6	0.1	0.3	1.2	3.8	1.1	0.6	34.5	38.0
024/06	<0.1	<0.1	<0.1	21.4	0.1	0.2	1.4	3.6	1.2	0.6	36.3	35.9
025/06	<0.1	<0.1	<0.1	21.6	0.1	0.2	1.2	3.5	1.0	0.7	37.0	34.3
026/06	<0.1	<0.1	<0.1	21.3	<0.1	0.1	1.4	3.2	0.4	0.7	43.2	27.8
027/06	<0.1	<0.1	<0.1	21.3	0.1	0.1	1.4	3.3	0.9	0.8	43.0	27.8

FLUE DUST DETERMINATION

The gravimetric method described on page 5-11, was employed to determine the dust content. See Figure 5-1 for a schematic drawing of the dust-measuring apparatus.

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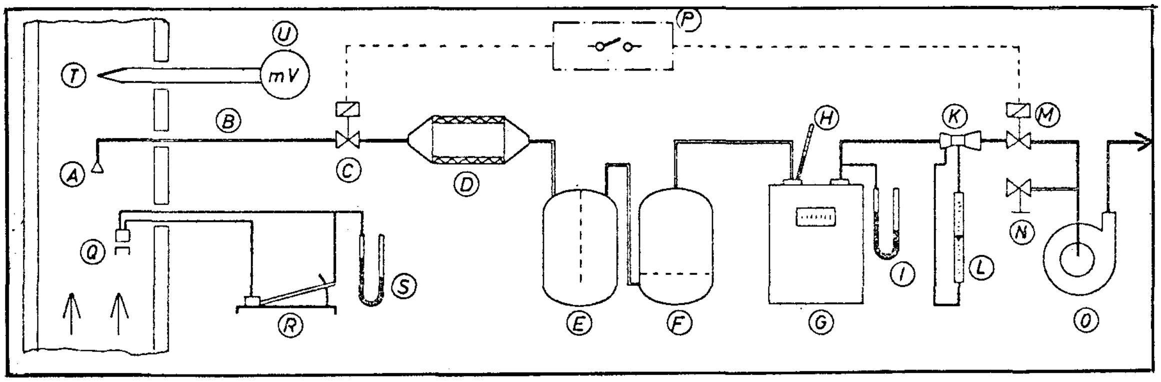
*The SO_3 values are only to be considered as approximate.



A	Probe nozzle
B	Probe
C	Shut-off valve, electrically operated
D	Dust column, electrically heated
E	Cooling column
F	Drying column
G	Flow meter
H	Thermometer
I	U-tube
K	Venturi or orifice plate
L	Rotameter or micromanometer
M	Shut-off valve, electrically operated
N	By-pass valve, manually operated
O	Sample suction fan
P	Controls for C and M with stop watches
Q	Pitot tube
R	Micromanometer
S	U-tube
T	Thermo element
U	Millivoltmeter

Figure 5-1. Dust Content Determination Arrangement of Test Equipment

Measuring Point

All dust content determinations were carried out at one location. The measuring point was located in the vertical part of the pipeline between the tube-boiler outlet and the washer-cooler inlet. The total length of the straight vertical section of the pipeline was 6 m. In order to develop a stable velocity profile, 3 m had to be reserved for the intake length. With a clear pipeline diameter of 490 mm, the intake length corresponded to a length of about 6 x D.

The pipeline ran outside the gasifier building, but the actual measurement connections extended into the building so that the measuring apparatus could be built under cover.

In addition to the dust determinations, this measuring point was used for all other essential measurements, namely:

- Water content determination (dew point measurement)
- Gas-velocity distribution
- Volume-flow measurement

- Temperature measurement
- Pressure measurement

Methods of Measurement

Gas Velocity Distribution. The distribution of gas velocities was measured in a plane. The pressure differential required for computing the gas velocity was determined by means of a Prandtl (or Pitot) tube and a Betz manometer.

The velocity distribution was determined by a grid or network measurement, whereby eight measuring points were established on the one available measuring plane. The measuring points lay on the centroid (gravity) circles of equi-area annula rings.

In the case of the pipeline whose inside diameter measured 490 mm, the following distances were computed between the measuring points and the inside wall of the pipe:

Gravity circle 1	15.7 mm
Gravity circle 2	51.5 mm
Gravity circle 3	95.1 mm
Gravity circle 4	158.3 mm
Gravity circle 5	331.7 mm
Gravity circle 6	394.9 mm
Gravity circle 7	438.6 mm
Gravity circle 8	474.3 mm

In each measurement, each gravity circle in the measuring plane was traversed twice. The static pressure was measured with a water-filled U-tube. The average gas-velocity in the pipeline was computed using the following equation:

$$\bar{v}_k = \sqrt{\Delta\bar{p}} \cdot \sqrt{2g/Q} \qquad (m/s) \qquad (5\text{-}1)^*$$

where $\Delta\bar{p}$ is the root-mean-square value of the pressure-differential.

In order to determine the volume flow Q_k in the gas line, the average gas velocity was multipled by cross-section area A of the pipe:

$$Q_k = A \cdot \bar{v}_k \qquad (m^3/s) \qquad (5\text{-}2)$$

Water Content of the Gas. Before beginning each dust measurement, the water content of the raw gas was determined by the wet and dry bulk thermometer method (psychrometric difference method). The determination is based on the fact that

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*Formula symbols and indices begin on page 5-20.

the cooling produced by evaporation, which occurred when an unsaturated stream of gas flowed over a water-soaked thermometer mercury bulb, is a measure of the degree of saturation of the flowing gas. A U-tube psychrometer, in which was incorporated a dry and a calibrated moistened mercury thermometer, was stationed in the gas stream. Because of the cooling produced by evaporation, the moistened thermometer showed a lower temperature than the dry one. The water content of the gas was determined from the temperature readings with the aid of a chart.

Gas Analyses

The samples for the required gas analyses were obtained by the K-K laboratory staff and analyzed in the Orsat apparatus.

Weighings

Both the empty and the dust-filled thimbles were placed in the drying oven each evening and dried overnight at 115°C. The bags were placed in a desiccator for two hours the next morning and were finally weighed on the same microbalance that was set up in a weighing room.

The weighed empty dust bags were sealed in plastic bags in order to prevent contamination before use. The dust-filled bags were likewise packed in plastic bags.

Temperature Measurement

The temperature of the gas at the measuring point was continuously measured by a thermocouple. The thermocouple voltage was read off on a millivoltmeter.

Pressure Measurement

Water-filled or mercury-filled U-tubes were used for all pressure measurements. The laboratory staff measured the barometric pressure at the beginning of the measurement.

Dust Measurement

In the gravimetric method of dust determination, a side stream had to be taken from the main flow. This stream was obtained using a suction nozzle located in the main stream. The arrangement was such that the velocity of the gas in the suction nozzle was the same as in the main stream. The principle of equi-velocity suction (isokinetic suction) was employed. During the gasification experiments, the suction nozzle was always located at the average velocity point in the main gas stream.

The volume flow V_s to be removed by the suction nozzle (or probe) was computed from the following equation:

$$V_s = 0.471 \cdot \bar{v}_k \cdot d_s^2 \quad \text{(Liters/min)} \qquad (5\text{-}3)$$

The diameter d_s of the probe was computed from the following expression:

$$d_s = 15.35 \sqrt{V_N \cdot k/\bar{v}_k \cdot st} \quad \text{(mm)} \qquad (5\text{-}4)$$

The volume flow computed from equation 5-4 had to be removed by the *probe* in the pipeline. At the gas meter where the volume flow could be measured, however, conditions were quite different from those in the main stream. Thus, a calculation based on conditions near the gas meter became necessary.
The following equation was used:

$$V_u = [0.3592\ V_s\ P_k/T_k - W_o/0.804]\ 2.782\ T_u/P_u \quad \text{(l/min)} \qquad (5\text{-}5)$$

The partial gas stream was cooled before arrival at the gas meter and dried with calcium chloride. Therefore the volume required at the gas meter is on a dry basis.

Because the increase in resistance in the dust collector caused the pressure P_u at the gas meter to vary during the course of a dust measurement, the previously computed volume flow had to be continuously regulated. The constancy of the suction was monitored with the aid of an aperture diaphragm and a differential pressure measuring device and was adjusted by modifying the rotational speed of the suction fan. Pre-prepared calculation sheets were used to calculate the necessary data.

Measurement Difficulties

The following measurements were carried out after constant operating conditions were attained in the installation:

- Determination of water content
- Gas velocity distribution
- Pressure temperature

- Dust measurements; several if possible

Difficulties were encountered in carrying out all of these measurements.

Determination of Water Content. Only a small volume flow is required for this measurement. The gas cools so much in the required hose connection from the gas line to the U-tube psychrometer that the water condenses out. In order to prevent the gas from falling below the dew point, a heated Ströhlein apparatus was connected between the gas line and the measuring apparatus. The other parts of the hose-connecting members were wrapped in heating tapes and additionally protected against the wind. The Ströhlein apparatus was fitted with a filter bag and thus simultaneously kept the dust out of the U-tube psychrometer.

Gas Velocity Distribution. Because of the high dust content of the gas, the holes in the Prandtl (or Pitot) tube rapidly became clogged. Therefore a flushing device was installed that made it possible to blow the Prandtl tube clean with nitrogen.

Because the volume flow fluctuated in the gas line, each gravity circle was traversed three or four times; the root-mean-square value was computed from all the individual test points.

Dust Measurements. Several difficulties had to be overcome during the actual dust measurements, that is, when the gas was drawn off by suction from the partial gas stream.

The suction nozzle became clogged with dust in the periods between the gasifying tests. As a result it was sometimes necessary to blow the suction probe clear via a shut-off valve connected to a nitrogen bottle.

Dust also settled in the suction probe while gas was being drawn off by suction from the partial gas stream. Because the suction probe could not be removed from the assembly, the dust could not be removed separately between two dust measurements and had to be added to the dust collected in the filter bag. The suction probe could not be removed for the following reasons:

- The gas is combustible and poisonous and thus would threaten the safety of the personnel.
- The shut-off slide valve mounted on the measurement connector (NW 150), which should serve as a lock (or charging value) in association with the measuring flange, became clogged with dust and could not be actuated.

- The time between taking two samples was too short.

As a result, the following procedure was adopted. The suction probe was blown clear with compressed nitrogen immediately before the beginning of each dust-sampling operation. When the plant was out of operation over extended periods of time—that is, when the entire system was out of commission and opened up—the probe was carefully removed and the dust was taken separately from the probe, dried, and weighed. A correction factor was then calculated for all dust measurements, using data from the last dust measurement prior to removal of the probe (see also the discussion under Dust Content, below).

Because of the extremely high dew point of the gas, the wash bottles and cooling and drying towers originally provided did not adequately provide dry gas at the gas meter. As a result, a cooling tank fitted with a water-cooled spiral tube was built and installed ahead of the gas meter.

Measured Values

The following table indicates which test data were recorded during the individual measurements:

Measurement Test Data	Determination of Water Content (Dew point)	Gas Velocity Measurement	Dust Measurement
Barometric Pressure	x	x	x
Gas analysis	x		
Clock time	x	x	x
Temperature, duct		x	x
Temperature, gas meter			x
Pressure, channel		x	x
Pressure, gas meter			x
Pressure, psychrometer	x		
Diff. Pressure, Pitot tube		x	
Diff. Pressure, orifice			x
Dust weighing			x

Evaluation

Gas Velocity. The average gas velocity in the main pipeline was computed in accordance with equation 5-1. The operating density Q must be introduced into this equation. The resulting equation is:

$$Q = 0.3592\ Q_{nf} \cdot P_k/T_k \qquad (5\text{-}6)$$

Water Content of the Gas. The water content of the gas or the dew point was computed in accordance with standard methods.

Dust Content. The dust contents were computed on *measured* values. As a result of several tests, it was established that dust deposited in the suction probe (see Dust Measurement, page 5-11). The amount of deposited dust taken from the probe was between 14.7 and 15.4 percent of the amount of dust in the thimble.

Thus, 15 percent was added to each computed value of dust content in order to obtain the true dust content of the raw gas.

Several dust content determinations were made during a gasification experiment. If the number of random samples was three or more, they were examined by a quality control technique. The evaluation procedure was based on the assumption of a normal distribution among the individual values. The assumption is admissible because the operating conditions were not appreciably modified during a gasification experiment.

If a small number of random samples was used to obtain a confidence level s = 98 percent, the test quantities and the limiting values for the sampling criterion were determined by the Graf-Hanning-Stange method.

After the quality test, the arithmetic mean was obtained from the current dust samples used in a gasifying experiment, and finally, the standard deviation was determined.

Results

The results of the dust measurements made in all the gasifying experiments are listed in Tables 5-8 and 5-9.

Column 1: Experiment number. The numbering of the dust measurements conforms with the number of the gasification experiment. The separate dust samples taken during an experiment are indicated by decimals.

Column 2: Date of the dust measurement.

Column 3: Clock time. This column lists the beginning and ending times for taking the individual samples.

Column 4: Duct temperature in °C. The temperature obtained during the taking of a dust sample is listed.

Column 5: Duct pressure in mm water column. This column lists the pressure in the main gas duct during the taking of a dust sample.

Column 6: Water content in g/$m^3{}_n$ dry*. The water content of the synthesis gas was determined before each dust measurement was begun.

Column 7: Dew point in °C. Dew point associated with the water content (column 6) is listed.

Column 8: Dust content in g/$m^3{}_n$ dry. The complete dust content is listed.

Column 9: Dust content, average value ($\bar{x} \pm s$) in g/$m^3{}_n$ dry. The mean value $\bar{x}$ is computed from the dust contents of the individual dust samples. In addition, the standard deviation s is listed.

Column 10: Empty

Column 11: Remarks. Basically, all the dust samplings are listed in the tables. Measurements that cannot be evaluated and broken-off measurements are noted in this column.

Column 12: Types of char. The types of char that were gasified during the research are listed here. The meanings of the symbols are as follows:

WK - Western Kentucky
PC - Pittsburgh Char

Columns 13 to 19: These columns contain data that were not evaluated but that are listed insofar as they were available for the sake of completeness.

Column 13: Volume flow in $m^3{}_n$ dry/h. This is the average volume flow of synthesis gas during the gasification experiment.

Column 14: Added steam in kg/h. Addition of steam to the gasifier is listed.

Column 15: O_2 quantity in $m^3{}_n$ dry/h. Volume-flow of oxygen blown into the gasifier is listed.

Columns 16 to 19: Gas analyses. The gas analyses listed in these columns were used as a basis for the calculations required for the dust measurements. The gas analyses were carried out before the dust measurements were begun.

Column 20: Density ϱ in kg/$m^3{}_n$. This column lists density of synthesis gas computed from the gas analyses (Columns 16 to 19).

Measurement Errors

Many factors contribute to the errors in measurement that can arise during the determination of the dust content in flowing gases. Because the result of the measurement, namely, the dust content, is a function of several partial results (such as, for example, volume flow and density), which are themselves functions

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*($m^3{}_n$ dry = Nm^3 dry)

Table 5-8

RESULTS OF THE DUST-CONTENT DETERMINATION EXPERIMENT 019-025

1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
DUST MEASUREMENT															GAS ANALYSES				
TEST NO.	DATE	CLOCK TIME	DUST TEMP. °C	DUCT PRESS. mm WS	WATER CONTENT g/m$^3_{ntr}$	DEW POINT °C	DUST CONTENT g/m$^3_{ntr}$	AVERAGE DUST CONTENT g/m$^3_{ntr}$		REMARKS	CHAR TYPE	VOLUME FLOW m$^3_{ntr}$/h	ADDED STEAM kg/h	QUANTITY O$_2$ m$^3_{ntr}$/h	CO$_2$ Vol.%	CO Vol.%	H$_2$ Vol.%	N$_2$ Vol.%	DENSITY kg/m^3_n
19.1	15.8.75	22.38-22.44	215	140	343	69.3	61.7				WK	4077	221	1797	12.0	68.2	13.4	6.4	1.1818
19.2		22.49-22.53	225	175			74.3												
19.3		23.05-23.11	226	195			64.6	66.8 ± 6.6											
20.1	18.8.75	-	239	147	419	72.4	-			Not computable	WK	4429	497	1914	16.4	60.3	15.7	7.6	1.1871
20.2		20.40-20.44	239	141			86.3												
20.3		20.50-20.54	239	139			72.4												
20.4		21.10-21.13	240	145			119.0												
20.5		21.18-21.19	239	145			76.0												
20.6		21.25-21.28	242	140			115.8	93.9 ± 22.1											
21.1	18.8.75	23.21-23.24	232	244	352	69.7	141.0				WK	4682	247.5	1863	10.2	65.0	17.6	7.2	1.1205
21.2		23.30-23.32	232	235			118.4												
21.3		23.55-23.57	230	235			131.4												
21.4		23.58-24.00	232	235			155.7	137.5 ± 14.4											
22.1	23.8.75	15.41-15.44	214	220	432	72.9	132.6				WK				12.6	66.1	15.6	5.7	1.1606
22.2		15.56-15.58	215	125			-			Not computable									
22.3		16.10-16.13	215	120			122.2												
22.4		16.28-16.31	214	120			119.4	124.7 ± 7.0											
23.1	20.8.75	18.00-18.04	231	165	558	76.7	82.9				PC				14.1	57.3	14.1	9.0	1.2289
23.2		18.12-18.15	230	183			91.5												
23.3		18.30-18.33	227	180			88.3												
23.4		18.45-18.48	227	195			88.9	87.8 ± 3.6											
24.1	20.8.75	21.00-21.02	219	200	443	73.1	145.5				PC				13.2	64.8	15.5	6.5	1.1666
24.2		21.15-21.17	220	215			140.2												
24.3		21.26-21.28	220	215			129.0												
24.4		21.40-21.42	218	215			128.5												
24.5		21.45-21.47	218	225			134.2	135.5 ± 7.3											
25.1	21.8.75	00.15-00.17	218	300	385	71.1	144.3				PC				7.8	69.6	17.0	5.6	1.1095
25.2		00.28-00.30	221	295			145.1												
25.3		00.47-00.49	221	307			142.6												
25.4		1.00-1.02	221	299			143.1	143.5 ± 1.2											

Data not completed in columns 13, 14, & 15 available in Tables 5-10, 5-11, 5-16, & 5-17.

Table 5-9

RESULTS OF THE DUST-CONTENT DETERMINATION EXPERIMENT 026-028

1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
DUST MEASUREMENT															GAS ANALYSES				
TEST NO.	DATE	CLOCK TIME	DUCT TEMP. °C	DUCT PRESS. mm WS	WATER CONTENT g/m^3_{ntr}	DEW POINT °C	DUST CONTENT g/m^3_{ntr}	AVERAGE DUST CONTENT g/m^3_{ntr}		REMARKS	CHAR TYPE	VOLUME FLOW m^3_{ntr}/h	ADDED STEAM kg/h	QUANTITY O_2 m^3_{ntr}/h	CO_2 Vol.%	CO Vol.%	H_2 Vol.%	N_2 Vol.%	DENSITY kg/m^3_n
26.1	21.8.75	13.15-13.18	215	180	370	70.5	116.5			Not computable	PC				10.4	65.6	18.4	5.8	1.1151
26.2		13.25-13.27	215	183			-												
26.3		13.44-13.46	215	185			123.0												
26.4		13.58-14.00	215	191			127.8	122.4 ± 5.8											
27.1	21.8.75	16.30-16.32	215	185	369	70.4	107.7				PC				10.8	66.9	17.1	5.2	1.1302
27.2		16.45-16.47	215	175			90.3												
27.3		17.00-17.02	218	189			86.0												
27.4		17.15-17.17	218	195			94.4												
27.5		17.30-17.32	215	185			88.9												
27.6		17.45-17.47	-	-			-			Not computable									
27.7		17.49-17.51	217	185			91.7	93.2 ± 7.7											
28.1	23.8.75	18.30-18.32	228	180	455	73.7	60.6				WK				20.5	58.0	16.1	5.4	1.2123
28.2		18.45-18.48	230	75			60.0												
28.3		19.00-19.02	226	110			60.0												
28.4		19.15-19.18	228	180			57.3	59.5 ± 1.5											

(such as temperature and pressure), the uncertainty in the measurement must be calculated in accordance with the laws of error propagation. Because of the spatial variations and the variations in time that occur in a dust determination, the standard deviations can only be determined approximately.

The possibilities of readily observable errors were considered individually for the following measurements:

Gas-Density Measurements. Errors made in determining the density of the gas were influenced by errors made in determining the composition of the gas, that is, in the Orsat analysis, as well as in the temperature and pressure measurements.

Water-Content Determination. If care is used in determining the moisture content of the gas by means of the U-tube psychrometer, errors are small. Errors, however, can occur during the calculation, because, for example, the tabulated values that enter into the calculation (such as saturation pressure and heat content) were interpolated linearly.

Gas Temperature Determination. Thermocouples were employed to measure the temperature. Allowance must be made for the effects of the leads, the cold junction temperature, and instrument errors in correcting the thermal emf's for departures from the measured values. The measured value of the gas temperature is relatively small compared with the absolute temperature that enters into the calculation. Because of this, the effect of a measurement error is small.

Gas-Pressure Determination. Errors arise in reading the barometric pressure and in measurement of the pressure of the gas stream. The gas pressure was measured with (water-filled) U-tubes; its value is small compared with the barometric pressure. Consequently, the influence of a measurement error on the value of the absolute gas pressure is small.

Gas-Velocity Determination.

> Errors Due to the Method of Measurement. The gas velocity in the main gas duct was measured with a Pitot tube. This method requires a determination of the velocity profile over the entire plane of measurement, but only the local velocity can be measured with one Pitot tube. While it is true that taking several separate measurements over a point grid makes it easier to draw conclusions concerning the spatial velocity profile, the determination is still affected with an unavoidable degree of uncertainty. The more irregular the velocity profile and the smaller the number of points at which measurements are made, the greater the uncertainty.

Errors Due to the Pitot Tube. The equipment error associated with the Pitot tube was given in the form of a correction coefficient. This coefficient is equal to 1 with the Prandtl (Pitot) tube employed, when the axis of the probe head does not deviate by more than ±5° from the direction of volume flow.

Errors During the Measurement of Dynamic Pressure. A Betz manometer was employed to measure the dynamic pressure. Errors are small if the apparatus is set up horizontally, if it is filled with bubble-free liquid, and if the zero is properly set.

Determination of Dust Content.

Errors Due to the Measuring Procedure. The actual average dust content can only be determined by the integration of instantaneous dust content over the entire plane of measurement and over the suction-sampling time. This procedure, however, is not possible in practice, since a suction sample can be taken at only one point (at the point of average gas velocity). Thus, a systematic error must be expected. This error becomes smaller if, as in the present case, the measurements are carried out in a vertical pipeline.

Errors During the Measurement of Partial Gas Flow. A gas meter whose readings were checked for accuracy was used to measure the partial gas flow.

Errors During Suction. During suction of the partial gas stream errors may arise as the result of the adjustments required to obtain a constant suction rate. Too high a suction rate leads to too low a dust content, and vice versa.

The aspirated dust can be deposited almost completely if a satisfactory dust-trapping apparatus and sampling filter material are used. Errors arise when allowance is not made for dust that had deposited in the Pitot tube. If the filter material is carefully removed and handled, the error becomes so small that it can be ignored.

Total Error. A count of the various error possibilities—especially in association with the various computation processes—shows that an error-propagation calculation would be quite time-consuming, especially when many of the individual errors must be estimated.

Previous measurements indicate that the total error in a case such as this, is approximately ± 10 percent.

Formula Symbols and Subscripts. The following symbols, quantities, and subscripts are used in this report:

Quantity	Units	Designation
A	m^2	area of cross-section
P	-	absolute pressure
Q	m^3/s; m^3/h	volume flow in main gas stream

Quantity	Units	Designation
S_i	mass/volume	dust content
S	kg/h	dust discharge
T	K	absolute temperature
V	m^3/s; 1/min	volume · flow
W_o	kg/m^3_n	water content
d	mm	diameter of suction nozzle
g	m/s^2	acceleration under gravity
m	mg	mass of trapped dust
p	mmWS(-WC); mm Hg	pressure
t	°C	temperature
v	m/s	gas velocity
β	-	Pitot tube coefficient
ρ	kg/m^3	density
τ	°C	dew point

Subscript	Designation
B	gas under operating conditions
Ges	total pressure
N	rated gas quantity for the dust-measuring apparatus
U	gas meter
b	atmospheric pressure; barometer
f	moist
k	main gas duct
n	normal state, 0°C, 760 torr, dry
s	suction nozzle
st	Ströhlein apparatus (15 m^3/h apparatus)
stat	static (pressure)
tr	dry

EVALUATION OF ALL MEASUREMENT AND ANALYSIS DATA

(Example: Pittsburgh Char experiment 027)

The evaluation includes the calculation of process data from measurement data, the setting up of characteristic values for gasification, and the checking of the results by setting up balances.

The evaluation is divided into the following operations:

- Computing volume flows from test data
- Computing mass flows from test data
- Computing errors
- Computing mass-flow of elements
- Setting up mass balances
- Setting up heat balances
- Obtaining characteristic data for the gasification process

Altogether, 10 experiments were carried out for different time periods and with different parameters. From the test data contained in the experiment records, an average value was derived from each evaluation, and the averages were used as a basis for the subsequent calculations. Because the mass flow of the raw (charged-in) char could not be measured (see Char Feed Rate in Section 1, page 1-6), the mass flow was computed theoretically from a carbon balance on streams exiting the gasifier. A preliminary mass balance was established from the computed individual mass flows.

The computation was carried out for each individual experiment in the manner described above. As a result, each evaluated experiment forms a self-enclosed unit. This form of calculation ensures that, in the event of unexpected results, the quantities used in the calculation and the method of calculation can be traced immediately and the necessary inferences drawn.

In some cases where test data could not be obtained because of an apparatus failure, the data could be estimated by obtaining an average from a corresponding test series.

The experimental data for the gas and water analyses were derived by chemical analysis. The methods employed are described in the first part of Section 5.

The amount of SO_2 determined in the wash water is larger than the amount of SO_2 present in the raw gas before the washer-cooler inlet. This difference can be attributed to the fact that in the SO_2 analysis we assumed that the total measured sulfur came from SO_2 compounds. Because this is clearly not the case, a recalculation is necessary.

The raw gas is analyzed after it leaves the final cooler, whereas the H_2S, HCN, SO_2, and COS are analyzed before entry into the washer-cooler. In order to allow for the composition changes in the washer-cooler and in order to obtain more accurate volume and mass flows, a short computer program was set up to adjust the data to a consistent basis. A brief description of the computer program is given in the following section on volume flows and measurement data.

With the aid of the error calculations, the average error was computed from the variations of the experimental data and within a series of measurements, from the reading errors, and from the apparatus errors. The average error for each mass flow in the provisional mass balance is used for weighting and adjusting the mass balance.

Because of the considerable amount of computation required, a complete calculation of the error was carried out for only a few experiments. For the remaining experiments, the average error values in the mass flow were employed.

The provisional mass balance was corrected in accordance with the error correction rules, where the average error of each individual mass flow is a measure for weighting the balance. The mass balance was adjusted in accordance with the weighting.

There are two methods of adjusting the balance. With the first method, the assumption is made that the carbon balance should remain unchanged, which causes all mass flows containing carbon to remain unchanged. The provisional mass balance is then adjusted accordingly. With the second method, the total balance is corrected and a new mass balance is set up using the corrected values. For the char investigations, only the first method was used.

