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U.S. Department of Energy
Federal Energy Technology Center

Refining and End Use Study of Coal Liquids

Contract No. DE-AC22-93PC91029

Quarterly Report
October - December 1996

DISCLAIMER

Section 1

Introduction and Summary

This report is Bechtel's twelth quarterly technical progress report and covers the period of October 1, 1996 through December 31, 1996.

1.1 Introduction

Bechtel, with Southwest Research Institute, Amoco Oil R&D, and the M.W. Kellogg Co. as subcontractors, initiated a study on November 1, 1993, for the U.S. Department of Energy's (DOE's) Pittsburgh Energy Technology Center (PETC) to determine the most cost effective and suitable combination of existing petroleum refinery processes needed to make specification transportation fuels or blending stocks, from direct and indirect coal liquefaction product liquids. This 47-month study, with an approved budget of $4.4 million dollars, is being performed under DOE Contract Number DE-AC22-93PC91029.

A key objective is to determine the most desirable ways of integrating coal liquefaction liquids into existing petroleum refineries to produce transportation fuels meeting current and future, e.g. year 2000, Clean Air Act Amendment (CAAA) standards. An integral part of the above objectives is to test the fuels or blends produced and compare them with established ASTM fuels. The comparison will include engine tests to ascertain compliance of the fuels produced with CAAA and other applicable fuel quality and performance standards.

The final part of the project includes a detailed economic evaluation of the cost of processing the coal liquids to their optimum products. The cost analyses is for the incremental processing cost; in other words, the feed is priced at zero dollars. The study reflects costs for operations using state of the art refinery technology; no capital costs for building new refineries is considered. Some modifications to the existing refinery may be required. Economy of scale dictates the minimum amount of feedstock that should be processed.

To enhance management of the study, the work has been divided into two parts, the Basic Program and Option 1.

The objectives of the Basic Program are to:

- Characterize the coal liquids
- Develop an optimized refinery configuration for processing indirect and direct coal liquids
- Develop a LP refinery model with the Process Industry Modeling System (PIMS) software.

The work has been divided into six tasks.

Task 1 - Development of a detailed project management plan for the Basic Program
Task 2 - Characterization of four coal liquid feeds supplied by DOE
Task 3 - Optimization of refinery processing configurations by linear programming
Task 4 - Pilot plant analysis of critical refinery process units to determine yield, product quality and cost assumptions. Petroleum cuts, neat coal liquids, and coal liquids/petroleum blends will be processed through the following process units: reforming, naphtha and distillate hydrotreating, catalytic cracking and hydrocracking.

- Task 5 -Development of the project management plan for Option 1
- Task 6 - Project management of the Basic Program and Option 1

The objectives of Option 1 are to:

- Confirm the validity of the optimization work of the Basic Program
- Produce large quantities of liquid transportation fuel blending stocks
- Conduct engine emission tests
- Determine the value and the processing costs of the coal liquids

This will be done by processing the coal liquids as determined by the optimization work, blending and characterizing the product liquids, and running engine emission tests of the blends. Option 1 has been divided into three tasks.

- Task 1 -Based on the pilot plant and linear programming optimization work of the Basic Program, production runs of pilot plants (hydrotreating, reforming, catalytic cracking, and hydrocracking) will be conducted to produce sufficient quantities for blending and engine testing.
- Task 2 -The pilot plant products will be blended, characterized, and engine tested
- Task 3 -An economic analysis will be conducted to determine the costs of processing the coal liquids through the existing refinery

Table 1-1 shows which organization has the primary responsibility for each task.

1.2 Summary

The major efforts conducted during the fourth quarter of 1996 were in the areas of:

- Option 1 blending
- Option 1 FCC production run

Introduction and Summary

Table 1-1 Project Task Primary Responsibility Chart

Task	Description	Bechtel	SwRI	Amoco	Kellogg
1	Project Management Plan (PMP) development	x			
2	Feed characterization		x		
3	Linear programming	x			
4	Pilot plant analysis -				
	Cat cracking of DL liquids				x
	Cat cracking of indirect wax			x	
	Hydrocracking of wax			x	
	Fractionation, reforming, hydrotreating, etc.		x		
5	Option 1 PMP development	x			
6	Project management	x			
Option 1 - Task 1	Pilot plant production -				
	Cat cracking of DL liquids and wax				x
	All other production work		x		
Option 1 - Task 2	Fuel blending, characterizing, engine testing		x		
Option 1 - Task 3	Economic analysis	x			

- x = key participant

Section 2

SwRI Activities

There was no project activity for this reporting period.

Section 3

Bechtel Activities

There was no project activity for this reporting period.

Section 4

Amoco Activities

There was no project activity for this reporting period.

Section 5

M.W. Kellogg Activities

(Editor's note: Information on the Option 1 - DL2 FCC production run was partially reported in the July-September 1996 quarterly progress report. A full discussion of the run is provided in this report.)

5.0 OPTION I - DL2 FCC PRODUCTION RUN

5.1 SUMMARY

A program was carried out on the FCC1 Pilot Plant at the Kellogg Technology Development Center (KTDC) to produce a large quantity of FCC liquid products for further testing and characterization. A total of 80 gallons of feedstock was processed at the targeted run conditions to produce about 60 gallons of total liquid product during 145 hours of lined out operation.

Previously, a MAT program had been carried out to compare the DL2 (sub-bituminous) coal liquid (heavy distillate) with a hydrotreated product of the DL2 that was prepared by Southwest Research Institute (SwRI). The effect of cracking temperature on yields and quality was also studied by making runs at both 970°F and 1010°F. An Amoco vacuum gas oil (VGO) was used as the baseline FCC feedstock.

The Kellogg pilot plant was then run at the conditions established for heat balance with the VGO at a cracking temperature of about 985°F. The same conditions were maintained while performing two additional "short" scoping runs with the DL2 liquid and with a blend of the two feeds containing 33.3 volume percent DL2.

