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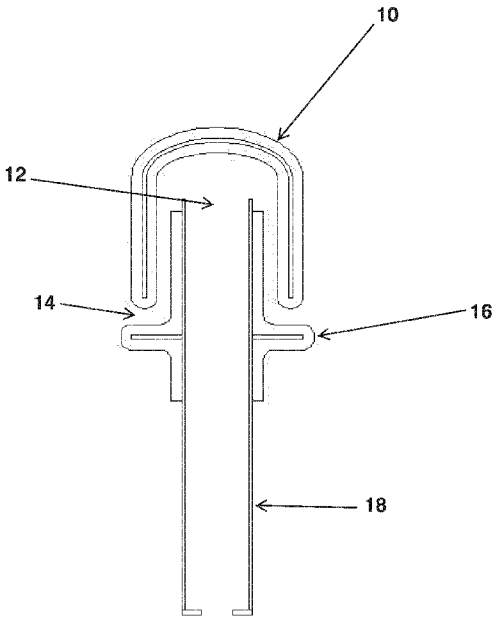
(54) Title: PROCESS FOR CONVERTING OXYGENATES TO OLEFINS

Figure 1



(57) Abstract: A process for converting oxygenates to olefins comprising: a) providing an oxygenate containing stream to an oxygenate to olefins conversion reactor; b) passing the oxygenate containing stream through a feed introduction system comprising one or more nozzles and one or more corresponding caps; c) contacting the oxygenate containing stream with a molecular sieve catalyst in the oxygenate to olefins conversion reactor to form an olefin containing product stream; and d) removing the product stream from the reactor.

WO 2015/063254 A1

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PROCESS FOR CONVERTING OXYGENATES TO OLEFINS

The present application claims the benefit of European Patent Application Serial No. 13191186.9, filed October 31, 2013.

Field of the Invention

The invention provides a process for converting oxygenates to olefins. The process includes passing the oxygenate containing stream through a feed introduction system comprising one or more nozzles and one or more corresponding caps.

Background of the Invention

Oxygenate-to-olefin ("OTO") processes are well described in the art. Typically, oxygenate-to-olefin processes are used to produce predominantly ethylene and propylene. An example of such an oxygenate-to-olefin process is described in US Patent Application Publication No. 2011/112344, which is herein incorporated by reference. The publication describes a process for the preparation of an olefin product comprising ethylene and/or propylene, comprising a step of converting an oxygenate feedstock in an oxygenate-to-olefins conversion system, comprising a reaction zone in which an oxygenate feedstock is contacted with an oxygenate conversion catalyst under oxygenate conversion conditions, to obtain a conversion effluent comprising ethylene and/or propylene.

U.S. Patent 6,737,556 describes a method and system for reducing the formation of metal catalyzed side-reaction byproducts formed in the feed vaporization and introduction system by forming and/or coating one or more of the heating devices, feed lines or feed introduction nozzles with a material that is resistant to the formation of metal catalyzed side reaction byproducts.

U.S Patent 7,034,196 describes a method and apparatus for reducing the amount of metal catalyzed side-reaction byproducts formed in the feed vaporization and introduction system of a methanol to olefin reactor system by maintaining a sufficiently low temperature in the feed vaporization and introduction system.

Summary of the Invention

The invention provides a process for converting oxygenates to olefins comprising: a) providing an oxygenate containing stream to an oxygenate to olefins conversion reactor; b) passing the oxygenate containing stream through a feed introduction system comprising one or

more nozzles and one or more corresponding caps; c) contacting the oxygenate containing stream with a molecular sieve catalyst in the oxygenate to olefins conversion reactor to form an olefin containing product stream; and d) removing the product stream from the reactor.

The invention further provides a system for converting oxygenates to olefins comprising: a) an oxygenate to olefins conversion reactor; b) one or more catalyst inlets for introducing catalyst into the reactor; c) one or more feed inlet nozzles located at the bottom of the reactor for introducing an oxygenate containing feed into the reactor; and d) one or more protective caps located above each of the feed inlet nozzles.

The invention also provides a feed introduction system for an oxygenates to olefins conversion system comprising: a) one or more feed inlet nozzles located in the bottom of an oxygenate to olefins conversion reactor; and b) one or more protective caps located above each of the feed inlet nozzles.

Brief Description of the Drawings

Figure 1 depicts one embodiment of the nozzle and cap according to the invention.

Figure 2 depicts one embodiment of a possible nozzle layout.

Figure 3 depicts one embodiment of a reactor with nozzles.

