MWK-MPR-18

RESEARCH AND DEVELOPMENT DEPARTMENT — KELLOGG

DEVELOPMENT OF KELLOGG COAL GASIFICATION PROCESS

Contract No. 14-01-0001-380

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Progress Report No. 18

APPROVED:

Project Manager

Director
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CONTENTS

I. INTRODUCTION

The objective of this contract with the Office of Coal Research is to develop the Kellogg Gasification Process to the point where it will be able, on a commercial scale, to convert coal into pipeline gas at a cost of 50¢/MSCF or hydrogen at 25¢/MSCF. Five raw materials are to be studied -- an anthracite, a high-volatile bituminous coal, a sub-bituminous coal, a lignite, and a char. Although Kellogg's experimental work will not extend beyond the production of raw synthesis gas, the overall project must make engineering evaluations for four ultimate end products -- pipeline gas, hydrogen, synthesis gas, and transport gas.

Basis for the Kellogg Gasification Process is the reaction between steam and fine coal in a molten salt bath to form synthesis gas, a mixture of hydrogen and carbon monoxide, according to the reaction:

$$\underset{\text{(coal)}}{C} + \underset{\text{(steam)}}{H_2O} + \text{heat} \longrightarrow \underbrace{H_2 + CO}_{\text{(synthesis gas)}}$$

The necessary heat of reaction is supplied by circulating a heated molten salt stream. In addition, the molten salt mixture is chosen to catalyze the gasification reaction so that it may be carried out at a relatively low temperature.

The program is divided into three phases of study extending over a five-year period. Phase I, which is now in progress, involves several concurrent efforts:

1. Bench-scale process research -- to investigate melt properties, reaction kinetics, and the effect of process variables.

2. Chemical engineering studies and development -- to determine the optimum process flowsheet and operating conditions and to coordinate experimental work with overall project objectives.

3. Mechanical development -- to find acceptable materials of construction and develop techniques for handling the molten salt and powdered coal.

Phase I will be concluded by the design of a pilot plant to gasify 24 tons of coal per day, if it is found that a pilot plant program is justified by the bench-scale experimentation and economic studies.

Phase II will be devoted largely to the construction and operation of a pilot plant to convert a variety of raw materials into raw synthesis gas. The effect of operating variables found to be significant in Phase I will be investigated to obtain data for design of a commercial plant.

Phase III will involve the detailed process design of a commercial plant to produce 250 million standard cubic feet a day of product gas, including cost estimates and projected economics for those areas of the country that appear to offer commercial possibilities.

Bench-scale process research continued during the month with experiments designed to investigate the effect of operating pressure on gasification rates and to study the factors affecting melt carryover. These results will be presented in the next monthly progress report.

11. SUMMARY

This progress report is the eighteenth since the contract was awarded. It is devoted exclusively to the engineering development program of the first phase of the contract.

A new conceptual process flowsheet limited to the gasification, combustion and ash removal sections of a plant capable of producing 250 million standard cubic feet per day of pipeline gas from bituminous coal by the Kellogg Gasification Process has been prepared. The investment for these sections has been estimated and a gas cost has been calculated for these sections and compared with that originally estimated by Kellogg in its proposal to OCR of November 1, 1962. From this study, a gas selling price in the range of 50¢/MSCF is expected to be a reasonable and achievable goal. This is based on the current estimate of about 35¢/MSCF for costs in the gasification, combustion and ash removal portions of the overall plant.

In arriving at these cost estimates, new conceptual mechanical designs have been prepared for all critical process areas in contact with melt (a mixture of molten sodium carbonate, coal and coal ash) or melt carry-over. These areas include the primary reactor vessel, overhead separators, melt quench tower, and transfer lines.

These designs are based on the results of an extensive corrosion testing program covering a wide range of refractories and alloys. The corrosion test medium was selected so as to reproduce closely the conditions anticipated in commercial operation. Thus, a system was set up containing molten sodium carbonate, coal ash and a graphite which underwent gasification or combustion at temperatures up to 1830°F. Graphite was used rather than coal to simplify the corrosion tests. Of the various materials tested, Monofrax A (a high-purity, high-density alumina melted at 3900°F and cast in relatively simple shapes) proved to be a suitable material. In total exposure of some 1200 hours in 4 tests, Monofrax A showed average corrosion rates of 0.14 to 0.18 inches per year under gasification conditions and 0.05 inches per year during a combustion test. Consequently Monofrax A was selected as the material of construction for the primary vessels and a conservative figure of 0.2 inches per year was used for design corrosion rate.

For the conceptual design Monofrax thickness and temperature gradient are chosen so that any melt tending to leak through or between the lining blocks will freeze within the confines of the corrosion-resistant material. With a corrosion allowance of 0.2 inches per year, a minimum ten-year lining life is anticipated with the design that has been adopted.

Nine gasification-combustion reactors, each 23 feet in inside diameter and 40 feet high, are needed to produce 250 MM SCFD of pipeline gas. Total cost of the Monofrax linings is estimated -- very approximately -- to be 6 million dollars. Cost of replacing the linings every ten years would be about 1c/MSCF of pipeline gas.

In addition to the conceptual designs, a preliminary study was made of the important control functions needed in the hot molten salt system. It was found that a feasible control system can be set up though much additional work is needed to complete its design.

Based on these designs, the corrosion data, and the estimated gas cost, it is concluded that the plant is both feasible and practical to build utilizing current technology, materials, and facilities. It should be emphasized that the designs are conceptual only; final designs will depend on the results of pilot-plant experience.

III. FEASIBILITY EVALUATION OF GASIFICATION AND ASH REMOVAL

The feasibility of designing and building a plant to produce 250 million SCFD of pipeline gas from coal by the Kellogg Gasification Process is best evaluated by investigating a full-scale design. Therefore, a process flowsheet for the gasification and ash removal sections of such a plant has been prepared, its basic control system outlined, and its investment estimated. From these figures a gas cost has been calculated. In conjunction with this process design, conceptual mechanical designs have been prepared for all critical areas in contact with carbonate melt or melt carry-over, including the primary reactor vessel, overhead separators, melt quench tower, and transfer lines. Based on these designs and the economic analysis, the plant appears both feasible and practical to build utilizing current technology, materials, and facilities.

It should be emphasized that the designs are conceptual and that final designs will be dependent on results of pilot-plant experience.

A. Suitability of Materials

Corrosion tests show Monofrax A to be the most consistently acceptable performer in contact with the liquid melt, melt interface, and vapor space at operating temperatures above 1800°F. Corrosion data on Monofrax A indicate an expected corrosion rate of less than 0.2 inches per year, which should be acceptable for the commercial application. The results of all the corrosion tests carried out with Monofrax A are given in Table 1. It should be noted that the corrosion rates for relatively long exposures to the gasification conditions are all below 0.2 inches per year and that the results are consistent in all the tests in the program. In addition, the corrosion rate under combustion conditions is substantially below that for gasification.

Several other materials, namely Zircofrax O, Chromex B, Ritex CB and Zirconia have performed satisfactorily at less severe conditions than currently contemplated for the commercial unit. None of these materials were found to be suitable for service at the 1800°F level in the molten salt environment, although the performance of Zirconia Y-1027 may have been adversely affected by the presence of metal alloys and other refractories that were concurrently tested in order to expeditiously screen a wide range of materials. Since Monofrax A has been unaffected by the presence of other materials and since it has an acceptable corrosion rate, it is clearly the material to be used.

TABLE I

MONOFRAX "A" TEST RESULTS

Operating Temperature: ~1830°F
Ash Content of Melt: ~10%

Test #	Duration	Atmosphere	Results	Corrosion Rate
3	242 Hrs.	Gasification	Satisfactory	0.14 in/yr Avg.
4	500 Hrs.	Gasification	Satisfactory	0.177 in/yr Avg.
7	395 Hrs.	Gasification	Satisfactory	0.135 in/yr Avg.
Oxid. #2	135 Hrs.	Combustion	Satisfactory	0.046 in/yr

Notes:
1. Corrosion rates are determined from change in dimensions.
2. In Tests # 3, 4, and 7 sample coupons were supported in the melt, in the gas space above the melt, and at the gas-melt interface. In oxidation Test #2, one sample was tested in the melt.
3. Tests 1, 2, 5, and 6 in this series were carried out to test materials other than Monofrax.

Monofrax A is manufactured by the Harbison-Carborundum Corporation. It is a high-purity alumina melted in an electric furnace at 3900°F and cast in graphite molds in relatively simple shapes suitable for block wall construction.

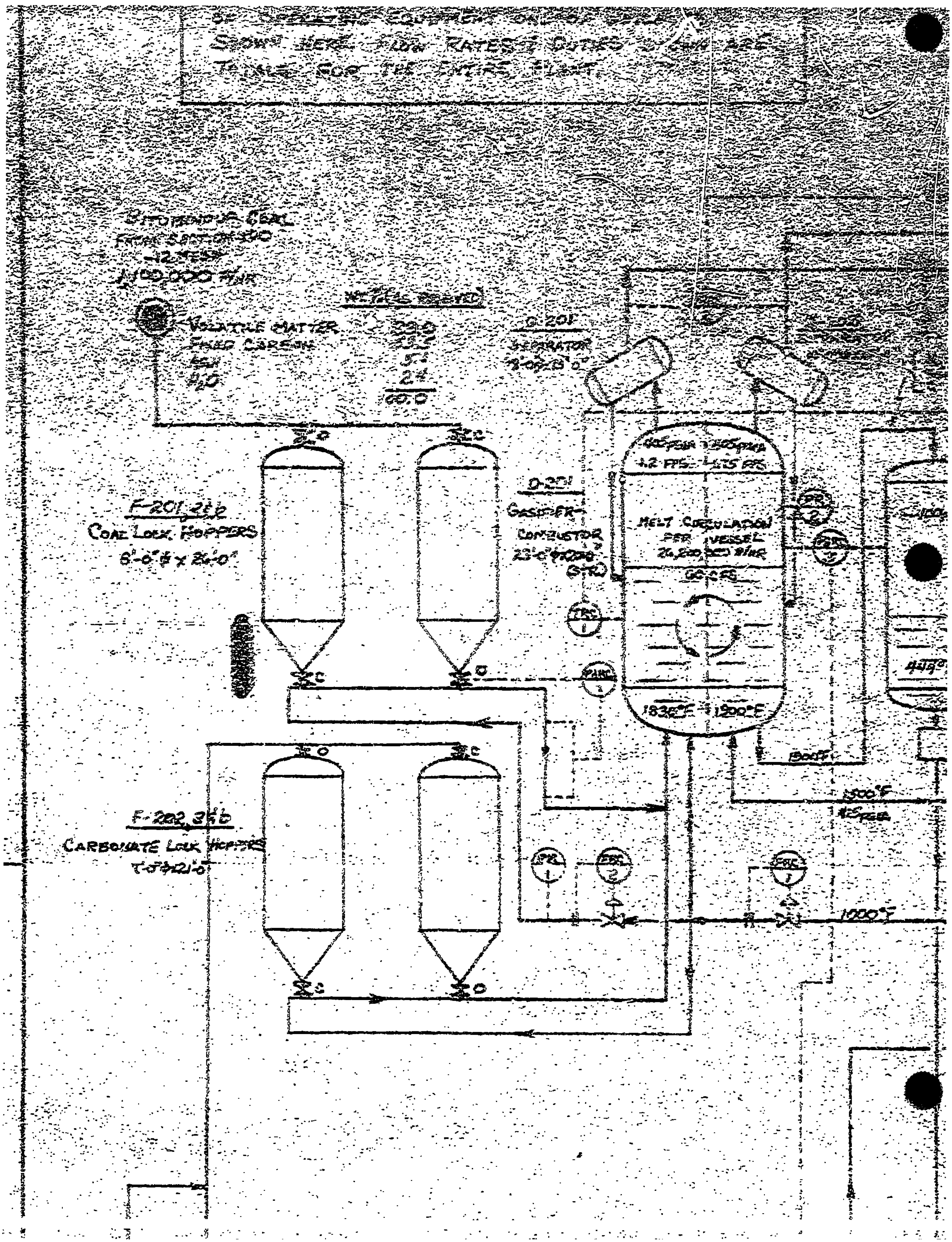
Monofrax A is now used in several commercial applications. Blocks of 10" - 12" thickness are used to line glass furnaces, where they last for four to five years. Several Texaco partial combustion generators lined with Monofrax A were operated commercially, but are now out of service for other reasons. A panel of Monofrax A is currently being tested in an operating blast furnace and appears to be performing satisfactorily. Blocks as large as three tons have been cast. Neglecting a casting void, which can be minimized or excluded by several devices, the product has very low porosity.

