

## 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 C<sub>1</sub>-C<sub>8</sub> 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 (C<sub>5</sub>-C<sub>50</sub>) and a small quantity of oxygenates and olefins. The gasoline range C<sub>5</sub>-C<sub>12</sub> product fraction consists of linear paraffins and olefins of low octane number. The distillate fraction, C<sub>12</sub>-C<sub>18</sub>, is an excellent quality fuel. The largest product fraction, C<sub>18</sub>+, 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

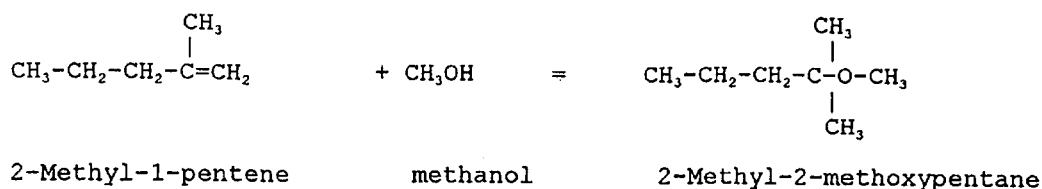
produce conventional gasoline and distillate fuels. There are no reported studies of the catalytic cracking processing of F-T liquids to produce C<sub>3</sub>-C<sub>6</sub> olefins as feedstocks for the synthesis of gasoline range ethers and alcohols.

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.<sup>(1,2)</sup> 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.

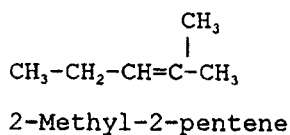
The reaction of olefin with methanol to form ether is catalyzed by acidic ion exchange resins at low temperatures (100-200°F) and moderate pressures (100-400 psig). These etherification reactions were discovered in 1907.<sup>(3)</sup> Detailed catalysis studies of these reactions have only recently appeared in the scientific literature.<sup>(4)</sup> 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.<sup>(4)</sup> 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 C<sub>6</sub> 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 C<sub>6</sub> olefins. The Etherol process<sup>(5)</sup> produces a mixed ether product from C<sub>4</sub>-C<sub>7</sub> reactive iso-olefins in naphtha. One report<sup>(6)</sup> provided some information about the reaction of C<sub>6</sub> iso-olefins with methanol to produce the higher ethers, MTHE's, methyl tertiary hexyl ethers. Several C<sub>6</sub> olefins can react with methanol to produce three different THME ethers:

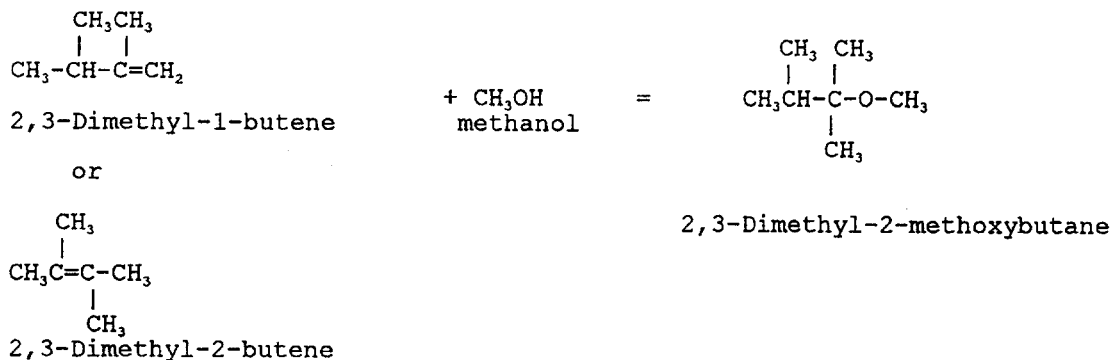
Scheme 1:



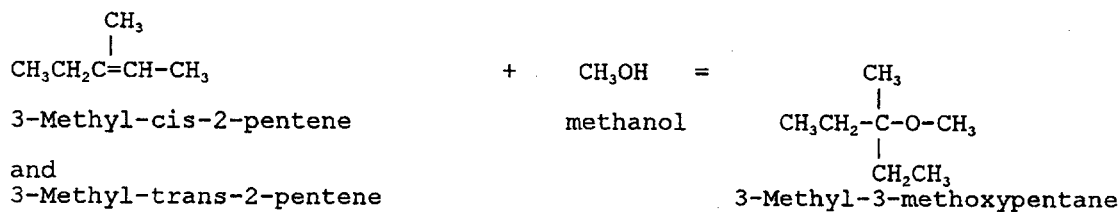
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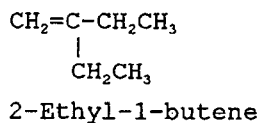
Scheme 2:



Scheme 3:



or



The reactions of these C<sub>6</sub> 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<sub>4</sub>-C<sub>8</sub> 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<sub>5</sub>-C<sub>8</sub> Ethers and C<sub>3</sub>-C<sub>8</sub> 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).

