DOE/PC/92121--T12

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HIGH CONVERSION OF COAL TO TRANSPORTATION FUELS
FOR THE FUTURE WITH LOW HC GAS PRODUCTION

Progress Report No. 11
(Covering the period April 1 - June 30, 1995)

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Prepared for the United States Department of Energy
under Contract No. DE-AC22-92PC92121

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Mark P. Dvorscak, 8-25-95

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July, 1995

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PERSONNEL

In addition to the Principal Investigators, research personnel participating in this project during the reporting period are:

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OBJECTIVES AND SCOPE OF THE RESEARCH PROJECT

An announced objective of the Department of Energy in funding this work, and other current research in coal liquefaction, is to produce a synthetic crude from coal at a cost lower than \$30.00 per barrel (Task A). A second objective, reflecting a recent change in direction in the synthetic fuels effort of DOE, is to produce a fuel which is low in aromatics, yet of sufficiently high octane number for use in the gasoline-burning transportation vehicles of today. To meet this second objective, research was proposed, and funding awarded, for conversion of the highly-aromatic liquid product from coal conversion to a product high in isoparaffins, which compounds in the gasoline range exhibit a high octane number (Task B).

Experimental coal liquefaction studies conducted in a batch microreactor in our laboratory have demonstrated potential for high conversions of coal to liquids with low yields of hydrocarbon (HC) gases, hence small consumption of hydrogen in the primary liquefaction step. Ratios of liquids/HC gases as high as 30/1, at liquid yields as high as 82% of the coal by

weight, have been achieved. The principal objective of this work is to examine how nearly we may approach these results in a continuous-flow system, at a size sufficient to evaluate the process concept for production of transportation fuels from coal.

A continuous-flow reactor system (1/2 inch inside diameter) is to be designed, constructed and operated. The system is to be computer-operated for process control and data logging, and is to be fully instrumented. The primary liquid products will be characterized by GC, FTIR, and GC/MS, to determine the types and quantities of the principal components produced under conditions of high liquids production with high ratios of liquids/HC gases. From these analyses, together with GC analyses of the HC gases, hydrogen consumption for the conversion to primary liquids will be calculated. Conversion of the aromatics of this liquid product to isoparaffins will be investigated, to examine the potential for producing a transportation fuel from coal with satisfactory octane rating but low in aromatic content.

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ACTIVITY FOR THE REPORTING PERIOD

TASK A

In the prior report No. 10, various difficulties with the fluidized bed were discussed, and the steps taken to correct the problems. Each modification corrected an aspect of the bed operation. It should be noted that such difficulties in operation of a fluidized bed of the size incorporated into this reactor system were not unexpected. The bed was designed based upon data from operation of a small fluidized bed of about 10-15% of the cross-sectional area of this bed. It was anticipated that modifications would be required.

Experiments in the micro-reactor which yielded ratios of liquids/hydrocarbon (HC) gases of 30/1, were obtained at a temperature of 500 °C, with a time in the heated reactor of 3 seconds. Operating the present reactor with a gas flow rate of 7.2 feet per second (about 1.4 times minimum turbulent flow velocity), the design capacity of the recycle compressor, would lead to a residence time in the reactor as presently configured of about 5 seconds. We also wanted to operate at 500 °C. The bed now appeared to be ready for an experimental run.

For the first experimental run, the coal feed hopper was charged with 200 grams of a Wyodak sub-bituminous coal, crushed and screened, with a fraction selected which passed through a 65 mesh screen (Standard Tyler Screen Series) and was retained on a 100 mesh screen. The heaters in both the hydrogen preheater section and the main reactor section were turned on. The bed appeared to be fluidizing in a satisfactory manner, and the

initial heat-up rates in both sections were in the anticipated range. At the rates observed, about four to five hours would be required for heat-up to 550 °C in the hydrogen preheater section and 500 °C in the main reactor section.