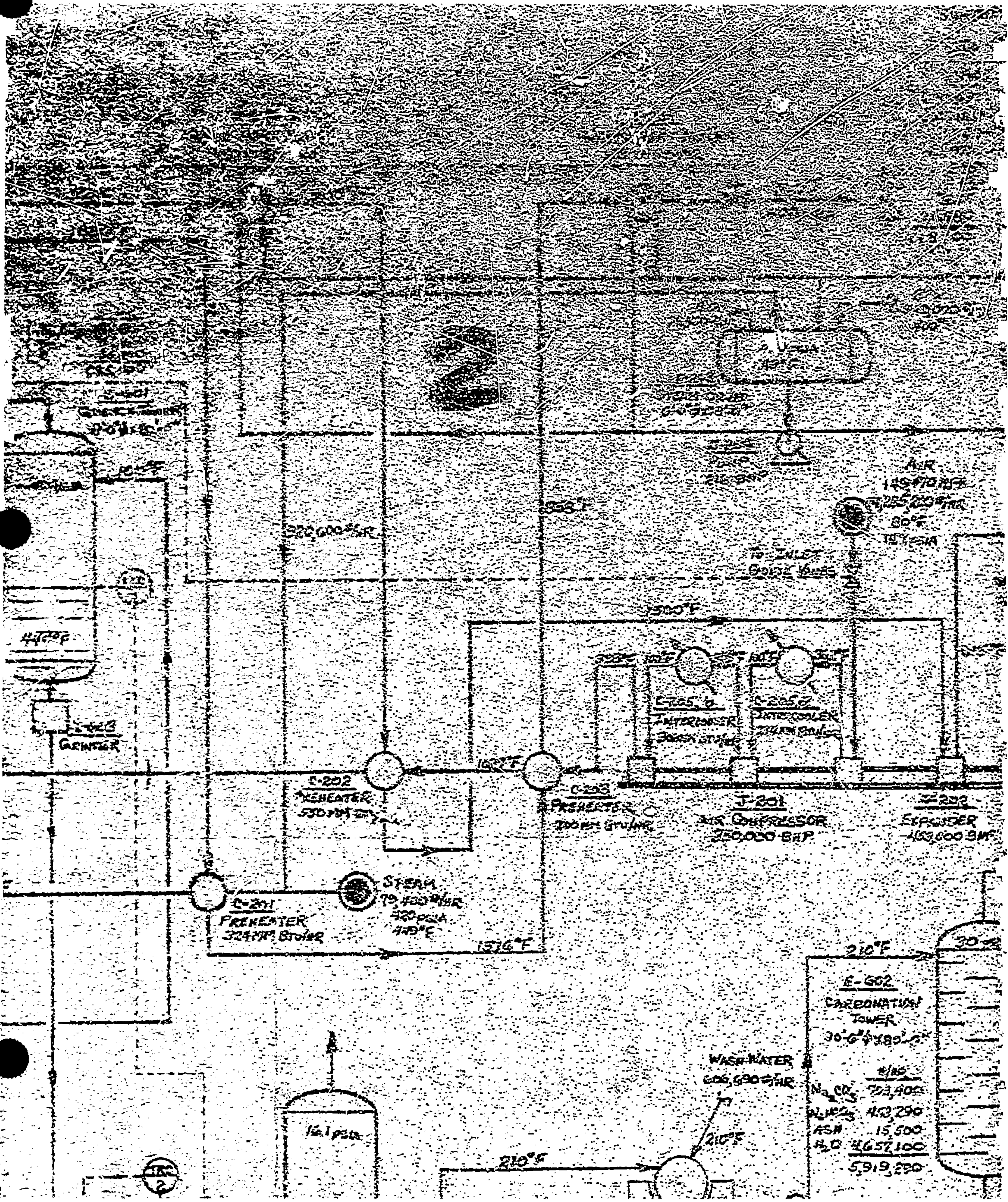
The price of Monofrax is in the order of $1,000 to $1,300 per average ton depending on whether regular cast blocks or DCL (Diamond Tread Blocks) are used. Initially, DCL blocks are recommended until commercial performance or pilot plant performance can be evaluated. Linings, in general, should be installed in a self-keying pattern without the use of binder. The temperature gradient through the Monofrax should be such as to assure freezing of the melt within the confines of the Monofrax blocks.

It should be noted here that although this present discussion has been concerned only with non-metallic materials of construction, an earlier test program indicated that no metallic materials currently available are capable of withstanding the conditions of the gasification system. Thus, process designs incorporating ideas such as heating the melt through a heat transfer surface to supply the heat of reaction have been discarded. To this degree, therefore, the proposed process design has been greatly influenced by the availability of acceptable materials of construction.

B. Process Description

A process flow diagram for the gasification and ash removal sections of a plant capable of producing 250,000,000 standard cubic feet per day (SCFD) of pipeline gas from bituminous coal is presented as Figure 1. Because of the large volume of gas produced, these sections are subdivided into nine parallel operating units. All flow rates and duties shown on this flowsheet and mentioned in this description are


SHOWN HERE. FLOW RATES & DUTIES SHOWN ARE
TOTALS FOR THE ENTIRE PLANT
BITUMINOUS COAL
VOLATILE MATTER
FIXED CARBON
ASH
H2O
D-201
SEPARATOR
D-201
GASIFIER-COMBUSTOR
MELT CIRCULATION
PER VESSEL
60 CFS
F-201, 2 & b
COAL LOCK HOPPERS
6'-0" Ø x 26'-0"
F-202, 3 & b
CARBONATE LOCK HOPPERS
1830°F
1900°F
1800°F
1500°F
1000°F



E-202
REHEATER
1632°F
E-203
PREHEATER
J-201
AIR COMPRESSOR
250,000 BHP
STEAM
79,400 #/HR
420 psia
449°F
E-201
PREHEATER
324 MM Btu/HR
1576°F
GRINDER
AIR
80°F
300°F
210°F
E-602
CARBONATION TOWER
30'-6" x 180'-0"
WASH WATER
606,690 #/HR
#/HR
Na_2CO_3 723,400
$NaHCO_3$ 453,290
ASH 15,500
H_2O 4,657,100
5,919,290
16.1 psia
210°F

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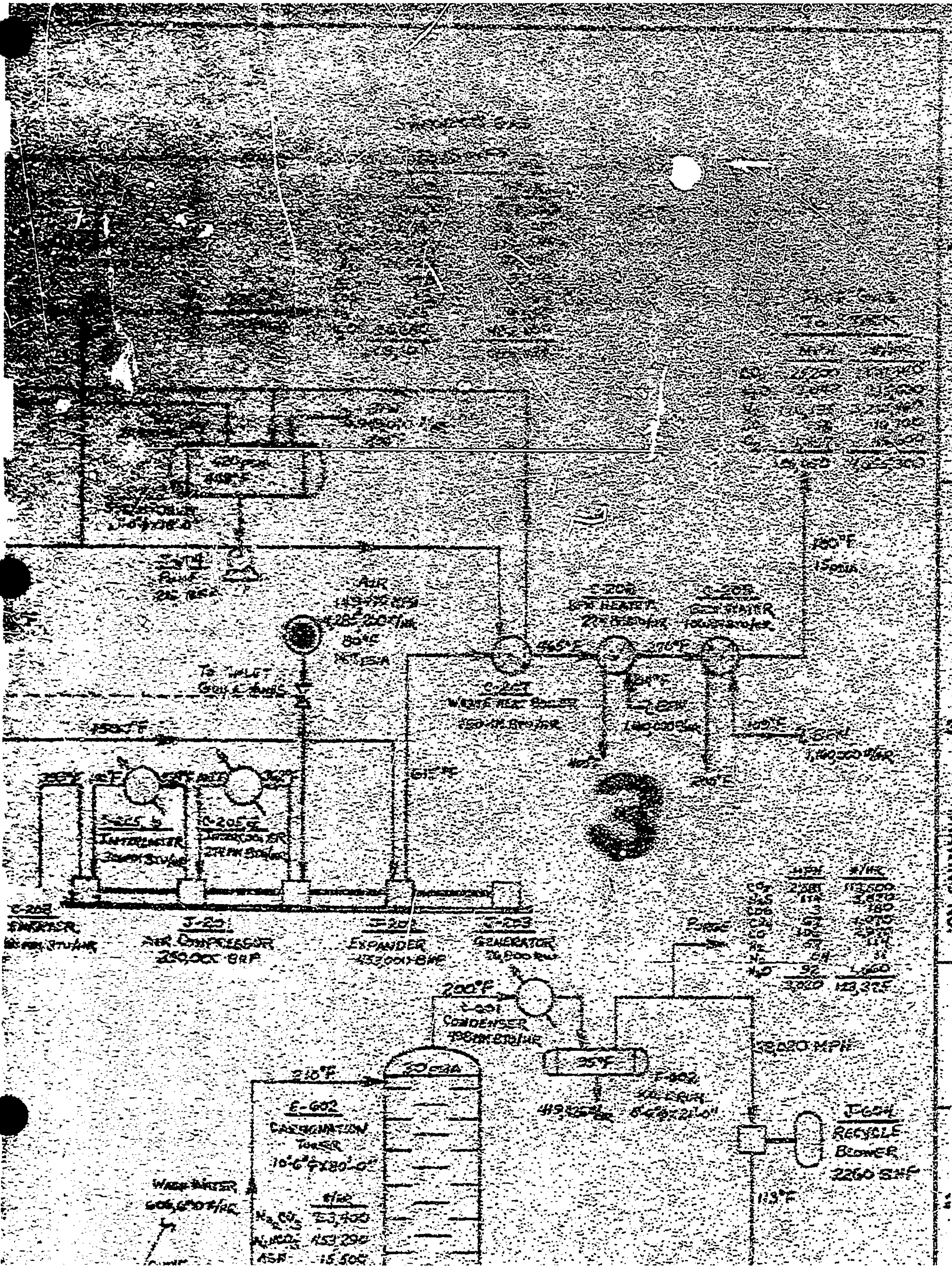

