MWK-MPR-23

RESEARCH AND DEVELOPMENT DEPARTMENT

DEVELOPMENT OF KELLOGG COAL GASIFICATION PROCESS

Contract No. 14-01-0001-380

June 30, 1966

Progress Report No. 23

APPROVED: _______________
Project Manager

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RESEARCH & DEVELOPMENT DEPARTMENT

CONTENTS

I. SUMMARY

This progress report is the twenty-third since the awarding of the contract. It is concerned with the first phase of the contract and summarizes the progress that has been made in the three principal areas now being studied: process research, chemical engineering studies, and mechanical development.

Gasification experiments were carried out for five feed materials and the reaction rates determined. The five materials used were a caking bituminous coal, a coke derived from that coal, a subbituminous coal, a lignite and an anthracite. The results of these experiments indicated that the gasification rate of bituminous coke is about the same as that of anthracite. Experiments with raw bituminous coal, however, resulted in a rate about 20 percent lower than that for either anthracite or coke. Since bituminous coal is generally expected to be more reactive than anthracite, there was a suspicion that the feed material was agglomerating either on the melt surface (due to top feeding) or in the inlet lines resulting in large particles and low rates. To test this theory, raw coal was pre-oxidized in air at low temperatures ($<450^{\circ}F$) until a non-agglomerating material was formed. When this oxidized coal was gasified, a rate about 40 percent higher than anthracite was obtained--tending to confirm the agglomeration theory. Lignite and subbituminous coal resulted in rates about 70 and 100 percent higher than anthracite, respectively.

A complete "process package" is presented for the production of 247 million standard cubic feet of pipeline gas a day from subbituminous coal. The package includes a process flowsheet, process description, material balance, utilities summary, capital cost estimate, and computation of gas production cost and selling price. Estimated capital investment is about 147 million dollars. Gas selling price is about 44¢/MSCF, based on $2/ton coal and the OCR standard procedure for calculating gas cost. The plant also produces about 203 megawatts of electricity which is sold as a by-product at 4.5 mills/kwh, yielding a credit of 8.9¢/MSCF of pipeline gas.

A study was made of the effect of reducing the melt depth in the gasifier from the design level of about 19 feet to 8 feet and 4 feet. Lowering the allowable depth caused an increase in number of gasifiers as well as in the amount of labor required to operate them. For 8 foot and 4 foot depths, gas cost is increased by about 3.9 and 10.2¢/MSCF, respectively.

Preparation of a "process package" for the manufacture of hydrogen from bituminous coal has been begun, and a preliminary processing sequence has been determined.

Corrosion Test #9, a long-term test under simulated gasification conditions, was interrupted after 407 hours by a failure of the support which held the test specimens in the melt. The specimens of Monofrax A which were being tested while under a compressive loading of 40 psi were cracked (but not corroded) when removed from the furnace after the support failure. However, up until that time the specimens had not shown any evidence of failure indicating that the failure of the support might have caused the specimens to fail.

Experiments were continued to study the melt quenching step required in the ash removal section. This work has shown that conditions are available under which little or no melt agglomeration has been detected.

II. PROCESS RESEARCH

A. Accomplishments

Gasification rates of five feed materials were determined in the 2-inch ID Inconel reactor, at two temperatures, 1640 and 1740°F, under about 3 atmospheres absolute steam pressure, with molten sodium carbonate melts containing 2 per cent of ash derived from the feed material. Other conditions of the runs were 1.0 ft./sec. superficial gas velocity, 4-inch quiescent bed height, and 4% carbon in bed initially (based upon the fixed carbon content of the feed). The results are presented graphically in Figure 1, and all the runs are summarized in Table I.

It is quite evident from Figure 1 that most of the process research information has been generated on the least reactive feedstock, namely, bituminous coal (Island Creek #27, W. Va.) or coke derived from this coal. In effect, the coke behaved like anthracite, normally considered to be the least reactive coal.

However, in testing bituminous coal (runs H-13, 20, 26) the gasification rates obtained were lower than for the coke derived from this coal and lower than those for gasification of anthracite--a very unexpected situation. Since this is a caking coal it was suspected that agglomeration of the charged 12/20 mesh material occurred when the coal hit the top of the hot bed or while passing down the hot reactor from the hopper. A qualitative visual experiment in a muffle furnace showed that of all the feedstocks only the bituminous coal agglomerated.

Oxidation of the bituminous coal in air at temperatures less than 450°F produced a non-agglomerating material which showed much higher gasification rates than the unoxidized coal (runs H-28 and 32). In run H-33, at 1840°F, suspicion of agglomeration from details of the run nullified its use. With bottom feeding of coal entrained in a gas, as envisioned for the commercial unit, agglomeration of a group of coal particles resting momentarily on the hot melt should not be a problem.

Elkol subbituminous coal appeared to be the most reactive material tested. It was 2.7 and 2.1 times more reactive than anthracite at 1640 and 1740°F, respectively.

In calculating these runs, it soon became evident that in order to obtain good material balances carbon initially in the bed had to be calculated on the basis of fixed carbon in the feedstock. About 7 to 14% of the total carbon shows up in the gas during devolatilization, while 14 to 26% of the total carbon is driven out as tar and lost when the solid hits the molten salt. Using fixed carbon in the feedstock weight balances are quite satisfactory.

In determining the first order kinetic constant for these runs on a fixed carbon basis, a considerable amount of carbon as mentioned above does not enter into the kinetics. In a commercial unit, this carbon, or at least most of it, would be gasified and hence would tend to increase the overall gasification rate. To this extent, the rate constants determined experimentally thus far are conservative and fall below the rate constants expected under continuous commercial operation.

The Arrhenius plot in Figure 1 has yielded the following apparent energies of activation for each feedstock at the conditions given.

	E_a - kcal
Elkol subbituminous coal	14
S. Beulah lignite	19
Oxidized W. Va. bituminous coal	27
Anthracite	28
W. Va. bituminous coke	34

B. Projections

Two other feedstocks remain to be tested, namely, a char from FMC (from an Illinois coal) and Renners Cove lignite. This will complete the evaluation of gasification rates of the different feedstocks. Work on combustion, particularly at pressure, will be resumed.

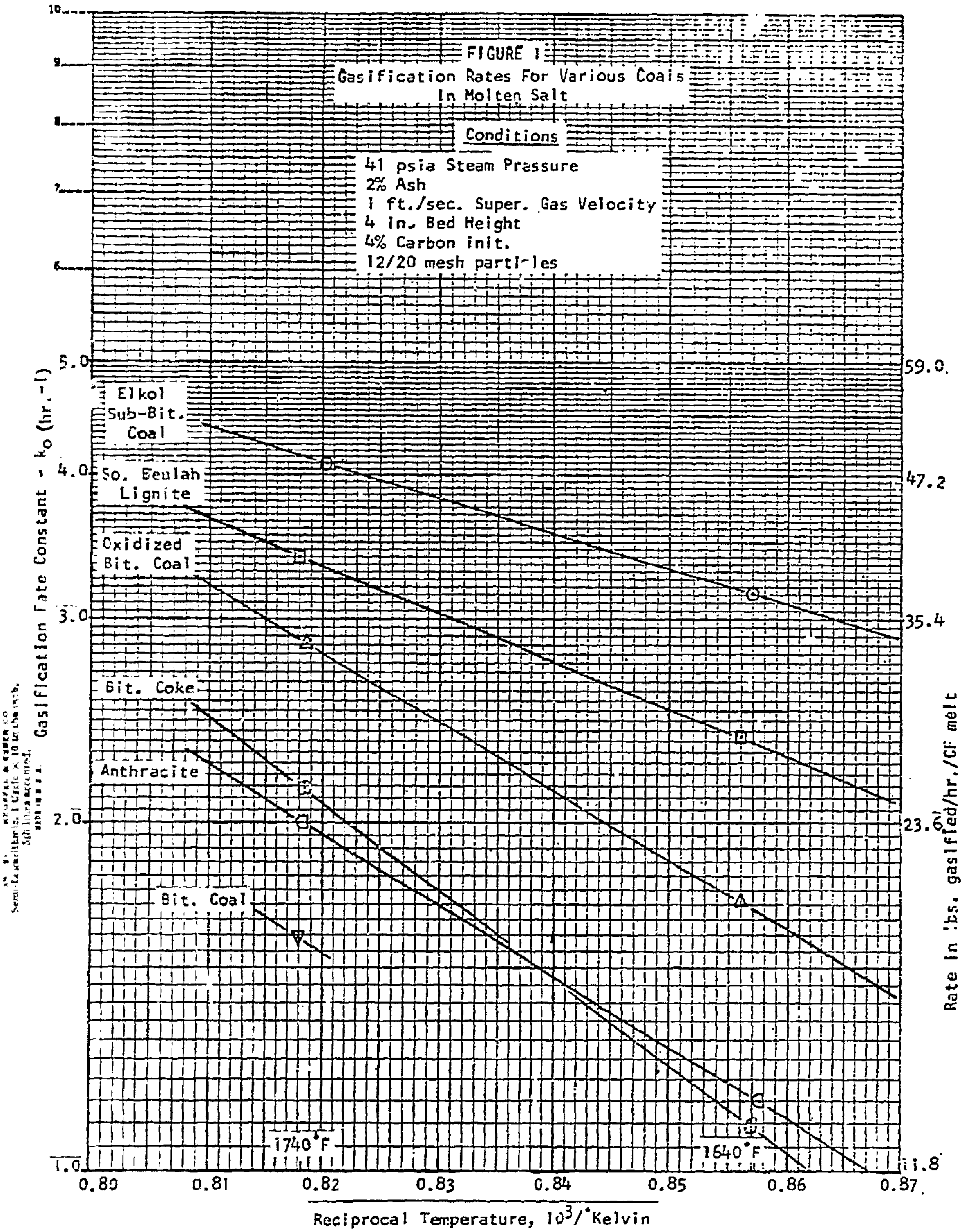
FIGURE 1
Gasification Rates For Various Coals In Molten Salt
Conditions
41 psia Steam Pressure
2% Ash
1 ft./sec. Super. Gas Velocity
4 In. Bed Height
4% Carbon init.
12/20 mesh particles
Elkol Sub-Bit. Coal
So. Beulah Lignite
Oxidized Bit. Coal
Bit. Coke
Anthracite
Bit. Coal
1740°F
1640°F
Gasification Rate Constant - k_o (hr.$^{-1}$)
Rate in lbs. gasified/hr./CF melt
Reciprocal Temperature, 10^3/°Kelvin
10
9
8
7
6
5.0
4.0
3.0
2.0
1.0
59.0
47.2
35.4
23.6
11.8
0.80
0.81
0.82
0.83
0.84
0.85
0.86
0.87

