

DESIGN, DEVELOPMENT, AND EVALUATION OF HYDROCARBON BASED FUELS FOR FUEL CELL ON-BOARD REFORMERS

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Abstract. The natural system for reforming naphtha is the well-known design of autothermal or partial oxidation (POX) reformers used in large stationary installations to generate refinery hydrogen supply. A pilot plant evaluation using a proprietary fuel reformer developed by Nuvera to compare the effectiveness of this system in terms of hydrogen yield and purity from candidate fuels including naphtha, methanol, and gasoline has been conducted. Fuels were prepared in Saudi Arabia by distillation from crude, hydrotreating, hydrocracking, and reforming in commercial refining units, and hydrogenating in a pilot plant. This presentation will outline the results from a comparative study of these and other candidate fuels for hydrogen generation using autothermal-partial oxidation reforming and water-shift reaction technologies. We will compare the fuel processor results from the standpoint of operations, hydrogen purity, and economics.

INTRODUCTION

Growing environmental demands and the need for higher efficiency vehicles (due to concerns regarding global warming) has triggered immense interest in fuel cells as propulsion systems for transportation vehicles. This has prompted an extensive search for new methods of delivering hydrogen fuel to these cells. One of the most likely options for delivering hydrogen fuel is to provide on-board reforming of liquid hydrocarbon fuels coupled with hydrogen separation. This option provides a safe fuel compatible with existing service station, supply, distribution, and storage infrastructure, as well as the most economic system of fueling. Low-volatility liquid hydrocarbon fuels may be stored in existing vehicular and service station tanks, pumped with existing equipment, and transported through existing pipelines and truck, marine, and rail tankers.

Gasoline has evolved over many decades to satisfy the requirements of the internal combustion engine. As the proposition of fuel cell vehicles becomes more tangible, fuel companies must consider how fuel structure specifications will change. For example, "octane" is a useless performance index for a fuel cell power platform, and in fact, typical octane enhancers actually complicate the process. To develop insight into this issue, Saudi Aramco and Nuvera (a new company borne out of the merger of Epyx Corporation and the De Nora Fuel Cell Group) have executed a collaborative research program to quantify hydrogen production efficiency for a range of candidate fuels.

Several candidate liquid fuels are already

available including gasoline, methanol, butane, ethanol, propanol, diesel, and naphtha. Naphtha is the most reasonable choice, based on its low volatility, high hydrogen content, ready availability, and low cost (1). Gasoline contains only two thirds of the hydrogen that naphtha contains, and costs almost 40 % more. Alcohols are also hydrogen deficient and expensive. In addition, most alcohols are hygroscopic, toxic and corrosive (2). Butane is too light for easy handling and containment, and diesel is not an ideal fuel for reformers because of its high sulfur content. Storage of compressed hydrogen poses a serious combustion and explosion hazard, which cannot be overlooked.

The natural system for reforming naphtha is the well-known design of autothermal or partial oxidation (POX) reformers used in large stationary installations to generate refinery hydrogen supply. A pilot plant evaluation using a proprietary fuel reformer to compare the effectiveness of this system in terms of hydrogen yield and purity from candidate fuels including naphtha, alcohols, and gasoline has been conducted. Fuels were prepared by distillation from crude, hydrotreating, hydrocracking, and reforming in commercial refining units, and hydrogenating in a pilot plant. This presentation will outline the results from a comparative study of these and other candidate fuels for hydrogen generation using autothermal-catalytic partial oxidation reforming and water-shift reaction technologies (3). We will compare the fuel processor results from the standpoint of operations, hydrogen purity, and economics.

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BACKGROUND

Naphtha comprises as much as 45 % of the liquid volume of a typical barrel of crude oil. As virgin naphtha it contains as much as 0.07 % sulfur, and 15-25 % aromatics (4). It's value averages \$0.40 per gallon, while gasoline is valued at \$0.56 per gallon and methanol is around \$1.00 (5). Because of its low energy density, methanol contains just half the energy contained in gasoline. Unleaded

gasoline, or reformat, contains only about 10 % hydrogen, while naphtha contains 16 % and methanol contains about 12.5 %. The hydrogen produced from reforming these three liquids completely to carbon dioxide and hydrogen depends on the carbon content, as well as the air/fuel and steam/fuel ratios, the hydrogen content and the density (6). Table 1 gives the values for these parameters and the net theoretical hydrogen yield for these fuels.