AIR
80°F
C-201
WASTE HEAT BOILER
BFW HEATER
150°F
J-201
AIR COMPRESSOR
250,000 BHP
EXPANDER
GENERATOR
PURGE
200°F
C-601
CONDENSER
210°F
E-602
CARBONATION TOWER
35°F
J-604
RECYCLE BLOWER
2260 BHP
133°F
WASH WATER
3

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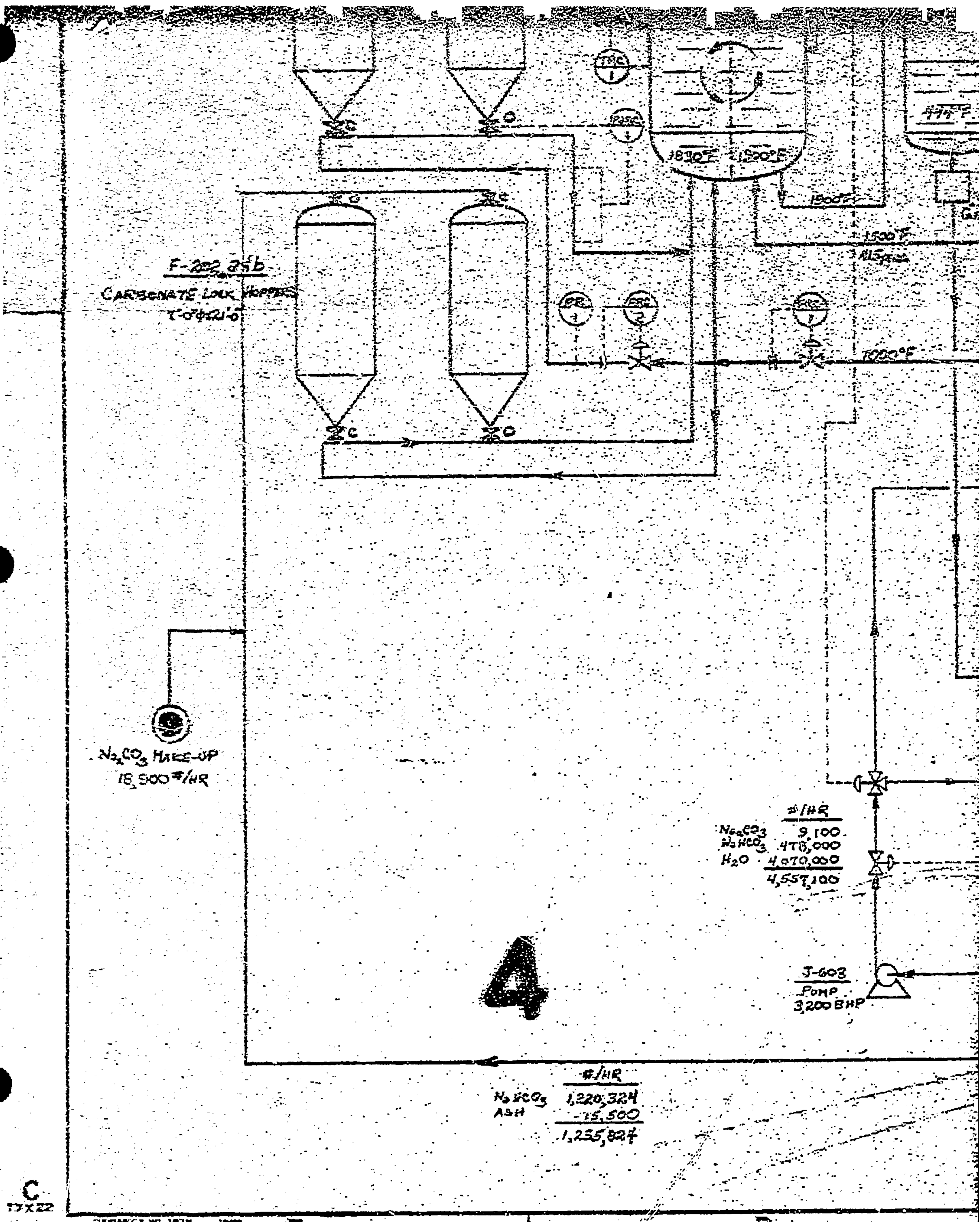
F-202 a&b
CARBONATE LOCK HOPPERS
7'-0"φ x 21'-0"
1830°F
1500°F
1500°F
1000°F
Na2CO3 MAKE-UP
18,500 #/HR
#/HR
Na2CO3 9,100
NaHCO3 478,000
H2O 4,070,000
4,557,100
J-603
PUMP
3,200 BHP
4
#/HR
Na2CO3 1,220,324
ASH 15,500
1,235,824
C
17 X 22

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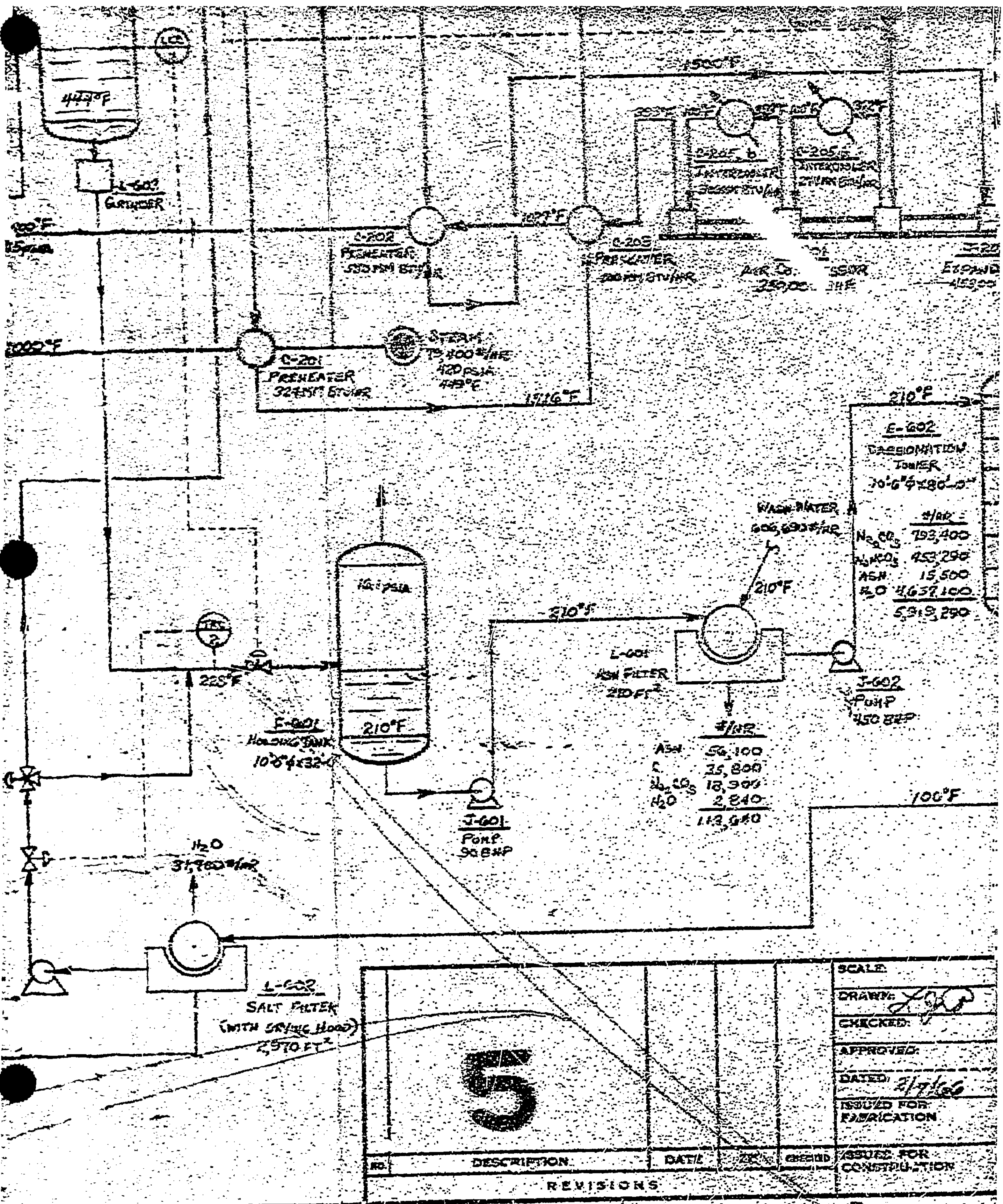

449°F
LCC
L-603
GRINDER
1500°F
C-205 b
INTERCOOLER
C-205 a
INTERCOOLER
C-202
PREHEATER
C-203
PREHEATER
AIR COMPRESSOR
EXPAND
C-201
PREHEATER
STEAM
420 PSIA
449°F
1916°F
210°F
E-602
CARBONATION TOWER
WASH WATER
#/HR
N_2CO_3 193,400
ASH 15,500
H_2O 4,657,100
5,919,290
210°F
L-601
ASH FILTER
210 FT²
J-602
PUMP
450 BHP
TRC
225°F
E-601
HOLDING TANK
210°F
#/HR
ASH 56,100
C 35,800
H_2O 2,840
113,940
J-601
PUMP
50 BHP
100°F
H_2O
L-602
SALT FILTER
5
SCALE
DRAWN
CHECKED
APPROVED
DATED
ISSUED FOR FABRICATION
ISSUED FOR CONSTRUCTION
NO.
DESCRIPTION
DATE
CHECKED
REVISIONS