When the temperature in the fluidized bed of the main reactor section reached about 375 °C, the heat-up rate slowed to a very low value. After about two more hours it appeared we may not reach the desired temperature of 500 °C, and it was apparent there was a problem. The heaters were turned off, and the bed allowed to cool to room temperature. Examination revealed that three heaters in the main reactor section and three heaters in the hydrogen preheater section had burned out. It was noted that although the bed overall appeared to be fluidizing properly, there was an accumulation of sand immediately around the heaters which had burned out which was not fluidizing. The result was overheating, and ultimate burnout, of those heaters.

Several weeks, growing into months, had been expended in attempts to obtain completely satisfactory operation of the fluidized bed. Although genuine progress was being realized, it was becoming urgent to begin to obtain good experimental data. A decision was made to remove all of the sand from both sections of the reactor system, operating both the hydrogen preheater and the main reactor section as separate ovens. The main reactor section was first operated in this manner, and it was found that heat-up rates were uniform throughout the length of the reactor tubes, with heat-up to 500 °C achieved in about four hours. The full system, including the hydrogen preheater section, was then

tested. It was realized that since cold hydrogen was being admitted to the hydrogen preheater, additional heat-up length may be required. An additional length of preheater tube was added to the front of the preheater, running the full length of the main reactor section, within that section, and return, before entering the coiled tube section of the preheater. This configuration allowed for heat-up of the hydrogen preheat section to 550 °C in the same time required for heat-up of the main reactor section, namely about four hours.

An experiment was then conducted with coal feed and hydrogen feed in a manner as planned. The experiment was conducted for 30 minutes. Everything seemed to be operating as planned. Ninety percent of the coal charged to the feed hopper passed through the reactor, namely 180 grams. Of this amount, 88 grams appeared as liquids, representing 49% by weight of the coal fed. The hydrocarbon gases were determined to be 3.4 grams, representing a ratio of liquids/HC gases of 26/1. The solids, following extraction of the liquids, appeared the same as the solids from a single pass in the microreactor. In those experiments, a second pass through the reactor produced an additional conversion of about 50% of the unreacted solids. It is anticipated that a second pass of the unreacted solids from this system, in a manner to simulate solids recycle, will lead to a total conversion of between 70% and 75%, with overall ratio of liquids/HC gases still in the range of 26/1, the same relationship of the two passes observed in the microreactor. We have not as yet attempted a run with the recycled unreacted solids.

In most work reported in the literature where small amounts of hydrocarbon gases are produced in a system of very high hydrogen flow, the amounts of gases are reported by difference. However, inasmuch as the ratio of liquids/HC gases is a major feature of this work, we must know accurately how much hydrocarbon gases are produced. In the gas chromatograph, the relative amounts of each HC gas can be accurately measured. There remains the need to know the absolute quantity represented by each peak. A standard was prepared containing amounts of each HC gas expected to be in our gas product, mixed with large amounts of hydrogen. This standard was then injected into the gas chromatograph, and the area under each peak related to the absolute amounts of each HC component. Comparison of these peaks with the peaks obtained from injection of the gases from our reactor permitted accurate calculation of the absolute amounts of each HC component in the gas. We therefore are confident of the ratios of liquids/HC gases reported from our experiments.

When work was commenced to design, fabricate and operate a system of the size of our present system, it was anticipated that we may obtain conversions of the order of 45-50% in a first pass, with an additional 20-25% (based upon the coal fed) in a second pass, thus simulating recycle of unreacted solids, for a total conversion of about 70%. We further anticipated an overall ratio of liquids/HC gases from the two passes of about 20/1. We have exceeded those expectations.

In the work of Exxon, involving the EDS liquefaction process, it was determined that a fully self-contained operation

of the reactor system could produce only about 45% of the feed coal as net product leading to a transportation fuel, especially gasoline. The remaining coal would be utilized in production of hydrogen, heat input into the system, and accommodating efficiency losses. In our system with conversions to liquids in a single pass representing about 50% of the feed, this conversion may be adequate without recycle of unreacted solids. It should be noted, as reported in the final report of our prior project, that a simulated distillation of the liquid products reveals a good quality of product, very low in asphaltenes or other high-boiling compounds in the primary product.