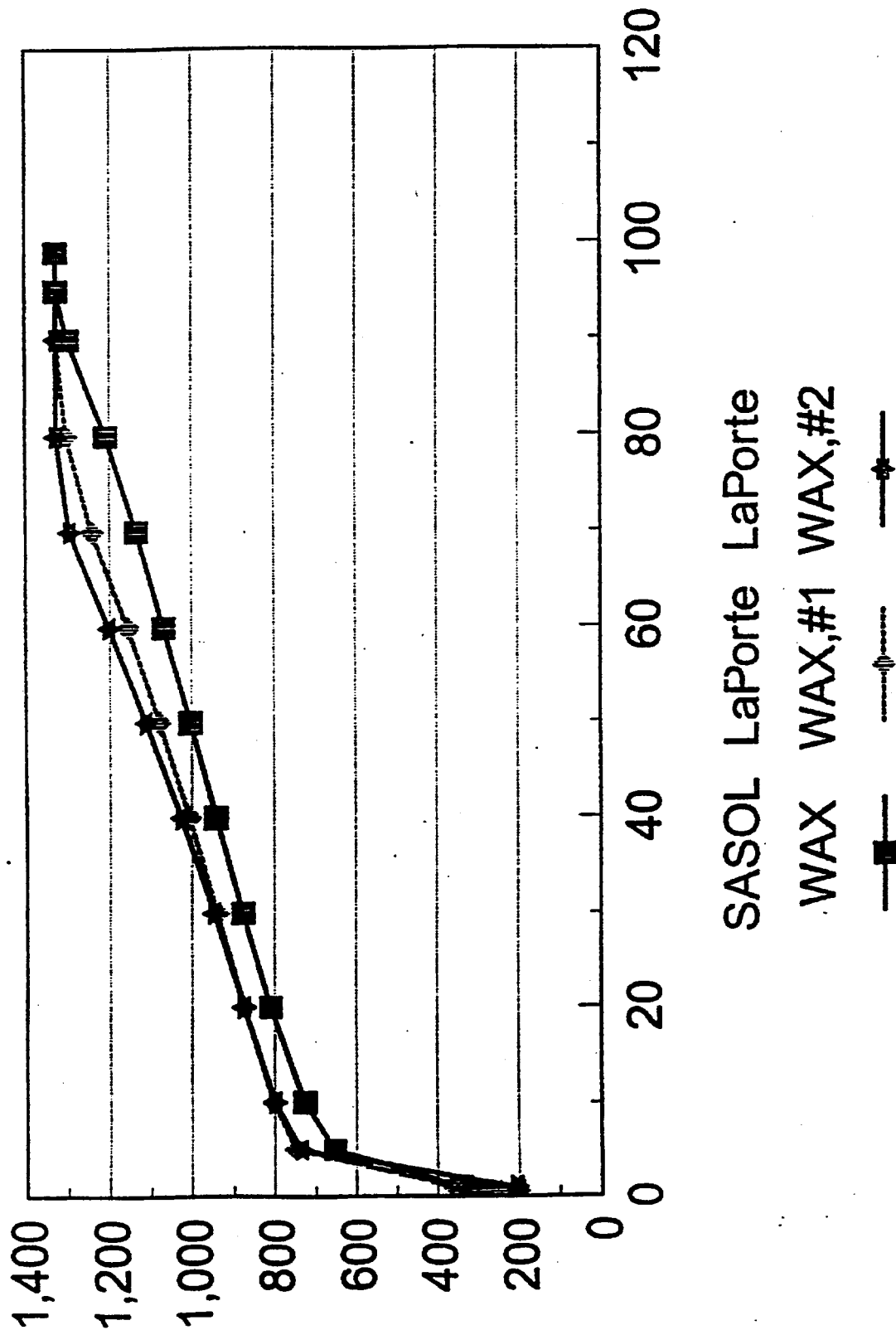
## 3.0 FEEDSTOCK CHARACTERIZATION and EQUIPMENT CALIBRATION (TASK 2)