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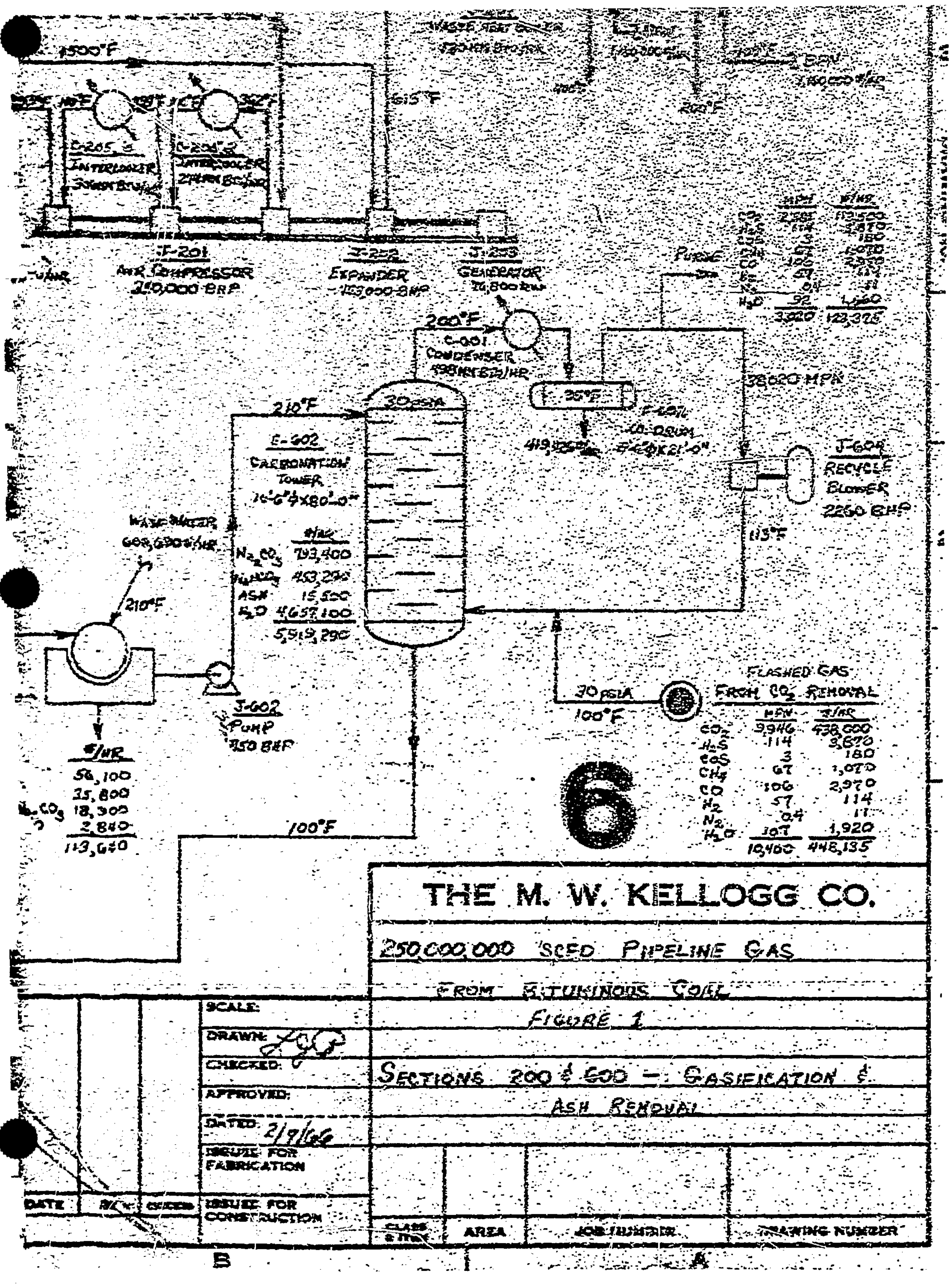

1500°F
615°F
WASTE HEAT BOILER
200°F
E-205
INTERCOOLER
E-204
INTERCOOLER
J-201
AIR COMPRESSOR
250,000 BHP
J-202
EXPANDER
153,000 BHP
J-203
GENERATOR
PURGE
200°F
C-601
CONDENSER
210°F
30 PSIA
35°F
F-601
KO DRUM
38,620 MPH
J-604
RECYCLE
BLOWER
2260 BHP
115°F
E-602
CARBONATION
TOWER
WASH WATER
210°F
J-602
PUMP
350 BHP
#/HR
56,100
35,800
18,300
2,840
113,040
100°F
30 PSIA
100°F
FLASHED GAS
FROM CO₂ REMOVAL
MPH
#/HR
CO₂ 9,946 438,000
H₂S 114 3,870
COS 3 180
CH₄ 67 1,070
CO 106 2,970
H₂ 57 114
N₂ 0.4 11
H₂O 107 1,920
10,400 448,135
6
THE M. W. KELLOGG CO.
250,000,000 SCFD PIPELINE GAS
FROM BITUMINOUS COAL
FIGURE 1
SECTIONS 200 & 600 — GASIFICATION & ASH REMOVAL
SCALE:
DRAWN:
CHECKED:
APPROVED:
DATED: 2/9/66
ISSUE FOR FABRICATION
ISSUE FOR CONSTRUCTION
DATE
BY
CHKD
CLASS & ITEM
AREA
JOB NUMBER
DRAWING NUMBER
B
A

total quantities for the nine units.

Bituminous coal from storage, ground to approximately -12 mesh in a preceeding section of the plant (Section 100), is fed to a set of lock hoppers, F-201,a&b, whose purpose is to receive the coal at atmospheric pressure and to deliver it to the process at the operating pressure of about 415 psia. Each of the lock hoppers operates on a 30-minute cycle comprising the following steps:

(1) Filling with coal at atmospheric pressure
(2) Pressurizing to about 450 psia with synthesis gas or product gas
(3) Discharging the coal into the steam line
(4) De-pressurizing

The lock hopper system is designed so that while one hopper is being filled and pressurized, the other is discharging coal into the process stream. This provides a continuous flow of coal to the gasifier.

Coal from F-201,a&b flows by gravity at the rate of 1,100,000 pounds per hour into a steam line where it is mixed with 420 psia steam superheated to 1 000°F in preheater C-201. The amount of steam used as a carrier gas is such that the coal will not be heated above about 500°F, thus preventing the coal from becoming plastic and sticky in the lines. Another stream of this 1000°F steam is used to pick up a recycle stream of $NaHCO_3$-Na_2CO_3 and make-up Na_2CO_3 from lock hoppers F-202, a&b. The two solids-containing steam streams then flow to the gasification section of gasifier-combustor D-201 together with the remainder of the 1000°F steam. The total steam fed to the gasifier is 1,000,000 pounds per hour. Of this, 920,600 pounds per hour is generated in waste heat boilers C-204, C-206, C-207 and 79,400 pounds per hour is generated in waste heat boilers in other sections of the plant.

In the gasifier, the steam and coal are heated to 1830°F by intimate contact with melt (a mixture of molten sodium carbonate, coal ash, and coal) and react according to the following reactions to produce synthesis gas.

$$C + H_2O \longrightarrow CO + H_2 \quad (1)$$
$$CO + H_2O \rightleftharpoons CO_2 + H_2 \quad (2)$$
$$C + 2H_2 \longrightarrow CH_4 \quad (3)$$
$$CO + 3H_2 \rightleftharpoons CH_4 + H_2O \quad (4)$$
$$C_mH_n \longrightarrow CH_4 + H_2 + C \quad (5)$$

The raw synthesis gas leaves the melt at about 1830°F and 405 psia and flows through separator G-201 designed to remove entrained coal particles and melt droplets. The gas is then cooled to 700°F in exchangers C-201, C-203 and C-204 and leaves the gasification section at the rate of 118,105 moles per hour.

The primary gasification reaction (reaction 1) is highly endothermic; the required heat of reaction is supplied as sensible heat of circulated melt. In the version shown in Figure 1, the gasifier-combustor is divided into two sections by a vertical wall which is perforated below the liquid level. By proper choice of gas velocity, a difference in degree of aeration of the melt in the two sections is induced. This causes melt to circulate from the synthesis gas section, where it provides the reaction heat, to the heating or combustion section. In this latter section heat is added to the melt by direct contact with hot combustion gases.

The flue gases which supply heat to the combustion section of the gasifier-combustor are generated by combustion of coal with air. Air for this purpose is compressed at the rate of 4,285,200 pounds per hour to 420 psia in compressor J-201 and is preheated to 1500°F in exchangers C-202 and C-203. Combustion occurs in direct contact with the melt, transferring heat to the reacting system efficiently.

The hot flue gases leave the bed at about 2200°F and 405 psia and pass through separator G-202 to remove entrained coal and melt. The gas then flows through exchangers C-206 and C-202 where it is cooled to 1500°F. This stream is expanded to substantially atmosphere pressure in J-202, which provides all the power required for air compression plus an additional 76,800 kw of electricity in generator J-203. Expanded flue gas at 615°F and about 18 psia is cooled to 180°F in exchangers C-207, C-208, and C-209 and is vented to the atmosphere at the rate of 4,655,300 pounds per hour.

The ash left in the melt by the combustion and gasification of the coal is allowed to build up to a level of 8 weight percent. A slipstream of the ash - carbon - Na_2CO_3 mixture is continuously withdrawn from D-201 and flows to E-601, where it is quenched to 444°F with a portion of a recycle solution saturated with $NaHCO_3$ at 100°F. Solid melt particles in the resulting slurry are ground in L-603 to facilitate dissolution of the melt stream. The remainder of the recycle $NaHCO_3$ stream is then mixed with this slurry cooling the mixture to 228°F. This stream is then flashed to 16.1 psia in F-601, where sufficient holding time is provided to dissolve the Na_2CO_3. The bottoms slurry from F-601 is filtered in L-601 to separate the ash and carbon (and some undissolved Na_2CO_3) from the solution. This residue is sent to disposal.

The filtrate from L-601 is pumped up to 30 psia in J-602 and is fed to carbonation tower E-602. In this tower the Na_2CO_3 is reacted with CO_2 from the gas purification system according to reaction (6).

$$Na_2CO_3 + CO_2 + H_2O \rightleftharpoons 2NaHCO_3 \quad (6)$$

Overhead gas from the tower at 200°F is cooled to 95°F in C-601 to condense water, a portion of it is purged to remove the impurities brought in with the CO_2 stream, and the remainder is recycled through J-604 to the tower. Fresh CO_2 is added to the tower at the rate of 10,400 moles per hour.