The results of these tests were reported in the January-March 1996 Technical Progress Report. These results were then employed to select the feed composition for the Option 1 production run in the FCC Pilot Plant. Approximately 43 volume percent of the feed was DL2 heavy distillate with the balance being the Amoco VGO.

The production run was completed over the course of two campaigns, one in August and the other in September. In each campaign, about half of the total feed was processed. Also in each campaign, a single two-hour period was saved separately and worked up for material balances and yields, including a breakdown of C_5 yields using PIANO analysis. A TBP distillation was performed and octane numbers obtained for one of the periods.

The remainder of total liquid product from the run was combined, debutanized, depentanized, and distilled in Kellogg's Batch Still. In addition to the IBP-120°F "C_5's" fraction produced during depentanization, cuts were made at 120-430°F, 430-650°F, and 650°F+. The light and heavy naphtha cuts were analyzed by PIANO and all samples were shipped to SwRI for further testing.

Section 5

M.W. Kellogg Activities

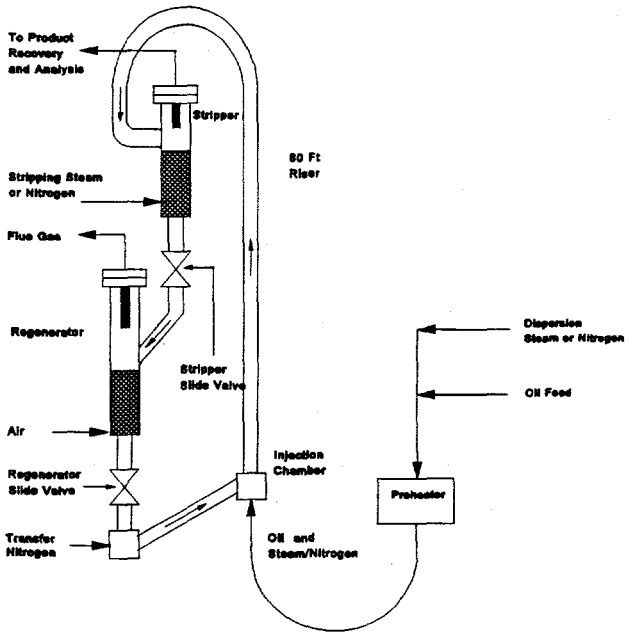
5.2 EXPERIMENTAL

5.2.1 FCC PILOT PLANT

General

The pilot plant used in this work (shown in Figure 5-1) is a nominal 1/3 barrel per day Riser FCC/Resid FCC unit consisting of cracking, stripping, and regeneration sections, all operating under pressure with catalyst continuously circulating between these sections. This unit is designed to handle a wide range of operating variables. The temperatures and flow rates are controlled to tight tolerances as required by the specifics of the test program.

Figure 5-1 - FCC Pilot Plant Flowsheet



Process Flows

Oil feed is introduced into the unit from a heated feed tank by a gear pump. After passing through an electrically-heated preheater. The hot feedstock stream (normal range: 150-700°F) is atomized with dispersion nitrogen in a proprietary injection nozzle and carries into the flowing catalyst stream.

Regenerated catalyst (1100-1450°F) passes through a slide valve which controls the catalyst circulation rate (20-50 lb/hr) and is transported via a short transfer line to the bottom of the riser with a flow of nitrogen. The dispersed feed stream and the catalyst stream mix in the 1/2" SCH 40 injection chamber, designed for use with heavy feeds to avoid coke deposition on the riser walls, and are transported up the riser.

Riser effluent enters the stripper tangentially. The solids are disengaged from the product oil vapors. The solids are then stripped with nitrogen to remove interstitial and adsorbed hydrocarbons from the catalyst. The catalyst then flows to the regenerator where coke is burned off. Upon leaving the unit, the regenerator flue gas is cooled to about 50°F to condense water of combustion. The remaining gas is then measured, analyzed, and vented. It is from these flue gas measurements that coke make is calculated.

From the disengager/stripper, the product oil vapors are partially condensed in two stages of cooling. Product is withdrawn continuously and collected hourly. The syncrude is analyzed off-line for dissolved gases to complete a full set of C_4 & lighter yields. Gas Chromatographic Simulated Distillation (GCSD) provides conversion and the yields of heavier fractions.

The uncondensed product gas exiting the low temperature receiver is then analyzed by an on-line G.C., measured by a wet test meter, and, finally, vented to a flare. Other on-line analytical instruments monitor the product gas density and the carbon dioxide and oxygen content of the regenerator flue gas.

Riser Description

The riser is a 1/4" SCH 40 stainless steel pipe, about 80 feet in length with provisions for temperature and pressure measurements at several points along its length. The riser is fitted with eight separate heating zones. The electrical heating wires are wound around a layer of insulation surrounding the riser. At the midpoint of each zone, skin thermocouples are welded to the riser outer surface. Internal thermocouples extend inside the riser to measure actual fluid temperatures. Additionally, jacket thermocouples are positioned in the insulation between the riser wall and the electrical windings at about the same elevation as both the skin and internal thermocouples. This design provides the flexibility to simulate adiabatic operation or to run isothermally. All runs in this program were made in the isothermal mode.

Disengager/Stripper Operation

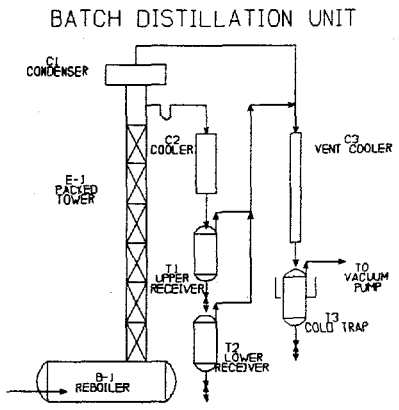
Product vapors exiting the riser (900-1050°F) enter the stripper tangentially into an annular space created by placing a collar within the stripper body. Catalyst is disengaged from the vapors by cyclonic forces and a change of direction as the vapors move downward in the annulus, then upward into the upper section of the stripper which contains a sintered-metal filter to prevent entrained catalyst particles from leaving the unit.

