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**THE SELECTIVE CATALYTIC CRACKING OF FISCHER-TROPSCH LIQUIDS TO
HIGH VALUE TRANSPORTATION FUELS**

Final Report

**By
M. M. Schwartz**

Work Performed Under Contract No. AC22-91PC90057

**For
U.S. Department of Energy
Pittsburgh Energy Technology Center
Pittsburgh, Pennsylvania**

**By
Amoco Oil Company
Naperville, Illinois**

MASTER

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FINAL REPORT

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FOR

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EXECUTIVE SUMMARY

Amoco Oil Company, under a contract with the United States Department of Energy, investigated a selective catalytic cracking process (FCC) to convert the Fischer-Tropsch (F-T) gasoline and wax fractions to high value transportation fuels. The primary tasks of this contract were to (1) optimize the catalyst and process conditions of the FCC process for maximum conversion of F-T wax into reactive olefins for later production of C₄-C₈ ethers, and (2) use the olefin-containing light naphtha obtained from FCC processing of the F-T wax as feedstock for the synthesis of ethers.

The catalytic cracking of F-T wax feedstocks gave high conversions with low activity catalysts and low process severities. HZSM-5 and beta zeolite catalysts gave higher yields of propylene, isobutylene, and isoamylenes but a lower gasoline yield than Y zeolite catalysts. Catalyst selection and process optimization will depend on product valuation. For a given catalyst and process condition, Sasol and LaPorte waxes gave similar conversions and product selectivities. The contaminant iron F-T catalyst fines in the LaPorte wax caused higher coke and hydrogen yields.

The net product values for catalytic cracking F-T wax feedstock in eight pilot plant runs were calculated with the iso-olefins valued as alkylate or gasoline (simple refinery configuration) and as feedstock to an ether unit (complex refinery configuration). Irrespective of catalyst, the net product values from catalytic cracking F-T wax feedstock were always higher when the iso-olefins were valued as feed to an ether unit rather than fuel. In a simple refinery configuration, additional experimental work would be required to identify the catalyst that gave the highest net product value. In a complex refinery configuration, HZSM-5 and beta zeolite catalysts gave higher net product values than Y zeolite catalyst because of their higher yields of iso-olefins. The net product values of the HZSM-5 and beta zeolite catalysts were additionally increased over the Y zeolite catalyst when propylene was valued as feedstock to a di-isopropyl ether unit rather than fuel gas.

The product value for hydrocracking F-T wax was calculated from literature data and compared with the product values from the eight pilot plant runs. The value for the hydrocracking case was below the values for catalytic cracking using a complex refinery configuration but were similar to the product values for catalytic cracking using a simple refinery configuration. If hydrocracking severity were higher, producing more naphtha such as in the all-gasoline mode operation, the product value from hydrocracking might be higher than from FCC in a simple refinery configuration.

The 200-°F product from catalytic cracking pilot plant runs was demonstrated to be a good feedstock for the production of mixed ethers by reaction with methanol using conventional technology.

1.0 BACKGROUND

Fischer-Tropsch synthesis technology produces liquid hydrocarbons from synthesis gas (hydrogen and carbon monoxide) derived from the gasification of coal. Domestic supplies of both high- and low- rank coals are extensive and represent a strategic resource to supplement dwindling petroleum reserves. The F-T technology has been practiced commercially at Sasol in South Africa since the mid-1950's. The F-T liquid product consists of a broad range of normal paraffins (C₅-C₅₀) and a small quantity of oxygenates and olefins. The gasoline range C₅-C₁₂ product fraction consists of linear paraffins and olefins of low octane number. The distillate fraction, C₁₂-C₁₈, is an excellent quality fuel. The largest product fraction, C₁₈+, is primarily wax and is useless as a transportation fuel. There are many studies on the upgrading of these F-T liquids. These products are further treated by conventional petroleum processes, such as hydrotreating, reforming and catalytic cracking to

produce conventional gasoline and distillate fuels. There are no reported studies of the catalytic cracking processing of F-T liquids to produce C₃-C₈ olefins as feedstocks for the synthesis of gasoline range ethers and alcohols.

Governmental regulations, most recently in the Clean Air Act Amendments of November 1990, have resulted in the phase-out of lead additives, lowering of the Reid vapor pressure of gasoline and in some geographical areas, the mandated use of oxygenates, i.e., alcohols and ethers. Methyl tertiary butyl ether (MTBE) and tertiary amyl methyl ether (TAME) are the most common oxygenates that are added to fuel because they reduce automotive carbon monoxide emissions, have high blending gasoline octane ratings, R+M/2, (MTBE=108, TAME=102) and have low Reid vapor pressure (MTBE=8-10, TAME=1). These ethers are produced commercially by the etherification of the appropriate olefin (isobutylene for MTBE and isoamylenes for TAME) with methanol. These olefins are derived from conventional petroleum processes such as catalytic cracking or steam/thermal reforming. MTBE has attracted the most attention in recent years. The growth rate for MTBE production could reach 25% per year by 1995.^(1,2) There is a growing need for alternative sources of olefins for ethers and alcohols as demand for these materials escalates beyond the capacity of conventional petroleum processes. The selective catalytic cracking task of this program addresses this need for an alternative olefin feedstock for these oxygenate syntheses.

The reaction of olefin with methanol to form ether is catalyzed by acidic ion exchange resins at low temperatures (100-200°F) and moderate pressures (100-400 psig). These etherification reactions were discovered in 1907.⁽³⁾ Detailed catalysis studies of these reactions have only recently appeared in the scientific literature.⁽⁴⁾ These reactions are mildly exothermic and equilibrium limited. At low temperatures, the equilibrium is shifted to the right. The original study of these reactions by Snamprogetti workers showed that the ratio of etherification rates for isobutylene and isoamylenes was 1.85.⁽⁴⁾ Both isomerization and etherification reactions were observed with the two reactive isoamylenes, 2-methyl-1-butene and 2-methyl-2-butene. The other isomer, 3-methyl-1-butene, was not reactive.

Reactive C₆ olefins, which are also potential feedstocks for ether production, are produced in significant quantities together with isobutylene and isoamylenes from the catalytic cracking of F-T wax. There is less information in the open literature about the etherification reactions of C₆ olefins. The Etherol process⁽⁵⁾ produces a mixed ether product from C₄-C₇ reactive iso-olefins in naphtha. One report⁽⁶⁾ provided some information about the reaction of C₆ iso-olefins with methanol to produce the higher ethers, MTHE's, methyl tertiary hexyl ethers. Several C₆ olefins can react with methanol to produce three different THME ethers:

Scheme 1:

Scheme 2:

Scheme 3:

The reactions of these C₆ olefins to the THME ethers could contribute significantly to the total ether yields from the FCC naphtha. The ether synthesis task of this program is to produce THME and TAME ethers from the light naphtha products of the pilot plant wax cracking studies. The reaction of isobutylene and methanol to form MTBE was not studied in this program because it is well known.

2.0 SCOPE OF WORK

2.1 Project Description

The objective of this program is to prepare high-value transportation fuels, including gasoline, distillate, and gasoline range ethers and alcohols from Fischer-Tropsch liquids using a selective catalytic cracking process. The C_4 - C_8 product olefins would then be etherified with methanol to prepare the target ethers. The gasoline and distillate products are also expected to be superior to conventional fuels because of the unique combination catalysts to be used in this process.

A two year, multi-task program will be used to accomplish the objective to develop a selective catalytic cracking process to produce premium transportation fuels, including ethers from Fischer-Tropsch gasoline and wax products.

2.2 Task Description

The program was split into one project planning and six experimental tasks, Tasks 1 and 2-7, respectively. Only the results of the six experimental tasks will be presented in this report. Specifically, the program consisted of the following tasks:

- 2.2.1. -- Project Management Plan.
- 2.2.2. -- Characterization of Feedstocks and Equipment Calibration (Task 2).
- 2.2.3. -- Catalytic Cracking Catalyst Screening Program (Task 3).
- 2.2.4. -- Pilot Plant Tests of the Optimized Catalyst and Process (Task 4).
- 2.2.5. -- Preparation of C_5 - C_8 Ethers and C_3 - C_8 Alcohols; subsequently modified to delete alcohols (Task 5).
- 2.2.6. -- Evaluation of Gasoline Blending Properties of Ethers and Alcohol Products; subsequently modified to exclude alcohols (Task 6).
- 2.2.7. -- Scoping Economic Evaluation of the Proposed Processes (Task 7).

3.0 FEEDSTOCK CHARACTERIZATION and EQUIPMENT CALIBRATION (TASK 2)

3.1 Feedstock Characterization

3.1.1 Sasol Wax

The DOE procured for Amoco a 55 gallon drum of F-T wax that was produced by Sasol, Limited, at their commercial Arge fixed-bed reactor. Figure 1 shows its high temperature simulated boiling point distribution curve. Figure 2 shows its carbon number distribution. Table I shows its viscosities at 135 and 150°C.

FIGURE 1

HIGH TEMPERATURE SIMULATED DISTILLATION
SASOL AND LAPORTE WAX FEEDSTOCKS

SASOL LaPorte LaPorte
WAX WAX,#1 WAX,#2

FIGURE 2

NORMAL PARAFFIN DISTRIBUTION

SASOL AND LaPorte WAX FEEDSTOCKS

WEIGHT PERCENT

CARBON NUMBER

LaPorte SASOL

WAX

WAX

—■—

—◆—

TABLE I

VISCOSITIES OF SASOL AND LAPORTE WAXES

	Sasol Wax	LaPorte Wax
Viscosity @ 135°C, cSt	6.65	64.07
Viscosity @ 150°C, cSt	5.28	23.30

3.1.2 Air Products Wax

The DOE also procured for Amoco F-T wax feedstock from the Liquid Phase Fischer-Tropsch (LPFT) synthesis demonstration run (19 day run, August 4-23, 1992) at the LaPorte, Texas, 0.7 T/D plant. Those runs used a silica-supported iron catalyst. The presence of some initial catalyst fines and some attrition in the reactor caused a significant contamination (2-4 wt.%) of the wax product with F-T catalyst.⁽⁷⁾

Replicate analyses of its simulated boiling point distribution are shown in Figure 1. Its boiling point distribution is similar to that of the Sasol wax, but the LaPorte wax (61% >1000°F) contains more high boiling material than the Sasol wax (52% >1000°F). The normal paraffin distributions for the two samples also reflect this difference, as Figure 2 shows. The viscosity of Laporte wax at 135 and 150°C is shown in Table I. The much higher viscosity of the LaPorte wax compared with the Sasol wax is probably due to its high concentration of iron catalyst fines because, as Figure 1 shows, the simulated distillation curves of the two waxes are similar.

Table II presents the solids content analyses of the LaPorte wax. The wax was ashed and the residue was analyzed by atomic absorption spectroscopy for individual metals content. The value of 2.46% solids (oxide basis) agrees with the average values reported by the contractor for these runs, 2-4%.⁽⁷⁾ A simple centrifuge experiment did not provide for a satisfactory separation of the F-T catalyst solids from the hot wax sample. The chemical analysis results of the centrifuged wax sample, which is also shown in Table II, indicates that the catalyst solids are distributed in an increasing gradient from top (3681 ppm) to bottom (9490 ppm). However, the ash composition values from this centrifuge experiment do not agree with the overall analysis. This discrepancy suggests that the catalyst may not be distributed homogeneously in the wax. No further efforts were made to separate the catalyst from the wax.

3.1.3 Fischer-Tropsch Naphtha from UOP Pilot Plant

UOP provided a 1 quart sample of light naphtha produced in their F-T pilot plant. This light naphtha was found to be a complex mixture of C₅-C₁₂ linear paraffins, olefins and oxygenates. The oxygen compounds include the normal alcohols, methanol, ethanol, propanol and aldehydes such as propanal and butanal. Table III gives the qualitative GC-MS analysis of this naphtha.

3.2 Equipment Modification

The high melting point of the F-T wax made it impossible to feed to the small scale catalytic cracking system, the micro yields unit (MYU). The MYU can only run liquid feedstocks at temperatures below about 150°F. The wax is still a solid at that temperature. A new high temperature feed system for the MYU unit was built to overcome this problem. The components of this system include a syringe heater, heat traced stainless steel lines, and a high temperature switching valve. This system delivered the wax feedstock in a reproducible manner. However, the test results showed a greater variability in the precision of the product distributions, and lower product recovery, than is typical of MYU tests

with normal gas oil. It is likely that the high boiling point of the wax (50% greater than 1000°F) caused this variability.

TABLE II
CHEMICAL ANALYSES OF LAPORTE WAX

Sample ID 15586-012

Chemical Composition (ppm)
of Ash:

Si	=	1,710	SiO ₂	=	3,659
K	=	1,350	K ₂ O	=	1,626
Fe	=	13,900	Fe ₂ O ₃	=	17,882
Cu	=	<u>1,150</u>	CuO	=	<u>1,439</u>
		18,110			24,606

4.0 CATALYTIC CRACKING CATALYST SCREENING PROGRAM (TASK 3)

Appendix A lists the MYU runs that were made for this contract. Except as indicated otherwise, conversion is the sum of light gases, all liquid products boiling below 430°F, and coke. In some cases conversion was determined from the simulated distillation curve of the total liquid product. Research Octane Number (RON) and Motor Octane Number (MON) from the MYU experiments are calculated values that were obtained from the gas chromatographic analysis of the 430-°F liquid product. This empirical correlation was derived for conventional gas oil feedstocks and has not been demonstrated to be valid for the products from the wax feedstock. The RON and MON values from the wax feedstocks are presented only as directional indicators of process and catalyst changes with that feed.

4.1 Sasol Wax Feedstock

Scoping runs were made to identify the process conditions to use with this feedstock; catalyst compositions were varied to optimize selectivity to light olefins for oxygenate production; and the effect of multiple catalyst regeneration was simulated.

4.1.1 Comparison with Gas Oil Feedstock

The initial MYU runs with Sasol wax feedstock used the same catalyst and process conditions that are used in control tests with conventional gas oil, the usual feedstock to commercial FCC units. The control catalyst was equilibrium ultrastable Y (USY), and the control conditions were

Summary report from ADAM for F-TGAS1.D
Sample =UOP FISCHER-TROPSCH PRODUCT - Bill Reagan

Peak	Assignment	Confidence
1 (19.74)	1-Propanol	90
2 (20.16)	Ethene, 1,2-dichloro-, (E)-	96
3 (20.89)	Butanal	80
4 (21.09)	2-Butanone	78
5 (21.54)	Acetic acid	50
6 (21.69)	1-Hexene	91
7 (22.27)	Hexane	78
8 (25.63)	1-Butanol	90
9 (27.02)	2-Pentanone	64
10 (27.72)	Pentanal	64
11 (27.88)	Cyclohexene	90
12 (28.77)	1-Heptene	91
13 (29.71)	Heptane	91
14 (31.50)	Ethane, 1,1-diethoxy-	78
15 (34.40)	Pentane, 1-chloro-	83
16 (35.49)	2-Propenoic acid, 6-methylheptyl e...	78
17 (36.06)	2-Hexanone	80
18 (36.96)	Hexanal	95
19 (37.99)	1-Octene	95
20 (38.66)	Acetic acid, butyl ester	72
21 (39.06)	Octane	94
22 (39.35)	Propane, 1,1-diethoxy-	74
23 (40.00)	Propane, 1-(1-ethoxyethoxy)-	72
24 (44.20)	1-Hexanol	83
25 (45.08)	6-METHYL-1-OCTENE	47
26 (45.94)	2-Heptanone	91
27 (46.95)	Heptanal	87
28 (47.78)	1-Nonene	94
29 (48.00)	Ethanol, 2,2-diethoxy-	43
30 (48.25)	Acetic acid, pentyl ester	74
31 (48.78)	Nonane	95
32 (53.39)	Heptanol	90
33 (54.91)	2-Octanone	91
34 (55.90)	Cyclohexanamine	53
35 (56.36)	1-Decene	95
36 (56.67)	Acetic acid, hexyl ester	83
37 (57.21)	Decane	95
38 (61.11)	1-Octanol	86
39 (62.39)	2-Nonanone	91

Summary report from ADAM for F-TGAS1.D
Sample =UOP FISCHER-TROPSCH PRODUCT - Bill Reagan

Peak	Assignment	Confidence
40 (63.42)	5-Undecene	93
41 (63.62)	Acetic acid, heptyl ester	56
42 (64.07)	Undecane	94
43 (67.19)	Nonanol	78
44 (68.27)	2-Decanone	94
45 (69.01)	1-Dodecene	96
46 (69.55)	Dodecane	94
47 (72.25)	1-Decanol	91
48 (73.72)	1-Tridecene	98
49 (74.16)	Tridecane	94
50 (77.48)	Cyclododecane	94
51 (77.79)	Tetradecane	94
52 (80.36)	11-Tricosene	86
53 (80.61)	Pentadecane	93

970°F reaction temperature and 3 catalyst to oil weight ratio. Table IV summarizes the comparison of the cracking data for these two feedstocks. The conventional gas oil feedstock gave about 62% conversion, 4.3% coke, and less than 12% C₃ and C₄ using these conditions, versus over 85% conversion, 2.2% coke, and over 41% C₃ and C₄ with the wax feedstock. It is likely that heavy aromatic molecules and catalyst poisons, such as aromatic nitrogen compounds, in the normal gas oil feedstocks limit the amount of conversion by normal FCC catalysts. The wax feedstock does not contain these types of molecules and the overall conversion and olefin yields for the wax were significantly higher than for gas oil feedstocks.

TABLE IV

MYU TESTS: F-T WAX VS. CONVENTIONAL GAS OIL FEEDSTOCKS

MYU Test Results : 970°F, 3 C/O, Equilibrium Catalyst
Product Distribution and Other Properties

	Gas Oil	Wax, Run 91-59-101
Conversion, wt%	61.6	88.1
Products, wt%:		
C ₂ -	2.6	1.8
C ₃ + C ₄	11.6	31.4
C ₅	6.9	15.9
C ₆ -430 F	36.2	36.8
430+°F	38.4	11.9
Coke	4.3	2.2
Isobutylene	1.3	4.5
Isoamylenes	2.3	5.8
Gasoline, RON	90.4	85.8
MON	79.8	77.6

4.1.2 Effect of Reaction Severity

Runs were made using lower reaction severities with the wax feedstock because lower reaction severity would lower overall conversion. It was believed that selectivity for the desirable olefin products (isobutylene and isoamylenes) would increase at lower conversion. Reaction severity is a function of catalyst activity (a measure of the active sites or zeolite concentration) and process conditions (such as temperature, catalyst to oil weight ratio and space velocity). The simplest process variable to change in the MYU test was the reaction temperature. A series of MYU runs was made with reaction temperatures of 880-970°F using two FCC catalysts, an equilibrium USY faujasite zeolite sample and a beta zeolite sample, that were steamed to lower their activities. As Figure 3 shows, the beta zeolite sample had a slightly higher wax conversion level than the Y zeolite sample, but the wax conversions of both catalysts varied little with temperature. Both catalysts showed greater than 85% conversion of the wax feedstock, even at the lowest reaction temperature of 880°F. Figure 4 shows that isobutylene production was essentially independent over this temperature range, but that the Y zeolite produced less than the beta zeolite.

The next major process variable studied was the catalyst to oil weight ratio (c/o). This variable was changed by diluting the starting catalyst with an inert material. This procedure kept the total catalyst bed volume in the reactor constant so that heat and mass transfer effects were minimized between different catalyst blends. The diluent used was a low

surface area, clay-based microsphere, obtained from a catalyst manufacturer.

Control runs with this diluent at two reaction temperatures, 880 and 970°F, showed that it was not inert. As Table V shows, the diluent gave a low wax conversion level of 15-20%. Additionally, the boiling point distribution of the 430+°F total liquid product from the two catalytic cracking runs made with diluent was substantially changed from that of the wax feedstock. As Figure 5 shows, a large fraction of the 1000+°F feed components was converted to 1000-°F products, including a significant amount of distillate (430-650°F). As expected, the liquid product from the 970°F reaction temperature had a higher portion of the lower boiling products than the 880°F product. The causes of these effects are thermal cracking and/or catalytic cracking by the clay based microspheres. Additional experiments would be required to determine the relative contributions of the two mechanisms.

TABLE V

MYU TESTS: EFFECT OF CATALYST DILUENT

Run No.	12	13
Reaction Temp, °F	880	970
Conversion, wt%	11.8	17.3
Products, wt%:		
C ₄ -	2.6	5.9
C ₅ -430°F	9.1	11.4
430-650°F	22.7	24.6
650+°F	65.5	58.1
Coke	0.08	0.07

Three steamed FCC catalysts, containing zeolite Y, zeolite beta, and zeolite HZSM-5, were tested for wax conversion and selectivity as a series of blends (50%, 25%, 12.5%, 6.25%) with the above-mentioned diluent. Table VI lists the process conditions that were used and the wide range of conversion values that were obtained. Even though there was scatter in

FIGURE 3

**SASOL WAX MYU DATA:
EFFECT OF TEMPERATURE ON CONVERSION**

CONVERSION, WEIGHT PERCENT

FIGURE 4

**SASOL WAX MYU DATA:
EFFECT OF TEMPERATURE ON ISOBUTYLENE YIELD**

iC4=, iC5=, WT. %

TABLE VI
MYU TESTS: EFFECT OF CATALYST TO OIL RATIO

Equilibrium Zeolite Y, CCC1397					Steamed Zeolite Beta, CCC1875					Steamed HZSM-5 CCC1891				
Run No.	% Catalyst	Cat/Oil	Temp, °F	Conversion, Wt%	Run No.	% Catalyst	Cat/Oil	Temp, °F	Conversion, Wt%	Run No.	% Catalyst	Cat/Oil	Temp, °F	Conversion, Wt%
14	50	1.5	970	86.0	4	100	3	970	90.9	32	100	3	880	93.6
15	50	1.5	970	85.1	5	100	3	970	91.8	33	100	3	880	90.4
16	50	1.5	970	86.2	7	100	3	940	91.4	50	25	0.75	880	84.1
17	50	1.5	970	86.1	9	100	3	910	91.4	51	25	0.75	880	87.7
18	25	0.75	970	84.3	11	100	3	880	92.4	52	12.5	0.375	880	88.3
19	25	0.75	970	84.4	26	25	0.75	880	91.2	53	12.5	0.375	880	85.6
20	25	0.75	970	84.3	27	25	0.75	880	89.6	54	6.25	0.1875	880	84.1
21	25	0.75	970	84.5	28	12.5	0.375	880	88.9	55	6.25	0.1875	880	83.8
22	25	0.75	880	70.1	29	12.5	0.375	880	84.2	64	3.12	0.09	880	65.5
23	25	0.75	880	82.6	36	6.25	0.187	880	64.7	65	3.12	0.09	880	66.7
24	25	0.75	880	81.8	37	6.25	0.187	880	80.3					
25	25	0.75	880	74.7	42	6.25	0.187	880	83.3					
30	12.5	0.375	880	77.6	43	6.25	0.187	880	58.0					
31	12.5	0.375	880	83.0										
34	6.25	0.187	880	55.1										
35	6.25	0.187	880	64.0										
38	6.25	0.187	880	71.5										
39	6.25	0.187	880	62.2										
40	6.25	0.187	880	69.1										
41	6.25	0.187	880	71.4										

the conversion levels of replicate runs, especially at the high dilution of the Y and beta samples, Figure 6 shows that the wax conversion levels for the three catalysts decreased with amount of diluent; there may have been feed by-pass with the low catalyst levels or some other physical test problem. The HZSM-5 zeolite catalyst was the most active of the three catalysts for wax conversion, and the beta sample had a higher wax conversion level than the Y zeolite catalyst. At the dilution level of 1:15 (6.25% zeolite catalyst) the HZSM-5 sample had wax conversion levels of over 80% while the Y and beta samples had wax conversions between 60 and 70%. At a dilution level of 1:31 (3.1% zeolite), the HZSM-5 zeolite had a wax conversion of 65%.

Substantial differences in product selectivity were observed for these three catalysts. Examples are shown in Figures 7-12. Our hypothesis that higher yields of desired isoamylenes and isobutylene would be obtained at lower wax conversion levels was found to be incorrect, especially for the Y and beta zeolite catalysts. Figures 7 and 8 show that the yields of isoamylenes and isobutylene, respectively, increased with increasing conversion and appear to maximize at a conversion level in the 80-85% range. This did not occur for the HZSM-5 sample where high isoamylenes and isobutylene yields were obtained at 65% conversion, with little change at the higher conversion levels. The HZSM-5 zeolite gave higher yields of isoamylenes and isobutylene at 65% conversion than the other two catalysts did at 80-85% conversion. However, the HZSM-5 catalyst also produced much higher yields of propylene (Figure 9), total C_3+C_4 gas (Figure 10), and much lower $C_5-430^\circ F$ gasoline (Figure 11) than the other two catalysts. The beta zeolite was superior to the Y zeolite sample for olefin production, in agreement with the temperature survey test results previously presented in Figure 4. The coke yields for these three catalysts were a function of conversion (Figure 12) and the c/o level (Figure 13). At the higher conversion and c/o levels, the HZSM-5 catalyst produced the lowest coke yields. However, at low conversion levels, the coke differences between the catalysts were small. Figure 14 shows the behavior of these three zeolite catalysts for aromatics formation as a function of wax feedstock conversion. Some of the scatter in these results arises from the lack of precision of the GC method for compound identification. However, qualitative trends are apparent. The aromatics yields for the three catalysts rose very slowly up to about 85% conversion. The aromatics formation for the HZSM-5 catalyst, initially lower than the other larger pore zeolite catalysts, increased dramatically above 85% conversion. Further changes in the aromatics yields for the other catalysts also occurred to a lesser extent. Increased reaction temperature and catalyst to oil ratio also caused increases in the yield of aromatics.

Table VII presents the detailed product distributions for the three catalysts at a similar conversion level of about 83%, all obtained at $880^\circ F$ reaction temperature. A trade-off of high light olefin yields with low gasoline yield is indicated. This is an important issue when comparing different catalysts and optimizing process conditions. Catalyst selection and process optimization will depend on product valuation. The detailed C_5 and C_4 isomer distributions in the liquid products from these three tests are compared in Figures 15 and 16, respectively. The low yields of isobutane and isopentane with the HZSM-5 compared with the other two zeolites probably reflect the smaller pore structure of the HZSM-5 zeolite.