In the following evaluation of the data, the formulas used in carrying out the computations are given for each computed point. The symbols are also given, along with the dimensions.

Calculation of Volume Flows and Measurement Data

Volumetric Flow Rate of Combined Effluents After Washer-Cooler Weir Measurements.*

$$\dot{V}_{WW} = \frac{2}{3} \cdot \mu \cdot b \cdot h \cdot \sqrt{2 \cdot g \cdot h} \qquad [m^3/s]$$

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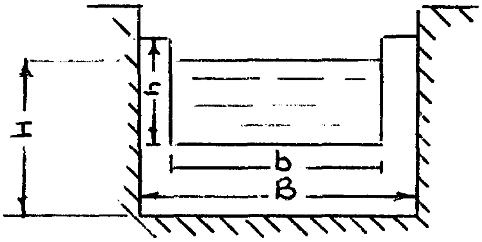
*See W. Kalide "Einführung in die Technische Strömungslehre."

$$\mu = \left[0.578 + 0.037\cdot\left(\frac{b}{B}\right)^2 + \frac{3.615 - 3\left(\frac{b}{B}\right)^2}{1000\cdot h + 1.6}\right]\cdot\left[1 + 0.5\cdot\left(\frac{b}{B}\right)^4\cdot\left(\frac{h}{H}\right)^2\right]$$



μ = contraction coefficient

b = width of rectangular cross-section of weir [m]

B = width of channel [m]

h = head of water in the rectangular cross-section [m]

H = total head (fall) of water [m]

For the measuring weir employed, we have:

$b = 0.300$ m

$B = 0.410$ m

$H - h = 0.250$ m

Head of water in rectangular weir	h: 98.2 mm
Amount of wash water	$\dot{V} = 61.34\ m^3/h$

Flow Rate of Effluent From Slag Extractor: Triangular Weir Measurement 80° - Weir.*

$$\dot{V}_{W,ASA} = \frac{8}{15}\cdot\mu\cdot\tan\frac{\alpha}{2}\cdot h^2\sqrt{2gh} \qquad [m^3/s]$$

α = angle of triangle

h = head of water in triangle [m]

μ = contraction coefficient = $0.565 + 0.0087\ h^{-\frac{1}{2}}$

Head of water in the triangle h: 131.8 mm

Ash discharge, wash water $\dot{V}_{W,ASA} = 26.53\ m^3/h$

Flow Rate of Washer-Cooler Effluent.

$$\dot{V}_{KW} = \dot{V}_{WW} - \dot{V}_{W,ASA} \qquad [m^3/h]$$

Total wash water, after washer-cooler	$\dot{V}_{WW}$: 61.34 m^3/h
Wash water, after slag discharge	$\dot{V}_{W,ASA}$: 26.53 m^3/h
Wash water in washer-cooler	$\dot{V}_{KW}$ = $\underline{34.81\ m^3/h}$

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*See THE COED PROCESS in Section 1.

Flow Rate of Quenching Water.

$$\dot{V}_Q = \dot{V}_{Q_1} + \dot{V}_{Q_2} \qquad [l/h]$$

- Measured Volume Flow

Volume flow of quenching water 1 $\dot{V}_{Q_1}$: 891.7 l/h

Volume flow of quenching water 2 $\dot{V}_{Q_2}$: 933.3 l/h

- Corrected Volume Flow

$$\dot{V}_{Q_z} = \beta_z \cdot \dot{V}_{Q_z} \qquad [m^3/h]$$

β_1 = 1.07 correction factor, rotameter 1

β_2 = 1.024 correction factor, rotameter 2

Volume flow of quenching water 1 $\dot{V}_{Q_1}$ = 954.1 l/h

Volume flow of quenching water 2 $\dot{V}_{Q_2}$ = 955.7 l/h

Total volume flow of quenching water $\dot{V}_Q$ = $\underline{1909.8}$ l/h

Flow Rate of Cooling Water for the Gasifier Jacket.

$$\dot{V}_{KW,V} = \dot{V}_{total} - \dot{V}_{RL} \qquad [m^3/h]$$

$$\dot{V}_{total} = K_2\sqrt{\Delta p \text{ [mm Hg]}} \qquad [m^3/h]$$

K_2 = 16.902 Diaphragm factor

$\dot{V}_{RL}$ = 3.4 m^3h cooling water for cooling the raw gas line

Pressure difference at measuring diaphragm Δp: 203.0 mm Hg

Total volume flow $\dot{V}_{total}$ = 240.8 m^3/h

Cooling water for raw gas line $\dot{V}_{RL}$ = 3.4 m^3/h

Volume flow of cooling water for gasifier jacket $\dot{V}_{KW,V}$ = 237.4 m^3/h

A comparison between the quench water sprayed into the gasifier outlet and the measured water content in the raw gas before entry into the washer-cooler yielded unexpected differences. These differences probably arose from the poor readability and inaccuracy of the flow apparatus for the quenching water. Because of this error, the water content of the raw gas was computed theoretically.

No additional correction was undertaken for the heat balance. Instead, the heat was balanced via the "lost" amounts.

All quantities of importance in the gasifying process are listed in tables according to the raw (charged-in) char.

Following is a list of all symbols and signs used in the analytical work in this section.

Symbols and Signs

Physical quantities

$\dot{V}$	m^3/h	volume flow
$\dot{m}$	kg/h	mass flow
ν	vol. %	volume fraction
w	weight %	mass fraction
n	mole %	mole fraction
p	mm Hg, bar	pressure
B	mm Hg	barometric pressure
R	kJ/kg°K	gas constant
c_p	kJ/kg°K	specific heat
t	°C	temperature
Δh	kJ/kg	specific enthalpy
ΔH	kJ/sec	enthalpy
Q	kJ/sec	total heat
H	kJ/sec	chemically useful heat
q	kJ/Nm^3	specific heat loss
kp	-	equilibrium constant
w'	mg/l	traces in wash water
w"	mg/Nm^3_{tr}	traces in raw gas
w^0	mg/l	traces in water
γ	kg/m^3	specific gravity
τ	°C	dew point
η	-	efficiency
ε	%	extent of steam decomposition
ξ	%	volumetric efficiency, packing factor

Material-identification symbols (as subscripts)

K	char
RG	raw gas
D	steam
SD	saturated steam
VS	combustion oxygen

FLST	flue dust
AS	ash
SL	slag
RV	residue, gasifier
KW	cooling water
WW	wash water
Q	quenching water

Plant parts (as subscripts)

ASA	ash discharge
KW	cooler-scrubber (or washer-cooler)
BL	diaphragm
Erh	O_2-heater
RK	tube boiler
V	gasifier

States (as subscripts)

N	normal state
tr	dry state
f	moist state
wf	water-free
Ein	inlet-entry
Aus	outlet, exhaust
Ges	total
Verlust	loss
B	Boudouard
W	water-gas
th	thermal

Flow Rate of Raw Gas. This is a preliminary calculation, for final value, refer to Corrected Mass Balance, page 5-44.

Calculation of Volumetric Flow Rate of Raw Gas Considering Absorption in the Washer-Cooler. A short computer program is used for the preliminary calculation of the volumetric flow rate of the raw gas. The input data to this program include the complete raw gas analysis before the washer-cooler inlet, measured absorption of H_2S, SO_2, and HCN, and the theoretically calculated absorption of CO_2 in the washer-cooler, as well as the pressure drop across the orifice meter in the raw gas line leaving the final cooler and the pressure and temperature at that location.

The amount of SO_2 in the wash water, as calculated from the water analysis, exceeds the amount found in the raw gas to the washer-cooler, because the wash water analysis was based on the assumption that all the iodine consumed in this analysis had reacted with sulfate.

The calculation for SO_2 absorption takes into consideration the fact that the iodine reacts in part with this sulfate and in part with sulfites. The exact distribution is based on the measured total sulfur, less the thiocyanate.

The computer program first calculates the raw gas volumetric and mass flow rates after the final cooler, based on the gas analysis before the washer-cooler inlet. The sum of the mass flow rates of the absorbed components is added to the mass flow rate. On the basis of the gas analysis before the washer-cooler inlet, a mass flow rate is obtained for each gas component before the washer-cooler inlet. Then the mass flow rates of the absorbed gas components are subtracted. The new mass flow rate is converted into parts by volume and is used to calculate a new volumetric flow rate. When the new and the old volumetric flow rates are equal, the iterative calculation is stopped and the result is printed. The result yields the mass flow rates of the raw gas components before the washer-cooler and after the final cooler.

The flow diagram around the washer-cooler is shown in Figure 5-2. The computer flow chart is shown in Figure 5-3.

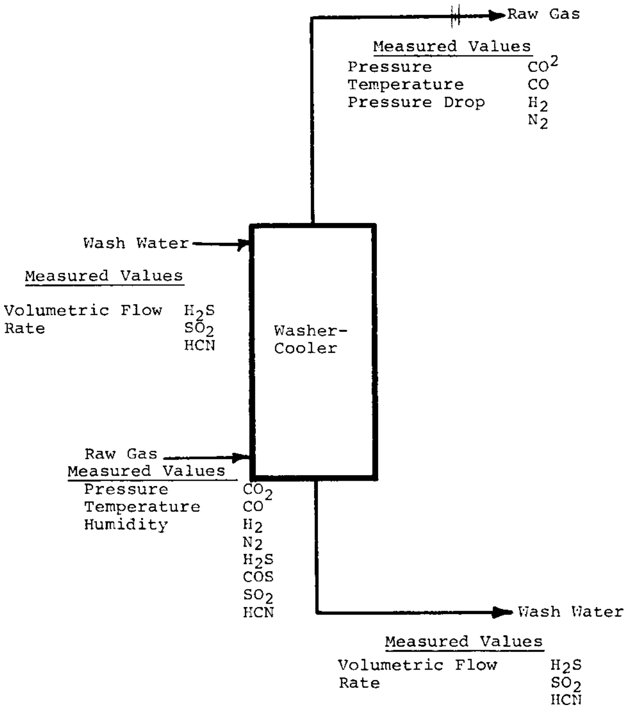

Figure 5-2. Analyses Performed for Washer-Cooler

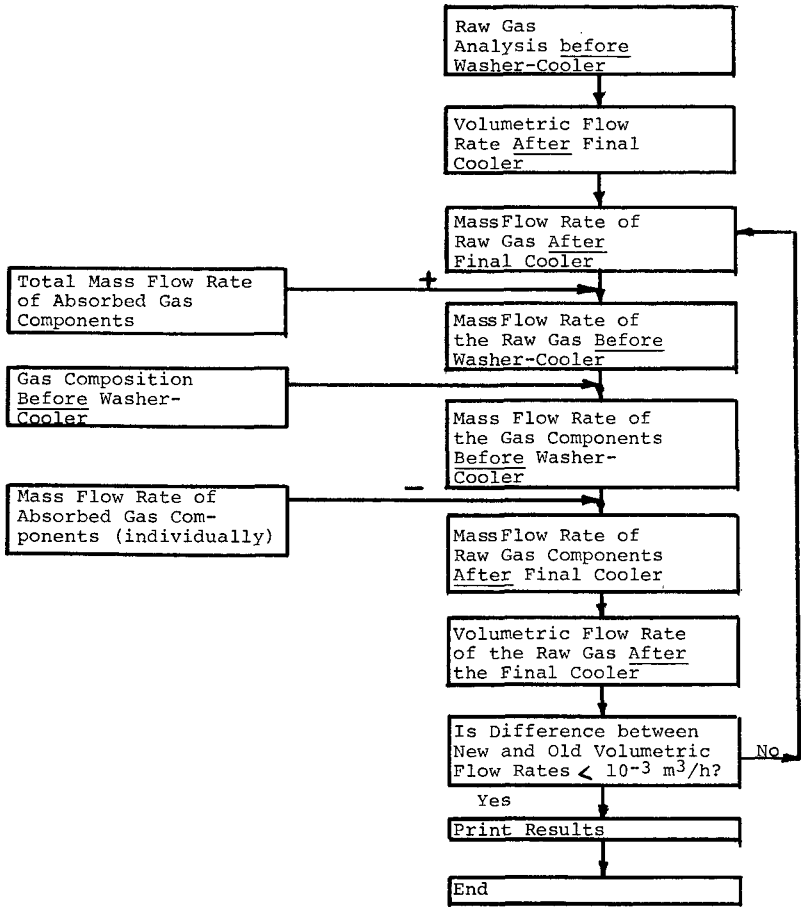

Figure 5-3. Computer Program Flow Chart for Determination of the Mass Balance Average Error

- CO_2-, H_2S-, SO_2-, and HCN-Absorption

—Measurement Data

Temperature, H_2O, washer-cooler inlet	$t_{H_2O,KW,E}$:	22.5 °C
Temperature, raw gas, washer-cooler outlet	$t_{RG,KW,A}$:	33.5 °C
Pressure, raw gas, washer-cooler inlet	$P_{EG,KW,E}$:	185.8 mmWC
Pressure, raw gas, washer-cooler outlet	$P_{RG,KW,A}$:	102.7 mmWC
Barometric pressure	B	731.0 mm Hg
CO_2 in the gas	ν_{CO_2}	10.6 vol. %
H_2O, washer-cooler	$\dot{V}_{H_2O,KW}$	34.81 m^3/h
CO_2 in H_2O before washer-cooler	w'_{CO_2}	9.2 mg/l
H_2S in the wash water, exit	w'_{H_2S}	80.90 mg/l
SO_2 in wash water, outlet	w'_{SO_2}	43.04 mg/l
HCN in wash water, outlet	w'_{HCN}	1.38 mg/l

—CO_2 absorption

Mean washer-cooler temperature

$t_m = 1/2(t_{H_2O} + t_{RG}) = \underline{28.0°C}$

Mean washer-cooler gas pressure

$P_m = 1/2(P_{RG,E} + P_{RG,A}) = \underline{10.61 \text{ mm HG}}$

Absolute pressure

$p = B + p_m$ $\quad$ $p = 741.61$ mm Hg

Partial pressure CO_2

$$P_{CO_2} = \frac{CO_2\ [\%]}{100} \cdot p \qquad P_{CO_2} = 78.6 \text{ mm Hg}$$

CO_2 absorption coefficient

at t_m and 760 mm Hg $\quad$ $\alpha\ CO_2 = 0.7026 \dfrac{Nm^3 CO_2}{m^3_{H_2O} \cdot 760 \text{ mm Hg}}$

Absorbed CO_2

$$\dot{V}'_{CO_2} = \alpha_{CO_2} \cdot \dot{V}_{H_2O,KW} \cdot P_{CO_2}, \qquad \dot{V}'_{CO_2} = 2.53 \text{ Nm}^3 CO_2/h$$

Density of CO_2 in raw gas $\quad$ $\gamma_{N_{CO_2}} = 1.9768$ kg CO_2/Nm_{tr}

Specific CO_2 content

$$W''_{CO_2} = \frac{V'_{CO_2}}{V_{H_2O}} \cdot \gamma_{N_{CO_2}} \cdot 10^3 \qquad W''_{CO_2} = 143.7 \text{ mg/l}$$

CO_2 dissolved in wash water

$$\dot{V}''_{CO_2} = \frac{V'_{CO_2}}{W''_{CO_2}} \cdot W'_{CO_2} \qquad \dot{V}''_{CO_2} = 0.16 \text{ Nm}^3 CO_2/h$$

CO_2 absorbed in wash water

$$\dot{V}°_{CO_2} = \dot{V}'_{CO_2} - \dot{V}''_{CO_2} \qquad \dot{V}°_{CO_2} = 2.37 \text{ Nm}^3 CO_2/h$$

CO_2 Absorption mass rate

$$\Delta\dot{m}_{CO_2} = \dot{V}_{CO_2} \cdot \gamma_{N,CO_2} \qquad \Delta\dot{m}_{CO_2} = 4.68 \text{ kg/h}$$

—H_2S Absorption

$$\Delta\dot{m}_{H_2S} = w'_{H_2S}[mg/l] \cdot \dot{V}_{H_2O}\ [m^3/h] \cdot 10^{-3}$$

$$\Delta\dot{V}_{N,H_2S} = \frac{\Delta\dot{m}_{H_2S}}{\gamma_{N,H_2S}} \qquad [Nm^3 tr/h]$$

Mass flow H_2S, absorption $\quad$ $\Delta\dot{m}_{H_2S} = 1.82$ kg/h

Normal density of H_2S in raw gas $\quad$ $\gamma_{N,H_2S} = 1.5392$ kg/Nm^3tr

H_2S volumetric absorption rate $\quad$ $\dot{V}_{N,H_2S} = 1.19$ Nm^3tr/h

—SO_2 Absorption

$$\Delta\dot{m}_{SO_2} = w'_{SO_2}\ [mg/l] \cdot \dot{V}_{H_2O}\ [m^3/h] \cdot 10^{-3}$$

$$\Delta\dot{V}_{N,SO_2} = \frac{\Delta\dot{m}_{SO_2}}{\gamma_{N,SO_2}} \qquad [Nm^3 tr/h]$$

SO_2 absorption mass rate	$\Delta\dot{m}_{SO_2} = 1.49$ kg/h
Normal density SO_2	$\gamma_{N,SO_2} = 2.9263$ kg/Nm^{3}tr
SO_2 volumetric absorption rate	$\dot{V}_{N,SO_2} = 0.51$ Nm^{3}tr/h

—HCN Absorption

$$\Delta\dot{m}_{HCN} = w'_{HCN}\ [mg/l] \cdot \dot{V}_{H_2O}\ [m^3/h] \cdot 10^{-3}$$

$$\Delta\dot{V}_{N,HCN} = \frac{\Delta\dot{m}_{HCN}}{\gamma_{N,HCN}} \qquad [Nm^3 tr/h]$$

HCN absorption mass rate	$\Delta\dot{m}_{HCN} = 0.05$ kg/h
Normal density HCN	$\gamma_{N,HCN} = 1.1893$ kg/Nm^{3}tr
HCN volumetric absorption rate	$\dot{V}_{N,HCN} = 0.04$ Nm^{3}tr/h

- Flow Rate of Raw Gas After the Final Cooler. This is the preliminary calculation, the final value is given under Corrected Mass Balance, page 5-44.

$$\dot{V}_{RG,f} = K_1\sqrt{\Delta p\ [mmWC]/\rho_{RG,f}\ [kg/m^3]} \qquad [m^3/h]$$

$K_1 = 788.09$, diaphragm factor

$$\rho_{RG,f} = \rho_{RG,tr} + \rho_{SD} \qquad [kg/m^3]$$

$$\rho_{RG,tr} = (P_{total} - P_{SD}) \cdot 10^5/R_{total} \cdot T \qquad [kg/m^3]$$

$$R_{total} = R/\Sigma\nu_i M_i \qquad [J/kg°K]$$

$R = 8314.3$ J/k mole, universal gas constant

$$p_{total} = B + p_{BL} \qquad [bar]$$

$$p_{SD} = f(t_{BL}) \qquad [bar]$$

$$\rho_{SD} = f(t_{BL}) \qquad [kg/m^3]$$

$$\dot{V}_{H_2O} = (p_{SD}/p_{total})\dot{V}_{RG,f} \qquad [m^3/h]$$

$$\dot{V}_{RG,tr} = \dot{V}_{RG,f} - \dot{V}_{H_2O} \qquad [m^3/h]$$

$$\dot{V}_{N,RG,tr} = \dot{V}_{RG,f} \cdot 0.3592(p_{total} - p_{SD})/T \qquad [Nm^3/h]$$

—Raw Gas Analyses, After the Final Cooler. The raw gas analyses are given in volume-percent from computer calculation.

CO_2 ν_{CO_2} : 10.60 Vol. %
CO ν_{CO} : 65.89 Vol. %
H_2 ν_{H_2} : 16.53 Vol. %
N_2 ν_{N_2} : 5.61 Vol. %
H_2S ν_{H_2S} : 1.13 Vol. %
COS ν_{COS} : 0.21 Vol. %
SO_2 ν_{SO_2} : 0.02 Vol. %
HCN ν_{HCN} : 0.014 Vol. %

Sum : 100.004 Vol. %

—Measuring Data at the Measuring Orifice After the Final Cooler

Temperature at the orifice t_{BL} : 23.4°C
Pressure at orifice p_{BL} : 103.3 mm WC
Barometric height B : 731.0 mm Hg
Pressure difference at diaphragm Δp : 55.5 mm WC

—Calculation Results, Raw Gas After the Final Cooler

Total pressure, absolute $p_{total} = 0.9847$ bar
Partial pressure, water vapor in raw gas $p_{SD} = 0.0288$ bar
Density, water vapor in raw gas $\rho_{SD} = 0.0211\ kg/m^3$
Gas constant of raw gas $R_{total} = 325.72$ J/kg°K
Density, raw gas, moist $\rho_{RG,f} = 1.0106\ kg/m^3$
Volume flow, raw gas, moist $\dot{V}_{RG,f} = 5840.40\ m^3/h$
Volume flow, H_2O in raw gas $\dot{V}_{H_2O,RG} = 170.82\ m^3/h$
Volume flow, raw gas, dry $\dot{V}_{RG,tr} = 5669.59\ m^3/h$
Normal volume, raw gas, dry $\dot{V}_{N,RG,tr} = 5071.25\ Nm^3tr/h$

- Flow Rate of Raw Gas to Washer-Cooler

$$\dot{V}_{RG,tr} = (\dot{m}_{RG,tr} \cdot R_{RG} \cdot T)/10^5 p_{total}$$

$$R_{total} = \Sigma(R_i m_i / m_{total})$$

$$p_{SD} = f(\tau)$$

$$\rho_{SD} = f(\tau)$$

$$\dot{V}_{H_2O} = p_{SD} \cdot \dot{V}_{RG,tr}/(p_{total} - p_{SD})$$

$$\dot{V}_{RG,f} = \dot{V}_{RG,tr} + \dot{V}_{H_2O}$$

—Computer Calculation of Mass Flow of Individual Components to Washer-Cooler

CO_2 mass flow in raw gas before washer-cooler	$\dot{m}_{CO_2}$	= 1056.94 kg/h
CO mass flow in raw gas before washer-cooler	$\dot{m}_{CO}$	= 4181.53 kg/h
H_2 mass flow in raw gas before washer-cooler	$\dot{m}_{H_2}$	= 75.50 kg/h
N_2 mass flow in raw gas before washer-cooler	$\dot{m}_{N_2}$	= 356.14 kg/h
H_2S mass flow in raw gas before washer-cooler	$\dot{m}_{H_2S}$	= 87.25 kg/h
COS mass flow in raw gas before washer-cooler	$\dot{m}_{COS}$	= 28.58 kg/h
SO_2 mass flow in raw gas before washer-cooler	$\dot{m}_{SO_2}$	= 2.90 kg/h
HCN mass flow in raw gas before washer-cooler	$\dot{m}_{HCN}$	= 0.86 kg/h
Mass flow, raw gas before washer-cooler	$\dot{m}_{RG,KW}$	= 5789.70 kg/h

—Gas Constants in J/kg°K

$R_{CO_2} = 188.92$ $\qquad R_{H_2S} = 243.90$

$R_{CO} = 296.84$ $\qquad R_{COS} = 138.42$

$R_{H_2} = 4124.40$ $\qquad R_{SO_2} = 129.80$

$R_{N_2} = 296.78$ $\qquad R_{HCN} = 307.61$

Gas constant, raw gas before washer-cooler $\quad R_{RG,KW} = 325.51$ J/kg°K

—Measurement Data

Temperature, raw gas, washer-cooler inlet	t_{RG}	: 216.3°C
Pressure, raw gas, washer-cooler outlet	p_{RG}	: 185.8 mm WC
Barometric height	B	: 731.0 mm Hg
Partial pressure, H_2O in raw gas	p_{SD}	: 0.3171 bar

—Calculation Results for Raw Gas to Washer-Cooler

Volume flow, raw gas, dry	$\dot{V}_{RG,tr}$	: 9291.26 m^3/h
Volume flow, H_2O in raw gas	$\dot{V}_{H_2O,RG}$	: 4360.41 m^3/h
Volume flow, raw gas, moist	$\dot{V}_{RG,f}$	: 13651.67 m^3/h
Normal volume flow, raw gas, dry	$\dot{V}_{N,RG,tr}$	: 5080.75 Nm^3/h

Volume Flow of Gasifying Oxygen. For the final value after error calculation, see page 5-44

$$\dot{V}_{VS,Z} = K_7 \cdot q_z\,[\%] \cdot \sqrt{K_G/\rho_{VS,Z}/100} \quad [m^3/h]$$

K_7 = 68.78 diaphragm constant

K_G = 166.5 Apparatus constant

z = burner subscript

$\rho_{VS} = \rho_{O_2} + \rho_{N_2} + \rho_{SD}$ [kg/m^3]

$\rho_i = \nu_i$ [%] $(P_{total} - P_{SD}) \cdot 10^3 / R_i \cdot T$ [kg/m^3]

i = gas component (O_2, N_2)

$P_{total} = B + p_{BL,z}$ (bar)

$p_{SD} = f(t_{BL})$

$\rho_{SD} = f(t_{BL})$

τ = 12°C evaluated saturation temperature of the water vapor in the gasifying oxygen

- Gas Analysis, Gasifying Oxygen, Dry

Volume fraction O_2	ν_{O_2} :	93.93 vol. %
Volume fraction N_2	ν_{N_2} :	6.07 vol. %
Sum		100.00 vol. %

The temperature at the orifice was obtained, allowing for temperature losses, in accordance with the formula

$$t_{BL} = t_{Erh} - (t_{Erh} - t_{VS})\ell_{BL}/\ell_{m_L} \quad [°C]$$

where ℓ_{m_L} = average heater-burner tube length
= 13 meters

and ℓ_{BL} = heater-diaphragm tube length
= 3 meters

Average temperature, heater t_{Erh} = 121°C

Average temperature, gasifier oxygen t_{VS} = 92.39°C

- Measurement Data

Barometric height	B = 731.0 mm Hg
Temperature at orifice	t_{BL} = 114.4°C
Vapor pressure at τ = 12°C	p_{SD} = 10.518 mm Hg
Specific weight (density), steam at τ = 12°C	ρ_{SD} = 0.0107 kg/m^3

- Calculation of Volume Flow per Burner

		Burner 1	Burner 2	Burner 3	Burner 4	
Static pressure at diaphragm	$p_{BL,z}$	227.5	194.17	212.7	230.7	mmHg
Flow-through	q_z	81.3	79.1	81.2	80.6	%
Density, gasifying oxygen	ρ_{VS}	1.2572	1.2134	1.2378	1.2614	kg/m^3
Volume flow	$\dot{V}_{VS,z}$	643.5	637.3	647.8	636.9	m^3/h
Normal volume flow	$\dot{V}_{N,VS,z}$	565.6	540.5	560.5	561.7	Nm$^3_{tr}$/h

- Calculation of Total Volume Flow of Gasifying Oxygen

  $\dot{V}_{VS} = \Sigma\dot{V}_{VS,z}$ [m^3/h]

  $\dot{V}_{H_2O} = (p_{SD}/p_{total,z})\dot{V}_{VS,z}$ [m^3/h]

  $\dot{V} = \Sigma\dot{V}_{H_2O,z}$ [m^3/h]

  $\dot{V}_{O_2} = (\dot{V}_{VS} - \dot{V}_{H_2O})\nu_{O_2}/100$ [m^3/h]

  $\dot{V}_{N_2} = \dot{V}_{VS} - (\dot{V}_{O_2} + \dot{V}_{H_2O})$ [m^3/h]

  Volume flow, gasifying oxygen $\dot{V}_{VS} = 2565.50\ m^3/h$

  Volume flow, H_2O in gasifying oxygen $\dot{V}_{H_2O,VS} = 28.49\ m^3/h$

  Volume flow, O_2 in gasifying oxygen $\dot{V}_{O_2} = 2383.01\ m^3/h$

  Volume flow, N_2 in gasifying oxygen $\dot{V}_{N_2} = 154.0\ m^3/h$

- Normal Volume Flow of Gasifying Oxygen

  Normal volume flow, gasifying oxygen $\dot{V}_{N,VS} = 2228.3\ Nm^3/h$

  Normal volume flow, H_2O $\dot{V}_{N,H_2O} = 24.72\ Nm^3/h$

  Normal volume flow, O_2 $\dot{V}_{N,O_2} = 2093.04\ Nm^3/h$

  Normal volume flow, N_2 $\dot{V}_{N,N_2} = 135.26\ Nm^3/h$

Calculation of Mass Flows From the Measurement Data

The following calculation is the preliminary result; the final value is given under Corrected Mass Balance, page 5-44.

