

PROCESS FOR CONVERTING GASEOUS ALKANES TO OLEFINS AND LIQUID HYDROCARBONS

5 REFERENCE TO RELATED PATENT APPLICATION:

This application is a continuation-in-part of copending U.S. patent application serial no. 11/101,886 filed on April 8, 2005 and entitled "Process for Converting Gaseous Alkanes to Liquid Hydrocarbons" which is a continuation-in-part of copending U.S. patent application serial no. 10/826,885
10 filed on April 16, 2004 and entitled "Process for Converting Gaseous Alkanes to Liquid Hydrocarbons".

BACKGROUND OF THE INVENTION

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FIELD OF THE INVENTION:

The present invention relates to a process for converting lower molecular weight, gaseous alkanes to olefins that are useful as monomers and intermediaries in the production of chemicals, such as lubricant and fuel
20 additives, and higher molecular weight hydrocarbons, and more particularly, to a process wherein a gas containing lower molecular weight alkanes is reacted with a dry bromine vapor to form alkyl bromides and hydrobromic acid which in turn are reacted over a crystalline alumino-silicate catalyst to form olefins and higher molecular weight hydrocarbons.

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DESCRIPTION OF RELATED ART:

Natural gas which is primarily composed of methane and other light alkanes has been discovered in large quantities throughout the world. Many of the locales in which natural gas has been discovered are far from
30 populated regions which have significant gas pipeline infrastructure or market demand for natural gas. Due to the low density of natural gas, transportation thereof in gaseous form by pipeline or as compressed gas in vessels is expensive. Accordingly, practical and economic limits exist to the distance over which natural gas may be transported in gaseous form exist. Cryogenic
35 liquefaction of natural gas (LNG) is often used to more economically transport

natural gas over large distances. However, this LNG process is expensive and there are limited regasification facilities in only a few countries that are equipped to import LNG.

Another use of methane found in natural gas is as feed to processes for the production of methanol. Methanol is made commercially via conversion of methane to synthesis gas (CO and H₂) at high temperatures (approximately 1000° C.) followed by synthesis at high pressures (approximately 100 atmospheres). There are several types of technologies for the production of synthesis gas (CO and H₂) from methane. Among these are steam-methane reforming (SMR), partial oxidation (POX), autothermal reforming (ATR), gas-heated reforming (GHR), and various combinations thereof. SMR and GHR operate at high pressures and temperatures, generally in excess of 600° C., and require expensive furnaces or reactors containing special heat and corrosion-resistant alloy tubes filled with expensive reforming catalyst. POX and ATR processes operate at high pressures and even higher temperatures, generally in excess of 1000° C. As there are no known practical metals or alloys that can operate at these temperatures, complex and costly refractory-lined reactors and high-pressure waste-heat boilers to quench & cool the synthesis gas effluent are required. Also, significant capital cost and large amounts of power are required for compression of oxygen or air to these high-pressure processes. Thus, due to the high temperatures and pressures involved, synthesis gas technology is expensive, resulting in a high cost methanol product which limits higher-value uses thereof, such as for chemical feedstocks and solvents. Furthermore production of synthesis gas is thermodynamically and chemically inefficient, producing large excesses of waste heat and unwanted carbon dioxide, which tends to lower the conversion efficiency of the overall process. Fischer-Tropsch Gas-to-Liquids (GTL) technology can also be used to convert synthesis gas to heavier liquid hydrocarbons, however investment cost for this process is even higher. In each case, the production of synthesis gas represents a large fraction of the capital costs for these methane conversion processes.

Numerous alternatives to the conventional production of synthesis gas

as a route to methanol or synthetic liquid hydrocarbons have been proposed. However, to date, none of these alternatives has attained commercial status for various reasons. Some of the previous alternative prior-art methods, such as disclosed in U.S. Patent Nos. 5,243,098 or 5,334,777 to Miller, teach
5 reacting a lower alkane, such as methane, with a metallic halide to form a metalous halide and hydrohalic acid which are in turn reduced with magnesium oxide to form the corresponding alkanol. However, halogenation of methane using chlorine as the preferred halogen results in poor selectivity to the monomethyl halide (CH_3Cl), resulting in unwanted by-products such as
10 CH_2Cl_2 and CHCl_3 which are difficult to convert or require severe limitation of conversion per pass and hence very high recycle rates..

Other prior art processes propose the catalytic chlorination or bromination of methane as an alternative to generation of synthesis gas (CO and H_2). To improve the selectivity of a methane halogenation step in an
15 overall process for the production of methanol, U.S. Patent No. 5,998,679 to Miller teaches the use of bromine, generated by thermal decomposition of a metal bromide, to brominate alkanes in the presence of excess alkanes, which results in improved selectivity to mono-halogenated intermediates such as methyl bromide. To avoid the drawbacks of utilizing fluidized beds of
20 moving solids, the process utilizes a circulating liquid mixture of metal chloride hydrates and metal bromides. Processes described in U.S. Patent Nos. 6,462,243 B1, US 6,472,572 B1, and US 6,525,230 to Grosso are also capable of attaining higher selectivity to mono-halogenated intermediates by the use of bromination. The resulting alkyl bromides intermediates such as
25 methyl bromide, are further converted to the corresponding alcohols and ethers, by reaction with metal oxides in circulating beds of moving solids. Another embodiment of U.S. Patent No. 6,525,230 avoids the drawbacks of moving beds by utilizing a zoned reactor vessel containing a fixed bed of metal oxide/metal bromide that is operated cyclically in four steps. These
30 processes also tend to produce substantial quantities of dimethylether (DME) along with any alcohol. While DME is a promising potential diesel engine fuel substitute, as of yet, there currently exists no substantial market for DME, and hence an expensive additional catalytic process conversion step would be

required to convert DME into a currently marketable product. Other processes have been proposed which circumvent the need for production of synthesis gas, such as U.S. Patent Nos. 4,655,893 and 4,467,130 to Olah in which methane is catalytically condensed into gasoline-range hydrocarbons via catalytic condensation using superacid catalysts. However, none of these
5 earlier alternative approaches have resulted in commercial processes.

It is known that substituted alkanes, in particular methanol, can be converted to olefins and gasoline boiling-range hydrocarbons over various forms of crystalline alumino-silicates also known as zeolites. In the Methanol
10 to Gasoline (MTG) process, a shape selective zeolite catalyst, ZSM-5, is used to convert methanol to gasoline. Coal or methane gas can thus be converted to methanol using conventional technology and subsequently converted to gasoline. However due to the high cost of methanol production, and at current or projected prices for gasoline, the MTG process is not considered
15 economically viable. Thus, a need exists for an economic process for the for the conversion of methane and other alkanes found in natural gas to olefins and higher molecular weight hydrocarbons which, due to their higher density and value, are more economically transported thereby significantly aiding development of remote natural gas reserves. A further need exists for a
20 process for converting alkanes present in natural gas which is relatively inexpensive, safe and simple in operation.

SUMMARY OF THE INVENTION

25 To achieve the foregoing and other objects, and in accordance with the purposes of the present invention, as embodied and broadly described herein, one characterization of the present invention is a process is provided for converting gaseous alkanes to to olefins and liquid hydrocarbons. A gaseous feed having lower molecular weight alkanes is reacted with bromine vapor to
30 form alkyl bromides and hydrobromic acid. The alkyl bromides and hydrobromic acid are then reacted in the presence of a synthetic crystalline alumino-silicate catalyst and at a temperature sufficient to form olefins and hydrobromic acid vapor.

In another characterization of the present invention, a process is provided for converting gaseous lower molecular weight alkanes to olefins which comprises reacting a gaseous feed containing lower molecular weight alkanes with bromine vapor to form alkyl bromides and hydrobromic acid. The alkyl bromides and hydrobromic acid are then reacted in the presence of a synthetic crystalline alumino-silicate catalyst to form olefins and hydrobromic acid. The process also includes converting the hydrobromic acid to bromine.

In still another characterization of the present invention, a process is provided for converting gaseous alkanes to olefins. A gaseous feed having lower molecular weight alkanes is reacted with bromine vapor to form alkyl bromides and hydrobromic acid. The alkyl bromides are reacted with hydrobromic acid in the presence of a synthetic crystalline alumino-silicate catalyst and at a temperature sufficient to form olefins and hydrobromic acid vapor. Hydrobromic acid vapor is removed from the olefins by reacting the hydrobromic acid vapor with a metal oxide to form a metal bromide and steam.

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BRIEF DESCRIPTION OF THE DRAWINGS

The accompanying drawings, which are incorporated in and form a part of the specification, illustrate the embodiments of the present invention and, together with the description, serve to explain the principles of the invention.