FIGURE 5

**SASOL WAX MYU DATA:
EFFECT OF DILUENT ON BOILING POINT DISTRIBUTION**

FIGURE 6

**SASOL WAX MYU DATA:
EFFECT OF DILUENT ON CONVERSION**

CONVERSION, WT. %

FIGURE 7

SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON ISOAMYLENE YIELD

FIGURE 8

SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON ISOBUTYLENE YIELD

FIGURE 9

**SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON PROPYLENE YIELD**

FIGURE 10

**SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON C3 AND C4 YIELD**

FIGURE 11

**SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON C5-430F NAPHTHA YIELD**

C5-430oF GASOLINE, WT. %

Y ZEOLITE BETA ZEOLITE ZSM-5 ZEOLITE

FIGURE 12

**SASOL MYU WAX DATA:
EFFECT OF CONVERSION ON COKE**

COKE, WT. %

Y ZEOLITE BETA ZEOLITE HZSM-5 ZEOLITE

TABLE VII
MYU TESTS: CATALYST COMPARISONS AT 880°F

Catalyst Type	Zeolite Y	Zeolite Beta	Zeolite HZSM-5
Run No.	031	042	055
Conversion, Wt%	83.0	83.3	83.8
Product Yields, Wt%:			
C ₂ ⁻	0.6	0.6	1.5
C ₃ ⁼	7.4	8.9	17.5
C ₃ ^o	0.8	0.9	2.7
C ₄ ⁼	13.2	17.7	27.4
C ₄ ^o	3.7	3.6	3.6
C ₅ ⁼	11.7	13.5	13.9
C ₅ ^o	3.6	2.2	2.0
C ₆ -430°F	41.7	35.8	15.3
430°F ⁺	17.0	16.8	16.2
Coke	0.3	0.2	0.1
Isobutylene	5.8	9.4	12.3
Isoamylenes	7.7	9.2	9.8
C ₅ -430°F	57.0	51.4	31.1
RON	85.2	84.4	84.4
MON	76.2	74.6	76.0

4.1.3 Non-zeolitic Catalysts

Amorphous, non-zeolitic FCC catalysts are no longer used in commercial units because they have lower activity than zeolite-based FCC catalysts. However, amorphous, low-activity catalysts may not be a disadvantage for F-T wax cracking because, as our tests have shown, it cracks readily under mild conditions. One example of this type of amorphous catalyst is the matrix only fraction of Davison Chemical Company's commercial Octacat D series of FCC catalysts. This material consists of an active alumina phase and inert clay components. Table VIII compares the wax cracking of a steamed sample of this catalyst with diluted zeolite Y and zeolite beta, with all the runs made at 970°F and 3 catalyst to oil ratio. Although the conversion levels of the three catalysts were not equivalent (they were between 71 and 78%), some qualitative observations are possible. The amorphous matrix sample had excellent light olefin selectivities, comparable to the beta zeolite and superior to the Y zeolite sample. As with the zeolite-based catalysts, the high yields of light olefins with the Octacat D matrix catalyst came at the expense of gasoline yield.

TABLE VIII
MYU TESTS: COMPARISON OF AMORPHOUS CATALYST AT 970°F

Catalyst Type	AD Matrix	Zeolite Y	Zeolite Beta
Run No.	048	046	044
C/O	3.0	0.2	0.2
Conversion, Wt%	72.6	70.9	77.8
Product Yields, Wt%:			
C ₂ ⁻	1.4	1.1	1.1
C ₃ ⁼	9.6	5.7	8.4
C ₃ ^o	1.0	0.7	0.8
C ₄ ⁼	16.4	9.7	14.9
C ₄ ^o	2.3	1.8	2.5
C ₅ ⁼	12.1	9.8	13.4
C ₅ ^o	1.7	1.8	1.5
C ₆ -430°F	27.4	40.2	35.1
430°F*	27.4	29.1	22.3
Coke	0.6	0.1	0.1
Isobutylene	7.2	4.2	7.6
Isoamylenes	8.1	5.9	8.0
C ₅ -430°F	41.2	51.9	50.0

4.1.4 Effect of HZSM-5 Source

The results described above for HZSM-5 were obtained using "ZCat Plus," which is Intercat Corporation's commercial FCC additive material. Two additional HZSM-5 type samples were tested to determine if the high yields of isobutylene and isoamylenes could be retained, but with lower propylene production. They were Davison's "Additive OH-S" and Intercat's "Isocat." A standard steam pre-treatment (100% steam at 1450°F for five hours) of these samples preceded the wax cracking tests. Table IX presents the results of these tests, all of which were obtained at the same process conditions of 880°F and 0.2 catalyst/oil ratio. Although there was considerable variability in the conversions, there were no major selectivity differences between the ZCat Plus and the other HZSM-5 catalysts; the ratio of propylene to isobutylene yields and the gasoline composition was similar for all samples.

4.1.5 HZSM-5 Zeolite Catalysts Containing Rare Earth

Table X shows the results of those tests with 1-3 % rare earth oxide added to HZSM-5 zeolite catalysts. At the same process conditions of 880°F and 0.2 catalyst/oil ratio, the addition of 1, 2, or 3% rare earth oxide to the base HZSM-5 catalyst (ZCat Plus) lowered conversion to about 80% but had no effect on selectivity.

Tests of these catalysts at 880°F and 0.9 catalyst to oil weight ratio showed that wax conversion level varied with the catalyst to oil ratio, as shown in Figure 17. The rare earth oxide treated HZSM-5 samples and the

FIGURE 13

**SASOL WAX MYU DATA:
EFFECT OF C/O ON COKE**

FIGURE 14

SASOL WAX MYU DATA: EFFECT OF CONVERSION ON AROMATICS FORMATION

AROMATICS, WT.% ON FEED

FIGURE 15

SASOL WAX MYU DATA: C5 ISOMER DISTRIBUTION

FIGURE 16

SASOL WAX MYU DATA: C4 ISOMER DISTRIBUTION

WEIGHT PERCENT

FIGURE 17

SASOL WAX MYU DATA: EFFECT OF C/O ON CONVERSION WITH HZSM-5 CATALYSTS

CONVERSION, WT%

TABLE IX
MYU TEST: VARIOUS HZSM-5 CATALYSTS
(TEST CONDITIONS: 880°F, 0.187 CATALYST/OIL RATIO)

Run No.	054	055	087	088	089	090	094	084	091	085	092	086	093
Catalyst Description	<-----CCC-1891-----> ZCat Plus Large Steamer					CCC-1891 ZCat Plus Small Steamer		14040-43-1 OH-S Small Steamer		14040-43-2 OH-S Small Steamer		14040-43-3 Isocat Small Steamer	
Conversion Wt%	90.2	91.9	92.5	86.5	86.9	78.5	90.5	84.7	87.8	83.9	89.8	65.9	69.4
Product Yields Wt%													
C ₂ -	1.6	1.5	2.8	1.6	2.2	1.2	2.1	1.9	1.7	1.7	3.0	0.9	0.8
C ₃ + C ₄	49.5	50.3	49.4	42.4	48.9	37.3	47.7	41.5	45.1	39.9	47.2	30.1	32.0
C ₅ -430°F	40.0	38.0	39.8	41.2	35.6	39.9	40.6	42.1	40.9	42.2	39.4	34.8	36.5
430°F+	8.8	10.0	7.5	13.5	13.1	21.1	9.5	15.6	12.2	16.1	9.3	34.1	30.6
Coke	0.1	0.2	0.5	1.3	0.2	0.1	0.1	0.2	0.1	0.1	0.2	0.1	0.1
C ₃ -	16.6	17.4	13.8	12.5	14.8	12.8	14.4	10.7	14.5	11.0	15.9	9.5	11.9
iC ₄ -	12.2	12.1	12.5	10.7	11.8	9.7	12.4	11.5	10.6	10.7	10.6	8.9	8.6
iC ₅ -	10.8	10.9	9.5	12.0	11.5	12.5	15.5	13.6	11.0	12.1	9.8	12.6	13.9
C ₅ -430°F													
RON	85.4	84.4	80.9	84.3	82.5	84.3	83.5	84.4	81.7	86.2	85.0	83.6	84.8
MON	76.4	76.0	74.2	75.1	74.8	74.9	75.2	75.0	74.2	75.7	76.5	74.3	75.3

WJR/lkv/94156
4/11/94

TABLE X

MYU TESTS: RARE EARTH/HZSM-5 CATALYSTS
Test Conditions: 880°F, 0.187 Catalyst/Oil Ratio

Run No.	096	097	098
Catalyst	14040-42-1	14040-42-2	14040-42-3
Description	1% ReO	2% ReO	3% ReO
Conversion, wt%	80.3	80.5	76.7
Product Yields, wt%:			
C ₂ -	1.2	1.3	1.1
C ₃ +C ₄	39.5	39.0	36.6
C ₅ -430°F	39.5	40.1	38.9
430°F ⁻¹	19.7	19.5	23.3
Coke	0.1	0.1	0.1
C ₃ =	13.3	12.7	11.8
iC ₄ =	9.9	9.5	9.5
iC ₅ =	10.1	10.8	10.1
C ₅ -430°F	83.3	83.7	83.3
RON	74.7	75.0	74.8
MON			

Isocat material were less active (lower wax conversion) than the other two commercial HZSM-5 samples, Intercat's ZCatPlus and Davison's OH-S. Catalytic activity is a function of the available Bronsted acid site concentration, or framework aluminum content. The lower catalytic activity of the Isocat sample is explained by its higher silica to alumina ratio (low framework alumina) HZSM-5 zeolite. Since the rare earth samples also showed lower activity, it is likely that the impregnation treatment lowered their acidity by lowering the framework aluminum content.

Figures 18-21 show there were no major selectivity differences in product yields of propylene, isobutylene, isoamylenes, and gasoline, respectively, among these HZSM-5 samples. The Isocat sample had lower propylene and isobutylene yields than the other HZSM-5 samples, but the differences were small. Figures 22 and 23 show that the total aromatic and olefin contents of the gasoline from the various catalysts varied with conversion. At >90% conversion levels, the olefins converted to aromatics.

This study of various HZSM-5 samples, with and without added rare earth, showed that there were differences in activity, but product selectivity changes were small and difficult to measure accurately.

4.1.6 Y Zeolite Catalysts Containing Rare Earth

A series of USY faujasite FCC catalyst samples with 0.3-1.5% rare earth was prepared and tested at two severity levels, catalyst to oil ratios of 0.75 and 0.1875 at 880°F. Table XI shows the results of those tests.

PROPYLENE, WT%

CONVERSION, WT%

OH-S#1 OH-S#2 ZCATPLUS ISOCAT 1%REO 2%REO 3%REO

SASOL WAX MYU DATA: EFFECT OF CONVERSION ON ISOBUTYLENE YIELD WITH HZSM-5 CATALYSTS

FIGURE 20

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON ISOAMYLENES YIELD WITH HZSM-5 CATALYSTS**

FIGURE 21

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON C5-430F NAPHTHA YIELD WITH HZSM-5 CATALYSTS**

FIGURE 22

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON AROMATICS FORMATION WITH HZSM-5 CATALYSTS**

FIGURE 23

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON OLEFIN YIELD WITH HZSM-5 CATALYSTS**

TABLE XI

MYU TESTS: RARE EARTH/USY CATALYSTS
 Test Conditions: 880°F, 0.75 Catalyst/Oil Ratio

Run No.	0.99	100	101	102	103
Catalyst Description	CCC-1701 ^(a)	CCC-1702 ^(a)	CCC-1703 ^(a)	CCC-1704 ^(a)	CCC-1705 ^(b)
Wt% Rare Earth	<0.04	0.27	0.34	0.96	1.49
Conversion, wt%	92.4	87.8	85.5	91.5	87.3
Product Yields, wt%:					
C ₂ -	0.8	0.6	0.6	0.8	0.7
C ₃ +C ₄	28.7	23.4	20.8	23.2	11.3
C ₅ -430°F	62.1	63.0	63.2	66.5	66.5
430°F ⁻¹	7.6	12.2	14.5	8.5	12.7
Coke	0.8	0.8	0.9	0.9	0.8
C ₃ =	7.6	6.6	5.8	6.5	5.3
iC ₄ =	5.9	4.5	4.1	4.3	3.8
iC ₅ =	7.8	6.4	5.6	6.4	5.9
RON	89.7	88.7	87.4	88.0	86.5
MON	78.0	76.3	74.5	76.0	74.0

(a) 100% steam; 1450°F; 8.5 hours

(b) 100% steam; 1450°F; 16 hours

(c) 100% steam; 1450°F; 8 hours

MMS/ml/94156
 4/11/94

Figures 24 summarizes the conversion data and Figures 25-31 summarize the selectivity data for the runs at both 0.75 and 0.1875 catalyst to oil ratio.

At 0.75 catalyst to oil ratio, in general, the addition of rare earth to conventional Y faujasite FCC catalysts lowered conversion, increased gasoline but lowered olefin yields, RON, and MON. These wax cracking test results with rare earth exchanged Y catalysts confirm the findings of other workers with regular gas oil feedstocks. There were wide variations in the wax conversion levels at 0.1875 catalyst to oil ratio that could not be resolved by running 4-6 tests on each of these samples. As Figures 18-24 show, there was too much scatter for differences between the catalysts to be distinguished. An attempt to resolve this difficulty was made by plotting selected points from the power law equations that were used to correlate the data for the various catalysts. Plots of this type for propylene, isobutylene, isoamylenes, and gasoline yield, in Figures 32-35, respectively, show that two of the high rare earth samples (1704, 1705) had the expected "rare earth effect" of higher gasoline and lower olefin yields. The other high rare earth sample (1706) had a product selectivity more like the low or no rare earth samples.

4.1.7 Effect of Coke Deposition

Although the coke yields from all the HZSM-5 catalysts were very low, <0.2 wt%, coke deposition is a major contributor to the rapid (but regenerable) catalyst deactivation in the fluid catalytic cracking process. The effects of coke deposition on activity and product selectivity were evaluated in a series of sequential MYU tests on the same coked HZSM-5 catalyst. This sequential experiment involved one HZSM-5 catalyst sample (CCC-1891) and five consecutive wax contact cracking tests. The catalyst sample remained in the reactor and the coke determination occurred only after the final wax cracking test. The test conditions were 880°F and 0.2 catalyst to oil ratio.

Figure 36 shows that wax conversion decreased with each subsequent test, which was expected due to the fouling of the active sites of the catalyst with coke. Note that the MYU conversion value for test #5 did not follow the conversion decrease and probably is in error.

Figures 37-41 show the effects of the multiple HZSM-5 wax cracking test run sequence on product selectivities for propylene, isobutylene, isoamylenes, gasoline, and C₃+C₄ gases, respectively. These figures show each of the "multiple" run product yield points and those of the other "individual" HZSM-5 catalyst runs. These results show that coke deposition had only a small impact on the wax cracking product selectivities of the HZSM-5 catalyst. This effect may be unique to the HZSM-5 type zeolite catalyst due to its medium pore geometry. Other zeolite catalysts may not respond to coke deposition in the same way.

4.1.8 Low Zeolite-content Catalysts

A series of FCC catalysts, with nominal zeolite concentrations of 10 and 40% of Y zeolite or 10% beta zeolite and a low activity matrix, was prepared and tested to further examine the relationships between type and amount of zeolite in the catalyst on wax conversion and selectivity. Table XII presents the nominal compositions of these catalyst samples, all of which were prepared using conventional raw materials and experimental

FIGURE 24

**SASOL WAX MYU DATA: EFFECT OF C/O ON CONVERSION
WITH RARE EARTH/USY CATALYSTS**

CONVERSION, WEIGHT PERCENT

FIGURE 25

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON PROPYLENE YIELD WITH RARE EARTH/USY CATALYSTS**

PROPYLENE, WT. %

FIGURE 26

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON PROPYLENE YIELD WITH RARE EARTH/USY CATALYSTS**

ISOBUTYLENE, WT.%

FIGURE 27

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON ISOAMYLENES YIELD WITH RARE EARTH/USY CATALYSTS**

ISOAMYLENES, WT.%

FIGURE 28

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON C5-430F YIELD WITH RARE EARTH/USY CATALYSTS**

C5-430oF, GASOLINE, WT. %

FIGURE 29

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON C3 AND C4 YIELD WITH RARE EARTH/USY CATALYSTS**

C3+C4 WEIGHT PERCENT

FIGURE 30

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON RON WITH RARE EARTH/USY CATALYSTS**

RESEARCH OCTANE, C5-430oF

FIGURE 31

**SASOL WAX MYU DATA: FISCHER-TROPSCH WAX
CATALYTIC CRACKING VARIABLE RARE EARTH Y CATALYSTS**

MOTOR OCTANE, C5-430oF

FIGURE 32

**SASOL WAX MYU DATA: POWER LAW PLOT
OF PROPYLENE YIELD VS. CONVERSION
WITH RARE EARTH/USY CATALYSTS**

FIGURE 33

**SASOL WAX MYU DATA: POWER LAW PLOT
OF ISOBUTYLENE YIELD
VS. CONVERSION WITH RARE EARTH/USY CATALYSTS**

FIGURE 34

SASOL WAX MYU DATA: POWER LAW PLOT
OF ISOAMYLENES YIELD
VS. CONVERSION WITH RARE EARTH/USY CATALYSTS

FIGURE 35

SASOL WAX MYU DATA: POWER LAW PLOT
OF C5-430F NAPHTHA YIELD
VS. CONVERSION WITH RARE EARTH/USY CATALYSTS

FIGURE 36

**SASOL WAX MYU DATA: EFFECT OF MULTIPLE RUNS
ON CONVERSION WITH HZSM-5 CATALYST**

FIGURE 37

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON PROPYLENE YIELD
IN MULTIPLE RUNS WITH HZSM-5 CATALYST**

FIGURE 38

SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON ISOBUTYLENE YIELD
IN MULTIPLE RUNS WITH HZSM-5 CATALYST

FIGURE 39

SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON ISOAMYLENES YIELD
IN MULTIPLE RUNS WITH HZSM-5 CATALYST

FIGURE 40

SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON C5-430F NAPHTHA YIELD
IN MULTIPLE RUNS WITH HZSM-5 CATALYST

FIGURE 41

SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON C3 AND C4 YIELD
IN MULTIPLE RUNS WITH HZSM-5 CATALYST

TABLE XII
CHEMICAL COMPOSITION OF ZEOLITE FCC CATALYSTS

Catalyst ID	Nominal Composition	Comment
9669-152	40% LZY-84 zeolite 20.8% Ludox AS-40 (silica sol) 4.2% Catapal alumina 35% Kaolin clay	Standard Y
9669-153	34.7% Ludox AS-40 (silica sol) 7% Catapal alumina 58.3% Kaolin clay	Matrix only
9669-154	10% LZY-84 20.8% Ludox AS-40 (silica sol) 4.2% Catapal alumina 65% Kaolin clay	Low Y zeolite
9669-155	10% Beta 20.8% Ludox AS-40 (silica sol) 4.2% Catapal alumina 65% Kaolin clay	Low Beta

techniques. The actual zeolite concentrations in the finished catalysts were not measured. The BET surface areas of these catalysts, after a steam treatment of 1450°F for 5 hours, are found in Table XIII. The surface areas and pore volumes decreased with the lower zeolite levels, as expected.

Runs were made with these catalysts using a catalyst to oil ratio of 0.75 and reaction temperatures of 970°F and 880°F. The conversions reported were calculated from the 430-°F fraction of the GC simulated distillation of the total liquid products. Figure 42 shows that the "SD" method gives a conversion value that is in reasonable agreement with conversion value that is obtained by the normal "MYU" method, which is the sum of product C- gas, gasoline and coke. The "SD" method also allowed calculation of the yields of light cycle oil (430-650°F) and heavy cycle oil (650+°F) products.

Figures 43-49 show that the yields of propylene, isobutylene, isoamylenes, gasoline, 430-650, 650-800, and 800+°F products, respectively, all increased with increasing conversion. No clear trends were seen that would distinguish between these catalysts.

4.2 LaPorte Wax Feedstock

Wax produced by the slurry phase Fischer-Tropsch process at the LaPorte plant currently contains over 2% catalyst contaminants, as Table II shows. That level of contamination is too high for it to be converted into transportation fuels using any fixed-bed process. However, it might be upgradable by the FCC process, which operates with a circulating catalyst inventory, if the F-T catalyst could be selectively removed from the FCC unit. These F-T fines will deposit on the external surfaces of the FCC catalyst microspheres. One possible removal method is to selectively attrit the F-T catalyst from the external surfaces of the FCC catalyst.

FIGURE 42

SASOL WAX MYU DATA: PARITY PLOT
OF MYU AND SIMULATED DISTILLATION METHODS
FOR CONVERSION CALCULATION

FIGURE 43

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON PROPYLENE YIELD
WITH LOW ZEOLITE CONTENT CATALYSTS**

a. Distinguishing Catalysts

b. Not Distinguishing Catalysts

FIGURE 44

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON ISOBUTYLENE YIELD
WITH LOW ZEOLITE CONTENT CATALYSTS**

a. Distinguishing Catalysts

ISOBUTYLENE, WT.%

b. Not Distinguishing Catalysts

ISOBUTYLENE, WT.%

FIGURE 45

SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON ISOAMYLENES YIELD
WITH LOW ZEOLITE CONTENT CATALYSTS

a. Distinguishing Catalysts

ISOAMYLENES, WT. %

b. Not Distinguishing Catalysts

ISOAMYLENES, WT. %

FIGURE 46

SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON C5-430F NAPHTHA YIELD
WITH LOW ZEOLITE CONTENT CATALYSTS

FIGURE 47

SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON LCCO YIELD
WITH LOW ZEOLITE CONTENT CATALYSTS

FIGURE 48

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON 650-800°F CYCLE OIL YIELD
WITH LOW ZEOLITE CONTENT CATALYSTS**

FIGURE 49

**SASOL WAX MYU DATA: EFFECT OF CONVERSION
ON 800+°F CYCLE OIL YIELD
WITH LOW ZEOLITE CONTENT CATALYSTS**

TABLE XIII

PHYSICAL PROPERTIES OF ZEOLITE FCC CATALYSTS
Steam Treatment: 1450°F, 5 hours, 100% steam

Catalyst ID	Description	BET Surface Area, m ² /g			Micropore Volume (cc/g)
		Total	Zeolite	Matrix	
14059-81-4	40% Beta	211	143	68	.066
14040-48-1	40% Y	253	183	70	.084
14040-48-2	40% Y	233	167	66	.077
14040-50-5	Matrix	55	3.0	52	.001
14040-50-6	10% Y	90	48	42	.022
14040-50-7	10% Beta	69	19	50	.009

This would be done with high velocity air jets in the regenerator of a commercial FCC unit. The fines would then be removed from the flue gas by conventional electrostatic precipitators or other collection devices. A series of laboratory experiments, which are described below, demonstrated the feasibility of the selective attrition concept. Further effort would have been required to optimize the selective attrition process, and no additional work was made to test this processing option under the present contract. The catalytic cracking properties of this LaPorte wax feedstock were measured and compared with those of the Sasol wax after it was determined that the FCC process could tolerate the high level of iron catalyst fines in the LaPorte wax.

4.2.1 Selective Attrition Experiments

F-T fines-contaminated FCC catalyst to be used in selective attrition experiments was prepared by performing ten individual cracking runs with the same commercial equilibrium USY FCC catalyst using LaPorte wax feedstock. The runs were made at 970°F and 5 catalyst to oil ratio. This wax cracking sequence resulted in a significant deposition of F-T catalyst fines from the wax onto the FCC catalyst. Table XIV, Part A shows that the iron content of the FCC catalyst increased from 0.42% to 1.05%. This F-T fines-contaminated catalyst was the feed for a laboratory attrition test. A high velocity air jet subjected this catalyst sample to severe attrition conditions. After the attrition test, the catalyst and fines were recovered and analyzed for contaminant metals.

Table XIV, Part B, No. 2 shows that the F-T catalyst fines were selectively attritted from the contaminated FCC catalyst into the fines. The iron content of the contaminated FCC catalyst decreased from 1.05% to 0.62% after the attrition experiment. The selective attrition technique, although technically feasible, would not be economically viable because a commercial F-T wax would not contain the high level of iron that was present in the LaPorte wax used in these experiments. The iron content of the fines generated in this experiment was nearly 3%. Table XIV, Part B, No. 1 shows the results of a control attrition experiment that used the base catalyst without F-T catalyst fines. The composition of the fines, especially the iron content, was similar to the starting catalyst, which

TABLE XIV
MYU TESTS: SELECTIVE ATTRITION EXPERIMENTS

	(ppm)				
	Si	K	Fe	Cu	Al
A. F-T catalyst fines deposition, chemical analyses of samples (from LaPorte wax):					
1. Base catalyst CCC-1397	226000	121	4200	26	258000
2. Treated catalyst: 5.0 catalyst, 10 g LaPorte wax in 1 g segments ID No. 9363005 (15586-030-2P)	225000	266	10500	610	231000
B. Selective attrition experiments:					
1. Control:					
a. Starting sample CCC-1397	226000	121	4200	26	258000
b. After attrition (15586-030-1B)	200000	108	4200	21	254000
c. Fines (15586-030-1F)	252000	204	4100	14	235000
2. F-T catalyst contaminator sample 9363005:					
a. Starting sample (15586-030-2F)	225000	266	10500	610	231000
b. After attrition (15586-030-2B)	212000	159	6200	289	267000
c. Fines (15586-030-2F)	219000	125	29600	2020	213000

suggests that no selective attrition occurred with the uncontaminated, control sample.

A brief scanning electron microscopy (SEM) study of the catalyst and fines samples from these selective attrition experiments supported the conclusions of the chemical analyses. The series of SEM photographs shown in Figures 50-53 illustrates the fate of the FCC catalyst in these laboratory tests. Figure 50 shows that the base FCC catalyst consisted of varied shaped microspheres with smooth, rounded surfaces, a morphology that is characteristic of the abrasive nature of the circulating catalyst inventory of a commercial FCC unit. Figures 51 and 52 show the composition of the FCC catalyst after the sequential wax cracking experiments. These figures show that the F-T catalyst fines - which are small bright spots, due to the high iron content - were deposited mainly on the external surfaces of the FCC catalyst microspheres. Figure 53 shows the FCC catalyst sample after the selective attrition experiment. Most of the external F-T catalyst fines were removed from the external surfaces of the FCC catalyst. Figures 54 and 55 show that the fines from the attrition experiment consisted largely of very small, high iron content fragments.

4.2.2 Catalyst Effects

Catalytic cracking tests of the LaPorte wax feedstock were made with USY, HZSM-5, and beta zeolite catalysts. Table XV presents the detailed results of the catalytic cracking tests with the LaPorte wax and new test runs with the Sasol wax. The type of FCC catalyst, but not feedstock, had a major impact upon product yields and quality. Both feedstocks gave similar conversions and yields of the major catalytic cracking products. However, the presence of the iron F-T fines in the LaPorte wax increased the coke and hydrogen gas yields. In addition, small increases occurred in the octane quality of the naphtha products from the LaPorte wax feedstock.

This series of MYU tests nominally used the same reaction temperature of 970°F. The scatter of conversion values for both feedstocks and catalyst combinations between 80-90% at catalyst to oil ratios of 0.9-1.0, shown in Figure 56, suggests that both feedstocks had similar conversion values. Further tests at a wider variety of catalyst to oil ratios would be required to verify this tentative conclusion.