Detailed Description of the Invention

The invention provides an improved process for converting the oxygenates to olefins, and it specifically provides an improved method of feeding the oxygenate containing feed into the reactor comprising passing the oxygenate containing stream through a feed introduction system comprising one or more nozzles and one or more corresponding caps.

The corresponding caps located above the nozzles prevent the nozzle from being in direct contact with reactor temperatures so that the feed nozzle is not heated to full reactor temperatures. This reduces the metal catalyzed side reactions that may occur at reactor temperatures with the oxygenate containing feed stream. The use of these caps allows for the nozzles to be located on the floor of the reactor which provides for more even dispersion than side entry nozzles that are typically used. This also provides for better gas-catalyst mixing.

The cap may comprise refractory and this prevents erosion of the cap by circulating catalyst. Another benefit is that the cap design has a relatively high cross-sectional area at the annulus where the gas exits the cap into the reactor. This provides for lower gas velocities as the cross-sectional area determines the velocity of a given flow of feed. This lower velocity is

acceptable because there is no required minimum exit velocity to keep the nozzles clear of catalyst unlike for grid type distributors.

The oxygenate to olefins process receives as a feedstock a stream comprising one or more oxygenates. An oxygenate is an organic compound that contains at least one oxygen atom. The oxygenate is preferably one or more alcohols, preferably aliphatic alcohols where the aliphatic moiety has from 1 to 20 carbon atoms, preferably from 1 to 10 carbon atoms, more preferably from 1 to 5 carbon atoms and most preferably from 1 to 4 carbon atoms. The alcohols that can be used as a feed to this process include lower straight and branched chain aliphatic alcohols. In addition, ethers and other oxygen containing organic molecules can be used. Suitable examples of oxygenates include methanol, ethanol, n-propanol, isopropanol, methyl ethyl ether, dimethyl ether, diethyl ether, di-isopropyl ether, formaldehyde, dimethyl carbonate, dimethyl ketone, acetic acid and mixtures thereof. In a preferred embodiment, the feedstock comprises one or more of methanol, ethanol, dimethyl ether, diethyl ether or a combination thereof, more preferably methanol or dimethyl ether and most preferably methanol.

In one embodiment, the oxygenate is obtained as a reaction product of synthesis gas. Synthesis gas can, for example, be generated from fossil fuels, such as from natural gas or oil, or from the gasification of coal. In another embodiment, the oxygenate is obtained from biomaterials, such as through fermentation.

The oxygenate feedstock can be obtained from a pre-reactor, which converts methanol at least partially into dimethylether and water. Water may be removed, by e.g., distillation. In this way, less water is present in the process of converting oxygenates to olefins, which has advantages for the process design and lowers the severity of hydrothermal conditions to which the catalyst is exposed.

The oxygenate to olefins process, may in certain embodiments, also receive an olefin co-feed. This co-feed may comprise olefins having carbon numbers of from 1 to 8, preferably from 3 to 6 and more preferably 4 or 5. Examples of suitable olefin co-feeds include butene, pentene and hexene.

Preferably, the oxygenate feed comprises one or more oxygenates and olefins, more preferably oxygenates and olefins in an oxygenate:olefin molar ratio in the range of from 1000:1 to 1:1, preferably 100:1 to 1:1. More preferably, in a oxygenate:olefin molar ratio in

the range of from 20:1 to 1:1, more preferably in the range of 18:1 to 1:1, still more preferably in the range of 15:1 to 1:1, even still more preferably in the range of 14:1 to 1:1. It is preferred to convert a C4 olefin, recycled from the oxygenate to olefins conversion reaction together with an oxygenate, to obtain a high yield of ethylene and propylene, therefore preferably at least one mole of oxygenate is provided for every mole of C4 olefin.

The olefin co-feed may also comprise paraffins. These paraffins may serve as diluents or in some cases they may participate in one or more of the reactions taking place in the presence of the catalyst. The paraffins may include alkanes having carbon numbers from 1 to 10, preferably from 3 to 6 and more preferably 4 or 5. The paraffins may be recycled from separation steps occurring downstream of the oxygenate to olefins conversion step.

The oxygenate to olefins process, may in certain embodiments, also receive a diluent co-feed to reduce the concentration of the oxygenates in the feed and suppress side reactions that lead primarily to high molecular weight products. The diluent should generally be non-reactive to the oxygenate feedstock or to the catalyst. Possible diluents include helium, argon, nitrogen, carbon monoxide, carbon dioxide, methane, water and mixtures thereof. The more preferred diluents are water and nitrogen with the most preferred being water.