### 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**

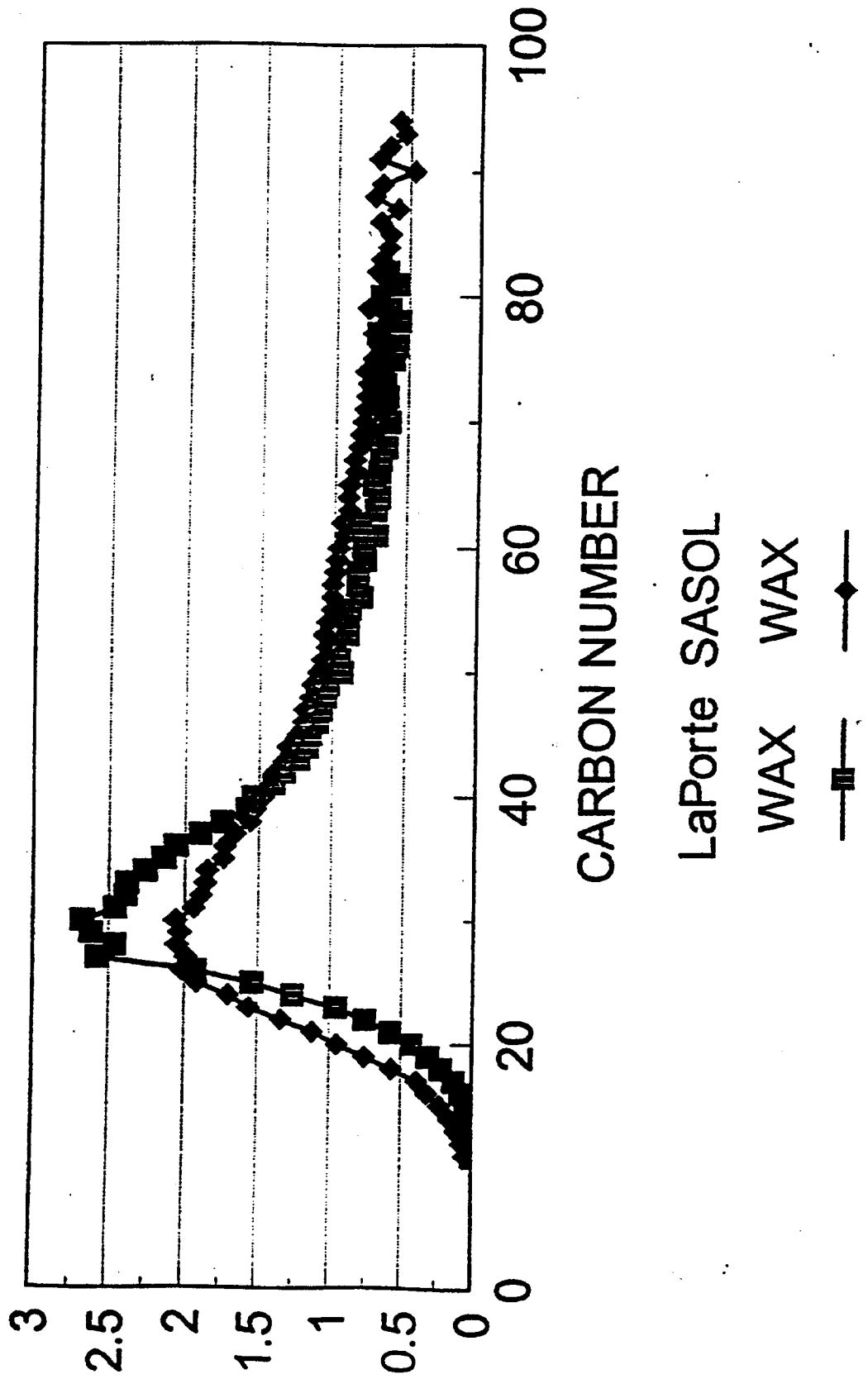


TABLE I

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.<sup>(7)</sup>

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%.<sup>(7)</sup> 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 C<sub>5</sub>-C<sub>12</sub> 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)	Si = 1,710	SiO <sub>2</sub> = 3,659
of Ash:	K = 1,350	K <sub>2</sub> O = 1,626
	Fe = 13,900	Fe <sub>2</sub> O <sub>3</sub> = 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.