Figures 57 to 66 present product yields versus conversion plots for the wax feedstock and catalyst combinations. Figures 57 and 58 show that the coke and hydrogen gas yields, respectively, were significantly higher for the LaPorte wax than the Sasol wax. The iron F-T catalyst fines in the LaPorte wax were responsible for the higher coke and dehydrogenation because the same result was obtained using Sasol wax feedstock with USY catalyst that had 1% iron Fischer-Tropsch catalyst fines deposited upon its surface (Table XIV, Run No. 21; from the selective attrition experiments). The HZSM-5 catalyst, CCC-1891, had a lower coke yield but similar hydrogen yield compared with the other two catalysts for the LaPorte wax tests. The intermediate pore structure of the HZSM-5 catalyst apparently inhibited coke formation even in the presence of the active dehydrogenation catalyst, iron F-T fines. Figures 59-65, respectively, show propylene, isobutylene, isoamylenes, gasoline, distillate, 650-800°F cycle oil, and 800+°F cycle oil selectivity plots for the two wax feedstocks and the three catalysts. The major differences in product

FIGURE 50

SEM PHOTOGRAPH
BASE CATALYST = CCC-1397
EQUILIBRIUM FCC CATALYST
IRON CONTENT = 4200 PPM

5586-230-1P

BSE

FIGURES 51, 52

SEM PHOTOGRAPHS
F-T CATALYST FINES CONTAMINATED FCC CATALYST
BEFORE SELECTIVE ATTRITION EXPERIMENT

ID NO. 15586-030-2P

IRON CONTENT = 10500 PPM

15586-030-2P, ESE

15586-030-2P, A-1

FIGURE 53

SEM PHOTOGRAPHS
F-T CATALYST FINES CONTAMINATED FCC CATALYST
AFTER SELECTIVE ATTRITION EXPERIMENT

ID NO. 15586-030-2B

IRON CONTENT = 6200 PPM

15586-030-2B BSE

FIGURES 54, 55

SEM PHOTOGRAPHS
CATALYST FROM SELECTIVE ATTRITION EXPERIMENT
ID NO. 15586-030-2F
IRON CONTENT = 29600 PPM

15586-030-2F BSE

15586-030-2F BSE A-1

FIGURE 56

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF C/O ON CONVERSION

CONVERSION, WT. %

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◇—	—*—	—□—	—◇—	—*—

FIGURE 57

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON COKE

COKE, WT.%

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◇—	—*—	—□—	—◇—	—*—

FIGURE 58

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON HYDROGEN YIELD

HYDROGEN, WT. %

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—●—	—*—	—□—	—△—	—×—

FIGURE 59

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON PROPYLENE YIELD

PROPYLENE, WT.%

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◆—	—*—	—□—	—◇—	—*—

FIGURE 60

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON ISOBUTYLENE YIELD

ISOBUTYLENE, WT.%

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◆—	—★—	—□—	—◇—	—✱—

FIGURE 61

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON ISOAMYLENES YIELD

ISOAMYLENES, WT.%

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◆—	—*—	—□—	—◇—	—*—

FIGURE 62

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON C5-430F NAPHTHA YIELD

C5-430oF, NAPHTHA, WT.%

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL

FIGURE 63

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON LCCO YIELD

430-650oF, WT.%

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◆—	—★—	—□—	—◇—	—*—

FIGURE 64

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON 650-800F CYCLE OIL YIELDS

650-800oF, WT.%

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◇—	—*—	—□—	—◇—	—*—

FIGURE 65

BOTTOMS SELECTIVITY SASOL AND LAPORTE WAX

800oF+, WT.%

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◆—	—★—	—□—	—◇—	—*—

FIGURE 66

LAPORTE VS SASOL WAX MYU DATA:
EFFECT OF CONVERSION ON RON

C5-430oF, RON

USY	HZSM5	BETA	USY	HZSM5	BETA
LAPORTE	LAPORTE	LAPORTE	SASOL	SASOL	SASOL
—■—	—◆—	—★—	—□—	—◇—	—✱—

TABLE XV

MYU TESTS: SASOL VS LAPORTE WAX FEEDSTOCKS
(Catalytic Cracking, 970F, 0.8 Catalyst/Oil)

Feed: LaPorte Wax				Weight Percent Product Yields									C ₃ -430°F		Gasoline Composition, Wt% on Feed				
Catalyst	Run #	*	Conv. Wt%	H ₂	C ₂ =	iC ₄ =	iC ₅ =	C ₅ -430°F	Coke	430-650°F	650-800°F	800°F+	RON	MON	P**	I**	A**	N**	O**
CCC 1397 (USY)	008	A	86.4	0.15	6.65	4.15	7.58	63.42	1.02	--	--	--	87.8	77.3	5.76	18.39	16.07	9.14	45.99
		B	85.6	--	6.66	4.33	7.72	64.88	1.03	11.40	2.65	0.37							
	017	A	84.9	0.16	7.01	3.67	6.17	62.87	1.32	--	--	--	87.8	77.2	5.30	18.6	19.9	10.9	41.8
		B	82.3	--	7.09	3.98	6.15	60.63	1.31	13.62	3.55	0.5							
CCC 1891 (HZSM-5)	006	A	81.8	0.19	20.55	9.14	6.15	28.45	0.75	--	--	--	92.0	80.1	10.17	5.30	28.79	5.67	42.7
		B	83.17	--	20.64	9.23	7.12	28.89	0.74	4.81	4.45	7.56							
	015	A	80.1	0.22	17.11	9.26	6.05	31.23	0.44	--	--	--	87.7	77.4	13.45	6.36	21.98	6.49	44.66
		B	81.7	--	17.0	9.29	6.72	31.89	0.45	6.39	4.91	6.99							
CCC 1875 (Beta)	007	A	88.2	0.16	9.6	7.69	8.83	57.18	1.19	--	--	--	88.2	76.4	6.66	10.55	11.73	9.56	54.24
		B	86.8	--	9.4	7.76	8.27	57.46	1.21	8.50	2.81	1.87							
	016	A	83.4	0.15	8.76	5.92	6.81	57.16	1.71	--	--	--	88.6	76.6	5.63	9.89	13.02	11.71	51.61
		B	79.4	--	8.72	6.00	7.19	53.08	1.73	13.34	4.18	3.1							
Feed: Sasol Wax																			
CCC 1397 (USY)	011	A	94.0	0.03	11.01	6.08	8.19	60.52	0.67	--	--	--	84.8	76.9	8.70	23.85	19.69	7.54	36.44
		B	93.01	--	11.11	6.18	6.87	60.42	0.68	6.18	.80	0							
	013	A	85.8	0.02	7.95	4.96	7.94	59.53	0.54	--	--	--	86.7	76.7	6.39	20.38	13.84	9.07	46.01
		B	81.0	--	7.97	5.13	8.14	57.22	0.54	10.47	5.67	2.81							

TABLE XV

MYU TESTS: SASOL VS LAPORTE WAX FEEDSTOCKS
(Catalytic Cracking, 970F, 0.8 Catalyst/Oil)
(Continued)

Feed: Sasol Wax				Weight Percent Product Yields									C ₃ -430°F		Gasoline Composition, Wt% on Feed				
Catalyst	Run #	*	Conv. Wt%	H ₂	C ₃ -	1C ₄ -	1C ₅ -	C ₃ -430°F	Coke	430-650°F	650-800°F	800°F+	RON	MON	P**	I**	A**	N**	O**
9363005*	021	A	86.2	0.16	9.89	4.16	4.57	53.11	2.60	--	--	--	86.4	78.7	--	--	--	--	--
		B	88.73	--	10.07	4.24	4.20	54.96	2.61	9.23	1.64	.397							
	022	A	75.7	0.02	9.03	4.43	3.49	49.38	0.95	--	--	--	85.3	76.5	6.93	20.48	25.52	12.37	31.76
		B	81.74	--	7.66	3.79	2.68	59.13	1.04	10.73	5.91	1.61							
	023	A	56.4	0.02	3.79	1.88	2.08	45.15	0.45	--	--	--	84.0	74.5	5.77	19.09	18.17	15.15	35.58
		B	62.2	--	3.68	1.95	4.40	51.54	.45	13.53	11.60	12.65							
	024	A	62.0	0.01	5.18	3.71	4.93	43.28	0.46	--	--	--	86.9	76.6	4.31	20.03	16.57	14.60	38.60
		B	66.42	--	5.29	4.05	6.58	48.09	.46	11.46	9.91	12.19							
	025	A	67.2	0.02	5.20	3.69	5.41	48.52	0.44	--	--	--	87.0	76.5	4.59	19.37	15.87	14.0	40.07
		B	66.1	--	5.33	3.96	6.76	47.61	0.45	12.44	10.03	11.42							
	026	A	78.2	0.02	6.34	4.42	6.70	56.5	0.43	--	--	--	87.0	76.4	5.03	20.41	12.86	11.43	43.9
		B	79.99	--	6.22	4.63	7.82	60.41	.435	9.597	6.49	3.91							
	027	A	61.6	0.01	4.65	2.78	4.02	46.88	0.38	--	--	--	86.0	75.5	4.66	19.44	15.64	14.19	39.73
		B	64.03	--	4.67	3.00	5.89	49.87	.38	13.34	10.67	11.97							
CCC 1891 (HZSM-5)	009	A	87.7		18.64	10.13	7.64	36.94	0.19	--	--	--	84.4	75.5	15.5	6.88	11.81	8.48	46.84
		B	90.35		18.56	10.13	7.85	39.87	.186	4.44	3.2	1.99							
	018	A	74.7		16.53	8.94	6.74	28.23	0.25	--	--	--	85.7	76.3	15.59	7.66	18.49	10.25	39.13
		B	79.86		16.58	9.01	6.96	33.00	.25	4.67	8.36	7.10							
CCC 1875 (Beta)	010	A	90.9		14.89	9.15	7.91	47.32	0.42	--	--	--	85.5	76.5	10.38	12.85	15.21	9.87	41.31
	5D	B	89.6		14.96	9.26	7.06	46.41	.44	6.29	2.70	1.38							
	012	A	79.8		10.78	7.59	7.72	46.25	0.38	--	--	--	86.0	75.8	8.85	13.37	12.47	11.60	48.08
	O	B	79.16		10.65	7.73	8.59	47.00	.40	8.39	7.11	5.33							
	019	A	57.2		6.74	4.19	3.95	37.04	0.42	--	--	--	84.6	75.8	7.98	14.68	19.34	15.85	33.16
		B	64.53		6.88	4.33	4.39	43.81	.436	13.05	10.16	12.25							

*A - MYU Conversion Calculation
B - Simulated Distillation Conversion Calculation

**P - paraffin
I - isoparaffin
A - aromatic
N - naphthene
O - olefin

MMS/lkv/93411
7/26/93

selectivities were from catalyst differences rather than feedstock differences. The HZSM-5 produced the highest yield of propylene, regardless of feedstock. The HZSM-5 and the beta zeolite catalysts had higher yields of isobutylene than the USY catalyst with both wax feedstocks. The higher light gas yields for the HZSM-5 and beta catalysts occurred at the expense of the naphtha (Figure 62) and distillate (Figure 63) yields. The 650°F+ cycle oil yields (Figures 64 and 65) did not vary significantly with wax feedstock or catalyst. Figure 66 shows that the RON of the naphtha products from the LaPorte wax feedstock were 1-2 octane numbers higher than the naphtha products from the Sasol wax. This may be another effect of the dehydrogenation activity of the iron F-T catalyst fines.

4.3 UOP Fischer-Tropsch Naphtha Feedstock

Reaction temperatures of 880 and 970°F were used with the following four catalysts: clay diluent only and a 6.3% blend of the Y, beta and HZSM-5 zeolite catalysts and diluent. The catalytic cracking of F-T naphtha gave low conversions, low light olefin yields, and resulted in high yields of water due to dehydration of oxygenates present in the naphtha. These results show that catalytic cracking is not a reasonable processing route for F-T naphtha.

The catalytic cracking runs at 880 and 970°F are summarized in Tables XVI and XVII, respectively. These tables show the distributions of the C₁ through C₅ hydrocarbons plus C₆+, which is the lumped total of products having carbon numbers of 6 and higher. The term "C5- conversion" used in Figures 67-69 is the sum of the C1-C5 hydrocarbon products (100 minus C₆+) in Tables XVI and XVII. Figure 67 shows that naphtha conversion to C₅- products was highest with the HZSM-5 catalyst. Figures 68 and 69, respectively, show that there were no selectivity differences between these USY, beta, and HZSM-5 catalysts for the production of either isobutylene or isoamylenes. The HZSM-5 catalyst gave the highest yields of isobutylene and isoamylenes products, but these increased yields of the desired olefins by the HZSM-5 catalyst were a function of its higher conversion rather than improved selectivity.

The liquid product consisted of two phases, water (and soluble, unconverted oxygenates) and hydrocarbons. The water originated from dehydration of the alcohols and aldehydes. The low water quantities in these MYU tests (~0.2 g) made quantitative analyses difficult. The detailed catalytic cracking behavior of this F-T gasoline is complex. Several of the important reactions are:

1. Dehydration of oxygenates to olefins and water.
2. Isomerization of 1-olefins to 2- or 3-olefins.
3. Cracking of olefins.
4. Cracking of normal paraffins.

5.0 PILOT PLANT TESTS (TASK 4)

Eleven catalytic cracking runs with the Sasol wax feedstock were made in the AU-2L pilot plant using a variety of catalysts and process conditions. The results of those runs are summarized in Table XVIII. Runs 939-1, 939-2, 939-4, and 945-2 used an equilibrium USY zeolite catalyst taken from one of Amoco's commercial FCC units. Run 939-5 used equilibrium USY zeolite catalyst that had been steamed to reduce its activity. Runs 940-1 and 940-2 used steamed 40% beta zeolite catalyst. Run 941-1 used a blend of 75% steamed equilibrium USY zeolite catalyst with 25% steamed HZSM-5 zeolite catalyst. Run 942-2 used a blend of 50% equilibrium USY zeolite catalyst with 50% diluent. Run 943-1 used steamed 10% USY zeolite catalyst. Run 944-2 used steamed 10% beta zeolite catalyst.

TABLE XVI
MYU TESTS: UOP F-T NAPHTHA FEEDSTOCK AT 880°F

Catalyst Type	Diluent Only	Zeolite Y	Zeolite Beta	Zeolite HZSM-5
Run No.	063	057	059	061
C/O	3.0	0.2	0.2	0.2
Product Yields, Wt%:				
Hydrocarbon	86	82	84	80
H ₂ O	14	18	16	20
Product Yields, HC Only, Wt%:				
H ₂	0.01	0.01	0.01	0.01
C ₁	0.07	0.04	0.04	0.03
C ₂	0.7	2.8	2.7	4.3
C ₂ ^o	0.31	0.2	0.2	0.1
C ₃	1.1	2.8	3.8	8.1
C ₃ ^o	0.13	0.1	0.1	0.1
C ₄	1.7	3.0	5.4	8.9
C ₄ ^o	0.3	0.1	0.3	0.2
C ₅	3.5	4.9	7.9	8.4
C ₅ ^o	0.4	0.4	0.5	0.4
C ₆ +	91.7	85.5	79.0	69.4
Coke	0.2	0.2	0.1	0.1
Isobutylene	0.58	0.50	1.6	3.0
Isoamylenes	0.24	1.23	2.9	4.5

WJR/ml/94156
4/11/94

TABLE XVII

MYU TESTS: UOP F-T NAPHTHA FEEDSTOCK AT 970°F

Catalyst Type	Zeolite Y		Zeolite Beta	Zeolite HZSM-5	
Run No.	068	069	070	071	072
Cat/Oil Ratio	0.2	0.2	0.2	0.2	0.1
Product Yields, Wt%:					
Hydrocarbon	85	83	85	84	86
Water	15	17	15	16	14
Product Yields, Hydrocarbon Only, Wt%:					
H ₂	0.02	0.03	0.04	0.03	0.04
C ₁	0.18	0.17	0.16	0.14	0.14
C ₂ ⁼	4.76	4.41	4.66	5.55	4.74
C ₂ ^o	0.43	0.43	0.46	0.28	0.36
C ₃ ⁼	4.28	4.13	6.44	11.75	8.80
C ₃ ^o	0.12	0.12	0.13	0.11	0.1
C ₄ ⁼	5.88	6.33	9.06	14.02	11.92
C ₄ ^o	0.25	0.26	0.28	0.14	0.25
C ₅ ⁼	8.81	9.54	12.13	14.91	14.06
C ₅ ^o	0.72	0.76	0.87	0.74	0.75
C ₆ ⁺	74.6	73.8	65.8	52.3	58.8
Coke	0.2	0.23	0.18	0.12	0.19
Isobutylene	1.22	1.31	3.15	5.61	3.93
Isoamylenes	3.04	3.32	5.84	8.93	7.63
RON	76.2	--	74.6	75.9	74.4
MON	69.0	70.3	68.3	69.4	68.6

WJR/lkv/94156
4/11/94

FIGURE 67

UOP F-T NAPHTHA MYU DATA:
EFFECT OF TEMPERATURE ON C5- CONVERSION

FIGURE 68

UOP F-T NAPHTHA MYU DATA: EFFECT OF C5- CONVERSION
ON ISOBUTYLENE YIELD

FIGURE 69

UOP F-T NAPHTHA MYU DATA: EFFECT OF C5- CONVERSION
ON ISOAMYLENES YIELD

ISOAMYLENES, WT.%

Y @ 880°F Y @ 970°F BETA @ 880°F
BETA @ 970°F ZSM-5 @ 880°F ZSM-5 @ 970°F

TABLE XVIII

FCC PILOT PLANT RUN DATA

Run No.	939-1	939-2	939-4	939-5	940-1	940-2	941-1	942-2	943-1	944-2	945-2
Run Conditions:											
Reaction Temp., F	944	932	882	879	934	910	965	937	890	897	888
c/o Ratio	5.16	4.06	2.29	2.25	5.08	3.35	2.84	1.57	3.08	3	2.57
WHSV, 1/hr	21	20.4	38.2	42.2	43.8	61.5	54.6	33.5	35.8	31.5	34.2
Catalyst	Eq. USY	Eq. USY	Eq. USY	Stm Eq. USY	Stm Beta	Stm Beta	75% Stm Eq USY 25% Stm HZSM-5	50% Eq USY 50% Diluent	Stm 10% USY	Stm 10% Beta	Eq. USY
Recovery, wt %	98.6	100.7	88.2	92.7	87.7	83.1	82.3	85.9	93.9	55.2	63
Conversion, vol %	93.5	93.7	83	85	96.6	96.5	89	90	89.3	89.7	84.9
Product Yields, wt %											
Hydrogen	0.04	0.04	0.03	0.02	0.02	0.01	0.02	0.02	0.01	0.01	0.03
Methane	0.36	0.33	0.18	0.14	0.1	0.07	0.09	0.16	0.11	0.13	0.23
Ethylene	0.48	0.42	0.26	0.21	0.66	0.5	1.01	0.34	0.22	0.31	0.33
Ethane	0.26	0.23	0.16	0.14	0.11	0.08	0.1	0.15	0.11	0.16	0.2
Propylene	9.26	8.2	7.28	6.28	13.93	13.68	16.03	8.94	6.63	10.94	8.57
Propane	1.86	1.67	1.07	0.9	2.11	1.81	2.47	1.25	0.93	1.27	1.48
Isobutane	7.93	7.23	4.15	3.4	9.04	7.66	3.4	5.02	3.97	4.14	5.59
n-Butane	2.08	1.84	1.21	0.99	2.58	2.09	1.92	1.3	0.99	1.16	1.17
i-Butene+1-Butene	7.38	6.7	7.68	6.75	12.46	12.99	12.95	7.67	6.25	9.62	5.18
t-2-Butene	4.19	3.74	3.69	3.19	5.65	5.44	5.33	3.72	2.27	3.5	2.92
c-2-Butene	3.12	2.77	2.67	2.31	4.16	3.98	3.76	2.68	2.04	2.44	2.14
Isopentane	8.49	8.65	3.38	3.35	5.11	3.73	2.16	3.97	3.66	1.93	3.18
n-Pentane	1.25	1.42	0.77	0.96	1.73	1.51	1.4	0.85	1.56	1.11	0.54
Pentenenes	7.37	9.5	6.3	10.54	10.15	10.24	12.47	7.27	11.45	9.65	4.83
C6-430 F Naphtha	37.11	38.98	46.89	46.08	27.34	31.53	24.81	45.36	48.53	43.59	47.15
430+ F Cycle Oils	6.47	6.27	16.69	14.06	3.59	3.67	11.6	10.42	10.12	9.63	14.65
Coke	2.34	2.01	0.61	0.68	1.2	1	0.47	0.88	0.55	0.43	1.81

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5.1 Conversion and Selectivity

Changes in the process variables of reaction temperature, catalyst to oil weight ratio, feed rate (WHSV), zeolite type, and catalyst activity had small effects on conversion. High conversions (83-97%) were always obtained. The activities of the USY zeolite catalysts were measured in relative microactivity tests (RMA), which is a standard measurement of FCC catalyst activity. The activities of the equilibrium USY catalyst, steamed equilibrium USY catalyst, and steamed 10% USY catalyst were 120, 85, and 77, respectively. The RMA test uses high sulfur heavy gas oil feedstock, and the value obtained with a beta zeolite catalyst cannot be compared with that of a USY zeolite catalyst. Beta zeolite catalysts have low RMAs but are very active with F-T wax feedstocks. Zeolite type and catalyst activity had a significant effect on product selectivity. Zeolites beta and HZSM-5 gave the higher yields of the target butenes and pentenes than zeolite Y. Zeolites beta and HZSM-5 also gave higher propylene but lower C₆-430°F naphtha yields than zeolite Y. Lowering the catalyst activity by steaming increased pentene and lowered coke yields. The steamed 10% beta zeolite catalyst with mild cracking conditions gave an excellent combination of olefin and naphtha yields.

Product recovery on this pilot plant is consistently over 98% with gas oil feedstock. In contrast, the product recoveries were between 55 and 101% with all the runs that were made with the wax feedstock. Feed rates with the wax were 23-25 g/m versus a standard gas oil feed rate of 13-15 g/m. The combination of wax feedstock and high feed rate flooded the condensing column, which caused the erratic, generally low recoveries in these runs. The cause of the low recoveries was not diagnosed until almost the end of the program. Elimination of this problem would have required a major revamp of the pilot plant which was not possible under the time limitations of this program. We believe that the low mass balances did not affect the conversion and selectivity data; the total liquid products that were recovered were representative of the process conditions and catalyst that were used.

The initial runs with equilibrium USY catalyst (runs 939-1 and 939-2) used process conditions similar to those normally used with petroleum gas oil feedstocks. The wax conversion level was about 94% with those conditions, confirming the high conversions that were observed in the small scale experiments. Lowering the severity of the process conditions (run 939-4) lowered the conversion to 83%, but additionally steaming the catalyst (run 939-5) while using the process conditions of run 939-4 gave essentially the same conversion (85%). The final run in this program, run 945-2, used the equilibrium USY catalyst with low process severity similar to run 949-4. There was excellent agreement in the conversion and product yields for runs 939-4 and 945-2, except for coke. Run 945-2 gave 1.8% coke, which was more in line with the 2-2.3% coke obtained in runs 39-1 and -2, than the 0.6% coke in run 939-4. A 50% dilution of the equilibrium Y zeolite catalyst with inert clay microsphere (run No. 942-2) did not lower conversion substantially. The use of added diluent to lower conversion was abandoned because the diluent had a higher particle density than normal FCC catalysts, which caused circulation problems in the pilot plant test of this catalyst mixture. One run (943-1) was also made with a steamed 10% USY zeolite catalyst using mild catalytic cracking severity. This run was made with the objective of using a less active catalyst plus mild cracking severity to get yield data at a lower conversion than was obtained in the run with steamed equilibrium USY (run 939-5). Lowering the activity had little effect; the conversion and product yields with the steamed 10% USY catalyst (RMA=77) were virtually identical to those with the steamed equilibrium USY catalyst (RMA=85) (run 939-5).

The two runs (940-1 and 940-2) with the steamed 40% beta zeolite catalyst confirmed the conclusions of the MYU tests: the beta zeolite converted a greater fraction of the wax feedstock to light olefins, including propylene, isobutylene and isoamylenes, than the steamed equilibrium USY zeolite catalyst; but the higher olefin yields were accompanied by a

significantly lower C₆-430°F naphtha yield (27-32% with steamed beta versus 46% with steamed equilibrium USY catalyst). One run (944-2) was made with steamed 10% beta zeolite catalyst using mild catalytic cracking process severity. This run was made with the objective of using a less active catalyst plus mild cracking severity to get yield data at a lower conversion than was obtained in the runs with steamed 40% beta (runs 940-1 and -2). Lowering the zeolite content lowered conversion--to about 90% versus 97% with 40% beta--and also had a beneficial effect on the product yields. High yields of olefins and propylene were obtained with the 10% beta catalyst, but with a higher C₆-430°F naphtha yield (44%) than the steamed 40% beta catalyst (27-32%).

The test results for the 25% steamed HZSM-5 catalyst mixture with steamed equilibrium USY (run 941-1) were similar to the steamed 40% beta zeolite. High yields of the target light olefins were produced, but gasoline yield was much lower than the Y zeolite catalyst. This trade-off of light olefin and gasoline yield can be adjusted by the amount of the HZSM-5 additive and the process conditions. Additional work would be needed to determine the optimal combination of HZSM-5 catalyst and process conditions.

5.2 Product Quality

The liquid products from runs 943-1 (steamed 10% USY catalyst), 939-05 (steamed equilibrium USY catalyst), 945-2 (equilibrium USY catalyst), and 944-2 (steamed 10% beta catalyst) were fractionated by an atmospheric distillation procedure (ASTM D-2892) to yield 430-°F and 430+°F fractions. Table XIX gives the following analyses for the IBP-430°F naphtha fractions: API, RON by gc, RON by engine test, MON by engine test, simulated distillation, and PIANO by gc. The 430+°F fractions were then distilled under vacuum (10mm Hg) to give 430-650°F distillate and 650+°F cycle oil fractions. Table XX lists the following properties of the 430-650°F distillates: API, pour point, cloud point, cetane index, C_A (by NMR), and simulated distillation. Table XXI gives the following properties for the 650+°F cycle oils: API, C_A (by NMR), and simulated distillation.