TABLE I

GASIFICATION OF VARIOUS COALS IN MOLTEN SODIUM CARBONATE(1)

Run No. H-	13	15	16	17	18	19	20	26	27	28	29	30	31	32	33
Date - 1966	5/26	6/1	6/3	6/3	6/7	6/7	6/9	6/10	6/14	6/14	6/16	6/21	6/22	6/24	6/28
Feed	Bit. Coal	Anthracite	Elkol Sub.	Bit. Coal	S. Beulah	Lignite	←Bit.	Coal→	Bit. Coke VI	Oxid. Bit. Coal	Anthracite	Elkol Sub. Bit	S.B. Lignite	Oxid.	Bit. Coal
% Fixed Carbon	58.3	82.5	48.5	48.5	42.9	42.9	58.3	58.3	-	61.4	82.5	48.5	42.9	58.3	58.3
% Total Carbon	81.3	82.5	73.5	73.5	66.4	66.4	81.3	81.3	93.2	81	82.5	73.5	66.4	81.3	81.3
% Vol. Matter	37.3	5.8	48.3	48.3	44.0	44.0	37.3	37.3	0.6	34.5	5.8	48.3	44.0	37.3	37.3
% Ash	3.9	11.7	3.2	3.2	13.1	13.1	3.9	3.9	6.2	4.1	11.7	3.2	13.1	3.9	3.9
gms. charged	29.6	20.5	23.5	35.6	26.0	40.2	29.6	29.6	19.0	28.1	20.5	35.6	40.2	29.6	29.6
mesh size	←12/20→														
Na_2CO_3 - gms.	(2)	405.7	405.	(2)	405.7	(2)	405.7	(2)	(2)	(2)	405.7	405.7	405.7	405.7	(2)
Ash - gms.	-	8.3	8.3	-	8.3	-	8.3	-	-	-	8.3	8.3	8.3	8.3	-
- % in Melt	2	2	2	2	2	2	2	2	2	2	2	2	2	2	2
% C in Melt - Init., Fixed C	4.0	3.9	2.7	4.0	2.6	4.0	4.0	4.0	4.0	4.0	3.9	4.0	4.0	4.0	4.0
Bed Height - In.	←4→														
Conditions							(3)	(4)							
Temp. - °F Ave.	1737	1740	1741	1735	1740	1741	1739	1741	1640	1739	1639	1641	1644	1644	1838
Pressure - psia	45.5	45.5	45.9	45.9	45.5	45.7	46.0	45.3	44.9	44.6	45.1	45.4	45.2	43.0	45.4
Steam Pressure psia	41.4	41.0	41.4	41.2	41.2	41.4	42.3	42.1	40.9	41.0	40.9	41.6	41.6	39.2	41.1
Gas in Steam	←N_2→														
Ft./Sec. Steam & Gas	1.03	1.04	1.04	1.05	1.03	1.03	1.02	1.00	1.03	1.01	1.03	0.98	0.97	0.98	1.08
Minutes to 0% CO	35	30	20	25	20	25	40	40	50	30	45	25	30	40	25
Minutes - Total Run	35	30	20	25	20	25	40	40	50	30	45	25	30	40	30
cc H_2O in /hr.	1192	1193	1197	1192	1200	1192	1196	1192	1197	1193	1197	1197	1193	1195	1196
cc N_2 in /min.	2470	2711	2729	3008	2605	2605	2154	1866	2392	2142	2585	2232	2124	2386	2565
Results															
% C Devolatilized	12.8	-	11.3	13.5	10.3	10.6	12.2	0	-	6.8	-	12.0	10.7	8.7	6.5
% C to Tar and Loss	15.6	-	23.9	21.4	25.7	22.8	14.1	14.0	-	21.5	-	20.7	22.7	19.6	28.3
% C to oxides and CH_4	71.6	90.3	64.8	65.1	64.0	66.6	73.7	81.7	90.5	71.7	97.9	67.3	66.6	71.7	65.2
% C to oxides - basis Fixed C	99.7	98.3	98.1	98.6	99.1	103.2	102.7	95.7	90.5	94.6	97.9	101.9	103.0	98.6	90.8
Gasif. Rate Constant - $hr.^{-1}$															
k_1 - Input	1.59	1.94	3.83	4.09	3.91	3.60	1.42	1.31	0.98	2.57	1.00	3.21	2.47	1.67	2.57
k_o - output	1.59	2.30	3.91	4.09	4.00	3.40	1.34	1.42	1.09	2.86	1.15	3.16	2.37	1.71	3.00
Rate-at 4% C in bed lbs C/hr./CF	19	21	46	48	47	40	16	17	13	34	14	37	28	20	35
Salt Carryover-gms	1.4	11.1	3.7	4.9	7.3	5.7	9.4	5.2	15.7	12.4	16.5	5.6	6.7	12.4	6.3

(1) Used 2-inch ID Inconel reactor. Bit. Coke VI made at 950°C.
(2) Reused melt from previous run plus makeup.
(3) Used 0.5 ft./sec. N_2 when coal added to get more initial mixing.
(4) No devolatilization, steam on when coal charged.

III. CHEMICAL ENGINEERING STUDIES AND DEVELOPMENT

A. Accomplishments

A complete "process package" has been developed for the production of 247 million standard cubic feet of pipeline gas a day from subbituminous coal using the Kellogg Molten Salt Gasification Process. The package includes a process flowsheet illustrating the conceptual design of the plant, a capital cost estimate, a computation of gas production cost and selling price, a material balance and a utilities summary. The components of this package are presented in the following sections:

1. Process Description

The process flowsheet for a plant capable of producing 247,000,000 SCFD of pipeline gas from subbituminous coal is presented as Figure 2. Flow rates and compositions of the various streams on Figure 2 are shown in Table II In addition, a section-by-section material balance is given in Table III. A brief description of the flowsheet follows. The details of the flowsheet which are not presented herein may be found in the process description previously given for bituminous coal.[1]

a. Section 100 - Coal Storage and Preparation

During eight hours each day, coal is received by truck or conveyor belt from an adjacent coal mine at the rate of 2250 tons per hour. The raw coal travels by belt conveyor to a coal distribution center, where about 750 tons per hour is dispatched for immediate use, and the remainder is conveyed to the storage area. The remainder of Section 100 is the same as described for the bituminous coal flowsheet.[1]

(1) Progress Report No. 21, Contract No. 14-01-0001-380, April 30, 1966.

b. Sections 200 and 600 - Gasification and Ash Removal

Coal from Section 100 is fed to lock hoppers F-201 a and b from which it flows to the gasifier-combustor D-201 at a rate of 1,524,700 pounds per hour using the same method described previously.(1) In the gasifier, the coal is contacted with 1,052,000 pounds per hour of steam, the entire amount of which is generated in waste heat boilers C-204, C-205 and C-206. An additional 51,500 pounds per hour of steam is generated therein and exported for use in other parts of the plant. Raw synthesis gas leaves the gasification section at the rate of 129,531 moles per hour.

Air for the combustion of coal in the melt is supplied at the rate of 4,924,510 pounds per hour. Flue gas from the combustion is cooled to 1500°F in exchangers C-205 and C-202. The gas is then expanded to 18 psia and 625°F which provides the energy for air compression plus an additional 92,100 kw of electricity generated in J-203. The expanded flue gas is then cooled to 325°F in exchangers C-206 and C-207 and is vented to the atmosphere at the rate of 5,539,400 pounds per hour.

The ash left in the melt by the gasification and combustion 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 treated in the same manner as described in Progress Report No. 21. The processing is the same up to the carbonation tower. Since there is sufficient CO_2 from gas purification to convert the required amount of Na_2CO_3 to $NaHCO_3$, no recycle is required. About 22,450 moles per hour of CO_2 from Section 400 are fed to the tower at 95°F and 35 psia. The CO_2 and water vapor from the top of the tower is vented to the atmosphere at 200°F. The remainder of Section 600 is as previously described.(1)

Because of the large volumes of gases processed, Sections 200 and 600 consist of nine parallel trains of operating equipment.

c. Section 300 - Shift Conversion

Synthesis gas leaves Section 200 at the rate of 129,531 moles per hour and flows to Section 300 where about 58 percent of it is fed to the shift converter D-301. Boiler feed water at 85°F is fed between beds of catalyst at the rate of 164,800 pounds per hour to absorb a portion of the heat of reaction. The remainder of the synthesis gas is bypassed and combined with the shift effluent. The combined stream, at 733°F and having a H_2/CO ratio of about 3.15/1, is cooled

(1) Progress Report No. 21, Contract No. 14-01-0001-380, April 30, 1966.

in C-301 to 470°F. The gas is then split into two streams, so that this high level heat can be used for heating two separate process streams. One portion of the gas passes through C-302 which provides the entire preheat (85°F to 400°F) for waste heat boilers C-204 and C-205. The other portion passes first through C-303 which provides high temperature preheat (275°F to 400°F) for the boiler feed water for C-301 and for an additional amount which is used by the Section 500 waste heat boilers. Next the gas passes through C-304 which provides low level preheat (125° to 275°F) for all Section 200 and 500 waste heat boilers, except C-204 and C-205. The two gas streams are recombined before F-302 and condensate is separated at 210°F and sent to Section 600 for use as filter wash water. Finally, the gas is cooled to 100°F in C-305 and scrubbed with water in E-301 to remove trace amounts of ammonia which might be present.