In the drawings:

FIG. 1 is a simplified block flow diagram of the process of the present invention;

FIG. 2 is a schematic view of one embodiment of the process of the present invention;

FIG. 3 is a schematic view of another embodiment of process of the present invention;

FIG. 4A is schematic view of another embodiment of the process of the present invention;

FIG. 4B is a schematic view of the embodiment of the process of the present invention illustrated in FIG. 4A depicting an alternative processing scheme which may be employed when oxygen is used in lieu of air in the oxidation stage;

FIG. 5A is a schematic view of the embodiment of the process of the present invention illustrated in FIG. 4A with the flow through the metal oxide beds being reversed;

FIG. 5B is a schematic view of the embodiment of the process of the present invention illustrated in FIG. 5A depicting an alternative processing scheme which may be employed when oxygen is used in lieu of air in the oxidation stage;

FIG. 6A is a schematic view of another embodiment of the process of the present invention;

FIG. 6B is a schematic view of the embodiment of the process of the present invention illustrated in FIG. 6A depicting an alternative processing scheme which may be employed when oxygen is used in lieu of air in the oxidation stage;

FIG. 7 is a schematic view of another embodiment of the process of the present invention;

FIG. 8 is a schematic view of the embodiment of the process of the present invention illustrated in FIG. 7 with the flow through the metal oxide beds being reversed; and

FIG. 9 is a schematic view of another embodiment of the process of the present invention.

DETAILED DESCRIPTION OF THE PREFERRED EMBODIMENTS

As utilized throughout this description, the term "lower molecular weight alkanes" refers to methane, ethane, propane, butane, pentane or mixtures thereof. As also utilized throughout this description, "alkyl bromides" refers to mono, di, and tri brominated alkanes. Also, the feed gas in lines 11 and 111

in the embodiments of the process of the present invention as illustrated in FIGS. 2 and 3, respectively, is preferably natural gas which may be treated to remove sulfur compounds and carbon dioxide. In any event, it is important to note that small amounts of carbon dioxide, e.g. less than about 2 mol %, can be tolerated in the feed gas to the process of the present invention.

A block flow diagram generally depicting the process of the present invention is illustrated in FIG. 1, while specific embodiments of the process of the present invention are illustrated in FIGS. 2 and 3. Referring to FIG. 2, a gas stream containing lower molecular weight alkanes, comprised of a mixture of a feed gas plus a recycled gas stream at a pressure in the range of about 1 bar to about 30 bar, is transported or conveyed via line, pipe or conduit 62, mixed with dry bromine liquid transported via line 25 and pump 24, and passed to heat exchanger 26 wherein the liquid bromine is vaporized. The mixture of lower molecular weight alkanes and dry bromine vapor is fed to reactor 30. Preferably, the molar ratio of lower molecular weight alkanes to dry bromine vapor in the mixture introduced into reactor 30 is in excess of 2.5:1. Reactor 30 has an inlet pre-heater zone 28 which heats the mixture to a reaction initiation temperature in the range of about 250° C. to about 400° C.

In first reactor 30, the lower molecular weight alkanes are reacted exothermically with dry bromine vapor at a relatively low temperature in the range of about 250° C. to about 600°C., and at a pressure in the range of about 1 bar to about 30 bar to produce gaseous alkyl bromides and hydrobromic acid vapors. The upper limit of the operating temperature range is greater than the upper limit of the reaction initiation temperature range to which the feed mixture is heated due to the exothermic nature of the bromination reaction. In the case of methane, the formation of methyl bromide occurs in accordance with the following general reaction:



This reaction occurs with a significantly high degree of selectivity to methyl bromide. For example, in the case of bromination of methane, a

methane to bromine ratio of about 4.5:1 increases the selectivity to the mono-halogenated methyl bromide. Small amounts of dibromomethane and tribromomethane are also formed in the bromination reaction. Higher alkanes, such as ethane, propane and butane, are also readily brominated
5 resulting in mono and multiple brominated species such as ethyl bromides, propyl bromides and butyl bromides. If an alkane to bromine ratio of significantly less than about 2.5 to 1 is utilized, a lower selectivity to methyl bromide occurs and significant formation of undesirable carbon soot is observed. Further, the dry bromine vapor that is feed into first reactor 30 is
10 substantially water-free. Applicant has discovered that elimination of substantially all water vapor from the bromination step in first reactor 30 substantially eliminates the formation of unwanted carbon dioxide thereby increasing the selectivity of alkane bromination to alkyl bromides and eliminating the large amount of waste heat generated in the formation of
15 carbon dioxide from alkanes.

The effluent that contains alkyl bromides and hydrobromic acid is withdrawn from the first reactor via line 31 and is partially cooled to a temperature in the range of about 150° C. to about 450° C. in heat exchanger 32 before flowing to a second reactor 34. In second reactor 34, the alkyl
20 bromides are reacted exothermically at a temperature range of from about 250° C. to about 500° C., and a pressure in the range of about 1 to 20 bar, over a fixed bed 33 of crystalline alumino-silicate catalyst, preferably a zeolite catalyst, and most preferably an X type or Y type zeolite catalyst. A preferred
25 zeolite is 10 X or Y type zeolite, although other zeolites with differing pore sizes and acidities, which are synthesized by varying the alumina-to-silica ratio may be used in the process of the present invention as will be evident to a skilled artisan. Although the zeolite catalyst is preferably used in a protonic form, a sodium form or a mixed protonic/sodium form, the zeolite may also be
30 modified by ion exchange with other alkali metal cations, such as Li, K or Cs, with alkali-earth metal cations, such as Mg, Ca, Sr or Ba, or with transition metal cations, such as Ni, Mn, V, W, or to the hydrogen form. These various alternative cations have an effect of shifting reaction selectivity. Other zeolite catalysts having varying pore sizes and acidities, which are synthesized by

varying the alumina-to-silica ratio may be used in the second reactor 34 as will be evident to a skilled artisan. In this reactor, the alkyl bromides are reacted to produce a mixture of olefins and various higher molecular weight hydrocarbons, and additional hydrobromic acid vapor.

5 The temperature at which the second reactor 34 is operated is an important parameter in determining the selectivity of the reaction to olefins and various higher molecular weight hydrocarbons. It is preferred to operated second reactor 34 at a temperature within the range of about 250° to 500° C. Temperatures above about 450° C. in the second reactor result in increased
10 yields of light hydrocarbons, such as undesirable methane and also deposition of coke, whereas lower temperatures increase yields of ethylene, propylene, butylene and heavier molecular weight hydrocarbon products. Also it is noted that methyl bromide appears to be more reactive over a lower temperature range relative to methyl chloride or other substituted methyl
15 compounds such as methanol. Notably, in the case of the alkyl bromide reaction over the preferred 10 X zeolite catalyst, cyclization reactions also occur such that the C7+ fractions contain substantial substituted aromatics. At increasing temperatures approaching 400° C., methyl bromide conversion increases towards 90% or greater, however selectivity towards C₅+ products
20 decreases and selectivity towards lighter products, particularly olefins increases. At temperatures exceeding 550° C., a high conversion of methyl bromide to methane and carbonaceous, coke occurs. In the preferred operating temperature range of between about 300° C and 450° C, as a byproduct of the reaction, a lesser amount of coke will build up on the catalyst
25 over time during operation, causing a decline in catalyst activity over a range of hours, up to hundreds of hours, depending on the reaction conditions and the composition of the feed gas. It is believed that higher reaction temperatures above about 400° C, associated with the formation of methane favor the thermal cracking of alkyl bromides and formation of carbon or coke
30 and hence an increase in the rate of deactivation of the catalyst. Conversely, temperatures at the lower end of the range, particularly below about 300° C. may also contribute to coking due to a reduced rate of desorption of heavier products from the catalyst. Hence, operating temperatures within the range of

about 250° C. to about 500° C., but preferably in the range of about 300° C. to about 450° C. in the second reactor 34 balance increased selectivity of the desired olefins and C₅+ products and lower rates of deactivation due to carbon formation, against higher conversion per pass, which minimizes the quantity of catalyst, recycle rates and equipment size required.

The catalyst may be periodically regenerated in situ, by isolating reactor 34 from the normal process flow, purging with an inert gas via line 70 at a pressure in a range from about 1 to about 5 bar at an elevated temperature in the range of about 400° C. to about 650° C. to remove unreacted material adsorbed on the catalyst insofar as is practical, and then subsequently oxidizing the deposited carbon to CO₂ by addition of air or inert gas-diluted oxygen to reactor 34 via line 70 at a pressure in the range of about 1 bar to about 5 bar at an elevated temperature in the range of about 400° C. to about 650° C. Carbon dioxide and residual air or inert gas is vented from reactor 34 via line 75 during the regeneration period.