Table 1. Hydrogen Moles Yield per Liter of Fuel

Fuel	% Hydrogen	% Carbon	Yield Moles H ₂ / Gram Fuel	Density	Yield Grams H ₂ / Liter of Fuel
Methanol	12.58	37.48	0.125	0.7866	98.4
Reformat	10.00	90.00	0.130	0.8028	104.4
Naphtha	16.00	84.00	0.150	0.7370	110.6

FUEL PREPARATION

We have selected a total of eight fuels for this study. Methanol was obtained as ACS reagent grade laboratory chemicals. California Phase II Reformulated Gasoline, containing 30 ppm sulfur, was acquired and tested by Nuvera. The next three fuels, light hydrotreated naphtha (LHTN), heavy hydrotreated naphtha (HHTN), and reformat, were obtained from a commercial, 350,000 BPD Arabian Gulf integrated refinery. The next two fuels are hydrogenated light hydro-treated naphtha (HD-LHTN) and hydrogenated heavy hydro-treated naphtha (HD-HHTN). These two fuels were prepared from the indicated refinery streams by hydrogenation in a one-gallon per day bench scale Zeton-Altamira AMI 2000 Pilot Plant. Hydrocracked naphtha, produced from a blend of vacuum gas oil (VGO)-deasphalted oil (DAO) feed in a commercial, Central Arabian Refinery Hydrocracker was also tested. The final fuel used in this comparison was California Phase II Reformulated Gasoline (30 ppm sulfur).

The candidate fuels produced in Saudi Arabia were analyzed and cross analyzed for typical properties using standard ASTM methods in various Saudi Aramco Laboratories. The properties which were analyzed include density, API gravity, sulfur, nitrogen, carbon, hydrogen, molecular weight, research octane number (RON), thermal heating value, PIONA analysis (paraffins, isoparaffins, olefins, naphthenes and aromatics), and boiling point distribution. These properties are listed in Table 2.

Methanol

Methanol is a volatile, flammable, toxic, hygroscopic, water miscible liquid produced by a number of processes, primarily from

natural gas. It has a boiling point of 64.7 °C, a density of 0.787, a flash point of 12°C, and autoignition temperature of 470 °C. The flammable limits are 6.0-36.5 volume % in air.

Methanol is used mainly as a chemical feedstock for the production of methyl tertiary butyl ether (MTBE). Current production is near 600,000 BPD, which represents only 1 % of today's fuel supply needs. Methanol may be steam reformed via endothermic reaction at temperatures as low as 250 °C. It has a high heat of vaporization, and steam reformers typically require a long start-up.

Hydro-treated Naphthas

The integrated refinery uses a 55,000 bpd naphtha hydrotreater (NHT) to process whole naphtha blend provided from an Arabian Light Crude Distillation Unit, a hydrocracker, a visbreaker, and light straight run naphtha (LSRN) from a naphtha splitter. Whole naphtha feed is mixed with hydrogen prior to entering a 545 °F pre-heater exchanger then a charge heater, which brings it to required reaction temperature of 600-650 °F.