Section 300 is designed as five parallel operating units.

d. Section 400 - Gas Purification

This section is the same as described in Progress Report No. 21.

e. Section 500 - Methane Synthesis

This section is the same as described in Progress Report No. 21.

f. Section 1100 - Offsites

Section 1100 (not shown in Figure 2) includes facilities for the same functions as listed and described in Progress Report No. 21.

Steam generation facilities consist solely of a start-up boiler capable of producing 550,000 pounds per hour of 420 psia steam at 1000°F and 155,000 pounds per hour of 420 psia saturated steam.

Electric power is produced at a rate of 115,000 kw at 13,800 volts by turbogenerators using condensing steam turbine drives. An electric substation is provided to reduce the voltage to 4160, 440, and 110 volts.

Facilities are also provided for circulating 306,000 gpm of cooling water, for treating 2720 gpm of boiler feedwater, and for deaerating 6450 gpm of boiler feedwater.

A complete utilities summary is included as Table IV.

2. Economics

The cost of producing 247,000,000 SCFD of pipeline gas from subbituminous coal is calculated in Tables V and VI, assuming 90 percent stream efficiency. The procedure used is in accordance with OCR's tentative standard for cost estimating of pipeline gas plants.(2)

Estimated capital investment is summarized in Table V. Shift catalyst and activated carbon are included in fixed investment because of their long lifetimes. Total capital investment is about $147,000,000.

Estimated operating expenses and gas selling price are shown in Table VI. Subbituminous coal is charged at $2 per ton. Total operating expense is calculated to be 34.4¢/MSCF, and gas selling price, based on a 20-year average return on equity capital of about 9.4 percent, is 43.5¢/MSCF.

It should be noted here that a credit of 8.9¢/MSCF is taken in the economics for the excess power. The rationale of this credit is the same as in Progress Report No. 21. The credit is taken for the excess power at 4.5 mills/kwh, the cost of producing it by burning subbituminous coal at $2 per ton.

3. Variation of Manufacturing Cost with Melt Height

Since increasing the melt bed depth has been shown to decrease the observed experimental gasification rates, a study has been made to determine the economic effects of reducing the bed depth from the 19 feet used in the flowsheet studies to 8 feet and to 4 feet. A new gasifier design (as yet unspecified) would be required for bed depths much below 4 feet, and therefore no estimate was made of the effect of reducing heights below this level.

If melt depth is reduced, the number of gasifiers must be correspondingly increased to keep the total melt volume constant. This, of course, results in an increased plant investment due primarily to the larger number of gasification units. This increase in number of operating trains, in turn, requires a larger amount of operating labor which further raises gas cost. The effects of reducing bed depth to 8 feet and 4 feet are increases in gas manufacturing cost of 3.9 and 10.2¢/MSCF, respectively. Thus, it appears that bed depth may be reduced to about 8 feet without seriously increasing gas cost, but if heights much below this are required to obtain the desired rates, the economics will begin to be seriously affected.

4. Hydrogen from Bituminous Coal

Estimated costs for producing hydrogen from bituminous coal, based on very preliminary process designs and cost estimates, were presented in a previous report.(3)

(2) OCR Tentative Standard for Cost Estimating of Investor-Owned Plants for Producing Pipeline Gas from Coal, June 4, 1965.

(3) Progress Report No. 6, Contract 14-01-0001-380, January 31, 1965, p. 9A.

Studies have been resumed on hydrogen in order to prepare a "process package" ... the ones prepared for producing pipeline gas.

The hydrogen content of the product gas will be 95%[4]. However, if the raw synthesis gas leaving the gasifier (assumed to have the same composition as for the case of pipeline gas from bituminous coal) is processed by shift conversion of CO followed by CO_2 removal without any methane removal, the product gas will contain about 10% methane. Thus, the methane content of the raw gas must be reduced to a more acceptable level. This could be done utilizing either partial oxidation with oxygen or catalytic steam reforming. Previous studies[3] have indicated the latter to be the more attractive of the two and so will be accepted for use in the present design. The use of steam reforming, however, requires that the feed gas to such a unit be sulfur free necessitating a sulfur removal step before reforming.

At the present stage, it appears that the following processing sequence will be used for the manufacture of hydrogen:

a. Gasification and Ash Removal
b. High Temperature Shift Conversion
c. CO_2 and Bulk Sulfur Removal
d. Complete Sulfur Removal
e. Steam Reforming of Methane
f. Low Temperature Shift Conversion
g. CO_2 Removal
h. Methanation of Carbon Oxides
i. Gas Compression

C. Projections

1. Pipeline Gas

Investment and operating costs will be determined for the manufacture of pipeline gas from anthracite coal and lignite.

2. Hydrogen

Preparation of the "process package" for the hydrogen-from-coal plant will continue.

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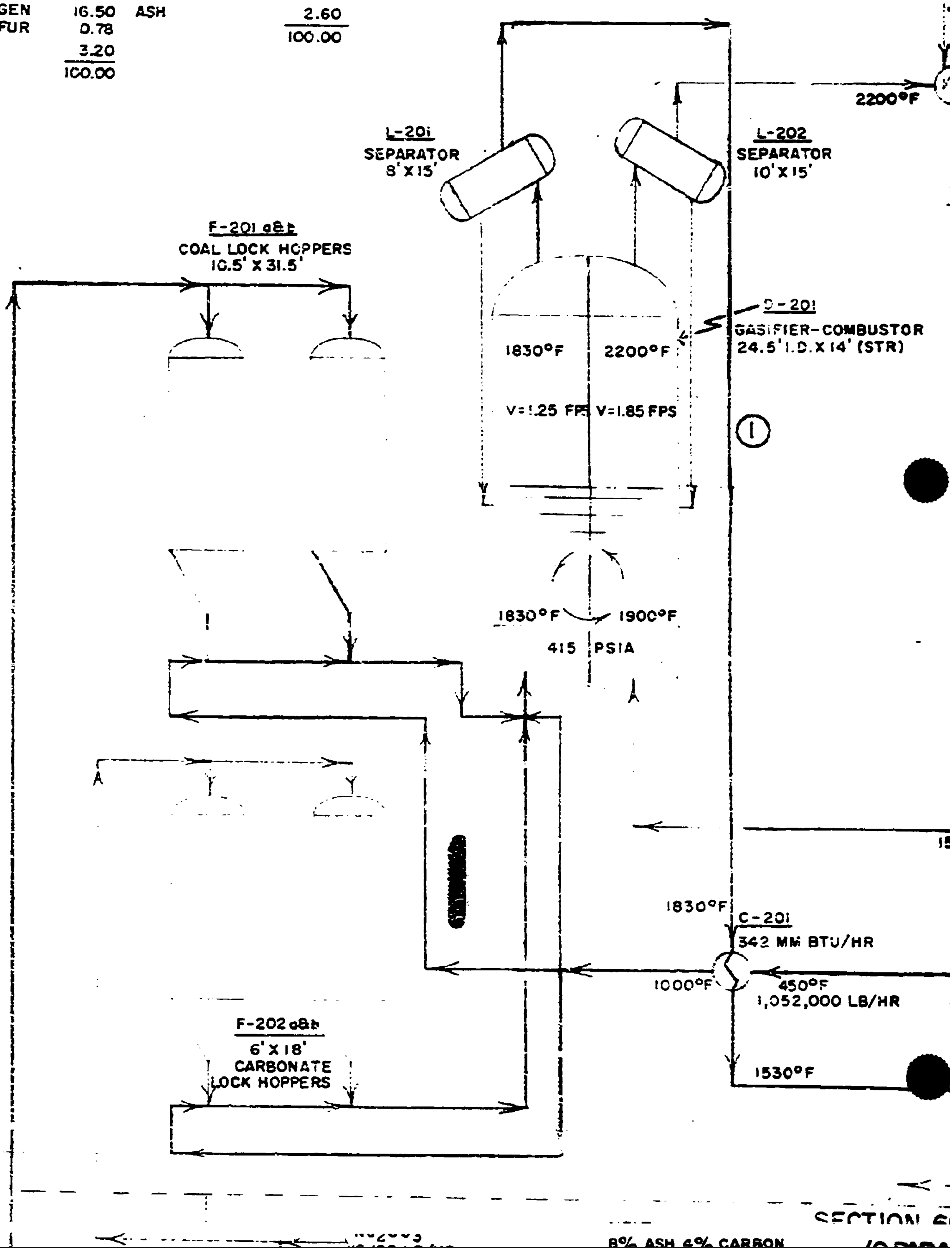
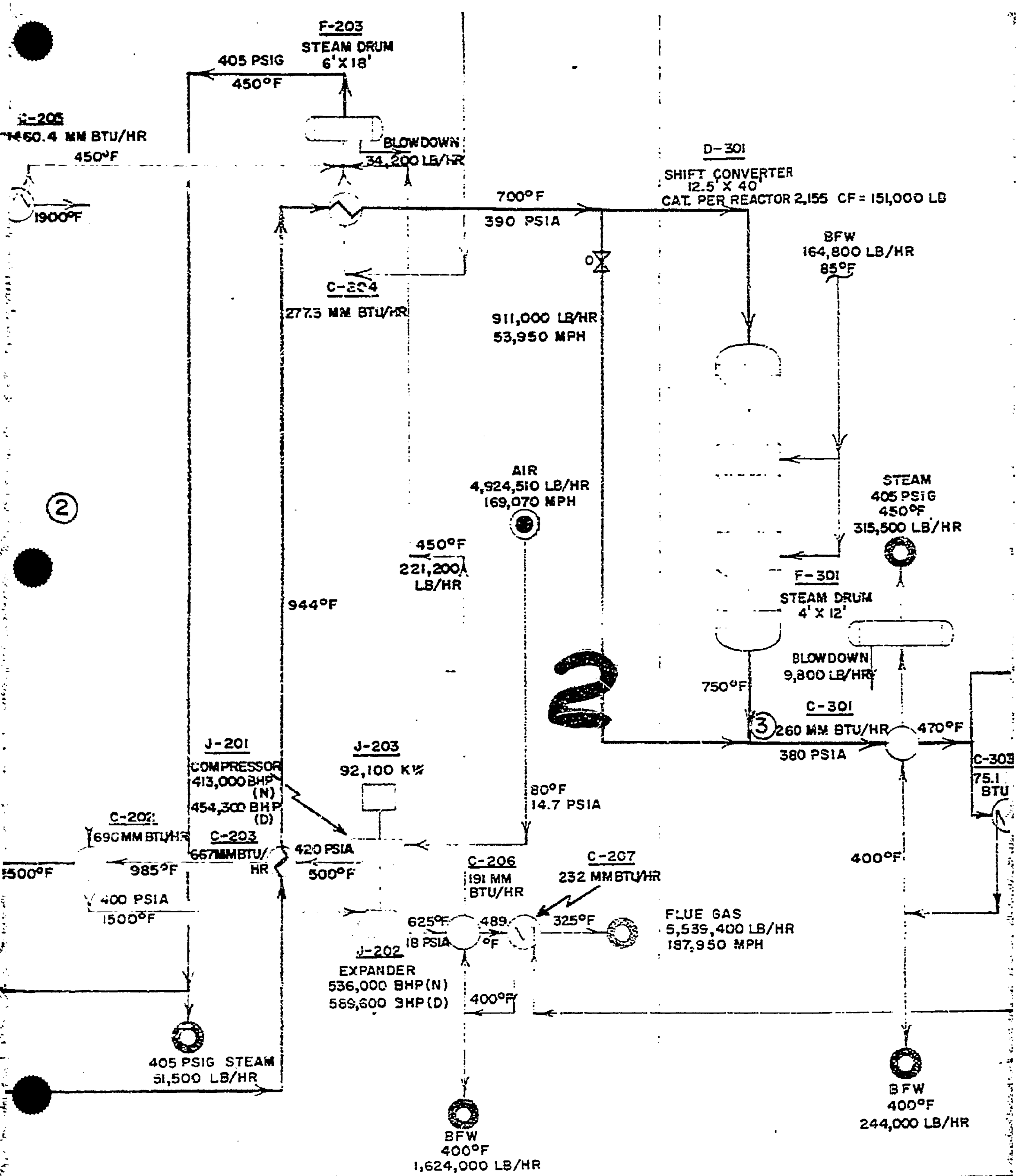
(3) Progress Report No. 6, Contract 14-01-0001-380, January 31, 1965,