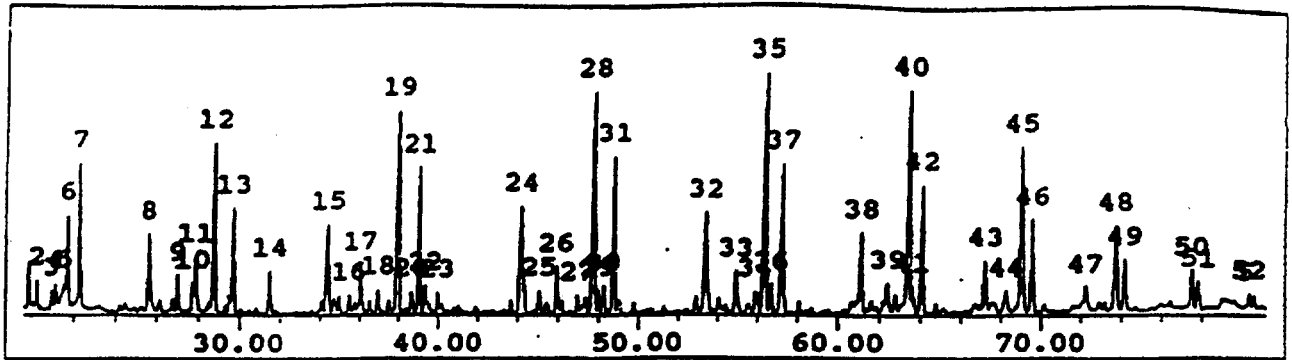
#### 4.1 Sasol Wax Feedstock

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% C<sub>3</sub> and C<sub>4</sub> using these conditions, versus over 85% conversion, 2.2% coke, and over 41% C<sub>3</sub> and C<sub>4</sub> 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 <sub>2</sub> -	2.6	1.8
C <sub>3</sub> + C <sub>4</sub>	11.6	31.4
C <sub>5</sub>	6.9	15.9
C <sub>6</sub> -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 <sub>4</sub> -	2.6	5.9
C <sub>5</sub> -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