The effluent which comprises olefins, the higher molecular weight hydrocarbons and hydrobromic acid is withdrawn from the second reactor 34 via line 35 and is cooled to a temperature in the range of 0 ° C. to about 100° C. in exchanger 36 and combined with vapor effluent in line 12 from hydrocarbon stripper 47, which contains feed gas and residual hydrocarbons stripped-out by contact with the feed gas in hydrocarbon stripper 47. The combined vapor mixture is passed to a scrubber 38 and contacted with a concentrated aqueous partially-oxidized metal bromide salt solution containing metal hydroxide and/or metal oxide and/or metal oxy-bromide species, which is transported to scrubber 38 via line 41. The preferred metal of the bromide salt is Fe(III), Cu(II) or Zn(II), or mixtures thereof, as these are less expensive and readily oxidize at lower temperatures in the range of about 120° C. to about 180° C., allowing the use of glass-lined or fluoropolymer-lined equipment; although Co(II), Ni(II), Mn(II), V(II), Cr(II) or other transition-metals which form oxidizable bromide salts may be used in the process of the present invention. Alternatively, alkaline-earth metals which also form oxidizable bromide salts, such as Ca (II) or Mg(II) may be used. Any liquid hydrocarbons condensed in scrubber 38 may be skimmed and withdrawn in

line 37 and added to liquid hydrocarbons exiting the product recovery unit 52 in line 54. Hydrobromic acid is dissolved in the aqueous solution and neutralized by the metal hydroxide and/or metal oxide and/or metal oxy-bromide species to yield metal bromide salt in solution and water which is removed from the scrubber 38 via line 44.

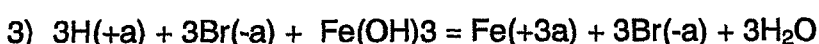
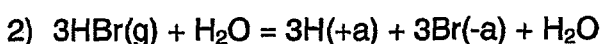
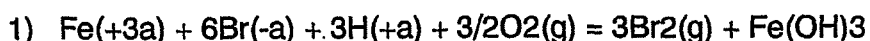
The residual vapor phase containing olefins and the higher molecular weight hydrocarbons that is removed as effluent from the scrubber 38 is forwarded via line 39 to dehydrator 50 to remove substantially all water via line 53 from the vapor stream. The water is then removed from the dehydrator 50 via line 53. The dried vapor stream containing olefins and the higher molecular weight hydrocarbons is further passed via line 51 to product recovery unit 52 to recover olefins and the C₅+ gasoline-range hydrocarbon fraction as a liquid product in line 54. Any conventional method of dehydration and liquids recovery, such as solid-bed dessicant adsorption followed by refrigerated condensation, cryogenic expansion, or circulating absorption oil or other solvent, as used to process natural gas or refinery gas streams, and to recover olefinic hydrocarbons, as will be evident to a skilled artisan, may be employed in the process of the present invention. The residual vapor effluent from product recovery unit 52 is then split into a purge stream 57 which may be utilized as fuel for the process and a recycled residual vapor which is compressed via compressor 58. The recycled residual vapor discharged from compressor 58 is split into two fractions. A first fraction that is equal to at least 2.5 times the feed gas molar volume is transported via line 62 and is combined with dry liquid bromine conveyed by pump 24, heated in exchanger 26 to vaporize the bromine and fed into first reactor 30. The second fraction is drawn off of line 62 via line 63 and is regulated by control valve 60, at a rate sufficient to dilute the alkyl bromide concentration to reactor 34 and absorb the heat of reaction such that reactor 34 is maintained at the selected operating temperature, preferably in the range of about 300° C. to about 450° C. in order to maximize conversion versus selectivity and to minimize the rate of catalyst deactivation due to the deposition of carbon. Thus, the dilution provided by the recycled vapor

effluent permits selectivity of bromination in the first reactor 30 to be controlled in addition to moderating the temperature in second reactor 34.

Water containing metal bromide salt in solution which is removed from scrubber 38 via line 44 is passed to hydrocarbon stripper 47 wherein residual dissolved hydrocarbons are stripped from the aqueous phase by contact with incoming feed gas transported via line 11. The stripped aqueous solution is transported from hydrocarbon stripper 47 via line 65 and is cooled to a temperature in the range of about 0° C. to about 70° C. in heat exchanger 46 and then passed to absorber 48 in which residual bromine is recovered from vent stream in line 67. The aqueous solution effluent from scrubber 48 is transported via line 49 to a heat exchanger 40 to be preheated to a temperature in the range of about 100° C. to about 600° C., and most preferably in the range of about 120° C. to about 180° C. and passed to third reactor 16. Oxygen or air is delivered via line 10 by blower or compressor 13 at a pressure in the range of about ambient to about 5 bar to bromine stripper 14 to strip residual bromine from water which is removed from stripper 14 in line 64 and is combined with water stream 53 from dehydrator 50 to form water effluent stream in line 56 which is removed from the process. The oxygen or air leaving bromine stripper 14 is fed via line 15 to reactor 16 which operates at a pressure in the range of about ambient to about 5 bar and at a temperature in the range of about 100° C. to about 600° C., but most preferably in the range of about 120° C. to about 180° C. so as to oxidize an aqueous metal bromide salt solution to yield elemental bromine and metal hydroxide and/or metal oxide and or metal oxy-bromide species. As stated above, although Co(II), Ni(II), Mn(II), V(II), Cr(II) or other transition-metals which form oxidizable bromide salts can be used, the preferred metal of the bromide salt is Fe(III), Cu(II), or Zn(II), or mixtures thereof, as these are less expensive and readily oxidize at lower temperatures in the range of about 120° C. to about 180° C., allowing the use of glass-lined or fluoropolymer-lined equipment. Alternatively, alkaline-earth metals which also form oxidizable bromide salts, such as Ca (II) or Mg(II) could be used.

Hydrobromic acid reacts with the metal hydroxide and/or metal oxide and/or metal oxy-bromide species so formed to once again yield the metal

bromide salt and water. Heat exchanger 18 in reactor 16 supplies heat to vaporize water and bromine. Thus, the overall reactions result in the net oxidation of hydrobromic acid produced in first reactor 30 and second reactor 34 to elemental bromine and steam in the liquid phase catalyzed by the metal
5 bromide/metal oxide or metal hydroxide operating in a catalytic cycle. In the case of the metal bromide being Fe(III)Br₃, the reactions are believed to be:



10 The elemental bromine and water and any residual oxygen or nitrogen (if air is utilized as the oxidant) leaving as vapor from the outlet of third reactor 16 via line 19, are cooled in condenser 20 at a temperature in the range of about 0° C. to about 70° C. and a pressure in the range of about ambient to 5 bar to condense the bromine and water and passed to three-phase separator
15 22. In three-phase separator 22, since liquid water has a limited solubility for bromine, on the order of about 3% by weight, any additional bromine which is condensed forms a separate, denser liquid bromine phase. The liquid bromine phase, however, has a notably lower solubility for water, on the order of less than 0.1%. Thus a substantially dry bromine vapor can be easily
20 obtained by condensing liquid bromine and water, decanting water by simple physical separation and subsequently re-vaporizing liquid bromine.

Liquid bromine is pumped in line 25 from three-phase separator 22 via pump 24 to a pressure sufficient to mix with vapor stream 62. Thus bromine is recovered and recycled within the process. The residual oxygen or nitrogen
25 and any residual bromine vapor which is not condensed exits three-phase separator 22 and is passed via line 23 to bromine scrubber 48, wherein residual bromine is recovered by solution into and by reaction with reduced metal bromides in the aqueous metal bromide solution stream 65. Water is removed from separator 22 via line 27 and introduced into stripper 14.

30 In another embodiment of the invention, referring to FIG. 3, a gas stream containing lower molecular weight alkanes, comprised of mixture of a feed gas plus a recycled gas stream at a pressure in the range of about 1 bar

to about 30 bar, is transported or conveyed via line, pipe or conduit 162, mixed with dry bromine liquid transported via pump 124 and passed to heat exchanger 126 wherein the liquid bromine is vaporized. The mixture of lower molecular weight alkanes and dry bromine vapor is fed to reactor 130.

5 Preferably, the molar ratio of lower molecular weight alkanes to dry bromine vapor in the mixture introduced into reactor 130 is in excess of 2.5:1 . Reactor 130 has an inlet pre-heater zone 128 which heats the mixture to a reaction initiation temperature in the range of about 250° C. to about 400° C.. In first reactor 130, the lower molecular weight alkanes are reacted

10 exothermically with dry bromine vapor at a relatively low temperature in the range of about 250° C. to about 600° C., and at a pressure in the range of about 1 bar to about 30 bar to produce gaseous alkyl bromides and hydrobromic acid vapors. The upper limit of the operating temperature range is greater than the upper limit of the reaction initiation temperature range to

15 which the feed mixture is heated due to the exothermic nature of the bromination reaction. In the case of methane, the formation of methyl bromide occurs in accordance with the following general reaction:



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This reaction occurs with a significantly high degree of selectivity to methyl bromide. For example, in the case of bromine reacting with a molar excess of methane at a methane to bromine ratio of 4.5:1, a high selectivity to the mono-halogenated methyl bromide occurs. Small amounts of

25 dibromomethane and tribromomethane are also formed in the bromination reaction. Higher alkanes, such as ethane, propane and butane, are also readily brominated resulting in mono and multiple brominated species such as ethyl bromides, propyl bromides and butyl bromides. If an alkane to bromine ratio of significantly less than 2.5 to 1 is utilized, substantially lower selectivity

30 to methyl bromide substantially occurs and significant formation of undesirable carbon soot is observed. Further, the dry bromine vapor that is feed into first reactor 130 is substantially water-free. Applicant has discovered

that elimination of substantially all water vapor from the bromination step in first reactor 130 substantially eliminates the formation of unwanted carbon dioxide thereby increasing the selectivity of alkane bromination to alkyl bromides and eliminating the large amount of waste heat generated in the
5 formation of carbon dioxide from alkanes.