5.2.1 Properties of 430-°F Product (Naphtha)

The PIANO analyses show that the *distribution of compound types* and the *carbon number distribution* within the compound types were strongly affected by both catalyst activity and catalyst type. Figure 70 shows the effect of activity for USY zeolite on the distribution of n-paraffins, i-paraffins, i/n-paraffin ratio, naphthenes, aromatics, and olefins. With increasing activity (steamed 10% Y with 77 RMA < steamed equilibrium Y with 85 RMA < equilibrium Y with 120 RMA), olefin content decreased while naphthene and aromatics contents increased. Increasing activity had no

ANALYSIS OF IBP-430F PRODUCTS FROM FCC PILOT PLANT RUNS

Run ID:	943-1	939-5	945-2	944-2
Catalyst Type:	<u>Stm'd 10% USY</u>	<u>Stm'd Eq USY</u>	<u>Eq USY</u>	<u>Stm'd 10 % Beta</u>
API	68.5	63.6	58	63.5
RON, gc	84.3	81.6	81.3	81.1
RON	90.3	86.7	81.3	85.4
MON	79.4	78.1	74.9	76.4
(RON+MON)/2	84.8	82.4	78.1	80.9
<u>Sim Dist., F</u>				
IBP	72		-22	-25
5 %	104		136	91
10 %	136		159	133
20 %	151		196	162
30 %	162		216	203
40 %	189		245	223
50 %	206		274	253
60 %	235		292	279
70 %	268		326	302
80 %	307		350	337
90 %	362		387	384
95 %	394		409	413
FBP	446		455	475
<u>PIANO, Distribution of Compound Types, wt %</u>				
n-Paraffins	5.41	5.05	5.56	7.18
i-Paraffins	23.23	20.4	24.88	16.02
i/n	4.3	4.04	4.5	2.2
Naphthenes	8.97	13.2	15.98	15.49
Aromatics	8.86	11.57	22.57	8.9
Olefins	51.17	46.03	26.95	46.87
Oxygenates	0.24	0.08	0.26	0.46
Unknowns	2.08	3.68	3.8	5.08

TABLE XIX (Cont)

ANALYSIS OF IBP-430F PRODUCTS FROM FCC PILOT PLANT RUNS

Run ID:	943-1	939-5	945-2	944-2
Allyl Type:	<u>Stm'd 10 % USY</u>	<u>Stm'd Eq USY</u>	<u>Eq USY</u>	<u>Stm'd 10 % Beta</u>
<u>PIANO, Distribution by Carbon Number, wt %</u>				
Propane	0.02	0.04	0.02	0.03
Butanes	1.78	1.1	1.43	2.83
Pentanes	17.87	6	4.13	11.06
Hexanes	30.07	25.95	10.68	15.62
Heptanes	20.98	23.49	23.08	21.34
Octanes	11.79	17.2	22.27	19.08
Nonanes	7.7	10.86	17.02	12.52
Decanes	5.31	7.81	11.82	8.84
C11's	2.04	3.17	4.68	3.01
C12's	0.37	0.7	1.06	0.59
C13's	0	0	0	0
Unknowns	2.08	3.68	3.8	5.08
<u>PIANO, Compound Types by Carbon Number, wt %</u>				
<u>Paraffins:</u>				
C4	0.12	0.08	0.012	0.19
C5	0.9	0.28	0.22	0.63
C6	1.92	1.33	0.79	1.39
C7	0.7	0.85	1.02	1.16
C8	0.78	1.13	1.4	1.74
C9	0.35	0.49	0.69	0.81
C10	0.24	0.35	0.54	0.53
C11	0.23	0.3	0.45	0.44
C12	0.18	0.24	0.32	0.3
<u>Isoparaffins:</u>				
C4	0.25	0.14	0.3	0.29
C5	3	0.86	1.24	1.29
C6	8.06	5.64	3.31	2
C7	6.06	5.6	7.21	3.06
C8	2.85	3.37	5.94	3.18
C9	1.06	1.39	2.12	2.12
C10	1.28	2.24	3.08	2.77
C11	0.68	1.1	1.61	1.25
C12	0	0.06	0.06	0.06
<u>Aromatics:</u>				
C6	0	0	0.14	0
C7	0.39	0.46	0.97	0.25
C8	1.92	2.33	5.28	1.7
C9	2.63	3.29	6.84	2.53
C10	2.79	3.72	6.54	3.18
C11	0.94	1.39	2.14	1.03
C12	0.2	0.36	0.64	0.21

TABLE XIX (Cont)

Table II, concl.

ANALYSIS OF IBP-430F PRODUCTS FROM FCC PILOT PLANT RUNSAnalysis of IBP-430 F Products from AU-2L Pilot Plant Runs

Run ID:	943-1	939-5	945-2	944-2
Catalyst Type:	<u>Stm'd 10 % USY</u>	<u>Stm'd Eq USY</u>	<u>Eq USY</u>	<u>Stm'd 10 % Beta</u>
<u>PIANO, Compound Types by Carbon Number, wt %</u>				
<u>Naphthenes:</u>				
C6	0.45	0.4	0.3	0.18
C7	0.84	1.29	1.38	0.59
C8	3	4.36	5.06	5.46
C9	3.57	5.47	7.3	6.86
C10	0.98	1.45	1.62	2.31
C11	0.1	0.21	0.33	0.1
C12	0	0	0	0
<u>Olefins:</u>				
C4	1.33	0.81	0.97	2.29
C5	13.94	4.86	2.67	9.09
C6	19.63	18.58	6.15	12.06
C7	12.82	15.29	12.27	15.92
C8	3.24	6	4.6	7
C9	0.09	0.21	0.06	0.2
C10	0.02	0.04	0.03	0.06
C11	0.09	0.17	0.15	0.2
C12	0	0.03	0.03	0.02

ANALYSIS OF 430-650F PRODUCTS FROM FCC PILOT PLANT RUNS

Run ID:	943-1	939-5	945-2	944-2
Catalyst Type:	<u>Stm'd 10 % USY</u>	<u>Stm'd Eq USY</u>	<u>Eq USY</u>	<u>Stm'd 10 % Beta</u>
API	38.9	39.5	36.9	42.2
Pour Pt, F	0	0	-6	-6
Cloud Pt, F	1	2	-1	-1
Cetane Index	56.3	58.9	50.8	60.9
C-sub-A (NMR), wt %	19.4		24.8	15
<u>Sim Dist, F</u>				
IBP	403	395	400	393
5 %	434	438	428	431
10 %	443	453	440	444
20 %	466	478	455	467
30 %	486	501	475	485
40 %	511	525	489	503
50 %	534	550	515	521
60 %	559	576	532	542
70 %	585	602	553	562
80 %	613	627	576	586
90 %	642	651	615	624
95 %	651	661	647	649
FBP	691	694	733	741

ANALYSIS OF 650+F PRODUCTS FROM FCC PILOT PLANT RUNS

Run ID:	943-1	939-5	945-2	944-2
Catalyst Type:	<u>Stm'd 10 % USY</u>	<u>Stm'd Eq USY</u>	<u>Eq USY</u>	<u>Stm'd 10 % Beta</u>
API	32.8	37.1	29.6	36.8
C-sub-A (NMR), wt %	16		24.1	10.6
<u>Sim Dist, F</u>				
IBP	236	273	233	224
5 %	670	397	604	600
10 %	690	664	625	624
20 %	704	693	649	649
30 %	717	711	672	674
40 %	730	726	695	697
50 %	741	741	712	715
60 %	759	758	730	734
70 %	781	776	752	762
80 %	808	802	786	794
90 %	856	845	838	844
95 %	910	897	895	892
FBP	1189	1265	1132	1155

FIGURE 70

**SASOL WAX FCC PILOT PLANT DATA:
EFFECT OF USY ZEOLITE CATALYST ACIDITY
ON PIANO DISTRIBUTION BY COMPOUND TYPE IN IBP-430F NAPHTHA**

FIGURE 71

**SASOL WAX FCC PILOT PLANT DATA:
EFFECT OF USY ZEOLITE CATALYST ACTIVITY
ON PIANO DISTRIBUTION BY CARBON NUMBER IN IBP-430F NAPHTHA**

effect on either *n*- or *i*-paraffin content. The naphtha from the 10% steamed USY catalyst contained 9% naphthenes, 9% aromatics, 51% olefins, and had a 4.3 ratio of *i*- to *n*-paraffins. The naphtha from the standard equilibrium USY catalyst contained 16% naphthenes, 23% aromatics, 27% olefins and had a 4.5 ratio of *i*- to *n*-paraffins. Comparison of the steamed 10% USY catalyst with the steamed 10% beta catalyst shows that catalyst type had a strong effect on naphtha compound type. The 10% steamed beta catalyst produced more naphthenes (15%) and had a 2.2 ratio of *i*- to *n*-paraffins.

Figure 71 shows the effect of activity for USY zeolite on the C₅ through C₁₂ distributions. With increasing activity (steamed 10% Y with 77 RMA < steamed equilibrium Y with 85 RMA < equilibrium Y with 120 RMA), the distribution shifted to heavier products. Naphtha from the 10% steamed USY catalyst was much lighter (38% C₅-C₆s and 8% C₁₀-C₁₃s) than the naphtha made with the standard equilibrium USY catalyst (15% C₅-C₆s and 18% C₁₀-C₁₃s). Comparison of the steamed 10% USY catalyst with the steamed 10% beta catalyst shows that catalyst type had an effect on carbon number distribution. The naphtha from the steamed 10% USY catalyst was also lighter than the naphtha from the steamed 10% beta catalyst (27% C₅-C₆s and 12% C₁₀-C₁₃s).

These PIANO results show that the primary products from cracking over the steamed 10% USY catalyst are olefins distributed around C₆, but that they increasingly cyclize, aromatize, and disproportionate into higher molecular weight products over the more active steamed equilibrium USY and standard equilibrium USY catalysts. Both catalyst and process parameters should influence this aromatics formation reaction. Although further studies will be required to elucidate the mechanism of aromatics formation, these results suggest the following direction for future work: modification of the catalyst to lower activity, such as by the addition of phosphorus,⁽⁸⁾ should be a successful strategy for producing more olefins and decreasing the formation of naphthenes and aromatics.

The steamed 10% USY catalyst produced naphtha with an API gravity of 68.5, which was much higher than the gravity produced by either the steamed 10% beta or standard equilibrium USY catalysts, which were 63.5 and 58, respectively. These gravities reflect the PIANO analyses which showed that the steamed 10% USY made low naphthenes and low aromatics, the steamed 10% beta made high naphthenes and low aromatics, and the standard equilibrium USY made high naphthenes and high aromatics.

The engine tests of RON and MON showed that both decreased with increasing acidity for the USY catalysts. The steamed 10% USY catalyst produced naphtha with an (R+M)/2 octane number of 84.8, versus 82.4 for the steamed equilibrium USY and 78.1 for the equilibrium USY catalyst. The steamed 10% USY catalyst gave a higher octane than the steamed 10% beta catalyst (80.9). These results are also explained by the PIANO analyses. The high octane number of the naphtha from the steamed 10% USY catalyst is produced by its high content of low molecular weight olefins. The naphtha from the steamed 10% beta catalyst was lower in octane number because it had lower total olefins, the olefins were of higher molecular weight, and it had a high naphthene content. The low octane number of the naphtha from the standard equilibrium USY catalyst was surprising because, although it had the lowest olefin content, it had the highest content of aromatics plus olefins. Apparently the increase in molecular weight of the olefins and aromatics over those made by the steamed 10% USY catalyst, and the high content on naphthenes, combined to lower the octane of the naphtha that was made with the standard equilibrium USY catalyst.

5.2.2 Properties of 430-650 °F Product (LCCO)

Any 430-650°F fraction is "distillate" and is the boiling range for products such as diesel fuel, jet fuel, kerosene, and heating oil. The 430-650°F product from catalytic cracking is called "light cat cycle oil," which is abbreviated LCCO. The cetane index rating is the major property

of interest for diesel oil. Cetane index, which is calculated from the API gravity and simulated distillation, is proportional to the paraffin content; it is a measurement of diesel fuel quality analogous to octane number for gasoline. Weight percent C_A (NMR) is usually used as the indicator of aromatic carbon content, but is really the amount of sp_2 -hybridized carbon. Because these samples are expected to have high olefin contents, the C_A might not be proportional to the aromatic carbon content. Additional analysis would be required to distinguish the olefinic from the aromatic carbon.

Table XX shows that the 10% beta catalyst produced LCCO with the highest API gravity (42.2) and cetane index (60.9) and lowest weight percent C_A (NMR) (15) of the four catalysts.

The effect of activity with the three USY catalysts on LCCO properties was not as unambiguous as the effect on naphtha properties. The highest activity catalyst, the standard equilibrium USY catalyst, produced LCCO with the lowest API gravity (36.9) and cetane index (50.8) and highest weight percent C_A (NMR) (24.8). However, the LCCO from the intermediate activity catalyst (steamed equilibrium USY) had a higher API and cetane index than the LCCO from the steamed 10% USY catalyst, 39.5 vs 38.9 °API and 58.9 vs 56.3 cetane index, respectively. As expected, the C_A (NMR) for the steamed 10% USY catalyst (19.4 wt%) was lower than the C_A (NMR) from the equilibrium USY catalyst, but the analysis of the LCCO from the steamed equilibrium USY catalyst is not available to confirm the trend.

Although the steamed 10% beta catalyst made the highest cetane index product, the LCCOs from all four catalysts would be excellent stock for blending into diesel fuel.

The relative ranking of the LCCO from the standard equilibrium USY catalyst--heaviest and most aromatic--is consistent with the trends observed in the naphtha (discussed above). However, the relative rankings of the APIs of the LCCOs from the steamed 10% USY and steamed 10% beta catalysts are reversed from those of the naphtha. This implies that the LCCO fraction is not simply partially converted paraffinic feed, but that the LCCO contains products of condensation, cyclization, and aromatization. The lower API gravity and higher C_A (NMR) of the LCCO made with the steamed 10% USY catalyst versus the steamed 10% beta catalyst is consistent with those processes.

The pour points (0 to -6°F) and cloud points (2 to -1°F) of all four products were not significantly different.

5.2.3 Properties of 650+°F Product (Heavy Cycle Oil)

Table XXI shows that the 10% steamed beta catalyst produced the 650+°F cycle oil with the highest API gravity (36.8) and lowest weight percent C_A (NMR) (10.6) of the four catalysts. Consistent with the above-mentioned effect of activity for the USY catalysts, the standard equilibrium USY catalyst produced 650+°F cycle oil with the highest C_A (NMR) (24.1) and lowest API gravity (29.9). LCCO from the steamed 10% USY catalyst had 16 wt % C_A (NMR) and a 32.8 API gravity. The steamed equilibrium USY catalyst gave LCCO with the highest (37.1) API. This result is not understood; it may be erroneously high because of the presence of 5% 400-°F material that was not removed in the distillation. The observed trends in API gravity and C_A (NMR) are consistent with those of the LCCOs from the steamed 10% USY, equilibrium USY, and steamed 10% beta catalysts: more condensation, cyclization, and aromatization (high API and low C_A (NMR)) with either USY-based catalyst than the beta-based catalyst, and more aromatization with the standard equilibrium USY catalyst than the steamed 10% USY catalyst.

6.0 PREPARATION OF C₅-C₈ ETHERS (TASK 5)

Table XXII identifies the three feedstocks, A-C, that were prepared by combining the 200-°F distillate from FCC cracking runs. Tables XXIII-XXV show the major product analyses of these light naphtha/methanol runs with feedstocks "A," "B," and "C," respectively. Table XXVI summarizes the iso-olefin conversions for the three feedstocks with two catalysts, with and without the presence of hydrogen gas.

6.1 Model Compound Runs

Scoping runs were made using the model compound 2-methyl-2-butene with methanol to produce TAME before using the naphtha that was made in the pilot plant. The initial scoping runs were made in pilot plant AU-6. Those runs were repeated in pilot plant AU-109 because it was a smaller unit. The limited volume of 200-°F pilot plant product that was to be the feedstock for the etherification runs required the smaller pilot plant. Figure 72 shows that equivalent results were obtained in pilot plants AU-6 and AU-109. The test conditions were: variable temperatures between 125 and 200°F; 200 PSIG unit pressure; 0.66 WHSV (olefin); 1.2 mole ratio of methanol/olefin; and Amberlyst 15 catalyst (a commercial etherification catalyst). Figure 72 shows that the reaction temperature had a strong influence on the etherification reaction. The decrease in TAME yield at 200°F versus 150°F shows that the reaction is equilibrium limited with these process conditions at a reaction temperature of about 150°F. Additional experiments would be needed to determine the reaction temperature between 125 and 150°F where equilibrium control begins, or to establish the process conditions for equilibrium control below 150°F.

6.2 Feedstock Variation

The primary feedstock for these etherification studies was the light naphtha obtained from the pilot plant catalytic cracking runs with the Fischer-Tropsch wax. Table XXII identifies the three feedstocks, A-C, that were prepared by combining the 200-°F distillate from FCC cracking runs. The iso-olefin contents of feeds "B" and "C" were higher than feed "A." Feed "B" had a high iso-olefin content because it was made using the olefin selective beta and HZSM-5 zeolite FCC catalysts in the pilot plant runs, Nos. 940-01,02 and 941-01. It is not known why the iso-olefin content was higher in "C" than in "A" because the same Y zeolite catalyst was used in the runs for both of those feeds.

TABLE XXII

HYDROCARBON COMPOSITION OF 200°F- NAPHTHAS FROM FCC PILOT PLANT RUNS

Feed ID:	92-0490-01A Feed A*	93-0024-01A Feed B	93-0024-01C Feed C
Pilot Plant Run Nos.	939-01, + 02 eq. USY catalyst conversion = 93.6%	940-01, 02 941-01 Beta/HZSM-5 catalyst Conversions = 90,96%	939-04 eq. USY catalyst conversion = 83%
Total Paraffins wt%	6.69	8.44	4.32
C ₄	0.18	0.37	0.16
C ₅	0.93	1.04	0.72
C ₆	4.03	4.50	2.35
C ₇	1.45	1.67	1.02
C ₈	0.08	0.69	0.06
C ₉	--	0.13	--
Total Iso-paraffins wt%	42.71	17.64	22.88
C ₄	0.32	0.56	0.28
C ₅	3.77	2.03	2.19
C ₆	22.94	6.22	10.55
C ₇	13.87	6.12	8.86
C ₈	1.82	2.16	0.99
C ₉	--	0.44	--
Total Aromatics wt%	1.74	2.62	0.35
C ₆	0.34	0	0
C ₇	1.34	1.03	0.35
C ₈	0.05	1.38	0
C ₉	--	0.22	--
Total Naphthenes wt%	3.96	5.55	3.16
C ₅	0.05	0.06	0.05
C ₆	1.23	0.92	0.71
C ₇	1.92	1.71	1.63
C ₈	0.75	1.82	0.80
C ₉	--	1.03	--
Total Olefins wt%	44.51	64.47	68.65
C ₃	0.01	0.11	0.04
C ₄	1.25	3.72	1.82
C ₅	8.701	12.10	12.03
C ₆	23.88	29.54	33.79
C ₇	10.36	15.65	19.95
C ₈	0.31	3.31	1.02
Reactive iso-olefins wt%			
C ₅ 's			
2-methyl-1-butene	1.25	2.15	1.76
2-methyl-2-butene	4.26	5.67	5.64
C ₆ 's			
2,3-dimethylbutene	0.8	0.73	0.97
2-methyl-1-pentene	2.35	2.49	3.02
2-methyl-2-pentene	4.01	5.27	5.46
3-methyl-trans-2-pentene	2.49	3.13	3.29
3-methyl-cis-2-pentene	3.98	5.48	5.35

*This light naphtha was used in the initial etherification runs reported in the December 1992 Monthly Report.xxii

TABLE XXIII

ETHERIFICATION RUNS WITH NAPHTHA "A"
 (200 PSIG, AMBERLYST 15, 2.9 GRAMS)
 FEED RATE: METHANOL, 1.37 G/HR; 200F, - NAPHTHA, 5.5 G/HR)

Run No.	Feed + MeOH	024-2	024-3	024-4	024-6	024-7	024-8
Temp, F		125	125	125	150	150	150
Wt% Product:							
C4-5 Olefins:							
1C4=	0.22761	0.44	0.408	0.443	0.477	0.467	0.466
3M1BUTENE	0.10959	0.101	0.094	0.104	0.103	0.103	0.104
2M1BUTENE	1.05375	0.161	0.146	0.153	0.108	0.109	0.101
2M2BUTENE	3.59118	2.719	2.403	2.247	1.267	1.237	1.208
C6 Olefins:							
3M1PENTENE	0.79242	0.774	0.769	0.791	0.795	0.81	0.796
23DMBUTENE	0.69126	0.293	0.296	0.313	0.116	0.113	0.115
4Mt2PENTENE	0.30348	0.334	0.326	0.352	0.341	0.361	0.353
4Mc2PENTENE	1.07904	1.048	1.047	1.087	1.072	1.094	1.083
2M1PENTENE	1.98105	0.657	0.667	0.701	0.253	0.25	0.245
HEXENE-1	0.6744	0.659	0.664	0.682	0.671	0.684	0.675
tHEXENE-3	1.46682	1.443	1.454	1.488	1.474	1.497	1.481
cHEXENE-3	2.28453	2.252	2.271	2.316	2.291	2.331	2.3
2M2PENTENE	3.38043	2.636	2.668	2.731	1.733	1.724	1.75
tHEXENE-2	0.06744	0.063	0.064	0.067	0.07	0.071	0.067
3Mt2PENTENE	2.09907	1.615	1.635	1.677	1.278	1.299	1.285
cHEXENE-2	1.22235	1.209	1.22	1.238	1.23	1.247	1.232
3Mc2PENTENE	3.35514	2.617	2.654	2.694	2.329	2.402	2.331
Oxygenates:							
MeOH	15.7	10.586	10.553	9.614	9.953	8.593	9.535
MTBE	0	0.362	0.379	0.396	0.407	0.401	0.394
TAME	0	4.193	4.264	4.027	5.271	5.356	5.299
THME1	0	0.557	0.563	0.556	0.978	1.002	0.981
THME2	0	2.904	2.931	2.929	4.771	4.937	4.815
THME3	0	2.341	2.361	2.353	3.361	3.397	3.42
nHEXANE	3.39729	4.792	4.708	4.207	3.552	3.53	3.449
TOLUENE	1.127091	1.132	1.154	1.16	1.146	1.165	1.154
2MPENTANE	9.9474	9.726	9.773	10.027	9.968	10.163	10.064
3MPENTANE	6.909228	6.877	6.899	7.039	6.96	7.095	7.011

MMS/ml/94156
 2/11/94

6.3 Etherification Runs

The above-mentioned light naphtha samples and methanol were the feedstock for a series of etherification runs at the same process conditions that were used for the earlier 2-methyl-2-butene/methanol tests. The reaction temperature was 150°F, except for a few runs at 125°F with feed "A." Tables XXIII-XXV show the major product analyses of these light naphtha/methanol runs with feedstocks "A," "B," and "C," respectively. The mixed ether product consisted of TAME and the three C₆ ethers, THME, (tertiary hexyl methyl ethers), which are 2-methyl-2-methoxypentane, 2,3-dimethyl-2-methoxybutane and 3-methyl-3-methoxypentane. Each run represents a weight balance period of at least 6 hours. There was good agreement in the product analyses for each test condition, except for the variability in the methanol analyses. In addition, the methanol and isobutylene components were not completely separated on the capillary GC column, a problem that did not allow for the quantification of the production of MTBE from the reaction of methanol and the minor amount of isobutylene in the naphtha sample. Table XXIII shows that higher conversions were obtained at 150°F than at 125°F reaction temperature, in agreement with the pure component 2-methyl-2-butene tests.

Different olefins have different reactivities and give different conversions during the etherification process. The conversion of each isomer can be calculated from the concentrations of feed and product that are given for each of the runs in Tables XXIII-XXV. Using data from Table XXIII, the effect of isomer and temperature on olefin reactivity is illustrated below in Table XXIIIa.

TABLE XXIIIa

REACTIVE ISO-OLEFINS CONVERSION TO ETHERS

Iso-olefin Component	125°F	150°F
C ₅	Wt%	Wt%
2-Methyl-1-butene	85.4	89.9
2-Methyl-2-butene	29.9	65.5
C ₆		
2,3 Dimethyl-1-butene	56.5	83.4
2-Methyl-1-pentene	65.9	87.5
2-Methyl-2-pentene	20.7	48.6
3-Methyl-cis-2-pentene	21.7	38.6
3-methyl-trans-2-pentene	20.8	29.8

The reaction products from the runs with feedstock "A" had a significant yellow color, especially at the higher reaction temperature of 150°F. The products from the pure component 2-methyl-1-butene tests were colorless. It is likely that polymerization of olefins to C₁₀+ hydrocarbon "color bodies" was responsible for the undesirable colored product. The color of the ether product could be a significant product quality issue. A possible solution would be to distill the ether product to remove the high molecular weight color bodies. However, the THME ethers boil at the high end of the product gasoline fraction, so such a separation might not be feasible. Another possible way to reduce color formation would be to use a catalyst that contained a hydrogenation metal component and hydrogen gas in the reactor.