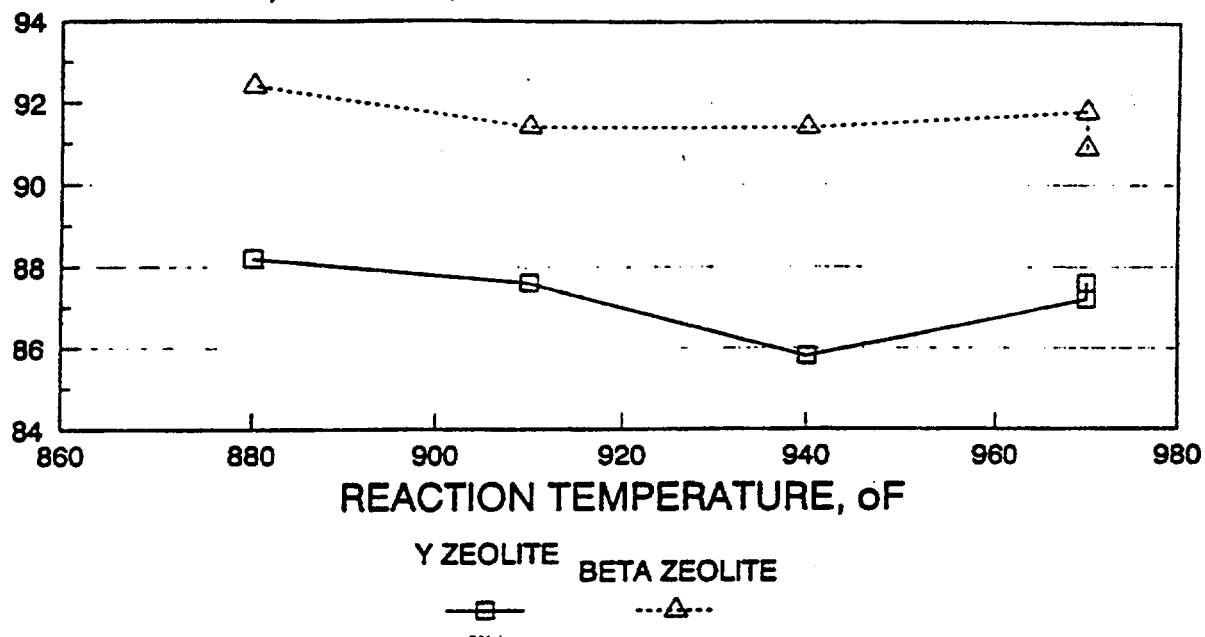


FIGURE 4

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

iC4=, iC5=, WT. %

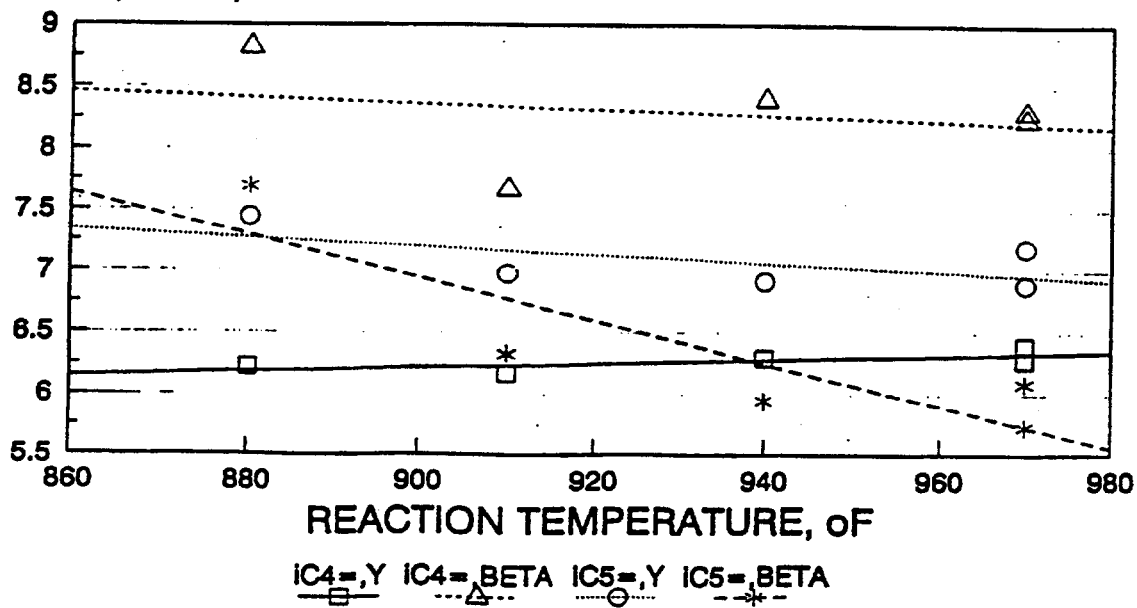


TABLE VI  
 MYU TESTS: EFFECT OF CATALYST TO OIL RATIO

Equilibrium Zeolite Y, CCCI397					Steamed Zeolite Beta, CCCI875					Steamed HZSM-5 CCCI891				
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<sub>3</sub>+C<sub>4</sub> gas (Figure 10), and much lower C<sub>5</sub>-430 °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°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<sub>3</sub> and C<sub>4</sub> 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**