The effluent that contains alkyl bromides and hydrobromic acid is withdrawn from the first reactor 130 via line 131 and is partially cooled to a temperature in the range of about 150° C. to 450° C. in heat exchanger 132 before flowing to a second reactor 134. In second reactor 134, the alkyl
10 bromides are reacted exothermically at a temperature range of from about 250° C. to about 500° C., and a pressure in the range of about 1 bar to 30 bar, over a fixed bed of crystalline alumino-silicate catalyst, preferably a zeolite catalyst, and most preferably an X type or Y type zeolite catalyst. A preferred
15 zeolite is 10 X or Y type zeolite, although other zeolites with differing pore sizes and acidities, which are synthesized by varying the alumina-to-silica ratio may be used in the process of the present invention as will be evident to a skilled artisan. Although the zeolite catalyst is preferably used in a protonic form, a sodium form or a mixed protonic/sodium form, such as Li, K or Cs, with alkali-earth metal cations, such as Mg, Ca, Sr or Ba, or with transition
20 metal cations, such as Ni, Mn, V, W, or to the hydrogen form. These various alternative cations have an effect of shifting reaction selectivity. Other zeolite catalysts having varying pore sizes and acidities, which are synthesized by varying the alumina-to-silica ratio may be used in the second reactor 134 as will be evident to a skilled artisan. In this reactor, the alkyl bromides are
25 reacted to produce a mixture of olefins and higher molecular weight hydrocarbons and additional hydrobromic acid vapor.

The temperature at which the second reactor 134 is operated is an important parameter in determining the selectivity of the reaction to olefins and various higher molecular weight liquid hydrocarbons. It is preferred to
30 operate second reactor 134 at a temperature within the range of about 250° C. to 500° C. , but more preferably within the range of about 300° C. to 450° C. Temperatures above about 450° C. in the second reactor result in increased yields of light hydrocarbons, such as undesirable methane and

carbonaceous coke, whereas lower temperatures increase yields of olefins such as ethylene, propylene and butylene and heavier molecular weight hydrocarbon products. Notably, in the case of alkyl bromides reacting over the preferred 10 X zeolite catalyst, cyclization reactions occur such that the

5 C₇+ fractions produced contain substantial substituted aromatics. At increasing temperatures approaching 400° C., methyl bromide conversion increases towards 90% or greater, however selectivity towards C₅+ products decreases and selectivity towards lighter products, particularly olefins, increases. At temperatures exceeding 550° C. almost complete conversion of

10 methyl bromide to methane and coke occurs. In the preferred range of operating temperatures of about 300° C. to 450° C., as a byproduct of the reaction, a small amount of carbon will build up on the catalyst over time during operation, causing a decline in catalyst activity over a range of several hundred hours, depending on the reaction conditions and feed gas

15 composition. It is observed that higher reaction temperatures above about 400° C favor the thermal cracking of alkyl bromides with formation of carbon and hence increases the rate of deactivation of the catalyst. Conversely, operation at the lower end of the temperature range, particularly below about 300° C may also promote coking, likely to the reduced rate of desorption of

20 hydrocarbon products. Hence, operating temperatures within the range of about 250° C. to 500° C. but more preferably in the range of about 300° C. to 450° C. in the second reactor 134 balance increased selectivity towards the desired olefin and C₅+ products and lower rates of deactivation due to carbon formation, against higher conversion per pass, which minimizes the quantity

25 of catalyst, recycle rates and equipment size required.

The catalyst may be periodically regenerated in situ, by isolating reactor 134 from the normal process flow, purging with an inert gas via line 170 at a pressure in the range of about 1 bar to about 5 bar and an elevated temperature in the range of 400° C. to 650° C. to remove unreacted material

30 adsorbed on the catalyst insofar as is practical, and then subsequently oxidizing the deposited carbon to CO₂ by addition of air or inert gas-diluted oxygen via line 170 to reactor 134 at a pressure in the range of about 1 bar to about 5 bar and an elevated temperature in the range of 400° C. to 650° C.

Carbon dioxide and residual air or inert gas are vented from reactor 134 via line 175 during the regeneration period.

The effluent which comprises olefins, the higher molecular weight hydrocarbons and hydrobromic acid is withdrawn from the second reactor 134
5 via line 135, cooled to a temperature in the range of about 0° C. to about 100° C. in exchanger 136, and combined with vapor effluent in line 112 from hydrocarbon stripper 147. The mixture is then passed to a scrubber 138 and contacted with a stripped, recirculated water that is transported to scrubber 138 in line 164 by any suitable means, such as pump 143, and is cooled to a
10 temperature in the range of about 0° C. to about 50° C. in heat exchanger 155. Any liquid hydrocarbon product condensed in scrubber 138 may be skimmed and withdrawn as stream 137 and added to liquid hydrocarbon product 154. Hydrobromic acid is dissolved in scrubber 138 in the aqueous solution which is removed from the scrubber 138 via line 144, and passed to
15 hydrocarbon stripper 147 wherein residual hydrocarbons dissolved in the aqueous solution are stripped-out by contact with feed gas 111. The stripped aqueous phase effluent from hydrocarbon stripper 147 is cooled to a temperature in the range of about 0° C. to about 50° C. in heat exchanger 146 and then passed via line 165 to absorber 148 in which residual bromine is
20 recovered from vent stream 167.

The residual vapor phase containing olefins and the higher molecular weight hydrocarbon is removed as effluent from the scrubber 138 and forwarded to dehydrator 150 to remove substantially all water from the gas stream. The water is then removed from the dehydrator 150 via line 153. The
25 dried gas stream containing olefins and the higher molecular weight hydrocarbons is further passed via line 151 to product recovery unit 152 to recover olefins and the C₅+ gasoline range hydrocarbon fraction as a liquid product in line 154. Any conventional method of dehydration and liquids recovery such as solid-bed dessicant adsorption followed by, for example,
30 refrigerated condensation, cryogenic expansion, or circulating absorption oil, or other solvents as used to process natural gas or refinery gas streams and recover olefinic hydrocarbons, as known to a skilled artisan, may be employed in the implementation of this invention. The residual vapor effluent from

product recovery unit 152 is then split into a purge stream 157 that may be utilized as fuel for the process and a recycled residual vapor which is compressed via compressor 158. The recycled residual vapor discharged from compressor 158 is split into two fractions. A first fraction that is equal to
5 at least 2.5 times the feed gas volume is transported via line 162, combined with the liquid bromine conveyed in line 125 and passed to heat exchanger 126 wherein the liquid bromine is vaporized and fed into first reactor 130. The second fraction which is drawn off line 162 via line 163 and is regulated by control valve 160, at a rate sufficient to dilute the alkyl bromide concentration
10 to reactor 134 and absorb the heat of reaction such that reactor 134 is maintained at the selected operating temperature, preferably in the range of about 300° C. to about 450° C. in order to maximize conversion vs. selectivity and to minimize the rate of catalyst deactivation due to the deposition of carbon. Thus, the dilution provided by the recycled vapor effluent permits
15 selectivity of bromination in the first reactor 130 to be controlled in addition to moderating the temperature in second reactor 134.

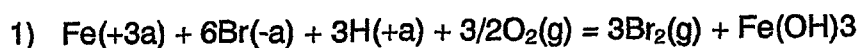
Oxygen, oxygen enriched air or air 110 is delivered via blower or compressor 113 at a pressure in the range of about ambient to about 5 bar to bromine stripper 114 to strip residual bromine from water which leaves
20 stripper 114 via line 164 and is divided into two portions. The first portion of the stripped water is recycled via line 164, cooled in heat exchanger 155 to a temperature in the range of about 20° C. to about 50° C., and maintained at a pressure sufficient to enter scrubber 138 by any suitable means, such as pump 143. The portion of water that is recycled is selected such that the
25 hydrobromic acid solution effluent removed from scrubber 138 via line 144 has a concentration in the range from about 10% to about 50% by weight hydrobromic acid, but more preferably in the range of about 30% to about 48% by weight to minimize the amount of water which must be vaporized in exchanger 141 and preheater 119 and to minimize the vapor pressure of HBr
30 over the resulting acid. A second portion of water from stripper 114 is removed from line 164 and the process via line 156.