(4) Progress Report No. 21, Contract 14-01-0001-380, April 30, 1966, p. 26.

ULTIMATE (DRY BASIS)	%
CARBON	73.50
HYDROGEN	4.52
NITROGEN	1.50
OXYGEN	16.50
SULFUR	0.78
ASH	3.20
	100.00

PROXIMATE	%
MOISTURE	18.79
VOLATILE MATTER	39.21
FIXED CARBON	39.40
ASH	2.60
	100.00





00: ASH REMOVAL
SECTION 500: METHANATION AND PRODUCT GAS COMPRESSION

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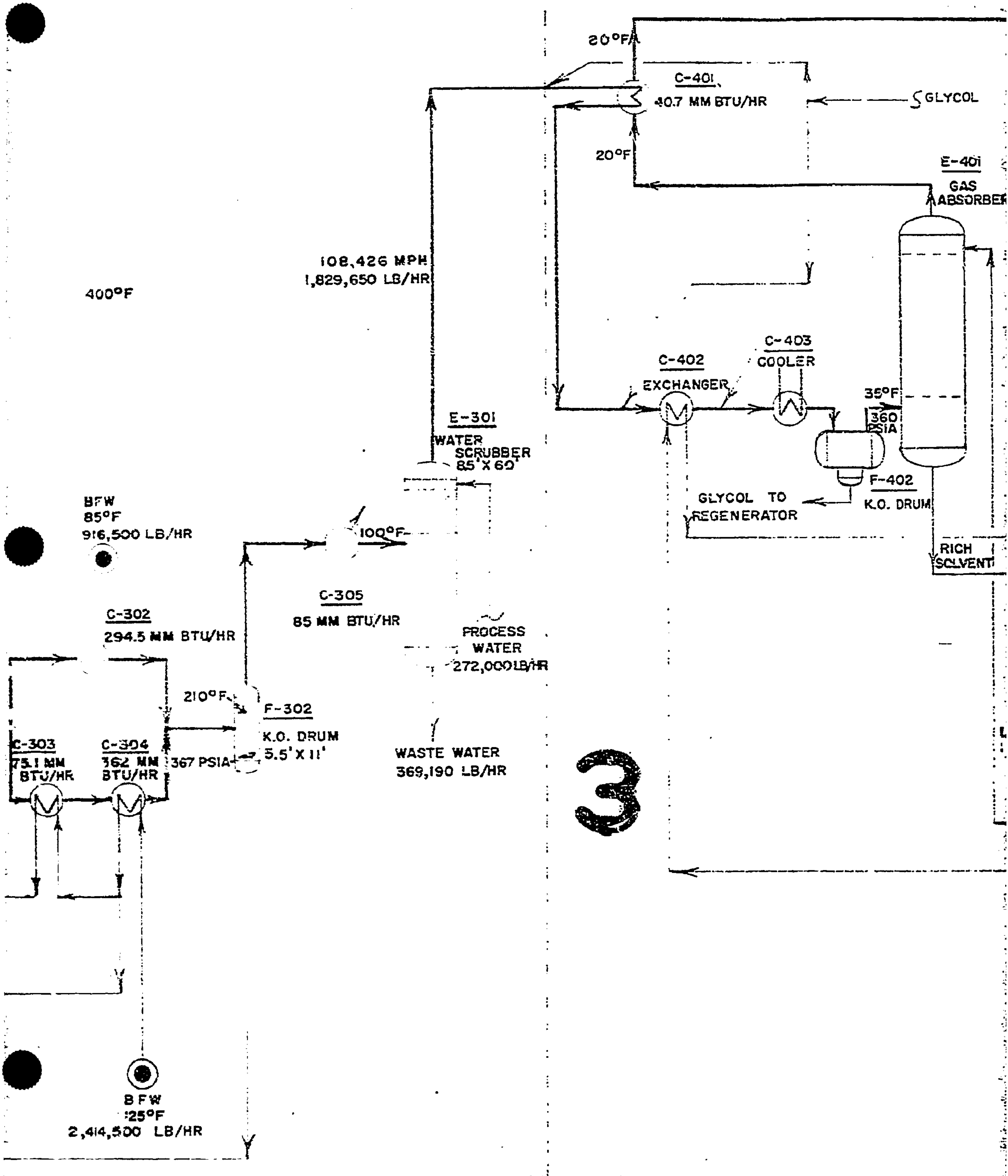

ON (5 PARALLEL TRAINS)