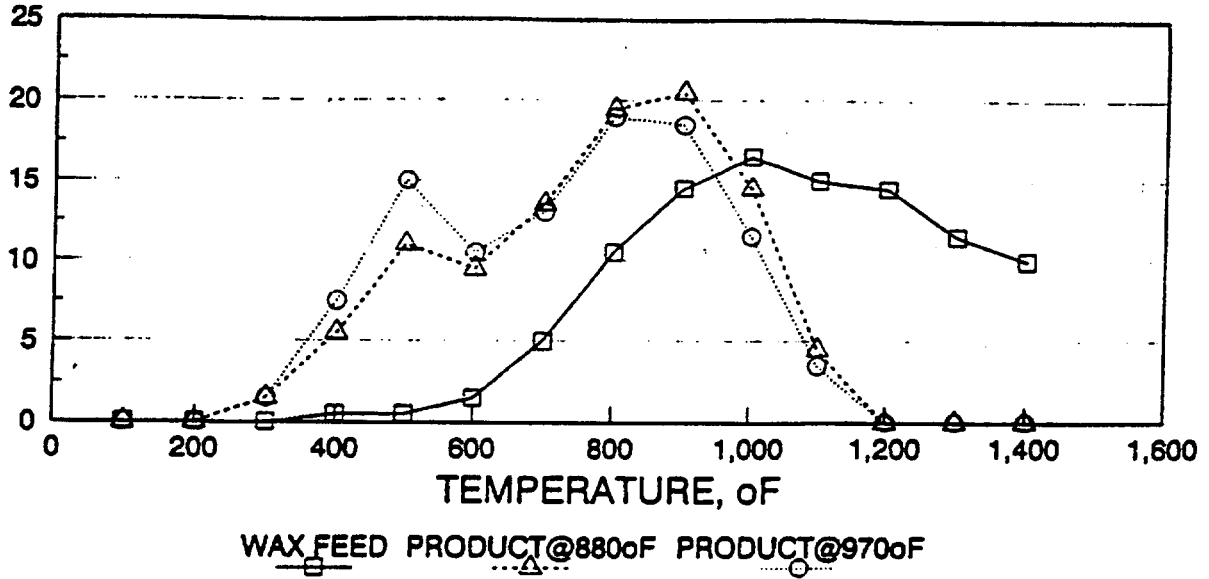


FIGURE 6

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

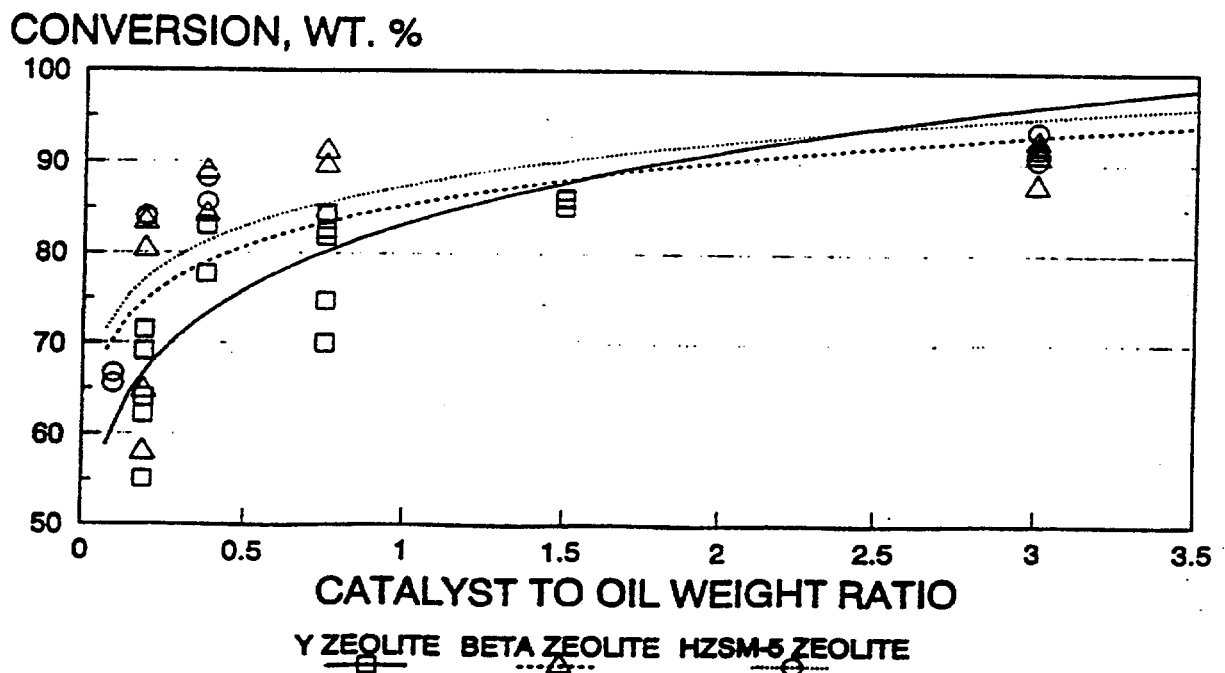




FIGURE 7

SASOL WAX MYU DATA:  
EFFECT OF CONVERSION ON ISOAMYLENE YIELD

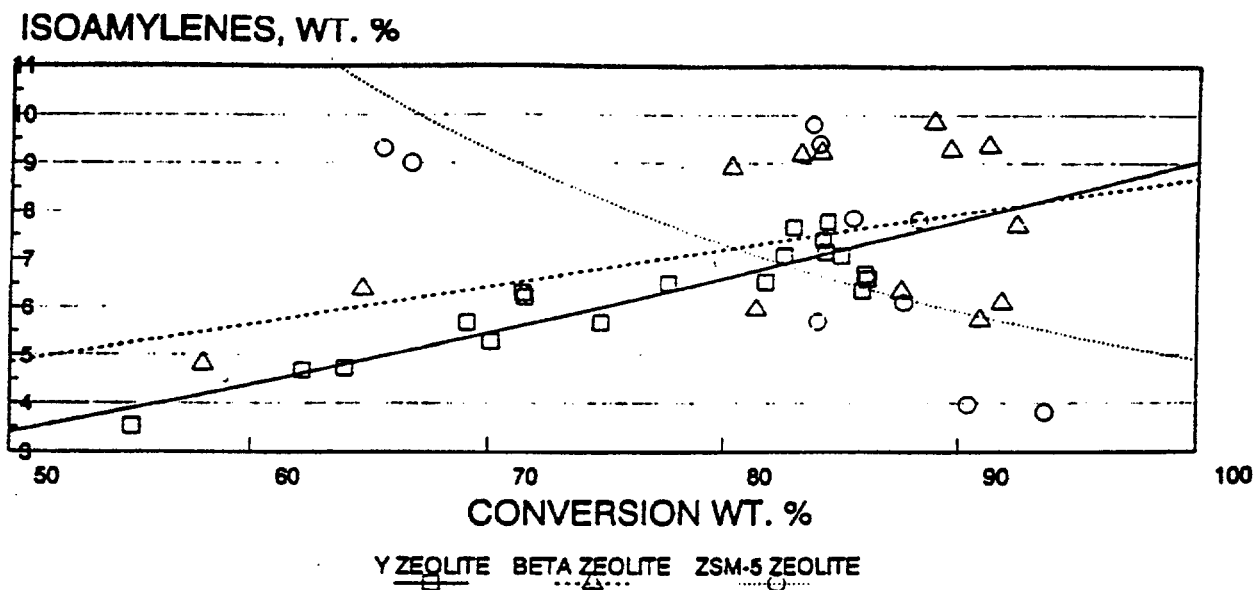