To reduce color formation, runs were made with feedstocks "B" and "C" using Bayer's commercial etherification catalyst, K2634, in addition to Amberlyst 15. The Bayer catalyst contains a noble metal in addition to

TABLE XXIV

ETHERIFICATION RUNS WITH NAPHTHA "B"
(200 PSIG, METHANOL 1.37 G/HR, 200°F- NAPHTHA 5.5 G/HR)

Run No.	Feed+MeOH	031-2	032-1	033-1	033-2	033-3	033-4
Temp		150°F	150°F	150°F	150°F	150°F	150°F
Catalyst		Amber15	Amber15	K2634	K2634	K2634	K2634
Product, Wt%							H2 added
C4-5							
Olefins:							
Isobutylene							
iC4=	1.036047	0.405	0.585	0.441	0.567	0.374	0.468
3M1BUTENE	0.225081	0.192	0.267	0.203	0.261	0.149	0.143
2M1BUTENE	1.81245	0.141	0.199	0.153	0.215	0.335	0.414
2M2BUTENE	4.778967	1.464	2.016	1.358	1.906	1.581	2.731
C6 OLEFINS							
3M1PENTENE	0.689574	0.66	0.767	0.6	0.748	0.423	0.429
23DMBUTENE	0.612861	0.512	0	0.038	0.048	0.03	0.083
4Mt2PENTENE	0.30348	0.293	0.368	0.289	0.363	0.221	0.262
4Mc2PENTENE	1.041105	0.995	1.159	0.912	1.136	0.69	0.929
2M1PENTENE	2.099913	0.276	0.329	0.239	0.335	0.424	0.583
HEXENE1	0.649953	0.622	0.696	0.544	0.68	0.368	0.387
tHEXENE3	1.620246	1.551	1.718	1.358	1.695	1.011	1.298
cHEXENE3	2.695914	2.569	2.633	2.086	2.601	1.594	2.207
2M2PENTENE	4.440081	1.81	2.08	1.497	2.008	1.46	2.478
tHEXENE2	0.06744	0.114	0	0.067	0.067	0.07	0.071
3Mt2PENTENE	2.640276	1.542	1.893	1.398	1.851	1.208	1.957
cHEXENE2	1.422141	1.359	1.482	1.169	1.468	0.863	1.04
3Mc2PENTENE	4.621326	2.969	3.378	2.436	3.289	2.127	3.521
OXYGENATES:							
MEOH	15.7	15.3	17.148	39.317	17.586	49.195	22.108
MTBE	0	1.558	2.031	1.459	1.991	0.927	1.835
TAME	0	6.17	6.328	4.452	6.338	2.853	4.644
THME1	0	1.172	0.67	0.466	0.677	0.386	0.52
THME2	0	5.535	4.369	2.866	4.397	2.45	3.542
THME3	0	4.142	3.003	2.148	3.113	1.782	2.376
NON-REACTIVE COMPOUNDS:							
nHEXANE	3.796872	3.25	3.598	2.883	3.484	2.48	4.653
TOLUENE	0.865761	0.851	0.746	0.499	0.739	0.461	0.714
2MPENTANE	3.153663	3.271	3.796	2.867	3.538	2.228	3.7
3MPENTANE	1.608444	1.746	1.972	1.455	1.794	1.121	1.878

WJF/ml/94156;4/11/94

TABLE XXV

ETHERIFICATION RUNS WITH NAPHTHA "c"
(200 PSIG, METHANOL 1.37 G/HR, 200°F- NAPHTHA 5.5 G/HR)

Run No.	Feed+MeOH	034-1	034-2	034-3	034-4
Temp		150°F	150°F	150°F	150°F
MEOHRATE		K2634	K2634	K2634	K2634
GASORATE		H2 added	H2 added	no H2	no H2
Product: C4-5 OL:					
Isobutylene iC4=	0.427401	0.3	0.206	0.435	0.3
3M1BUTENE	0.15174	0.019	0.008	0.111	0.159
2M1BUTENE	1.485366	0.348	0.302	0.213	0.151
2M2BUTENE	4.752834	3.095	2.778	2.094	1.595
C6 OLEFINS:					
3M1PENTENE	1.012443	0.073	0.052	0.709	1.012
23DMBUTENE	0.820239	0.115	0.115	0.115	0.115
4Mt2PENTENE	0.418971	0.145	0.123	0.458	0.46
4Mc2PENTENE	1.353858	0.751	0.651	1.481	1.404
2M1PENTENE	2.54586	0.54	0.548	0.351	0.318
HEXENE1	0.96945	0.067	0.048	0.407	0.957
tHEXENE3	2.128575	0.891	0.786	2.009	2.167
cHEXENE3	3.339966	1.884	1.692	3.589	3.43
2M2PENTENE	4.606152	3.087	3.219	2.281	2.187
tHEXENE2	0.06744	0.067	0	0.067	0.067
3Mt2PENTENE	2.774313	2.185	2.215	1.868	1.834
cHEXENE2	1.797276	0.588	0.518	1.609	1.817
3Mc2PENTENE	4.509207	4.125	4.237	3.399	3.29
OXYGENATES:					
MEOH	15.7	15.88	11.402	13.721	10.7
MTBE	0	1.318	1.117	1.082	1.049
TAME	0	4.555	4.854	5.838	6.819
THME1	0	0.494	0.597	0.679	0.893
THME2	0	3.876	4.932	5.012	6.114
THME3	0	2.087	2.589	3.366	4.166
NON-REACTIVE COMPOUNDS:					
nHEXANE	1.978521	7.371	7.847	2.943	2.151
TOLUENE	0.292521	0.375	0.335	0.269	0.325
2MPETANE	4.517637	5.961	6.157	5.253	4.731
3MPETANE	3.250608	3.691	3.973	3.571	3.316

TABLE XXVI
SUMMARY OF ETHERIFICATION RUN DATA

	Run No.	Reaction Temp	Catalyst	Research Octane Number*	Paraffins	Iso- paraffins	Aromatics	Naphthenes	Olefins	Oxygenates	Unknowns
Feed A	92-0490-01A			80.92	6.689	42.712	1.736	3.956	44.507	0.071	0.33
	15586-024-2	125°F	Amberlyst 15	82.09	7.804	40.230	1.985	4.474	33.762	11.406	0.34
	15586-024-6	150°F	Amberlyst 15	83.76	6.463	40.429	2.247	4.576	29.671	16.294	0.32
	15586-024-8	150°F	Amberlyst 15	83.88	6.337	40.466	2.263	4.586	29.58	16.436	0.33
Feed B	93-0024-01A			83.12	8.437	17.637	2.623	5.549	64.472	0.17	1.11
	15586-031-2	150°F	Amberlyst 15	87.43	7.417	17.989	3.687	6.424	41.847	21.815	0.821
	15586-033-1	150°F	Bayer K2634	87.48	7.381	17.340	3.691	6.583	41.716	22.445	0.844
	15586-033-3	150°F	Bayer K2634	85.78	8.205	17.668	3.62	6.512	45.889	17.277	0.83
Feed C	93-0024-01C			84.56	4.315	22.881	0.353	3.161	68.651	0.15	0.49
	15586-034-1	150°F	Bayer H2 K2634	79.47	14.535	24.733	2.405	5.436	35.369	17.332	0.19
	15586-034-3	150°F	Bayer no H2 K2634	85.78	5.921	22.489	1.977	4.497	43.51	21.192	0.415

*Calculated

WR/lkv/94156
4/11/94

FIGURE 72

**EFFECT OF REACTION TEMPERATURE ON ETHERIFICATION
OF 2-METHYL-2-BUTENE WITH METHANOL**

the strong acid functionality. The noble metal is available for olefin isomerization and diolefin saturation, in the presence of hydrogen.

When hydrogen gas was present with the Bayer catalyst (run No. 034-1, Table XXV), the color of the product improved, but there was a major loss of iso-olefin conversion because of concurrent hydrogenation of both reactive iso-olefins and linear olefins. This was an undesirable result since both the production of ethers and the octane number of the product decreased significantly. As Table XXVI shows, this run with added hydrogen gas had a lower research octane number (79.5) than the feedstock (84.6) or run No. 034-3 which was made without added hydrogen (85.8). This octane loss was due to the conversion of high octane olefins to low octane paraffins.

Table XXVI, which summarizes the etherification data at 150°F where the reaction was under equilibrium control, shows that iso-olefin conversions were similar for the three feedstocks and two catalysts, and in the absence of hydrogen gas, the light naphtha fractions from the catalytic cracking of Fischer-Tropsch wax were excellent feedstocks for the synthesis of TAME and THME. Table XXVI shows that the calculated research octane values for the products of these etherification runs were 2-4 numbers higher than the starting light naphtha feedstocks. As expected, this octane increase depended to some extent upon the concentrations of the ethers in the product.

7.0 EVALUATION OF GASOLINE BLENDING PROPERTIES OF ETHERS (TASK 6)

The previous section described RON data that were calculated from gas chromatography analyses of the mixed ether product from methanol etherification of the 200-°F fraction from the pilot plant catalytic cracking runs of Fischer-Tropsch wax feedstock. Measured RON and Motor Octane Numbers (MON) were obtained on those mixed ethers products in engine tests using 10% blends of the mixed ether product with unleaded regular gasoline (ULR). The octane number from the engine test of the blend is the observed octane number. The blend octane number is the octane number of the ether component. The blend octane number of the ethers was calculated from the observed octane number of the blend by using Equation 1, which assumes that the volumes of ethers and unleaded regular gasoline blend linearly. These RON and MON data are given in Table XXVII, which also gives the RON data shown in Table XXVI that were calculated from gas chromatography analyses.

Equation 1:

$$\text{Observed Octane} = (\text{Vol. Fr. Additive})(\text{Blend Value Additive}) + (\text{Vol. Fr. ULR})(\text{Obs. Octane ULR})$$

where Vol. Fr. ULR = 0.90, RON ULR = 91.9, and MON ULR = 82.3

The calculated RON values for the products of the etherification runs were 2-4 numbers higher than the starting light naphtha feedstocks. As expected, this octane increase depended to some extent upon the concentrations of the ethers in the product. However, as Figure 73 shows, there was only fair agreement between the measured blending RON and the RON calculated by gas chromatography, even after editing the data to exclude outliers from the engine tests. Figure 73 excluded the data for additive 92-0490-1A because the MON was higher than the RON, which is impossible. We believe these results reflect engine test measurement errors because of very small amounts of samples available for analysis. Although the measurement of octane numbers in engine tests is usually very accurate, the data we obtained on the mixed ether blends is extremely sensitive to small errors because only enough etherified product was available for a single evaluation as a 10% solution, and the octane numbers of the unleaded regular blend stock were very similar to those of the ether blends.

8.0 ECONOMIC EVALUATIONS (TASK 7)

8.1 Product Value of FCC Pilot Plant Runs

8.1.1 Propylene Valued as Fuel Gas

Product values were calculated for eight pilot plant runs (939-1, -2, -4, -5, 940-1, -2, 941-1, and 941-2) using the product yields given in Table XVIII plus the product yields of reactive (for the production of methyl ethers) C₅ and C₆ olefin isomers that are shown in Table XXVIII. Tables XXIX-XXXVI show the results of the calculations for the eight runs, respectively. The rate basis for all the analyses was 283,687 lb/hr, which is consistent with the F-T wax rate used in the "Baseline Design/Economics for Advanced Fischer-Tropsch Technology."⁽⁹⁾ Net product values (which accounts for the external energy required to maintain heat balance) were calculated for both simple (no ether unit) and complex (contains ether unit) refinery configurations. The price structure was the 1989 average spot price, PAD III obtained from the National Petroleum Council. All the iso-olefins were valued 100% as etherification feedstocks although they are only partially converted to ethers. Table XXXVII summarizes the net product values for simple and complex refineries, and the difference between the two, for the data in Tables XXIX-XXXVI.

Table XXXVII shows that irrespective of catalyst or reaction conditions, the net product values (\$/d) for a simple refinery, which ranged from about \$555,500 to \$584,500, were always lower than the net product values for a complex refinery, which ranged from about \$605,600 to \$653,300. The net product values from the complex refinery configuration were higher because isobutylene was valued at 85.4 cpq as MTBE unit vs 63.8 cpq as alkylation product and the isoamylenes and isohexenes were valued at ~ 69 cpq as ethers vs 59.2 cpq as gasoline. Additional experimental data would be required to select the best catalyst if the products were valued for a simple refinery configuration. The HZSM-5 and beta zeolite catalysts would be better catalysts than USY zeolite catalyst in a complex refinery configuration because of their higher yields of iso-olefins. The complex refinery values of the beta and HZSM-5 catalyst runs were about \$630,000-653,000/d vs \$606,000-626,000/d for the USY catalyst runs.

8.1.2 Propylene Valued as Feedstock to Di-isopropyl Ether Unit

The product values in Tables XXVIII-XXXVI were recalculated with propylene valued at 42.7 cpq as feedstock to a diisopropyl ether unit⁽¹⁰⁾ (DIPE) rather than 17.1 cpq as fuel gas. Table XXXVIII compares the product values for a complex refinery that were calculated with the propylene valued as feedstock to a DIPE unit rather than as fuel gas. The propylene value was computed using the properties⁽¹⁰⁾ shown in Table XXXIX. Gasoline product values were calculated based on octane, RVP, and unit operating costs. The complete economic evaluation is shown in Appendix B.

Valuing propylene as DIPE feedstock raised the value of the FCU products by between 4% and 10.2%, depending on yield. The net product values from the HZSM-5 and beta zeolite catalysts increased by 8.4-10.2% (to \$694,000-708,000/d) vs 4-6.1% (to \$635,000-653,000/d) for the USY catalysts. The net product values from the HZSM-5 and beta zeolite catalysts were additionally increased over the Y zeolite catalyst because high yields of propylene accompanied the high yields of iso-olefins.

FIGURE 73

ENGINE TEST RON VS CALCULATED RON FOR MIXED ETHER REACTION PRODUCTS

GC Calculated RON of Ethers

TABLE XXVII

RON AND MON OF PRODUCTS OBTAINED FROM LIGHT NAPHTHA ETHERIFICATION RUNS

Engine Test Measurements of RON and MON for
Products Obtained from Light Naphtha
Etherification Runs

	Run No.	Reaction Temp.	Catalyst	Olefins	Oxygenates	GC Calc. Research Octane Number *	Blend Value RON	Blend Value MON	Observed RON*	Observed MON*
Feed A	92-0490-01A			44.507	0.071	80.92	66.9	67.3	89.4	80.8
	15586-024-2	125°F	Amberlyst 15	33.762	11.406	82.09	77.9	74.3	90.5	81.5
	15586-024-6	150°F	Amberlyst 15	29.671	16.294	83.76	79.9	75.3	90.7	81.6
	15586-024-8	150°F	Amberlyst 15	29.58	16.436	83.88	--	--	--	--
Feed B	93-0024-01A			64.472	0.17	83.12	83.9	--	91.1	--
	15586-031-2	150°F	Amberlyst 15	41.847	21.815	87.43	87.9	78.3	91.5	81.9
	15586-033-1	150°F	Bayer K2634	41.716	22.445	87.48	85.9	79.3	91.3	82.0
	15586-033-3	150°F	Bayer K2634	45.889	17.277	85.78	84.9	74.3	91.2	81.5
Feed C	93-0024-01C			68.651	0.15	84.56	95.9	80.3	92.3	82.1
	15586-034-1	150°F	Bayer H2 K2634	35.369	17.332	79.47	88.9	81.3	91.6	82.2
	15586-034-3	150°F	Bayer no H2 K2634	43.51	21.192	85.78	95.9	85.3	92.3	82.6

*Observed Octane = (Volume Fraction Additive) (Blend Value Additive) + (Volume Fraction Unleaded Regular Gasoline) (Observed Octane Unleaded Regular Gasoline)

Volume fraction unleaded regular gasoline = 0.90
 RON unleaded regular gasoline = 91.9
 MON unleaded regular gasoline = 82.3

MMS/lkv/94156
 4/11/94

TABLE XXVIII

C4-C6 ISOMER DISTRIBUTION OF FCC PILOT PLANT PRODUCTS

RUN NO.	939-1	939-2	939-4	939-5	940-1	940-2	941-1	942-2
CATALYST	eq Y	eq Y	eq Y	eq Y	BETA	BETA	Y+ZSM	Y BLD
CONVERSION, WT%	93.5	93.7	83	85	96.6	96.5	89	90.5
COMPONENT, WT%								
HYDROGEN	0.04	0.04	0.03	0.02	0.02	0.01	0.02	0.02
METHANE	0.36	0.33	0.18	0.14	0.1	0.07	0.09	0.16
ETHYLENE	0.48	0.42	0.26	0.21	0.66	0.5	1.01	0.34
ETHANE	0.26	0.23	0.16	0.14	0.11	0.08	0.1	0.15
PROPYLENE	9.26	8.2	7.28	6.28	13.93	13.68	16.03	8.94
PROPANE	1.86	1.67	1.07	0.9	2.11	1.81	2.47	1.25
i-BUTANE	7.93	7.23	4.15	3.4	9.04	7.66	3.4	5.02
n-BUTANE	2.08	1.84	1.21	0.99	2.58	2.09	1.92	1.3
1-BUTENE	1.47	1.32	1.4	1.22	2.22	2.24	2.2	1.41
i-BUTYLENE	5.91	5.38	6.28	5.53	10.24	10.75	10.75	6.26
t-2-BUTENE	4.19	3.74	3.69	3.19	5.65	5.44	5.33	3.72
c-2-BUTENE	3.12	2.77	2.67	2.31	4.16	3.98	3.76	2.68
i-PENTANE	8.49	8.65	3.38	3.35	5.11	3.73	2.16	3.97
n-PENTANE	1.25	1.42	0.77	0.96	1.73	1.51	1.4	0.85
3M-1-BUTENE	0.086	0.166	0.092	0.168	0.187	0.214	0.279	0.126
2M-1-BUTENE	0.993	1.47	0.936	1.516	1.68	1.77	2.3	1.13
2M-2-BUTENE	3.78	4.57	2.99	5.077	5.03	4.896	5.94	3.36
1-PENTENE	0.297	0.47	0.312	0.505	0.476	0.505	0.62	0.378
t-2-PENTENE	1.4	1.8	1.248	2.069	1.768	1.826	2.145	1.45
c-2-PENTENE	0.806	1.01	0.7166	1.203	1	1.029	1.18	0.819
2,3-DM-1-BUTENE	0.187	0.173	0.33	0.369	0.193	0.147	0.085	0.22
2M-1-PENTENE	0.48	0.525	0.7177	0.803	0.579	0.507	0.266	0.57
2M-2-PENTENE	0.812	0.919	1.287	1.419	1.065	1.078	0.682	1.136
c-3M-2-PENTENE	0.812	0.923	1.306	1.419	1.078	1.169	0.84	1.21
t-3M-2-PENTENE	0.52	0.588	0.822	0.907	0.689	0.693	0.448	0.733
C6-430oF *	34.29	35.85	39.43	41.16	23.74	27.94	22.49	41.49
430-650oF	4.97	5.12	10.08	8.95	2.72	2.96	7.47	7.47
650oF+	1.5	1.15	6.61	5.1	0.87	0.71	4.13	2.95
SUBTOTAL	97.642	97.976	99.4046	99.308	98.731	98.99	99.514	99.113
COKE	2.34	2.01	0.61	0.68	1.2	1	0.47	0.88
GRAND TOTAL	99.982	99.986	100.0146	99.988	99.931	99.99	99.984	99.993

*C₆ fraction contains paraffins and non-ether-forming olefins.

TABLE XXIX

ECONOMIC ANALYSIS OF FCC PILOT PLANT RUN 939-1: USY CATALYST

Fischer Tropesch Wax Economics
Pilot Plant Results of Wax Run Through an FCU
03/19/93

Rate Basis: 283,657 lb/hr
Run No. 939-1

Component	normalized wt % Yield	lb/hr	BBL/Day	<---Simple Configuration--->				<---Complex Configuration--->			
				cpg	\$/Day	Valued as		cpg	\$/Day	Valued as	
Hydrogen	0.040	113	111	6.0	279	Fuel Gas	2	6.0	279	Fuel Gas	2
Methane	0.360	1,021	233	10.8	1,059	Fuel Gas	2	10.8	1,059	Fuel Gas	2
Ethylene	0.480	1,362	252	12.8	1,356	Fuel Gas	2	12.8	1,356	Fuel Gas	2
Ethane	0.260	738	142	12.1	721	Fuel Gas	2	12.1	721	Fuel Gas	2
Propylene	9.263	26,274	3,450	17.1	24,777	Fuel Gas	2	17.1	24,777	Fuel Gas	2
Propane	1.861	5,277	713	16.8	5,028	Fuel Gas	2	16.8	5,028	Fuel Gas	2
1-Butane	7.932	22,500	2,739	37.2	42,792	Alkylation	3	37.2	42,792	Alkylation	3
n-Butane	2.081	5,902	692	29.8	8,662	Gasoline	5	29.8	8,662	Gasoline	5
1-Butene	1.470	4,171	475	63.9	12,756	Alkylation	3	63.9	12,756	Alkylation	3
1-Butylene	5.912	16,769	1,914	63.8	51,289	Alkylation	3	85.4	68,654	Ether Unit	4
1-2-Butene	4.191	11,888	1,336	64.8	36,349	Alkylation	3	64.8	36,349	Alkylation	3
c-2-Butene	3.121	8,852	968	66.6	27,065	Alkylation	3	66.6	27,065	Alkylation	3
1-Pentane	8.492	24,089	2,652	49.4	55,017	Gasoline	5	49.4	55,017	Gasoline	5
n-Pentane	1.250	3,547	385	33.8	5,469	Gasoline	5	33.8	5,469	Gasoline	5
3M-1-Butene	0.086	244	27	50.3	560	Gasoline	5	53.6	597	Alkylation	3
2M-1-Butene	0.993	2,817	294	50.3	6,216	Gasoline	5	86.5	10,689	Ether Unit	4
2M-2-Butene	3.781	10,725	1,101	50.3	23,268	Gasoline	5	88.0	40,707	Ether Unit	4
1-Pentene	0.297	843	89	50.3	1,889	Gasoline	5	54.9	2,062	Alkylation	3
1-2-Pentene	1.400	3,972	417	50.3	8,803	Gasoline	5	55.5	9,713	Alkylation	3
c-2-Pentene	0.806	2,287	240	50.3	5,068	Gasoline	5	55.5	5,592	Alkylation	3
2,3-diM-1-Butene	0.187	531	53	59.2	1,326	Gasoline	5	68.6	1,537	Ether Unit	4
2-M-1-Pentene	0.480	1,362	137	59.2	3,395	Gasoline	5	68.7	3,940	Ether Unit	4
2-M-2-Pentene	0.812	2,304	229	59.2	5,689	Gasoline	5	69.4	6,669	Ether Unit	4
c-3-M-2-Pentene	0.812	2,304	227	59.2	5,634	Gasoline	5	70.1	6,671	Ether Unit	4
1-3-M-2-Pentene	0.520	1,475	144	59.2	3,586	Gasoline	5	70.5	4,270	Ether Unit	4
C6-430	34.299	97,292	8,416	59.7	211,018	Gasoline	5	59.7	211,018	Gasoline	5
430-650	4.971	14,102	1,019	52.1	22,296	Diesel	6	52.1	22,296	Diesel	6
650+	1.500	4,256	275	31.8	3,666	No 6 FO	7	31.8	3,666	No 6 FO	7
Sub-total	97.659	277,018	28,728		575,033				619,411		
Coke	2.341	6,639									
Grand Total	100.000	283,657									
Coke Amount for Heat Balance, wt % lb/hr		5% 14,929									
Coke Deficit, lb/hr		8,290									
MMBTU/Day		3,382									
\$/Day		6,765			(6,765)				(6,765)		
Net \$/Day					568,269				612,646		

TABLE XXX

ECONOMIC ANALYSIS OF FCC PILOT PLANT RUN 939-2: USY CATALYST

Fischer Tropsch Wax Economics

Rate Basis: 283,657 lb/hr

Pilot Plant Results of Wax Run Through an FCU

Run No. 939-2

03/19/93

Component	normalized wt % Yield	lb/hr	BBL/Day	<---Simple Configuration--->				<---Complex Configuration--->			
				cpg	\$/Day	Valued as		cpg	\$/Day	Valued as	
Hydrogen	0.040	113	111	6.0	279	Fuel Gas	2	6.0	279	Fuel Gas	2
Methane	0.330	936	214	10.8	970	Fuel Gas	2	10.8	970	Fuel Gas	2
Ethylene	0.420	1,192	221	12.8	1,187	Fuel Gas	2	12.8	1,187	Fuel Gas	2
Ethane	0.230	653	125	12.1	638	Fuel Gas	2	12.1	638	Fuel Gas	2
Propylene	8.201	23,264	3,055	17.1	21,939	Fuel Gas	2	17.1	21,939	Fuel Gas	2
Propane	1.670	4,738	640	16.8	4,514	Fuel Gas	2	16.8	4,514	Fuel Gas	2
1-Butane	7.231	20,512	2,497	37.2	39,010	Alkylation	3	37.2	39,010	Alkylation	3
n-Butane	1.840	5,220	612	29.8	7,661	Gasoline	5	29.8	7,661	Gasoline	5
1-Butene	1.320	3,745	427	63.9	11,454	Alkylation	3	63.9	11,454	Alkylation	3
1-Butylene	5.381	15,263	1,742	63.8	46,685	Alkylation	3	85.4	62,490	Ether Unit	4
1-2-Butene	3.741	10,610	1,192	64.8	32,442	Alkylation	3	64.8	32,442	Alkylation	3
c-2-Butene	2.770	7,859	859	66.6	24,026	Alkylation	3	66.6	24,026	Alkylation	3
1-Pentane	8.651	24,540	2,701	49.4	56,048	Gasoline	5	49.4	56,048	Gasoline	5
n-Pentane	1.420	4,029	438	33.8	6,212	Gasoline	5	33.8	6,212	Gasoline	5
3M-1-Butene	0.166	471	51	50.3	1,081	Gasoline	5	53.6	1,151	Alkylation	3
2M-1-Butene	1.470	4,170	435	50.3	9,200	Gasoline	5	86.5	15,822	Ether Unit	4
2M-2-Butene	4.571	12,965	1,331	50.3	28,127	Gasoline	5	88.0	49,209	Ether Unit	4
1-Pentene	0.470	1,333	141	50.3	2,989	Gasoline	5	54.9	3,262	Alkylation	3
1-2-Pentene	1.800	5,107	536	50.3	11,317	Gasoline	5	55.5	12,487	Alkylation	3
c-2-Pentene	1.010	2,865	301	50.3	6,350	Gasoline	5	55.5	7,007	Alkylation	3
2,3-diM-1-Butene	0.173	491	49	59.2	1,227	Gasoline	5	68.6	1,421	Ether Unit	4
2-M-1-Pentene	0.525	1,489	149	59.2	3,713	Gasoline	5	68.7	4,308	Ether Unit	4
2-M-2-Pentene	0.919	2,607	259	59.2	6,438	Gasoline	5	69.4	7,547	Ether Unit	4
c-3-M-2-Pentene	0.923	2,619	258	59.2	6,404	Gasoline	5	70.1	7,583	Ether Unit	4
1-3-M-2-Pentene	0.588	1,668	163	59.2	4,054	Gasoline	5	70.5	4,828	Ether Unit	4
C6-430	35.856	101,707	8,798	59.7	220,594	Gasoline	5	59.7	220,594	Gasoline	5
430-650	5.121	14,526	1,050	52.1	22,967	Diesel	6	52.1	22,967	Diesel	6
650+	1.150	3,263	210	31.8	2,811	No 6 FO	7	31.8	2,811	No 6 FO	7
Sub-total	97.990	277,955	28,565		580,334				629,866		
Coke	2.010	5,702									
Grand Total	100.000	283,657									
Coke Amount for Heat Balance, wt % lb/hr		5% 14,929									
Coke Deficit, lb/hr		9,227									
MMBTU/Day		3,765									
\$/Day		7,529			(7,529)				(7,529)		
Net \$/Day					572,805				622,337		