Mass Flow of Raw Gas.

- Mass Flow of Raw Gas After Final Cooler. The computer calculation gives:

  Mass flow of CO_2 in the raw gas $\dot{m}_{CO_2}$ = 1052.26 kg/h

  Mass flow of CO in the raw gas $\dot{m}_{CO}$ = 4181.47 kg/h

  Mass flow of H_2 in the raw gas $\dot{m}_{H_2}$ = 75.50 kg/h

  Mass flow of N_2 in the raw gas $\dot{m}_{N_2}$ = 356.14 kg/h

  Mass flow of H_2S in the raw gas $\dot{m}_{H_2S}$ = 84.44 kg/h

  Mass flow of COS in the raw gas $\dot{m}_{COS}$ = 28.58 kg/h

  Mass flow of SO_2 in the raw gas $\dot{m}_{SO_2}$ = 1.41 kg/h

  Mass flow of HCN in the raw gas $\dot{m}_{HCN}$ = 0.81 kg/h

  Mass flow, raw gas, dry $\dot{m}_{RG}$ = 5780.61 kg/h

- Mass Flow of Raw Gas to Washer-Cooler. The computer calculation gives:

Mass flow of CO_2 in the raw gas	$\dot{m}_{CO_2}$ =	1056.94 kg/h
Mass flow of CO in the raw gas	$\dot{m}_{CO}$ =	4181.53 kg/h
Mass flow of H_2 in the raw gas	$\dot{m}_{H_2}$ =	75.50 kg/h
Mass flow of N_2 in the raw gas	$\dot{m}_{N_2}$ =	356.14 kg/h
Mass flow of H_2S in the raw gas	$\dot{m}_{H_2S}$ =	87.25 kg/h
Mass flow of COS in the raw gas	$\dot{m}_{COS}$ =	28.58 kg/h
Mass flow of SO_2 in the raw gas	$\dot{m}_{SO_2}$ =	2.90 kg/h
Mass flow of HCN in the raw gas	$\dot{m}_{HCN}$ =	0.86 kg/h
Mass flow, raw gas, dry	$\dot{m}_{RG}$ =	5789.70 kg/h

- Composition of Raw Gas to Washer-Cooler in Weight-Percent

Percent by weight of CO_2 in the raw gas	W_{CO_2} =	18.26 wt. %
Percent by weight of CO in the raw gas	W_{CO} =	72.22 wt. %
Percent by weight of H_2 in the raw gas	W_{H_2} =	1.30 wt. %
Percent by weight of N_2 in the raw gas	W_{N_2} =	6.15 wt. %
Percent by weight of H_2S in the raw gas	W_{H_2S} =	1.51 wt. %
Percent by weight of COS in the raw gas	W_{COS} =	0.494 wt. %
Percent by weight of SO_2 in the raw gas	W_{SO_2} =	0.050 wt. %
Percent by weight of HCN in the raw gas	W_{HCN} =	0.015 wt. %
Total		99.999 wt. %

- Composition of Raw Gas From Washer-Cooler in Weight-Percent

Percent by weight of CO_2 in the raw gas	W_{CO_2} =	18.20 wt. %
Percent by weight of CO in the raw gas	W_{CO} =	72.34 wt. %
Percent by weight of H_2 in the raw gas	W_{H_2} =	1.31 wt. %
Percent by weight of N_2 in the raw gas	W_{N_2} =	6.16 wt. %
Percent by weight of H_2S in the raw gas	W_{H_2S} =	1.46 wt. %
Percent by weight of COS in the raw gas	W_{COS} =	0.494 wt. %
Percent by weight of SO_2 in the raw gas	W_{SO_2} =	0.024 wt. %
Percent by weight of HCN in the raw gas	W_{HCN} =	0.014 wt. %
Total		100.02 wt. %

Mass Flow of Gasifying Oxygen. The following is a preliminary calculation; for final values, see Corrected Mass Balance, page 5-44.

$$\dot{m}_{VS,z} = \rho_{VS,z} \cdot \dot{V}_{VS,z}$$

$$\dot{m}_{VS} = \Sigma \dot{m}_{VS,z}$$

$$\dot{m}_{H_2O,VS} = \rho''_{H_2O} \cdot \dot{V}_{H_2O}$$

$$\dot{m}_{O_2,VS} = \frac{\nu_{O_2} \cdot M_{O_2}}{\Sigma \nu_i \cdot M_i} \cdot (\dot{m}_{VS} - \dot{m}_{H_2O})$$

$$\dot{m}_{N_2,VS} = \dot{m}_{VS} - (\dot{m}_{O_2} + \dot{m}_{H_2O})$$

	Burner 1	Burner 2	Burner 3	Burner 4	
Volume flow $\dot{V}_{VS}$	634.5	637.3	647.8	636.9	m^3/h
Density ρ_{VS}	1.2572	1.2134	1.2378	1.2614	kg/m^3
Mass flow $\dot{m}_{VS}$	797.69	773.30	801.85	803.39	kg/h

Total mass flow, gasifying oxygen	$\dot{m}_{VS}$ = 3176.23 kg/h
Mass flow H_2O in gasifying oxygen	$\dot{m}_{H_2O,VS}$ = 25.85 kg/h
Mass flow, O_2 in gasifying oxygen	$\dot{m}_{O_2,VS}$ = 2981.65 kg/h
Mass flow, N_2 in gasifying oxygen	$\dot{m}_{N_2,VS}$ = 168.72 kg/h

Mass Flow of Solids From Washer-Cooler.

$$\dot{m}_{FST,KW} = w_{FST} \; [g/l] \cdot \dot{V}_{H_2O,KW} \; [m^3/h]$$

Solids content, washer-cooler effluent	w_{FST} :	18.01 [g/l]
Flow rate of effluent washer-cooler	$\dot{V}_{H_2O,KW}$:	34.81 $[m^3/h]$

Mass flow, solids, from washer-cooler

$\dot{m}_{FST,KW}$ = 626.93 kg/h

Mass Flow of Solids From Slag Extractor.

$$\dot{m}_{ASA} = w_{ASA} \; [g/l] \cdot \dot{V}_{H_2O,ASA} \; [m^3/h]$$

Solids content, slag extractor	w_{ASA} :	1.50 [g/l]
Wash water, slag extractor	$\dot{V}_{H_2O,ASA}$:	26.53 $[m^3/h]$

Mass flow, solids, ash

$\dot{m}_{ASA}$ = 39.80 kg/h

Mass Flow of Flue Dust to Washer-Cooler.

$\dot{m}_{FLST} = 472.66$ kg/h

Mass Flow of Flue Dust From Tubular Boiler Dust Leg.

$$\dot{m}_{FLST,RK} = \frac{\xi_{FLST}\ [\%]}{100} \cdot \frac{\dot{V}_{FLST,RK}\ [m^3]}{t\ [h]} \cdot \rho_{FLST,RK}\ [kg/m^3]$$

Total volume, dust bag $\dot{V}_{STS}$ - m^3

Total duration of test t - h

Filling fraction, dust bag $\xi_{FLST,RK}$ - %

Density, dust $\rho_{FLST,RK}$ - kg/m^3

Mass flow, flue dust, tube boiler

$\dot{m}_{FLST,RK} =$ -- kg/h

Mass Flow of Slag.

$\dot{m}_{SL} = m_{SL}$ [kg] /t [h]

Mass, slag, total $m_{SL} = 141.2$ kg

Duration of test, total $t = 1.08$ h

Water content, slag $w_{H_2O,SL} = 13.8$ wt. %

Mass flow, slag

$\dot{m}_{SL} = 112.70$ kg/h

Mass Flow of Gasifier Residue.

$$\dot{m}_{RV} = \frac{\xi_{ZV}\ [\%]}{100} \cdot \frac{m_{RV}\ [kg]}{t\ [h]}$$

Mass or residue corresponding to 100% $m_{RV} = 950$ kg

Residual contents, gasifier $\xi_{RV} = 60$ %

Duration of test, total $t = 11$ h

Mass flow, gasifier residue

$\dot{m}_{RV} = 51.82$ kg/h

Mass Flow of Water From Char Feed Coal.

$$\dot{m}_{H_2O,K} = \frac{w_{H_2O,K}}{100} \cdot \dot{m}_K\ [kg/h]$$

Mass flow, char feed $\dot{m}_K = 3146.48$ kg/h

Moisture content, in char $w_{H_2O,K} = 1.63$ wt. %

Mass flow, water in char

$$\dot{m}_{H_2O,K} = 52.18 \text{ kg/h}$$

Mass Flow of Water in Gas, Less Quenching Water.

Volume flow, H_2O in gas $\dot{V}$ - m^3/h

Dew point, H_2O in gas τ - °C

Density, H_2O in gas ρ' - kg/m^3

Mass flow, H_2O in gas

$m =$ -- kg/h

Mass Flow of Water From Gas, Less Quenching Water.

$$\dot{m}_{H_2O,RG} = \dot{m}_{H_2O,VS} + \dot{m}_{H_2O,K} + \dot{m}_D + \dot{m}_{H,K} \cdot 8.9365 - \dot{m}_{H_2,RG} \cdot 8.9365$$
$$- \dot{m}_{H_2S,RG} \cdot 0.5286 - \dot{m}_{HCN,RG} \cdot 0.6666$$

Water from gasifying oxygen $\dot{m}_{H_2O,VS}$: 25.85 kg/h

Water from char $\dot{m}_{H_2O,K}$: 52.18 kg/h

Steam addition $\dot{m}_D$: 364.40 kg/h

Hydrogen in char 8.9365 $\dot{m}_{H,K}$: 53.49 kg/h 8.9365

Hydrogen in raw gas 8.9365 $\dot{m}_{H_2,RG}$: 75.50 kg/h 8.9365

H_2S in raw gas 0.5286 $\dot{m}_{H_2S,RG}$: 87.25 kg/h 0.5286

HCN in raw gas 0.6666 $\dot{m}_{HCN,RG}$: 0.86 kg/h 0.6666

Water in raw gas (without quenching water)

$$\dot{m}_{H_2O,RG} = 199.04 \text{ kg/h}$$

Calculation of Steam Addition.

$$\dot{m}_D = 4.137 \sqrt{0.25\, q_D^2/V_{SD}} \quad [\text{kg/h}]$$

$$V_{SD} = f(p_{abs})$$

$$p_{abs} = B + p_D$$

Steam pressure at mixing nozzle p_D : 4.9 kg/cm^2

Barometric pressure B : 731.0 mm Hg

Absolute pressure p_{abs} : 5.57 bar

Volume of saturated steam v_{SD} : 0.3385 m^3/kg

Flow-through, meter 1 q_1 : 26.5 %

Flow-through, meter 2 q_2 : 25.5 %

Flow-through, meter 3 q_3 : 25.0 %

Flow-through, meter 4 q_4 : 25.5 %

Mass flow, steam, mixer 1 $\dot{m}_{D,1}$: 94.22 kg/h

Mass flow, steam, mixer 2 $\dot{m}_{D,2}$: 90.66 kg/h

Mass flow, steam, mixer 3 $\dot{m}_{D,4}$: 88.88 kg/h

Mass flow, steam, mixer 4 $\dot{m}_{D,4}$: 90.66 kg/h

Total mass flow of steam

$\dot{m}_D$ = 364.42 kg/h

Mass Flow of Cooling Water for Gasifier Jacket.

$$\dot{m}_{H_2O,V} = \rho_{H_2O,V} \cdot \dot{V}_{H_2O,V} \quad [kg/h]$$

Volume flow, cooling water, gasifier $\dot{V}_{H_2O,V}$ = 237.4 m^3/h (Page 5-25)

Temperature, cooling water, inlet $t_{H_2O,V}$ = 41.52°C

Density, cooling water, gasifier $\rho_{H_2O,V}$ = 991.66 kg/m^3

Mass, flow, cooling water, gasifier jacket

$\dot{m}_{H_2O,V}$ = 235421 kg/h

Mass Flow of Char Feed.

$$\dot{m}_K = \dot{m}_{C,K}/w_{C,K} \quad [kg/h]$$

$$\dot{m}_{C,K} = \dot{m}_{C,RG} + \dot{m}_{C,FST,KW} + \dot{m}_{C,FLST} + \dot{m}_{C,FLST,RK} + \dot{m}_{C,ASA} + \dot{m}_{C,SL} + \dot{m}_{C,RV}$$

The use of the quantities $\dot{m}_{C,FST,KW}$ and $\dot{m}_{C,FLST,RK}$ is dependent on the test procedure.

- C-Mass Flow. This is the preliminary calculation; for the final value, see Corrected Mass Balance, page 5-44.

—Mass Flow of Carbon in Raw Gas to Washer-Cooler.

Mass fraction:

CO_2	w_{CO_2} :	18.26 wt.%
CO	w_{CO} :	72.22 wt.%
COS	w_{COS} :	0.494 wt.%
HCN	w_{HCN} :	0.015 wt.%

C-fraction in gas components (see corrected value on page 5-45)

Mass flow of raw gas to washer-cooler

$$\dot{m}_{RG,KW} = 5789.70 \text{ kg/h}$$

C-mass flow from raw gas:

CO_2	$\dot{m}_{C,CO_2}$:	288.44	kg/h
CO	$\dot{m}_{C,CO}$:	1793.04	kg/h
COS	$\dot{m}_{C,COS}$:	5.71	kg/h
HCN	$\dot{m}_{C,HCN}$:	0.38	kg/h
Raw gas	$\dot{m}_{C,RG}$:	2087.57	kg/h

—Mass Flow of Carbon in Solids From Washer-Cooler. This calculation is not required, since the mass flow of carbon is calculated in the flue dust.

C-mass fraction in solids	$w_{C,FST}$	- wt. %
Mass flow, solids	$\dot{m}_{FST}$	- kg/h
C-mass flow, solids	$\dot{m}_{C,FST,KW}$	- kg/h

—Mass Flow of Carbon in Flue Dust.

C-mass fraction in flue dust	$w_{C,FLST}$	: 51.2 wt. %
Mass flow in flue dust	$\dot{m}_{FLST}$	: 472.66 kg/h
C-mass flow in flue dust	$\dot{m}_{C,FLST}$	: 242.0 kg/h

—Mass Flow of Carbon in Flue Dust Boiler Dust Leg. This calculation is not required.

C-mass fraction, flue dust RK	$w_{C,RK}$	- wt. %
Mass flow, flue dust RK	$\dot{m}_{RK}$	- kg/h
C-mass flow, flue dust RK	$\dot{m}_{C,FLST,RK}$	- kg/h

—Mass Flow of Carbon in Ash Discharge.

C-mass fraction	$w_{C,ASA}$	: 72.9 kg/h
Mass flow rate	$\dot{m}_{ASA}$	: 39.8 kg/h
C-mass flow rate	$\dot{m}_{C,ASA}$	: 29.01 kg/h

—Mass Flow Rate of Carbon in Slag.

C-mass fraction, slag	$w_{C,SL}$	: 6.7 wt. %
Mass flow, slag	$\dot{m}_{SL}$	: 112.70 kg/h
C-mass flow, slag	$\dot{m}_{C,SL}$	: 7.55 kg/h

—Mass Flow of Carbon in Gasifier Residue.

C-mass fraction	$w_{C,RV}$	: 72.9 wt. %
Mass flow rate	$\dot{m}_{RV}$	: 51.82 kg/h
C-mass flow rate	$\dot{m}_{C,RV}$	: 37.78 kg/h

—Mass Flow of Carbon in Char Feed.

C-mass flow, raw gas	$\dot{m}_{C,RG}$:	2087.57	kg/h
C-mass flow, solids, from washer-cooler	$\dot{m}_{C,FST,KW}$:	-	kg/h
C-mass flow, flue dust	$\dot{m}_{C.FLST}$:	242.00	kg/h
C-mass flow, flue dust, from tubular boiler dust leg	$\dot{m}_{C,FLST,RK}$	-	kg/h
C-mass flow, ash discharge slag extractor	$\dot{m}_{C,ASA}$:	29.01	kg/h
C-mass flow, slag	$\dot{m}_{C,SL}$:	7.55	kg/h
C-mass flow, residue, in gasifier heel	$\dot{m}_{C,RV}$:	37.78	kg/h
C-mass flow, in char feed	$\dot{m}_{C,K}$:	2403.91	kg/h

- Mass Flow of Char Feed

C-mass fraction, in char	$w_{C,K}$	: 76.4 wt. %
C-mass flow, in char	$\dot{m}_{C,K}$	: 2403.91 kg/h

Mass flow, char feed

$$\dot{m}_K = 3146.48 \text{ kg/h}$$

- Provisional Mass Balance

—Mass Flow, Gasifier Inlet

Char, dry	$\dot{m}_K$:	3146.48	kg/h
Oxygen, pure	$\dot{m}_{O_2}$:	2981.65	kg/h
Nitrogen	$\dot{m}_{N_2}$:	162.72	kg/h
Water from gasifier oxygen	$\dot{m}_{H_2O,VS}$:	25.85	kg/h
Water from char	$\dot{m}_{H_2O,K}$:	52.18	kg/h
Steam addition	$\dot{m}_D$:	364.40	kg/h
Total, mass flow, inlet	$\dot{m}_{Ein}$:	6733.28	kg/h

—Mass Flow, Gasifier Outlet

Raw gas before washer-cooler	$\dot{m}_{RG}$:	5789.70	kg/h
Water in raw gas (excluding quenching water)	$\dot{m}_{H_2O,RG}$:	199.04	kg/h
Solids, washer-cooler	$\dot{m}_{FST, KW}$:	-	
Flue dust	$\dot{m}_{FLST}$:	472.66	kg/h
Flue dust in tubular boiler dust leg	$\dot{m}_{FLST,RK}$:	-	
Ash discharge	$\dot{m}_{ASA}$:	39.8	kg/h
Slag	$\dot{m}_{SL}$:	112.70	kg/h
Residue, gasifier	$\dot{m}_{RV}$:	51.82	kg/h
Total, mass flow outlet	$\dot{m}_{Aus}$ =	6665.72	kg/h

$\delta = \dot{m}_{Ein} - \dot{m}_{Aus}$ [kg/h]

(in) (out)

= 67.56 [kg/h]

Calculation of Errors

The error calculation presented here is an abbreviation of a detailed and comprehensive error calculation. The average percentage errors of the individual mass flows were determined from four series of experiments and the average value determined. This constituted the conversion factor for the average error of each individual mass flow.

Mass Flow, Gasifier Inlet.

Char	$\dot{m}_K$	3146.48 ± 0.0	kg/h
Oxygen, pure	$\dot{m}_{O_2}$	2981.65 ± 164.0	kg/h
Nitrogen	$\dot{m}_{N_2}$	162.72 ± 9.8	kg/h
Water from gasifier O_2	$\dot{m}_{H_2O,VS}$	25.85 ± 1.6	kg/h
Water from char	$\dot{m}_{H_2O,K}$	52.18 ± 6.3	kg/h
Steam addition	$\dot{m}_D$	364.4 ± 29.2	kg/h
Total	$\dot{m}_{total,in}$	= 6733.28	kg/h

Mass Flow, Gasifier Outlet.

Raw gas before washer-cooler	$\dot{m}_{RG}$	5789.70 ± 0.0 kg/h
Water in raw gas (without quenching water)	$\dot{m}_{H_2O,RG}$	199.04 ± 10.9 kg/h
Flue dust	$\dot{m}_{FLST}$	472.66 ± 0.0 kg/h
Ash discharge	$\dot{m}_{ASA}$	39.80 ± 0.0 kg/h
Slag	$\dot{m}_{SL}$	112.70 ± 0.0 kg/h
Gasifier residue	$\dot{m}_{RV}$	51.82 ± 0.0 kg/h
Total	$\dot{m}_{total,out}$	= 6665.72 kg/h

Difference: $\dot{m}_{total,in} - \dot{m}_{total,out} = \delta = 67.56$ kg/h

Corrected Mass Balance.

$$\delta = \Sigma\dot{m}_{i,in} - \Sigma\dot{m}_{i,out}$$

$$\dot{m}_{i,corrected} = \dot{m}_{i,in} = s_i^2\delta/\Sigma s_i^2 \qquad \text{(inlet)}$$

$$-\dot{m}_{i,corrected} = -\dot{m}_{i,out} - s_i^2\delta/\Sigma s_i^2 \qquad \text{(outlet)}$$

- Corrected Mass Flow, Gasifier Inlet

Char	$\dot{m}_K$	3146.48	kg/h
Oxygen, pure	$\dot{m}_{O_2}$	2916.77	kg/h
Nitrogen	$\dot{m}_{N_2}$	162.72	kg/h
Water from gasifier O_2	$\dot{m}_{H_2O,VS}$	25.84	kg/h
Water from char	$\dot{m}_{H_2O,K}$	52.08	kg/h
Steam addition	$\dot{m}_D$	362.34	kg/h
Mass flow, inlet	$\dot{m}_{total,in}$	6666.23	kg/h

- Corrected Mass Flow, Gasifier Outlet

Raw gas before washer-cooler	$\dot{m}_{RG}$	5789.70	kg/h
Water in raw gas (without quenching water)	$\dot{m}_{H_2O,RG}$	199.33	kg/h
Flue dust	$\dot{m}_{FLST}$	472.66	kg/h
Ash discharge	$\dot{m}_{ASA}$	39.80	kg/h
Slag	$\dot{m}_{SL}$	112.70	kg/h
Gasifier residue	$\dot{m}_{RV}$	51.82	kg/h
Mass flow, outlet	$\dot{m}_{total,out}$	6666.01	kg/h

Calculation of Mass Flows of Elements

This calculation was computed from the corrected mass balance, page 5-44. The mass balance resulting from the test data was corrected on the basis of the error calculation. With the corrected values, a calculation was made of the mass flows of the gas components and of the elements, based on the assumption that the volume fraction and the weight fraction had not varied.

$$\dot{m}_i = (1/100) w_i \, [\%] \cdot \dot{m}_{total}$$

$$\dot{m}_{j,i} = \dot{m}_i \, [kgi/h] \cdot n_{j,i} \, [kg \; j/k \; mole_i] \cdot (1/M_i) \, [k \; mole_i/kgi]$$

where i = gas component

j = element component (C, H, O, S)

n = element j in kg per k mole of gas component i

C, H, O, S-Mass Flow in Raw Gas to the Washer-Cooler (corrected value).*

Corrected mass flow, raw gas before washer-cooler	$\dot{m}_{RG}$	5789.70	kg/h
CO_2 mass fraction in raw gas before washer-cooler	w_{CO_2}	18.26	wt. %
CO mass fraction in raw gas before washer-cooler	w_{CO}	72.22	wt. %
H_2 mass fraction in raw gas before washer-cooler	w_{H_2}	1.30	wt. %
N_2 mass fraction in raw gas before washer-cooler	w_{N_2}	6.15	wt. %
H_2S mass fraction in raw gas before washer-cooler	w_{H_2S}	1.51	wt. %
COS mass fraction in raw gas before washer-cooler	w_{COS}	0.494	wt. %
SO_2 mass fraction in raw gas before washer-cooler	w_{SO_2}	0.050	wt. %
HCN mass fraction in raw gas before washer-cooler	w_{HCN}	0.015	wt. %
Total	w =	99.999	wt. %

- C-Mass Flow in Raw Gas Before Washer-Cooler
  —C-Fraction in:

CO_2	n_{C,CO_2}	12.01 kgC/k mole CO_2
CO	$n_{C,CO}$	12.01 kgC/k mole CO
COS	$n_{C,COS}$	12.01 kgC/k mole COS
HCN	$n_{C,HCN}$	12.01 kgC/k mole HCN

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* See page 5-36.