FIGURE 8

SASOL WAX MYU DATA:  
EFFECT OF CONVERSION ON ISOBUTYLENE YIELD

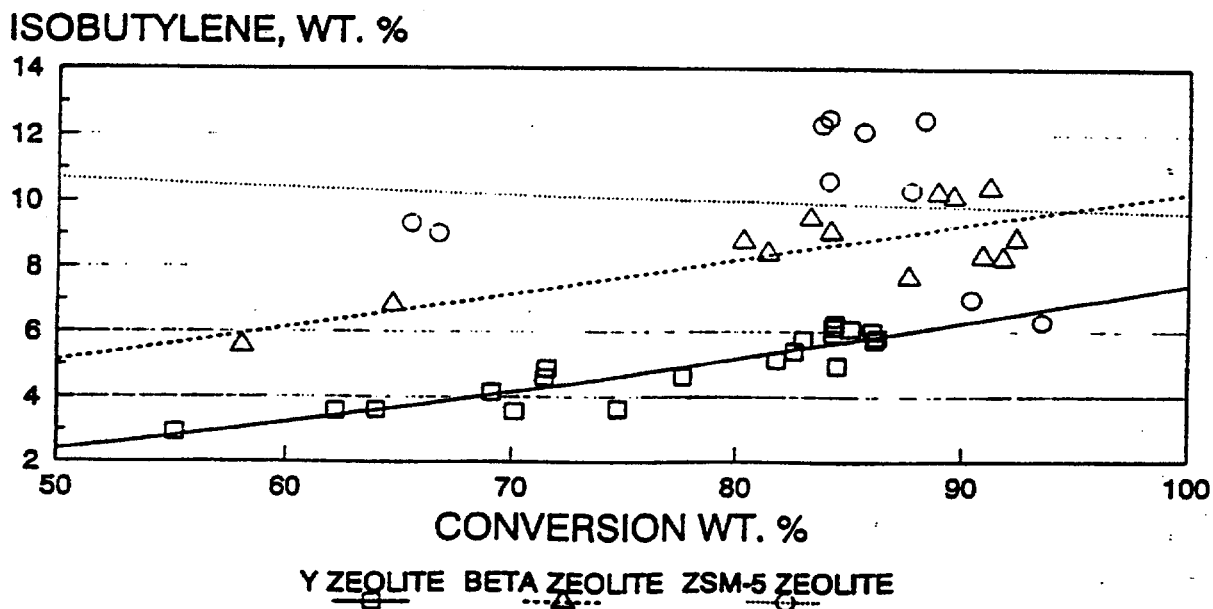


FIGURE 9

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

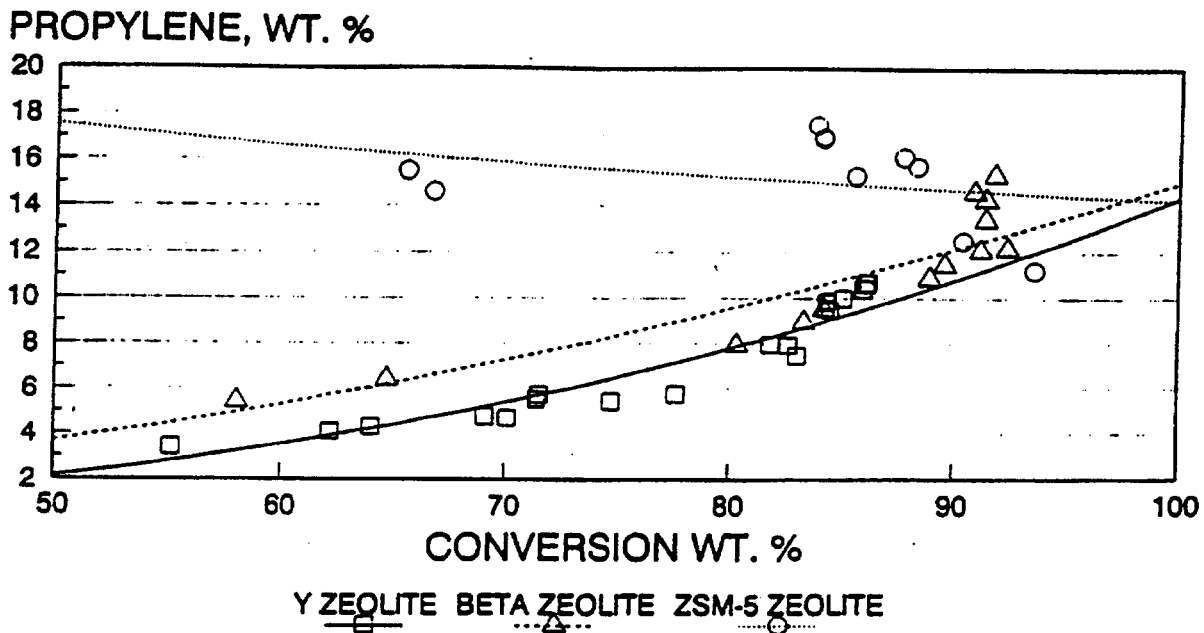


FIGURE 10