TABLE XXXI

ECONOMIC ANALYSIS OF FCC PILOT PLANT RUN 939-4: USY CATALYST

Fischer Tropsch Wax Economics

Rate Basis: 283,657 lb/hr

Pilot Plant Results of Wax Run Through an FCU

Run No. 939-4

03/19/93

Component	normalized wt % Yield	lb/hr	BBL/Day	<---Simple Configuration--->			Valued as	<---Complex Configuration--->			Valued as
				cpg	\$/Day			cpg	\$/Day		
Hydrogen	0.030	85	83	6.0	209	Fuel Gas	2	6.0	209	Fuel Gas	2
Methane	0.180	510	117	10.8	529	Fuel Gas	2	10.8	529	Fuel Gas	2
Ethylene	0.260	737	137	12.8	734	Fuel Gas	2	12.8	734	Fuel Gas	2
Ethane	0.160	454	87	12.1	443	Fuel Gas	2	12.1	443	Fuel Gas	2
Propylene	7.279	20,647	2,711	17.1	19,471	Fuel Gas	2	17.1	19,471	Fuel Gas	2
Propane	1.070	3,035	410	16.8	2,891	Fuel Gas	2	16.8	2,891	Fuel Gas	2
i-Butane	4.149	11,770	1,433	37.2	22,384	Alkylation	3	37.2	22,384	Alkylation	3
n-Butane	1.210	3,432	402	29.8	5,036	Gasoline	5	29.8	5,036	Gasoline	5
1-Butene	1.400	3,971	452	63.9	12,144	Alkylation	3	63.9	12,144	Alkylation	3
1-Butylene	6.279	17,811	2,033	63.8	54,476	Alkylation	3	85.4	72,920	Ether Unit	4
1-2-Butene	3.689	10,465	1,176	64.8	31,997	Alkylation	3	64.8	31,997	Alkylation	3
c-2-Butene	2.670	7,572	828	66.6	23,151	Alkylation	3	66.6	23,151	Alkylation	3
1-Pentane	3.379	9,586	1,055	49.4	21,893	Gasoline	5	49.4	21,893	Gasoline	5
n-Pentane	0.770	2,184	237	33.8	3,367	Gasoline	5	33.8	3,367	Gasoline	5
3M-1-Butene	0.092	261	28	50.3	599	Gasoline	5	53.6	638	Alkylation	3
2M-1-Butene	0.936	2,655	277	50.3	5,856	Gasoline	5	86.5	10,071	Ether Unit	4
2M-2-Butene	2.989	8,480	871	50.3	18,397	Gasoline	5	88.0	32,185	Ether Unit	4
1-Pentene	0.312	885	94	50.3	1,983	Gasoline	5	54.9	2,165	Alkylation	3
1-2-Pentene	1.248	3,539	371	50.3	7,844	Gasoline	5	55.5	8,655	Alkylation	3
c-2-Pentene	0.716	2,032	213	50.3	4,504	Gasoline	5	55.5	4,970	Alkylation	3
2,3-diM-1-Butene	0.330	936	94	59.2	2,339	Gasoline	5	68.6	2,710	Ether Unit	4
2-M-1-Pentene	0.718	2,035	204	59.2	5,074	Gasoline	5	68.7	5,888	Ether Unit	4
2-M-2-Pentene	1.287	3,650	362	59.2	9,013	Gasoline	5	69.4	10,566	Ether Unit	4
c-3-M-2-Pentene	1.306	3,704	364	59.2	9,058	Gasoline	5	70.1	10,725	Ether Unit	4
1-3-M-2-Pentene	0.822	2,331	228	59.2	5,666	Gasoline	5	70.5	6,747	Ether Unit	4
C6-430	39.423	111,827	9,673	59.7	242,542	Gasoline	5	59.7	242,542	Gasoline	5
430-650	10.078	28,588	2,066	52.1	45,201	Diesel	6	52.1	45,201	Diesel	6
650+	6.609	18,746	1,209	31.8	16,149	No 6 FO	7	31.8	16,149	No 6 FO	7
Sub-total	99.390	281,927	27,216		572,952				616,383		
Coke	0.610	1,730									
Grand Total	100.000	283,657									
Coke Amount for Heat Balance, wt % lb/hr		5% 14,929									
Coke Deficit, lb/hr MMBTU/Day \$/Day		13,199 5,385 10,771			(10,771)				(10,771)		
Net \$/Day					562,181				605,612		

TABLE XXXII

ECONOMIC ANALYSIS OF FCC PILOT PLANT RUN 939-5: USY CATALYST

Fischer Tropsch Wax Economics

Rate Basis: 283,657 lb_{min}

Pilot Plant Results of Wax Run Through an FCU

Run No. 939-5

03/19/93

Component	normalized wt % Yield	lb/hr	BBL/Day	<---Simple Configuration--->			Valued as	<---Complex Configuration--->			Valued as
				cpg	\$/Day			cpg	\$/Day		
Hydrogen	0.020	57	55	6.0	140	Fuel Gas	2	6.0	140	Fuel Gas	2
Methane	0.140	397	91	10.8	412	Fuel Gas	2	10.8	412	Fuel Gas	2
Ethylene	0.210	596	110	12.8	593	Fuel Gas	2	12.8	593	Fuel Gas	2
Ethane	0.140	397	76	12.1	388	Fuel Gas	2	12.1	388	Fuel Gas	2
Propylene	6.281	17,816	2,339	17.1	16,802	Fuel Gas	2	17.1	16,802	Fuel Gas	2
Propane	0.900	2,553	345	16.8	2,432	Fuel Gas	2	16.8	2,432	Fuel Gas	2
1-Butane	3.401	9,646	1,174	37.2	18,345	Alkylation	3	37.2	18,345	Alkylation	3
n-Butane	0.990	2,809	329	29.8	4,122	Gasoline	5	29.8	4,122	Gasoline	5
1-Butene	1.220	3,461	394	63.9	10,586	Alkylation	3	63.9	10,586	Alkylation	3
1-Butylene	5.531	15,689	1,791	63.8	47,986	Alkylation	3	63.8	47,986	Alkylation	3
1-2-Butene	3.190	9,050	1,017	64.8	27,671	Alkylation	3	64.8	27,671	Alkylation	3
c-2-Butene	2.310	6,553	716	66.6	20,036	Alkylation	3	66.6	20,036	Alkylation	3
1-Pentane	3.351	9,504	1,046	49.4	21,706	Gasoline	5	49.4	21,706	Gasoline	5
n-Pentane	0.960	2,724	296	33.8	4,199	Gasoline	5	33.8	4,199	Gasoline	5
3M-1-Butene	0.168	477	52	50.3	1,094	Gasoline	5	53.6	1,165	Alkylation	3
2M-1-Butene	1.516	4,301	449	50.3	9,488	Gasoline	5	86.5	16,316	Ether Unit	4
2M-2-Butene	5.078	14,403	1,479	50.3	31,247	Gasoline	5	88.0	54,667	Ether Unit	4
1-Pentene	0.505	1,433	152	50.3	3,211	Gasoline	5	54.9	3,505	Alkylation	3
1-2-Pentene	2.069	5,870	616	50.3	13,008	Gasoline	5	55.5	14,353	Alkylation	3
c-2-Pentene	1.203	3,413	358	50.3	7,564	Gasoline	5	55.5	8,346	Alkylation	3
2,3-diM-1-Butene	0.369	1,047	105	59.2	2,616	Gasoline	5	68.6	3,032	Ether Unit	4
2-M-1-Pentene	0.803	2,278	228	59.2	5,678	Gasoline	5	68.7	6,590	Ether Unit	4
2-M-2-Pentene	1.419	4,026	400	59.2	9,940	Gasoline	5	69.4	11,653	Ether Unit	4
c-3-M-2-Pentene	1.419	4,026	396	59.2	9,845	Gasoline	5	70.1	11,657	Ether Unit	4
1-3-M-2-Pentene	0.907	2,573	252	59.2	6,254	Gasoline	5	70.5	7,447	Ether Unit	4
C6-430	41.166	116,771	10,101	59.7	253,266	Gasoline	5	59.7	253,266	Gasoline	5
430-650	8.951	25,391	1,835	52.1	40,147	Diesel	6	52.1	40,147	Diesel	6
650+	5.101	14,469	933	31.8	12,464	No 6 FO	7	31.8	12,464	No 6 FO	7
Sub-total	99.320	281,728	27,136		581,240				636,272		
Coke	0.680	1,929									
Grand Total	100.000	283,657									
Coke Amount for Heat Balance, wt % lb/hr		5% 14,929									
Coke Deficit, lb/hr MMBTU/Day \$/Day		13,000 5,304 10,608			(10,608)				(10,608)		

Net \$/Day

570,631

625,664

TABLE XXXIII

ECONOMIC ANALYSIS OF FCC PILOT PLANT RUN 940-1: STEAMED BETA CATALYST

Fischer Tropsch Wax Economics

Rate Basis: 283,657 lb/hr

Pilot Plant Results of Wax Run Through an FCU

Run No. 940-1

03/19/93

Component	normalized wt % Yield	lb/hr	BBL/Day	<---Simple Configuration--->				<---Complex Configuration--->			
				cpg	\$/Day	Valued as		cpg	\$/Day	Valued as	
Hydrogen	0.020	57	55	6.0	140	Fuel Gas	2	6.0	140	Fuel Gas	2
Methane	0.100	284	65	10.8	294	Fuel Gas	2	10.8	294	Fuel Gas	2
Ethylene	0.660	1,873	347	12.8	1,866	Fuel Gas	2	12.8	1,866	Fuel Gas	2
Ethane	0.110	312	60	12.1	305	Fuel Gas	2	12.1	305	Fuel Gas	2
Propylene	13.939	39,539	5,192	17.1	37,287	Fuel Gas	2	17.1	37,287	Fuel Gas	2
Propane	2.111	5,989	809	16.8	5,706	Fuel Gas	2	16.8	5,706	Fuel Gas	2
1-Butane	9.046	25,659	3,123	37.2	48,800	Alkylation	3	37.2	48,800	Alkylation	3
n-Butane	2.582	7,323	859	29.8	10,748	Gasoline	5	29.8	10,748	Gasoline	5
1-Butene	2.221	6,301	718	63.9	19,272	Alkylation	3	63.9	19,272	Alkylation	3
1-Butylene	10.247	29,065	3,318	63.8	88,901	Alkylation	3	85.4	118,999	Ether Unit	4
1-2-Butene	5.654	16,037	1,802	64.8	49,034	Alkylation	3	64.8	49,034	Alkylation	3
c-2-Butene	4.163	11,808	1,291	66.6	36,100	Alkylation	3	66.6	36,100	Alkylation	3
1-Pentane	5.113	14,504	1,597	49.4	33,127	Gasoline	5	49.4	33,127	Gasoline	5
n-Pentane	1.731	4,910	533	33.8	7,571	Gasoline	5	33.8	7,571	Gasoline	5
3M-1-Butene	0.187	531	58	50.3	1,218	Gasoline	5	53.6	1,298	Alkylation	3
2M-1-Butene	1.681	4,769	498	50.3	10,520	Gasoline	5	86.5	18,091	Ether Unit	4
2M-2-Butene	5.033	14,277	1,466	50.3	30,974	Gasoline	5	88.0	54,188	Ether Unit	4
1-Pentene	0.476	1,351	143	50.3	3,028	Gasoline	5	54.9	3,305	Alkylation	3
1-2-Pentene	1.769	5,018	526	50.3	11,122	Gasoline	5	55.5	12,271	Alkylation	3
c-2-Pentene	1.001	2,838	298	50.3	6,290	Gasoline	5	55.5	6,941	Alkylation	3
2,3-diM-1-Butene	0.193	548	55	59.2	1,369	Gasoline	5	68.6	1,586	Ether Unit	4
2-M-1-Pentene	0.579	1,643	165	59.2	4,096	Gasoline	5	68.7	4,754	Ether Unit	4
2-M-2-Pentene	1.066	3,023	300	59.2	7,464	Gasoline	5	69.4	8,750	Ether Unit	4
c-3-M-2-Pentene	1.079	3,060	301	59.2	7,483	Gasoline	5	70.1	8,860	Ether Unit	4
1-3-M-2-Pentene	0.689	1,956	191	59.2	4,753	Gasoline	5	70.5	5,660	Ether Unit	4
C6-430	23.755	67,384	5,829	59.7	146,150	Gasoline	5	59.7	146,150	Gasoline	5
430-650	2.722	7,720	558	52.1	12,207	Diesel	6	52.1	12,207	Diesel	6
650+	0.871	2,469	159	31.8	2,127	No 6 FO	7	31.8	2,127	No 6 FO	7
Sub-total	98.799	280,251	30,315		587,951				655,438		
Coke	1.201	3,406									
Grand Total	100.000	283,657									
Coke Amount for Heat Balance, wt % lb/hr		5% 14,929									
Coke Delch, lb/hr		11,523									
MMBTU/Day		4,701									
\$/Day		9,403			(9,403)				(9,403)		
Net \$/Day					578,548				646,035		

TABLE XXXIV

ECONOMIC ANALYSIS OF FCC PILOT PLANT RUN 940-2: STEAMED BETA CATALYST

Fischer Tropsch Wax Economics

Rate Basis: 283,657 lb/hr

Pilot Plant Results of Wax Run Through an FCU

Run No. 940-2

03/19/93

Component	normalized wt % Yield	lb/hr	BBL/Day	<---Simple Configuration--->			<---Complex Configuration--->		
				cpg	\$/Day	Valued as	cpg	\$/Day	Valued as
Hydrogen	0.010	28	28	6.0	70	Fuel Gas	2	6.0	70
Methane	0.070	199	45	10.8	206	Fuel Gas	2	10.8	206
Ethylene	0.500	1,418	263	12.8	1,413	Fuel Gas	2	12.8	1,413
Ethane	0.080	227	44	12.1	222	Fuel Gas	2	12.1	222
Propylene	13.681	38,807	5,096	17.1	36,596	Fuel Gas	2	17.1	36,596
Propane	1.810	5,134	693	16.8	4,891	Fuel Gas	2	16.8	4,891
i-Butane	7.660	21,729	2,645	37.2	41,326	Alkylation	3	37.2	41,326
n-Butane	2.090	5,929	695	29.8	8,701	Gasoline	5	29.8	8,701
1-Butene	2.240	6,354	724	63.9	19,434	Alkylation	3	63.9	19,434
i-Butylene	10.751	30,495	3,481	63.8	93,273	Alkylation	3	85.4	124,852
1-2-Butene	5.440	15,432	1,734	64.8	47,183	Alkylation	3	64.8	47,183
c-2-Butene	3.980	11,290	1,234	66.6	34,518	Alkylation	3	66.6	34,518
1-Pentane	3.730	10,581	1,165	49.4	24,166	Gasoline	5	49.4	24,166
n-Pentane	1.510	4,283	465	33.8	6,605	Gasoline	5	33.8	6,605
3M-1-Butene	0.214	607	66	50.3	1,393	Gasoline	5	53.6	1,484
2M-1-Butene	1.770	5,021	524	50.3	11,077	Gasoline	5	86.5	19,049
2M-2-Butene	4.896	13,889	1,426	50.3	30,131	Gasoline	5	88.0	52,714
1-Pentene	0.505	1,433	152	50.3	3,211	Gasoline	5	54.9	3,505
1-2-Pentene	1.826	5,180	543	50.3	11,480	Gasoline	5	55.5	12,666
c-2-Pentene	1.029	2,919	306	50.3	6,469	Gasoline	5	55.5	7,138
2,3-diM-1-Butene	0.147	417	42	59.2	1,042	Gasoline	5	68.6	1,208
2-M-1-Pentene	0.507	1,438	144	59.2	3,585	Gasoline	5	68.7	4,160
2-M-2-Pentene	1.078	3,058	304	59.2	7,551	Gasoline	5	69.4	8,852
c-3-M-2-Pentene	1.169	3,316	326	59.2	8,109	Gasoline	5	70.1	9,603
1-3-M-2-Pentene	0.693	1,966	192	59.2	4,778	Gasoline	5	70.5	5,690
C6-430	27.942	79,259	6,856	59.7	171,905	Gasoline	5	59.7	171,905
430-650	2.960	8,397	607	52.1	13,276	Diesel	6	52.1	13,276
650+	0.710	2,014	130	31.8	1,735	No 6 FO	7	31.8	1,735
Sub-total	99.000	280,820	29,930		594,346			663,167	
Coke	1.000	2,837							
Grand Total	100.000	283,657							
Coke Amount for Heat Balance, wt % lb/hr		5% 14,929							
Coke Defct, lb/hr		12,093							
MMBTU/Day		4,934							
\$/Day		9,868			(9,868)			(9,868)	
Net \$/Day					584,479			653,299	

TABLE XXXV

**ECONOMIC ANALYSIS OF FCC PILOT PLANT RUN 941-1:
BLEND 25% STEAMED HZSM-5 WITH USY CATALYST**

Fischer Tropsch Wax Economics
Pilot Plant Results of Wax Run Through an FCU
03/19/93

Rate Basis: 283,657 lb/hr
Run No. 941-1

Component	normalized wt % Yield	lb/hr	BBL/Day	<---Simple Configuration--->			<---Complex Configuration--->			
				cpg	\$/Day	Valued as	cpg	\$/Day	Valued as	
Hydrogen	0.020	57	55	6.0	140	Fuel Gas	2	6.0	140	Fuel Gas 2
Methane	0.090	255	58	10.8	265	Fuel Gas	2	10.8	265	Fuel Gas 2
Ethylene	1.010	2,865	531	12.8	2,854	Fuel Gas	2	12.8	2,854	Fuel Gas 2
Ethane	0.100	284	55	12.1	277	Fuel Gas	2	12.1	277	Fuel Gas 2
Propylene	16.032	45,477	5,971	17.1	42,887	Fuel Gas	2	17.1	42,887	Fuel Gas 2
Propane	2.470	7,007	946	16.8	6,676	Fuel Gas	2	16.8	6,676	Fuel Gas 2
i-Butane	3.401	9,646	1,174	37.2	18,345	Alkylation	3	37.2	18,345	Alkylation 3
n-Butane	1.920	5,447	639	29.8	7,994	Gasoline	5	29.8	7,994	Gasoline 5
1-Butene	2.200	6,241	711	63.9	19,089	Alkylation	3	63.9	19,089	Alkylation 3
i-Butylene	10.752	30,498	3,481	63.8	93,282	Alkylation	3	85.4	124,863	Ether Unit 4
i-2-Butene	5.331	15,121	1,699	64.8	46,233	Alkylation	3	64.8	46,233	Alkylation 3
c-2-Butene	3.761	10,667	1,166	66.6	32,613	Alkylation	3	66.6	32,613	Alkylation 3
i-Pentane	2.160	6,128	675	49.4	13,996	Gasoline	5	49.4	13,996	Gasoline 5
n-Pentane	1.400	3,972	431	33.8	6,124	Gasoline	5	33.8	6,124	Gasoline 5
3M-1-Butene	0.279	792	86	50.3	1,816	Gasoline	5	53.6	1,935	Alkylation 3
2M-1-Butene	2.300	6,525	681	50.3	14,395	Gasoline	5	86.5	24,755	Ether Unit 4
2M-2-Butene	5.941	16,852	1,731	50.3	36,559	Gasoline	5	88.0	63,960	Ether Unit 4
1-Pentene	0.620	1,759	187	50.3	3,942	Gasoline	5	54.9	4,303	Alkylation 3
i-2-Pentene	2.145	6,085	638	50.3	13,486	Gasoline	5	55.5	14,880	Alkylation 3
c-2-Pentene	1.180	3,348	351	50.3	7,419	Gasoline	5	55.5	8,186	Alkylation 3
2,3-dim-1-Butene	0.085	241	24	59.2	603	Gasoline	5	68.6	698	Ether Unit 4
2-M-1-Pentene	0.266	755	76	59.2	1,881	Gasoline	5	68.7	2,183	Ether Unit 4
2-M-2-Pentene	0.682	1,935	192	59.2	4,778	Gasoline	5	69.4	5,601	Ether Unit 4
c-3-M-2-Pentene	0.840	2,383	234	59.2	5,828	Gasoline	5	70.1	6,901	Ether Unit 4
i-3-M-2-Pentene	0.448	1,271	124	59.2	3,089	Gasoline	5	70.5	3,679	Ether Unit 4
C6-430	22.493	63,804	5,519	59.7	138,385	Gasoline	5	59.7	138,385	Gasoline 5
430-650	7.471	21,192	1,531	52.1	33,508	Diesel	6	52.1	33,508	Diesel 6
650+	4.131	11,717	756	31.8	10,093	No 6 FO	7	31.8	10,093	No 6 FO 7
Sub-total	99.530	282,324	29,723		566,555			641,421		
Coke	0.470	1,333								
Grand Total	100.000	283,657								
Coke Amount for Heat Balance, wt % lb/hr		5% 14,929								
Coke Deficit, lb/hr MMBTU/Day \$/Day		13,596 5,547 11,094			(11,094)			(11,094)		
Net \$/Day					555,461			630,327		

TABLE XXXVI

**ECONOMIC ANALYSIS OF FCC PILOT PLANT RUN 942-2:
BLEND 50% USY WITH DILUENT CATALYST**

Fischer Tropsch Wax Economics
Pilot Plant Results of Wax Run Through an FCU
03/19/93

Rate Basis: 283,657 lb/hr
Run No. 942-2

Component	normalized wt % Yield	lb/hr	BBL/Day	<---Simple Configuration--->			<---Complex Configuration--->		
				cpg	\$/Day	Valued as	cpg	\$/Day	Valued as
Hydrogen	0.020	57	55	6.0	140	Fuel Gas 2	6.0	140	Fuel Gas 2
Methane	0.160	454	104	10.8	471	Fuel Gas 2	10.8	471	Fuel Gas 2
Ethylene	0.340	965	179	12.8	961	Fuel Gas 2	12.8	961	Fuel Gas 2
Ethane	0.150	426	82	12.1	416	Fuel Gas 2	12.1	416	Fuel Gas 2
Propylene	8.941	25,361	3,330	17.1	23,916	Fuel Gas 2	17.1	23,916	Fuel Gas 2
Propane	1.250	3,546	479	16.8	3,378	Fuel Gas 2	16.8	3,378	Fuel Gas 2
i-Butane	5.020	14,241	1,733	37.2	27,084	Alkylation 3	37.2	27,084	Alkylation 3
n-Butane	1.300	3,688	432	29.8	5,412	Gasoline 5	29.8	5,412	Gasoline 5
1-Butene	1.410	4,000	456	63.9	12,233	Alkylation 3	63.9	12,233	Alkylation 3
i-Butylene	6.261	17,758	2,027	63.8	54,316	Alkylation 3	85.4	72,706	Ether Unit 4
1-2-Butene	3.720	10,553	1,186	64.8	32,266	Alkylation 3	64.8	32,266	Alkylation 3
c-2-Butene	2.680	7,603	831	66.6	23,244	Alkylation 3	66.6	23,244	Alkylation 3
1-Pentane	3.970	11,262	1,240	49.4	25,722	Gasoline 5	49.4	25,722	Gasoline 5
n-Pentane	0.850	2,411	262	33.8	3,718	Gasoline 5	33.8	3,718	Gasoline 5
3M-1-Butene	0.126	357	39	50.3	820	Gasoline 5	53.6	874	Alkylation 3
2M-1-Butene	1.130	3,206	335	50.3	7,072	Gasoline 5	86.5	12,161	Ether Unit 4
2M-2-Butene	3.360	9,532	979	50.3	20,678	Gasoline 5	88.0	36,177	Ether Unit 4
1-Pentene	0.378	1,072	114	50.3	2,403	Gasoline 5	54.9	2,623	Alkylation 3
1-2-Pentene	1.450	4,113	432	50.3	9,116	Gasoline 5	55.5	10,058	Alkylation 3
c-2-Pentene	0.819	2,323	244	50.3	5,149	Gasoline 5	55.5	5,681	Alkylation 3
2,3-diM-1-Butene	0.220	624	63	59.2	1,560	Gasoline 5	68.6	1,807	Ether Unit 4
2-M-1-Pentene	0.570	1,617	162	59.2	4,031	Gasoline 5	68.7	4,677	Ether Unit 4
2-M-2-Pentene	1.136	3,223	320	59.2	7,957	Gasoline 5	69.4	9,328	Ether Unit 4
c-3-M-2-Pentene	1.210	3,433	338	59.2	8,394	Gasoline 5	70.1	9,940	Ether Unit 4
1-3-M-2-Pentene	0.733	2,079	203	59.2	5,054	Gasoline 5	70.5	6,018	Ether Unit 4
C6-430	41.493	117,699	10,181	59.7	255,278	Gasoline 5	59.7	255,278	Gasoline 5
430-650	7.471	21,191	1,531	52.1	33,506	Diesel 6	52.1	33,506	Diesel 6
650+	2.950	8,369	540	31.8	7,209	No 6 FO 7	31.8	7,209	No 6 FO 7
Sub-total	99.120	281,161	27,874		581,503			627,004	
Coke	0.880	2,496							
Grand Total	100.000	283,657							
Coke Amount for Heat Balance, wt %		5%							
lb/hr		14,929							
Coke Deficit, lb/hr		12,433							
MMBTU/Day		5,073							
\$/Day		10,145			(10,145)			(10,145)	
Net \$/Day					571,358			616,859	

TABLE XXXVII

**SUMMARY OF NET PRODUCT VALUES FOR FCC PILOT PLANT RUNS
WITH PROPYLENE VALUED AS FEEDSTOCK TO A DIISOPROPYL ETHER UNIT**

Run Number	<-----Net \$/Day----->		Percent Increase
	Propylene valued as fuel	Propylene valued as DIPE feedstock	
939-1	612,646	649,740	6.1%
939-2	622,337	655,180	5.3%
939-4	605,612	634,761	4.8%
939-5	625,664	650,817	4.0%
940-1	646,035	701,856	8.6%
940-2	653,299	708,086	8.4%
941-1	630,327	694,532	10.2%
942-2	616,859	652,663	5.8%

TABLE XXXVIII

SUMMARY OF NET PRODUCT VALUES FOR FCC PILOT PLANT RUNS

Run Number	Catalyst	Net Product Value, \$/D		
		Simple Refinery	Complex Refinery	(Complex-Simple)
939-1	Eq. USY	568,269	612,646	44,377
939-2	Eq. USY	572,805	622,337	49,532
939-4	Eq. USY	562,181	605,612	43,431
939-5	Stmd Eq. USY	570,631	625,664	55,033
940-1	Stmd Beta	578,548	646,035	67,487
940-2	Stmd Beta	584,479	653,299	68,820
941-1	75% Stmd Eq USY; 25% Stmd HZSM-5	555,461	630,327	74,866
942-2	50% Eq USY; 50% Diluent	571,358	616,859	45,501

MSS/ml/94156
4/11/94

TABLE XXXIX
PROPYLENE PROPERTIES USED IN ECONOMIC ANALYSIS

Property	Value
MON, Octane	97
RON, Octane	110
RVP, psi	5.0
Density, lb/gal.	6.4
Operating Cost, cents/gal.	30

DIPE economics are very sensitive to propylene feedstock cost. In a refinery where propylene was higher-valued as feed for alkylation or as polymer grade propylene, the manufacture of DIPE would have a smaller effect on FCU product value.

8.2 Calculation of Hydrocracking Product Values and Comparison With Pilot Plant FCC Product Values

An economic analysis of hydrocracking F-T wax was made. The economics for using Fischer-Tropsch wax as feedstock for a mild-severity hydrocracking process are given in Table XL. Figure 74 shows the refinery configuration which set the basis for the economic evaluation. Light ends from the hydrocracker were valued as fuel gas, i-butane as alkylation feed, n-butane as gasoline, all C₅ and C₆ hydrocrackate as isomerization feed, all C₇ to 350°F endpoint hydrocrackate as naphtha reformer feed, all 350+°F hydrocrackate as diesel fuel, and all off-gas from the naphtha reformer at its heating value. The hydrocracking product values were calculated using yield and density data consistent with the Bechtel/DOE computer modelling project for the hydrocracking of F-T liquids.⁽⁹⁾ The price structure used for the hydrocracking case is shown in Table XL, and is the same as was used with the FCC cases. Table XLI summarizes the yield information and calculates the value of the hydrocrackate products, in terms of dollars per day.