—Mass Flow From Raw Gas Before Washer-Cooler

Mass flow CO_2	$\dot{m}_{CO_2,RG}$	1057.20 kg/h
Mass flow CO	$\dot{m}_{CO,RG}$	4181.32 kg/h
Mass flow COS	$\dot{m}_{COS,RG}$	28.60 kg/h
Mass flow HCN	$\dot{m}_{HCN,RG}$	0.87 kg/h

—C-Mass Flow in Raw Gas

C-mass flow from CO_2	$\dot{m}_{C,CO_2}$	288.62 kg/h
C-mass flow from CO	$\dot{m}_{C,CO}$	1793.79 kg/h
C-mass flow from COS	$\dot{m}_{C,COS}$	5.72 kg/h
C-mass flow from HCN	$\dot{m}_{C,HCN}$	0.39 kg/h
C-mass flow, raw gas	$\dot{m}_{C,RG}$	= 2088.51 kg/h

- H-Mass Flow in the Raw Gas Before Washer-Cooler

—H-fraction in:

H_2	n_{H,H_2}	2.016 kgH/k mole H_2
H_2S	n_{H,H_2S}	2.016 kgH/k mole H_2S
HCN	$n_{H,HCN}$	1.008 kgH/k mole HCN

—Mass Flow of Raw Gas

Mass flow H_2	$\dot{m}_{H_2,RG}$	75.27 kg/h
Mass flow H_2S	$\dot{m}_{H_2S,RG}$	87.42 kg/h
Mass flow HCN	$\dot{m}_{HCN,RG}$	0.87 kg/h

—H-Mass Flow in Raw Gas

H-mass flow from H_2	$\dot{m}_{H,H_2}$	75.27 kg/h
H-mass flow from H_2S	$\dot{m}_{H,H_2S}$	5.16 kg/h
H-mass flow from HCN	$\dot{m}_{H,HCN}$	0.03 kg/h
H-mass flow, raw gas	$\dot{m}_{H,RG}$	80.46 kg/h

- O-Mass Flow in the Raw Gas Before Washer-Cooler

—O-fraction in:

CO_2	n_{O,CO_2}	32.00 kgO/k mole CO_2
CO	$n_{O,CO}$	16.00 kgO/k mole CO
SO_2	n_{O,SO_2}	32.00 kgO/k mole SO_2
COS	$n_{O,COS}$	16.0 kgO/k mole COS

—Mass Flow in Raw Gas

Mass flow CO_2	$\dot{m}_{CO_2,RG}$	1057.20 kg/h
Mass flow CO	$\dot{m}_{CO,RG}$	4181.32 kg/h

Mass flow SO_2	$\dot{m}_{SO_2,RG}$	2.89 kg/h
Mass flow COS	$\dot{m}_{COS,RG}$	28.60 kg/h

—O-Mass Flow in Raw Gas

O-mass flow from CO_2	$\dot{m}_{O,CO_2}$	768.87 kg/h
O-mass flow from CO	$\dot{m}_{O,CO}$	2589.33 kg/h
O-mass flow from SO_2	$\dot{m}_{O,SO_2}$	1.45 kg/h
O-mass flow from COS	$\dot{m}_{O,COS}$	7.63 kg/h
O-mass flow in raw gas	$\dot{m}_{O,RG}$	3167.27 kg/h

- S-Mass Flow in the Raw Gas to the Washer-Cooler

—S-fraction in:

SO_2	n_{S,SO_2}	32.06 kgS/k mole SO_2
H_2S	n_{S,H_2S}	32.06 kgS/k mole H_2S
COS	$n_{S,COS}$	32.06 kgS/k mole COS

—Mass Flow in Raw Gas

Mass Flow SO_2	$\dot{m}_{SO_2,RG}$	2.89 kg/h
Mass flow H_2S	$\dot{m}_{H_2S,RG}$	87.42 kg/h
Mass flow COS	$\dot{m}_{COS,RG}$	28.60 kg/h

—S-Mass Flow in Raw Gas

S-mass flow from SO_2	$\dot{m}_{S,SO_2}$	1.45 kg/h
S-mass flow from H_2S	$\dot{m}_{S,H_2S}$	82.25 kg/h
S-mass flow from COS	$\dot{m}_{S,COS}$	15.28 kg/h
S-mass flow, raw gas	$\dot{m}_{S,RG}$	98.98 kg/h

H, O-Mass Flow From Gasifier Oxygen.

$$\dot{m}_{j,i} = \dot{m}_i \text{ [kgi/h]} \cdot n_{j,i} \text{ [kgj/kgi]} \cdot (1/M_i) \text{ [k mole i/kgi]}$$

where i = gas component

j = component element (H, O)

M_{H_2} = 2.016 kg/k mole

M_{O_2} = 32.00 kg/k mole

- O-Mass Flow as O_2 From Gasifier Oxygen

O-fraction in O_2	n_{O,O_2}	32.00 kgO/k mole O_2	
Mass flow O_2	$\dot{m}_{O_2}$	2916.77 kg/h	(Page 5-44)
O-mass flow from O_2	$\dot{m}_{O,O_2}$	2916.77 kg/h	

- O-Mass Flow From H_2O in Gasifier Oxygen

O-fraction in H_2O	n_{O,H_2O}	16.00 kgO/k mole H_2O	
Mass flow H_2O in gasifier oxygen	$\dot{m}_{H_2O,VS}$	25.84 kg/h	(Page 5-44)
O-mass flow from H_2O in gasifier oxygen	$\dot{m}_{O,H_2O,VS}$	22.95 kg/h	

- H-Mass Flow From H_2O in Gasifier Oxygen

H-fraction in H_2O	n_{H,H_2O}	2.016 kgH/k mole H_2O
Mass flow H_2O in gasifier oxygen	$\dot{m}_{H_2O,VS}$	25.84 kg/h
H-mass flow from H_2O in gasifier oxygen	$\dot{m}_{H,H_2O,VS}$	2.89 kg/h

<u>C, H, O, N, S-Mass Flow of Flue Dust at the Dust-Measuring Location.</u>

$$\dot{m}_{j,FLST} = (1/100)\, w_{j,FLST} \cdot \dot{m}_{FLST} \quad [kg/h]$$

- Mass Flow of Dry Flue Dust, Measuring Location

  $\dot{m}_{FLST} = 472.66$ kg/h (Page 5-44)

- Element Analysis of Flue Dust

H_2O	w_{H_2O}	-	wt. %
C	w_C	51.2	wt. % wf
H	w_H	0.1	wt. % wf
O	w_O	0.0	wt. % wf
N	w_N	0.0	wt. % wf
S	w_S	2.7	wt. % wf
Ash	w_{AS}	46.1	wt. % wf

- C-Mass Flow
  $\dot{m}_{C,FLST} = 242.00$ kg/h
- H-Mass Flow
  $\dot{m}_{H,FLST} = 0.47$ kg/h
- O-Mass Flow
  $\dot{m}_{O,FLST} = 0.00$ kg/h
- N-Mass Flow
  $\dot{m}_{N,FLST} = 0.00$ kg/h
- S-Mass Flow
  $\dot{m}_{S,FLST} = 12.76$ kg/h
- Ash Mass Flow
  $\dot{m}_{AS,FLST} = 217.90$ kg/h

C, H, O, N, S-Mass Flow of Solids Suspended From the Washer-Cooler Effluent (Not Used in This Calculation).

$$\dot{m}_{j,FST} = (1/100) w_{j,FST} \cdot \dot{m}_{FST} \quad [kg/h]$$

- Mass Flow of Dry Solids
  $\dot{m}_{FST}$ - kg/h
- Element Analysis of the Solids From the Washer-Cooler*

H_2O	w_{H_2O}	-	wt. %
C	w_C	-	wt. % wf
H	w_H	-	wt. % wf
O	w_O	-	wt. % wf
N	w_N	-	wt. % wf
S	w_S	-	wt. % wf
Ash	w_{AS}	-	wt. % wf

- C-Mass Flow
  $\dot{m}_{C,FST}$ - kg/h
- H-Mass Flow
  $\dot{m}_{H,FST}$ - kg/h
- O-Mass Flow
  $\dot{m}_{O,FST}$ - kg/h
- N-Mass Flow
  $\dot{m}_{N,FST}$ - kg/h
- S-Mass Flow
  $\dot{m}_{S,FST}$ - kg/h
- Ash Mass Flow
  $\dot{m}_{AS,FST}$ - kg/h

C, H, O, N, S Mass Flow of Solids in Slag Extractor Effluent.

$$\dot{m}_{j,ASA} = (1/100) w_{j,ASA} \cdot \dot{m}_{ASA} \quad [kg/h]$$

- Mass Flow of Dry Solids
  $\dot{m}_{ASA}$ = 39.80 kg/h (Page 5-44)
- Element Analysis of Solids

H_2O	w_{H_2O}	-	wt. %
C	w_C	72.9	wt. % wf
H	w_H	0.2	wt. % wf

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*Not required because the Dust Analysis from the gas stream is used for the Mass Balance.

O	w_O	0.35	wt. % wf
N	w_N	0.25	wt. % wf
S	w_S	2.2	wt. % wf
Ash	w_{AS}	24.0	wt. % wf

- C-Mass Flow
  $\dot{m}_{C,ASA} = 29.1$ kg/h
- H-Mass Flow
  $\dot{m}_{H,ASA} = 0.08$ kg/h
- O-Mass Flow
  $\dot{m}_{O,ASA} = 0.14$ kg/h
- N-Mass Flow
  $\dot{m}_{N,ASA} = 0.14$ kg/h
- S-Mass Flow
  $\dot{m}_{S,ASA} = 0.88$ kg/h
- Ash Mass Flow
  $\dot{m}_{AS,ASA} = 9.55$ kg/h

C, H, O, N, S-Mass Flow From the Slag.

$$\dot{m}_{j,SL} = (1/100) w_{j,SL} \cdot \dot{m}_{SL} \quad [kg/h]$$

- Mass Flow of Dry Slag
  $\dot{m}_{SL} = 112.70$ kg/h (Page 5-44)
- Element Analysis of Slag

H_2O	w_{H_2O}	-	wt. %
C	w_C	6.7	wt. % wf
H	w_H	0.0	wt. % wf
O	w_O	0.0	wt. % wf
N	w_N	0.0	wt. % wf
S	w_S	1.9	wt. % wf
Ash	w_{AS}	91.4	wt. % wf

- C-Mass Flow
  $\dot{m}_{C,SL} = 7.55$ kg/h
- H-Mass Flow
  $\dot{m}_{H,SL} = 0.00$ kg/h
- O-Mass Flow
  $\dot{m}_{O,SL} = 0.00$ kg/h
- N-Mass Flow
  $\dot{m}_{N,SL} = 0.00$ kg/h

- S-Mass Flow
  $\dot{m}_{S,SL} = 2.14$ kg/h
- Ash Mass Flow
  $\dot{m}_{AS,SL} = 103.01$ kg/h

C, H, O, N, S-Mass Flow From the Residue in the Gasifier.

$$\dot{m}_{j,RV} = (1/100) w_{j,RV} \cdot \dot{m}_{RV} \quad [\text{kg/h}]$$

- Mass Flow of Dry Residue
  $\dot{m}_{RV} = 51.82$ kg/h (Page 5-44)
- Element Analysis of Residue

H_2O	w_{H_2O}	-	wt. %
C	w_C	72.9	wt. % wf
H	w_H	0.2	wt. % wf
O	w_O	0.35	wt. % wf
N	w_N	0.35	wt. % wf
S	w_S	2.2	wt. % wf
Ash	w_{AS}	24.0	wt. % wf

- C-Mass Flow
  $\dot{m}_{C,RV} = 37.78$ kg/h
- H-Mass Flow
  $\dot{m}_{H,RV} = 0.10$ kg/h
- O-Mass Flow
  $\dot{m}_{O,RV} = 0.18$ kg/h
- N-Mass Flow
  $\dot{m}_{N,RV} = 0.18$ kg/h
- S-Mass Flow
  $\dot{m}_{S,RV} = 1.14$ kg/h
- Ash Mass Flow
  $\dot{m}_{AS,RV} = 12.44$ kg/h

C, H, O, N, S-Mass Flow in the Char Feed.

$$\dot{m}_{j,K} = (1/100) w_{j,K} \cdot \dot{m}_K \quad [\text{kg/h}]$$

- C-Mass Flow of the Char Feed
  $\dot{m}_{C,K} = \dot{m}_{C,RG} + \dot{m}_{C,FLST} + \dot{m}_{C,FST} + \dot{m}_{C,ASA} + \dot{m}_{C,SL} + \dot{m}_{C,RV}$

C-mass flow, raw gas	$\dot{m}_{C,RG}$	2088.51 kg/h	(Pages 5-46 to 5-51)
C-mass flow, flue dust	$\dot{m}_{C,FLST}$	242.00 kg/h	
C-mass flow, solids	$\dot{m}_{C,FST}$	- kg/h	
C-mass flow, ash discharge, slag extractor	$\dot{m}_{C,ASA}$	29.1 kg/h	
C-mass flow, slag	$\dot{m}_{C,SL}$	7.55 kg/h	
C-mass flow, residue	$\dot{m}_{C,RV}$	37.78 kg/h	
C-mass flow of char feed	$\dot{m}_{C,K}$ =	2403.91 kg/h	

- Element Analysis of Char Feed

H_2O	w_{H_2O}	1.63 wt. %
C	w_C	76.4 wt. % wf
H	w_H	1.7 wt. % wf
O	w_O	3.3 wt. % wf
N	w_N	1.4 wt. % wf
S	w_S	4.0 wt. % wf
Ash	w_{AS}	13.2 wt. % wf

- Mass Flow of the Char Feed

  $\dot{m}_K = 100\ \dot{m}_{C,K}/w_{C,K}$ [kg/h]

C-wt. fraction of charged-in dry char	$w_{C,K}$	76.4 wt. % wf
Mass flow of charged-in dry char	$\dot{m}_K$	3146.48 kg/h

- C-Mass Flow, Char

  $\dot{m}_{C,K} = 2403.91$ kg/h

- H-Mass Flow, Char

  $\dot{m}_{H,K} = 53.49$ kg/h

- O-Mass Flow, Char

  $\dot{m}_{O,K} = 103.83$ kg/h

- N-Mass Flow, Char

  $\dot{m}_{N,K} = 44.05$ kg/h

- S-Mass Flow, Char

  $\dot{m}_{S,K} = 125.86$ kg/h

- Ash Mass Flow, Char

  $\dot{m}_{AS,K} = 415.34$ kg/h

<u>H, O-Mass Flow in the Water Content of the Char Feed.</u>

$$\dot{m}_{j,H_2O,K} = \dot{m}_{H_2O,K}\ [\text{kg/h}] \cdot n_{j,i}\ [\text{kgi/k mole i}] \cdot (1/M_i)\ [\text{k mole i/kgi}]$$

where i = component material

j = component element

n = fraction of element in kg/kmole material component i

M = molecular weight

- H-Mass Flow From H_2O of the Char

  Mass flow, H_2O in char $\dot{m}_{H_2O,K}$ 52.08 kg/h (Page 5-44)

  H-fraction in the H_2O n_{H,H_2O} 2.016 kgH/kmole H_2O

  H-mass flow from H_2O in char

  $\dot{m}_{H,H_2O,K} = 5.83$ kg/h

- O-Mass Flow in H_2O in the Char

  Mass flow, H_2O in char $\dot{m}_{H_2O,K}$ 52.08 kg/h (Page 5-44)

  O-fraction in the H_2O n_{O,H_2O} 16.00 kgO/kmole H_2O

  O-mass flow from H_2O in char

  $\dot{m}_{O,H_2O,K} = 46.25$ kg/h

H, O-Mass Flow From the Water Content of the Raw Gas to the Washer-Cooler (Without Quenching Water.

$\dot{m}_{j,H_2O,RG} = \dot{m}'_{H_2O,RG}$ [kg/h] · $n_{j,i}$ [kgj/kmole i] · [1/M_i) [kmole i/kgi]

- H-Mass Flow From H_2O in the Raw Gas

  Mass flow, H_2O in the raw gas $\dot{m}'_{H_2O,RG}$: 199.33 kg/h (Page 5-44)

  H-fraction in H_2O n_{H,H_2O} : 2.016 kgH/kmole H_2O

  H-mass flow from H_2O in the raw gas, without quenching water

  $\dot{m}'_{H,H_2O,RG} = 22.31$ kg/h

- O-Mass Flow From H_2O in the Raw Gas

  Mass flow of H_2O in the raw gas $\dot{m}'_{H_2O,RG}$: 199.33 kg/h (Page 5-44)

  O-fraction H_2O n_{O,H_2O} : 16.0 kgO/kmole H_2O

  O-mass flow from H_2O in the raw gas, without quenching water

  $\dot{m}'_{O,H_2O,RG} = 177.02$ kg/h

H, O-Mass Flow From Steam Addition.

$\dot{m}_{j,H_2O,D} = \dot{m}_D$ [kg/h] · $n_{j,i}$ [kgi/kmole i] · (1/M_i) [kmole i/kgi]

- H-Mass Flow From Steam Addition

  Mass flow, steam addition $\dot{m}_D$: 362.34 kg/h (Page 5-44)

  H fraction in the H_2O $n_{H,D}$: 2.016 kgH/kmole H_2O

  H-mass flow from steam addition

  $\dot{m}_{H,D} = 40.55$ kg/h

- O-Mass Flow From Steam Addition

  Mass flow, steam addition $\dot{m}_D$: 362.34 kg/h (Page 5-44)

  O-fraction in H_2O $n_{O,D}$: 16.00 kgO/kmole H_2O

  O-mass flow from steam addition

  $\dot{m}_{O,D} = 321.79$ kg/h

Mass Balance

Mass Flow, Inlet.

Char	$\dot{m}_K$	: 3146.48	kg/h
Oxygen (pure)	$\dot{m}_{O_2}$	: 2916.77	kg/h
Nitrogen	$\dot{m}_{N_2}$	: 162.72	kg/h
Water from oxygen	$\dot{m}_{H_2O,O_2}$	: 25.84	kg/h
Water from char	$\dot{m}_{H_2O,K}$	: 52.08	kg/h
Steam	$\dot{m}_D$	: 362.34	kg/h
Total	$\dot{m}_{in}$	: 6666.23	kg/h

Mass Flow, Outlet.

Raw gas before washer-cooler	$\dot{m}_{RG}$	: 5789.70	kg/h
Water in raw gas, without quenching water	$\dot{m}_{H_2O,RG}$	: 199.33	kg/h
Flue dust	$\dot{m}_{FLST}$	: 472.66	kg/h
Ash discharge	$\dot{m}_{ASA}$	: 39.80	kg/h
Slag	$\dot{m}_{SL}$	: 112.70	kg/h
Gasifier residue	$\dot{m}_{RV}$	: 51.82	kg/h
Total	$\dot{m}_{out}$	: 6666.01	kg/h

Carbon Balance

Mass Flow, Carbon, Inlet.

C-char	$\dot{m}_{C,K}$	: 2403.91	kg/h	(Page 5-52)
C-total	$\dot{m}_{C,in}$	: 2403.91	kg/h	

Mass Flow, Carbon, Outlet.

C-raw gas before washer-cooler	$\dot{m}_{C,RG}$:	2088.51	kg/h	(Page 5-46)
C-flue dust	$\dot{m}_{C,FLST}$:	242.00	kg/h	(Page 5-48)
C-ash discharge	$\dot{m}_{C,ASA}$:	29.10	kg/h	(Page 5-50)
C-slag	$\dot{m}_{C,SL}$:	7.55	kg/h	(Page 5-50)
C-gasifier residue	$\dot{m}_{C,RV}$:	37.78	kg/h	(Page 5-51)
C-total	$\dot{m}_{C,out}$:	2404.94	kg/h	

Difference: $\dot{m}_{C,in} - \dot{m}_{C,out} = \Delta\dot{m}_C = -1.03$ kg/h

Ratio: $\dot{m}_{C,out} : \dot{m}_{C,in} = \eta_{C,balance} = 1.000$

Hydrogen Balance

Mass Flow, Hydrogen (H), Inlet.

H-char	$\dot{m}_{H,K}$:	53.49	kg/h	(Page 5-52)
H-water in char	$\dot{m}_{H,H_2O,K}$:	5.83	kg/h	(Page 5-53)
H-water in oxygen	$\dot{m}_{H,H_2O,VS}$:	2.89	kg/h	(Page 5-48)
H-steam addition	$\dot{m}_{H,D}$:	40.55	kg/h	(Page 5-53)
H-total	$\dot{m}_{H,in}$:	102.76	kg/h	

Mass Flow, Hydrogen (H), Outlet.

H-raw gas before washer-cooler	$\dot{m}_{H,RG}$:	80.46	kg/h	(Page 5-46)
H-water in raw gas, without quenching water	$\dot{m}_{H,H_2O,RG}$:	22.31	kg/h	(Page 5-53)
H-flue dust	$\dot{m}_{H,FLST}$:	0.47	kg/h	(Page 5-48)
H-ash discharge	$\dot{m}_{H,ASA}$:	0.08	kg/h	(Page 5-50)
H-slag	$\dot{m}_{H,SL}$:	0.00	kg/h	(Page 5-50)
H-gasifier residue	$\dot{m}_{H,RV}$:	0.10	kg/h	(Page 5-51)
H-total	$\dot{m}_{H,out}$:	103.42	kg/h	

Difference: $\dot{m}_{H,in} - \dot{m}_{H,out} = \Delta\dot{m}_H = -0.66$ kg/h

Ratio: $\dot{m}_{H,out} : \dot{m}_{H,in} = \eta_{H,balance} = 1.006$

Oxygen Balance

Mass Flow, Oxygen (O), Inlet.

O-char	$\dot{m}_{O,K}$	: 103.83	kg/h	(Page 5-52)
O-oxygen	$\dot{m}_{O,O_2}$	: 2916.77	kg/h	(Page 5-47)
O-water in char	$\dot{m}_{O,H_2O,K}$	: 46.25	kg/h	(Page 5-53)
O-water in oxygen	$\dot{m}_{O,H_2O,VS}$	: 22.95	kg/h	(Page 5-48)
O-steam addition	$\dot{m}_{O,D}$	: 321.79	kg/h	(Page 5-54)
O-total	$\dot{m}_{O,in}$	: 3411.59	kg/h	

Mass Flow Oxygen (O), Outlet.

O-raw gas before washer-cooler	$\dot{m}_{O,RG}$	: 3167.27	kg/h	(Page 5-47)
O-water in raw gas	$\dot{m}_{O,H_2O,RG}$	: 177.02	kg/h	(Page 5-53)
O-flue dust	$\dot{m}_{O,FLST}$	: 0.00	kg/h	(Page 5-48)
O-ash discharge	$\dot{m}_{O,ASA}$	: 0.14	kg/h	(Page 5-50)
O-slag	$\dot{m}_{O,SL}$	: 0.00	kg/h	(Page 5-50)
O-gasifier residue	$\dot{m}_{O,RV}$	: 0.18	kg/h	(Page 5-51)
O-total	$\dot{m}_{O,out}$	: 3344.61	kg/h	

Difference: $\dot{m}_{O,in} - \dot{m}_{O,out} = \Delta\dot{m}_{O} = 66.98$ kg/h

Ratio: $\dot{m}_{O,out} : \dot{m}_{O,in} = \eta_{O,balance} = 0.980$

Sulfur Balance

Mass Flow, Sulfur (S), Inlet.

S-char	$\dot{m}_{S,K}$	: 125.86 kg/h	(Page 5-52)
S-total	$\dot{m}_{S,in}$	: 125.86 kg/h	

Mass Flow, Sulfur (S), Outlet.

S-raw gas before washer-cooler	$\dot{m}_{S,RG}$	: 98.98	kg/h	(Page 5-47)
S-flue dust	$\dot{m}_{S,FLST}$	: 12.76	kg/h	(Page 5-48)
S-ash discharge	$\dot{m}_{S,ASA}$	: 0.88	kg/h	(Page 5-50)
S-slag	$\dot{m}_{S,SL}$	: 2.14	kg/h	(Page 5-51)
S-gasifier residue	$\dot{m}_{S,RV}$	: 1.14	kg/h	(page 5-51)
S-total	$\dot{m}_{S,out}$	: 115.90	kg/h	

Difference: $\dot{m}_{S,in} - \dot{m}_{S,out} = \Delta\dot{m}_S = 9.96$ kg/h

Ratio: $\dot{m}_{S,out} : \dot{m}_{S,in} = \eta_{S,balance} = 0.921$

Heat Balance (For Test Run 027)

The balance limits are illustrated in Figure 5-4.

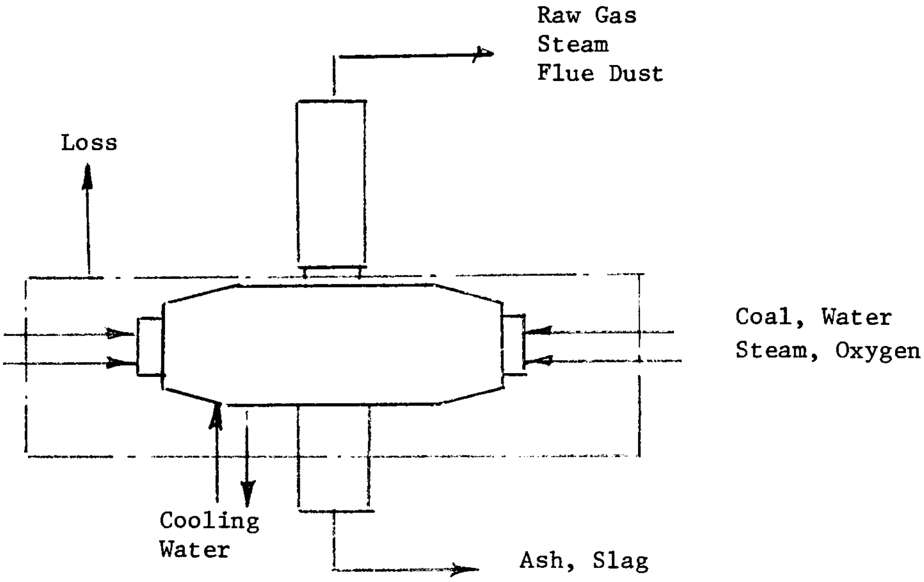

Figure 5-4. Sketch of Balance Limits

Added Enthalpy.

$$\Delta H_i = \dot{m}_i \cdot c_{pi} \Big|_0^{t_i} (t_i - t_o)$$

i = component

Reference temperature t_o is 0°C,

- Enthalpy of Char Feed

Initial temperature of char	$t_K = 20°C$ (assumed)
Average specific heat of char	$c_{p_{m,K}} = 0.7392$ kJ/kg°K (Page 5-58)
Mass flow of char	$\dot{m}_K = 3146.48$ kg/h (Page 5-44)
Enthalpy of charged-in (raw) char, inlet	$\Delta H_K = 12.92$ kJ/sec
	$\Delta H_K = 1.11 \cdot 10^4$ kcal/h

—Average Specific Heat of Char, Flue Dust, Ash

$$c_{p_{m,i}} \Big|_0^t = (1/100)w_{AS,i}\ [\%]\ C_{p_{AS}} \dot{+} (1/100)w_{c,i}\ [\%]\ c_{p_C}\ [\mathrm{kJ/kg°K}]$$

$w_{C,i}$ = mass fraction of carbon in the material (coke, flue dust, etc.)

c_{p_C} = average specific heat of graphite

c_{p_C} = 0.80 kJ/kg°K

$w_{AS,i}$ = mass fraction of ash in material (coke, flue dust, etc.)

$c_{p_{AS}}$ = average specific heat of ash = quartz-sand

$c_{p_{AS}}$ = 0.97 kJ/kg°K

—Average Specific Heat of Charged-in Char

Percent mass fraction of carbon in the char	$w_{C,K}$ = 46.7 wt. %
Percent mass fraction of ash in the char	$w_{AS,K}$ = 13.2 wt. %
Average specific heat of char	$c_{p_{m,K}}$ = 0.7392 kJ/kg°K

—Average Specific Heat of Flue Dust

Percent mass fraction of carbon in flue dust	$w_{C,FLST}$ = 51.2 wt. %
Percent mass fraction of ash in flue dust	$w_{AS,FLST}$ = 46.1 wt. %
Average specific heat of flue dust	$c_{p_{m,FLST}}$ = 0.8568 kJ/kg°K

—Average Specific Heat of Ash

Percent mass fraction of carbon in the ash	$w_{C,ASA}$ = 72.9 wt. %
Percent mass fraction of ash in ash	$w_{AS,ASA}$ = 24.0 wt. %
Average specific heat of ash	$c_{p_{m,ASA}}$ = 0.8160 kJ/kg°K

—Average Specific Heat of Slag

Percent mass fraction of carbon in the slag	$w_{C,SL}$ = 6.7 wt. %
Percent mass fraction of ash in the slag	$w_{AS,SL}$ = 91.4 wt. %
Average specific heat of the slag	$c_{p_{m,SL}}$ = 0.9402 kJ/kg°K

- Enthalpy of Water in Char

Inlet temperature of water in char	t_{H_2O} = 20°C (assumed)
Average specific heat of water	$c_{p_w}\Big\|_0^t$ = 4.186 kG/kg°K (Steam Tables)
Mass flow, water in char	$\dot{m}_{H_2O,K}$ = 52.08 kg/h (Page 5-44)
Enthalpy of water from char	$\Delta H_{H_2O,K}$ = 1.21 kJ/sec
	$\Delta H_{H_2O,K} = 1.04 \cdot 10^3$ kcal/h

- Enthalpy of O_2, Inlet

Inlet temperature of O_2	t_{O_2} = 114.4°C (Page 5-33)
Average specific heat of O_2	$c_{p_{O_2}}\Big\|_0^t$ = 2.710 kJ/kg°K

Mass flow O_2 $\dot{m}_{O_2} = 2916.77$ kg/h (Page 5-44)

Enthalpy O_2 $\Delta H_{O_2} = 317.06$ kJ/sec

$\Delta H_{O_2} = 2.73 \cdot 10^5$ kcal/h

- Enthalpy of N_2, Inlet

  Inlet temperature $t_{N_2} = 114.4°C$ (Page 5-33)

  Average specific heat N_2 $c_{p_{N_2}}\Big|_0^t = 1.025$ kJ/kg°K

  Mass flow, N_2 $\dot{m}_{N_2} = 162.72$ kg/h (Page 5-44)

  Enthalpy of N_2, inlet $\Delta H_{N_2} = 5.30$ kJ/sec

  $\Delta H_{N_2} = 4.56 \cdot 10^3$ kcal/h

- Enthalpy of Water Vapor in Gasifying Oxygen

  Inlet temperature $t_{H_2O,VS} = 114.4°C$ (Page 5-33)

  Specific enthalpy of steam $\Delta h_{SD} = 2710$ kJ/kg (Steam Tables)

  Mass flow, steam in VS $\dot{m}_{H_2O,VS} = 25.84$ kg/h (Page 5-44)

  Enthalpy of steam in VS $\Delta H_{H_2O,VS} = 19.45$ kJ/sec

  $\Delta H_{H_2O,VS} = 1.67 \cdot 10^4$ kcal/h

- Enthalpy of Steam Addition, Inlet

  Pressure of saturated steam $p_D = 5.57$ bars

  Specific enthalpy of saturated steam $\Delta h_D = 2752$ kJ/kg (Steam Tables)

  Mass flow, steam addition $\dot{m}_D = 364.4$ kg/h

  Enthalpy of steam addition $\Delta H_D = 278.56$ kJ/sec

  $\Delta H_D = 2.4 \cdot 10^5$ kcal/h

- Enthalpy of Cooling Water, Inlet

  Inlet temperature $t = 41.52°C$

  Specific enthalpy of cooling water $c_{p_{H_2O}} = 4.186$ kJ/kg°K (Steam Tables)

  Mass flow, cooling water $\dot{m}_{KW} = 235421$ kg/h (Page 5-40)

  Enthalpy of cooling water $\Delta H_{KW} = 11365.78$ kJ/sec

  $\Delta H_{KW} = 9.77 \cdot 10^6$ kcal/h

Enthalpy Leaving.