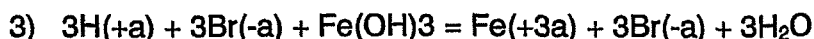
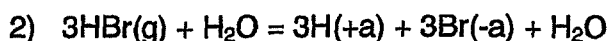
The dissolved hydrobromic acid that is contained in the aqueous solution effluent from scrubber 148 is transported via line 149 and is

combined with the oxygen, oxygen enriched air or air leaving bromine stripper 114 in line 115. The combined aqueous solution effluent and oxygen, oxygen enriched air or air is passed to a first side of heat exchanger 141 and through preheater 119 wherein the mixture is preheated to a temperature in the range of about 100° C. to about 600° C. and most preferably in the range of about 120° C. to about 180° C. and passed to third reactor 117 that contains a metal bromide salt or metal oxide. The preferred metal of the bromide salt or metal oxide is Fe(III), Cu(II) or Zn(II) although Co(II), Ni(II), Mn(II), V(II), Cr(II) or other transition-metals which form oxidizable bromide salts can be used.

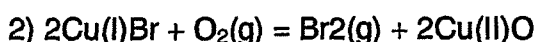
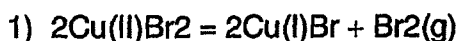
Alternatively, alkaline-earth metals which also form oxidizable bromide salts, such as Ca (II) or Mg(II) could be used. The metal bromide salt in the oxidation reactor 117 can be utilized as a concentrated aqueous solution or preferably, the concentrated aqueous salt solution may be imbibed into a porous, high surface area, acid resistant inert support such as a silica gel.

More preferably, the oxide form of the metal in a range of 10 to 20% by weight is deposited on an inert support such as alumina with a specific surface area in the range of 50 to 200 m²/g. The oxidation reactor 117 operates at a pressure in the range of about ambient to about 5 bar and at a temperature in the range of about 100° C. to 600° C., but most preferably in the range of about 120° C. to 180° C.; therein, the metal bromide is oxidized by oxygen, yielding elemental bromine and metal hydroxide, metal oxide or metal oxy-bromide species or, metal oxides in the case of the supported metal bromide salt or metal oxide operated at higher temperatures and lower pressures at which water may primarily exist as a vapor. In either case, the hydrobromic acid reacts with the metal hydroxide, metal oxy-bromide or metal oxide species and is neutralized, restoring the metal bromide salt and yielding water. Thus, the overall reaction results in the net oxidation of hydrobromic acid produced in first reactor 130 and second reactor 134 to elemental bromine and steam, catalyzed by the metal bromide/metal hydroxide or metal oxide operating in a catalytic cycle. In the case of the metal bromide being Fe(III)Br₃ in an aqueous solution and operated in a pressure and temperature range in which water may exist as a liquid the reactions are believed to be:





In the case of the metal bromide being Cu(II)Br_2 supported on an inert support and operated at higher temperature and lower pressure conditions at which water primarily exists as a vapor, the reactions are believed to be:



The elemental bromine and water and any residual oxygen or nitrogen (if air or oxygen enriched air is utilized as the oxidant) leaving as vapor from the outlet of third reactor 117, are cooled in the second side of exchanger 141 and condenser 120 to a temperature in the range of about 0° C. to about 70° C. wherein the bromine and water are condensed and passed to three-phase separator 122. In three-phase separator 122, since liquid water has a limited solubility for bromine, on the order of about 3% by weight, any additional bromine which is condensed forms a separate, denser liquid bromine phase. The liquid bromine phase, however, has a notably lower solubility for water, on the order of less than 0.1%. Thus, a substantially dry bromine vapor can be easily obtained by condensing liquid bromine and water, decanting water by simple physical separation and subsequently re-vaporizing liquid bromine. It is important to operate at conditions that result in the near complete reaction of HBr so as to avoid significant residual HBr in the condensed liquid bromine and water, as HBr increases the miscibility of bromine in the aqueous phase, and at sufficiently high concentrations, results in a single ternary liquid phase.

Liquid bromine is pumped from three-phase separator 122 via pump 124 to a pressure sufficient to mix with vapor stream 162. Thus the bromine is recovered and recycled within the process. The residual air, oxygen enriched air or oxygen and any bromine vapor which is not condensed exits three-phase separator 122 and is passed via line 123 to bromine scrubber 148, wherein residual bromine is recovered by dissolution into hydrobromic acid solution stream conveyed to scrubber 148 via line 165. Water is

removed from the three-phase separator 122 via line 129 and passed to stripper 114.

Thus, in accordance with all embodiments of the present invention set forth above, the metal bromide/metal hydroxide, metal oxy-bromide or metal
5 oxide operates in a catalytic cycle allowing bromine to be easily recycled within the process. The metal bromide is readily oxidized by oxygen, oxygen enriched air or air either in the aqueous phase or the vapor phase at temperatures in the range of about 100° C. to about 600° C. and most preferably in the range of about 120° C. to about 180° C. to yield elemental
10 bromine vapor and metal hydroxide, metal oxy-bromide or metal oxide. Operation at temperatures below about 180° C. is advantageous, thereby allowing the use of low-cost corrosion-resistant fluoropolymer-lined equipment. Hydrobromic acid is neutralized by reaction with the metal hydroxide or metal oxide yielding steam and the metal bromide.

15 The elemental bromine vapor and steam are condensed and easily separated in the liquid phase by simple physical separation, yielding substantially dry bromine. The absence of significant water allows selective bromination of alkanes, without production of CO₂ and the subsequent efficient and selective reactions of alkyl bromides to primarily C₂ to C₄ olefins
20 and heavier products, the C₅+ fraction of which contains substantial branched alkanes and substituted aromatics. Byproduct hydrobromic acid vapor from the bromination reaction and subsequent reaction in reactor 134 are readily dissolved into an aqueous phase and neutralized by the metal hydroxide or metal oxide species resulting from oxidation of the metal bromide.

25 In accordance with another embodiment of the process of the present invention illustrated in FIG. 4A, the alkyl bromination and alkyl bromide conversion stages are operated in a substantially similar manner to those corresponding stages described with respect to FIGS. 2 and 3 above. More particularly, a gas stream containing lower molecular weight alkanes,
30 comprised of mixture of a feed gas and a recycled gas stream at a pressure in the range of about 1 bar to about 30 bar, is transported or conveyed via line, pipe or conduits 262 and 211, respectively, and mixed with dry bromine liquid in line 225. The resultant mixture is transported via pump 224 and passed to

heat exchanger 226 wherein the liquid bromine is vaporized. The mixture of lower molecular weight alkanes and dry bromine vapor is fed to reactor 230. Preferably, the molar ratio of lower molecular weight alkanes to dry bromine vapor in the mixture introduced into reactor 230 is in excess of 2.5:1. Reactor
5 230 has an inlet pre-heater zone 228 which heats the mixture to a reaction initiation temperature in the range of 250° C. to 400° C. In first reactor 230, the lower molecular weight alkanes are reacted exothermically with dry bromine vapor at a relatively low temperature in the range of about 250° C. to about 600° C., and at a pressure in the range of about 1 bar to about 30 bar to
10 produce gaseous alkyl bromides and hydrobromic acid vapors. The upper limit of the operating temperature range is greater than the upper limit of the reaction initiation temperature range to which the feed mixture is heated due to the exothermic nature of the bromination reaction. In the case of methane, the formation of methyl bromide occurs in accordance with the following
15 general reaction:



This reaction occurs with a significantly high degree of selectivity to methyl bromide. For example, in the case of bromine reacting with a molar excess of
20 methane at a methane to bromine ratio of 4.5:1, a high selectivity to the mono-halogenated methyl bromide occurs. Small amounts of dibromomethane and tribromomethane are also formed in the bromination reaction. Higher alkanes, such as ethane, propane and butane, are also readily brominated resulting in mono and multiple brominated species such
25 as ethyl bromides, propyl bromides and butyl bromides. If an alkane to bromine ratio of significantly less than 2.5 to 1 is utilized, substantially lower selectivity to methyl bromide occurs and significant formation of undesirable carbon soot is observed. Further, the dry bromine vapor that is feed into first reactor 230 is substantially water-free. Applicant has discovered that
30 elimination of substantially all water vapor from the bromination step in first reactor 230 substantially eliminates the formation of unwanted carbon dioxide thereby increasing the selectivity of alkane bromination to alkyl bromides and

eliminating the large amount of waste heat generated in the formation of carbon dioxide from alkanes.