The catalyst passes downward into the stripping section. A bed of spent catalyst is maintained while stripping nitrogen passes countercurrently from its inlet, which is located by the entrance to the stripper standpipe.

Regenerator Operation

The regenerator controls the level of carbon on regenerated catalyst. The regenerator includes both electrical windings and an air-cooled jacket to permit the riser to run in heat balance independently of coke make. The air jacket also allows operation with higher carbon residue-containing feeds than would otherwise be possible. By adjusting the regenerator temperature to about 1325°F and maintaining oxygen content of the flue gas between about 6 and 12 percent, carbon on regenerated catalyst levels below 0.1 weight percent are produced.

Figure 5-2 - Batch Distillation



5.2.2 Batch Still

The Batch Still (Figure 5-2) consists of a 55 gallon reboiler, a 6 inch diameter by 15 foot stainless steel column packed

with 1/2 inch intalox saddles, an overhead condenser with reflux splitter, and a train of product coolers to collect the distillate.

The reboiler is electrically heated with boil-up rate automatically controlled by column pressure drop. Operation from atmospheric pressure to 5 mm Hg absolute is possible.

5.2.3 PILOT PLANT OPERATIONS

The production run consisted of several periods of operation, each at the same operating targets. Operation continued until 80 gallons of feedstock were processed at lined out conditions.

The feedstock was a blend of Amoco Heavy Vacuum Gas Oil and Direct Coal Liquid Distillate (42.6 vol% or 43.1 wt%). The catalyst used in the entire production run came from a sample of equilibrium Vektor-50 obtained from Conoco (Billings, MT). The catalyst charge to the FCC1 pilot plant was replaced weekly.

In all there were 145 hours spread over nine periods ranging from 2-29 "good hours" of lined out operation. During the first and seventh periods, two hours of product were saved and worked up separately for official material balances and yields. The first, in August, is called H-2038-4. The later run, in September, is H-2038-4A.

Once the desired amount of feedstock had been processed, the total liquid product from all but the four "official hours" were combined and charged to the Batch Still. The still was run at atmospheric pressure and total reflux to accomplish the debutanization. After the unit's temperature profile stabilized, indicating that the debutanization was complete, the reflux timer was set at 5:1. Depentanization was deemed complete once the overhead vapor temperature reached 120°F and the C_5-rich light naphtha product was collected.

Distillation continued at 5:1 reflux ratio until the overhead vapor temperature reached 430°F. At this time the heavy naphtha cut was withdrawn from the product receivers.

After cooling the reboiler contents, vacuum was pulled to 20 mm Hg and distillation was resumed at a 2:1 reflux ratio. Distillation continued until the overhead vapor temperature reached 415°F, equivalent to 650°F at atmospheric pressure. At this point, the reboiler temperature was 525°F, well below temperatures where cracking of the bottoms would be a concern.

5.3 RESULTS

5.3.1 FCC PILOT PLANT

The pilot plant operating conditions were previously established to produce the heat-balanced coke yield with the Amoco VGO at the baseline operating temperature (H-2038-1) and these conditions were maintained while running the DL-2 liquid (H-2038-2) and the current blend of the VGO and DL-2. A summary of the operating conditions and product yields is given in Table 5-1. Conversion and liquid product yields are calculated based upon GCSD data.

Section 5

M.W. Kellogg Activities

TABLE 5-1 — PILOT PLANT YIELDS SUMMARY

RUN NUMBER: H-2038-	1	2	4	4A
CATALYST/OIL RATIO	10.0	10.2	10.4	10.6
TEMPERATURES, DEG F:				
OIL PREHEAT	224	222	229	237
CATALYST INLET	1254	1252	1250	1252
RISER AVERAGE	984	985	983	984
MATERIAL BALANCE:				
CLOSURE, WT%	99.90	101.86	101.87	101.15
CONVERSION, WT%	77.10	76.96	75.63	76.05
TOTAL C2 & LIGHTER	2.93	2.14	2.53	2.18
TOTAL C3'S	7.06	6.02	6.81	6.82
TOTAL C4'S	12.18	10.17	11.79	11.78
TOTAL GASOLINE	49.89	54.97	50.08	51.00
TOTAL CYCLE OIL	22.90	23.04	24.37	23.95
COKE	5.04	3.66	4.42	4.27

Of special interest in this portion of the study are the individual C_5's yields. These are summarized in Table 5-2. Tables listing other detailed component yields and operating conditions for these FCC pilot plant runs are given in the Appendix to this report.

TABLE 5-2 — C5'S YIELDS

RUN NUMBER: H-2038-	4	4A	AVG
3-Methylbutene-1	0.09	0.11	0.10
i-Pentane	5.09	5.88	5.49
Pentene-1	0.21	0.23	0.22
2-Methylbutene-1	0.49	0.54	0.52
n-Pentane	0.58	0.64	0.61
Isoprene	0.04	0.03	0.04
t-Pentene-2	0.53	0.58	0.56
c-Pentene-2	0.30	0.33	0.32

2-Methylbutene-2	0.89	0.97	0.93
1t,3-Pentadiene	0.03	0.02	0.03
1c,3-Pentadiene	0.07	0.06	0.07
Cyclopentene	0.23	0.24	0.24
Cyclopentane	0.00	0.03	0.02
TOTAL C5'S, WT%	8.55	9.66	9.10

5.3.2 BATCH STILL

The charge to the Batch Still was distilled to produce four liquid products, IBP-120°F, 120-430°F, 430-650°F, and 650°F+. Also, based upon debutanization of samples from runs H-2038-4 & 4A, there would be an expected loss on debutanization of about 2.86 wt% of the charge.