The operating temperature at the bottom of E-602 is 100°F. At this temperature the $NaHCO_3$ concentration exceeds the solubility limit and $NaHCO_3$ (along with some ash) is precipitated. The resulting slurry is filtered in L-602, the $NaHCO_3$ solution being pumped up to 400 psia and recycled to E-601. The filter residue ($NaHCO_3$) is dried by hot gases flowing through the drying hood and is returned to lock hoppers F-202, a&b to be recycled to the gasifier-combustor.

The raw synthesis gas produced in this section subsequently undergoes a single stage of water-gas shift, followed by CO_2- and sulfur-removal steps. The purified synthesis gas ($CO + H_2$) is then reacted over a catalyst to produce a methane-rich product having a heating value of about 910 Btu/SCF. The compressed product gas is delivered to the gas mains at 1000 psig at the rate of 250 million SCFD.

C. Process Control

Means for controlling some of the major process variables in the gasification and ash removal sections are illustrated in Figure 1. It should be emphasized that the instruments shown are intended for controlling the critical variables associated with gasification, combustion, and melt handling under normal, steady-state conditions only. Instrumentation and controls required for start-up and normal and emergency shut-downs have not been included in the present discussion. A thorough analysis of start-up procedures and possible malfunctions, which is necessary to develop a complete instrumentation plan, will be undertaken at a later date.

1. Steam-Coal Feed Rates

Flow rate of total process steam to the gasifier is recorded and controlled by FRC-1. This rate is independently set to yield the desired steam/coal ratio for gasification.

A portion of the total gasifier steam, the quantity being set independently and then controlled by FRC-2, is used to convey feed coal from lock hoppers F-201 to the gasifier. Pressure at the pickup point is measured and recorded by PR-1. Pressure drop through a fixed length of transfer line, as measured and recorded by PdRC-1, is an indication of coal loading in the steam. This measurement, together with the known steam pressure and flow rate, is used to control the flow rate of coal from the lock hopper. Ultimately, the coal rate (hence the desired setting on the differential pressure controller) is set by the flow rate and analysis of the down-stream synthesis gas.

Although it is not shown in Figure 1, the same control set-up used for coal would also be used to control the addition of $Na_2\ CO_3$-$Na\ HCO_3$ from F-202.

2. Gasifier Temperature

The temperature of the melt in the gasifier is controlled by regulating the flow rate of combustion air. Melt temperature is measured and recorded by TRC-1, and is used to position the inlet guide vanes of the centrifugal air compressor, J-201. As the melt temperature varies, the controller will adjust these vanes to permit more or less air to flow to the combustor, thereby generating more or less heat and raising or lowering the melt temperature.

3. Melt Circulation Rate

With the reactor design shown in Figure 1 the rate of melt circulation between gasifier and combustor is not controlled independently. Rather, this rate of flow will depend on operating conditions and the free area provided in the wall separating the two compartments. Differential pressure between the gasification and combustion sections will be controlled by PdRC-2 at a level consistent with the desired melt circulation rate and an acceptable rate of gas leakage. The pressure of the flue gas will be allowed to "float" (in effect, the atmosphere exerts a back-pressure on the expander and serves as a control valve), while the pressure of the synthesis gas will be controlled downstream (where the temperature is lower) to give the specified differential.

4. Melt Withdrawal Rate

The rate of melt withdrawal from the gasifier-combustor, D-201, to the quench tower, E-601, is controlled by the differential pressure between these two vessels. This differential is measured and recorded by PdRC-3 and is used to control the flow rate of quench solution to E-601. For example, if the pressure in E-601 becomes lower than its specified level (yielding a melt withdrawal rate which is greater than desired), the flow rate of quench solution will be reduced, resulting in a higher temperature, hence higher pressure in E-601 (which will restore melt flow rate to the desired level). Pressure level in gasifier-combustor D-201 is recorded by PR-2.

The remainder of the quench stream is by-passed around E-601 and is combined with the hot slurry after grinding. The temperature of this mix-stream is continuously recorded by TRC-2 and is used to control the total flow rate of quench solution.

The rate of melt removal is checked periodically by analyzing the residues from L-601 and L-602 together with the quench solution to insure that ash concentration in D-201 is remaining constant at the desired level. If this level does change, the pressure differential between D-201 and E-601 must be re-set to increase or reduce the melt withdrawal rate, as necessary.

The liquid level in E-601 is held constant by liquid level controller LLC-1 attached to the let-down valve preceding F-601.

D. Reactor Design

For the design capacity, nine (9) large reactors approximately 23 feet in diameter and 40 feet high are required to contain the primary reaction. Figure 1 shows the general process flow requirements and temperature-pressure levels in these reactors.

Although the conceptual design discussed herein is based on the use of a single vessel separated by a dividing wall, it must be recognized that two separate vessels could be used. The design considerations applied to the separated vessel are generally applicable to a two-vessel system. Since the divided vessel presents a more difficult design problem it has been chosen for consideration in this feasibility study. Figure 2 provides a conceptual mechanical design for the reactors.

Inlet lines to the reactor are unlined eight-inch Inconel pipe with 0.5-inch wall thickness, embedded in the center dividing wall. Six air inlet lines and six coal-steam-carbonate lines are provided. The air lines are sized on the basis of approximately 150 feet per second. The coal-steam-carbonate lines are sized on the basis of approximately 30 feet per second to minimize erosion and particle damage. This velocity is sufficient to transport the solids particles

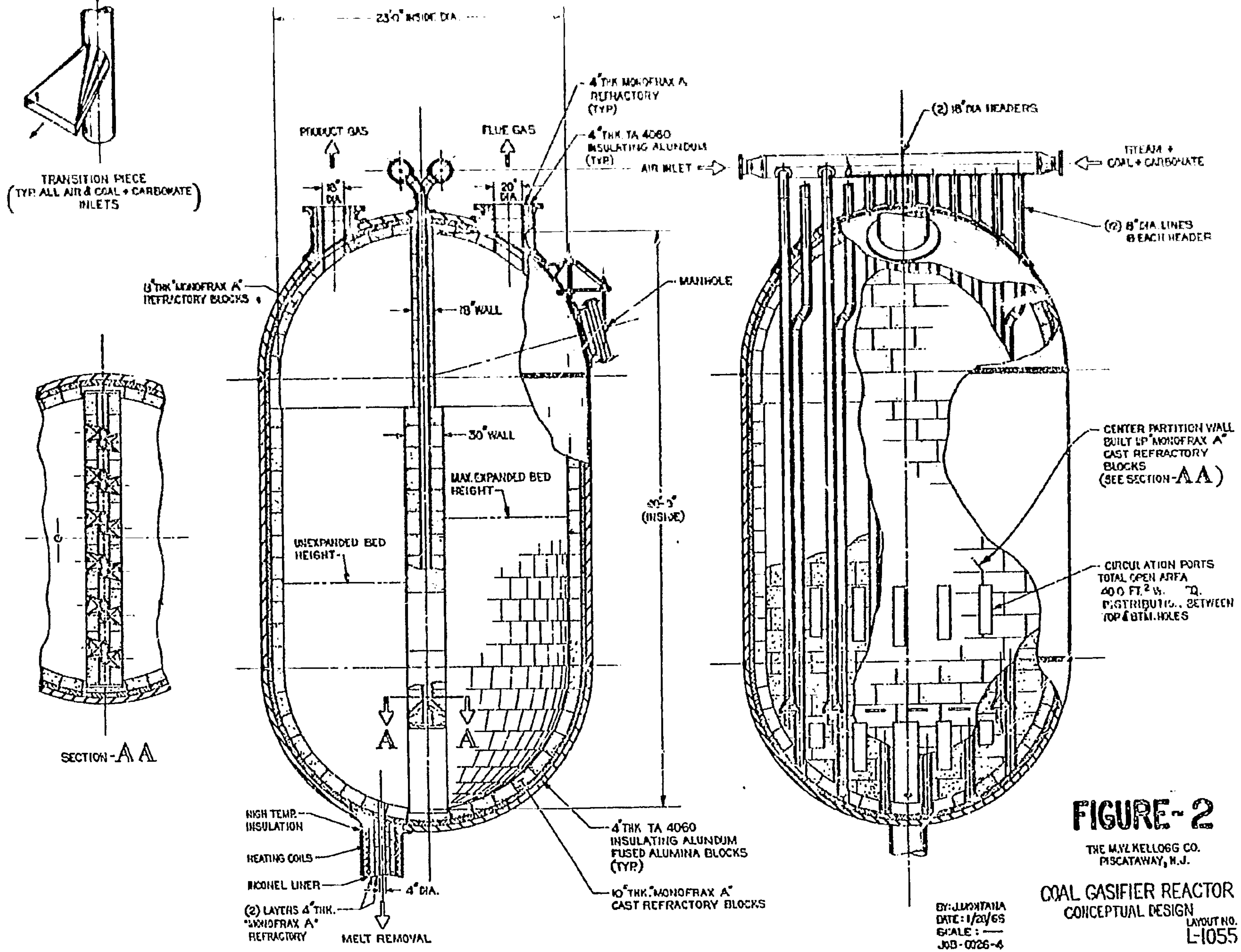

FIGURE-2

THE M.W. KELLOGG CO.
PISCATAWAY, N.J.

COAL GASIFIER REACTOR
CONCEPTUAL DESIGN

LAYOUT NO.
L-1055

THE M. W. KELLOGG COMPANY
A DIVISION OF PULLMAN INCORPORATED
RESEARCH & DEVELOPMENT DEPARTMENT

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at the anticipated loading of approximately one pound per cubic foot. It may prove desirable to introduce the coal in just a portion of the steam. Thus one or two of the feed lines would carry a steam-coal-carbonate mixture, and the remaining four or five would carry only steam.

Both the product gas and flue gas lines are internally lined with four inches of Monofrax A backed by four inches of insulating alumina similar to Norton 4060. The high density and strength at temperature of Monofrax A allows both lines to be sized on the basis of approximately 150 feet per second for short runs into internally-lined solids separators (covered elsewhere in this report). An 18-inch ID product line and a 20-inch ID flue gas line are provided in the design. Both outlets are located at the top of the vessel on their respective sides of the separating partition.

The melt removal line is sized to withdraw the entire melt contents of the vessel in less than two hours, if emergency shutdown is required. The removal rate is limited by the capacities of the melt quench tower and dissolving drums to quench and process the hot melt.

A four-inch ID internally-lined, externally-heated, double-walled pipe is provided to prevent freezing melt from plugging the removal line. Under normal operating conditions external heat is unnecessary since the withdrawal flow (one to two feet per second) is sufficient to allow for heat loss in this line with less than 100°F temperature drop in the melt (approximately 1830 to 1730°F). This is well above the melt solidifying temperature of approximately 1540°F. Emergency melt removal will result in a 10 fps flow in this line.