The NHT is a two-stage system with each reactor charged with 1647 cubic feet of cobalt/molybdenum on alumina extrudate NHT catalyst (7). In the reactors, desulfurization, denitrification, and olefin saturation takes place. Little change in aromatics content is observed. A cold hydrogen quench is operated between stages to control exotherms. Product is cooled in the feed exchangers, and washed by water injection, then enters the product separator where gas, liquid hydrocarbons, and sour water are separated. The liquid hydrocarbon stream enters the Naphtha Stripper Column (NSC), where hydrogen sulfide is withdrawn to amine treatment, LNG are condensed in the receiver then pumped to LNG treatment, and stripper

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bottoms, consisting of hydro-treated whole naphtha is sent to the Naphtha Splitter Column. In this column, the feed is split into light hydrotreated naphtha (LHTN) and heavy hydrotreated naphtha (HHTN). The split is maintained with a 200 °F overhead, 350 °F bottoms, and a 70 % reflux rate. Light naphtha (105-180°F) is cooled in the condenser and sent to LHTN storage. Heavy product (180-360°F) is cooled with the stripper feed exchanger then sent to the Platformer feed tank.

Reformate

Reformate is prepared in the Continuous Catalyst Regeneration Platformer (CCR) unit which converts the HHTN to 101 RON reformate with maximum Reid Vapor Pressure of 3.5 psi, as a blending stock for the gasoline pool (7). This unit, which consists of the Platformer and Continuous Catalyst Regeneration units, can process up to 40,000 BPD of HHTN, with a Liquid Volume Yield of 80 % reformate as well as hydrogen and LPG products. The CCR continuously regenerates Platformer catalyst. Air is used to burn off coke deposited on the catalyst surface, and the catalyst is circulated continuously in a closed loop between the reactors and the regenerator. The catalyst activity and product properties and yields are maintained at a constant level throughout the catalyst life. The catalyst has a life-span of about 3 years (8). Hydrogen chloride generated from the combustion gas is scrubbed in the Exhaust Gas Treater (EGT). Hydrogen chloride in the hydrogen recycle gas is removed by the Net Gas Treater (NGT).

The feed is treated during startup and at other times in the Chemical Injector pretreater section with chloride and sulfiding agents as required. Feed HHTN is then mixed with hydrogen-rich recycle gas in the Combined Feed Exchanger where it is heated to 980 °F by exchanging with the reactor effluent, then raised to 1020 °F in the Charge Heater before entering the first reactor stage. The Platformer consists of a four-reactor stack topped by the Top Surge Catalyst Hopper. Each reactor is of radial flow design to minimize pressure drop. The reactors are charged with platinum on alumina spheres with volumes of 800, 1000, 1200, and 1800 cubic feet respectively.

Feed enters the top of the first reactor, is distributed to the outside of the bed then flows to the center, up through the center pipe, then out the top of the reactor. Catalyst is withdrawn from the reactor boot and sent to regeneration. Inter-heaters between reactors make up heat lost due to the endothermic reforming reactions. Product is cooled in the

Combined Feed Exchangers then enters the Product Condensers. The product then enters the flash drum where hydrogen is evolved to suction of the recycle compressor and the Net Gas Treater. The product stream next enters the Recovery system which maximizes recovery of propane and higher gasses, then goes to the debutanizer which separates LNG from pentanes plus product (reformate) (9).

Hydrogenated Naphthas

Hydrotreated light and heavy naphthas are excellent fuel candidates for fuel processors, with high hydrogen and low sulfur and nitrogen content. They still contain aromatics, however, and can be improved as feeds to fuel processors by further treatment (see Table 3). We have recently agreed with Akzo Nobel to evaluate their new Ketjenfine (KF-200) Aromatics Saturation Catalyst (10). This is a promoted noble metal catalyst developed by Akzo Nobel and Fina which was designed for use in deep aromatics saturation of naphtha, kerosene, and low sulfur diesel. The evaluation was performed on LHTN and HHTN in a 135 ml/hour Zeton Altamira AMI-2000 Pilot Plant. The AMI 2000 pilot plant was designed for the study of catalytic reactions utilizing both gases and liquid feeds. The system was designed to offer flexibility in its control and operation. It includes a feed delivery system, which allows precise control of up to two gases and one liquid. A condenser / separator module allows for the collection of liquid products. The gas effluent is automatically routed to a gas chromatograph for analysis. The unit was operated at a WABT of 518 °F and hydrogen pressure of 1100 psi. The reactor is 0.5 meter by 1.7 cm ID, and was charged with 100 milliliters of KF-200. The catalyst was diluted 2:1 with inert silicon carbide pellets to ensure good temperature distribution. The hydrogen-oil ratio was maintained at 700:1, with a LHSV of 1.5. The average operating conditions and material balance were calculated on a daily basis. Gas samples were analyzed by gas chromatography and the molecular weight of the gas was calculated. C_5^+ in the gas was calculated and added back to the liquid product to calculate the liquid volume yield. If the operating conditions and material balance were acceptable (between 98% and 102%), a daily composite sample was prepared. Every composite sample was analyzed for density, distillation and PIONA. The distillation results give the boiling point distribution and the PIONA analysis indicates the weight percent of paraffins, iso-paraffins, olefins, naphthenes, and aromatics and the calculated RON, C/H ratio, and average molecular weight.