The actual results (also assuming the expected loss of C_4 & lighter materials), compared to those that would be predicted based upon GCSD analysis, are given in Table 5-3. Overall the agreement is good.

TABLE 5-3 — BATCH STILL RESULTS

Yields, wt%	Actual	GCSD
C_4 & Lighter	2.86	2.15
IBP-120°F	5.18	5.92
120-430°F	58.24	57.48
430-650°F	23.27	27.65
650°F+	8.31	6.80
Balance	97.86	100.00

The IBP-120°F (light naphtha) and 120-430°F (heavy naphtha) cuts were analyzed by PIANO to observe the C_5's distributions and check the purity of the light naphtha or C_5's (IBP-120°F) cut. Analysis showed that the C_5's cut retained about 6.6 wt% C_4's and had about 3.0 wt% C_6's. The heavy naphtha contained about 1.3 wt% C_5's. These results are comparable to what has been observed in the past for FCC products distilled on the lab true boiling point (TBP) still after debutanization in the lab debut still.

M.W. Kellogg Activities

5.4 DISCUSSION

5.4.1 FCC PILOT PLANT

The main objective of this phase was to process 80 gallons of blended feedstock, at the operating targets previously identified, to produce sufficient product materials for further study. Feedstock was prepared in four batches measuring 25.0-25.3 API.

Table 5-4 again summarizes the yields of the two recent runs with blended feed (H-2038-4 & 4A) and the two original runs with the individual components. Also of interest is the comparison between a weighted average of the yields from the two runs with neat feeds and those of the runs with blended feeds. For essentially every component and boiling range, the results of the runs are within the limits of the reproducibility of the unit.

In working up the new runs, it was discovered that a change in the chromatographic response factor for ethylene had occurred between the times of the older and newer sets of runs. Since the new factor is more reliable, it has been applied to the older data, as well. Only the ethylene yields are significantly different than had previously been reported, raising the total C_2 and lighter yields by less than 0.5 wt%.

TABLE 5-4 — COMPARISON OF BLEND AND CALCULATED BLEND RESULTS

RUN NUMBER: H-2038-	1	2	4	4A	AVG	CALC
FEED	VGO	DL2	BLEND	BLEND	BLEND	BLEND
TOTAL C2 & LIGHTER	2.93	2.14	2.53	2.18	2.36	2.59
TOTAL C3'S	7.06	6.02	6.81	6.82	6.82	6.61
TOTAL C4'S	12.18	10.17	11.79	11.78	11.79	11.31
TOTAL GASOLINE	49.89	54.97	50.08	51.00	50.54	52.08
TOTAL CYCLE OIL	22.90	23.04	24.37	23.95	24.16	22.96
COKE	5.04	3.66	4.42	4.27	4.35	4.45
CONVERSION, WT%	77.10	76.96	75.63	76.05	75.84	77.04

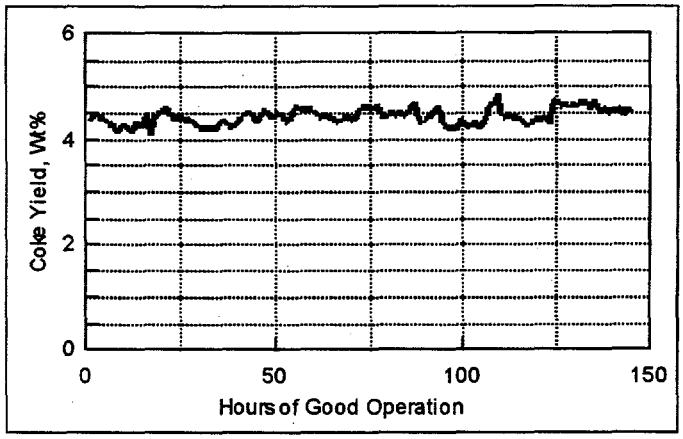
Run conditions which were used with the 80-foot riser remained (nominally) a C/O=10.0, riser temperature of 985°F, and catalyst preheat of 1250°F. Figure 5-3 shows the coke yields over the 145 hours of good operation. Coke yields averaged 4.44 ± 0.14 wt%.

Figures 5-4, 5-5, and 5-6 plot some key temperatures: catalyst temperature, riser average temperature, and oil inlet temperature, over the 145 good hours of operation. Averages for these variables were:

Catalyst Temperature = 1251 ± 3°F
Riser Average Temperature = 984 ± 1°F
Oil Inlet Temperature = 232 ± 4°F

The variations in these temperatures are well within acceptable tolerances. The oil inlet temperature, which shows the greatest short-term variations, has the least effect on unit stability, especially at high C/O, such as was employed in these runs.

Figure 5-3 - DL2 Production Run Coke Yield



5.4.2 BATCH STILL

Operation of the still was assigned run number H-2038-11. The quality of the light and heavy naphthas prepared in this run look quite good in terms of separation, that is, the overlap between the C_4, C_5, and C_6 components. However, the overall recovery of C_5's was less than might have been anticipated.

Table 5-5 lists the grams of each C_5 component expected to be in the charge to the still and the amounts recovered in the IBP-120°F (Cut 1) and 120-430°F (Cut 2) fractions.