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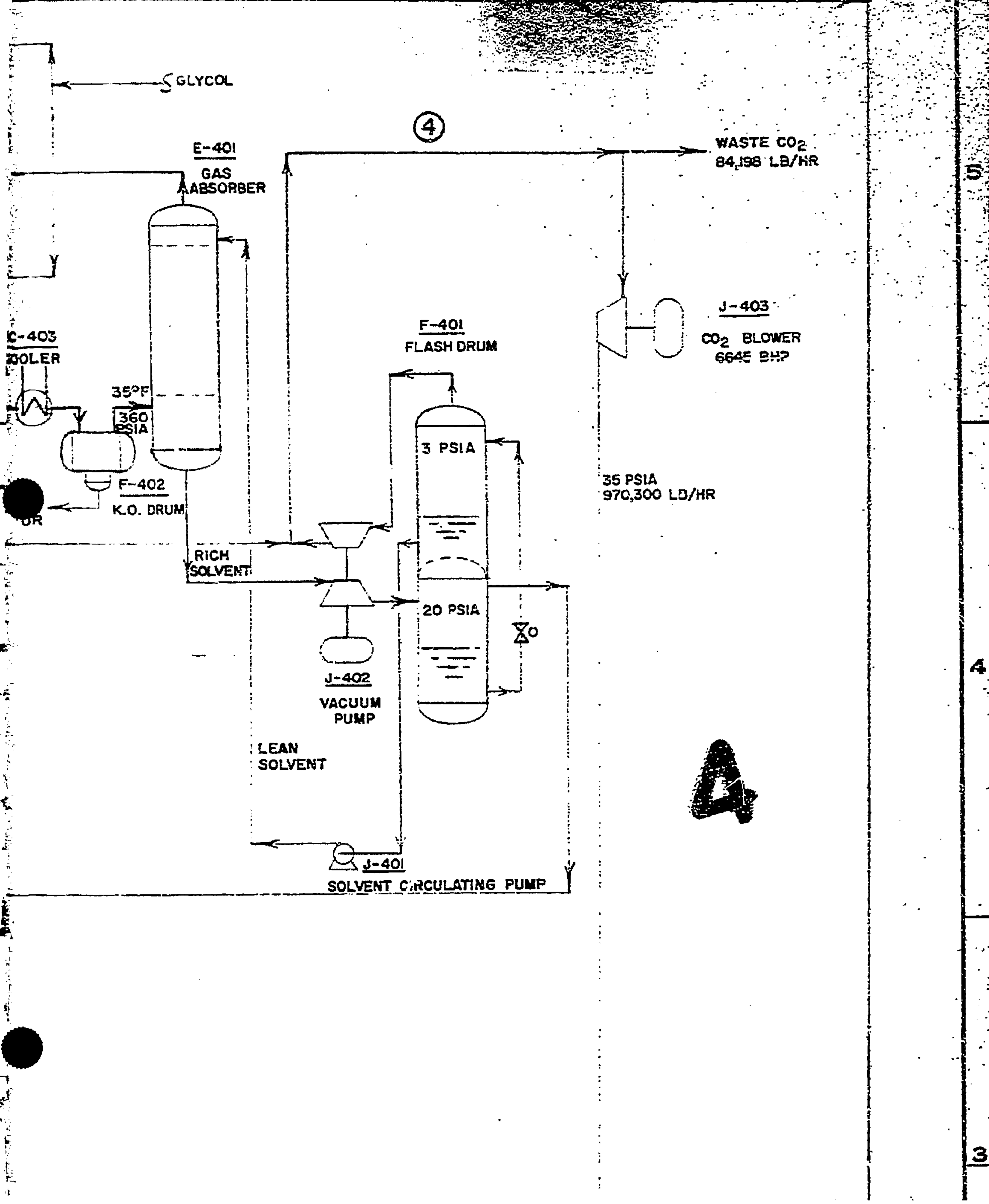
…TION 400: GAS PURIFICATION



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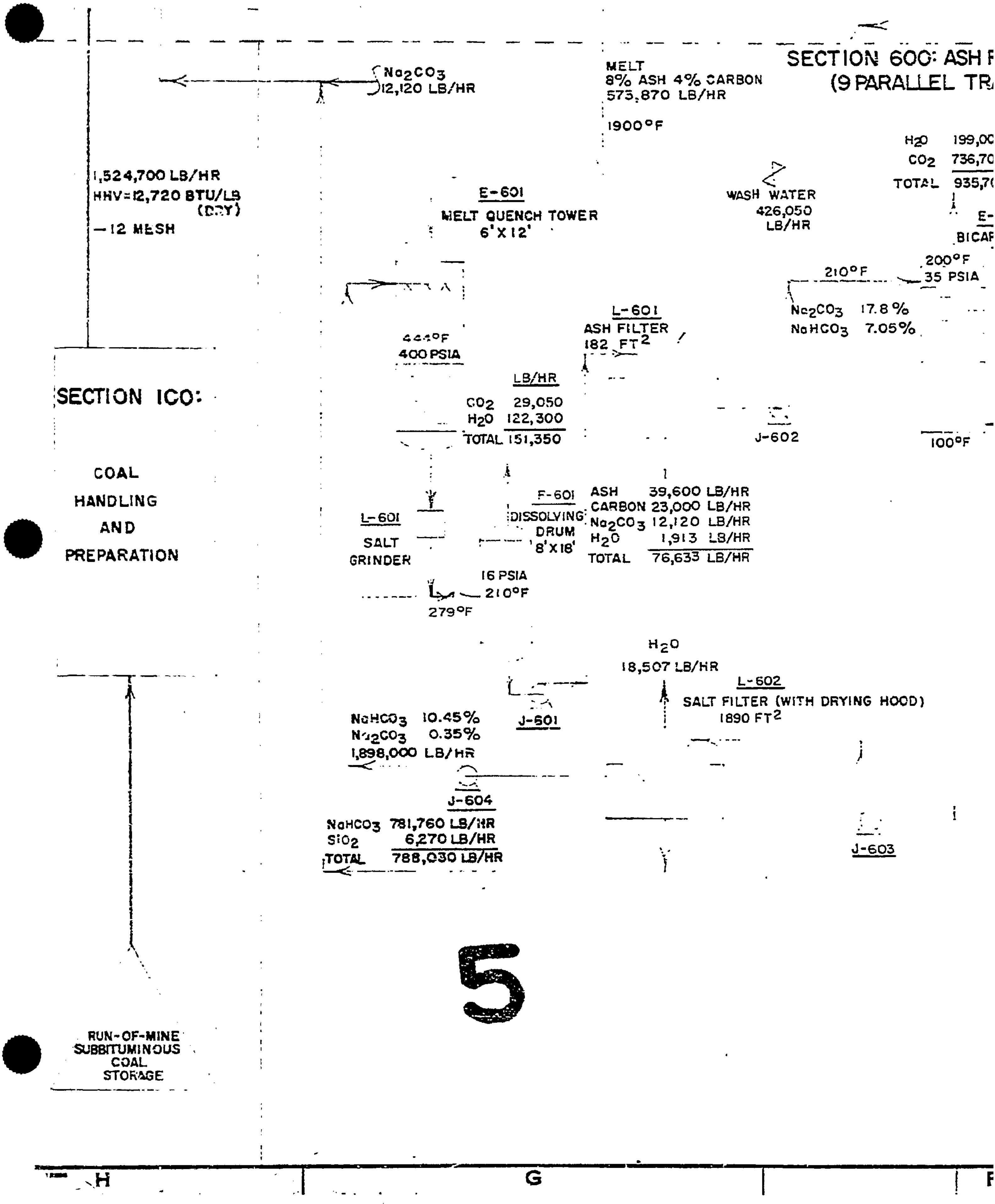
SECTION 600: ASH F
(9 PARALLEL TR
Na_2CO_3
12,120 LB/HR
MELT
8% ASH 4% CARBON
573,870 LB/HR
1900°F
H_2O 199,00
CO_2 736,70
TOTAL 935,7
1,524,700 LB/HR
HHV=12,720 BTU/LB (DRY)
—12 MESH
E-601
MELT QUENCH TOWER
6'X12'
WASH WATER
426,050 LB/HR
E-
BICAR
200°F
35 PSIA
210°F
Na_2CO_3 17.8%
$NaHCO_3$ 7.05%
L-601
ASH FILTER
182 FT2
444°F
400 PSIA
SECTION 100:
LB/HR
CO_2 29,050
H_2O 122,300
TOTAL 151,350
J-602
100°F
COAL
HANDLING
AND
PREPARATION
F-601
DISSOLVING DRUM
8'X18'
ASH 39,600 LB/HR
CARBON 23,000 LB/HR
Na_2CO_3 12,120 LB/HR
H_2O 1,913 LB/HR
TOTAL 76,633 LB/HR
L-601
SALT
GRINDER
16 PSIA
210°F
279°F
H_2O
18,507 LB/HR
L-602
SALT FILTER (WITH DRYING HOOD)
1890 FT2
J-601
$NaHCO_3$ 10.45%
Na_2CO_3 0.35%
1,898,000 LB/HR
J-604
$NaHCO_3$ 781,760 LB/HR
SiO_2 6,270 LB/HR
TOTAL 788,030 LB/HR
J-603
RUN-OF-MINE
SUBBITUMINOUS
COAL
STORAGE
H
G
F