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

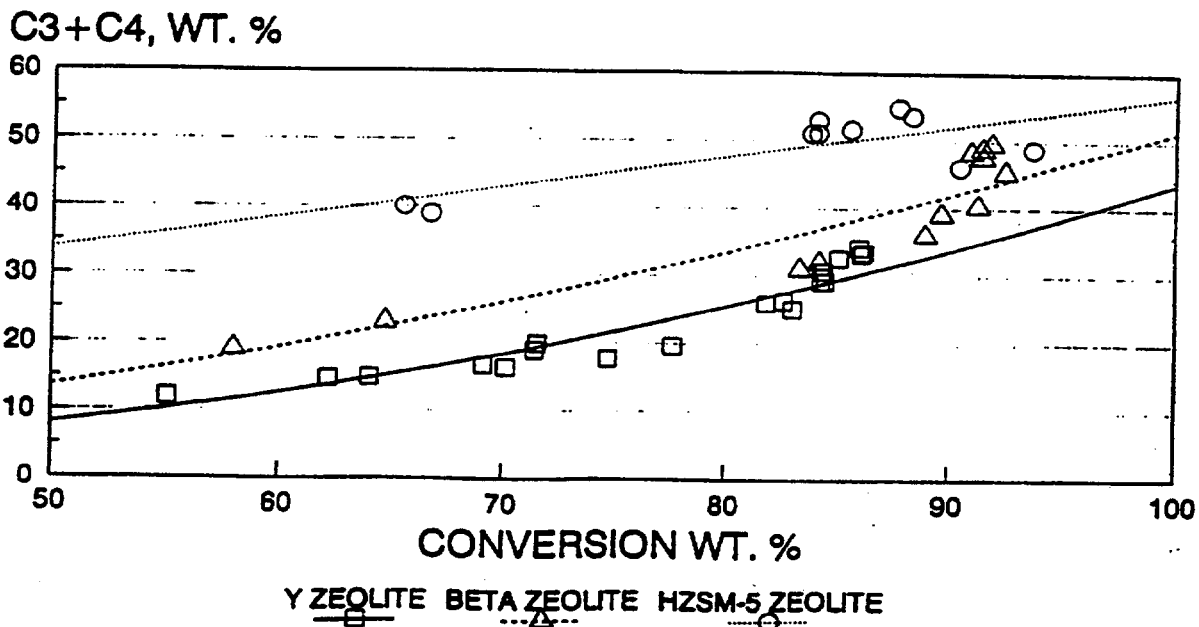


FIGURE 11

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

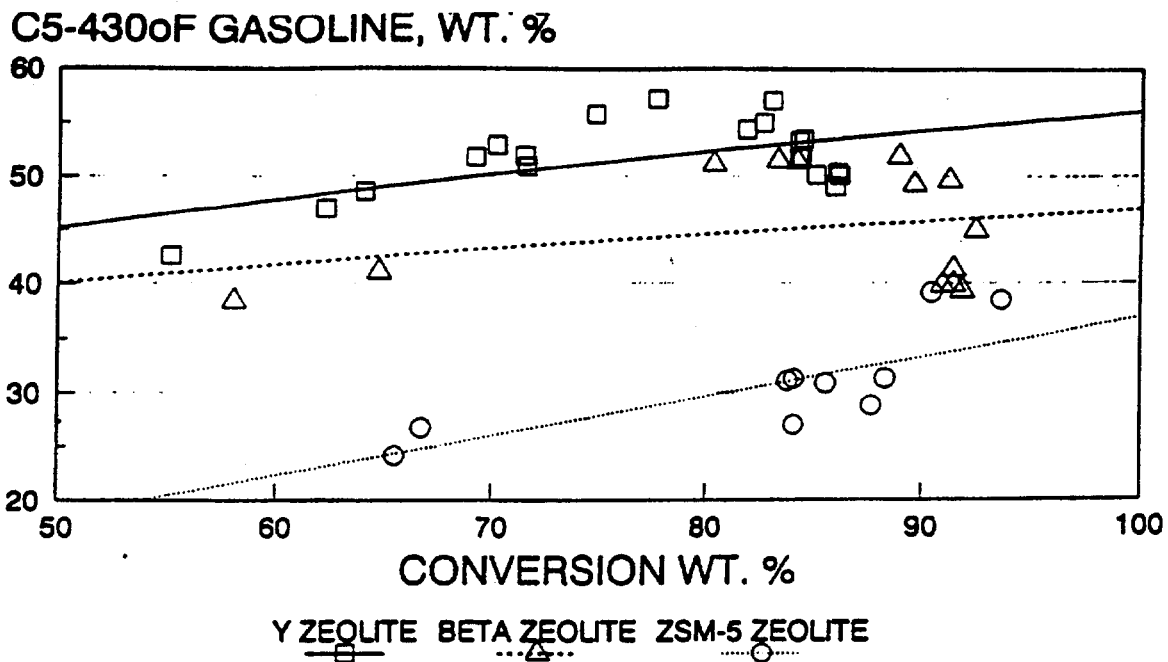


FIGURE 12

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