With consideration to the process requirements, all valves or similar mechanisms have been eliminated from direct contact with the molten melt. A more detailed discussion of the instrumentation and control of the major process variables has already been presented in Section C of this report.

In an emergency, the quench tower can be isolated by allowing the back pressure to rise sufficiently to stop the melt flow and allowing the melt to freeze in the transfer line. This will provide a sort of solid plug valve. The quench vessel could then be depressurized for inspection or other work. Operation could be resumed by simply repressurizing the quench tower and melting the transfer line plug by the electrical heaters provided.

The internal lining construction is designed to provide for a minimum of ten years life before replacement is required. The Monofrax A block lining in the melt and melt splash areas has a minimum thickness of 10 inches. Heat transfer calculations have established that the temperature gradient through the Monofrax A is such as to insure that melt penetrating between the blocks will solidify within seven inches of the melt-block surface. If the corrosion rate is taken as 0.2 inches per year (greater than the experimentally-determined corrosion rate), eight inches of block will be left after ten years of service. Thus, after ten years, a margin of safety of one-inch in block thickness will remain before liquid melt may be expected to reach the back-up insulation. Since the temperature gradient will increase with loss of material from the block face, and since the corrosion rate is expected to be less than 0.2 inches per year, the service life of the proposed lining may well be between 12 and 15 years. Backing up the Monofrax liner is a four-inch layer of insulating alumina blocks set with suitable binder to prevent any short-term leakage of melt from reaching the carbon steel shell.

The anticipated operating shell temperature of the reactor is about 600°F, which is well below the freezing temperature of the melt. Attachment or built-up areas of the shell should be suitably cooled to the 600°F level. The underside of the vessel must be adequately ventilated to prevent hot spots from occurring. Steam lancing or water spray cooling should be provided for emergency treatment of hot spots.

Eight-inch thick blocks of Monofrax are used above the melt or melt splash zone for corrosion and structural strength reasons. Four inches of insulating alumina is also used as back-up in this area. The overhead arch is self-supporting and self-keyed.

The proposed construction offers maximum corrosion protection to the pressure shell while maintaining process heat loss at less than 1% of the reaction heat input.

E. Dividing Wall Construction

The center dividing partition is of stepped, self-keyed construction. The partition wall is 30-inches thick below the melt and melt splash level to provide structural strength to resist the forces of sloshing melt.

Above the melt splash level the wall thickness is reduced to 18 inches. The construction of the wall is such that with the stiffening effect of the air and coal-steam-carbonate lines, sloshing forces of the melt bed equal to the differential head developed by one-half the total bed height can be withstood. While this criterion is reasonable, it is possible to make the wall stronger by using heavier-walled pipe or rearranging the piping within the wall.

Aside from the stiffening effect in the dividing wall, entry of the inlet lines through the partition wall rather than the bottom of the reactor offers several other advantages.

First, the lines will tend to be self-clearing in that gravity is in their favor. Any melt entering these lines will be forced out by the inlet flow. If the lines should become clogged with a plug of melt, the stoppage of inlet flow will cause the lines to be heated to a greater extent by the melt in contact with the wall, causing the plug to remelt.

Second, inlet flow can be directed sideways or down without complicated vessel internals. This should improve contacting in the beds and reduce the possibility of solids or liquid carry-over.

Third, the inlet lines help to key the wall blocks and the wall to the reactor shell. The coal-steam-carbonate lines are extended to the bottom head lining to accomplish this and to provide a self-sustaining deadman to reduce inlet ercsion at the final turn into the reactor. The deadman will be filled with a refractory plug to prevent a trapped accumulation of coal and carbonate particles.

Expansion of the inlet lines is accommodated by allowing the inlet lines to grow into the reactor when heated. Inlet line cavaties are filled with a compressible insulation to provide short time sealing against melt filling the cavities and binding the pipe. In normal operation heat transfer calculations indicate the colder inlet flows will cause the melt to freeze in the Monofrax wall block prior to reaching the pipe cavities.

If pilot plant experience indicates that expansion of these lines is a problem with this configuration, the inlet lines could be fixed at the bottom and expansion joints or packed slip joints used. Since the inlet lines are constructed of Inconel, short-time exposure to the melt at temperature can be tolerated.

F. Circulation in the Reactor

Approximately 40 square feet of hole area has been provided in the center partition of the reactor to allow melt circulation from one side to the other. This is necessary to transfer the required reaction heat from the combustion side to the reaction side. Although small model tests with a 200-centipoise-viscosity simulated melt are planned to study the reactor circulation, pilot-plant tests with actual melt will be necessary to confirm any preliminary results. The hole size can be adjusted as required. Additional design information is required to determine the extent of cross-contamination of the vapors on each side of the partition resulting from leakage through the partition and entrainment in the circulating melt. If necessary to reduce the contamination, separate vessels could be used for the reaction and combustion process connected by U-bend transfer lines. Such a design would obviously be more expensive and complicated and heat loss would be somewhat increased. The design, however, also appears technically feasible.

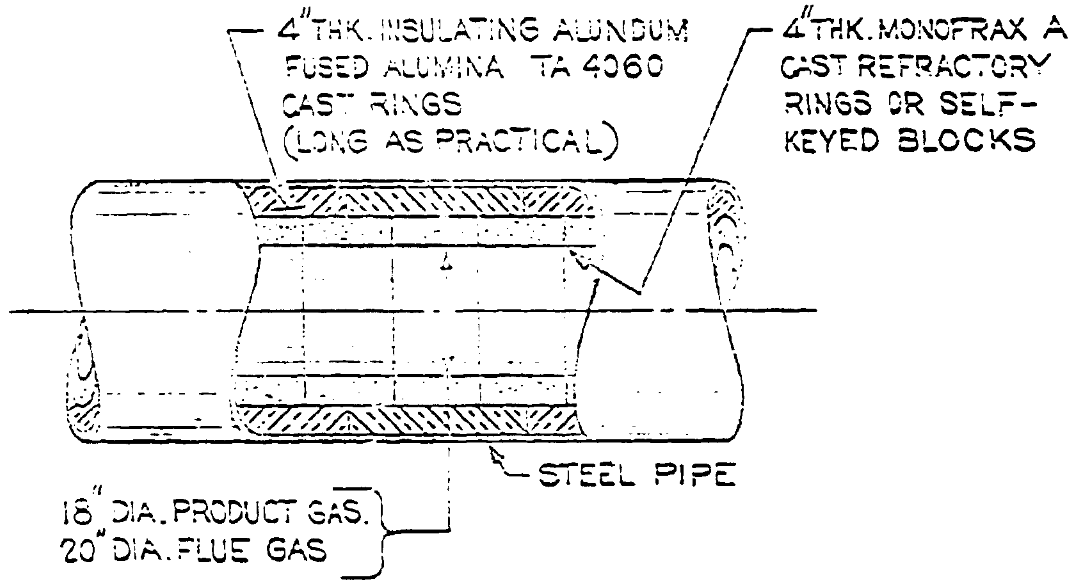
G. Reactor Construction

The reactor vessel shells, while large and of relatively high pressure for their size, can be fabricated with current technology and facilities. The shell may be shop-fabricated or combination shop and field-fabricated, depending on transportation facilities and location of the commercial plant. Internal linings must be field-fabricated due to the nature of their construction. Access to the inside of the vessel is provided through one or two large manholes in the vessel head, depending on final cross-flow hole configuration in the partition wall.

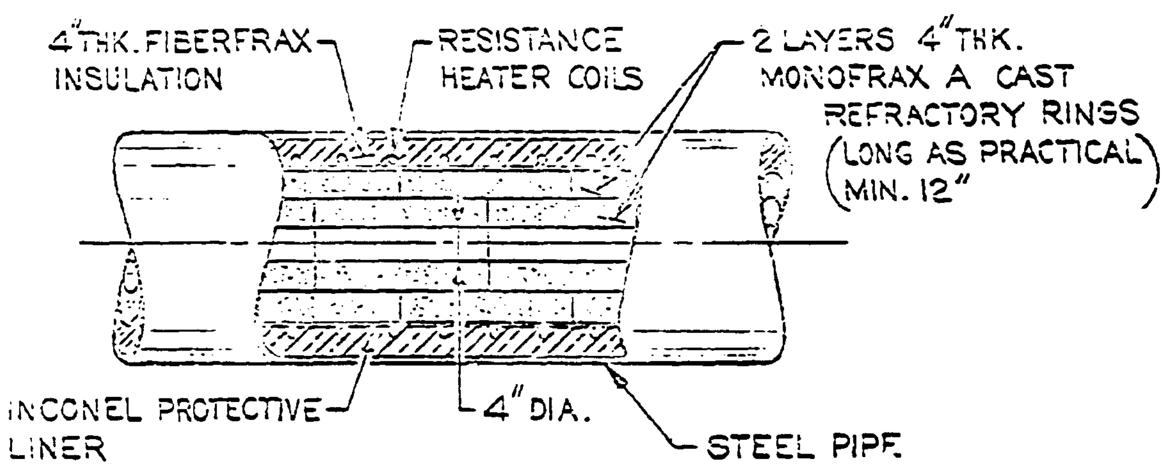
H. Transfer Line Construction

Typical constructions of the various critical transfer lines are shown on Figure 3. Generally the product and flue gas transfer lines are internally lined with four inches of Monofrax A followed by four inches of insulating alumina. Construction of the Monofrax lining is of self-keyed block or cast ring sections. The insulating alumina is in fused cast ring sections of lengths as long as practical to prevent melt infiltration to the transfer lines. These lines should be kept as short and as straight as possible to minimize cost and erosion of the lining surfaces.

TRANSFER LINE CONSTRUCTION



PRODUCT GAS & FLUE GAS LINE (TYP.)



MELT TRANSFER LINE

FIGURE-3

J.M. 2/5/63

The melt transfer line is lined with two staggered rows of four-inch thick ring sections of Monofrax A. This is followed in turn by an Inconel protective pipe, electrical resistance heating coils, a high-temperature insulation such as Fiberfrax O, and the carbon steel pressure shell.

I. Melt Quench Tower

Figure 4 shows the proposed construction of the melt quench tower. The tower is essentially a water-wall drum with a Monofrax distributor cone set on top of a water spray stand pipe. The eight-foot diameter by twenty-foot vessel provides a minimum holding time and steam disengaging space. A partially-dissolved slurry of solidified melt particles is removed from the bottom of the tower and transferred to a dissolving and settling drum by back-pressure-controlled blowdown. Sufficient quench water is provided to completely eliminate the net production of steam since steam produced locally by water contacting the melt is condensed within the quench tower.