Although a very good experimental run was conducted, we are still encountering problems. In the next experimental run, a poor material balance was observed. When the system was opened, it was observed that a large amount of solids, with liquids entrained, had accumulated in the condenser. We have been forced to reduce the water flow in the condenser, hence the amount of cooling to occur there, almost to zero. We also found some coal hold-up in the early part of the reactor tube, about one foot from the point of entrance of the coal into the reactor tube. We have redesigned the hydrogen injection nozzle in a way which we believe will correct that problem. We anticipate that our next experimental run will yield a good material balance.

The data reported above were obtained at a system pressure of 1180 psig. Inasmuch as the microreactor revealed it is possible to conduct the chemistry of this conversion approach at pressures as low as 500 psia, without much loss in conversion or

in ratio of liquids/HC gases, our next experiment is to be conducted at a pressure of 500 psig. We remain optimistic about the potential of this reactor system approach in this pressure region, and the economic advantages which this represents.

TASK B

Experimental work is continuing toward the objective of producing from coal-derived liquids a transportation fuel exhibiting a high octane number (e.g. greater than 85) but low in aromatics. In the previous report, No. 10, the data obtained to date relative to production of isoparaffins were summarized. Using catalysts identified to date, the yields of isoparaffins produced from aromatics in a single stage are very low. Data obtained to date indicate that decalins are a necessary intermediate in the route to formation of isoparaffins, or at least both rings of the two-ring starting material must ultimately be hydrogenated. The data from these experiments do not allow for determining how much cracking of the tetralins may occur to yield substituted single-ring aromatics. In a two-stage process, utilizing decalins as feed to the second stage, yields of isoparaffins are still under 10% by weight of the starting aromatic materials. At the temperatures which produce the highest yields of isoparaffins, dehydrogenation of the decalins to aromatics is high.

More promising than the yields of isoparaffins in experiments utilizing catalysts identified to date are the yields of substituted cycloalkanes. Examples of octane numbers of some

of the substituted cycloalkanes are presented in Table 1, together with iso-octane as a reference.

Table 1

Blending Octane Numbers of Some Cycloalkanes

<u>Compound</u>	<u>Research Octane No.</u>	<u>Motor Octane No.</u>	<u>R + M 2</u>
1,1-Dimethyl- cyclopropane	116	108	112
1,2-Dimethyl- cyclopropane	122	116	119
Cyclopentane	141	141	141
Cyclohexane	110	97	104
Methylcyclo- pentane	107	99	103
Methylcyclo- hexane	104	84	94
Ethylcyclo- pentane	74	63	69
Ethylcyclo- hexane	43	40	42
1,1-Dimethyl- cyclopentane	98	102	100
1,1-Dimethyl- cyclohexane	95	94	94
1,3-Dimethyl- cyclopentane	90	75	83
1,2-Dimethyl- cyclohexane	84	85	84
1,3-Dimethyl- cyclohexane	67	64	66
2,2,4-Trimethyl- pentane (Ref.) (Isooctane)	100	100	100

Source: ASTM Special Technical Publication No. 225

In general, these compounds exhibit octane numbers higher than those of most of the isoparaffins. A few general trends may be observed in the above few compounds, trends which also extend over a wider range of compounds beyond those in the above table:

- - The pentanes generally exhibit a slightly higher octane number than the hexanes;

- - The highest octane numbers are observed when the substitute group is a methyl group, with the octane number dropping rapidly as the length of the substitute group increases beyond a single carbon atom;

- - The octane number is slightly higher when the molecule contains only one methyl group than with two methyl groups;

(Note: This trend is reversed from the trend with aromatics, where the octane number of xylenes is generally higher than for toluene.)

- - The octane number is higher when both methyl groups are attached to the same carbon atom, than when the methyl groups are on separate carbon atoms.