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Table 3 gives a summary of operating conditions, and important properties for the two hydrotreated naphtha feeds and products. The hydrogenation experiments were effective in saturating most aromatic compounds at moderately low temperature. In the case of LHTN the conversion of aromatics to naphthenes was 100% effective as indicated in Table III. In the case of the HHTN the aromatics content dropped from 16% to 1%. Operating the pilot plant at a temperature 20 °C higher resulted in even greater conversion (<0.5% aromatics). Due to time constraints, hydrogenated, heavy, hydrotreated naphtha (HD HHTN) was not tested in the Nuvera MPR facility.

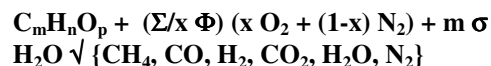
Hydrocracked Naphtha

This fuel was produced from the hydrotreating/hydrocracking of an Arabian Light vacuum gas oil (VGO)-deasphalted oil (DAO) feed in a commercial Hydrocracker. The reactor is two-stage with an amorphous cobalt-molybdenum catalyst in the first stage and a zeolite based catalyst in the second stage. This is a high pressure unit operating at 2500 psi hydrogen in the hydrotreater stage and 2800 psi in the hydrocracker stage (11). The hydrocracker operates in the indirect recycle mode with 20 % product-fractionator bottoms returned to the crude fractionator and a small purge stream withdrawn to fuel. Products from the unit include 15 % light gasses, 35 % naphtha, and 45 % distillate. The heavy naphtha fraction has a RON value of 68 and contains 5-50 ppm sulfur. Other properties are

found in Table 2. The heavy naphtha product is usually combined with straight run naphtha feed to hydrotreating and reforming units.

FUEL PROCESSING

Fuel processing is a general term which has been adopted to describe the chemical engineering steps required to extract hydrogen from primary fuel sources, for use in fuel cell power applications. The general stoichiometry may be expressed as follows:



Where : $C_m H_n O_p$ is the fuel

$\Sigma = (m+n/4-p/2) =$ stoichiometric coefficient of oxidant mixture for complete combustion

$x =$ mole fraction of oxygen in the oxidant mixture (e.g. air)

$\Phi =$ equivalence ratio

$\sigma =$ molar steam/carbon ratio

The product set will invariably contain other species, e.g. higher alkanes, olefins, aromatics, ammonia, etc., but for any truly feasible operating conditions, the indicated products will predominate. Four special classes of fuel processing may be readily identified, depending on the utilization of air and steam, as shown in Table 4.

Table 4. Special Classes of Fuel Processing

Fuel Processing Class	Oxidant/air	Water/steam
Thermal cracking	No	No
Partial oxidation	Yes	No
Steam reforming	No	Yes
Autothermal reforming	Yes	Yes

In the present study, we confine our attention to autothermal reforming (ATR), the presently preferred technology for vehicular applications. The performance objective of the reforming step is to break the fuel down into syngas: a mixture of H_2 , CO , H_2O , and CO_2 . The higher the relative concentrations of H_2 and CO , the higher the “quality” of the syngas. There are two primary quantities of interest,

$$\eta_{syn} = (H_2 + CO \text{ out}) \times LHV_{H_2} / [(Fuel \text{ in}) \times LHV_{Fuel}]$$

and implicitly assumes the complete conversion of carbon monoxide to carbon dioxide and hydrogen – this is a valid approximation because the *water gas shift*

which are related, in the present testing: Syngas production efficiency of the reforming step, and hydrogen concentration in the product gas. The syngas production efficiency (η_{syn}) is defined as the amount of hydrogen energy produced per unit of primary fuel energy consumed:

stage, a downstream unit operation, is very robust for facilitating this conversion.