5

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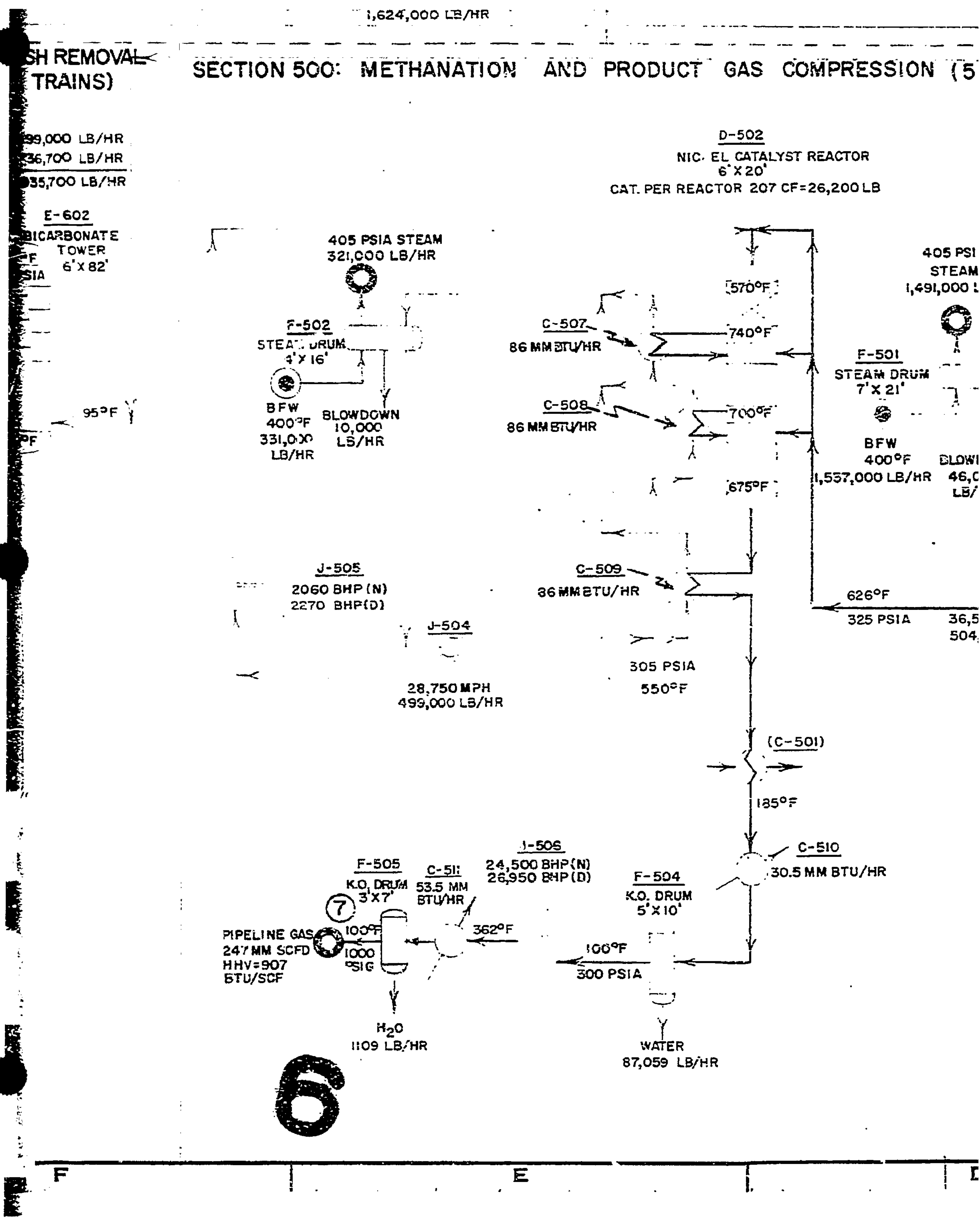
SECTION 500: METHANATION AND PRODUCT GAS COMPRESSION (5



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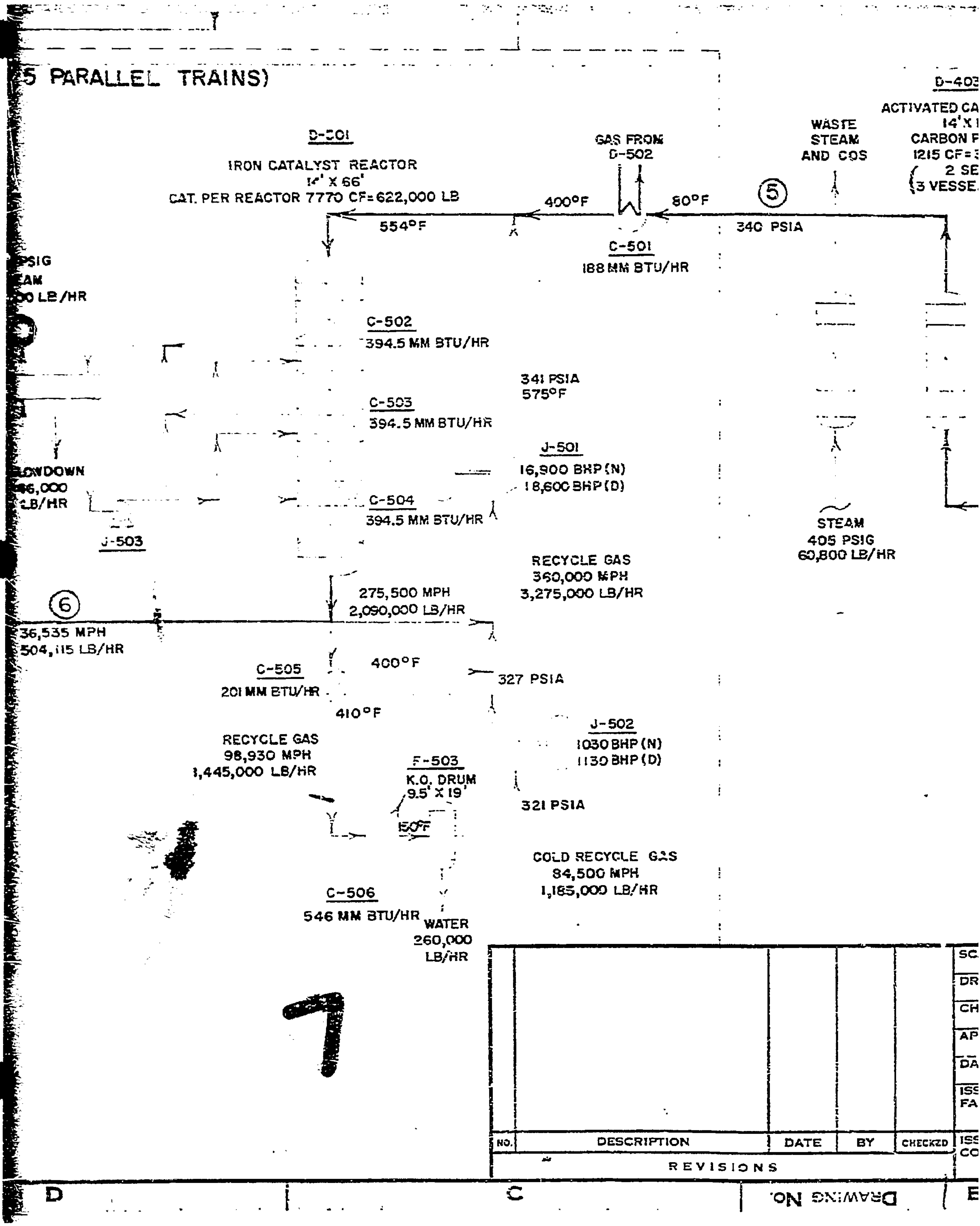

(5 PARALLEL TRAINS)
D-501
IRON CATALYST REACTOR
14' X 66'
CAT. PER REACTOR 7770 CF=622,000 LB
GAS FROM
D-502
400°F
80°F
554°F
C-501
188 MM BTU/HR
340 PSIA
WASTE
STEAM
AND COS
D-403
ACTIVATED CA
CARBON F
1215 CF=
2 SE
3 VESSE
PSIG
LB/HR
C-502
394.5 MM BTU/HR
341 PSIA
575°F
C-503
394.5 MM BTU/HR
J-501
16,900 BHP (N)
18,600 BHP (D)
BLOWDOWN
LB/HR
C-504
394.5 MM BTU/HR
J-503
STEAM
405 PSIG
60,800 LB/HR
RECYCLE GAS
360,000 MPH
3,275,000 LB/HR
275,500 MPH
2,090,000 LB/HR
36,535 MPH
504,115 LB/HR
C-505
201 MM BTU/HR
400°F
410°F
327 PSIA
J-502
1030 BHP (N)
1130 BHP (D)
RECYCLE GAS
98,930 MPH
1,445,000 LB/HR
F-503
K.O. DRUM
9.5' X 19'
150°F
321 PSIA
COLD RECYCLE GAS
84,500 MPH
1,185,000 LB/HR
C-506
546 MM BTU/HR
WATER
260,000
LB/HR
NO.
DESCRIPTION
DATE
BY
CHECKED
REVISIONS
D
C
E
DRAWING NO.

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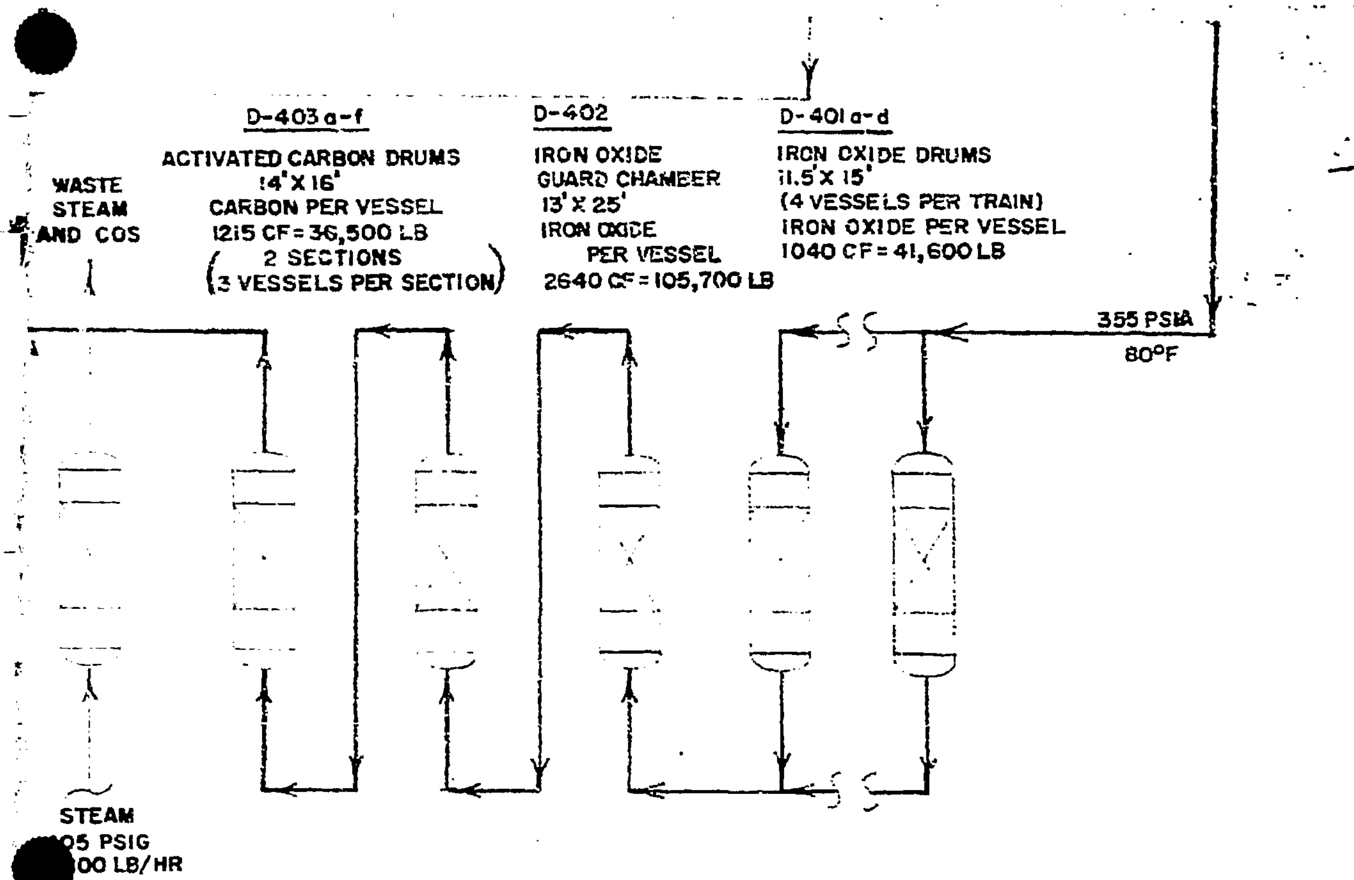