Figure 75 plots the "simple refinery" and "complex refinery" product values for the eight FCC pilot plant runs that were given in Table XXXVII plus the single-yield hydrocracker product value line. Since both FCU and hydrocracker diesel are sulfur-free, neither processing scheme offers an advantage concerning sulfur. Figure 75 shows that the product value for the hydrocracker process were similar to the FCC product values for a simple refinery configuration. The hydrocracker product value used was based on a simple refinery configuration when mild severity hydrocracking produced 31.4 wt% of C₅ to 350°F hydrocrackate plus alkylate and 65.4 wt% of diesel fuel, while catalytic cracking produced 68.2-80.8 wt% of C₅-430°F gasoline plus alkylate and only 2.7-9.0 wt% of diesel. Diesel fuel has a lower value than gasoline. If hydrocracking severity were higher, producing more naphtha such as in the all-gasoline mode operation, the product value from hydrocracking might be higher than from FCC in a simple refinery configuration. The FCU diesel had an average cetane index of 58.6. For Amoco, the incremental cetane value is between \$0.0/cetane-barrel and \$0.04/cetane-barrel, depending on season and location. Using an average value of \$0.02/cetane barrel and a cetane of 75 for hydrocracker diesel, the price increment for hydrocracker diesel was calculated to be 0.78 cpg. This calculated value was viewed as too small to incorporate into the economic analysis. As a sensitivity, a 10% increase in diesel product value for the hydrocracker case would raise the value of the hydrocracker products by 6.2%. This interaction would be

TABLE XL

FISCHER-TROPSCH WAX HYDROCRACKING CASE: PRICE STRUCTURE CALCULATIONS

Component	Price	Properties	
gasoline (ULR)	55.70 cpg	9.0 RVP	87.0 (R + M)/2
FO #2	52.10 cpg		
FO #6	31.8 cpg		
MTBE	86.00 cpg	9.0 RVP	108.0 (R + M)/2
iC4	37.20 cpg	87.0 RVP	97.6 (R + M)/2
nC4	29.80 cpg	65.0 RVP	90.6 (R + M)/2
N. Gas	2.00 \$/MMBTU		
Power	2.28 ¢/kw-hr		
H2SO4	75.00 \$/ton		
Caustic	395.00 \$/ton		
Methanol	50.00 cpg	32.04 MW	278.418 lb/gal
Incremental Property Values	1.44 ¢/octane-gal (0.55)¢/RVP-gal		
Alkylate	62.49 cpg	7.3 RVP	91.1 (R + M)/2
Isomerate	46.43 cpg	10.1 RVP	81.0 (R + M)/2
Reformate	63.13 cpg	3.4 RVP	90.0 (R + M)/2

Component	BTU/Gal	Value To Fuel, cpg	RVP, psi	Octane (R + M)/2	Value to ULR, cpg	Value to Alky, cpg	Value to Isom, cpg	Value to Reformer, cpg	Value to Diesel, cpg
Hydrogen	30,160	6.03							
Methane	53,754	10.75							
Ethane	60,592	12.12							
Propane	84,100	16.82							
i-Butane	91,830	18.37	87.0	97.6	27.71	37.20			
n-Butane	95,603	19.12	65.0	90.6	29.80				
i-Pentane	100,540	20.11	22.0	87.7	49.44		43.39		
n-Pentane	101,770	20.35	17.0	74.9	33.81		43.80		
i-Hexane	106,608	21.32	8.0	77.2	42.11		48.48		
n-Hexane	107,113	21.42	5.0	25.4	(30.96)		48.79		
C7-300 F HC								60.69	
300 F - 350 F HC								63.49	
350 F - 500 F HC									52.10
500 F - 700 F HC									52.10

TABLE XLI

ECONOMIC ANALYSIS OF F-T WAX HYDROCRACKER CASE

Rate Basis: 283,657 lb/hr

Component	normalized wt % Yield	lb/hr	BBL/Day	cpg	\$/Day	Valued as	
Hydrogen	0.000	0	0	6.0	0	Fuel Gas	2
Methane	0.049	139	32	10.8	145	Fuel Gas	2
Ethane	0.049	139	27	12.1	136	Fuel Gas	2
Propane	1.460	4,141	559	16.8	3,945	Fuel Gas	2
i-Butane	1.934	5,486	668	37.2	10,433	Alkylation	3
n-Butane	1.569	4,451	522	29.8	6,532	Gasoline	6
i-Pentane	2.407	6,828	752	43.4	13,700	Isom Unit	4
n-Pentane	2.033	5,766	626	43.8	11,522	Isom Unit	4
i-Hexane	3.828	10,858	1,068	48.5	21,754	Isom Unit	4
n-Hexane	2.348	6,661	651	48.8	13,344	Isom Unit	4
C7 - 300 F HC	14.287	40,526	3,949	60.7	100,678	Reformer	5
300 F - 350 F HC	4.588	13,014	1,220	63.5	32,550	Reformer	5
350 F - 500 F HC	20.918	59,336	5,379	52.1	117,711	Diesel	7
500 F - 700 F HC	44.530	126,312	10,961	52.1	239,849	Diesel	7
Total	100.000	283,657	26,414		572,300		

FIGURE 74

**SCHEMATIC OF BASIS REFINERY CONFIGURATION
FOR HYDROCRACKING CASE ECONOMICS**

FIGURE 75

**NET PRODUCT VALUE FOR FCC PILOT PLANT RUNS
AND HYDROCRACKER CASE**

Simple Refinery Complex Refinery Hydrocracker Case

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much smaller for the FCU cases, because the diesel yields are much smaller.

Figure 75 shows that hydrocracker values are always lower than the FCC product values for a complex refinery configuration. Since hydrocracking does not generate olefins, catalytic cracking is preferred because of its high yields of iso-olefins for use as feedstocks for ether production.

For the reasons discussed above in section 8.1.2, valuation of the propylene from the FCC process as feedstock for a DIPE unit rather than as fuel would further increase the value of the FCC process over the hydrocracking process in a complex refinery configuration.

9.0 CONCLUSIONS

The catalytic cracking of F-T wax feedstocks gave high conversions with low activity catalysts and low process severities. Catalyst selection and process optimization will depend on product valuation. HZSM-5 and beta zeolite catalysts gave higher yields of propylene, isobutylene, and isoamylenes but a lower gasoline yield than Y zeolite catalysts. For a given catalyst and process condition, Sasol and LaPorte waxes gave similar conversions and product selectivities. The contaminant iron F-T catalyst fines in the LaPorte wax caused higher coke and hydrogen yields.

The net product values for catalytic cracking F-T wax feedstock in eight pilot plant runs were calculated with the iso-olefins valued as alkylate or gasoline (simple refinery configuration) and as feedstock to an ether unit (complex refinery configuration). Irrespective of catalyst, the net product values from catalytic cracking F-T wax feedstock were always higher when the iso-olefins were valued as feed to an ether unit rather than fuel. In a simple refinery configuration, additional experimental work would be required to identify the catalyst that gave the highest net product value. In a complex refinery configuration, HZSM-5 and beta zeolite catalysts gave higher net product values than Y zeolite catalyst because of their higher yields of iso-olefins. The net product values of the HZSM-5 and beta zeolite catalysts were additionally increased over the Y zeolite catalyst when propylene was valued as feedstock to a di-isopropyl ether unit rather than fuel gas.

The product value for mild-severity hydrocracking F-T wax was calculated from literature data and compared with the product values from the eight pilot plant runs. The value for the hydrocracking case was below the values for catalytic cracking using a complex refinery configuration but were similar to the product values for catalytic cracking using a simple refinery configuration. If hydrocracking severity were higher, producing more naphtha such as in the all-gasoline mode operation, the product value from hydrocracking might be higher than from FCC in a simple refinery configuration.

The 200-°F product from catalytic cracking pilot plant runs was demonstrated to be a good feedstock for the production of mixed ethers by reaction with methanol using conventional technology.

10.0 REFERENCES

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APPENDIX A

MYU Run Data

APPENDIX A -- MYU Run Data

Run Number	Feedstock	Catalyst Identification	Catalyst Type	C/O Ratio	Reaction Temperature, °F	Recovery Wt%	Conversion Wt%	Coke Wt%
91-59-098	Sasol Wax	CCC-408	Eq. USY	3.94	971	76.8	90.2	2.99
91-59-099	Sasol Wax	CCC-408	Eq. USY	3.72	971	81.8	99.1	2.06
91-59-100	Sasol Wax	CCC-408	Eq. USY	2.72	971	111.1	85.3	1.85
91-59-101	Sasol Wax	CCC-408	Eq. USY	3.06	970	97.6	88.1	2.21
92-59-001	Sasol Wax	CCC-408	Eq. USY	3.16	971	92.7	88.2	3.03
92-59-002	Sasol Wax	st. CCC-1397	St. Eq. USY	3.20	971	94.3	87.6	0.97
92-59-003	Sasol Wax	st. CCC-1397	St. Eq. USY	3.27	972	91.7	87.2	1.18
92-59-004	Sasol Wax	st. CCC-1875	St. Beta	3.26	972	92.2	90.9	0.82
92-59-005	Sasol Wax	st. CCC-1875	St. Beta	3.37	972	89.5	91.8	0.91
92-59-006	Sasol Wax	st. 1397	St. Eq. USY	3.35	939	89.7	85.8	1.13
92-59-007	Sasol Wax	st. CCC-1875	St. Beta	3.29	939	91.6	91.4	1.08
92-59-008	Sasol Wax	st. 1397	St. Eq. USY	3.28	909	91.7	87.6	1.59
92-59-009	Sasol Wax	st. 1875	St. Beta	3.31	909	90.7	91.4	1.25
92-59-010	Sasol Wax	st. 1397	St. Eq. USY	3.32	880	90.2	88.2	1.84
92-59-011	Sasol Wax	CCC-1875	St. Beta	3.16	879	94.3	92.4	1.29
92-59-014	Sasol Wax	1397(1:1-1872)	Eq. USY	1.62	970	86.8	86	1.11
92-59-015	Sasol Wax	1397(1:1-1872)	Eq. USY	1.55	969	90.7	85.1	0.91
92-59-016	Sasol Wax	1397(1:1-1872)	Eq. USY	1.58	969	88.8	86.2	1.05
92-59-017	Sasol Wax	1397(1:1w1872)	Eq. USY	1.63	969	89.9	86.1	1.04
92-59-018	Sasol Wax	1397(1:3w1872)	Eq. USY	0.85	969	86.7	84.3	0.5
92-59-019	Sasol Wax	1397(1:3w1872)	Eq. USY	0.84	969	87.1	84.4	0.57
92-59-020	Sasol Wax	1397(1:3w1872)	Eq. USY	0.82	970	93.6	93.6	0.55
92-59-021	Sasol Wax	1397(1:3w1872)	Eq. USY	0.79	969	93.5	84.5	0.52
92-59-022	Sasol Wax	1397(1:3w1872)	Eq. USY	0.76	881	91.9	70.1	0.6
92-59-023	Sasol Wax	1397(1:3w1872)	Eq. USY	0.80	881	87.1	82.6	0.71
92-59-024	Sasol Wax	1397(1:3w1872)	Eq. USY	0.89	880	84.1	81.8	0.86
92-59-025	Sasol Wax	1397(1:3w1872)	Eq. USY	0.82	881	92.1	74.7	0.77
92-59-026	Sasol Wax	1875(1:3w1872)	St. Beta	0.89	881	85.2	91.2	0.5
92-59-027	Sasol Wax	1875(1:3w1872)	St. Beta	0.93	880	83.4	89.6	0.47
92-59-028	Sasol Wax	1875(1:7w1872)	St. Beta	0.45	880	88.3	88.9	0.35
92-59-029	Sasol Wax	1875(1:7w1872)	St. Beta	0.46	881	89.2	84.2	0.28
92-59-030	Sasol Wax	1397(1:7w1872)	Eq. USY	0.43	880	91.7	77.6	0.32
92-59-031	Sasol Wax	1397(1:7w1872)	Eq. USY	0.43	880	86.8	83	0.31
92-59-032	Sasol Wax	CCC-1891	St. HZSM-5	3.58	880	83	93.6	0.35
92-59-033	Sasol Wax	CCC-1891	St. HZSM-5	3.65	880	81.5	90.4	0.38

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Propylene Wt%	Total C3 Wt%	Isobutylene Wt%	Total C4 Wt%	Isoamylenes Wt%	C5-430 deg. F Wt%	RON	MON	Conversion by Sim. Dis., Wt%	Comments
91-59-098	10.31	12.23	3.5	18.19	3.51	54.41	85.7	78.2		
91-59-099	9.86	11.69	3.83	19.83	5.82	63.3	86.9	78		
91-59-100	7.08	8.32	2.55	13.68	4.79	59.83	84.7	77.3		
91-59-101	9.95	11.39	4.49	19.99	5.79	51.16	85.8	77.6		
92-59-001	11.42	13.26	3.24	15.88	4.46	53.08	83.5	77.6		
92-59-002	11.04	12.34	6.4	21.39	7.19	51.18	86.1	77.9		
92-59-003	10.94	12.3	6.28	21.15	6.89	50.78	85.7	77.4		
92-59-004	14.6	16.5	8.3	31.84	5.73	39.85	86	78.6		
92-59-005	15.35	17.36	8.24	32.26	6.09	39.35	87.1	79.2		
92-59-006	10.42	11.67	6.3	21.64	6.92	50.12	85.8	77.5		
92-59-007	13.4	15.32	8.4	33.56	5.95	39.88	86.1	78.5		
92-59-008	10.96	12.46	6.17	22.66	6.89	49.55	86.2	78.5		
92-59-009	14.3	16.33	7.67	31.26	6.33	41.27	86.6	79		
92-59-010	9.88	11.26	6.22	22.94	7.44	51.02	87.2	78.7		
92-59-011	12.14	13.89	8.82	31.53	7.69	44.89	87.3	78.8		
92-59-014	10.4	11.7	6	22.58	6.35	49.02	85.9	78.4		
92-59-015	9.98	11.17	6.1	21.57	7.07	50.11	87.3	78.7		
92-59-016	10.66	11.92	5.81	21.7	6.59	50.06	85.8	78.3		
92-59-017	10.62	11.85	5.73	21.44	6.69	50.31	85.9	78.3		
92-59-018	9.8	10.81	6.14	19.95	7.4	51.74	86	77.3	91.8	
92-59-019	9.88	10.92	6.26	19.79	7.14	51.76	85.6	77.1	91.8	
92-59-020	9.56	10.58	5.93	18.61	7.4	53.27	85.9	77.3	91.7	
92-59-021	9.46	10.43	4.96	18.87	7.78	53.46	85.9	77.2	92.2	
92-59-022	4.69	5.23	3.56	11.13	5.29	52.83	86	75.9	77.3	
92-59-023	7.61	8.82	5.41	17.38	7.08	54.95	85.3	77	89.8	
92-59-024	7.96	8.88	5.13	17.02	6.52	54.34	85.2	77.1	89.5	
92-59-025	5.44	6.06	3.64	11.65	5.67	55.74	85.8	76.4	82.8	
92-59-026	12.06	13.2	10.37	27.2	9.35	49.59	83.8	76	97.1	
92-59-027	11.44	12.54	10.1	26.7	9.27	49.23	85.6	76.6	95.1	
92-59-028	10.85	11.82	10.23	24.33	9.85	51.78	84.2	75.2		
92-59-029	9.53	10.39	9.03	21.68	9.22	51.35	84.5	75.1		
92-59-030	5.75	6.37	4.63	13.26	6.49	57.2	85.2	75.7		
92-59-031	7.45	8.25	5.78	16.81	7.66	56.99	85.2	76.2	89.2	
92-59-032	11.16	19.08	6.31	29.77	3.81	38.49	96.4	83		
92-59-033	12.46	18.93	7.02	27.23	3.97	39.18	93	81.2		

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Feedstock	Catalyst Identification	Catalyst Type	C/O Ratio	Reaction Temperature, °F	Recovery Wt%	Conversion Wt%	Coke Wt%
92-59-034	Sasol Wax	1397(1:15w1872)	Eq. USY	0.21	880	90.9	55.1	0.17
92-59-035	Sasol Wax	1397(1:15w1872)	Eq. USY	0.23	880	85.6	64	0.18
92-59-036	Sasol Wax	1875(1:15w1872)	St. Beta	0.22	881	86.6	64.7	0.17
92-59-037	Sasol Wax	1875(1:15w1872)	St. Beta	0.22	880	87.1	80.3	0.17
92-59-038	Sasol Wax	1397(1:15w1872)	Eq. USY	0.24	880	85.2	71.6	0.23
92-59-039	Sasol Wax	1397(1:15w1872)	Eq. USY	0.21	880	90.4	62.2	0.16
92-59-040	Sasol Wax	1397(1:15w1872)	Eq. USY	0.22	880	88	69.2	0.19
92-59-041	Sasol Wax	1397(1:15w1872)	Eq. USY	0.22	880	88.8	71.6	0.19
92-59-042	Sasol Wax	1875(1:15w1872)	St. Beta	0.23	880	83.1	83.3	0.19
92-59-043	Sasol Wax	1875(1:15w1872)	St. Beta	0.22	879	89.5	58	0.18
92-59-044	Sasol Wax	1875(1:15w1872)	St. Beta	0.20	969	91.7	76.7	0.09
92-59-045	Sasol Wax	1875(1:15w1872)	St. Beta	0.21	970	89.7	80.9	0.09
92-59-046	Sasol Wax	1397(1:15w1872)	Eq. USY	0.20	970	95.5	70.9	0.12
92-59-047	Sasol Wax	1397(1:15w1872)	Eq. USY	0.20	969	94.9	71.8	0.12
92-59-048	Sasol Wax	CCC 1894	Octacat Matrix	3.30	970	92	72.6	0.61
92-59-049	Sasol Wax	CCC 1894	Octacat Matrix	3.27	970	92.8	73	0.64
92-59-050	Sasol Wax	CCC 1891 (1:3w1872)	St. HZSM-5	0.95	880	81.8	84.1	0.19
92-59-051	Sasol Wax	1891 (1:3w1872)	St. HZSM-5	0.91	880	84.7	87.7	0.19
92-59-052	Sasol Wax	1891 (1:7w1872)	St. HZSM-5	0.48	880	89.3	88.3	0.16
92-59-053	Sasol Wax	1891 (1:7w1872)	St. HZSM-5	0.48	879	80.6	85.6	0.15
92-59-054	Sasol Wax	1891 (1:15w1872)	St. HZSM-5	0.22	880	88	84.1	0.11
92-59-055	UOP Naph.	1891 (1:15w1872)	St. HZSM-5	0.22	880	83.8	85.9	0.13
92-59-056	UOP Naph.	1397(1:15d1872)	Eq. USY	0.20	881	92.3	76.2	0.17
92-59-057	UOP Naph.	1397(1:15d1872)	Eq. USY	0.20	880	93.4	75	0.16
92-59-058	UOP Naph.	1875(1:15d1872)	St. Beta	0.21	880	90.9	79.1	0.14
92-59-059	UOP Naph.	1875(1:15d1872)	St. Beta	0.20	881	91.2	79.2	0.12
92-59-060	UOP Naph.	1891 (1:15/1872)	St. HZSM-5	0.21	881	90.4	61.3	0.13
92-59-061	UOP Naph.	1891 (1:15/1872)	St. HZSM-5	0.21	880	90.7	62.9	0.1
92-59-062	UOP Naph.	1872	Diluent	3.20	880	95	696.8	0.18
92-59-063	UOP Naph.	1872	Diluent	3.18	880	95.3	73	0.16
92-59-064	Sasol Wax	1875(1:31w1872)	St. Beta	0.12	881	80.1	65.5	0.16
92-59-065	Sasol Wax	1875(1:31w1872)	St. Beta	0.12	879	84.2	66.7	0.11
92-59-068	UOP Naph.	1397(1:15d1872)	Eq. USY	0.23	879	90.2	68.5	0.14
92-59-069	UOP Naph.	1397(1:15d1872)	Eq. USY	0.21	970	88.5	100.1	0.23
92-59-070	UOP Naph.	1875(1:15d1872)	St. Beta	0.22	969	87.7	101.6	0.18
92-59-071	UOP Naph.	1891 (1:15d1872)	St. HZSM-5	0.21	969	90	100.1	0.12

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Propylene Wt%	Total C3 Wt%	Isobutylene Wt%	Total C4 Wt%	Isoamylenes Wt%	C5-430 deg. F Wt%	RON	MON	Conversion by Sim. Dis., Wt%	Comments
92-59-034	3.4	3.82	2.93	8.16	3.53	42.54	84.9	74.3	57.6	
92-59-035	4.29	4.79	3.61	10.11	4.73	48.53	84.9	74.6	72.8	
92-59-036	6.4	7.04	6.83	15.98	6.37	41.03	84.3	74.5		
92-59-037	7.92	8.71	8.77	19.82	8.91	51.12	84.2	74.5		
92-59-038	5.72	6.36	4.88	13.52	6.21	50.98	85.1	75.3		
92-59-039	4.07	4.55	3.58	10.14	4.68	46.92	85	74.8		
92-59-040	4.76	5.28	4.15	11.45	5.68	51.83	85.6	75.1		
92-59-041	5.53	6.15	4.61	12.74	6.3	51.98	84.5	75		
92-59-042	8.94	9.79	9.45	21.24	9.18	51.44	84.4	74.6		
92-59-043	5.41	5.97	5.56	13.01	4.81	38.41	83.4	73.8		
92-59-044	8.35	9.17	7.61	17.37	5.09	48.89	83.6	73.6		
92-59-045	9.7	10.58	8.69	19.94	5.91	49.07	82.8	73.7		
92-59-046	5.73	6.39	4.22	11.49	5.94	51.85	86.1	75	77.7	
92-59-047	6.16	6.84	4.38	12.02	6.25	51.62	86.2	75.3	83	
92-59-048	9.64	10.65	7.21	18.66	8.08	41.22	88.6	76.8		
92-59-049	10.53	11.61	7.38	19.52	7.59	39.75	87.9	76.9		
92-59-050	16.9	21.4	10.62	31.57	5.72	26.95	83.5	76.3		
92-59-051	16.05	20.39	10.31	34.55	6.09	28.74	87.4	78.2		
92-59-052	15.69	19.23	12.46	34.47	7.81	31.25	84.4	76.4		
92-59-053	15.27	18.64	12.09	33.08	7.85	30.82	86.6	77.5		
92-59-054	17.01	19.71	12.52	31.43	9.44	31.21	85.4	76.4		
92-59-055	17.48	20.18	12.28	30.91	9.75	31.06	84.4	76		
92-59-056	2.98	3.04	0.47	2.92	1.02	66.84	77.3	69.6		
92-59-057	2.81	2.86	0.5	3.09	1.23	65.87	76.8	69.5		
92-59-058	3.71	3.78	1.49	5.17	2.75	67.12	75.8	68.9		
92-59-059	3.83	3.9	1.64	5.64	2.93	66.65	75.3	68.7		
92-59-060	8.22	8.29	2.84	8.53	4.24	39.93	74.6	68.6		
92-59-061	8.12	8.19	2.95	9.03	4.54	41.08	75.6	69.2		
92-59-062	0.98	1.11	0.44	1.66	0.2	692.83	77.6	66.2		
92-59-063	1.1	1.23	0.58	1.94	0.24	68.61	74.1	65.1		
92-59-064	15.54	17.5	9.26	22.62	6.69	24.16	84.2	75.1		
92-59-065	14.62	16.41	9.02	22.54	7.88	26.7	85.2	75.9		
92-59-068	8.19	9.33	6.48	17.86	8.6	40.43	83.6	74.3		
92-59-069	4.13	4.25	1.31	6.59	3.32	83.99	79.4	70.3		
92-59-070	6.44	6.58	3.15	9.34	5.84	80.15	74.6	68.3		
92-59-071	11.75	11.86	5.91	14.16	8.93	67.95	75.9	69.4		

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Feedstock	Catalyst Identification	Catalyst Type	C/O Ratio	Reaction Temperature, F	Recovery Wt%	Conversion Wt%	Coke Wt%
92-59-072	UOP Naph.	1891 (1:15d1872)	St. HZSM-5	0.21	969	86.4	99.8	0.19
92-59-084	Sasol Wax	14040- 43-1 (1:15w1872)	St. HZSM-5	0.25	882	80	88.9	0.16
92-59-085	Sasol Wax	14040- 43-2 (1:15w1872)	St. HZSM-5	0.23	881	78.7	89.6	0.13
92-59-086	Sasol Wax	14040- 43-3 (1:15W1872)	St. HZSM-5	0.23	879	0.2	68.5	0.14
92-59-087	Sasol Wax	1891(1:15W1872)	St. HZSM-5	0.27	880	72.1	91.9	0.14
92-59-088	Sasol Wax	1891(1:15W1872)	St. HZSM-5	0.24	880	76.8	89.8	1.28
92-59-089	Sasol Wax	1891(1:15W1872)	St. HZSM-5	0.24	881	76.9	92.8	0.14
92-59-090	Sasol Wax	14040- 45-1 (1:15 1872)	St. HZSM-5	0.22	880	78.8	83.8	0.13
92-59-091	Sasol Wax	14040- 43-1 (1:15w1872)	St. HZSM-5	0.24	880	80.3	90	0.12
92-59-092	Sasol Wax	14040- 43-2 (1:15w1872)	St. HZSM-5	0.25	881	73.1	95.1	0.15
92-59-093	Sasol Wax	14040- 43-3 (1:15W1872)	St. HZSM-5	0.22	880	83.7	75.7	0.12
92-59-094	Sasol Wax	14040- 45-1 (1:15w1872)	St. HZSM-5	0.24	881	78.6	117.7	0.15
92-59-095	Sasol Wax	14040- 44-4 (1:15w1872)	St. Mordenite	0.18	880	107.4	37.7	0.1
92-59-096	Sasol Wax	14040- 42-1 (1:15w1872)	Reo/HZSM-5	0.24	880	77.3	85	0.14
92-59-097	Sasol Wax	14040- 42-2 (1:15w1872)	Reo/HZSM-5	0.24	880	80.4	85.1	0.13
92-59-098	Sasol Wax	14040- 42-3 (1:15W1872)	Reo/HZSM-5	0.22	880	80.6	82.1	0.13
92-59-099	Sasol Wax	1701 (1:3w1872)	St. USY	0.89	880	81.9	91.6	0.83
92-59-100	Sasol Wax	1702 (1:3w1872)	St. Reo/USY	0.82	880	99.6	87.9	0.79
92-59-101	Sasol Wax	1703 (1:3w1872)	St. Reo/USY	0.81	880	90.5	84.3	0.8
92-59-102	Sasol Wax	1704 (1:3w1872)	St. Reo/USY	0.86	881	86.4	90.8	0.94
92-59-103	Sasol Wax	1705 (1:3w1872)	St. Reo/USY	0.82	878	89.3	90.1	0.83
92-59-104	Sasol Wax	1706 (1:3w1872)	St. Reo/USY	0.90	880	80.3	94.2	0.13
92-59-106	Sasol Wax	1891 (1:31d1872)	St. HZSM-5	0.12	880	75	83.8	0.14
92-59-107	Sasol Wax	14040- 45-1 (1:31)	St. HZSM-5	0.11	880	79.1	83.3	0.15
92-59-108	Sasol Wax	14040- 43-1 (1:31 1872)	St. HZSM-5	0.11	880	80.1	80.4	0.13
92-59-109	Sasol Wax	14040- 43-2 (1:31 1872)	St. HZSM-5	0.11	880	83.1	90.2	0.12
92-59-110	Sasol Wax	14040- 43-3 (1:31 1872)	St. HZSM-5	0.10	880	88.3	46.7	0.16
92-59-111	Sasol Wax	14040- 42-1 (1:31d1872)	Reo/HZSM-5	0.10	880	85.8	64.2	0.17
92-59-112	Sasol Wax	14040- 42-2 (1:31d1872)	Reo/HZSM-5	0.10	880	86.5	59.9	0.14
92-59-113	Sasol Wax	14040- 42-3 (1:31d1872)	Reo/HZSM-5	0.11	880	84.3	56.5	0.15
92-59-114	Sasol Wax	1891 (1:31d1872)	St. HZSM-5	0.12	880	76.4	80.5	0.17
92-59-115	Sasol Wax	14040- 45-1 (1:31d1872)	St. HZSM-5	0.11	877	82.6	84.7	0.13
92-59-116	Sasol Wax	14040- 43-1 (1:31 1872)	St. HZSM-5	0.12	880	75.8	82.6	0.15
92-59-117	Sasol Wax	14040- 43-2 (1:31 1872)	St. HZSM-5	0.13	880	68.8	85.8	0.17
92-59-118	Sasol Wax	14040- 43-3 (1:31 1872)	St. HZSM-5	0.11	880	81.9	41.4	0.15
92-59-119	Sasol Wax	14040- 42-1 (1:31d1872)	Reo/HZSM-5	0.11	881	67.7	79.8	0.14