Figure 5-4 - DL2 Production Run Catalyst Temperature

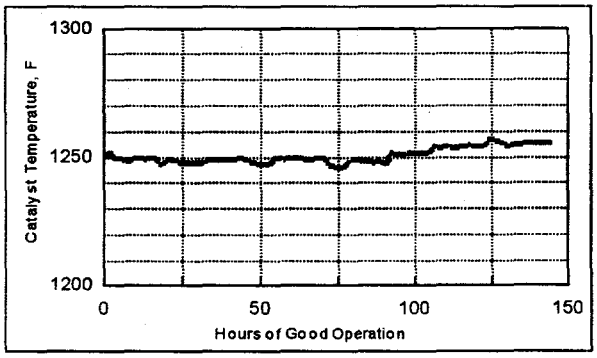


Figure 5-5 - DL2 Production Run Riser Temperature

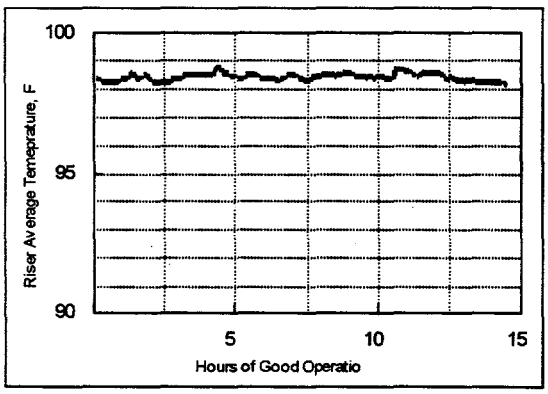
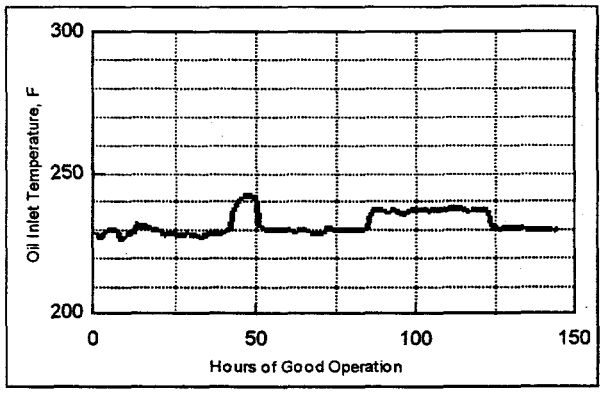


Figure 5-6 - DL2 Production Run Oil Inlet Temperature



M.W. Kellogg Activities

TABLE 5-5 — C_5'S RECOVERY

RUN NUMBER: H-2038-11	Charge	Cut 1	Cut 2	Total	Rec in
	grams	grams	grams	Rec %	Cut1%
3-Methylbutene-1	115	59	3	54	95
i-Pentane	7115	4679	335	70	93
Pentene-1	331	227	24	76	91
2-Methylbutene-1	802	541	62	75	90
n-Pentane	1000	709	123	83	85
Isoprene	63	40	7	75	85
t-Pentene-2	953	686	138	86	83
c-Pentene-2	544	390	87	88	82
2-Methylbutene-2	1693	1121	343	86	77
1t,3-Pentadiene	45	28	12	90	70
1c,3-Pentadiene	131	67	12	61	85
Cyclopentene	482	236	207	92	53
Cyclopentane	0	118	0	>100	100
TOTAL	13273	8901	1353	77	87

Overall recovery of C_5's was about 77%, with about 87% of this being recovered in the IBP-120°F cut. The trend in percentages recovered generally follows the boiling points of the components, with the least recovery of the lightest materials. Likewise, the trend in recovery of C_5's compounds in the IBP-120°F is greater for the lighter components, as would be expected.It is possible that the overall recovery was actually better than the numbers show. The IBP-120°F cut is so volatile that handling losses can become a factor. Losses may have occurred while taking a sample from the product can or even when putting some of the sample into the vial for injection into the PIANO analyzer. Kellogg recommended that SwRI recheck the analysis before proceeding with other tests.

GASOLINE OCTANE NUMBERS

Samples of the condensed total liquid products from runs H-2038-1 and H-2038-2 were debutanized at KTDC and sent to SwRI for fractionation and determination of the octane numbers of the IBP-430°F gasolines. For run H-2038-4A the FCC product was debutanized and TBP distilled at KTDC. Octane numbers were measured at Core Laboratories.

Table 5-6 summarizes results based upon TBP distillations, in both weight and volume percent yields. The agreement between the results with blended feedstock versus those calculated on a weighted average of the two components is reasonably good, especially taking into account that the distillations were performed at different laboratories. A table listing componential yields is included in the Appendix to this report.

M.W. Kellogg Activities

TABLE 5-6 —PRODUCT YIELDS BASIS TBP DATA

RUN NUMBER: H-2038-	1		2		4A		CALC	
	WT%	VOL%	WT%	VOL%	WT%	VOL%	WT%	VOL%
TOTAL C2 & LIGHTER	2.93	---	2.14	---	2.18	---	2.59	---
TOTAL C3'S	7.06	12.19	6.02	10.62	6.82	11.87	6.61	11.52
TOTAL C4'S	12.18	18.39	10.17	15.80	11.78	17.94	11.31	17.29
TOTAL GASOLINE	51.16	61.09	56.20	66.17	51.72	61.59	53.33	63.25
TOTAL CYCLE OIL	21.63	19.08	21.81	19.76	23.23	20.49	21.71	19.37
COKE	5.04	---	3.66	---	4.27	---	4.45	---
CONVERSION, WT%	78.37	80.92	78.19	80.24	76.77	79.51	78.29	80.63

Table 5-7 gives the octane numbers for the total gasoline product from each run. The numbers that were found for the IBP-430°F liquids were adjusted for the C_5+ in the product gases using octane blending values and component specific gravities. That the octane numbers for H-2038-4A differ from the calculated weighted averages of those from runs H-2038-1 & 2 is probably a reflection of a combination of differences in conversion, laboratory distillations, and the use of different laboratories for measuring the octane numbers, themselves. For run H-2038-3 from the older run series, distilled and tested by SwRI, the RON of 92.4 and MON of 83.5 agreed well with calculated values of 92.8 RON and 83.5 MON. In that run, the level of DL2 in the blended feedstock was 33.3 vol%.