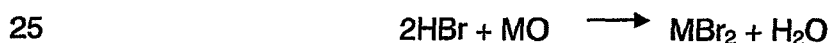
The effluent that contains alkyl bromides and hydrobromic acid is withdrawn from the first reactor 230 via line 231 and is partially cooled to a temperature in the range of about 150° C. to 450° C. in heat exchanger 232 before flowing to a second reactor 234. In second reactor 234, the alkyl bromides are reacted exothermically at a temperature range of from about 250° C. to about 500° C., and a pressure in the range of about 1 bar to 30 bar, over a fixed bed of crystalline alumino-silicate catalyst, preferably a zeolite catalyst, and most preferably, an X type or Y type zeolite catalyst. A preferred zeolite is 10 X or Y type zeolite, although other zeolites with differing pore sizes and acidities, which are synthesized by varying the alumina-to-silica ratio may be used in the process of the present invention as will be evident to a skilled artisan. Although the zeolite catalyst is preferably used in a protonic form, a sodium form or a mixed protonic/sodium form, the zeolite may also be modified by ion exchange with other alkali metal cations, such as Li, K or Cs, with alkali-earth metal cations, such as Mg, Ca, Sr or Ba, with transition metal cations, such as Ni, Mn, V, W, or to the hydrogen form. These various alternative cations have an effect of shifting reaction selectivity. Other zeolite catalysts having varying pore sizes and acidities, which are synthesized by varying the alumina-to-silica ratio may be used in the second reactor 234 as will be evident to a skilled artisan. In this reactor, the alkyl bromides are reacted to produce a mixture of higher molecular weight hydrocarbon products and additional hydrobromic acid vapor.

The temperature at which the second reactor 234 is operated is an important parameter in determining the selectivity of the reaction to olefins and various higher molecular weight liquid hydrocarbons. It is preferred to operate second reactor 234 at a temperature within the range of about 250° C. to about 500° C., but more preferably within the range of about 300° C. to about 450° C. Temperatures above about 450° C. in the second reactor result in increased yields of light hydrocarbons, such as undesirable methane and carbonaceous coke, whereas lower temperatures increase yields of olefins and heavier molecular weight hydrocarbon products. Notably, in the

case of alkyl bromides reacting over the preferred 10 X zeolite catalyst, cyclization reactions occur such that the C₇+ fractions produced contain substantial substituted aromatics. At increasing temperatures approaching 400° C., methyl bromide conversion increases towards 90% or greater, however selectivity towards C₅+ products decreases and selectivity towards lighter products, particularly olefins, increases. At temperatures exceeding 550° C. almost complete conversion of methyl bromide to methane and coke occurs. In the preferred range of operating temperatures of about 300° C. to 450° C., as a byproduct of the reaction, a small amount of carbon will build up on the catalyst over time during operation, causing a decline in catalyst activity over a range of several hundred hours, depending on the reaction conditions and feed gas composition. It is observed that higher reaction temperatures above about 400° C favor the thermal cracking of alkyl bromides with formation of carbon and hence increases the rate of deactivation of the catalyst. Conversely, operation at the lower end of the temperature range, particularly below about 300° C may also promote coking, likely to the reduced rate of desorption of hydrocarbon products. Hence, operating temperatures within the range of about 250° C. to 500° C. but more preferably in the range of about 300° C. to 450° C. in the second reactor 234 balance increased selectivity towards the desired olefin and C₅+ products and lower rates of deactivation due to carbon formation, against higher conversion per pass, which minimizes the quantity of catalyst, recycle rates and equipment size required.

The catalyst may be periodically regenerated in situ, by isolating reactor 234 from the normal process flow, purging with an inert gas via line 270 at a pressure in the range of about 1 bar to about 5 bar and an elevated temperature in the range of about 400° C. to about 650° C. to remove unreacted material adsorbed on the catalyst insofar as is practical, and then subsequently oxidizing the deposited carbon to CO₂ by addition of air or inert gas-diluted oxygen via line 270 to reactor 234 at a pressure in the range of about 1 bar to about 5 bar and an elevated temperature in the range of about 400° C. to about 650° C. Carbon dioxide and residual air or inert gas are vented from reactor 234 via line 275 during the regeneration period.

The effluent which comprises olefins, the higher molecular weight hydrocarbons and hydrobromic acid is withdrawn from the second reactor 234 via line 235 and cooled to a temperature in the range of about 100° C. to about 600° C. in exchanger 236. As illustrated in FIG. 4A, the cooled effluent is transported via lines 235 and 241 with valve 238 in the opened position and valves 239 and 243 in the closed position and introduced into a vessel or reactor 240 containing a bed 298 of a solid phase metal oxide. The metal of the metal oxide is selected from magnesium (Mg), calcium (Ca), vanadium (V), chromium (Cr), manganese (Mn), iron (Fe), cobalt (Co), nickel (Ni), copper (Cu), zinc (Zn), or tin (Sn). The metal is selected for the impact of its physical and thermodynamic properties relative to the desired temperature of operation, and also for potential environmental and health impacts and cost. Preferably, magnesium, copper and iron are employed as the metal, with magnesium being the most preferred. These metals have the property of not only forming oxides but bromide salts as well, with the reactions being reversible in a temperature range of less than about 500° C. The solid metal oxide is preferably immobilized on a suitable attrition-resistant support, for example a synthetic amorphous silica, such as Davicat Grade 57, manufactured by Davison Catalysts of Columbia, Maryland. Or more preferably, an alumina support with a specific surface area of about 50 to 200 m²/g. In reactor 240, hydrobromic acid is reacted with the metal oxide at temperatures below about 600° C. and preferably between about 100° C. to about 500° C. in accordance with the following general formula wherein M represents the metal:



The steam resulting from this reaction is transported together with olefins and the high molecular hydrocarbons in line 244, 218 and 216 via opened valve 219 to heat exchanger 220 wherein the mixture is cooled to a temperature in the range of about 0° C. to about 70° C. This cooled mixture is forwarded to dehydrator 250 to remove substantially all water from the gas stream. The water is then removed from the dehydrator 250 via line 253. The dried gas stream containing olefins and the higher molecular weight hydrocarbons is further passed via line 251 to product recovery unit 252 to recover olefins and

the C₅+ fraction as a liquid product in line 254. Any conventional method of dehydration and liquids recovery such as solid-bed dessicant adsorption followed by, for example, refrigerated condensation, cryogenic expansion, or circulating absorption oil or other solvent, as used to process natural gas or refinery gas streams and recover olefinic hydrocarbons, as known to a skilled artisan, may be employed in the implementation of this invention. The residual vapor effluent from product recovery unit 252 is then split into a purge stream 257 that may be utilized as fuel for the process and a recycled residual vapor which is compressed via compressor 258. The recycled residual vapor discharged from compressor 258 is split into two fractions. A first fraction that is equal to at least 1.5 times the feed gas volume is transported via line 262, combined with the liquid bromine and feed gas conveyed in line 225 and passed to heat exchanger 226 wherein the liquid bromine is vaporized and fed into first reactor 230 in a manner as described above. The second fraction which is drawn off line 262 via line 263 and is regulated by control valve 260, at a rate sufficient to dilute the alkyl bromide concentration to reactor 234 and absorb the heat of reaction such that reactor 234 is maintained at the selected operating temperature, preferably in the range of about 300° C. to about 450° C. in order to maximize conversion vs. selectivity and to minimize the rate of catalyst deactivation due to the deposition of carbon. Thus, the dilution provided by the recycled vapor effluent permits selectivity of bromination in the first reactor 230 to be controlled in addition to moderating the temperature in second reactor 234.

Oxygen, oxygen enriched air or air 210 is delivered via blower or compressor 213 at a pressure in the range of about ambient to about 10 bar to bromine via line 214, line 215 and valve 249 through heat exchanger 215, wherein oxygen, oxygen enriched air or air is preheated to a temperature in the range of about 100° C. to about 500° C. to a second vessel or reactor 246 containing a bed 299 of a solid phase metal bromide. Oxygen reacts with the metal bromide in accordance with the following general reaction wherein M represents the metal:



In this manner, a dry, substantially HBr free bromine vapor is produced thereby eliminating the need for subsequent separation of water or hydrobromic acid from the liquid bromine. Reactor 246 is operated below 600° C., and more preferably between about 300° C. to about 500° C. The resultant bromine vapor is transported from reactor 246 via line 247, valve 248 and line 242 to heat exchanger or condenser 221 where the bromine is condensed into a liquid. The liquid bromine is transported via line 242 to separator 222 wherein liquid bromine is removed via line 225 and transported via line 225 to heat exchanger 226 and first reactor 230 by any suitable means, such as by pump 224. The residual air or unreacted oxygen is transported from separator 222 via line 227 to a bromine scrubbing unit 223, such as venturi scrubbing system containing a suitable solvent, or suitable solid adsorbant medium, as selected by a skilled artisan, wherein the remaining bromine is captured. The captured bromine is desorbed from the scrubbing solvent or adsorbant by heating or other suitable means and the thus recovered bromine transported via line 212 to line 225. The scrubbed air or oxygen is vented via line 229. In this manner, nitrogen and any other substantially non-reactive components are removed from the system of the present invention and thereby not permitted to enter the hydrocarbon-containing portion of the process; also loss of bromine to the surrounding environment is avoided.

One advantage of removing the HBr by chemical reaction in accordance with this embodiment, rather than by simple physical solubility, is the substantially complete scavenging of the HBr to low levels at higher process temperatures. Another distinct advantage is the elimination of water from the bromine removed thereby eliminating the need for separation of bromine and water phases and for stripping of residual bromine from the water phase.