- Enthalpy of Raw Gas at Gasifier Exit

  $$\Delta H_{RG} = \Sigma \Delta H_i$$

  $$\Delta H_i = m_i c_{p,i}\Big|_0^{t_i} (t_i - t_o)$$

Because they were present only in small amounts, the gas components COS, SO_2, and HCN were included in the calculation as CO_2. Temperature-dependence of the average specific heats:

$$c_{p,CO_2}\ [\mathrm{kJ/kg°K}] = 0.8557 + 3.4150 \cdot 10^{-4} \cdot t + 7.8471 \cdot 10^{-8} \cdot t^2$$

$$c_{p,CO}\ \text{"} = 1.0143 + 1.3749 \cdot 10^{-4} \cdot t - 2.2865 \cdot 10^{-8} \cdot t^2$$

$$c_{p,H_2}\ \text{"} = 14.3670 + 2.2633 \cdot 10^{-4} \cdot t + 19.310 \cdot 10^{-8} \cdot t^2$$

$$c_{p,N_2}\ \text{"} = 1.0117 + 1.2177 \cdot 10^{-4} \cdot t - 1.7183 \cdot 10^{-8} \cdot t^2$$

$$c_{p,H_2S}\ \text{"} = 0.9532 + 1.9014 \cdot 10^{-4} \cdot t - 1.4002 \cdot 10^{-8} \cdot t^2$$

Temperature of raw gas $t_{RG} = 1503.2°C$

Specific heat CO_2 $c_{p,CO_2}\Big|_0^t = 1.5463$ kJ/kg°K

Specific heat CO $c_{p,CO}\Big|_0^t = 1.1693$ kJ/kg°K

Specific heat H_2 $c_{p,H_2}\Big|_0^t = 15.1435$ kJ/kg°K

Specific heat N_2 $c_{p,N_2}\Big|_0^t = 1.559$ kJ/kg°K

Specific heat H_2S $c_{p,H_2S}\Big|_0^t = 1.2074$ kJ/kg°K

CO_2 mass flow in the raw gas before washer-cooler $\dot{m}_{CO_2,RG} = 1057.20$ kg/h (Page 5-46)

CO mass flow in the raw gas before washer-cooler $\dot{m}_{CO,RG} = 4181.32$ kg/h (Page 5-46)

H_2 mass flow in the raw gas before washer-cooler $\dot{m}_{H_2,RG} = 75.27$ kg/h (Page 5-46)

N_2 mass flow in the raw gas before washer-cooler $\dot{m}_{N_2,RG} = 356.07$ kg/h

H_2S mass flow in the raw gas before washer-cooler $\dot{m}_{H_2S,RG} = 87.42$ kg/h (Page 5-46)

Enthalpy of the CO_2 in raw gas $\Delta H_{CO_2} = 682.60$ kJ/sec

Enthalpy of the CO in raw gas $\Delta H_{CO} = 2041.52$ kJ/sec

Enthalpy of the H_2 in raw gas $\Delta H_{H_2} = 475.95$ kJ/sec

Enthaply of the N_2 in raw gas $\Delta H_{N_2} = 171.86$ kJ/sec

Enthalpy of the H_2S in raw gas $\Delta H_{H_2S} = 44.07$ kJ/sec

Enthalpy of raw gas, gasifier exit $\Delta H_{RG} = 3416.0$ kJ/sec

$\Delta H_{RG} = 2.94 \cdot 10^6$ kcal/h

- Enthalpy of Water Vapor in Raw Gas

  Temperature, steam, exit $t_{H_2O,RG}$ = 1503.2°C (Dust measurement)

  Partial pressure of steam p_{SD} = - bar

  Specific enthalpy at (t, p) $c_{p_{RG}}$ = 2.3025 kJ/kg (Steam Tables)

  Mass flow, steam, without quenching water $\dot{m}_{H_2O,RG}$ = 199.33 kg/h (Page 5-44)

  Enthalpy of steam in raw gas $\Delta H_{H_2O,RG}$ = 191.6 kJ/sec

  $\Delta H_{H_2O,RG}$ = $1.65 \cdot 10^5$ kcal/h

- Enthalpy of Flue Dust, Exit

  Temperature, flue dust t_{FLST} = 1503.2°C (Dust measurement)

  Specific heat, flue dust $c_{p_{FLST}}$ = 0.8568 kJ/kg°K (Page 5-58)

  Mass flow, flue dust $\dot{m}_{FLST}$ = 472.66 kg/h (Page 5-44)

  Enthalpy of flue dust, exit ΔH_{FLST} = 169.1 kJ/sec

  ΔH_{FLST} = $1.45 \cdot 10^5$ kcal/h

- Enthalpy of Ash at Ash Discharge

  Temperature, ash t_{ASA} = 1503.2°C (Assumed)

  Specific heat, ash $c_{p_{ASA}}$ = 0.8160 kJ/kg°K (Page 5-58)

  Mass flow, ash $\dot{m}_{ASA}$ = 39.80 kg/h (Page 5-44)

  Enthalpy of ash in ash discharge ΔH_{ASA} = 13.6 kJ/sec

  ΔH_{ASA} = $1.17 \cdot 10^4$ kcal/h

- Enthalpy of Slag, Exit

  Temperature, slag in slag leg t_{SL} = 1300°C (Melting point)

  Specific heat, slag $c_{p_{SL}}$ = 0.9402 kJ/kg°K (Page 5-58)

  Mass flow, slag $\dot{m}_{SL}$ = 112.70 kg/h (Page 5-44)

  Enthalpy of slag, exit ΔH_{SL} = 38.3 kJ/sec

  ΔH_{SL} = $3.29 \cdot 10^4$ kcal/h

- Enthalpy of Cooling Water, Gasifier Exit

  Temperature, cooling water, exit t = 51.24°C

  Specific enthalpy H_2O c_{H_2O} = 4.186 kJ/kg°K (Steam Tables)

  Mass flow, cooling water $\dot{m}_{KW}$ = 235,421 kg/h (Page 5-40)

  Enthalpy of cooling water, exit ΔH_{KW} = 14026.6 kJ/sec

  ΔH_{KW} = $1.206 \cdot 10^7$ kcal/h

Chemically Bound Heat.

$$H_i = m_i \cdot H_{o,i}$$

- Chemically Bound Heat of Char Feed, Inlet
  Upper heating value, char $H_{o,K}$ = 28210.66 kJ/kg
  Mass flow, charged-in char $\dot{m}_K$ = 3146.48 kg/h (Page 5-44)
  Chemically bound heat, charged-in char H_K = 24656.74 kJ/sec
  H_K = $2.1201 \cdot 10^7$ kcal/h

- Chemically Bound Heat, Raw Gas, Exit
  Upper heating value, CO $H_{o,CO}$ = $10.09 \cdot 10^3$ kJ/kg
  Mass flow, CO $\dot{m}_{CO}$ = 4181.32 kg/h (Page 5-46)
  Chemically bound heat, CO H_{CO} = 11719.3 kJ/sec
  H_{CO} = $1.01 \cdot 10^7$ kcal/h
  Upper heating value, H_2 H_{o,H_2} = $142.00 \cdot 10^3$ kJ/kg
  Mass flow, H_2 $\dot{m}_{H_2}$ = 75.27 kg/h (Page 5-46)
  Chemically bound heat, H_2 H_{H_2} = 2968.9 kJ/sec
  H_{H_2} = $2.55 \cdot 10^6$ kcal/h
  Upper heating value H_2S H_{o,H_2S} = $16.547 \cdot 10^3$ kJ/kg
  Mass flow, H_2S $\dot{m}_{H_2S}$ = 87.42 kg/h (Page 5-46)
  Chemically bound heat, H_2S H_{H_2S} = $3.45 \cdot 10^5$ kcal/h
  ΣH_{RG} = 15090.0 kJ/h

- Chemically Bound Heat, Flue Dust, Exit
  Heating value of carbon $H_{O,C}$ = $33.87 \cdot 10^3$ kJ/kg
  C-fraction, flue dust $w_{C,FLST}$ = 51.2 wt. %
  Mass flow, flue dust $\dot{m}_{FLST}$ = 472.66 kg/h (Page 5-44)
  Chemically bound heat, flue dust H_{FLST} = 2274.8 kJ/sec
  H_{FLST} = $1.96 \cdot 10^6$ kcal/h

- Chemically Bound Heat, Ash in Slag Extractor, Exit
  Heating value of carbon $H_{O,C}$ = $33.87 \cdot 10^3$ kJ/kg
  C-fraction, ash discharge $w_{C,ASA}$ = 72.9 wt. %
  Mass flow, ash discharge $\dot{m}_{ASA}$ = 39.8 kg/h (Page 5-44)
  Chemically bound heat H_{ASA} = 372.7 kJ/sec
  H_{ASA} = $2.34 \cdot 10^5$ kcal/h

- Chemically Bound Heat, Slag, Exit
  Heating value of carbon $H_{O,C} = 33.87 \cdot 10^3$ kJ/kg
  C-fraction, slag $w_{C,SL} = 6.7$ wt. %
  Mass flow, slag $\dot{m}_{SL} = 112.70$ kg/h (Page 5-44)
  Chemically bound heat, slag $H_{SL} = 71.0$ kJ/sec
  $H_{SL} = 6.10 \cdot 10^4$ kcal/h

Heat Balance, Total Heat.

$$Q = \Sigma(\Delta H_i + H_i)$$

- Heat Flow, Inlet

Total heat, in char feed	Q_K	= 24669.66	kJ/sec
Total heat, H_2O from char	$Q_{H_2O,K}$	= 1.21	kJ/sec
Total heat, O_2	Q_{O_2}	= 317.06	kJ/sec
Total heat, N_2	Q_{N_2}	= 5.30	kJ/sec
Total heat, H_2O in oxygen feed	$Q_{H_2O,VS}$	= 19.45	kJ/sec
Total heat, steam addition	Q_D	= 278.56	kJ/sec
Total heat, cooling water	Q_{KW}	= 11365.78	kJ/sec
Total heat, inlet	Q_{in}	= 36657.02	kJ/sec

- Heat Flow, Exit

Total heat, raw gas	Q_{RG}	= 18506.02	kJ/sec
Total heat, H_2O in raw gas	$Q_{H_2O,RG}$	= 191.6	kJ/sec
Total heat, flue dust	Q_{FLST}	= 2443.9	kJ/sec
Total heat, ash discharge	Q_{ASA}	= 286.3	kJ/sec
Total heat, slag	Q_{SL}	= 109.2	kJ/sec
Total heat, cooling water	Q_{KW}	= 14026.6	kJ/sec
Total heat, losses	Q_{VL}	= 1093.4	kJ/sec
Total heat, exit	Q_{out}	= 36657.02	kJ/sec

Ratio: $Q_{VL} : Q_{in} = \eta_Q = 0.0298$

Gasification Results

Degree of C-Gasification.

$$\eta_C = 100\ \dot{m}_{C,RG}/\dot{m}_{C,K} \quad [\%]$$

C-mass flow, raw gas $\dot{m}_{C,RG} = 2088.51$ kg/h (Page 5-46)
C-mass flow, char $\dot{m}_{C,K} = 2403.91$ kg/h (Page 5-52)

$\eta_C = 86.88$ [%]

Degree of Steam Decomposition.

$$\varepsilon = 100(1 - \dot{m}_{H_2O,out}/\dot{m}_{H_2O,in}) \quad [\%]$$

- Mass Flow, H_2O, Inlet

  H_2O-char $\dot{m}_{H_2O,K}$ = 52.08 kg/h (Page 5-44)

  H_2O-oxygen $\dot{m}_{H_2O,O_2}$ = 25.84 kg/h (Page 5-44)

  H_2O-steam $\dot{m}_D$ = 362.34 kg/h (Page 5-44)

  Total H_2O $\dot{m}_{H_2O,in}$ = 440.26 kg/h

- Mass Flow, H_2O, Exit

  H_2O-raw gas (without quenching water) $\dot{m}_{H_2O,RG}$ = 199.33 kg/h (Page 5-44)

  Total H_2O $\dot{m}_{H_2O,out}$ = 199.33 kg/h

ε = 54.72 [%]

Degree of Hydrogen Decomposition.

$$\eta_H = 100\ \dot{m}_{H,RG}/\dot{m}_{H,in} \quad [\%]$$

H-mass flow, raw gas, exit (from H_2, H_2S, and HCN) $\dot{m}_{H,RG}$ = 80.46 kg/h

H-mass flow, inlet $\dot{m}_{H,in}$ = 102.76 kg/h

η_H = 78.30%

Degree of Hydrogen Formation.

$$\eta'_H = 100\ \dot{m}_{H,RG}/\dot{m}_{H,K} \quad [\%]$$

H-mass flow, raw gas, exit (from H_2, H_2S and HCN) $\dot{m}_{H,RG}$ = 80.46 kg/h

H-mass flow, char $\dot{m}_{H,K}$ = 53.49 kg/h

η'_H = 150.42%

Efficiency of Thermal Gasification (Cold Gas Efficiency).

$$\eta_{th,V} = H_{O,gas}/H_{O,K}$$

Chemically bound heat, char $H_{O,K}$ = 24656.7 kJ/sec

Chemically bound heat, raw gas $H_{O,gas}$ = 15090.0 kJ/sec

Thermal gasification efficiency $\eta_{th,V}$ = 0.6120

Specific Heat Loss of Gasifier.

$$q = \Delta Q_{KW}/\dot{V}_{N,(CO + H_2)}$$

Enthalpy, cooling water, gasifier inlet	$Q_{KW,in}$	= 11365.8 kJ/sec
Enthalpy, cooling water, gasifier exit	$Q_{KW,out}$	= 14026.6 kJ/sec
Increase in enthalpy, cooling water, gasifier	ΔQ_{KW}	= 2660.0 kJ/sec
	ΔQ_{KW}	= $2.288 \cdot 10^6$ kcal/h
Volume percent $CO + H_2$	$\nu_{CO + H_2}$	= 82.42 vol. %
Normal volume flow, raw gas from washer-cooler	$\dot{V}_{N,RG}$	= 3080.57 Nm^3/h
Normal volume flow $CO + H_2$ before washer-cooler	$\dot{V}_{N,(CO + H_2)}$	= 4198 Nm^3/h
Specific heat loss, gasifier	q	= 2287.2 kJ/Nm^3
	q	= 546.4 $kcal/Nm^3$

Tabular Representation of the Char Tests

Western Kentucky Char.

Tables 5-10 to 5-15 list in consecutive order the data and results from the five test runs with Western Kentucky char. All tests were made under steady and stable operating conditions in order to ensure reliable data. The time required to reach stable conditions and to adjust the screw feed for steady flow is not included as part of the test period.

Table 5-10 covers the analyses and feed rates of the char, oxygen, and steam to the gasifier.

Table 5-11 gives the raw gas analysis, including trace elements, and flow rate after the gasifier but before the washer-cooler.

Table 5-12 lists the solid discharges from the gasifier, giving the analyses and mass flow rates for flue dust, ash discharge, slag discharge, and build-up of gasifier residue.

Table 5-13 gives the once-through wash water analysis, both to and from the washer-cooler. The analysis includes flow rates, sulfur analysis as H_2S and SO_2, and HCN, both fixed and free.

Table 5-14 gives specific gasification criteria; the gasifier exit temperature; the useful gas make; the specific heat loss; the ratio by weight of crude gas, oxygen, and steam to char; the useful volume of gas to char weight; and the carbon conversion, thermal gasifying efficiency, and heating value of the char.

Table 5-15 gives the H-formation (hydrogen in the gas/hydrogen in the char), the oxygen and preliminary mass balances (these two balances are of primary importance in assessing the overall accuracy of the test results), the sulfur balance, and heat-loss percentage.

Pittsburgh Char.

Tables 5-16 to 5-21 summarize the data and results for the Pittsburgh char in the same order as the Western Kentucky char.

Table 5-10

ANALYSES AND COMPUTED RESULTS RAW MATERIAL FOR GASIFYING WESTERN KENTUCKY CHAR

COAL	TEST NO.	DURATION OF TEST	CHAR									GASIFYING OXYGEN				STEAM
			ANALYSES							MASS FLOW	HEATING VALUE	ANALYSES		VOL. FLOW	MASS FLOW	MASS FLOW
			w_{H_2O}	w_C	w_H	w_O	w_N	w_S	w_{AS}	$\dot{m}_K$	$H_{O,K}$	O_2	N_2	V_{N,O_2}	$\dot{m}_{O_2}$*	$\dot{m}_D$
		h	wt%	wt%	wt%	wt%	wt%	wt%	wt%	kg/h	kcal /kg	Vol.%	Vol.%	Nm^3/h	kg/h	kg/h
W.Kentucky	019	1.50	1.63	75.0	1.7	2.3	1.6	3.0	16.4	2463	6600	93.35	7.65	1797	2427	229
W.Kentucky	020	2.00	2.33	77.1	1.6	1.9	1.7	3.0	14.7	2667	6771	89.7	10.3	1914	2718	517
W.Kentucky	021	2.00	2.33	77.1	1.6	1.9	1.7	3.0	14.7	3047	6771	89.0	11.0	1863	2656	259
W.Kentucky	022	2.00	2.25	75.6	1.7	2.4	1.6	3.2	15.5	2939	6662	94.5	5.5	1880	2605	365
W.Kentucky	028	2.00	2.25	75.6	1.7	2.4	1.6	3.2	15.5	2630	6662	94.0	6.0	2088	2870	715

*Pure oxygen

Table 5-11

ANALYSES AND COMPUTED RESULTS RAW GAS FROM GASIFICATION, WESTERN KENTUCKY CHAR

CHAR	TEST NO.	RAW GAS BEFORE WASHER-COOLER												TRACES IN RAW GAS				
		Analyses									Vol. Flow	Mass Flow	Heating Value	Specific Mass Flow				
		H_2O	CO_2	CO	H_2	N_2	H_2S	COS	SO_2	HCN	$V_{N,RG}$	$\dot{m}_{RG}$	$H_{O,RG}$	w'_{H_2S}	w'_{COS}	w'_{SO_2}	w'_{HCN}	w'_S
		Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Nm^3/h	kg/h	kcal/kg	mg/Nm^3_{tr}	mg/Nm^3_{tr}	mg/Nm^3_{tr}	mg/Nm^3_{tr}	mg/Nm^3_{tr}
W.Kentucky	019	5.46	12.10	66.74	13.76	6.34	0.91	0.12	0.02	0.01	4088	4828	2101.5	16841	4615	706	59.0	1759
W.Kentucky	020	9.97	16.68	59.10	15.62	7.59	0.86	0.12	0.02	0.006	4443	5301	1933.8	17589	4362	1020	100.7	1738
W.Kentucky	021	4.04	9.01	66.87	15.66	7.40	0.91	0.12	0.02	0.009	4689	5338	2237.3	17294	4978	1392	123.4	1784
W.Kentucky	022	8.10	10.73	67.07	15.71	5.38	0.91	0.15	0.04	0.007	4465	5136	2219.6	18296	4067	1189	86.0	2168
W.Kentucky	028	14.71	21.33	56.16	15.89	5.66	0.79	0.12	0.06	0.005	4472	5470	1816.0	17327	5541	708	173.4	1664

Table 5-12

ANALYSES AND COMPUTED RESULTS SOLIDS FROM GASIFICATION, WESTERN KENTUCKY CHAR

CHAR	TEST NO.	FLUE DUST				ASH DISCHARGE				SLAG				GASIFIER RESIDUE			
		Analysis, Dry			Mass Flow	Analysis, Dry			Mass Flow	Analysis, Dry			Mass Flow	Analysis, Dry			Mass Flow
		w_C	w_S	w_{AS}	$\dot{m}_{FLST}$	w_C	w_S	w_{AS}	$\dot{m}_{ASA}$	w_C	w_S	w_{AS}	$\dot{m}_{SL}$	w_C	w_S	w_{AS}	$\dot{m}_{RV}$
		Wt.%	Wt.%	Wt.%	kg/h	Wt.%	Wt.%	Wt.%	kg/h	Wt.%	Wt.%	Wt.%	kg/h	Wt.%	Wt.%	Wt.%	kg/h
W.Kentucky	019	34.4	1.2	64.2	272	63.1	1.4	34.8	33	3.0	0.5	96.5	16.1	63.1	1.4	34.8	19.0
W.Kentucky	020	45.8	1.2	52.9	416	71.2	1.4	25.4	58	7.0	0.7	92.3	81.4	71.2	1.4	25.4	13.7
W.Kentucky	021	59.6	1.7	38.5	644	77.1	1.6	20.4	103	5.9	0.5	93.6	90.0	77.1	1.6	20.4	16.3
W.Kentucky	022	54.2	2.0	43.6	554	71.6	1.8	26.5	52	8.2	2.1	89.7	62.0	71.6	1.8	26.5	16.3
W.Kentucky	028	33.3	1.2	65.3	266	65.9	1.7	32.3	32	10.2	1.0	88.8	62.0	65.9	1.7	32.3	16.3

Table 5-13

ANALYSES AND COMPUTED RESULTS WASH WATER
WESTERN KENTUCKY CHAR

CHAR	TEST NO.	WASH WATER FOR WASHER-COOLER								
		Inlet				Exit				
		Specific Mass Flow				Specific Mass Flow				Volume Flow
		$W^{\circ}H_2S$	$W^{\circ}HCN$ free	$W^{\circ}HCN$ fixed	$W^{\circ}SO_2$	$W''H_2S$	$W''HCN$ free	$W''HCN$ fixed	$W''SO_2$	$\dot{V}H_2O,KW$
		mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	m^3/h
W. Kentucky	019	-	0.002	1	-	30.72	1.27	12.42	49.52	38.41
W. Kentucky	020	-	0.002	1	-	34.95	1.68	7.56	144.98	33.06
W. Kentucky	021	-	0.002	1	-	93.90	2.59	10.80	22.29	32.99
W. Kentucky	022	-	0.002	1	-	128.30	2.46	3.60	41.25	34.74
W. Kentucky	028	-	0.002	1	-	28.50	0.95	10.00	124.79	32.50

Table 5-14

COMPUTED RESULTS WESTERN KENTUCKY CHAR

CHAR	TEST NO.	TEMP.	USEFUL GAS	RESULTS							
		Gasifier	Volume Flow	Specific Heat Loss	Crude Gas / Char	O_2 / Char	Steam / Char	CO + H_2 / Char	Carbon Gasification	Thermal Gasifying Efficiency	HO,RG / Char
		tv	CO + H_2	q							
		°C	Nm^3/h	kcal/ Nm^3CO + H_2	kg/kg	kg/kg	kg/kg	Nm^3/kg	%	%	kJ/kg
W. Kentucky	019	1590	3290	770.3	1.9602	0.9854	0.0930	1.3358	93.71	62.42	17246
W. Kentucky	020	1383	3320	823.5	1.9876	1.0191	0.1939	1.2735	88.02	56.77	16091
W. Kentucky	021	1238	3870	604.5	1.7519	0.8717	0.0850	1.2701	81.44	57.88	16406
W. Kentucky	022	1409	3696	633.2	1.7475	0.8864	0.1242	1.2576	84.07	58.22	16237
W. Kentucky	028	1490	3222	962.2	2.0798	1.0913	0.2719	1.2251	93.65	56.70	15812

Table 5-15

COMPUTED RESULTS EFFICIENCIES, WESTERN KENTUCKY CHAR

CHAR	TEST NO.	EFFICIENCY				
		H-Formation	O-Balance	S-Balance	Q_{loss}/Q_{in}	Preliminary Mass Balance
		H %	O %	S %	Heat Loss %	%
W. Kentucky	019	127.94	103.8	88.5	0.4	97.37
W. Kentucky	020	153.43	99.6	88.1	2.6	99.70
W. Kentucky	021	142.84	98.7	92.0	1.5	99.89
W. Kentucky	022	132.75	103.4	89.3	2.8	98.62
W. Kentucky	028	149.03	102.0	79.0	2.7	97.95

Table 5-16

ANALYSES AND COMPUTED RESULTS RAW MATERIALS FOR GASIFYING, PITTSBURGH CHAR

COAL	TEST NO.	DURATION OF TEST	CHAR									GASIFYING OXYGEN				STEAM
			ANALYSES							MASS FLOW	HEAT VALUE	ANALYSES		VOL. FLOW	MASS FLOW	MASS FLOW
			w_{H_2O}	w_C	w_H	w_O	w_N	w_S	w_{AS}	$\dot{m}_K$	$H_{O,K}$	O_2	N_2	$\dot{V}_{N,O_2}$	$\dot{m}_{O_2}$ *	$\dot{m}_D$
		h	Wt.%	Wt.%	Wt.%	Wt.%	Wt.%	Wt.%	Wt.%	kg/h	kcal/kg	Vol.%	Vol.%	Nm^3/h	kg/h	kg/h
Pittsburgh	023	2.00	1.80	75.0	1.9	2.3	1.4	4.1	15.3	2495	6723	92.60	7.40	1860	2498	697
Pittsburgh	024	2.00	1.80	75.0	1.9	2.3	1.4	4.1	15.3	3009	6723	93.65	6.35	1922	2714	428
Pittsburgh	025	2.00	1.80	75.0	1.9	2.3	1.4	4.1	15.3	3245	6723	93.65	6.35	1916	2607	254
Pittsburgh	026	2.50	1.63	76.4	1.7	3.3	1.4	4.0	13.2	3244	6738	93.70	6.30	1964	2778	374
Pittsburgh	027	2.50	1.63	76.4	1.7	3.3	1.4	4.0	13.2	3146	6738	93.93	6.07	2093	2917	362