TABLE 5-7 —GASOLINE OCTANE NUMBERS

RUN NUMBER: H-2038-	1		2		4A		CALC	
	RON	MON	RON	MON	RON	MON	RON	MON
OCTANE NUMBERS	92.6	83.3	93.1	84.0	91.5	82.9	92.8	83.6

M.W. Kellogg Activities

Section 5
Appendix

TABLE A

OPERATING CONDITIONS

FEEDSTOCK:		F-9819	F-9888	F-9819/9888	F-9819/9888	F-9819/9888
CATALYST:		F-9804	F-9804	F-9804	F-9804	F-9804
RUN NUMBER:	H-2038-	1	2	3	4	4A
DATE:		3/25/96	3/25/96	3/26/96	8/13/96	9/5/96
OIL FEED RATE , GRAM/HR		1924	1887	1912	1920	1935
CATALYST RATE , LB/HR		42.5	42.3	43.3	44.0	45.0
CATALYST/OIL RATIO		10.0	10.2	10.3	10.4	10.6
MATERIAL BALANCE:						
CLOSURE, WT%		99.90	101.86	100.71	101.87	101.15
GASOLINE, WT%		49.89	54.97	51.10	50.08	51.00
CONVERSION, WT%		77.10	76.96	76.15	75.63	76.05
COKE YIELD, WT%		5.04	3.66	4.51	4.42	4.27
SELECTIVITY, W/W		0.65	0.71	0.67	0.66	0.67
C/(1-C), W/W		3.37	3.34	3.19	3.10	3.18
RISER OUTLET PRESSURE , PSIG		35.0	35.0	35.0	35.0	35.0
TEMPERATURES, DEG F:						
OIL PREHEAT		224	222	221	229	237
CATALYST INLET		1254	1252	1255	1250	1252
RISER PROFILE, FT						
0.58 (MIXING ZONE)		984	997	994	994	979
5.47		992	992	992	984	982
9.22		985	986	986	984	985
17.10		985	987	986	984	985
19.18		979	981	978	982	976
22.87		989	991	990	989	991
26.12		980	982	980	981	N.G.
33.99		991	993	992	983	987
36.08		986	988	987	984	979
41.33		982	982	981	982	N.G.
45.08		982	980	978	984	986
50.92		990	991	991	982	988
55.64		991	992	992	986	987
60.45		974	975	975	967	975
64.78		987	988	987	984	986
69.53		974	974	974	989	986
75.45		987	987	987	983	987
RISER AVERAGE (EX MIX ZONE)		984	985	985	983	984

TABLE B

PRODUCT YIELD SPECTRUM
(USING GCSD DATA)

NORMALIZED, BASIS FRESH FEED, WT%

FEEDSTOCK:		F-9819	F-9888	F-9819/9888	33.7 wt% DL2	F-9819/9888	F-9819/9888		42.6 wt% DL2
CATALYST:		F-9804	F-9804	F-9804	Weighted Avg	F-9804	F-9804	Average	Weighted Avg
RUN NUMBER:	H-2038-	1	2	3	1&2	4	4A	4 & 4A	1&2
H2S		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2		0.13	0.17	0.14	0.14	0.16	0.15	0.16	0.15
CH4		1.13	0.75	1.00	1.00	1.07	1.02	1.05	0.97
C2H4		0.87	0.71	0.80	0.82	0.69	0.53	0.61	0.80
C2H6		0.80	0.51	0.70	0.70	0.61	0.48	0.55	0.68
C3H6		5.65	4.28	5.13	5.19	5.34	5.32	5.33	5.06
C3H8		1.41	1.74	1.51	1.52	1.47	1.50	1.49	1.55
C4H6		0.09	0.04	0.03	0.07	0.07	0.04	0.06	0.07
1-C4H8		2.02	1.12	1.59	1.72	1.39	1.34	1.37	1.63
I-C4H8		1.92	0.73	1.27	1.52	1.00	1.05	1.03	1.41
T-2-C4H8		1.71	1.06	1.46	1.49	1.67	1.65	1.66	1.43
C-2-C4H8		0.94	0.59	0.84	0.82	1.29	1.29	1.29	0.79
IC4H10		4.24	5.03	4.71	4.51	5.13	5.16	5.15	4.58
NC4H10		1.26	1.60	1.36	1.37	1.24	1.25	1.25	1.41
C5+ IN GAS		4.22	3.94	4.38	4.13	5.40	5.94	5.67	4.10
IBP-430 F		45.67	51.03	46.72	47.48	44.68	45.06	44.87	47.98
430-650 F		17.21	19.33	18.84	17.93	19.61	19.18	19.40	18.13
650+ F		5.69	3.70	5.01	5.02	4.76	4.76	4.76	4.83
COKE		5.04	3.66	4.51	4.57	4.42	4.27	4.35	4.45
TOTAL		100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
SUMMARY									
TOTAL C2 & LIGHTER		2.93	2.14	2.64	2.66	2.53	2.18	2.36	2.59
TOTAL C3'S		7.06	6.02	6.64	6.71	6.81	6.82	6.82	6.61
TOTAL C4'S		12.18	10.17	11.26	11.50	11.79	11.78	11.79	11.31
TOTAL GASOLINE		49.89	54.97	51.10	51.60	50.08	51.00	50.54	52.08
TOTAL CYCLE OIL		22.90	23.04	23.85	22.95	24.37	23.95	24.16	22.96
COKE		5.04	3.66	4.51	4.57	4.42	4.27	4.35	4.45
CONVERSION		77.10	76.96	76.15	77.05	75.63	76.05	75.84	77.04

TABLE C

PRODUCT YIELD SPECTRUM
(Using GCSD & PIANO Input)