For modest pressures (e.g. < 3 atm), sufficiently high temperatures (e.g. > 750 °C), and typical operating parameters, the

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equilibrium methane concentration in the product set will be low, as shown in Figure 1. For example, at 3 atm and 750 °C, the equilibrium methane content of the product gases (at $\Phi = 3.75$, $\sigma = 2.0$) is about 0.2 mol% on a dry basis. As pressure is increased, the temperature must be increased to maintain the same equilibrium methane level, in accord

with Le Chatelier's principle. For the purpose of developing *preliminary* performance expectations, one can take methane out of the product set altogether.

A stoichiometric analysis was conducted on this basis, and will be discussed subsequently in the context of the results.

Table 5. Saudi Aramco-Nuvera Autothermal Reforming Test matrix

Test ID	Fuel	Equivalence Ratio*
Bench 1	Methanol	3.00, 3.25, 3.50, 3.75
Bench 2	Cal Phase II RFG	3.00, 3.25, 3.50, 3.75
SA 1	Hydrogenated, Light HT Naphtha (HD LHTN)	3.00, 3.25, 3.50, 3.75
SA 2	Reformate (sulfur-free gasoline)	3.00, 3.25, 3.50, 3.75
SA 3	Heavy HT Naphtha (HHTN)	3.00, 3.25, 3.50, 3.75
SA 4	Light HT Naphtha (LHTN)	3.00, 3.25, 3.50, 3.75
SA 5	Hydrocracker Naphtha	3.00, 3.25, 3.50, 3.75

* Equivalence ratio = (Fuel/Air) / (Fuel/Air)_{Stoichiometric}

EXPERIMENTAL APPROACH

The fuels tested comprise a subset of the fuels discussed in the earlier portion of this paper. The only experimental parameter that was adjusted was the equivalence ratio – the test matrix is shown in Table 5.

The feed preheat was kept approximately constant at 650 °F (343 °C) and the steam/carbon ratio was fixed at 2.1. The operating pressure was 3 atm. The firing rate was 80 kW_{th} in all cases, based on the lower heating values of the fuels. The reforming catalyst used in the tests was a base metal material on a granular alumina substrate.

All testing was done at Nuvera's Modular Pressurized Reactor (MPR) facility, shown in Figure 2.

The MPR is an assemblage of thoroughly instrumented, thermally controlled pipe spools, each of which houses a specific process function. Due to its disintegrated nature, different parameters can be varied independently and their effects can be observed in isolation. In the present study, the preheat, mixing, and reforming spools were used. All gas samples were collected at the *exit* of the reforming bed.

Multiple samples were collected at every condition to ensure repeatability. Collected samples were conditioned and sent to a specially configured gas chromatograph (GC) for analysis. The GC was calibrated in a composition range typical of the reformer section exit. All operating conditions were monitored for carbon formation using a differential pressure gage across the reforming bed and an inline sample filter.

RESULTS AND DISCUSSION

All fuels were tested over a range of equivalence ratios, from 3.00 to 3.75. None of the fuels exhibited carbon formation for the conditions tested. GC analyses showed repeatability to within 5% in all cases.

Because of their impact on temperature, heat losses from the system were estimated. A number of comparisons of experimentally measured steady-state temperature fields to equilibrium predictions suggest that heat losses on the fixed thermal input of 80 kW_{th} were between 3 and 5 kW over the test range. In an actual integrated reactor system, they would be less.