NOTE: ALL QUANTITIES ARE TOTALS FOR ALL PARALLEL TRAINS UNLESS OTHERWISE NOTED.

THE M. W. KELLOGG CO.

PIPELINE GAS FROM SUBBITUMINOUS COAL

247 MM SCFD

OFFICE OF COAL RESEARCH

CONTRACT NO. 14-01-0001-380

8

SCALE:	NONE
DRAWN:	EAP
CHECKED:	WHH
APPROVED:	3-25-66
DATED:	
ISSUED FOR FABRICATION	
ISSUED FOR CONSTRUCTION	

DATE	BY	CHECKED

CLASS & ITEM	AREA	JOB NUMBER	DRAWING NUMBER
		L.O. 6026	6026-3-4

DRAWING B 3 A

TABLE 11

PROCESS STREAM COMPOSITIONS

247 MM SCFD PIPELINE GAS FROM SUBBITUMINOUS COAL

Stream No.	①			②			③		
Temperature, °F	1830			1900			733		
Pressure, psia	405			405			380		
Flow Rate	Lb/Hr	Mols/Hr	Mol %	Lb/Hr	Mols/Hr	Mol %	Lb/Hr	Mols/Hr	Mol %
CO	878,000	31,420	24.24	-----	------	-----	538,000	19,210	13.86
CO_2	540,910	12,480	9.63	1,329,080	30,150	16.02	1,075,150	24,450	17.64
CH_4	72,720	4,545	3.51	-----	------	-----	72,720	4,545	3.28
H_2	96,610	48,300	37.25	-----	------	-----	120,870	60,435	43.62
H_2O	582,000	32,300	24.92	392,450	21,822	11.62	528,300	29,350	21.18
N_2	11,200	400	0.31	3,744,400	133,800	71.21	11,200	400	0.29
H_2S	5,560	163	0.13	-----	------	-----	5,560	163	0.12
COS	1,090	18	0.01	-----	------	-----	1,090	18	0.01
O_2	-------	------	-----	65,790	2,058	1.09	------	-----	-----
SO_2	-------	------	-----	7,680	120	0.06	------	-----	-----
C_2^+	-------	------	-----	-----	------	-----	------	-----	-----
Total	2,188,090	129,626	100.00	5,539,400	187,950	100.00	2,352,890	138,571	100.00

Stream No.	[illegible]			⑤		
Temperature, °F	[illegible]			[illegible]0		
Pressure, psia	[illegible]			[illegible]		
Flow Rate	Lb/Hr	Mols/Hr	Mol %	Lb/Hr	Mols/Hr	Mol %
CO	5,545	198	0.81	532,455	19,012	22.35
CO_2	1,037,740	23,600	96.74	37,410	850	1.00
CH_4	192	12	0.05	72,528	4,533	5.32
H_2	348	174	0.71	120,522	60,261	70.81
H_2O	4,305	239	0.98	755	42	0.05
N_2	-----	----	----	11,200	400	0.47
H_2S	5,287	156	0.64	-----	-----	-----
COS	1,081	18	0.07	-----	-----	-----
O_2	-----	----	----	-----	-----	-----
SO_2	-----	----	----	-----	-----	-----
C_2^+	-----	----	----	-----	-----	-----
Total	1,054,498	24,397	100.[illegible]	774,670	85,098	100.00

Stream No.	⑥			⑦		
Temperature, °F	626			100		
Pressure, psia	325			1000		
Flow Rate	Lb/Hr	Mols/Hr	Mol %	Lb/Hr	Mols/Hr	Mol %
CO	7,455	282	0.77	770	27.5	0.10
CO_2	38,110	1,094	3.00	31,250	711	2.62
CH_4	244,920	15,313	41.90	378,300	23,660	87.02
H_2	18,422	9,211	25.20	4,723	2,361.5	8.70
H_2O	133,500	7,410	20.30	459	25.4	0.09
N_2	11,200	400	1.10	11,200	400	1.47
H_2S	-----	-----	-----	-----	-----	-----
COS	-----	-----	-----	-----	-----	-----
O_2	-----	-----	-----	-----	-----	-----
SO_2	-----	-----	-----	-----	-----	-----
C_2^+	50,500	2,825	7.74	-----	-----	-----
Total	504,115	36,535	100.00	426,702	27,185.4	100.00

TABLE III

Section-by-Section Material Balance

247 MM SCFD Pipeline Gas From Subbituminous Coal

Section 200

Input			Pounds per Hour
Coal from Section 100			1,524,700
Air			4,924,510
Steam			1,052,000
Salt from Section 600			
$NaHCO_3$	781,760		
Na_2CO_3	12,120		
SiO_2	6,270		
			800,150
		Total	8,301,360
Output			**Pounds per Hour**
Gas to Section 300			2,188,090
Melt to Section 600			573,870
Flue Gas			5,539,400
		Total	8,301,360

Section 300

Input		Pounds per Hour
Gas from Section 200		2,188,090
Boiler Feed Water to Shift Convertor D-301		164,800
Process Water to Scrubber E-301		272,000
	Total	2,624,890

Output		Pounds per Hour
Gas to Section 400		1,829,650
Condensate from F-302 to L-601		426,050
Waste Water from Scrubber E-301		369,190
	Total	2,624,890

Section 400

Input		Pounds per Hour
Gas from Section 300		1,829,650
Regeneration Steam to Activated Carbon Drums, D-503		60,800
	Total	1,890,450

Output		Pounds per Hour
Gas to Section 500		774,870
CO_2 from F-401 and F-402 to E-602	970,300	
to Stack	84,198	
		1,054,498
H_2S reacted in D-401		273
Steam & COS from Activated Carbon Drums, D-503		60,809
	Total	1,890,450

Section 500

Input	Pounds per Hour
Gas from Section 400	774,870

Output		Pounds per Hour
Pipeline Gas Product		426,702
Condensate from F-503		260,000
Condensate from F-504		87,059
Condensate from F-505		1,109
	Total	774,870

Section 600

Input			Pounds per Hour
Melt from D-201			573,870
Wash water from F-302			426,050
CO_2 from Section 400			970,300
		Total	1,970,220

Output			Pounds per Hour
Salt to F-202			
$NaHCO_3$	781,760		
SiO_2	6,270		788,030
Ash from L-601			39,600
Waste CO_2			
From F-601	29,050		
From E-602	736,700		765,750
Steam			
From F-601	122,300		
From E-602	199,000		
From L-601	1,913		
From L-602	18,507		341,720
Na_2CO_3 with residue from L-601			12,120
Carbon with residue from L-601			23,000
		Total	1,970,220

TABLE IV

Utilities Summary

247 MM SCFD Pipeline Gas From Subbituminous Coal

Steam (420 psia, 450°F)

A. Generation

Section	Item		Normal Production Pounds per Hour
200	C-204	Flue Gas WHB	555,000
	C-205	Gasifier Effluent WHB	334,000
	C-206	Expander Exhaust WHB	214,500
300	C-301	Shift WHB	315,500
500	C-502	Iron Cat. Reactor Cooling Coils	497,000
	C-503	Iron Cat. Reactor Cooling Coils	497,000
	C-504	Iron Cat. Reactor Cooling Coils	497,000
	C-507	Nickel Cat. Reactor Cooling Coils	107,000
	C-508	Nickel Cat. Reactor Cooling Coils	107,000
	C-509	Nickel Cat. Reactor Cooling Coils	107,000
		Total	3,231,000

B. Consumption

Section	Item		Normal Consumption Pounds Per Hour
200	D-201	Gasification	1,052,000
400	J-401	Solvent Pump	198,250
	J-402	CO_2 Blower	61,100
	D-403	Activated Carbon Regeneration	60,800
	Other Section 400		29,000
500	J-501	Iron Cat. Reactor Hot Recycle Gas Compressor	155,500
	J-502	Iron Cat. Reactor Cold Recycle Gas Compressor	9,500
	J-505	Nickel Cat. Reactor Recycle Gas Compressor	19,000
	J-511	Product Gas Compressor	226,000
1100	N-1101	Turbogenerator	1,419,850
		Total	3,231,000