NORMALIZED, BASIS FRESH FEED, WT%

FEEDSTOCK:		F-9819/ F-9888		
CATALYST:		F-9804	F-9804	Average
RUN NUMBER:	H-2038-	4	4A	4 & 4A
H2S		0.00	0.00	0.00
H2		0.16	0.15	0.16
CH4		1.07	1.02	1.05
C2H4		0.69	0.53	0.61
C2H6		0.61	0.48	0.55
C3H6		5.34	5.32	5.33
C3H8		1.47	1.50	1.49
C4H6		0.07	0.04	0.06
1-C4H8		1.39	1.34	1.37
I-C4H8		1.00	1.05	1.03
T-2-C4H8		1.67	1.65	1.66
C-2-C4H8		1.29	1.29	1.29
IC4H10		5.13	5.16	5.15
NC4H10		1.24	1.25	1.25
3-Methylbutene-1		0.09	0.11	0.10
i-Pentane		5.09	5.88	5.49
Pentene-1		0.21	0.23	0.22
2-Methylbutene-1		0.49	0.54	0.52
n-Pentane		0.58	0.64	0.61
Isoprene		0.04	0.03	0.04
t-Pentene-2		0.53	0.58	0.56
c-Pentene-2		0.30	0.33	0.32
2-Methylbutene-2		0.89	0.97	0.93
1t,3-Pentadiene		0.03	0.02	0.03
1c,3-Pentadiene		0.07	0.06	0.07
Cyclopentene		0.23	0.24	0.24
Cyclopentane		0.00	0.03	0.02
C6-430 F		41.53	41.34	41.44
430-650 F		19.61	19.18	19.40
650+ F		4.76	4.76	4.76
COKE		4.42	4.27	4.35
TOTAL		100.00	100.00	100.00
SUMMARY				
TOTAL C2 & LIGHTER		2.53	2.18	2.36
TOTAL C3'S		6.81	6.82	6.82
TOTAL C4'S		11.79	11.78	11.79
TOTAL GASOLINE		50.08	51.00	50.54
TOTAL CYCLE OIL		24.37	23.95	24.16
COKE		4.42	4.27	4.35
CONVERSION		75.63	76.05	75.84

TABLE D

PRODUCT YIELD SPECTRUM

NORMALIZED, BASIS FRESH FEED (TBP DATA)

FEEDSTOCK:		F-9819		F-9888		F-9819/9888		33.7 wt% DL2		F-9819/9888		42.6 wt% DL2	
CATALYST:		F-9804		F-9804		F-9804		Weighted Avg		F-9804		Weighted Avg	
RUN NUMBER:	H-2038-	1		2		3		1&2		4A		1&2	
		Wt%	Vol%	Wt%	Vol%	Wt%	Vol%	Wt%	Vol%	Wt%	Vol%	Wt%	Vol%
H2S		0.00	----	0.00	----	0.00	----	0.00	----	0.00	----	0.00	----
H2		0.13	----	0.17	----	0.14	----	0.14	----	0.15	----	0.15	----
CH4		1.13	----	0.75	----	1.00	----	1.00	----	1.02	----	0.97	----
C2H4		0.87	----	0.71	----	0.80	----	0.82	----	0.53	----	0.80	----
C2H6		0.80	----	0.51	----	0.70	----	0.70	----	0.48	----	0.68	----
C3H6		5.65	9.70	4.26	7.49	5.13	8.86	5.19	8.96	5.32	9.20	5.06	8.76
C3H8		1.41	2.49	1.74	3.13	1.51	2.68	1.52	2.70	1.50	2.67	1.55	2.76
C4H6		0.09	0.13	0.04	0.06	0.03	0.04	0.07	0.10	0.04	0.06	0.07	0.10
1-C4H8		2.02	3.03	1.12	1.71	1.59	2.40	1.72	2.59	1.34	2.02	1.63	2.47
I-C4H8		1.92	2.88	0.73	1.12	1.27	1.92	1.52	2.29	1.05	1.59	1.41	2.13
T-2-C4H8		1.71	2.51	1.06	1.59	1.46	2.16	1.49	2.20	1.65	2.44	1.43	2.12
C-2-C4H8		0.94	1.34	0.59	0.86	0.84	1.20	0.82	1.18	1.29	1.85	0.79	1.13
IC4H10		4.24	6.51	5.03	7.87	4.71	7.27	4.51	6.96	5.16	7.98	4.58	7.09
NC4H10		1.26	2.01	1.60	2.60	1.36	2.18	1.37	2.20	1.25	2.00	1.41	2.26
C5+ IN GAS		4.22	5.92	3.94	5.60	4.38	6.20	4.13	5.82	5.94	8.37	4.10	5.79
IBP-430 F		46.94	55.17	52.26	60.56	48.41	56.66	48.73	56.97	45.78	53.22	49.23	57.47
430-650 F		----	----	----	----	----	----	----	----	18.34	16.68	----	----
650+ F		21.63	19.08	21.81	19.76	22.16	20.22	21.69	19.31	4.89	3.81	21.71	19.37
COKE		5.04	----	3.66	----	4.51	----	4.57	----	4.27	----	4.45	----
TOTAL		100.00	110.76	100.00	112.35	100.00	111.80	100.00	111.29	100.00	111.89	100.00	111.44
SUMMARY		Wt%	Vol%	Wt%	Vol%	Wt%	Vol%	Wt%	Vol%	Wt%	Vol%	Wt%	Vol%
TOTAL C2 & LIGHTER		2.93	----	2.14	----	2.64	----	2.66	----	2.18	----	2.59	----
TOTAL C3'S		7.06	12.19	6.02	10.62	6.64	11.55	6.71	11.66	6.82	11.87	6.61	11.52
TOTAL C4'S		12.18	18.39	10.17	15.80	11.26	17.17	11.50	17.53	11.78	17.94	11.31	17.29
TOTAL GASOLINE		51.16	61.09	56.20	66.17	52.79	62.85	52.86	62.79	51.72	61.59	53.33	63.25
TOTAL CYCLE OIL		21.63	19.08	21.81	19.76	22.16	20.22	21.69	19.31	23.23	20.49	21.71	19.37
COKE		5.04	----	3.66	----	4.51	----	4.57	----	4.27	----	4.45	----
CONVERSION		78.37	80.92	78.19	80.24	77.84	79.78	78.31	80.69	76.77	79.51	78.29	80.63

Section 6

Project Management

6.1 Reports and Schedules

The milestone schedule and status for the Basic Program and Option 1 is shown in Figure 6-1.