*Pure oxygen

Table 5-17

ANALYSES AND COMPUTED RESULTS RAW GAS FROM GASIFICATION, PITTSBURGH CHAR

CHAR	TEST NO.	RAW GAS BEFORE WASHER-COOLER												TRACES IN RAW GAS				
		Analysis									Vol. Flow	Mass Flow	Heating Value	Specific Mass Flow				
		H_2O	CO_2	CO	H_2	N_2	H_2S	COS	SO_2	HCN	$V_{N,RG}$	$\dot{m}_{RG}$	$H_{O,RG}$	w'_{H_2S}	w'_{COS}	w'_{SO_2}	w'_{HCN}	w'_S
		Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Vol.%	Nm^3/h	kg/h	kcal/kg	mg/Nm^3_{tr}	mg/Nm^3_{tr}	mg/Nm^3_{tr}	mg/Nm^3_{tr}	mg/Nm^3_{tr}
Pittsburgh	023	15.92	18.59	56.64	17.52	5.96	1.09	0.17	0.02	0.005	4070	4831	1943.3	16841	4615	706	59.04	2463
Pittsburgh	024	9.56	11.04	65.18	16.63	5.80	1.14	0.16	0.04	0.009	4506	5150	2220.7	17589	4362	1020	100.7	2327
Pittsburgh	025	4.98	5.87	70.56	17.01	5.19	1.12	0.18	0.05	0.010	4776	5261	2460.9	17294	4978	1392	123.4	2657
Pittsburgh	026	3.93	8.16	66.69	17.64	6.12	1.19	0.15	0.04	0.007	5057	5616	2357.6	18296	4067	1189	86.0	2171
Pittsburgh	027	4.65	10.59	65.83	16.59	5.61	1.13	0.21	0.02	0.001	5081	5790	2241.2	172327	5541	708	173.4	2957

Table 5-18

ANALYSES AND COMPUTED RESULTS SOLIDS FROM GASIFICATION, PITTSBURGH CHAR

COAL	TEST NO.	FLUE DUST				ASH DISCHARGE				SLAG				GASIFIER RESIDUE			
		Analysis, Dry			Mass Flow	Analysis, Dry			Mass Flow	Analysis, Dry			Mass Flow	Analysis, Dry			Mass Flow
		w_C	w_S	w_{AS}	$\dot{m}_{FLST}$	w_C	w_S	w_{AS}	$\dot{m}_{ASA}$	w_C	w_S	w_{AS}	$\dot{m}_{SL}$	w_C	w_S	w_{AS}	$\dot{m}_{RV}$
		Wt. %	Wt. %	Wt. %	kg/h	Wt. %	Wt. %	Wt. %	kg/h	Wt. %	Wt. %	Wt. %	kg/h	Wt. %	Wt. %	Wt. %	kg/h
Pittsburgh	023	45.6	2.4	51.8	357	67.9	1.9	29.9	46	7.0	0.9	92.1	40	67.9	1.9	29.9	52
Pittsburgh	024	50.8	2.4	51.8	608	74.6	2.3	23.0	75	5.3	1.0	93.7	127	74.6	2.3	23.0	52
Pittsburgh	025	54.3	3.4	42.0	684	75.3	2.6	21.5	73	6.8	1.2	92.0	73	75.3	2.6	21.5	52
Pittsburgh	026	55.8	3.0	41.2	618	76.0	2.4	20.6	70	6.5	1.4	92.1	97	76.0	2.4	20.6	52
Pittsburgh	027	51.2	2.7	46.1	473	72.9	2.2	24.0	40	6.7	1.9	91.4	113	72.9	2.2	24.0	52

Table 5-19

ANALYSES AND COMPUTED RESULTS WASH WATER PITTSBURGH CHAR

CHAR	TEST NO.	WASH WATER FOR WASHER-COOLER								
		Inlet				Exit				
		Specific Mass Flow				Specific Mass Flow				Volume Flow
		W°H_2S	W°HCN free	W°HCN fixed	W°SO_2	W"H_2S	W"HCN free	W"HCN fixed	W"SO_2	$\dot{V}$ H_2O,KW
		mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	m^3/h
Pittsburgh	023	-	0.002	1	-	60.70	1.64	7.92	96.21	36.90
Pittsburgh	024	-	0.002	1	-	280.00	1.81	8.58	45.46	35.06
Pittsburgh	025	-	0.002	1	-	134.70	1.69	-	53.33	32.97
pittsburgh	026	-	0.002	1	-	117.30	1.94	5.32	28.60	31.63
Pittsburgh	027	-	0.002	1	-	80.90	1.38	4.71	43.04	34.81

Table 5-20

COMPUTED RESULTS-PITTSBURGH CHAR

CHAR	TEST NO.	TEMP.	USEFUL GAS	RESULTS							
		Gasifier	Vol. Flow	Specific Heat Loss	Crude Gas/Char	O_2/Char	Steam/Char	CO + H_2/Char	Carbon Gasification	Thermal Gasifying Efficency	HO,RG/Char
		tv	CO + H_2	q							
		°C	Nm^3/h	kcal/Nm^3 CO + H_2	kg/kg	kg/kg	kg/kg	Nm^3/kg	%	%	kJ/kg
Pittsburgh	023	1439	3018	900.7	1.9363	1.0012	0.2794	1.2096	88.10	55.97	15753
Pittsburgh	024	1413	3686	752.2	1.7115	0.9020	0.1422	1.2250	81.8	56.52	15910
Pittsburgh	025	1442	4182	459.8	1.6213	0.8034	0.0783	1.2888	80.69	59.33	16700
Pittsburgh	026	1411	4265	456.6	1.7312	0.8564	0.1153	1.3147	82.15	60.55	17082
Pittsburgh	027	1503	4188	546.4	1.8404	0.9272	0.1151	1.3312	86.88	61.20	17265

Table 5-21

COMPUTED RESULTS EFFICIENCIES PITTSBURGH CHAR

CHAR	TEST NO.	EFFICIENCY				
		H-Formation	O-Balance	S-Balance	Q_{loss}/Q_{in}	Preliminary Mass Balance
		H	O	S	Heat Loss %	%
Pittsburgh	023	142.97	101.6	083.6	2.46	96.81
Pittsburgh	024	125.21	098.1	085.4	1.94	99.43
Pittsburgh	025	125.59	100.6	090.2	2.84	97.90
Pittsburgh	026	154.41	096.6	094.7	2.35	99.53
Pittsburgh	027	150.42	098.0	099.1	2.98	99.00

Section 6

DISCUSSION OF TEST RESULTS

Because two COED chars, Western Kentucky and Pittsburgh, have similar elemental analyses and yield similar results, they will be discussed together in this section.

All together, five experiments were carried out on Pittsburgh char and five on Western Kentucky char. For the Pittsburgh char, the ratio kg O_2/kg char was investigated over the range 0.80 to 1.0 approximately, and the ratio kg steam/kg char over the range 0.08 to 0.28, approximately. The corresponding ranges investigated for Western Kentucky char were 0.88 to 1.1 and 0.03 to 0.27, respectively. A long-term test of over 30 hours was carried out with Western Kentucky char.

The most important results obtained for Western Kentucky char are presented in Tables 5-10 to 5-15, and those for Pittsburgh char are presented in Tables 5-16 to 5-21. Figures 6-1 to 6-10 contain graphical presentations of the characteristic gasification data, which will be discussed below.

COMPOSITION OF RAW GAS AS A FUNCTION OF THE O_2/CHAR RATIO

Figures 6-1 and 6-2 show the composition of the raw gas as a function of the O_2/char ratio. The CO_2 percentage of the raw gas increases with increasing O_2/char ratio, while the CO percentage decreases in the same proportion. The H_2 fraction remains almost constant.

GASIFIER TEMPERATURE AS A FUNCTION OF THE O_2/CHAR RATIO

Figures 6-3 and 6-4 show the dependence of the gasification temperature on the O_2/char ratio, with the steam/char ratio as a parameter. An increased addition of steam at the same O_2/char ratio lowers the temperature of the gasifier. Correspondingly, the temperature rises with greater oxygen supplied at a constant steam/char ratio.

GAS MASS FLOWS PER kg CHAR AS A FUNCTION OF THE O_2/CHAR RATIO

Figures 6-5 and 6-6 illustrate the dependence of the raw gas produced per kg of char on the O_2/char ratio. The mass flow of raw gas produced increases with increasing O_2/char ratio. In the steam/char ratio range considered, no decisive influence of steam addition is to be seen. It will be noted, however, that there is an increase in the production of CO_2 with increasing O_2/char ratio (see Figures 6-1 and 6-2).

The mass flow of useful gas produced, $CO + H_2$, per kg of char clearly has a maximum and is influenced by the temperature in the gasifier, which, apart from the O_2/char ratio, is a function of the steam/char ratio.

QUANTITIES OF GAS PRODUCED PER kg OF CHAR AS A FUNCTION OF THE O_2/CHAR RATIO

Figures 6-7 and 6-8 illustrate the volume of gas produced per kg of char as a function of the O_2/char ratio. The same relationships are obtained as described in the preceeding section on gas mass flows.

The yield of $CO + H_2$/kg char is the most important characteristic for the production of synthesis gas. Because the combustible constituents of the gas produced are almost exclusively $CO + H_2$, this index is also a measure of the calorific yield per kg of char.

GASIFICATION OF C AS A FUNCTION OF THE O_2/CHAR RATIO

The ratio of carbon in the raw gas to the carbon in the gasified char (C-conversion) is plotted as a function of the O_2/char ratio in Figures 6-9 and 6-10. The C-conversion increases with increasing O_2 supply, but increasing amounts of CO_2 are also formed.

CONCLUSIONS CONCERNING THE GASIFICATION OF COED CHARS

The large-scale experiments carried out at Puentes have shown that COED char can be gasified with good results in Koppers-Totzek gasifiers. To a large extent, the gas produced in the experiments was additionally processed to ammonia in the existing installations.

On the basis of the research results and the experience of Krupp-Koppers G.M.B.H., Essen, a useful-gas yield of about 1.4 Nm^3 of $CO + H_2$ per kg of char can be expected in a large modern Koppers-Totzek gasifier.

GRINDING OF CHARS

The grinding of the COED chars was undertaken in the existing grinding installation of the Puentes plant, which had been designed for the grinding of lignite. To achieve a ground char with the degree of fineness required for satisfactory gasification, the grinding installation had to be operated at a reduced throughput. The energy consumptions observed with this mode of operation were practically indistinguishable from the energy consumption at no load. Thus, no conclusions were possible concerning the specific energy demand for the grinding of char on site. As a result, it can, however, be stated that the grinding of the chars to the desired fineness depends only on a properly designed ball mill and classification installations.

GASIFIER LINING

For purposes of gasifying both petroleum cokes and COED chars, a corundum lining was selected for the gasifier so that operating temperatures, above those normally required for lignite, could be investigated. Due to the extreme range of test conditions, the corundum lining was partially washed away by the liquid slag made during gasification of the chars.

On the basis of Krupp-Koppers G.M.B.H., Essen, research and operating experience, and knowledge gained from the tests, a suitable lining and operating practice would be employed in future gasification of chars to resist the slag and associated temperature conditions.

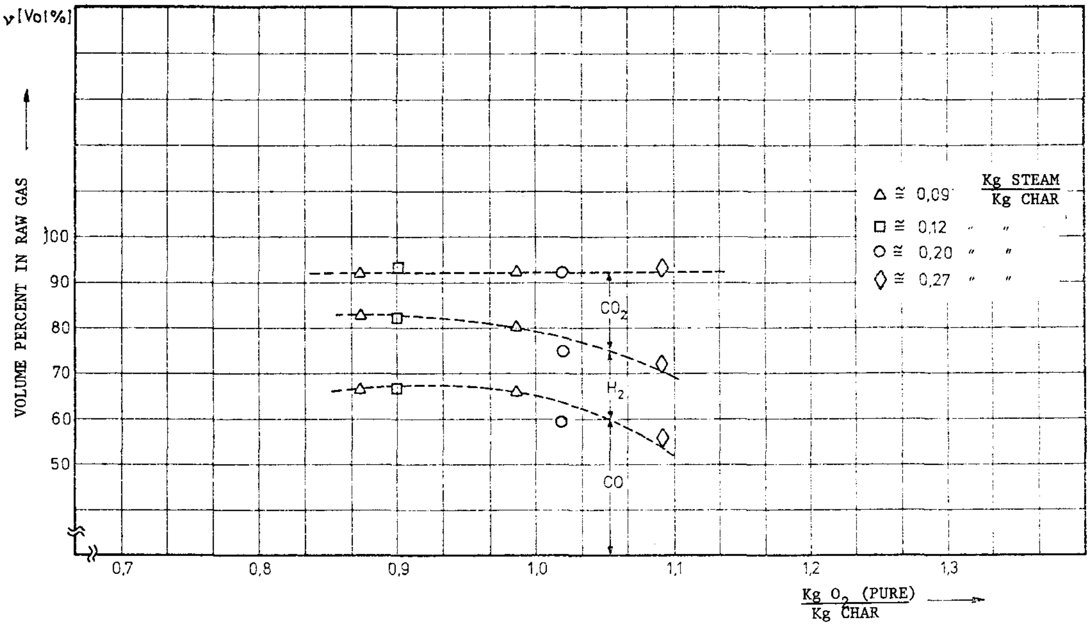

Figure 6-1. Composition of Raw Gas (CO_2, H_2, and CO, Volume %), as a Function of O_2/Char Ratio, Western Kentucky Char

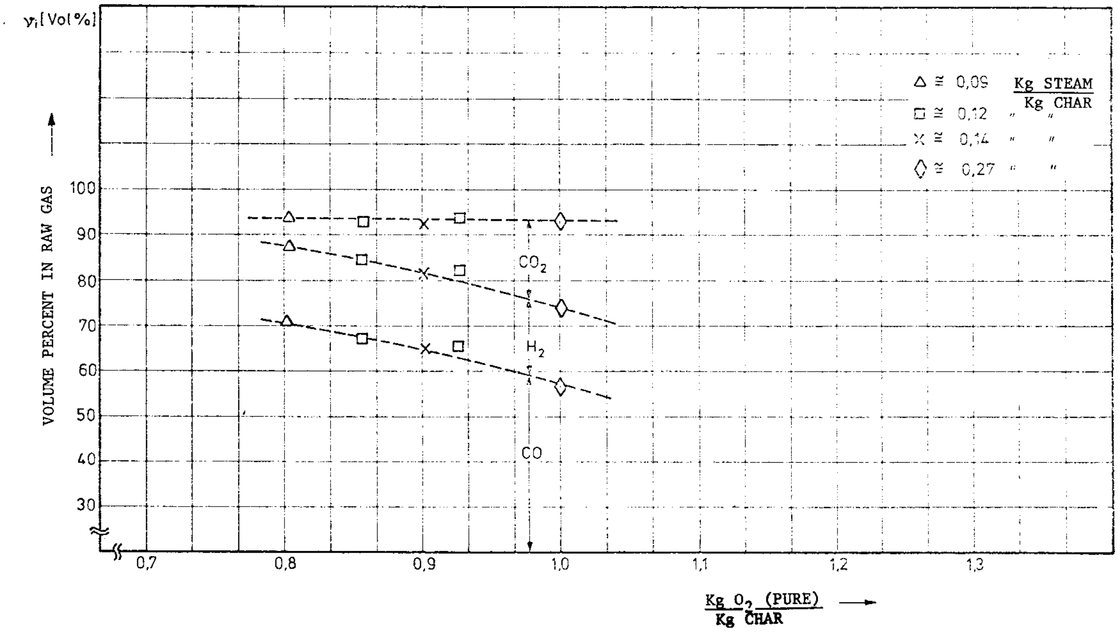

Figure 6-2. Composition of Raw Gas (CO_2, H_2, and CO, Volume %), as a Function of O_2/Char Ratio

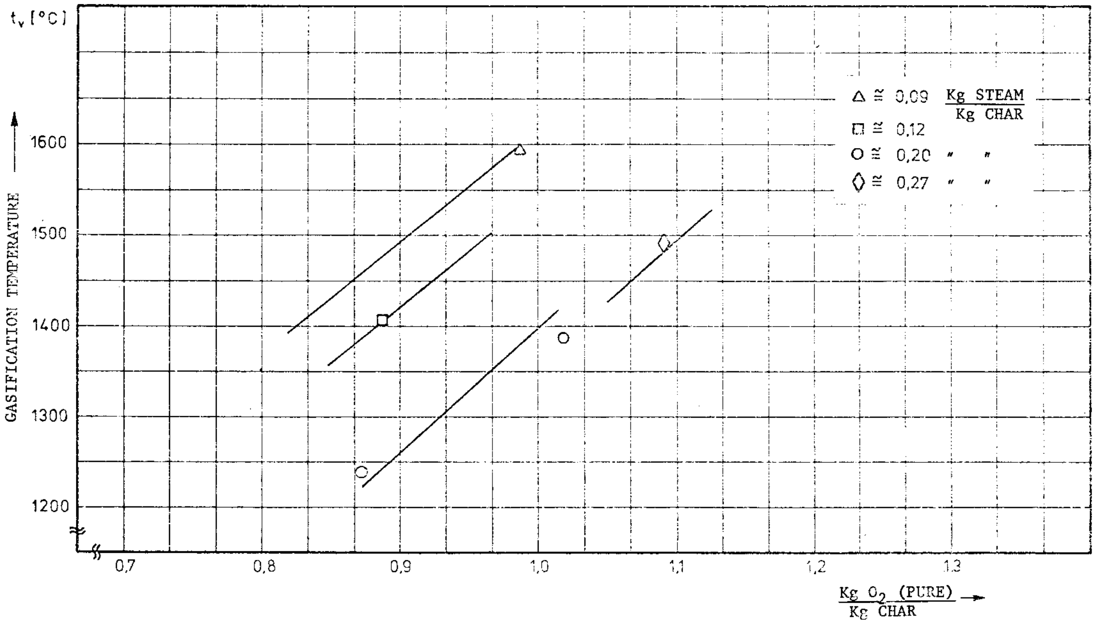

Figure 6-3. Gasifying Temperature as a Function of O_2/Char Ratio, Western Kentucky Char

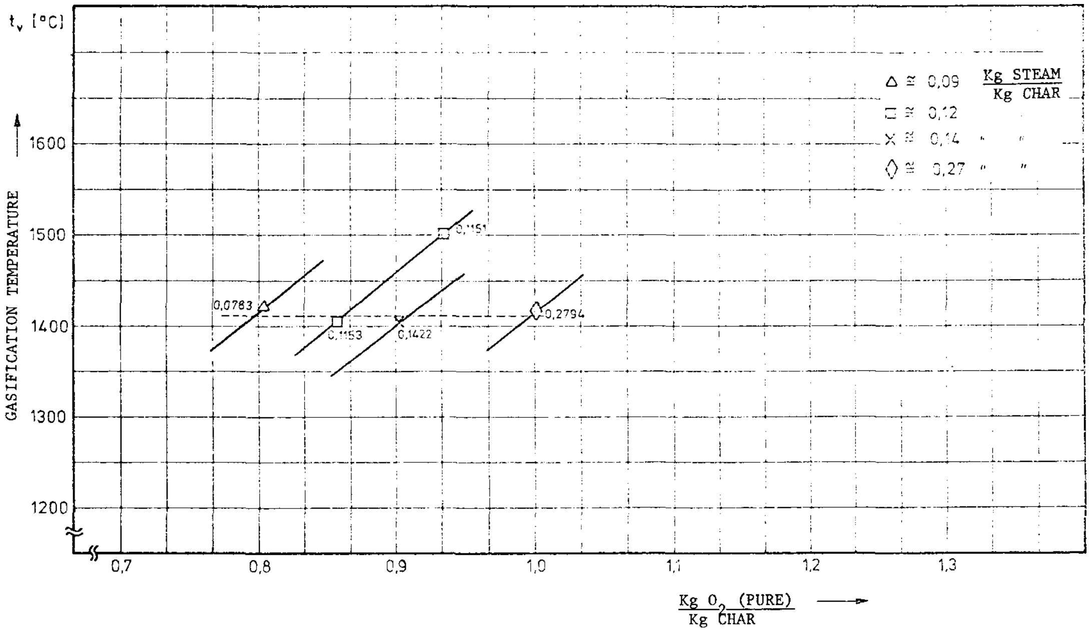

Figure 6-4. Gasifying Temperature as a Function of O_2/Char Ratio, Pittsburgh Char

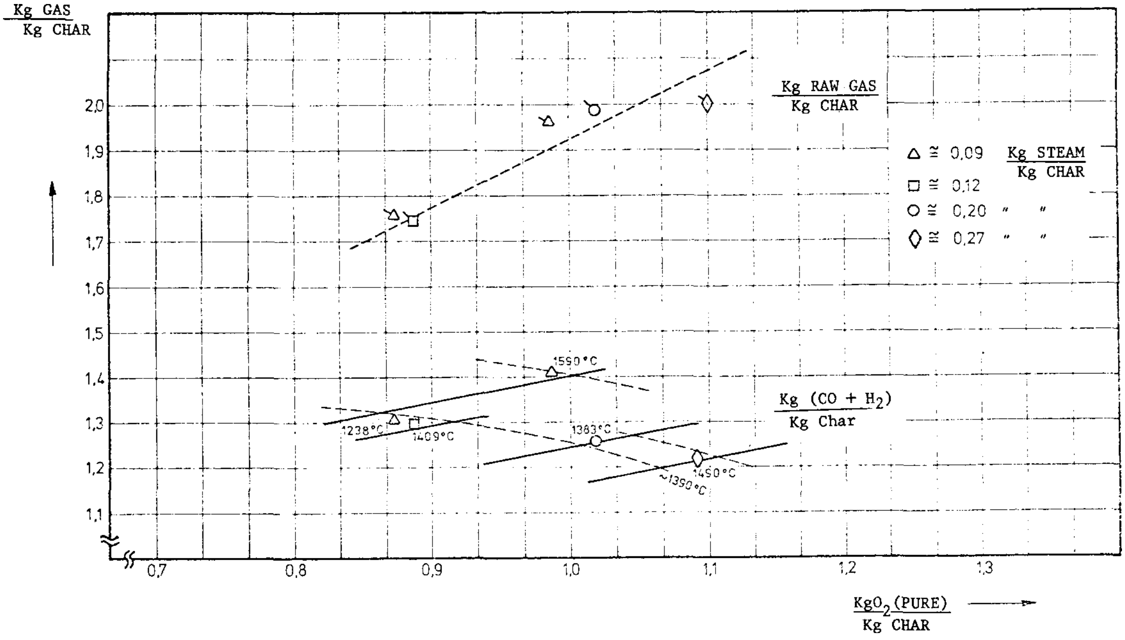

Figure 6-5. Gas Mass Flow as a Function of O_2/Char Ratio, Western Kentucky Char

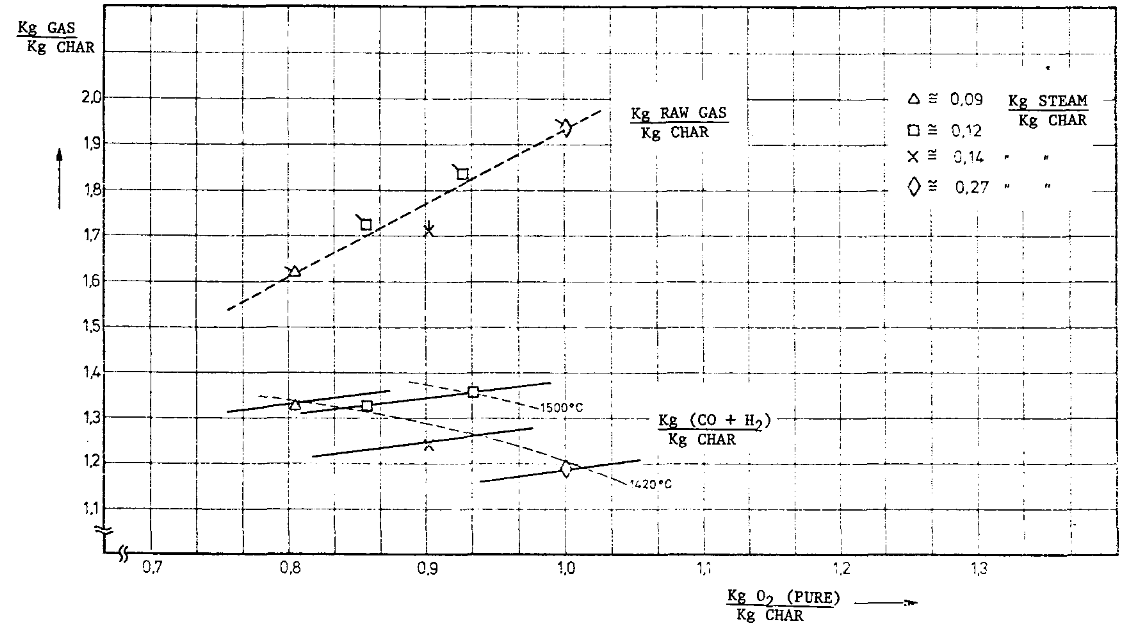

Figure 6-6. Gas Mass Flow as a Function of O_2/Char Ratio, Pittsburgh Char

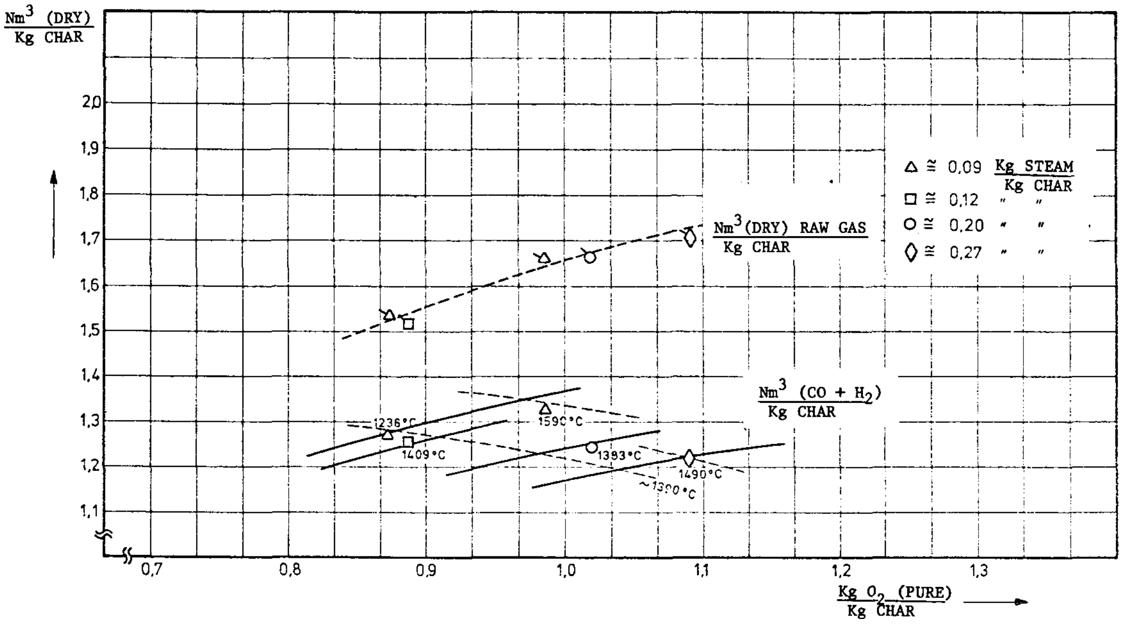

Figure 6-7. Gas Volume Generated per kg Char as a Function of O_2/Char Ratio, Western Kentucky Char