J. Overhead Melt Separators

The product gas and flue gas lines from the reactor may carry over entrained particles of reactor melt. To eliminate this from the exit streams, external overhead separators are included in the feasibility design. Figure 5 shows one possible configuration of these separators. The product gas separator is approximately eight feet in diameter by twenty feet in length. Corresponding size for the flue gas separator is ten feet by twenty feet. The collected melt is returned to the reactor in a transfer line similar in design to the melt removal line previously described. Separated vapors or gas are sent to downstream heat exchangers where the heat is recovered against reactor inlet gases.

Pilot plant operation will be needed to confirm the requirement for these separators and establish the degree of baffling required for acceptable separation.

K. Lining Costs

Cost of the Monofrax for lining the vessels and transfer lines is estimated -- very approximately -- at about 6 million dollars. If DCL blocks (diamond-cut blocks) were needed, the cost would increase by about 50%. Cost of replacing the linings every ten years would be about 1¢/MSCF of pipeline gas, even if DCL blocks were used.

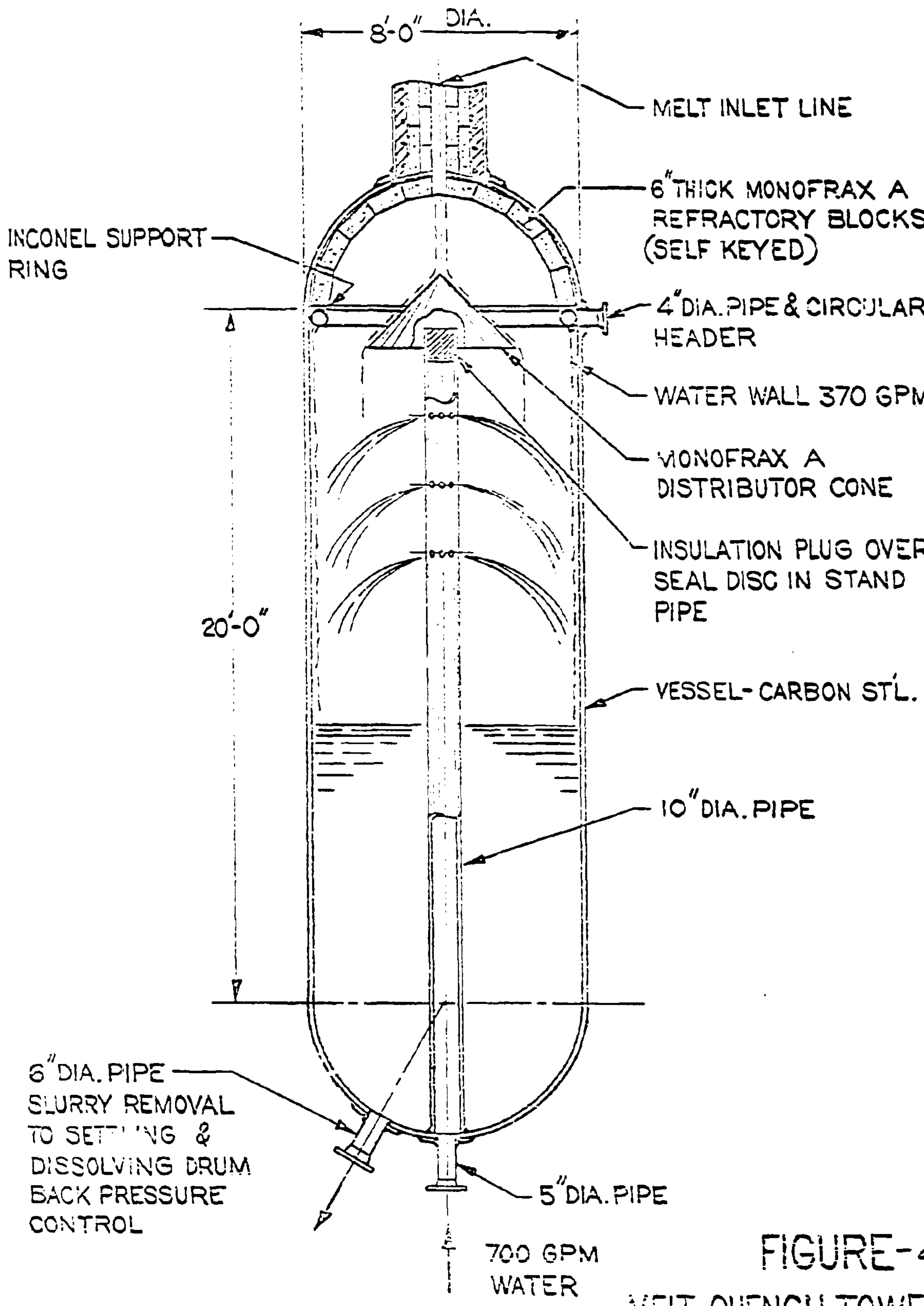


FIGURE-4
MELT QUENCH TOWER

J.M. 2/1/66

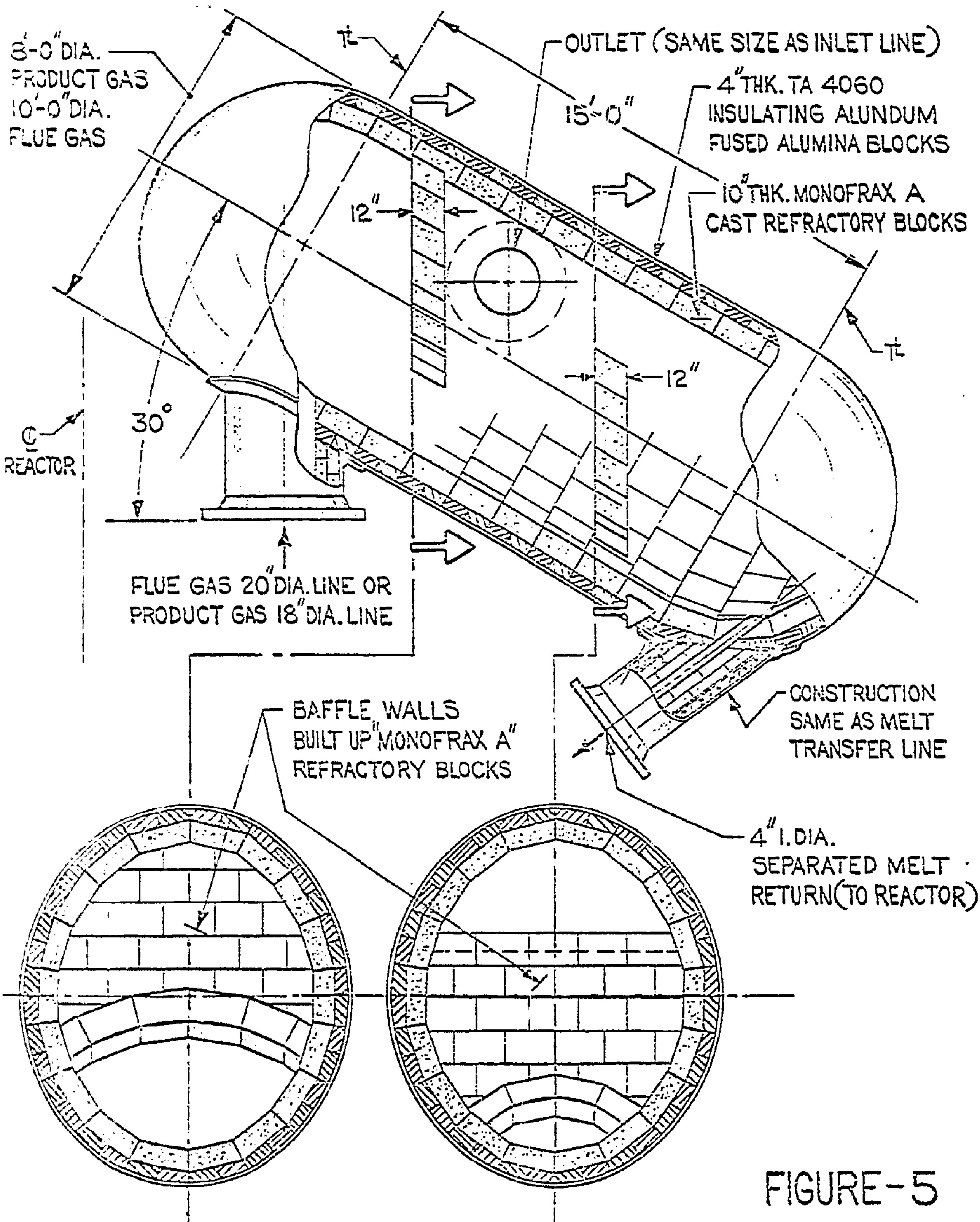


FIGURE-5
OVERHEAD SEPARATOR

J.M. 2/2/63 SCALE: 1/4

L. Economics

Estimated fixed investment for the gasification and ash removal sections of a 250-million SCFD pipeline gas plant utilizing the Kellogg Gasification Process is $73,500,000. Using this figure and the materials, utilities, and labor requirements for these sections, the cost of raw synthesis gas expressed as cents per MSCF of pipeline gas has been calculated and is presented in Table 2. Costs have been estimated according to the OCR Tentative Standard dated June 4, 1965. It should be emphasized that these costs represent only the contributions of gasification and ash removal to the total manufacturing cost and selling price of the pipeline gas.

An economic comparison has been made of this present scheme with that originally proposed to OCR on November 1, 1962. (1). Although this latter plant is based only on 90 million SCFD, the comparison should still be valid because at these large capacities little, if any, economic advantage will be realized by increasing the capacity. Estimated investment for the gasification section of this plant was $23,500,000. Once again, gas cost has been calculated and is presented in Table 3.

As can be seen by comparing Tables 2 and 3, the present economics compare quite favorably with those of the proposal. Thus, although the process is at a much higher state of development today than at the time of the original proposal, the costs have not changed significantly. It appears as if the goal of a pipeline gas selling price in the range of 50¢/MSCF is within reach.

(1) OCR Proposal No. 219, "Proposal For the Development of the Kellogg Coal Gasification Process", November 1, 1962.

TABLE 2

ECONOMIC SUMMARY FOR GASIFICATION AND ASH REMOVAL SECTIONS

COST OF RAW SYNTHESIS GAS EXPRESSED AS CENTS/MSCF OF PIPELINE GAS

Bases: 250,000,000 SCFD of Pipeline Gas from Bituminous Coal
90 Percent Stream Efficiency

Total Fixed Investment = $73,500,000.