For reasons of simplicity, it is of interest to examine the yields of desirable product from a single stage reactor system. Data are presented using a model compound, methylnaphthalene, as being somewhat representative of molecular types found in primary coal liquids, particularly liquids from coals most responsive to conversion to high yields of liquids. These coals are subbituminous and high-volatile bituminous coals, where the average size of molecules in such liquids ranges between two and three rings in a condensed configuration.

In the prior report, No. 10, experimental data were presented utilizing 11 separate catalysts. Three of those catalysts contained only a hydrogenation function, and the remaining eight exhibited both a hydrogenation and a cracking function. The data presented therein demonstrate the difficulty encountered in attempts to convert aromatics in a single stage. Table 1 from that report is incorporated into this report as Table 2, for ease in following the discussion presented below.

The aromatic compound must be hydrogenated before cracking can occur. If the desired non-aromatic compounds are to be obtained, ultimately both rings must be hydrogenated. Hydrogenation to decalins occurred most extensively in the temperature range 300 °C to 350 °C. At 400 °C, high yields of tetralins appeared in the hydrogenation product, but for most of the catalysts examined only small amounts of decalins were present. However, for most of the catalysts exhibiting both hydrogenation and cracking functions, cracking did not become significant at temperatures below 350 °C.

Results from a few of the catalysts examined are presented in Table 3 herein, using 1-methylnaphthalene as feed. In terms of molecular size (boiling range) and octane number, the substituted cyclopentanes and cyclohexanes are very desirable components for gasoline. The most effective catalyst among those examined for production of these materials in a single stage is NiW/SiO₂-Al₂O₃, identified in the table as Ni-4301. With conversions to liquids being essentially 100%, this catalyst yielded 37% of the 1-methylnaphthalene by weight as substituted

cyclopentanes and cyclohexanes in 10 hours at 325 °C, 30% in 7 hours at 350 °C, and 14.5% in 5 hours at 375 °C. It is noted that as the temperature increased, the amount of aromatics in the product increased, representing 13.6% of the product at 325 °C, 23.5% at 350 °C, and 46.3% at 375 °C. These increased amounts of aromatics are principally the result of increased temperature.

An experiment is now to be performed, using the same catalyst, at 10 hours time, but at 300 °C, where it is expected the yields of substituted cyclopentanes and cyclohexanes may be higher, with perhaps very little or no aromatics. (As a basis for this expectation, see Table 4, Run No. 59, with decalins as feed.) If the product could be free of aromatics, it may be practical to use a single stage reactor system for production of these products of high octane number (Table 1), because of easier separation of the product.

In addition to the yields of substituted cyclopentanes and cyclohexanes, there are significant yields of other cycloalkanes, some of which are in the gasoline boiling range and with high octane numbers. These compounds may find use in gasoline.

The data obtained indicate that decalins are intermediate compounds in the formation of either isoparaffins or cycloalkanes from aromatics. Experiments have been conducted with methyldecalins as feed. In preparation for these experiments, it was desired to obtain methyldecalins in a reasonably pure form. It was observed that, with $\text{NiMo/TiO}_2\text{-Al}_2\text{O}_3$ as catalyst, and reaction time of 10 hours, 1-methylnaphthalene could be reacted almost completely to methyldecalins, achieving yields as high as

99.5% by weight of the 1-methylnaphthalene. These experiments are reported as the first six entries in Table 3.

Table 4 presents data for conversion of methyldecalins, using proprietary Mobil catalysts (identified in the table as Pt/REX and Pd/REX), with reaction time of one hour and temperatures ranging from 300 °C to 450 °C. It is noted that, at a temperature of 300 °C with the platinum catalyst (Run 59), conversion of the methyldecalins to lighter liquid products is 66.9%, with substituted cyclopentanes and cyclohexanes representing 48.4% of those liquid products by weight, and other cycloalkanes an additional 32.9%, some of which may be good gasoline components. Of particular interest also is the lack of aromatics in the products, making possible ease of separation of the product. With the palladium catalyst (Run 74) the yield of substituted cyclopentanes and cyclohexanes is 41.1% of the lighter liquid product, with an additional 40.8% other cycloalkanes, and again with no aromatics in the product.