Reactors 240 and 246 may be operated in a cyclic fashion. As illustrated in FIG. 4A, valves 238 and 219 are operated in the open mode to permit hydrobromic acid to be removed from the effluent that is withdrawn from the second reactor 234, while valves 248 and 249 are operated in the open mode to permit air, oxygen enriched air or oxygen to flow through

reactor 246 to oxidize the solid metal bromide contained therein. Once significant conversion of the metal oxide and metal bromide in reactors 240 and 246, respectively, has occurred, these valves are closed. At this point, bed 299 in reactor 246 is a bed of substantially solid metal bromide; while bed 298 in reactor 240 is substantially solid metal oxide. As illustrated in FIG. 5A, valves 245 and 243 are then opened to permit oxygen, oxygen enriched air or air to flow through reactor 240 to oxidize the solid metal bromide contained therein, while valves 239 and 217 are opened to permit effluent which comprises olefins, the higher molecular weight hydrocarbons and hydrobromic acid that is withdrawn from the second reactor 234 to be introduced into reactor 246. The reactors are operated in this manner until significant conversion of the metal oxide and metal bromide in reactors 246 and 240, respectively, has occurred and then the reactors are cycled back to the flow schematic illustrated in FIG. 4A by opening and closing valves as previously discussed.

When oxygen is utilized as the oxidizing gas transported in via line 210 to the reactor being used to oxidize the solid metal bromide contained therein, the embodiment of the process of the present invention illustrated in FIGS. 4A and 5A can be modified such that the bromine vapor produced from either reactor 246 (FIG. 4B) or 240 (FIG. 5B) is transported via lines 242 and 225 directly to first reactor 230. Since oxygen is reactive and will not build up in the system, the need to condense the bromine vapor to a liquid to remove unreactive components, such as nitrogen, is obviated. Compressor 213 is not illustrated in FIGS. 4B and 5B as substantially all commercial sources of oxygen, such as a commercial air separator unit, will provide oxygen to line 210 at the required pressure. If not, a compressor 213 could be utilized to achieve such pressure as will be evident to a skilled artisan.

In the embodiment of the present invention illustrated in FIG. 6A, the beds of solid metal oxide particles and solid metal bromide particles contained in reactors 240 and 246, respectively, are fluidized and are connected in the manner described below to provide for continuous operation of the beds without the need to provide for equipment, such as valves, to change flow direction to and from each reactor. In accordance with this embodiment, the

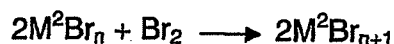
effluent which comprises olefins, the higher molecular weight hydrocarbons and hydrobromic acid is withdrawn from the second reactor 234 via line 235, cooled to a temperature in the range of about 100° C. to about 500° C. in exchanger 236, and introduced into the bottom of reactor 240 which contains
5 a bed 298 of solid metal oxide particles. The flow of this introduced fluid induces the particles in bed 298 to move upwardly within reactor 240 as the hydrobromic acid is reacted with the metal oxide in the manner as described above with respect to FIG. 4A. At or near the top of the bed 298, the particles
10 which contain substantially solid metal bromide on the attrition-resistant support due to the substantially complete reaction of the solid metal oxide with hydrobromic acid in reactor 240 are withdrawn via a weir or cyclone or other conventional means of solid/gas separation, flow by gravity down line 259 and are introduced at or near the bottom of the bed 299 of solid metal bromide particles in reactor 246. In the embodiment illustrated in FIG. 6A, oxygen,
15 oxygen enriched air or air 210 is delivered via blower or compressor 213 at a pressure in the range of about ambient to about 10 bar, transported via line 214 through heat exchanger 215, wherein the oxygen, oxygen enriched air or air is preheated to a temperature in the range of about 100° C. to about 500° C. and introduced into second vessel or reactor 246 below bed 299 of a solid
20 phase metal bromide. Oxygen reacts with the metal bromide in the manner described above with respect to FIG. 4A to produce a dry, substantially HBr free bromine vapor. The flow of this introduced gas induces the particles in bed 299 to flow upwardly within reactor 246 as oxygen is reacted with the metal bromide. At or near the top of the bed 298, the particles which contain
25 substantially solid metal oxide on the attrition-resistant support due to the substantially complete reaction of the solid metal bromide with oxygen in reactor 246 are withdrawn via a weir or cyclone or other conventional means of solid/gas separation, flow by gravity down line 264 and are introduced at or near the bottom of the bed 298 of solid metal oxide particles in reactor 240. In
30 this manner, reactors 240 and 246 can be operated continuously without changing the parameters of operation.

In the embodiment illustrated in FIG. 6B, oxygen is utilized as the oxidizing gas and is transported in via line 210 to reactor 246. Accordingly,

the embodiment of the process of the present invention illustrated in FIG. 6A is modified such that the bromine vapor produced from reactor 246 is transported via lines 242 and 225 directly to first reactor 230. Since oxygen is reactive and will not build up in the system, the need to condense the bromine vapor to a liquid to remove unreactive components, such as nitrogen, is obviated. Compressor 213 is not illustrated in FIG. 6B as substantially all commercial sources of oxygen, such as a commercial air separator unit, will provide oxygen to line 210 at the required pressure. If not, a compressor 213 could be utilized to achieve such pressure as will be evident to a skilled artisan.

In accordance with another embodiment of the process of the present invention that is illustrated in FIG. 7, the alkyl bromination and alkyl bromide conversion stages are operated in a substantially similar manner to those corresponding stages described in detail with respect to FIG. 4A except as discussed below. Residual air or oxygen and bromine vapor emanating from reactor 246 is transported via line 247, valve 248 and line 242 and valve 300 to heat exchanger or condenser 221 wherein the bromine-containing gas is cooled to a temperature in the range of about 30° C. to about 300° C. The bromine-containing vapor is then transported via line 242 to vessel or reactor 320 containing a bed 322 of a solid phase metal bromide in a reduced valence state. The metal of the metal bromide in a reduced valence state is selected from copper (Cu), iron (Fe), or molybdenum (Mo). The metal is selected for the impact of its physical and thermodynamic properties relative to the desired temperature of operation, and also for potential environmental and health impacts and cost. Preferably, copper or iron are employed as the metal, with copper being the most preferred. The solid metal bromide is preferably immobilized on a suitable attrition-resistant support, for example a synthetic amorphous silica, such as Davicat Grade 57, manufactured by Davison Catalysts of Columbia, Maryland. More preferably the metal is deposited in oxide form in a range of about 10 to 20 wt% on an alumina support with a specific surface area in the range of about 50 to 200 m²/g. In reactor 320, bromine vapor is reacted with the solid phase metal bromide, preferably retained on a suitable attrition-resistant support at temperatures

below about 300° C. and preferably between about 30° C. to about 200° C. in accordance with the following general formula wherein M² represents the metal:



- 5 In this manner, bromine is stored as a second metal bromide, i.e. 2M²Br_{n+1}, in reactor 320 while the resultant vapor containing residual air or oxygen is vented from reactor 320 via line 324, valve 326 and line 318.

The gas stream containing lower molecular weight alkanes, comprised of mixture of a feed gas (line 211) and a recycled gas stream, is transported
10 or conveyed via line 262, heat exchanger 352, wherein the gas stream is preheated to a temperature in the range of about 150° C. to about 600° C, valve 304 and line 302 to a second vessel or reactor 310 containing a bed 312 of a solid phase metal bromide in an oxidized valence state. The metal of the metal bromide in an oxidized valence state is selected from copper (Cu),
15 iron (Fe), or molybdenum (Mo). The metal is selected for the impact of its physical and thermodynamic properties relative to the desired temperature of operation, and also for potential environmental and health impacts and cost. Preferably, copper or iron are employed as the metal, with copper being the most preferred. The solid metal bromide in an oxidized state is preferably
20 immobilized on a suitable attrition-resistant support, for example a synthetic amorphous silica such as Davicat Grade 57, manufactured by Davison Catalysts of Columbia, Maryland. More preferably the metal is deposited in an oxide state in a range of 10 to 20wt% supported on an alumina support with a specific surface area of about 50 to 200 m²/g. The temperature of the gas
25 stream is from about 150° C. to about 600° C., and preferably from about 200° C. to about 450° C. In second reactor 310, the temperature of the gas stream thermally decomposes the solid phase metal bromide in an oxidized valence state to yield elemental bromine vapor and a solid metal bromide in a reduced state in accordance with the following general formula wherein M² represents
30 the metal:



The resultant bromine vapor is transported with the gas stream containing lower molecular weight alkanes via lines 314, 315, valve 317, line 330, heat exchanger 226 prior to being introduced into alkyl bromination reactor 230.