Item	Unit Cost	Units/MSCF PLG	¢/MSCF PLG
Coal	$4/ton	0.0528 tons	21.1
Na_2CO_3	1.55¢/lb.	1.81 lb.	2.8
Operating Labor	$3.20/man-hour	0.00384 man-hours	1.2
Steam	27¢/1000 lb.	12.4 lb.	0.3
Power	0.5¢/KWH	7.37 KWH	(3.7) credit
Boiler Feed Water	25¢/1000 gal.	11.0 gal.	0.3
Cooling Water	1.5¢/1000 gal.	415 gal.	0.6
Maintenance @ 3% of bare cost per year			2.3
Supplies @ 15% of maintenance			0.3
Supervision @ 10% of operating labor			0.1
Payroll Overhead @ 10% of operating labor + supervision			0.1
General Overhead @ 50% of (operating labor + supervision + maintenance + supplies)			2.0
Plant Operating Expenses			27.4
Depreciation @ 5% of fixed investment per year			4.4
Local Taxes and Insurance @ 3% of fixed investment per year			2.7
Sub-total			34.5
Contingencies (2% of sub-total)			0.7
TOTAL MANUFACTURING COST			35.2
GAS SELLING PRICE (Based on 20-year average revenue requirement)			40.3

TABLE 3

ECONOMIC SUMMARY FOR GASIFICATION AND ASH REMOVAL SECTIONS

COST OF RAW SYNTHESIS GAS EXPRESSED AS CENTS/MSCF OF PIPELINE GAS

Bases: 90,000,000 SCFD of Pipeline Gas from Bituminous Coal
90 Percent Stream Efficiency

Investment and Utilities taken from OCR Proposal No. 219, November 1, 1962
Total Fixed Investment = $23,500,000

Item	Unit Cost	Units/MSCF PLG	¢/MSCF PLG
Coal	$4/ton	0.047 tons	18.8
Na_2CO_3	1.55¢/lb.	0.58 lb.	0.9
Operating Labor	$3.20/man-hour	0.003 man-hours	1.0
Steam	27¢/1000 lb.	31.2 lb.	0.8
Power	0.5¢/KWH	5.28 KWH	2.6
Boiler Feed Water	25¢/1000 gal.	10.6 gal.	0.3
Cooling Water	1.5¢/1000 gal.	61.9 gal.	0.1
Maintenance @ 3% of bare cost per year			2.0
Supplies @15% of maintenance			0.3
Supervision @ 10% of operating labor			0.1
Payroll Overhead @ 10% of operating labor + supervision			0.1
General Overhead @ 50% of (operating labor + supervision + maintenance + supplies)			1.7
Plant Operating Expenses			28.7
Depreciation @ 5% of fixed investment per year			4.0
Local Taxes and Insurance @ 3% of fixed investment per year			2.4
Sub-total			35.1
Contingencies (2% of sub-total)			0.7
TOTAL MANUFACTURING COST			35.8
GAS SELLING PRICE (Based on 20-year average revenue requirement)			40.3

IV. MANPOWER AND COST ESTIMATES

Figure 6 shows the projected manpower breakdown for Phase I for 1966 as well as the actual effort that was made. It can be seen that a 14.5 man-effort was made during January.

Figure 7 shows the expenditures during January. For the month $24,036 was expended, not including fee and G & A. The total expenditures through January were $339,532. Including fee and G & A the total expenditures were $390,130. This is 65% of the encumbered funds.

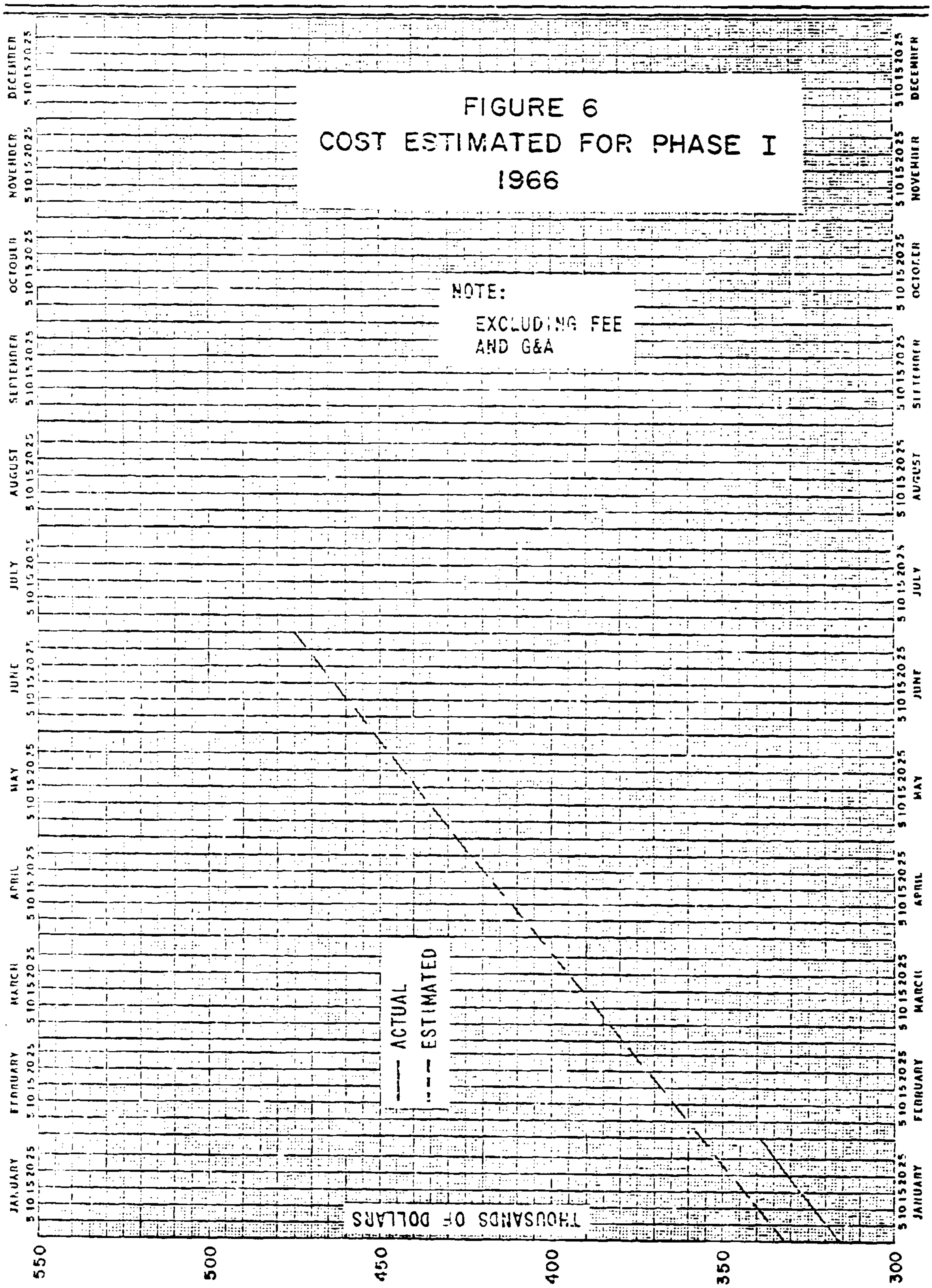


FIGURE 6
COST ESTIMATED FOR PHASE I
1966

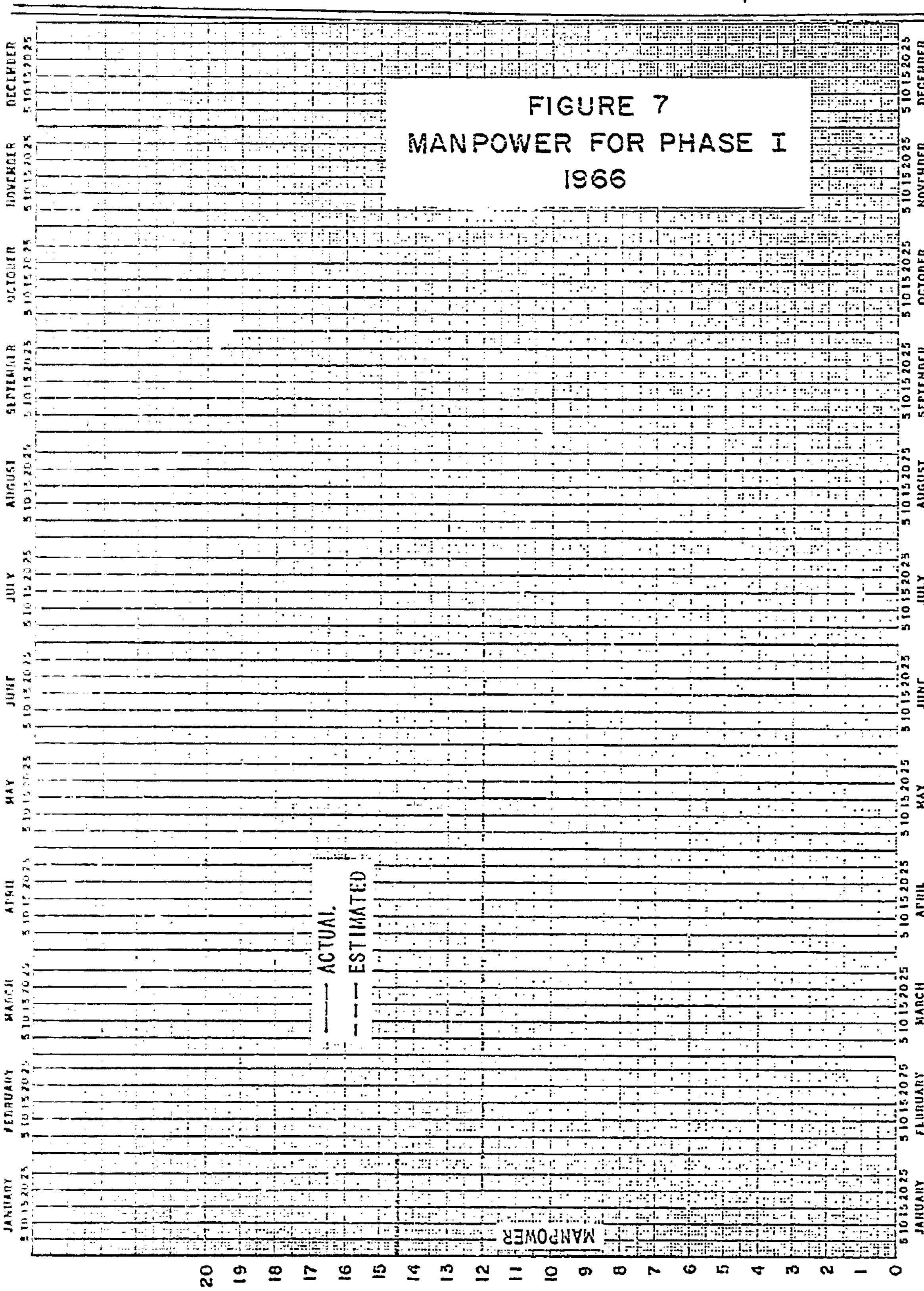


FIGURE 7
MANPOWER FOR PHASE I
1966