It is noted that conversion of 1-methylnaphthalene to decalins increased as reaction time increased, reaching a value of nearly 100% in 10 hours. An experiment is to be conducted, increasing the time to times longer than one hour, which is expected to increase the conversions, leading to higher yields of substituted cyclopentanes and cyclohexanes.

It is noted that with both catalysts, as temperature increases, an increasing amount of the decalin feed is dehydrogenated to form aromatics, the aromatics becoming the dominant component in the product at a temperature of 375 °C.

Experiments reported in Progress Report No. 10 reveal that, at temperatures below 300 °C, both the hydrogenation and the cracking are too slow to be of practical value for this work. It appears that the optimum temperature for production of cycloalkanes, whether in a single stage or a two-stage process, would be in the range of 300 to 325 °C. At temperatures above these values, either the hydrogenation step is insufficient, or dehydrogenation of the hydrogenated intermediate to yield aromatics is too extensive, considering that a product essentially free of aromatics is desired.

Experiments are continuing to attempt to optimize the reaction parameters for the two-stage approach. Although we have not abandoned the idea of a single-stage process, it now appears that a two-stage process will be the more practical.

Table 2: Product distribution for the hydrocracking of 1-methylnaphthalene by using several catalysts.

			Product Distribution, wt. %								
			Cyclanes		Cyclenes	Alkanes		Alkenes	Alkynes	Aromatics	
			Methyl-decalins	Others		Branched	Normal			Methyl-tetralins	Others
Run	Catalyst	T, °C									
7	PtMo/SiO ₂ Al ₂ O ₃ (sulfided)	450	0.13	--	--	--	--	--	--	16.85	83.02
19	NiMo/TiO ₂ Al ₂ O ₃ (sulfided)	450	1.20	0.94	--	0.47	0.19	--	--	12.56	84.65
17	NiMo/SiO ₂ Al ₂ O ₃ (sulfided)	400	17.77	0.52	--	--	--	--	--	59.52	22.19
5	PtMo/SiO ₂ Al ₂ O ₃ (sulfided)	400	8.99	0.16	--	--	--	--	--	71.03	19.82
6	RhMo/SiO ₂ Al ₂ O ₃ (sulfided)	400	15.76	0.60	--	--	--	--	--	64.25	19.39
8	NiMo/TiO ₂ Al ₂ O ₃ (sulfided)	400	29.47	0.82	--	--	--	--	--	54.55	15.15
14	C-344 ⁽¹⁾	400	--	--	--	--	--	--	--	52.40	47.59
15	C-344:SiO ₂ Al ₂ O ₃ (4:1 w:w)	400	--	--	--	--	--	--	--	48.53	51.56
16	Pt(0.5%)/Al ₂ O ₃ ⁽²⁾	400	16.57	--	--	--	--	--	--	67.50	15.92
18	NiW/SiO ₂ Al ₂ O ₃ ⁽²⁾ (sulfided)	400	5.35	4.37	0.37	0.31	0.10	--	--	45.86	43.63
21	HT-400 ⁽³⁾ (sulfided)	400	20.72	--	--	--	--	--	--	62.31	16.97
20	NiMo/TiO ₂ Al ₂ O ₃ (sulfided)	350	33.32	--	--	--	--	--	--	64.39	2.28

Reaction Conditions : Feed = 1-methylnaphthalene ; Initial H₂ pressure = 1000 psig ; Reaction time = 3 hours. All MMo catalysts were prepared by incipient wetness impregnation and contain 25 mmol of M and 77 mmol of Mo/100 g SiO₂Al₂O₃ or 77 mmol of Mo/100 g TiO₂Al₂O₃.