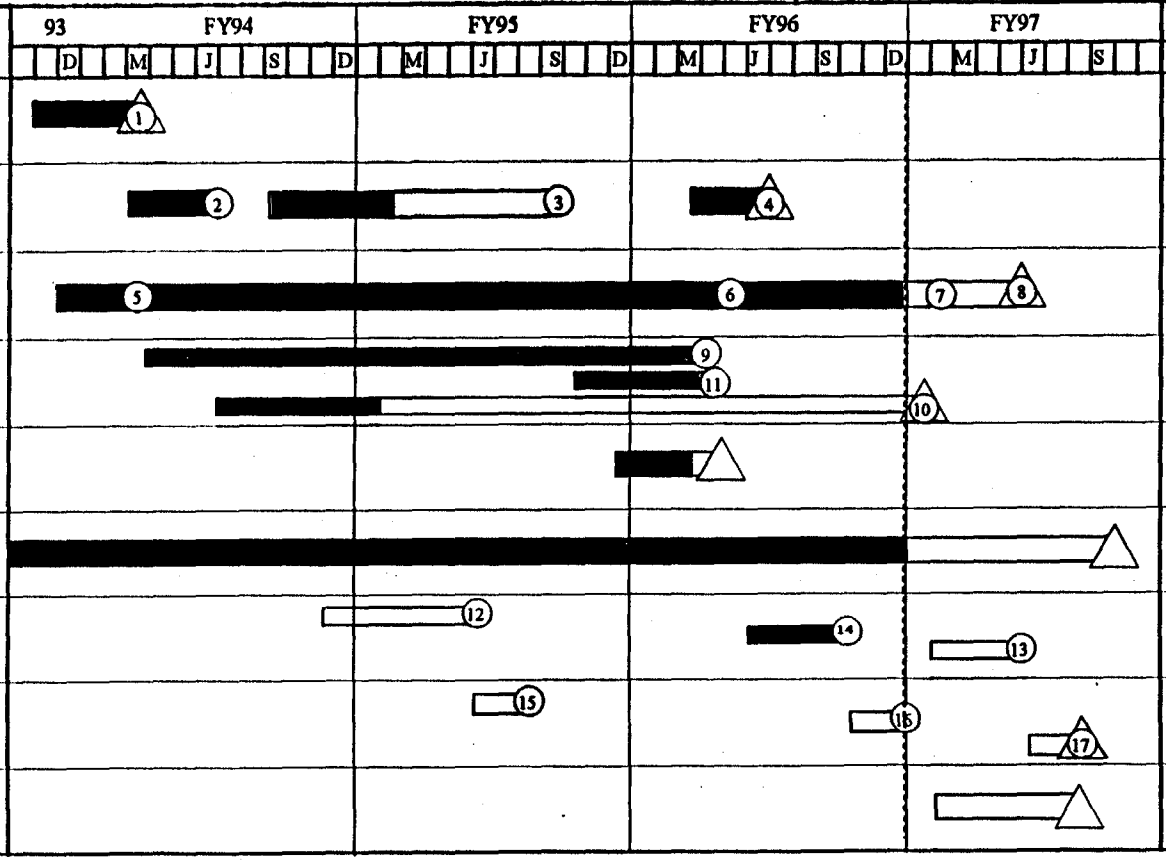
DOE F1332.3
(11.84)

FORM APPROVED
OMB NO 1901.1400

Figure 6-1
Milestone Schedule for Basic Program & Option 1

☐ PLAN ■ STATUS REPORT

1. TITLE	Refining and End Use Study of Coal Liquids	2. REPORTING PERIOD 10/1/96 to 12/31/96	3. IDENTIFICATION NUMBER DE-AC22-93PC91029
4. PARTICIPANT NAME AND ADDRESS	Bechtel Corporation 50 Beale Street San Francisco, CA 94105		5. START DATE 11/1/93 6. ESTIMATED COMPLETION DATE 9/30/97



7. ELEMENT CODE	8. REPORTING ELEMENT	10. PERCENT COMPLETE a. Plan	b. Actual
Task 1	Project Work Plan	100	100
Task 2	Feed Characterization	100	70
Task 3	Linear Programming (LP) Analysis	98	82
Task 4	Pilot Plant Analysis	100	75
Task 5	Option 1 Work Plan	100	60
Task 6	Administration Task	79	79
Option 1 Task 1	Pilot Plant Analysis (Produce Fuels)	72	30
Option 1 Task 2	Characterization, Blending, and Testing	0	0
Option 1 Task 3	Economic Study	0	0

1 Submit Final Work Plan
2 Characterize DL1 liquid
3 Characterize IL liquid
4 Characterize DL2 liquid
5 Develop LP model
6 Input DL pilot plant data
7 Input IL pilot plant data
8 Conduct evaluation runs
9 Conduct DL1 pilot plant tests
10 Conduct IL pilot plant tests
11 Conduct DL2 pilot plant tests
12 Production runs for DL1 (deleted from program)
13 Production runs for IL
14 Production runs for DL2
15 ASTM tests for DL1 (deleted from program)
16 ASTM tests for IL
17 ASTM tests for DL2

11. SIGNATURE OF PARTICIPANT'S PROJECT MANAGER AND DATE

Cliff Lowe 2/4/97