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Propylene Wt%	Total C3 Wt%	Isobutylene Wt%	Total C4 Wt%	Isoamylenes Wt%	C5-430 deg. F Wt%	RON	MON	Conversion by Sim. Dis., Wt%	Comments
92-59-072	8.8	8.9	3.93	12.16	7.63	73.24	74.4	68.6		
92-59-084	9.41	11.28	10.07	28.1	11.67	47.62	84.4	75		
92-59-085	10.38	12.48	9.3	26.36	10.74	49.02	86.2	75.7		
92-59-086	8.19	9.33	6.48	17.86	8.6	40.43	83.6	74.3		
92-59-087	13	15.8	11.74	31.1	8.73	42.23	80.9	74.2		
92-59-088	11.65	13.75	10.35	26.87	10.73	46.33	84.3	75.1		
92-59-089	15.11	17.82	12.19	31.44	10.69	41.14	82.5	74.8		
92-59-090	12.86	14.73	9.48	24.2	10.23	43.54	84.3	74.9		
92-59-091	14.79	17.23	10.75	27.38	9.96	43.47	81.7	74.2		
92-59-092	17.98	21.77	11.94	32.75	9.25	36.98	85	76.5		
92-59-093	11.94	13.33	8.05	20.19	10.25	41.28	84.8	75.3		
92-59-094	14.96	17.72	13.03	33.46	14.98	64.14	82.3	74.3		
92-59-095	1.74	2.11	1.54	4.45	1.98	30.4	76.2	70.2		
92-59-096	13.45	15.35	10.05	24.79	11.22	43.54	83.3	74.7		
92-59-097	13.01	14.86	10.16	25.09	11.63	43.94	83.7	75		
92-59-098	12.19	13.92	9.85	24.28	11.08	42.67	83.3	74.8		
92-59-099	7.67	8.57	5.9	19.58	7.74	61.85	89.8	78.1		
92-59-100	7.28	8.14	4.92	16.77	6.57	61.5	88.8	76.5		
92-59-101	5.99	6.72	4.23	14.39	6.02	61.79	87.7	74.9		
92-59-102	6.65	7.45	4.36	15.83	6.48	65.75	88.1	76.1		
92-59-103	5.56	6.21	4.01	14.05	6.12	68.34	86.6	74.1		
92-59-104	7.37	8.2	6.16	19.14	8.52	65.94	88.8	76.7		
92-59-106	13.83	15.84	10.14	25.65	10.62	40.74	84.2	75.2		
92-59-107	13.7	15.48	9.4	23.51	11.19	43.05	84.3	75.2		
92-59-108	12.31	14.01	8.96	22.2	10.6	42.98	83.6	74.7		
92-59-109	14.18	16.48	10.74	27.24	11.56	44.67	83.9	75.1		
92-59-110	6.8	7.45	3.96	11.01	4.61	27.49	83.4	74		
92-59-111	10.99	12.26	7.18	18.07	7.38	32.82	83.5	74.5		
92-59-112	9.84	10.9	6.1	15.64	6.97	32.39	85.9	74.4		
92-59-113	9.87	10.9	6.02	15.49	6.04	29.16	83.2	74.3		
92-59-114	15.32	17.53	10.38	25.84	8.97	35.13	81.5	74.3		
92-59-115	14.91	17.03	10.9	27.05	11.31	39.09	84.6	75.6		
92-59-116	13.57	15.58	10.45	25.92	10.97	39.49	84.1	75.4		
92-59-117	14.58	17.72	9.24	24.7	10.04	41.12	83.3	75		
92-59-118	6.81	7.61	3.91	10.52	3.76	22.48	80.6	73.1		
92-59-119	11.59	12.87	7.35	18.34	8.74	35.57	84.8	75.2		

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Feedstock	Catalyst Identification	Catalyst Type	C/O Ratio	Reaction Temperature, F	Recovery F Wt%	Conversion Wt%	Coke Wt%
92-59-120	Sasol Wax	14040- 42-2 (1:31d1872)	Reo/HZSM-5	0.11	880	83.7	67.6	0.15
92-59-121	Sasol Wax	14040- 42-3 (1:31d1872)	Reo/HZSM-5	0.11	880	81.2	67.9	0.15
92-59-122	Sasol Wax	1701 (1:15d1872)	St. USY	0.22	881	81.7	86.1	0.15
92-59-123	Sasol Wax	1702 (1:15d1872)	St. Reo/USY	0.21	881	85.6	85.3	0.23
92-59-124	Sasol Wax	1703 (1:15d1872)	St. Reo/USY	0.20	881	85.7	83.8	0.26
92-59-125	Sasol Wax	1704 (1:15d1872)	St. Reo/USY	0.22	880	84.9	89.9	0.32
92-59-126	Sasol Wax	1705 (1:15d1872)	St. Reo/USY	0.22	879	86.3	36.7	0.17
92-59-127	Sasol Wax	1706 (1:15d1872)	St. Reo/USY	0.20	879	87	69.1	0.23
92-59-128	Sasol Wax	1701 (1:15d1872)	St. USY	0.23	882	85	84.9	0.32
92-59-129	Sasol Wax	1702 (1:15d1872)	St. Reo/USY	0.20	879	91.3	52.4	0.2
92-59-130	Sasol Wax	1703 (1:15d1872)	St. Reo/USY	0.20	878	86.6	75.5	0.23
92-59-131	Sasol Wax	1704 (1:15d1872)	St. Reo/USY	0.26	880	68.2	46.2	0.27
92-59-132	Sasol Wax	1705 (1:15d1872)	St. Reo/USY	0.21	880	87.7	85.7	1.39
92-59-133	Sasol Wax	1706 (1:15d1872)	St. Reo/USY	0.21	880	91.3	44.1	0.17
92-59-134	Sasol Wax	1701 (1:15d1872)	St. USY	0.20	878	93.8	52.3	0.19
92-59-135	Sasol Wax	1702 (1:15d1872)	St. Reo/USY	0.20	877	92.1	39.7	0.2
92-59-136	Sasol Wax	1703 (1:15d1872)	St. Reo/USY	0.22	884	89.8	66.5	0.25
92-59-137	Sasol Wax	1701 (1:3d1872)	St. USY	0.86	881	89	80.1	0.71
92-59-138	Sasol Wax	1702 (1:3d1872)	St. Reo/USY	0.81	880	97.7	83	0.7
92-59-139	Sasol Wax	1703 (1:3d1872)	St. Reo/USY	0.85	880	90.9	81.3	0.76
92-59-140	Sasol Wax	1704 (1:3d1872)	St. Reo/USY	0.81	879	97.1	71.6	0.76
92-59-141	Sasol Wax	1705 (1:3d1872)	St. Reo/USY	0.78	881	93.7	59.7	0.67
92-59-142	Sasol Wax	1706 (1:3d1872)	St. Reo/USY	0.87	880	90.7	62.7	0.56
92-59-143	Sasol Wax	1701 (1:3d1872)	St. USY	0.83	880	89.1	62.9	0.54
92-59-149	Sasol Wax	1701 (1:3d1872)	St. USY	0.86	969	87.4	80.3	0.54
92-59-150	Sasol Wax	1702 (1:3d1872)	St. Reo/USY	0.86	970	87.9	89.3	0.63
92-59-151	Sasol Wax	1703 (1:3d1872)	St. Reo/USY	0.88	969	85.7	90.4	0.6
92-59-152	Sasol Wax	1704 (1:3d1872)	St. Reo/USY	0.85	970	91	87.5	0.22
92-59-153	Sasol Wax	1705 (1:3d1872)	St. Reo/USY	0.98	970	83	97.9	0.8
92-59-154	Sasol Wax	1706 (1:3d1872)	St. Reo/USY	0.89	969	85.7	78.9	0.73
92-59-155	Sasol Wax	14040- 48-1 (1:3d1872)	St. USY	0.85	970	86.7	76.1	0.57
92-59-156	Sasol Wax	14040- 48-2 (1:3d1872)	St. USY	0.85	970	87.5	84.5	0.46
92-59-157	Sasol Wax	9669-153 (1:3d)	Matrix Only	0.83	969	91.4	47.9	0.13
92-59-158	Sasol Wax	14040- 50-5	St. Matrix	3.43	969	89.8	63.3	0.31
92-59-159	Sasol Wax	14040- 50-6 (1:3d)	St. 10% USY	0.83	970	90.9	78.5	0.2
92-59-160	Sasol Wax	14040- 50-7 (1:3d1872)	St. 10% Beta	0.85	970	88.3	64.7	0.18

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Propylene Wt%	Total C3 Wt%	Isobutylene Wt%	Total C4 Wt%	Isoamylenes Wt%	C5-430 deg. F Wt%	RON	MON	Conversion by Sim. Dis., Wt%	Comments
92-59-120	11.67	12.95	7.76	19.45	8.74	34.22	85	75.5		
92-59-121	11	12.18	7.03	17.69	8.02	37.11	82	73.9		
92-59-122	4.94	5.37	5.14	12.88	8.15	67.3	87	75.3		
92-59-123	4.88	5.43	4.74	12.74	7.18	66.21	86.3	74.6		
92-59-124	4.1	4.58	4.17	11.09	6.97	67.31	85.7	73.7		
92-59-125	4.57	5.11	4.02	11.56	6.7	72.37	86.2	74.2		
92-59-126	1.88	2.1	1.7	4.94	0	29.25	88.2	76.4		
92-59-127	3.5	3.94	3.31	9.15	4.78	55.32	86.3	74.5		
92-59-128	5.5	6.11	5.03	13.69	7.3	64.15	87.1	75.2		
92-59-129	2.86	3.16	2.74	7.82	3.35	40.94	86	74.3		
92-59-130	4.17	4.66	4.02	10.83	6.02	59.22	86.6	74.8		
92-59-131	2.43	2.7	2.15	6.43	2.84	36.55	86.9	74.7		
92-59-132	4.49	4.98	4.05	11.57	6.69	67.25	86.8	75.3		
92-59-133	2.26	2.52	2.2	6.37	2.82	34.76	86.1	74.6		
92-59-134	2.56	2.84	2.43	6.77	3.41	42.26	86.6	74.4		
92-59-135	1.98	2.21	1.69	5.01	2.2	32.02	85.5	74.4		
92-59-136	3.54	3.91	3.11	8.6	4.32	53.29	86.8	75		
92-59-137	4.99	5.55	3.66	11.92	6.46	61.49	90.7	77.7		
92-59-138	5.33	5.97	3.93	12.87	6.28	61.48	89.8	76.8		
92-59-139	4.26	4.72	3.28	10.49	6.19	63.37	88.7	76.3		
92-59-140	2.84	3.17	2	7.73	4.3	58.29	87.4	75.5		
92-59-141	2.56	2.87	1.57	6.97	0	48.93	91.3	77.1		
92-59-142	2.89	3.22	2.48	7.78	4.34	50.88	88.6	76.1		
92-59-143	3.01	3.32	2.07	6.62	2.92	52.19	87.1	75.1		
92-59-149	5.8	6.37	3.57	11.8	6.23	60.69	87.8	76.2		
92-59-150	7.35	8.14	4.47	15.43	7.73	63.88	87.6	76.4		
92-59-151	7.4	8.18	4.52	15.74	7.96	64.73	87.6	67.4		
92-59-152	5.6	6.2	3.94	12.45	7.36	67.37	86.8	75		
92-59-153	5.98	6.66	3.23	12.27	5.89	62.07	87	76		
92-59-154	7.2	7.98	4.17	13.79	5.64	55.29	87.5	76.2		
92-59-155	5.91	6.54	3.76	12.4	6.23	55.61	88.7	76.7		
92-59-156	8.11	8.92	4.94	16.39	7.48	57.47	88.2	76.9		
92-59-157	2.58	3	2.18	6.15	2.89	37.43	86	74		
92-59-158	4.96	5.47	4.04	10	5.6	46.75	87.5	74.9		
92-59-159	6.75	7.42	4.93	13.29	7.14	56.57	85.2	75.2		
92-59-160	6.43	7.1	5.91	14.02	5.55	42.5	83	73.7		

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Feedstock	Catalyst Identification	Catalyst Type	C/O Ratio	Reaction Temperature, °F	Recovery Wt%	Conversion Wt%	Coke Wt%
92-59-161	Sasol Wax	1397 (1:3d1872)	Eq. USY	0.81	969	94.6	83.2	0.53
92-59-162	Sasol Wax	1891 (1:15d1872)	St. HZSM-5	0.22	880	87	72.3	0.34
92-59-163	Sasol Wax	1891 (1:15d1872)	St. HZSM-5	0.22	880	87.1	65.1	0.34
92-59-164	Sasol Wax	1891 (1:15d1872)	St. HZSM-5	0.22	880	86.9	64.8	0.33
92-59-165	Sasol Wax	1891 (1:15d1872)	St. HZSM-5	0.22	880	87	61.8	0.33
92-59-166	Sasol Wax	1891 (1:15d1872)	St. HZSM-5	0.22	880	86.5	6.66	0.34
92-59-173	Sasol Wax	14040- 50-8 (1:3d1872)	St. USY	0.83	970	90.3	86.3	0.42
92-59-174	Sasol Wax	14040-/50-9 9669-153 #2st (1:3	St. Matrix	0.84	969	87.5	58.1	0.15
92-59-175	Sasol Wax	14040-/50-10 9669-154 (1:3187	St. 10% USY	3.21	969	92.7	70.6	0.18
92-59-176	Sasol Wax	14040-/50-11 9669-155 (1:3187	St. 10% Beta	0.84	970	92.2	55.9	0.14
92-59-177	Sasol Wax	14040- 50-8 (1:3d1872)	St. USY	0.85	880	87.8	85.7	0.61
92-59-178	Sasol Wax	14040- 50-9 (1:3d1872)	St. Matrix	0.81	881	90.6	43.1	0.3
92-59-179	Sasol Wax	14040- 50-10 (1:3d1872)	St. 10% USY	0.81	881	92.3	87	0.38
92-59-180	Sasol Wax	14040- 50-11 (1:3d1872)	St. 10% Beta	0.78	880	94.4	34.8	0.24
93-59-005	LaPorte	1397	Eq. USlt	1.07	974	93.1	92.3	3.01
93-59-006	LaPorte	1891 (1:3d1872)	St. HZSM-5	0.96	970	81.1	81.8	0.75
93-59-007	LaPorte	1875 (1:3d1872)	St. Beta	0.99	970	83.5	88.2	1.19
93-59-008	LaPorte	1397 (1:3d1872)	Eq. USY	0.87	972	87.2	86.4	87.2
93-59-009	LaPorte	1891 (1:3d1872)	St. HZSM-5	0.97	970	84.1	87.7	0.19
93-59-010	Sasol Wax	1875 (1:3d1872)	St. Beta	1.23	972	88.6	90.9	0.42
93-59-011	Sasol Wax	1397 (1:3d1872)	Eq. USY	1.17	971	92.4	94	0.67
93-59-012	Sasol Wax	1875 (1:3d1872)	St. Beta	0.86	970	91.8	79.8	0.38
93-59-013	Sasol Wax	1397 (1:3d1872)	Eq. USY	0.88	969	94	85.8	0.54
93-59-015	LaPorte	1891 (1:3d1872)	St. HZSM-5	1.00	970	77.7	80.1	0.44
93-59-016	LaPorte	1875 (1:3d1872)	St. Beta	0.86	971	88	83.4	1.71
93-59-017	LaPorte	1397 (1:3d1872)	Eq. USY	0.91	971	88.9	84.9	1.32
93-59-018	Sasol Wax	1891 (1:3d1872)	St. HZSM-5	1.32	971	65.6	74.7	0.25
93-59-019	Sasol Wax	1875 (1:3d1872)	St. Beta	1.15	969	77.5	57.2	0.42
93-59-021	Sasol Wax	9363005 (CCC1397)	Eq. USY	5.00	969	89.9	86.2	2.6
93-59-022	Sasol Wax	1397 (1:3d1872)	Eq. USY	1.59	969	69.4	75.7	0.95
93-59-023	Sasol Wax	1397 (1:3d1872)	Eq. USY	0.56	970	91.9	56.4	0.45
93-59-024	Sasol Wax	1397 (1:3d1872)	Eq. USY	0.75	970	83.1	62	0.46
93-59-025	Sasol Wax	1397 (1:3d1872)	Eq. USY	0.77	970	86.3	67.2	0.44
93-59-026	Sasol Wax	1397 (1:3d1872)	Eq. USY	0.81	970	89.3	78.2	0.43
93-59-027	Sasol Wax	1397 (1:3d1872)	Eq. USY	0.71	970	92.2	61.6	0.38

APPENDIX A -- MYU Run Data (Cont'd)

Run Number	Propylene Wt%	Total C3 Wt%	Isobutylene Wt%	Total C4 Wt%	Isoamylenes Wt%	C5-430 deg. F Wt%	RON	MON	Conversion by Sim. Dis., Wt%	Comments
92-59-161	7.89	8.74	4.81	15.74	7.21	57.28	85.1	76		
92-59-162	15.71	17.92	9.18	24.88	7.38	27.93	86.1	76.5		
92-59-163	13.72	15.58	8.15	22.11	6.8	26.08	85.9	76.6		
92-59-164	13.18	14.87	7.51	20.19	6.75	28.53	86.3	76.3		
92-59-165	12.85	14.42	7.33	19.76	6.56	26.49	86.3	76.4		
92-59-166	12.52	14	7.07	18.91	6.63	32.61	84.7	75.2		
92-59-173	7.38	8.17	4.99	15.07	7.54	61.67	85.7	76		
92-59-174	6.18	6.81	5.55	13.64	5.09	36.62	83.6	74		
92-59-175	5.11	5.6	3.62	9.88	4.75	54.21	85.3	74.7		
92-59-176	3.91	4.46	3.15	8.46	3.96	41.68	85.8	74.3		
92-59-177	5.51	6.12	4.3	13.1	6.91	65.48	84.3	75.2		
92-59-178	2.82	3.14	3.14	7.69	3.17	31.68	83.7	73.5		
92-59-179	6.62	7.34	5.99	16	8.86	62.65	86.1	75.7		
92-59-180	1.71	1.98	1.85	4.95	2.57	27.33	84.8	73.5		
93-59-005	2.82	3.31	3.1	12.63	8.68	72.91	89.6	79.6		10 runs
93-59-006	20.55	22.99	9.14	25.29	6.15	28.45	92	80.1		
93-59-007	9.6	10.37	7.69	18.23	8.83	57.18	88.3	76.4		
93-59-008	86.4	7.33	4.15	13.48	7.58	63.42	87.8	77.3		
93-59-009	18.64	21.33	10.13	26.93	7.64	36.94	84.4	75.5		
93-59-010	14.89	16.33	9.15	25.67	7.91	47.32	85.5	76.5		
93-59-011	11.01	12.1	6.08	19.65	8.19	60.52	84.8	76.9		
93-59-012	10.78	11.82	7.59	20.48	7.72	46.25	86	75.8		
93-59-013	7.95	8.74	4.96	16.3	7.94	59.53	86.7	76.7		
93-59-015	17.11	19.39	9.26	24.74	6.05	31.23	87.7	77.4		
93-59-016	8.76	9.43	5.92	14.07	6.81	57.16	88.6	76.6		
93-59-017	7.01	7.7	3.67	11.9	6.17	62.87	87.8	77.2		
93-59-018	16.53	18.79	8.94	24.33	6.74	28.23	85.7	76.3		
93-59-019	6.74	7.38	4.19	11.61	3.95	37.04	84.6	75.8		
93-59-021	9.89	11.23	4.16	17.86	4.57	53.11	86.4	78.7		
93-59-022	9.03	10	4.43	14.52	3.49	49.38	85.3	76.5		
93-59-023	3.79	4.28	1.88	5.89	2.08	45.15	84	74.5		
93-59-024	5.18	5.75	3.71	11.97	4.93	43.28	86.9	76.6		
93-59-025	5.2	5.75	3.69	11.93	5.41	48.52	87	76.5		
93-59-026	6.34	7	4.42	13.7	6.7	56.5	87	76.4		
93-59-027	4.65	5.17	2.78	8.68	4.02	46.88	86	75.5		

APPENDIX B

Economic Calculations

March 22, 1994

M. M. Schwartz
Mail Station H-4

INFORMATION:

**CATALYTIC CRACKING OF FISCHER-TROPSCH LIQUIDS TO TRANSPORTATION FUELS
PRODUCT VALUE CALCULATIONS -- PROJECT 70-6184-45**

As requested by the U.S. Department of Energy, this letter details additional information and sample calculations used in determining the values of refining intermediates. This information was previously used in the economic evaluation of the catalytic cracking of Fischer-Tropsch liquids to transportation fuels. Sample calculations for propylene value as a feedstock to a di-isopropyl ether (DIPE) unit and iso-butylene value as a feedstock to an MTBE unit are given.

Barring any additional questions from the DOE, this concludes economic analysis work for this DOE contract.

SUMMARY

A breakdown of the costs for DIPE, MTBE and TAME are shown in Table I, attached. Details for DIPE and MTBE costs follow.

BASIC PRICE AND PROPERTY INFORMATION

The pertinent price and product quality information that was assumed is shown in the table below:

Component	price, cpg	RVP, psi	Octane, (R+M)/2
Unleaded regular gasoline (ULR)	55.7	9.0	87.0
MTBE	86.0	9.0	108.0
n-Butane	29.8	65.0	90.55
Methanol	50.0		

OCTANE AND RVP VALUE

To value refinery intermediates, it is necessary to limit properties from which to determine value. In this study, octane and RVP properties were chosen. Using the price difference between ULR and MTBE, one can calculate the value of octane, by solving for x in the equation below:

$$86.0 - 55.7 = (108.0 - 87.0) * x$$

Solving the above equation for x gives an octane value of 1.44 ¢/octane-gallon.

From the price difference between n-butane and ULR, and using the octane value calculated above, a value for RVP can be determined from the equation:

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$$29.8-55.7=(90.55-87)*1.44+(65-9)*x$$

Solving for x gives an RVP value of -0.55 ¢/RVP-gallon.

These two values are used in subsequent calculations to determine the value of refining intermediates.

DIPE VALUE CALCULATION

From the May 25, 1992 issue of the *Oil and Gas Journal*, the following information was obtained for DIPE production from propylene:

Item	value
DIPE RVP, psi	5.0
DIPE RON	110
DIPE MON	97
DIPE R+M/2	103.5
DIPE density, lb/gallon	6.4
operating costs, cpg	30.0

Using the values for octane and RVP that were previously calculated, a value for DIPE can be calculated by solving for x in the equation below:

$$x=55.7+(103.5-87)*1.44+(5-9)*(-0.55)$$

yielding a DIPE value of 81.7 cpg.

To determine the value of propylene to a DIPE plant, the kinetics of the chemical reaction must be considered. The equation for producing DIPE from propylene is shown below:

From molecular weights, 84 pounds of propylene plus 18 pounds of water yield 102 pounds of DIPE. On a volume basis, 1.211 gallons of propylene yield 1.0 gallon of DIPE. To calculate propylene value to a DIPE unit, consider DIPE value and operating costs and solve for x in the equation shown below:

$$1.211*x+30=81.7$$

The above equation yields 42.7 cpg for propylene value to a DIPE unit.

ISO-BUTYLENE VALUE TO AN MTBE UNIT

From the assumed price structure, MTBE has a value of 86 cpg. The kinetics of the iso-butylene to MTBE reaction are:

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From molecular weights, 56.1 pounds of iso-butylene plus 32.0 pounds of methanol produce 88.1 pounds of MTBE. On a volume basis, 0.792 gallons of iso-butylene plus 0.341 gallons of methanol are required to produce one gallon of MTBE. Operating costs for an MTBE unit were obtained from the November 1992 issue of *Hydrocarbon Processing*, and are shown in the table below:

Utility	consumption/bbl of ether produced	utility price
Electricity, KW-hr	1.0	2.28 ¢/kw-hr
all steam, pounds	296.0	\$2.0/MM BTU

Using the above information gives an MTBE operating unit cost of 1.28 cpg of ether produced. Note that this does not include the methanol cost. To determine the value of iso-butylene to an MTBE unit, consider MTBE value and all operating costs, and solve for x in the equation below:

$$0.792 * x + 0.341 * 50 + 1.28 = 86$$

The value for iso-butylene determined from the above equation is 85.4 cpg.

Values for other refinery intermediates was determined by a similar method. Please advise if you have any questions.

Jim Nicholas

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Attachment

C. D. Sorrentino, H-4
File 240.1.008

TABLE I--SUMMARY OF ETHER COSTS

Ether	Value/cost to gasoline, cpg	Refining Intermediate	Methanol	Operating Cost, cpg
DIPE	81.7	Source: Propylene Amount/gallon of ether: 1.211 Unit price, cpg: 42.7 Cost/gallon of ether: 51.7	Amount/gallon of ether: 0.0 Unit Price, cpg: 50.0 Cost/gallon of ether: 0.0	30.0
MTBE	86.0	Source: iso-butylene Amount/gallon of ether: 0.792 Unit price, cpg: 85.4 Cost/gallon of ether: 67.6	Amount/gallon of ether: 0.341 Unit Price, cpg: 50.0 Cost/gallon of ether: 17.1	1.3
TAME	84.0	Source: 2-methyl-1-butene Amount/gallon of ether: 0.785 Unit price, cpg: 86.5 Cost/gallon of ether: 67.9	Amount/gallon of ether: 0.296 Unit Price, cpg: 50.0 Cost/gallon of ether: 14.8	1.3