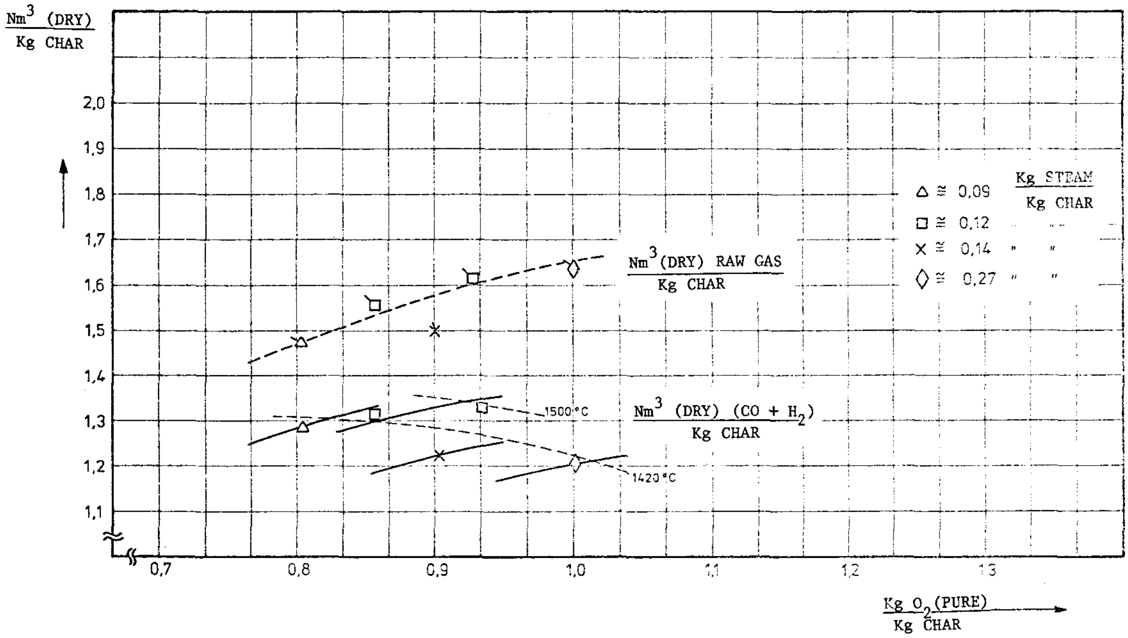

Figure 6-8. Gas Volume Generated per kg Char as a Function of O_2/Char Ratio, Pittsburgh Char

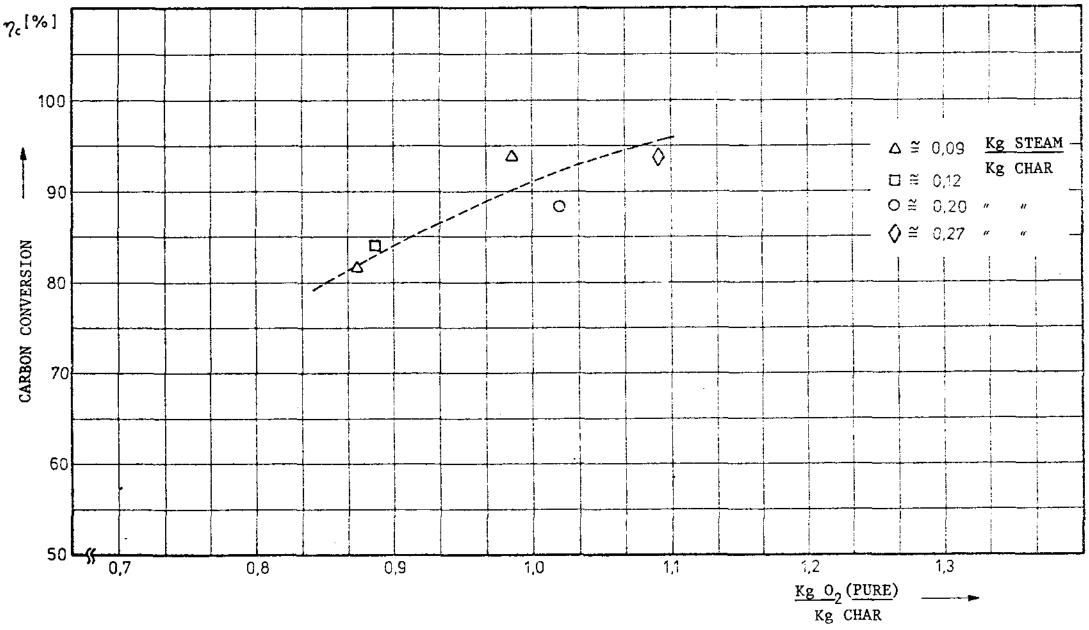

Figure 6-9. Carbon Conversion as a Function of O_2/Char Ratio, Western Kentucky Char

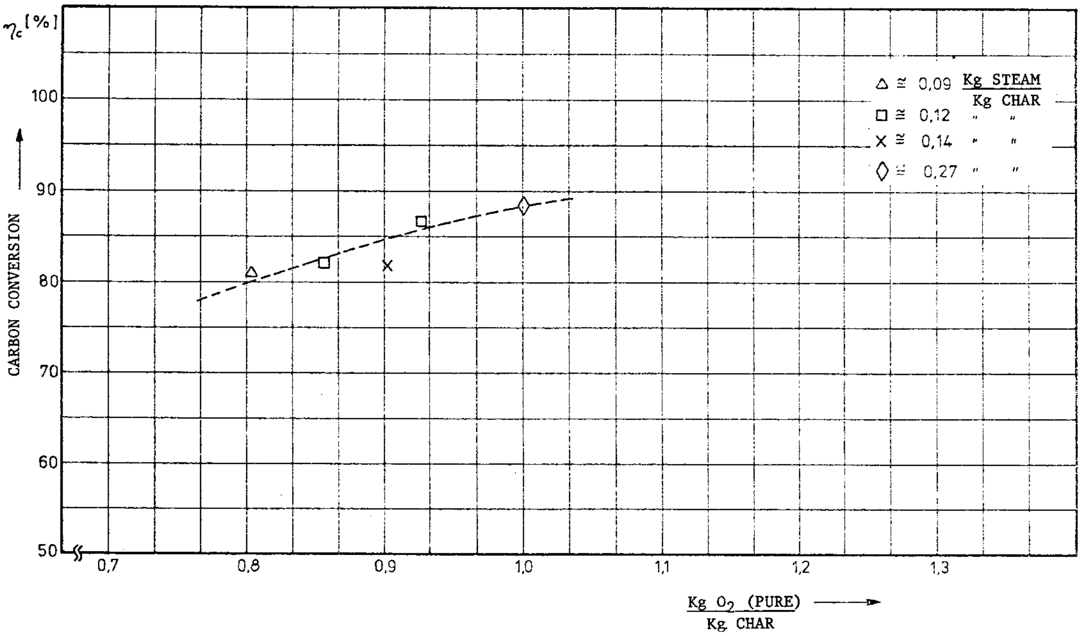

Figure 6-10. Carbon Conversion as a Function of O_2/Char Ratio, Pittsburgh Char

Section 7

PREDICTION OF LARGE-SCALE COMMERCIAL OPERATION THROUGH COMPUTER SIMULATION

The Puentes demonstration was essential to prove the practicality of gasifying COED char on a commercial scale. The choice of the Puentes plant, however, limited commercial testing to gasifiers with a capacity of 3 to 4 tons per hour.

The commercialization of COED-KT within the States could require plants with a capacity of 10,000 to 20,000 tons of coal throughput per day. Gasifiers even larger than the 12-ton-per-hour units now operating in South Africa would be required to handle the quantity of char produced from this much coal. At present, although no limit on size has been set, the largest proposed gasifier by Krupp-Koppers, a four-headed gasifier, could produce 55,000 Nm^3/h total gas (about 35,000 s.c.f.m. [wet]).

To predict the performance of this large gasifier, criteria from the Puentes demonstration were used as an input to the K-T coal gasification program (see the computer runs at the end of this section). The Puentes data was modified to reflect a modern operation.

Gas Analysis Vol. %	Puentes Gasifier (Test 27)	Simulated Large Commercial Gasifier
CO_2	10.59	12.39
CO	65.83	62.19
H_2	16.59	22.36
H_2S	1.13	1.49
COS	0.21	0.17
N_2	5.61	1.40
Gas/Char Thermal Efficiency %	61.20	62.66

CONCLUSION

As with most thermochemical processes, increase in size has many beneficial attributes. The computer simulations show that with a larger commercial gasifier for char gasification we can anticipate an improved gas quantity, a reduction in oxygen demand, and an improved char-to-gas efficiency. The larger scale reduces heat loss associated with unit throughput, and the additional heat available is utilized in the process to dissociate more steam. The additional steam dissociation releases both hydrogen and oxygen, thus increasing the gas yield, decreasing the oxygen requirements, and improving the coal-to-gas efficiency.

SUMMARY

The Puentes demonstration adequately demonstrated the commercial practicality of gasifying COED char. Indeed, after the minor modifications outlined in this report (to convert this older plant), the char handled more easily than any other material gasified, and exhibited first-class control features. The char gas was easily shifted and more productive than the plant's lignite gas. As a result, the Spanish plant used COED char with routine equipment for the production of ammonia.

The sulfur in the gas in the form of hydrogen sulfide or carbonyl sulfide can be readily removed by many commercial physical and chemical absorbent processes. Thus, as either a chemical synthesis gas or as a fuel gas (to replace natural gas), the clean gas is ideally suited for the demands of industry and can meet the stringent requirements for a clean environment.

Prior to commercialization, no further design problems would be anticipated with grinding, handling, gasifying, gas clean up, or sulfur removal. However, the individual slag characteristics of all coals should be checked for flow and wearing properties. For commercial use, every operable feature of the COED-KT process has been proven.

Some comparisons of the actual Puentes operation and the computer simulation are given as follows:

COMPARISON OF PUENTES AND LARGE-SCALE COMMERCIAL GASIFICATION

Western Kentucky Char

	Puentes Gasifier (Test 19)	Simulated Large Commercial Gasifier
Operating Temp.	2893°F	2730°F (assumed)
Carbon Conversion	93.71%	92% (assumed)
Oxygen Purity	93.35%	98% (assumed)
Char Feed (tons/h)	2.71	32.38 (computed)
Gas Make* (mscf/ton char)	45.112	47.840 (computed)
Oxygen (tons/ton char)	0.9854	0.90 (computed)
Steam (tons/ton char)	0.0930	0.32 (computed)
Gas Analysis Volume %		
CO_2	12.1	11.72 (computed)
CO	66.74	62.84 (computed)
H_2	13.76	22.74 (computed)
H_2S	0.91	1.11 (computed)
COS	0.12	0.126 (computed)
N_2	6.34	1.46 (computed)
Gas/Char Thermal Efficiency %	62.42	64.83 (computed)

Pittsburgh Char

	Puentes Gasifier (Test 27)	Simulated Large Commercial Gasifier
Operating Temp.	2736°F	2730°F (assumed)
Carbon Conversion	86.88%	90% (assumed)
Oxygen Purity	93.93%	98% (assumed)
Char Feed (tons/h)	3.47	32.29 (computed)
Gas Make* (mscf/ton char)	44.958	47.840 (computed)
Oxygen (tons/ton char)	0.9272	0.88 (computed)
Steam (tons/ton char)	0.1151	0.32 (computed)

The following computer-simulated data for large-scale operation are intended only as a guide and imply no guarantee.

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*CO + H_2 only

COMPUTER SIMULATION RUN NO. 1

Commercial Gasification of COED Char
One 4 Headed Gasifier

Western Kentucky Char

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      KOPPERS COMPANY, INC.           ENGINEERING AND CONSTRUCTION DIVISION         ** CONFIDENTIAL **        PAGE  1

                                         KOPPERS FUEL GASIFICATION PROCESS                                    RUN NO 01
                                   HP-615                                                                     WINTRELL
                   COMMERCIAL GASIFICATION OF COED CHAR-WESTERN KENTUCKY ONE 4-HEADED-GASIFIER               07/21/76
                                                     ************         1----------------1        1-------------------1
                                                     *          *7-------71 NET CALORIFIC 17-- 11 --71 GAS TO WASHER   1
                                                     *          *         1 VALUE   264.7 1   DRY    1 COOLERS   31554. 1
                                                     *          *         1----------------1 BASIS   1-------------------1
                                                     *          *                BTUS/SCF                      SCFH
                 1---------------1                   *  WASTE   *
                 1  BOILER FEED 17------- 12 -------7*  HEAT    *             1-------------------1
                 1     WATER     1                   *  BOILER  *7---- 13 ----71 SUPERHEATED STEAM 1
                 1     159.  GPM 1                   *          *             1       75625.  LB/HR1
                 1---------------1                   *          *             1-------------------1
                                                     *          *
                                                     *          *             1-------------------1
                 1---------------1                   *          *7-----------71 TO BLOWDOWN TANK  1
                 1   GAS SPRAY   1                   *          *             1      3781.   LB/HR1
                 1 COOLING WATER 17------- 9 ------7V *         *             1-------------------1
                 1    40.    GPM 1                 1 *          *
                 1---------------1                 1 *          *
                                                   1   *          *
                                                   1    *           *                      PROCESS EFFICIENCY = 81.79 PCT
                                                   1    * * 10        *                    GAS CALORIFIC VALUE=   12029.
                 1---------------1                 7---7*   *       *                             MILLIONS BTUS/DAY
                 1  BOILER FEED  1                      *SPRAY*    *
                 1     WATER    17--- 7 ---7V           *     * *
                 1    123.  GPM  1          1           *     *
                 1---------------1          1           *     *                          1-------------1
                                            1           *  6  *         7--------8 --------71 LOW PRESSURE 1
                                            V  *************************  A                1 STEAM= 58522.I LB/HR
                                            *  *                        *  *               1-------------1
       COAL=   777. NT/D              *     *                        *     *
            V        1A               *     *                        *     *
            1        FLUX=    0.   NT/D  *  *        GASIFIER        *     *          1-------------1
            1        ADDITIVES    17-----7* *                        *     *7---------------71 TO BLOWDOWN 1
            V            V         1     *  *                        *     *           1 TANK=  2926.I LB/HR
     1-------------1     V         1     *  *                        *     *           1-------------1
     1 DRIED  COAL 17------ 1 ---71         *************************
     1 PREPARATION 1    777. NT/D 1                *     *
     1-------------1               1                *     *
                                   1                *     *
                                   1                *     *
       1-------1     244.99 NT/D  1                *     *
       1 STEAM 17---------- 2 ---71                *******
       1-------1                   1                   V
                                   1                   1                         *
                                   1                   1                       * *
     1-------------1               1                   1                     *   *
     1   OXYGEN    1  699.76 NT/D  1                   1                   *     *
     1 FACILITIES 17------- 3 ---71                   1                 *       *             1-------------------1
     1   98.0 PCT  1                                   V               *         *7---- 4 -----71 SLAG=   44.9 NT/D 1
     1-------------1                        ************************             *             1-------------------1
                                            *                                  *
                       1---------------1    *                                *
                       1  WATER FROM  17- 5 -7*  SLAG QUENCH TANK          *           1------------------------------1
                       1 COOLING TOWER 1    *                            *7--- 14 ---71    WATER =       162.   GPM 1
                       1 =    163.  GPM 1   *                          *               1 TO CLARIFIER                 1
                       1---------------1    ************************                   1------------------------------1
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COMPUTER SIMULATION RUN NO. 1 (continued)

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KOPPERS COMPANY, INC.          ENGINEERING AND CONSTRUCTION DIVISION          ** CONFIDENTIAL **        PAGE  2
                                      KOPPERS FUEL GASIFICATION PROCESS                                 RUN NO 01
                              MP-615                                                                    WINTRELL
                 COMMERCIAL GASIFICATION OF COED CHAR-WESTERN KENTUCKY ONE 4 HEADED GASIFIER            07/21/76

                                      COAL GASIFICATION MATERIAL ANALYSIS
                                      ---- ------------ -------- --------

                                                   MATERIAL CHARGED
                                          FLUX     -------- -------
   FUEL AS RECEIVED    FUEL ASH                                 PERCENT      OXYGEN SUPPLY        STEAM SUPPLY
           PERCENT              PERCENT   IRON OXIDE           =  0.00
 CARBON     = 75.24  SILICA     = 46.06   SILICA               =  0.00       OXYGEN  (PCT)= 98.00 PRESSURE    =  15. PSIG
 HYDROGEN   =  1.78  ALUMINA    = 22.80   ALUMINA              =  0.00
 NITROGEN   =  1.58  LIME       =  2.71   LIME                 =  0.00       NITROGEN(PCT)=  2.00 TEMPERATURE= 250. DEG-F
 SULPHUR    =  3.07  MAGNESIA   =  0.00   MAGNESIA             =  0.00
 OXYGEN     =  2.87  IRON OXIDE = 28.45   CALCIUM CARBONATE    =  0.00       TEMPERATURE  = 220.F ENTHALPY  =1164. BTU/LB
 ASH        = 14.45                       MAGNESIUM CARBONATE  =  0.00
 MOISTURE   =  1.00                       SULPHUR              =  0.00
 CHLORINE   =  0.00                       MOISTURE             =  0.00
 COST=   0.00 DLS/NT                      COST =   0.00 DLS/NT               COST=  0.00 DLS/NT    COST=  0.00 DLS/NT

                                                   GASIFIER OPERATION
                        FUEL DATA                  -------- ---------       GASIFIER SHELL    STEAM DATA

                   PREHEAT              =   100. DEG-F                  FEED WATER TEMP.    =  230.0 DEG-F
                   MOISTURE             =     1. PERCENT                STEAM TEMPERATURE   =  274.0 DEG-F
                   CARBON GASIFIED      =  92.00 PERCENT                STEAM PRESSURE      =   29.7 PSIG
                   ASH TO GAS           =  60.00 PERCENT                ENTHALPY            = 1172.0 BTUS/LB
                                                                        SHELL HEAT LOSS     =  95.00 PERCENT
                                                                        STEAM CREDIT        =   0.00 DLS/NT
                     GASIFIER
                                                                           WASTE HEAT BOILER
                   FLAME TEMPERATURE    =  2730. DEG-F
                   GAS PRESSURE         =  15.30 PSIA                   FEED WATER TEMP.    =  230.0 DEG-F
                   HEAT LOSS            =    60. MMBTUS/HR              STEAM TEMPERATURE   =  900.0 DEG-F
                                                                        STEAM PRESSURE      =  910.0 PSIG
                     EQUILIBRIUM RATIO                                  ENTHALPY            = 1450.0 BTUS/LB
                      (PCO2)(PH2)   = 0.383                             BOILER EFFICIENCY   =  66.26 PERCENT
                      ----------- (PRACTICAL)                           GAS EXIT TEMPERATURE=   300. DEG-F
                      (PCO)(PH2O)                                       STEAM CREDIT        =   0.00 DLS/NT

                          SPRAY WATER                                      SLAG COOLING WATER

                   INLET TEMPERATURE =   110. DEG-F                     TEMPERATURE RISE    =    35. DEG-F
                   GAS PRESSURE      =  15.30 PSIA                      INLET TEMPERATURE   =    80. DEG-F
                   GAS TEMPERATURE   =  2011. DEG-F
                                                   MATERIAL PRODUCED
                                                   -------- --------
     SLAG ANALYSIS          DUST ANALYSIS        GAS (WET)  BOILER EXIT           GAS (DRY)  BOILER EXIT
           PERCENT                PERCENT            -            PERCENT                         PERCENT
 SILICA     = 46.05  SILICA     = 27.18   CARBON MONOXIDE      = 47.34       CARBON MONOXIDE    = 62.84
 ALUMINA    = 22.80  ALUMINA    = 13.46   CARBON DIOXIDE       =  8.82       CARBON DIOXIDE     = 11.72
 LIME       =  2.71  LIME       =  1.60   HYDROGEN             = 17.00       HYDROGEN           = 22.74
 MAGNESIA   =  0.00  MAGNESIA   =  0.00   HYDROGEN SULPHIDE    =  0.93       HYDROGEN SULPHIDE  =  1.11
 IRON OXIDE = 28.44  IRON OXIDE = 16.79   NITROGEN             =  1.10       NITROGEN           =  1.46
 BASICITY   = 0.039  CARBON     = 40.97   MOISTURE             = 24.81       CARBONYL SULPHIDE  = 0.126
                                                                             CALORIFIC VALUE    = 264.7 BTU/SCF
                                                                                                    NET DRY
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COMPUTER SIMULATION RUN NO. 1 (continued)

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KOPPERS COMPANY, INC.          ENGINEERING AND CONSTRUCTION DIVISION          ** CONFIDENTIAL **          PAGE  3

                                        KOPPERS FUEL GASIFICATION PROCESS                                 RUN NO 01
                              MR-615                                                                      WINTRELL
                    COMMERCIAL GASIFICATION OF COED CHAR-WESTERN KENTUCKY ONE - HEADED GASIFIER           07/21/76

                          -1- (**)      -1A-       -2-                    -3-                -4-     -5-              -6-
                          GASIFIER      FLUX     STEAM TO             OXYGEN TO              SLAG   QUENCH        GAS OUT OF
                            FEED                 GASIFIER              GASIFIER                     WATER          GASIFIER
                       LBS/HR   WT(*)  LBS/HR  LBS/HR  MOL/HR  LBS/HR  MOL/HR VOL(*)  LBS/HR  LBS/HR    LBS/HR  MOL/HR   VOL(*)
                       ------   -- -   ------  ------  ------  ------  ------ --- -  ------  ------    ------  ------   --- -
 CARBON MONOXIDE                                                                                       88097.   3145.   56.73
 CARBON DIOXIDE                                                                                        25778.    586.   10.56
 HYDROGEN               1152.    1.78                                                                   2277.   1129.   20.37
 NITROGEN               1023.    1.58                                  1024.    36.5    2.0             2046.     73.    1.32
 H2S + COS                                                                                              2112.     62.    1.12
 HYDROGEN CHLORIDE                                                                                         0.      0.    0.00
 H2O                     647.    1.00                                                    374.   81665.   9896.    549.    9.91
 ASH                    9354.   14.45                                                   3742.            5613.
 CARBON                48704.   75.25                                                                    3895.    324.
 SULPHUR                1987.    3.07
 CHLORINE                  0.   0.000
 OXYGEN                 1858.    2.87                                 57290.   1790.   98.0
 STEAM                                          20416.   1133.
 FLUX                                     0.
 TOTAL                 64726.  100.00     0.    20416.   1133.   58314.   1827.  100.0   3742.   81665. 130205.   5545.  100.00
                                                                                                        (***)   (***)
 TEMPERATURE(DEG-F)                               250.                                            80.            2730.
 PRESSURE(PSIA)                                  29.70                                                           15.30
 VOLUME(SCFM)                                                            11540.                                 35023. (NF)
 VOLUME(ACFM)                                    4837.                                                         206373.
 GPM                                                                                             163.
                          -7-       -8-       -9-             -10-                      -11-              -12-     -13-     -14-
                        BFW TO    STEAM     SPRAY                                                       BFW TO    STEAM   SLAG COOL
                       GASIFIER  GASIFIER  COOLING       GAS TO WASTE              GAS OUT WASTE        W.HEAT   W.HEAT    WATER
                       JACKETS   JACKETS    WATER        HEAT BOILERS              HEAT BOILERS         BOILER   BOILER    RETURN
                        LBS/HR    LBS/HR   LBS/HR   LBS/HR   MOL/HR  VOL(*)   LBS/HR   MOL/HR  VOL(*)   LBS/HR   LBS/HR    LBS/HR
                        ------    ------   ------   ------   ------  --- -    ------   ------  --- -    ------   ------    ------
 CARBON MONOXIDE                                    88097.   3145.   47.34    88097.   3145.   47.34
 CARBON DIOXIDE                                     25776.    586.    8.82    25776.    586.    8.82
 HYDROGEN                                            2277.   1129.   17.00     2277.   1129.   17.00
 NITROGEN                                            2046.     73.    1.10     2046.     73.    1.10
 H2S + COS                                           2112.     62.   0.933     2112.     62.   0.933
 HYDROGEN CHLORIDE
 H2O                    61448.             19793.   29688.   1648.   24.81    29688.   1648.   24.81    79406.             81291.
 ASH                                                 5613.                     5613.
 CARBON                                              3896.    324.             3896.    324.
 SULPHUR
 CHLORINE
 OXYGEN
 STEAM                            58522.                                                                         75625.
 FLUX
 TOTAL                  61448.    58522.   19793.  149997.   6643.  100.00   149997.   6643.  100.00   79406.   75625.    81291.
                                                    (***)    (***)            (***)    (***)
 TEMPERATURE,DEG-F        230.      274.     110.            2011.                       300.             230.     900.      115.
 PRESSURE(PSIA)                      44.                     15.30                      15.20                      925.
 VOLUME(SCFM)                     20518.                    41963. (NF)                41963. (NF)
 VOLUME(ACFM)                      9587.                   191533.                     59297.                     1102.
 GPM                      123.               40.                                                          159.               162.
  (*) PERCENTAGES,   (**) REF. CIRCLED NUMBERS ON SCHEMATIC DRAWING,PAGE 1,   (***) GAS ONLY,   (NF) NOT FEASIBLE,
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COMPUTER SIMULATION RUN NO. 1 (continued)

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KOPPERS COMPANY, INC.         ENGINEERING AND CONSTRUCTION DIVISION         ** CONFIDENTIAL **         PAGE  4

                              KOPPERS FUEL GASIFICATION PROCESS                                        RUN NO 01
                            MP-615                                                                     WINTRELL
              COMMERCIAL GASIFICATION OF COED CHAR-WESTERN KENTUCKY ONE 4 HEADED GASIFIER              07/21/76

                                                 HEAT BALANCE
           HEAT IN                               ---- -------                  HEAT OUT
           ---- --                                                             ---- ---

           GASIFIER                AS-RECEIVED FUEL                            GASIFIER                 AS RECEIVED FUEL
           --------                    BTU/NT                                  --------                     BTU/NT

   SENSIBLE HEAT IN COAL                 =    10654.         SENSIBLE HEAT IN FLUXED ASH               =    222747.
   SENSIBLE HEAT IN OXYGEN               =    57800.         SENSIBLE HEAT IN UNBURNT CARBON           =    126558.
   SENSIBLE + LATENT HEAT IN STEAM       =   705205.         SENSIBLE + LATENT HEAT IN EXIT GAS        =   4134481.
   CARBON OXIDATION TO CARBON MONOXIDE   =  5177548.         HEAT FOR DISSOCIATION OF LIMESTONE        =         0.
   CARBON OXIDATION TO CARBON DIOXIDE    =  3166030.         HEAT TO DISSOCIATE STEAM                  =   2355095.
   HYDROGEN TO HYDROGEN SULPHIDE         =    16616.         HEAT TO STEAM FROM GASIFIER JACKET        =   1761106.
   ELEMENT BREAKDOWN                     =-  441176.         HEAT LOSS THROUGH GASIFIER JACKET,ETC=        92690.

          TOTAL HEAT INTO GASIFIER       =  8692676.                 TOTAL HEAT OUT OF GASIFIER =      8692676.