(1) C-344 : CoMo/Al₂O₃ from Criterion Catalysts

(2) from Strem Chemicals

(3) HT-400 : CoMo/Al₂O₃ from Harshaw Catalysts

Table 3: Product distribution for the hydrogenation/hydrocracking of 1-methylnaphthalene by using several catalysts.

Run	Catalyst	P ₁₁ ⁽²⁾ , psig	t, h	Catalyst/ Feed, w/w	Total Conver., wt %	Liquid, g ⁽⁸⁾	Gas ⁽⁹⁾ g ⁽⁸⁾	Product Distribution, wt. % ⁽¹⁾							
								Methyl- decalins	Substituted Cyclohexanes and Cyclopentanes	Others Cyclo- alkanes	Branched Alkanes	Normal Alkanes	Unsatd. Comps.	Methyl- tetralin s	Aromatics
22	NiMo/TiO ₂ Al ₂ O ₃	1000	10	10/1	100.0	36.5	3.5	95.1	--	--	--	--	--	4.9	--
24	NiMo/TiO ₂ Al ₂ O ₃	1000	10	10/1	100.0	36.5	3.5	97.2	2.0	--	--	--	--	0.8	--
39	NiMo/TiO ₂ Al ₂ O ₃	1500	10	10/1	100.0	38.4	3.7	98.6	0.5	--	--	--	--	0.9	--
54	NiMo/TiO ₂ Al ₂ O ₃	1000	10	5/1	100.0	36.2	3.8	99.5	0.5	--	--	--	--	--	--
56	NiMo/TiO ₂ Al ₂ O ₃	1520	8	5/1	100.0	36.0	4.0	92.5	0.7	--	--	--	--	6.8	--
58	NiMo/TiO ₂ Al ₂ O ₃	1500	8	4/1	100.0	34.8	5.2	96.1	0.9	--	--	--	--	3.0	--
27	Pt(0.5%)/Al ₂ O ₃ ⁽⁶⁾	800	5	10/1	17.5	--	--	--	3.6	--	--	--	--	89.9	6.5
23a	Ni-4301 ⁽³⁾	1000	5	10/1	98.7	--	--	22.8	3.4	1.5	--	0.2	1.5	40.0	30.6
82	Ni-4301 ⁽³⁾	1000	10	10/1	99.9	38.5	1.5	15.3	37.0	18.9	1.6	0.2	12.2	1.2	13.6
83 ⁽⁴⁾	Ni-4301 ⁽³⁾	1000	7	10/1	100.0	38.3	1.7	13.5	30.0	12.4	3.8	0.9	11.7	4.2	23.5
84 ⁽⁵⁾	Ni-4301 ⁽³⁾	1000	5	10/1	100.0	38.4	1.6	6.4	14.5	6.6	2.2	0.5	10.0	13.5	46.3
85	Pt/REX ⁽⁷⁾	1000	10	10/1	100.0	37.6	1.4	1.4	1.2	1.3	0.5	--	3.9	36.9	54.8
86	Pd/REX ⁽⁷⁾	1000	10	10/1	100.0	37.9	1.9	2.0	3.7	2.3	1.1	0.3	5.6	25.6	59.4

Reactions were performed at 325°C. All MMo catalysts were prepared by incipient wetness impregnation and contain 25 mmol of M and 77 mmol of Mo/100 g SiO₂Al₂O₃ or 77 mmol of Mo/100 g TiO₂Al₂O₃. Catalysts except Pt/REX and Pd/REX were sulfided prior to use. (1) Calculated on converted material. (2) Initial pressure. (3) Ni-4301 : NiW/SiO₂Al₂O₃ from Harshaw Catalysts. (4) Reaction was performed at 350°C. (5) Reaction was performed at 375°C. (6) from Strem Chemicals. (7) from Mobil. (8) for 40 g of feed. (9) from difference for 40 g of feed

Table 4 : Liquid product distribution for the hydrocracking of methyldecalins by using several catalysts.