At 3 atm and above 700 °C, all fuels are expected to evidence a high level of conversion of atomic carbon to CO and CO₂ on the basis of equilibrium, i.e. methane concentrations should be less than 1.0 mol%. For all but two of the 24 state points studied, the bed exit temperature was indeed in excess of 700 °C (1292 °F), as shown in Figure 3. However, as can be seen in the plot of methane slip, Figure 4, there were very clear departures from equilibrium.

The influence of temperature on the kinetics is apparent, as shown in Figure 4. As equivalence ratio is increased, compositions are changing but more importantly reaction temperature is falling, and accordingly, methane levels are rising because of *kinetics* limitations. This is the key point:

Methane levels are rising with equivalence ratio *not* because of the shift in equilibrium, but rather because of the retarded kinetics associated with reduced temperatures.

The methane slip has a dramatic effect on syngas production efficiency and hydrogen

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concentration levels, as shown in Figures 5 and 6, because each methane molecule represents three potential hydrogen molecules, two from the atomic hydrogen, and one from the conversion of carbon as CO shift. While the data suggests methanol and gasoline have the highest syngas production efficiency, this is contrary to expectation.

An equilibrium stoichiometric analysis was done based on the properties of the different fuels. The results for predicted syngas production efficiency are plotted in Figure 7.

According to Figure 7, hydrocracked naphtha and hydrogenated, light, hydrotreated, naphtha (HD LHTN) have the highest syngas production efficiency, the complete opposite of the experimental findings. Nonetheless, note that methanol is essentially at its theoretical limit, consistent with its low methane slip as shown in Figure 4.

The present testing involved variation of only two parameters: fuel and equivalence ratio. For identical preheats and stoichiometries, different fuels will achieve different peak and equilibrium temperatures. These temperatures will have a very profound effect on the rate of approach to equilibrium, as is indicated clearly by the experimental data. For the baseline experimental conditions chosen, the reforming bed was sized to conservatively achieve full conversion of methanol, which was demonstrated. However, the bed turned out to be quite undersized (space velocity too aggressive) for the fuels with lower reaction temperatures. A detailed study of fuel energetics is required to resolve these effects, and a systems analysis is required to define the preferred reformer operating conditions for each fuel individually. Neither has yet been undertaken in the present work.

Hydrocracked Naphtha, a sulfur-bearing fuel, yielded the lowest experimental efficiencies. This fuel was run last in the experimental program. For this fuel, sulfur-induced degradation of the catalyst may have contributed to elevated methane and higher hydrocarbon slippage. This could be part of the reason for the lower hydrogen production and sagging efficiency numbers for this fuel. A reference run was made at the conclusion of testing to investigate this hypothesis, and some modest evidence of deactivation was indeed found.

CONCLUSION

In the present work, we have tested four candidate naphtha based fuels (Table 2) with

two reference fuels in the Nuvera multi fuel processing facility in the interest of demonstrating syngas production via autothermal reforming. This objective was easily met, without any carbon formation and with acceptable, but not optimal, efficiencies in all cases.

Caution is necessary in interpreting experimental results from test programs wherein fuels are changed but test conditions are kept constant. Kinetics limitations associated with the choice of test state points can lead to misguided conclusions, as was described in the comparison of the experimental results to those derived from chemical reaction equilibrium. A proper comparison of fuel performance requires detailed study: an in-depth systems analysis, and parametric testing in preheat level, steam/carbon ratio, and fuel sulfur loading. We intend to follow this approach in the next stages of this study and to evaluate additional candidate fuel cell fuels that have been conditioned in the Saudi Aramco R&D Center.

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KEY PHRASES AND INDEXING

The following are keywords which apply to this paper:- Fuels, Fuel Cells, , Efficiency, Equivalence Ratio, Autothermal Reforming, Hydrogen, Naphtha, Methanol, Gasoline, Partial Oxidation, Hydro-treating, Hydro-cracking, Hydrogenation.