           SPRAY CHAMBER                                             SPRAY CHAMBER
           ----- -------                                             ----- -------

   SENSIBLE + LATENT HEAT IN ENTRY GAS   =  4134481.         SENSIBLE + LATENT HEAT OF EXIT GAS        =   4238974.
   SENSIBLE HEAT IN FLUXED ASH           =   133648.         SENSIBLE HEAT IN FLUXED ASH               =     89821.
   SENSIBLE HEAT IN UNBURNT CARBON       =   126558.         SENSIBLE HEAT IN UNBURNT CARBON           =     86071.
   SENSIBLE HEAT IN SPRAY WATER          =    20180.

       TOTAL HEAT INTO SPRAY CHAMBER     =  4414867.             TOTAL HEAT OUT OF SPRAY CHAMBER =         4414867.

           WASTE HEAT BOILER                                         WASTE HEAT BOILER
           ----- ---- ------                                         ----- ---- ------

   SENSIBLE + LATENT HEAT OF ENTRY GAS   =  4238974.         SENSIBLE + LATENT HEAT OF EXIT GAS        =   1321725.
   SENSIBLE HEAT IN FLUXED ASH           =    89821.         SENSIBLE HEAT IN FLUXED ASH               =      7535.
   SENSIBLE HEAT IN UNBURNT CARBON       =    86071.         SENSIBLE HEAT IN UNBURNT CARBON           =      6268.
                                                             SENSIBLE + LATENT HEAT IN STEAM           =   2925371.
                                                             HEAT LOSS FROM WASTE HEAT BOILER          =    153967.

    TOTAL HEAT INTO WASTE HEAT BOILER    =  4414867.         TOTAL HEAT OUT OF WASTE HEAT BOILER =     4414867.

           SLAG QUENCH TANK                                          SLAG QUENCH TANK
           ---- ------ ----                                          ---- ------ ----

   SENSIBLE HEAT IN SLAG                 =    89099.         SENSIBLE HEAT IN SLAG                     =       788.
   SENSIBLE HEAT IN WATER                =     7570.         SENSIBLE HEAT IN WATER                    =     95441.
                                                             SENSIBLE + LATENT HEAT IN FLASH STEAM=              0.

    TOTAL HEAT INTO SLAG QUENCH TANKS    =    96668.         TOTAL HEAT OUT OF SLAG QUENCH TANKS =        96229.

           PROCESS HEAT INPUT                                  PROCESS HEAT OUTPUT (AVAILABLE)
           ------- ---- -----                                  ------- ---- ------ -----------

   HEAT POTENTIAL IN COAL (NET)          = 23887650.         TOTAL HEAT IN STEAM GENERATED             =   4686478.
   SENSIBLE + LATENT HEAT INPUTS         =   773658.         CALORIFIC VALUE OF EXIT GAS               =  15485227.

    TOTAL HEAT ENERGY SUPPLIED           = 24661308.         TOTAL HEAT ENERGY AVAILABLE               =  20171705.
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COMPUTER SIMULATION RUN NO. 1 (continued)

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KOPPERS COMPANY, INC.        ENGINEERING AND CONSTRUCTION DIVISION           ** CONFIDENTIAL **        PAGE  5

                                  KOPPERS FUEL GASIFICATION PROCESS                                    RUN NO 01
                              MR=615                                                                   WINTRELL
                  COMMERCIAL GASIFICATION OF COED CHAR-WESTERN KENTUCKY-ONE 4 HEADED GASIFIER          07/21/76


                                  SUPPLEMENTARY COAL GASIFICATION INFORMATION
                                  ------------- ---- ------------ -----------

      COAL GASIFICATION RAW MATERIAL COSTS                      ADDITIONAL COAL GASIFICATION DATA

                                       DOLLARS/DAY
     COST OF FUEL                       =        0.        SCF OF OXYGEN SUPPLIED PER LB OF CARBON =  13.9
     COST OF INJECTED STEAM             =        0.        SCF OF OXYGEN TOTAL PER LB OF CARBON    =  14.4
     COST OF OXYGEN                     =        0.        LB OF STEAM TO GASIFIER PER LB OF CARBON= 0.419
     COST OF FLUXES                     =        0.        DUST CONTENT OF EXIT BOILER GAS,GR/ACF  =  18.7
     COST OF BOILER FEED WATER          =        0.                   DULONGS FORMULA
     COST FOR RAW WATER                 =        0.        BTUS/NT OF COAL AS RECEIVED (HCV)       =23897627.
                                                           COAL/GAS EFFICIENCY                     = 64.825 PERCENT
     GASIFIER SHELL STEAM     (CREDIT)=          0.
     WASTE HEAT BOILER STEAM  (CREDIT)=          0.

     COST OF GAS                        =        0.
     COST PER MILLION BTUS              =0.0000000 DOLLARS
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COMPUTER SIMULATION RUN NO. 2

Commercial Gasification of COED Char One 4 Headed Gasifier

Pittsburgh Char

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     KOPPERS COMPANY, INC.             ENGINEERING AND CONSTRUCTION DIVISION          ** CONFIDENTIAL **         PAGE  1
                                         KOPPERS FUEL GASIFICATION PROCESS                                       RUN NO 02
                                    MR-615                                                                        WINTRELL
                    COMMERCIAL GASIFICATION OF COED CHAR - PITTSBURGH ONE 4 HE:DED GASIFIER                      07/21/76
                                                       *************         1---------------1          1------------------1
                                                       *           *7-------71 NET CALORIFIC 17-- 11 --71 GAS TO WASHER    1
                                                       *           *         1 VALUE   261.6 1   DRY    1 COOLERS   31386. 1
                                                       *           *         1---------------1  BASIS   1------------------1
                                                       *           *                BTUS/SCF                        SCFM
                 1---------------1                     *  WASTE    *
                 1   BOILER FEED 17-------- 12 -------7*  HEAT     *              1--------------------1
                 1     WATER     1                     * BOILER    *7---- 13 ----71 SUPERHEATED STEAM  1
                 1     161.  GPM 1                     *           *              1       76721.  LB/HR1
                 1---------------1                     *           *              1--------------------1
                                                       *           *
                                                       *           *              1--------------------1
                 1---------------1                     *           *7------------71 TO BLOWDOWN TANK   1
                 1   GAS SPRAY   1                     *           *              1       3836.   LB/HR1
                 1 COOLING WATER 17-------- 9 ------7V *           *              1--------------------1
                 1     40.   GPM 1                   1 *           *
                 1---------------1                   1 *           *
                                                     1   *           *
                                                     1     *           *                       PROCESS EFFICIENCY = 79.53 PCT
                                                     1     * * 10        *                     GAS CALORIFIC VALUE=     11824.
                 1---------------1                  7----7*    *       *                                    MILLIONS BTUS/DAY
                 1  BOILER FEED  1                         *SPRAY*   *
                 1     WATER     17---- 7 ---7V            *     * *
                 1     123.  GPM 1            1            *     *
                 1---------------1            1            *     *                                1--------------1
                                              1            *  6  *            7------- 8 --------71 LOW PRESSURE 1
                                              V  **************************   A                   1 STEAM= 58522.I LB/HR
                                                                                                  1--------------1
       COAL=   799. NT/D                   *     *                         *     *
            V        1A                    *     *                         *     *
            1       FLUX=    0.   NT/D     *     *       GASIFIER          *     *                 1-------------1
            1       ADDITIVES     17------7*     *                         *     *7---------------71 TO BLOWDOWN 1
            V           V         1        *     *                         *     *                 1 TANK=  2926.I LB/HR
     1-------------1    V         1                                                                1-------------1
     1 DRIED  COAL 17------ 1 ---71              **************************
     1 PREPARATION 1    799. NT/D 1                        *     *
     1-------------1              1                        *     *
                                  1                        *     *
                                  1                        *     *
       1-------1      251.92 NT/D 1                        *     *
       1 STEAM 17---------- 2 ---71                        *******
       1-------1                  1                           V
                                  1                           1                         *
                                  1                           1                       * *
     1-------------1              1                           1                     *   *
     1    OXYGEN   1  700.62 NT/D 1                           1                   *     *
     1 FACILITIES 17------- 3 ---71                           1                 *       *              1-------------------1
     1   98.0 PCT  1                                          V               *         *7---- 4 -----71 SLAG=    47.1 NT/D1
     1-------------1                               **************************           *              1-------------------1
                                                   *                                  *
                         1----------------1        *                                *
                         1   WATER FROM   17-- 5 -7*  SLAG QUENCH TANK            *          1-----------------------------1
                         1 COOLING TOWER  1        *                            *7--- 14 ---71    WATER  =      170.   GPM 1
                         1      171.  GPM 1        *                          *              1 TO CLARIFIER                1
                         1----------------1        **************************                1-----------------------------1
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COMPUTER SIMULATION RUN NO. 2 (continued)

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KOPPERS COMPANY, INC.         ENGINEERING AND CONSTRUCTION DIVISION          ** CONFIDENTIAL **        PAGE  2

                                   KOPPERS FUEL GASIFICATION PROCESS                                    RUN NO 02
                    MP=615                                                                              WINTRELL
             COMMERCIAL GASIFICATION OF COED CHAR - PITTSBURGH ONE 4 HEADED GASIFIER                    07/21/76

                                   COAL GASIFICATION MATERIAL ANALYSIS
                                   ---- ------------ -------- --------

                                               MATERIAL CHARGED
                                         FLUX  -------- -------
  FUEL AS RECEIVED    FUEL ASH                                  PERCENT      OXYGEN SUPPLY         STEAM SUPPLY
           PERCENT              PERCENT  IRON OXIDE           =  0.00
CARBON    = 74.45  SILICA     = 44.65    SILICA               =  0.00      OXYGEN  (PCT)= 98.00 PRESSURE    =  15. PSIG
HYDROGEN  =  1.88  ALUMINA    = 21.63    ALUMINA              =  0.00
NITROGEN  =  1.39  LIME       =  2.03    LIME                 =  0.00      NITROGEN(PCT)=  2.00 TEMPERATURE= 250. DEG-F
SULPHUR   =  3.96  MAGNESIA   =  0.00    MAGNESIA             =  0.00
OXYGEN    =  2.77  IRON OXIDE = 31.69    CALCIUM CARBONATE    =  0.00      TEMPERATURE  = 220.F ENTHALPY  =1164. BTU/LB
ASH       = 14.75                        MAGNESIUM CARBONATE  =  0.00
MOISTURE  =  1.00                        SULPHUR              =  0.00
CHLORINE  =  0.00                        MOISTURE             =  0.00
COST=   0.00 DLS/NT                      COST =    0.00 DLS/NT             COST=  0.00 DLS/NT       COST=  0.00 DLS/NT

                                               GASIFIER OPERATION
                   FUEL DATA                   -------- ---------       GASIFIER SHELL    STEAM DATA

                   PREHEAT           =    100. DEG-F                   FEED WATER TEMP.      =  230.0 DEG-F
                   MOISTURE          =      1. PERCENT                 STEAM TEMPERATURE     =  274.0 DEG-F
                   CARBON GASIFIED   =   90.00 PERCENT                 STEAM PRESSURE        =   29.7 PSIG
                   ASH TO GAS        =   60.00 PERCENT                 ENTHALPY              = 1172.0 BTUS/LB
                                                                       SHELL HEAT LOSS       =  95.00 PERCENT
                                                                       STEAM CREDIT          =   0.00 DLS/NT
                   GASIFIER
                                                                             WASTE HEAT BOILER
                   FLAME TEMPERATURE =  2730. DEG-F
                   GAS PRESSURE      =  15.30 PSIA                     FEED WATER TEMP.      =  230.0 DEG-F
                   HEAT LOSS         =     60. MMBTUS/HR               STEAM TEMPERATURE     =  900.0 DEG-F
                                                                       STEAM PRESSURE        =  910.0 PSIG
                   EQUILIBRIUM RATIO                                   ENTHALPY              = 1450.0 BTUS/LB
                   (PCO2)(PH2)    = 0.383                              BOILER EFFICIENCY     =  66.13 PERCENT
                   -----------(PRACTICAL)                              GAS EXIT TEMPERATURE=    300. DEG-F
                   (PCO)(PH2O)                                         STEAM CREDIT          =   0.00 DLS/NT

                        SPRAY WATER                                        SLAG COOLING WATER

                   INLET TEMPERATURE =   110. DEG-F                    TEMPERATURE RISE      =    35. DEG-F
                   GAS PRESSURE      =  15.30 PSIA                     INLET TEMPERATURE     =    80. DEG-F
                   GAS TEMPERATURE   =  2011. DEG-F
                                             MATERIAL PRODUCED
                                             -------- --------
     SLAG ANALYSIS        DUST ANALYSIS          GAS (WET)  BOILER EXIT             GAS (DRY)  BOILER EXIT
           PERCENT              PERCENT                      PERCENT                            PERCENT
SILICA     = 44.65   SILICA     = 24.25   CARBON MONOXIDE    = 46.54     CARBON MONOXIDE    = 62.19
ALUMINA    = 21.63   ALUMINA    = 11.75   CARBON DIOXIDE     =  9.25     CARBON DIOXIDE     = 12.39
LIME       =  2.03   LIME       =  1.10   HYDROGEN           = 16.56     HYDROGEN           = 22.36
MAGNESIA   =  0.00   MAGNESIA   =  0.00   HYDROGEN SULPHIDE  =  1.24     HYDROGEN SULPHIDE  =  1.49
IRON OXIDE = 31.69   IRON OXIDE = 17.21   NITROGEN           =  1.05     NITROGEN           =  1.40
BASICITY   = 0.031   CARBON     = 45.69   MOISTURE           = 25.37     CARBONYL SULPHIDE  = 0.170
                                                                         CALORIFIC VALUE    = 261.6 BTU/SCF
                                                                                              NET DRY
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COMPUTER SIMULATION RUN NO. 2 (continued)

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KOPPERS COMPANY, INC.            ENGINEERING AND CONSTRUCTION DIVISION          ** CONFIDENTIAL **      PAGE  3
                                      KOPPERS FUEL GASIFICATION PROCESS                                 RUN NO 02
                                  MR-61                                                                 WINTRELL
                COMMERCIAL GASIFICATION OF COED CHAR - PITTSBURGH ONE 4 HEADED GASIFIER                 07/21/76

                           -1- (**)       -1A-          -2-                   -3-                -4-      -5-               -6-
                          GASIFIER        FLUX       STEAM TO             OXYGEN TO             SLAG    QUENCH          GAS OUT OF
                            FEED                     GASIFIER             GASIFIER                      WATER            GASIFIER
                         LBS/HR    WT(*)   LBS/HR   LBS/HR   MOL/HR   LBS/HR   MOL/HR   VOL(*)   LBS/HR   LBS/HR   LBS/HR   MOL/HR   VOL(*)
                         ------    -----   ------   ------   ------   ------   ------   ------   ------   ------   ------   ------   ------
CARBON MONOXIDE                                                                                                    86792.    3099.    55.92
CARBON DIOXIDE                                                                                                     27102.     616.    11.11
HYDROGEN                  1119.     1.68                                                                            2223.    1103.    19.90
NITROGEN                   925.     1.39                               1025.     36.6      2.0                      1950.      70.     1.26
H2S + COS                                                                                                           2802.      82.     1.48
HYDROGEN CHLORIDE                                                                                                     0.       0.     0.00
H2O                        666.     1.00                                                    393.   85411.   10310.     572.    10.33
ASH                       9820.    14.75                                                   3928.            5892.
CARBON                   49567.    74.45                                                                    4957.     413.
SULPHUR                   2636.     3.96
CHLORINE                     0.    0.000
OXYGEN                    1844.     2.77                              57361.    1793.     98.0
STEAM                                             20994.    1165.
FLUX                                         0.
TOTAL                    66578.   100.00       0.   20994.    1165.   58385.    1829.    100.0    3928.   85411.  131180.    5541.   100.00
                                                                                                                    (***)    (***)
TEMPERATURE(DEG-F)                                  250.                                                  80.              2730.
PRESSURE(PSIA)                                     29.70                                                                   15.30
VOLUME(SCFM)                                                                   11554.                                     35001. (NF)
VOLUME(ACFM)                                       4974.                                                                 206242.
GPM                                                                                                      171.
                        -7-      -8-      -9-          -10-                       -11-              -12-     -13-     -14-
                        BFW TO   STEAM    SPRAY                                                     BFW TO   STEAM   SLAG COOL
                      GASIFIER GASIFIER COOLING      GAS TO WASTE              GAS OUT WASTE         W.HEAT   W.HEAT  WATER
                       JACKETS  JACKETS    WATER     HEAT BOILERS              HEAT BOILERS         BOILER   BOILER   RETURN
                         LBS/HR   LBS/HR   LBS/HR   LBS/HR   MOL/HR   VOL(*)   LBS/HR   MOL/HR   VOL(*)   LBS/HR   LBS/HR   LBS/HR
                         ------   ------   ------   ------   ------   ------   ------   ------   ------   ------   ------   ------
CARBON MONOXIDE                                     86792.    3099.    46.54   86792.    3099.    46.54
CARBON DIOXIDE                                      27102.     616.     9.25   27102.     616.     9.25
HYDROGEN                                             2223.    1103.    16.56    2223.    1103.    16.56
NITROGEN                                             1950.      70.     1.05    1950.      70.     1.05
H2S + COS                                            2802.      82.    1.235    2802.      82.    1.235
HYDROGEN CHLORIDE
H2O                      61448.            20120.   30430.    1689.    25.37   30430.    1689.    25.37   80557.            85018.
ASH                                                  5892.                      5892.
CARBON                                               4957.     413.             4957.     413.
SULPHUR
CHLORINE
OXYGEN
STEAM                             58522.                                                                     76721.
FLUX
TOTAL                    61448.   58522.   20120.  151300.    6658.   100.00  151300.    6658.   100.00   80557.   76721.   85018.
                                                     (***)    (***)             (***)    (***)
TEMPERATURE,DEG-F          230.     274.     110.             2011.                       300.              230.     900.     115.
PRESSURE(PSIA)                       44.                      15.30                      15.20                       925.
VOLUME(SCFM)                      20518.                     42056. (NF)                42056. (NF)
VOLUME(ACFM)                       9587.                    191955.                     59428.                      1118.
GPM                        123.               40.                                                       161.              170.
  (*) PERCENTAGES,   (**) REF. CIRCLED NUMBERS ON SCHEMATIC DRAWING,PAGE 1,   (***) GAS ONLY,   (NF) NOT FEASIBLE,
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COMPUTER SIMULATION RUN NO. 2 (continued)

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KOPPERS COMPANY, INC.          ENGINEERING AND CONSTRUCTION DIVISION          ** CONFIDENTIAL **          PAGE  4

                                   KOPPERS FUEL GASIFICATION PROCESS                                      RUN NO C2
                         MR-615                                                                           WINTRELL
                COMMERCIAL GASIFICATION OF COED CHAR - PITTSBURGH ONE 4 HEADED GASIFIER                   07/21/76

                                                  HEAT BALANCE
                HEAT IN                           ---- -------               HEAT OUT
                ---- --                                                      ---- ---

                GASIFIER             AS RECEIVED FUEL                        GASIFIER             AS RECEIVED FUEL
                --------                 BTU/NT                              --------                 BTU/NT

     SENSIBLE HEAT IN COAL                  =    10515.      SENSIBLE HEAT IN FLUXED ASH                =   226504.
     SENSIBLE HEAT IN OXYGEN                =    56266.      SENSIBLE HEAT IN UNBURNT CARBON            =   156536.
     SENSIBLE + LATENT HEAT IN STEAM        =   705069.      SENSIBLE + LATENT HEAT IN EXIT GAS         =  4050918.
     CARBON OXIDATION TO CARBON MONOXIDE =     4959457.      HEAT FOR DISSOCIATION OF LIMESTONE         =        0.
     CARBON OXIDATION TO CARBON DIOXIDE  =     3236606.      HEAT TO DISSOCIATE STEAM                   =  2327175.
     HYDROGEN TO HYDROGEN SULPHIDE          =    21433.      HEAT TO STEAM FROM GASIFIER JACKET         =  1712288.
     ELEMENT BREAKDOWN                      =-  425804.      HEAT LOSS THROUGH GASIFIER JACKET,ETC=       90120.

           TOTAL HEAT INTO GASIFIER         =  8563542.               TOTAL HEAT OUT OF GASIFIER =  8563542.

               SPRAY CHAMBER                                                SPRAY CHAMBER
               ----- -------                                                ----- -------

     SENSIBLE + LATENT HEAT IN ENTRY GAS =     4050918.      SENSIBLE + LATENT HEAT OF EXIT GAS         =  4165589.
     SENSIBLE HEAT IN FLUXED ASH            =   135902.      SENSIBLE HEAT IN FLUXED ASH                =    91254.
     SENSIBLE HEAT IN UNBURNT CARBON        =   156536.      SENSIBLE HEAT IN UNBURNT CARBON            =   106460.
     SENSIBLE HEAT IN SPRAY WATER           =    19945.

        TOTAL HEAT INTO SPRAY CHAMBER       =  4363302.           TOTAL HEAT OUT OF SPRAY CHAMBER =  4363302.

               WASTE HEAT BOILER                                          WASTE HEAT BOILER
               ----- ---- ------                                          ----- ---- ------

     SENSIBLE + LATENT HEAT OF ENTRY GAS =     4165589.      SENSIBLE + LATENT HEAT OF EXIT GAS         =  1310530.
     SENSIBLE HEAT IN FLUXED ASH            =    91254.      SENSIBLE HEAT IN FLUXED ASH                =     7647.
     SENSIBLE HEAT IN UNBURNT CARBON        =   106460.      SENSIBLE HEAT IN UNBURNT CARBON            =     7753.
                                                             SENSIBLE + LATENT HEAT IN STEAM            =  2885504.
                                                             HEAT LOSS FROM WASTE HEAT BOILER           =   151869.

       TOTAL HEAT INTO WASTE HEAT BOILER    =  4363302.        TOTAL HEAT OUT OF WASTE HEAT BOILER =  4363302.

               SLAG QUENCH TANK                                           SLAG QUENCH TANK
               ---- ------ ----                                           ---- ------ ----

     SENSIBLE HEAT IN SLAG                  =    90602.      SENSIBLE HEAT IN SLAG                      =      800.
     SENSIBLE HEAT IN WATER                 =     7697.      SENSIBLE HEAT IN WATER                     =    97051.
                                                             SENSIBLE + LATENT HEAT IN FLASH STEAM=           0.

       TOTAL HEAT INTO SLAG QUENCH TANKS    =    98299.        TOTAL HEAT OUT OF SLAG QUENCH TANKS =     97851.

               PROCESS HEAT INPUT                               PROCESS HEAT OUTPUT (AVAILABLE)
               ------- ---- -----                               ------- ---- ------ -----------

     HEAT POTENTIAL IN COAL (NET)           = 23619128.      TOTAL HEAT IN STEAM GENERATED              =  4597792.
     SENSIBLE + LATENT HEAT INPUTS          =   771850.      CALORIFIC VALUE OF EXIT GAS                = 14800032.

       TOTAL HEAT ENERGY SUPPLIED           = 24390978.        TOTAL HEAT ENERGY AVAILABLE              = 19397824.
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COMPUTER SIMULATION RUN NO. 2 (continued)

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KOPPERS COMPANY, INC.        ENGINEERING AND CONSTRUCTION DIVISION        ** CONFIDENTIAL **      PAGE  5

                                     KOPPERS FUEL GASIFICATION PROCESS                            RUN NO 02
                                MB=615                                                            WINTRELL
                      COMMERCIAL GASIFICATION OF COED CHAR - PITTSBURGH ONE 4 HEADED GASIFIER     07/21/76

                                     SUPPLEMENTARY COAL GASIFICATION INFORMATION
                                     ------------- ---- ------------ -----------

      COAL GASIFICATION RAW MATERIAL COSTS                 ADDITIONAL COAL GASIFICATION DATA

                                   DOLLARS/DAY
     COST OF FUEL                  =        0.      SCF OF OXYGEN SUPPLIED PER LB OF CARBON =  13.7
     COST OF INJECTED STEAM        =        0.      SCF OF OXYGEN TOTAL PER LB OF CARBON    =  14.1
     COST OF OXYGEN                =        0.      LB OF STEAM TO GASIFIER PER LB OF CARBON= 0.424
     COST OF FLUXES                =        0.      DUST CONTENT OF EXIT BOILER GAS,GR/ACF  =  21.3
     COST OF BOILER FEED WATER     =        0.                DULONGS FORMULA
     COST FOR RAW WATER            =        0.      BTUS/NT OF COAL AS RECEIVED (HCV)       =23631373.
                                                    COAL/GAS EFFICIENCY                     = 62.661 PERCENT
     GASIFIER SHELL STEAM    (CREDIT)=      0.
     WASTE HEAT BOILER STEAM (CREDIT)=      0.

     COST OF GAS                   =        0.
     COST PER MILLION BTUS         =0.0000000 DOLLARS
```

Section 8

CONVERSION FACTORS

To Convert	Into	Multiply by
Pressures:		
atu (std. atmosphere)	Pounds/sq in (psia)	14.696
bar	Pounds/sq in	14.50
kg/cm^2	Pounds/sq in	14.223
Lengths:		
m (meters)	Feet	3.2808
m	Inches	39.3708
cm	Inches	0.393708
mm	Inches	0.0393708
μ (microns)	Inches	0.0393708×10^{-3}
Weights:		
Ton (metric)	Tons (short)	1.1023
Ton (metric)	Pounds	2205.0
kg	Pounds	2.20462
g (grams)	Grains	15.432
mg	Grains	0.015432
Areas:		
cm^2	Sq inches	0.1550
m^2	Sq feet	10.7639
m^2	Sq inches	1550.15
Velocities:		
m/s	Ft/min	196.8503
m/s	Ft/sec	3.2808
m/s^2	Ft/sec^2	3.2808

To Convert	Into	Multiply by
Volumes and Flow Rates:		
l (liters)	Gallons	0.26417
l	Cu ft	3.53146×10^{-2}
l/min	Cu ft/sec	5.8858×10^{-4}
l/min	Gallons/sec	4.4028×10^{-3}
l/h	Cu ft/sec	9.8096×10^{-6}
l/h	Gallons/sec	7.3381×10^{-5}
m^3/s	Cu ft/sec	35.3145
m^3/s	Gallons/sec	264.1720
m^3/h	Cu ft/hr	35.3147
m^3/h	Gallons/hr	264.1720
m^3/h	Acre-feet/hr	0.8107×10^{-4}
Nm^3/h (0°C, 760 mm Hg, dry)	SCF/hr (60°F, 30 in Hg, dry)	37.2281
kg/h	Pounds/hr	2.20462
Weights of Volumes:		
kg/m^3	Pounds/cu ft	6.24278×10^{-2}
kg/Nm^3 (0°C, 760 mm Hg, dry)	Pounds/SCF (60°F, 30 in Hg, dry)	5.92187×10^{-2}
mg/l	Pounds/cu ft	6.2447×10^{-6}
mg/l	Parts/million	1.00
mg/Nm^3 (dry)	Pounds/SCF (dry)	5.920×10^{-8}
mg/Nm^3 (dry)	Grains/SCF (dry)	4.14539×10^{-4}
Heat Quantities:		
Kcal	Btu	3.96832
[one cal = 4.1868 joules (J)]		
Kcal	KJ	4.1868
KJ	Btu	0.947817
Kcal/h	Btu/hr	3.9683
Kcal/h	KJ/s	1.1629×10^{-3}
KJ/s	Btu/hr	3.4121×10^{-3}
KJ/kg	Btu/pound	0.42992
KJ/Nm^3 (dry)	Btu/SCF (dry)	2.5459×10^{-2}