Run	Catalyst	P _{in} (initial) psig	T, °C	t, h	Total Conv., wt%	Liquid, g	Gas, g	Product Distribution, wt. %						
								Substituted Cyclohexanes and Cyclopentanes	Other Cyclo- alkanes	Alkanes		Olefins and Cycloolefins	Aromatics	
										Branched	Normal		Methyl- tetralins	Others
25	NiMo/TiO ₂ Al ₂ O ₃	1000	380	3	8.7	--	--	37.6	--	--	--	--	14.4	48.0
40	HPC-G416 ⁽¹⁾	1200	400	3	16.5	--	--	24.2	2.0	--	--	--	20.4	53.4
41	NiMo/SiO ₂ Al ₂ O ₃ ⁽²⁾	600	450	0.8	41.7	--	--	13.5	1.5	0.4	0.3	--	24.4	59.9
42	NiMo/SiO ₂ Al ₂ O ₃ ⁽²⁾	600	450	3	59.8	--	--	13.6	3.0	1.6	0.6	0.7	32.0	48.5
45	NiMo/TiO ₂ Al ₂ O ₃	600	450	3	50.6	34.1	5.9	14.1	3.8	1.2	0.8	--	29.3	50.8
44	ZrO ₂ /SO ₄ ⁻²	600	450	3	45.7	32.6	7.4	26.4	7.3	4.9	0.7	--	2.4	58.3
46	NiW/SiO ₂ Al ₂ O ₃ ⁽³⁾	600	450	3	98.0	36.4	3.6	18.4	0.7	6.6	1.9	2.0	2.8	67.6
47	NiW/SiO ₂ Al ₂ O ₃ ⁽³⁾ :SiO ₂ Al ₂ O ₃ (3:1 w.w)	600	450	3	97.7	35.9	4.1	18.7	0.7	2.7	0.8	2.3	2.2	72.6
48	Ni-4301 ⁽⁴⁾	600	450	3	99.0	30.1	9.7	22.3	0.4	10.2	2.6	1.9	1.3	61.3
59	Pt/REX ⁽⁵⁾	700	300	1	66.9	35.5	4.9	48.4	32.9	2.8	0.2	15.6	--	--
60	Pt/REX ⁽⁵⁾	700	350	1	93.7	34.2	5.7	53.2	4.4	8.2	2.1	10.6	1.9	19.6
62	Pt/REX ⁽⁵⁾	700	400	1	99.8	32.7	7.1	38.9	0.6	7.8	1.6	3.9	1.2	46.0
79	Pt/REX ⁽⁵⁾	700	450	1	100.0	25.3	14.4	23.8	0.2	10.6	3.2	2.5	0.8	58.9
74	Pd/REX ⁽⁵⁾	700	300	1	40.0	35.6	4.1	41.1	40.8	1.8	0.3	16.0	--	--
75	Pd/REX ⁽⁵⁾	700	350	1	94.9	33.7	6.2	55.0	2.9	6.2	1.5	8.8	1.2	24.4
76	Pd/REX ⁽⁵⁾	700	400	1	99.7	30.9	8.8	38.4	0.4	8.2	1.7	3.3	1.0	47.0
80	Pd/REX ⁽⁵⁾	700	450	1	99.7	25.7	14.1	20.0	0.2	9.5	3.0	3.1	0.7	57.5

All MMo catalysts were prepared by incipient wetness impregnation and contain 25 mmol of M and 77 mmol of Mo/100 g SiO₂Al₂O₃ or 77 mmol of Mo/100 g TiO₂Al₂O₃. Catalysts except ZrO₂/SO₄⁻², Pt/REX and Pd/REX were sulfided prior to use.

(1) NiMo/TiO₂Al₂O₃ from Engelhard Catalysts (2) 16.3% MoO₃, 3.6% NiO ; from Engelhard (3) from Strem Chemicals (4) from Harshaw Catalysts (5) from Mobil