Electric Power

A. Generation

Section	Item		Normal Generation KW
200	N-201	Expander-driven generator	92,100
1100	N-1101	Turbogenerator	115,200
		Total	207,300

B. Consumption

Section	Item		Normal Consumption HP	Normal Consumption KW
100	L-103	Bradford Breaker	229	171
	L-104	Hammermill	3,435	2,565
	Other Equipment		100	75
500	Miscellaneous Pumps		80	60
600	J-604	Quench Tower Pumps	1,305	975
	J-602	Bicarbonate Tower Pumps	360	269
	Other Equipment		70	52
		Total	5,579	4,167

Cooling Water

A. Production

Section	Item		GPM
1100	L-1101	Cooling Towers	305,260

B. Consumption

Section	Item		Temp. Rise °F	GPM
200	C-208	Intercooler	15	61,100
	C-209	Intercooler	15	51,600
		Total Section 200		112,700
300	C-306	Cooler	15	11,130
		Total Section 300		11,310
400	C-404	Aftercooler	15	140
	J-401	Surface Condenser	30	11,560
		Total Section 400		11,700
500	C-506	Cooler	15	72,700
	C-510	Cooler	15	4,055
	C-511	Aftercooler	15	7,110
	J-501	Surface Condenser	30	4,010
	J-502	" "	30	175
	J-505	" "	30	350
	J-506	" "	30	4,160
		Total Section 500		92,550
1100	N-1101	Surface Condenser	30	77,000

Cooling Water Consumption Summary

Section	Title	GPM
200	Gasification	112,[illegible]
300	Shift Conversion	11,3[illegible]
400	Gas Purification	11,700
500	Methane Synthesis	92,[illegible]
1100	Offsite Facilities	77,000
	Total Consumption	305,260

C. Cooling Water Balance

	GPM
Recirculated Water	298,050
Makeup Water from J-1101	28,850
Total Water to Cooling Towers	326,900
Water losses in Cooling Tower	21,660
Total Water to Process	305,260
Warm water returned to river	7,210
Recirculated water	298,050

Boiler Feed Water

I. 85-125°F

A. Production

Section	Item	Normal Production Pounds per Hour
1100	Feedwater Makeup and Condensate Return	3,495,800

B. Consumption

Section	Item		Normal Consumption Pounds per Hour
300	D-301	Shift Converter	164,800
	C-302	BFW Heater	916,500
	C-304	BFW Heater	2,414,500
		Total	3,495,800

II. 275°F, 430 psia

A. Production

Section	Item		Normal Production Pounds per Hour
300	C-304	BFW Heater	2,414,500

B. Consumption

Section	Item		Normal Consumption Pounds per Hour
200	C-207	BFW Heater	1,845,200
300	C-303	BFW Heater	569,300
		Total	2,414,500

III. 400°F. 420 psia

A. Production

Section	Item		Normal Production Pounds per Hour
200	C-207	BFW Heater	1,8[illegible],000
300	C-302	BFW Heater	91[illegible],000
	C-303	BFW Heater	5[illegible],000
		Total	3,331,000

B. Consumption

Section	Item		Normal Consumption Pounds per Hour
200	F-203	Steam Drum	1,137,700
300	F-301	Steam Drum	325,300
500	F-501	Steam Drum	1,537,000
	F-502	Steam Drum	331,000
		Total	3,331,000

IV. Boiler Feed Water Balance

A. Losses from System

Section	Item		Pounds per Hour
200	D-201	Gasifier Steam	1,052,000
	F-203	Steam Drum Blowdown	34,200
300	D-301	Shift Converter Quench	164,800
	F-301	Steam Drum Blowdown	9,800
400	D-403	Activated Carbon Regeneration	60,800
		Other Steam	29,000
500	F-501	Steam Drum Blowdown	46,000
	F-502	Steam Drum Blowdown	10,000
		Total	1,406,600

B. Recirculation

Feed Water Makeup	1,406,600
Condensate from Surface Condensers	2,089,200
Total Boiler Feed Water	3,495,800

Process Water

Consumption

Section	Item		Normal Consumption Pounds per Hour
300	E-301	Scrubber	272,000

TABLE V

INVESTMENT SUMMARY

PIPELINE GAS FROM SUBBITUMINOUS COAL

Basis: 247,000,000 SCFD of Pipeline Gas
90% Stream Efficiency

SECTION	TITLE	BARE COST *
100	Coal Storage and Preparation	$ 5,017,000
200	Gasification	60,135,000
300	Shift Conversion	3,714,000
400	Gas Purification	19,631,000
500	Methane Synthesis	9,604,000
600	Ash Removal	3,165,000
1100	Offsite Facilities	16,931,000
	Total Bare Cost	$118,197,000
	Interest during construction and contractors' overhead and profit	23,639,000
	TOTAL FIXED INVESTMENT	$141,836,000

Working Capital

30 Days Coal Inventory	$ 1,090,000
30 Days Carbonate Inventory	135,000
30 Days Catalyst Inventory	79,000
Catalyst Charge	430,000
Accounts Receivable at 11% of Total Operating Expense	3,091,000
Total Working Capital	$ 4,825,000

TOTAL CAPITAL INVESTMENT	$146,661,000

*Bare cost includes materials, freight, construction labor, field administration and supervision, insurance during construction, cost of tools, field office expense and cost of home office engineering and procurement.

TABLE VI

PIPELINE GAS FROM SUBBITUMINOUS COAL

ESTIMATED ANNUAL OPERATING COST

AND GAS SELLING PRICE

Basis: 247,000,000 SCFD of Pipeline Gas
90% Stream Efficiency

ITEM	$/YEAR	¢/MSCF
Subbituminous Coal at $2/Ton	$12,080,000	14.8
Sodium Carbonate Makeup at 1.55¢/lb.	1,490,000	1.8
Miscellaneous Chemicals	174,000	0.2
Sponge Iron Makeup	37,000	0.05
Methanation Catalyst Makeup	657,000	0.8
Direct Operating Labor at $3.20/man-hr.	1,648,000	2.0
Power Credit at 4.5 mills/kwh	-7,240,000	-8.9
Maintenance at 3% of bare cost	3,540,000	4.3
Supplies at 15% of maintenance	531,000	0.7
Supervision at 10% of operating labor	165,000	0.2
Payroll Overhead at 10% of operating labor and supervision	181,000	0.2
General Overhead at 50% of maintenance + supplies + operating labor + supervision	2,942,000	3.6
Plant Operating Expenses	$16,205,000	19.8
Depreciation at 5% of fixed investment	7,092,000	8.7
Local Taxes & insurance at 3% of fixed investment	4,255,000	5.2
Subtotal	27,552,000	33.7
Contingencies at 2%	551,000	0.7
TOTAL OPERATING EXPENSE	$28,103,000	34.4

20 YEAR AVERAGE REVENUE REQUIREMENT $35,542,000

AVERAGE GAS SELLING PRICE 43.5¢/MSCF

IV. MECHANICAL DEVELOPMENT

A. Accomplishments

1. Environmental Testing of High Temperature Materials

Gasification Corrosion Test #9, using a simulated gasification condition, was terminated at 407 hours during this report period. This test was conducted to investigate what the corrosion problems might be at the joint between two blocks of Monofrax A. Two blocks of this material, each a cube one inch on a side, were subjected to a compressive loading of 40 psi while at 1840°F and in a simulated gasification situation. (The compositions of the melt and gas supplied are given in Table VII). At intervals during the test, the samples were removed and examined. Each examination found the blocks in good condition, hence, they were returned to the test. However, at 407 hours into the test, the specimen support failed, and upon examination both specimens were found to have cracked in a direction parallel to the applied stress; there was still little evidence of corrosion. There is a possibility that the failure of the support also caused the specimens to fail, inasmuch as both failures seem to have occurred at about the same time.

2. Mechanical Characteristics Testing

The melt quench tests have been continued during this report period. Melt material at 1840°F has been quenched in cold, still water with little or no agglomeration. The melt was gravity-fed through holes up to 3/4" in diameter and allowed to fall into some nine inches of water. The resulting material closely resembled popcorn in general size and shape; some large clinkers were observed when the depth of water was reduced to less than about four inches.

A circulation system for studying the flow parameters of a simulated melt material is currently nearing completion. This test system is designed to allow measurement of flows attainable with an "air-lift" of the type proposed for the molten-salt system.

The 5-3/4 inch reactor test facility is nearing completion. The furnace was received recently, and was erected on the previously fabricated support frame; the reactor itself is now also in place. Final wiring and controls are all that is required prior to initial testing of this facility.

B. Projections

1. Environmental Testing of High Temperature Materials

The simulated Gasification Corrosion Test will be restarted as soon as new specimens can be prepared and the facilities made ready.

2. Mechanical Characteristics Testing

Work on the circulation test facility will continue, as will theoretical work to predict flow of the melt material under various conditions. The test facility when complete will be used to check the validity of the theoretical predictions.

Installation of the controls and accessory items on the 5-3/4 inch test facility will continue, and should be complete about mid-July.

TABLE VII

MELT AND FEED GAS COMPOSITION

FOR SIMULATED GASIFICATION CORROSION TESTING

Feed Gas Composition	Mol %
H_2	16
CO	5
CO_2	5
N_2	44
H_2O	30
	100%

Initial Melt Composition	Wt. %
Na_2CO_3	87
Ash	10
Na_2S	3
	100%