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Table 2. Properties of Fuels and Feedstocks for Fuel Conditioning

Fuel	Methanol	Light Naphtha		Heavy Naphtha		Refined Naphtha	
Treatment	None	Hydro-treated (LHTN)	Hydrogenated (HD LHTN)	Hydro-treated (HHTN)	Hydrogenated (HD HHTN)	Reformed (Platformate)	Hydro-cracked (HCN)
Density (g/cc)	0.791	0.659	0.658	0.737	0.714	0.807	0.727
API gravity @ 60°F		83.9	84.0	61.2	64.3	44.7	54.0
Sulfur ppm	0	<1	< 1	<1	<1	<1	6
Nitrogen ppm	0	ND	ND	ND	ND	ND	ND
% Hydrogen	12.5	16.17	16.41	14.49	15.31	11.18	13.60
% Carbon	37.5	83.83	83.69	85.51	84.69	88.82	86.40
MW	32.05	81.0	81.45	108.6	107.5	101.6	107.9
RON		61.5	61.2	54.0	48.8	101.0	72.9
Heating Value (Btu/lb)		19,500	19,700	19,900	20,000	18,600	18,800
PIONA (wt. %)							
Paraffins	0	51.9	51.9	32.7	32.1	8.26	7.10
Isoparaffins	0	39.8	40.7	32.8	35.8	18.4	34.6
Olefins	0	0.00	0.00	0.17	0.00	1.87	0.10
Naphthenes	0	6.44	7.39	18.4	29.9	1.69	32.6
Aromatics	0	1.88	0.00	15.7	1.09	68.8	24.0
Others	100.00	0.00	0.00	0.26	1.31	0.86	1.65
SIMDIS (°F)	148.2						
5 %		98	97.4	152	152	130	163.9
10 %		99	98.1	166	166	153	188.9
20 %		129	99.1	191	190	187	204
30 %		131	130.4	206	200	232	225.8
40 %		133	132	222	222	233	242.5
50 %		141	139.4	238	238	277	258.4
60 %		151	142.7	256	252	282	280.6
70 %		153	153	277	279	292	292.8
80 %		154	154.1	292	295	324	320.5
90 %		156	155.7	320	333	347	344
95 %		167	166.5	336	358	378	362.6
FBP		202	262.3	379	398	485	398.0

Table 3. Conditions and Properties for Hydrogenation Pilot Plant Study

Conditions	Light Naphtha		Heavy Naphtha	
	Hydrotreated (LHTN)	Hydrogenated (HD LHTN)	Hydrotreated (HHTN)	Hydrogenated (HD HHTN)
Catalyst	UOP S-12H	AKZO KF200	UOP S-12H	AKZO KF200
Temperature °F	600	572	600	608
Pressure	300	1100	300	1100
Hydrogen/HC	500	700	500	700
LHSV	1.5	1.5	1.5	1.5
LV% Yield	97	101	99	102
Properties				
Specific Gravity	0.659	0.658	0.737	0.714
API Gravity @ 60°F	83.9	84.0	61.2	64.3
Sulfur ppm	<1	< 1	<1	<1
Nitrogen ppm	ND	ND	ND	ND
Hydrogen wt %	16.17	16.41	14.49	15.31
Carbon wt %	83.83	83.69	85.51	84.69
MW	81.0	81.45	108.6	107.5
RON	61.5	61.2	54.0	48.8
Heating Value (Btu/lb)	19,500	19,700	19,900	20,000
PIONA (wt%)				
Paraffins	51.9	51.9	32.7	32.1
Isoparaffins	39.8	40.7	32.8	35.8
Olefins	0.00	0.00	0.17	0.00
Naphthenes	6.44	7.39	18.4	29.9
Aromatics	1.88	0.00	15.7	1.09
Unidentified	0.00	0.00	0.26	1.31
SIMDIS °F				
5 %	98	97.4	152	152
10 %	99	98.1	166	166
20 %	129	99.1	191	190
30 %	131	130.4	206	200
40 %	133	132	222	222
50 %	141	139.4	238	238
60 %	151	142.7	256	252
70 %	153	153	277	279
80 %	154	154.1	292	295
90 %	156	155.7	320	333
95 %	167	166.5	336	358
FBP	202	262.3	379	398