Reactors 310 and 320 may be operated in a cyclic fashion. As illustrated in FIG. 7, valve 304 is operated in the open mode to permit the gas stream containing lower molecular weight alkanes to be transported to the second reactor 310, while valve 317 is operated in the open mode to permit this gas stream with bromine vapor that is generated in reactor 310 to be transported to alkyl bromination reactor 230. Likewise, valve 306 is operated in the open mode to permit bromine vapor from reactor 246 to be transported to reactor 320, while valve 326 is operated in the open mode to permit residual air or oxygen to be vented from reactor 320. Once significant conversion of the reduced metal bromide and oxidized metal bromide in reactors 320 and 310, respectively, to the corresponding oxidized and reduced states has occurred, these valves are closed as illustrated in FIG. 8. At this point, bed 322 in reactor 320 is a bed of substantially metal bromide in an oxidized state, while bed 312 in reactor 310 is substantially metal bromide in a reduced state. As illustrated in FIG. 8, valves 304, 317, 306 and 326 are closed, and then valves 308 and 332 are opened to permit the gas stream containing lower molecular weight alkanes to be transported or conveyed via lines 262, heat exchanger 352, wherein gas stream is heated to a range of about 150° C to about 600° C, valve 308 and line, 309 to reactor 320 to thermally decompose the solid phase metal bromide in an oxidized valence state to yield elemental bromine vapor and a solid metal bromide in a reduced state. Valve 332 is also opened to permit the resultant bromine vapor to be transported with the gas stream containing lower molecular weight alkanes via lines 324 and 330 and heat exchanger 226 prior to being introduced into alkyl bromination reactor 230. In addition, valve 300 is opened to permit bromine vapor emanating from reactor 246 to be transported via line 242 through exchanger 221 into reactor 310 wherein the solid phase metal bromide in a reduced valence state reacts with bromine to effectively store bromine as a metal bromide. In addition, valve 316 is opened to permit the resulting gas, which is substantially devoid of bromine to be vented via lines 314 and 318.

The reactors are operated in this manner until significant conversion of the beds of reduced metal bromide and oxidized metal bromide in reactors 310 and 320, respectively, to the corresponding oxidized and reduced states has occurred and then the reactors are cycled back to the flow schematic
5 illustrated in FIG. 7 by opening and closing valves as previously discussed.

In the embodiment of the present invention illustrated in FIG. 9, the beds 312 and 322 contained in reactors 310 and 320, respectively, are fluidized and are connected in the manner described below to provide for continuous operation of the beds without the need to provide for equipment,
10 such as valves, to change flow direction to and from each reactor. In accordance with this embodiment, the bromine-containing gas withdrawn from the reactor 246 via line 242 is cooled to a temperature in the range of about 30° C. to about 300° C. in exchangers 370 and 372, and introduced into the bottom of reactor 320 which contains a moving solid bed 322 in a fluidized
15 state. The flow of this introduced fluid induces the particles in bed 322 to flow upwardly within reactor 320 as the bromine vapor is reacted with the reduced metal bromide entering the bottom of bed 322 in the manner as described above with respect to FIG. 7. At or near the top of the bed 322, the particles which contain substantially oxidized metal bromide on the attrition-resistant
20 support due to the substantially complete reaction of the reduced metal bromide with bromine vapor in reactor 320 are withdrawn via a weir, cyclone or other conventional means of solid/gas separation, flow by gravity down line 359 and are introduced at or near the bottom of the bed 312 in reactor 310. In the embodiment illustrated in FIG. 9, the gas stream containing lower
25 molecular weight alkanes, comprised of mixture of a feed gas (line 211) and a recycled gas stream, is transported or conveyed via line 262 and heat exchanger 352 wherein the gas stream is heated to a range of about 150° C. to about 600° C. and introduced into reactor 310. The heated gas stream thermally decomposes the solid phase metal bromide in an oxidized valence
30 state present entering at or near the bottom of bed 312 to yield elemental bromine vapor and a solid metal bromide in a reduced state. The flow of this introduced gas induces the particles in bed 312 to flow upwardly within reactor 310 as the oxidized metal bromide is thermally decomposed. At or near the

top of the bed 312, the particles which contain substantially reduced solid metal bromide on the attrition-resistant support due to the substantially complete thermal decomposition in reactor 310 are withdrawn via a weir or cyclone or other conventional means of gas/solid separation and flow by gravity down line 364 and introduced at or near the bottom of the bed 322 of particles in reactor 310. In this manner, reactors 310 and 320 can be operated continuously with changing the parameters of operation.

The process of the present invention is less expensive than conventional process since it operates at low pressures in the range of about 1 bar to about 30 bar and at relatively low temperatures in the range of about 20 ° C. to about 600 ° C. for the gas phase, and preferably about 20 ° C. to about 180 ° C. for the liquid phase. These operating conditions permit the use of less expensive equipment of relatively simple design that are constructed from readily available metal alloys or glass-lined equipment for the gas phase and polymer-lined or glass-lined vessels, piping and pumps for the liquid phase. The process of the present invention is also more efficient because less energy is required for operation and the production of excessive carbon dioxide as an unwanted byproduct is minimized. The process is capable of directly producing a mixed hydrocarbon product containing various molecular-weight components in the liquefied petroleum gas (LPG), olefin and motor gasoline fuels range that have substantial aromatic content thereby significantly increasing the octane value of the gasoline-range fuel components.

While the foregoing preferred embodiments of the invention have been described and shown, it is understood that the alternatives and modifications, such as those suggested and others, may be made thereto and fall within the scope of the invention.

CLAIMS

1. A process for converting gaseous alkanes to olefins and liquid hydrocarbons comprising:

 reacting a gaseous feed having lower molecular weight alkanes with bromine vapor to form alkyl bromides and hydrobromic acid; and

 reacting said alkyl bromides and hydrobromic acid in the presence of a synthetic crystalline alumino-silicate catalyst and at a temperature sufficient to form olefins and hydrobromic acid vapor.

2. The process of claim 1 wherein said bromine vapor is substantially dry, thereby avoiding the formation of significant carbon dioxide along with said alkyl bromides.

3. The process of claim 1 wherein said gaseous feed is natural gas.

4. The process of claim 3 wherein said natural gas is treated to remove substantially all of the carbon dioxide and sulfur compounds therefrom prior to reacting with said bromine vapor.

5. The process of claim 1 wherein said temperature is from about 250° C. to about 500° C.

6. The process of claim 5 wherein said temperature is from about 300° C. to about 450° C.

7. The process of claim 1 wherein said crystalline alumino-silicate catalyst is a zeolite catalyst.

8. The process of claim 7 wherein said zeolite catalyst is an X or Y type zeolite catalyst.

9. The process of claim 8 wherein said zeolite catalyst is a 10 X zeolite catalyst.

10. The process of claim 1 further comprising:

 removing said hydrobromic acid vapor from said olefins by neutralization reaction with an aqueous solution containing reaction products obtained by oxidizing an aqueous solution containing a metal bromide salt, the metal of said metal bromide salt being selected from Cu, Zn, Fe, Co, Ni, Mn, Ca or Mg bromide.

11. The process of claim 1 wherein said bromine vapor is produced by oxidizing an aqueous metal bromide salt solution, the metal of said metal bromide salt being selected from Cu, Zn, Fe, Co, Ni, Mn, Ca, or Mg.
12. A process for converting gaseous lower molecular weight alkanes to olefins comprising:
- reacting a gaseous feed containing lower molecular weight alkanes with bromine vapor to form alkyl bromides and hydrobromic acid;
 - reacting said alkyl bromides and hydrobromic acid in the presence of a synthetic crystalline alumino-silicate catalyst to form olefins and hydrobromic acid; and
 - converting said hydrobromic acid to bromine.
13. The process of claim 12 further comprising:
- dehydrating said olefins.
14. The process of claim 12 further comprising:
- recycling said bromine that is converted from said hydrobromic acid to said step of reacting with said gaseous feed, said bromine being recycled as a vapor.
15. A process for converting gaseous alkanes to olefins comprising:
- reacting a gaseous feed having lower molecular weight alkanes with bromine vapor to form alkyl bromides and hydrobromic acid;
 - reacting said alkyl bromides and hydrobromic acid in the presence of a synthetic crystalline alumino-silicate catalyst and at a temperature sufficient to form olefins and hydrobromic acid vapor; and
 - removing said hydrobromic acid vapor from said olefins by reacting said hydrobromic acid vapor with a metal oxide to form a metal bromide and steam.
16. The process of claim 15 wherein said the metal of said metal oxide is magnesium, calcium, vanadium, chromium, manganese, iron, cobalt, nickel, copper, zinc or tin.
17. The process of claim 15 wherein said metal oxide is supported on a solid carrier.
18. The process of claim 15 wherein said metal oxide is contained in a bed in a reactor.

19. The process of claim 15 further comprising:

reacting said metal bromide with an oxygen containing gas to obtain a metal oxide and said bromine vapor.

20. The process of claim 19 wherein said bromine vapor is used in the step of reacting said gaseous feed having lower molecular weight alkanes.