The diluent may be used in either liquid or vapor form. The diluent may be added to the feedstock before or at the time of entering the reactor or added separately to the reactor or added with the catalyst. In one embodiment, the diluents is added in an amount in the range of from 1 to 90 mole percent, more preferably from 1 to 80 mole percent, more preferably from 5 to 50 mole percent, most preferably from 5 to 40 mole percent.

During the conversion of the oxygenates in the oxygenate to olefins conversion reactor, steam is produced as a by-product, which serves as an in-situ produced diluent. Typically, additional steam is added as diluent. The amount of additional diluent that needs to be added depends on the in-situ water make, which in turn depends on the composition of the oxygenate feed. Where the diluent provided to the reactor is water or steam, the molar ratio of oxygenate to diluent is between 10:1 and 1:20.

The oxygenate feed is contacted with the catalyst at a temperature in the range of from 200 to 1000 °C, preferably of from 300 to 800 °C, more preferably of from 350 to 700 °C, even more preferably of from 450 to 650°C. The feed may be contacted with the catalyst at a temperature in the range of from 530 to 620 °C, or preferably of from 580 to 610 °C. The feed

may be contacted with the catalyst at a pressure in the range of from 0.1 kPa (1 mbar) to 5 MPa (50 bar), preferably of from 100 kPa (1 bar) to 1.5 MPa (15 bar), more preferably of from 100 kPa (1 bar) to 300 kPa (3 bar). Reference herein to pressures is to absolute pressures.

A wide range of WHSV for the feedstock may be used. WHSV is defined as the mass of the feed (excluding diluents) per hour per mass of catalyst. The WHSV should preferably be in the range of from 1 hr^{-1} to 5000 hr^{-1}.

The process takes place in a reactor and the catalyst may be present in the form of a fixed bed, a moving bed, a fluidized bed, a dense fluidized bed, a fast or turbulent fluidized bed, a circulating fluidized bed. In addition, riser reactors, hybrid reactors or other reactor types known to those skilled in the art may be used. In another embodiment, more than one of these reactor types may be used in series. In one embodiment, the reactor is a riser reactor. The advantage of a riser reactor is that it allows for very accurate control of the contact time of the feed with the catalyst, as riser reactors exhibit a flow of catalyst and reactants through the reactor that approaches plug flow.

The feedstocks described above are converted primarily into olefins. The olefins produced from the feedstock typically have from 2 to 30 carbon atoms, preferably from 2 to 8 carbon atoms, more preferably from 2 to 6 carbon atoms, most preferably ethylene and/or propylene. In addition to these olefins, diolefins having from 4 to 18 carbon atoms, conjugated or nonconjugated dienes, polyenes, vinyl monomers and cyclic olefins may be produced in the reaction.

In a preferred embodiment, the feedstock, preferably one or more oxygenates, is converted in the presence of a molecular sieve catalyst into olefins having from 2 to 6 carbon atoms. Preferably the oxygenate is methanol, and the olefins are ethylene and/or propylene.

The products from the reactor are typically separated and/or purified to prepare separate product streams in a recovery system. Such systems typically comprise one or more separation, fractionation or distillation towers, columns, and splitters and other associated equipment, for example, various condensers, heat exchangers, refrigeration systems or chill trains, compressors, knock-out drums or pots, pumps and the like.

The recovery system may include a demethanizer, a deethanizer, a depropanizer, a wash tower often referred to as a caustic wash tower and/or quench tower, absorbers,

adsorbers, membranes, an ethylene-ethane splitter, a propylene-propane splitter, a butene-butane splitter and the like.

Typically in the recovery system, additional products, by-products and/or contaminants may be formed along with the preferred olefin products. The preferred products, ethylene and propylene are preferably separated and purified for use in derivative processes such as polymerization processes.

In addition to the propylene and ethylene, the products may comprise C4+ olefins, paraffins and aromatics that may be further reacted, recycled or otherwise further treated to increase the yield of the desired products and/or other valuable products. C4+ olefins may be recycled to the oxygenate to olefins conversion reaction or fed to a separate reactor for cracking. The paraffins may also be cracked in a separate reactor, and/or removed from the system to be used elsewhere or possibly as fuel.

Although less desired, the product will typically comprise some aromatic compounds such as benzene, toluene and xylenes. Although it is not the primary aim of the process, xylenes can be seen as a valuable product. Xylenes may be formed in the OTO process by the alkylation of benzene and, in particular, toluene with oxygenates such as methanol. Therefore, in a preferred embodiment, a separate fraction comprising aromatics, in particular benzene, toluene and xylenes is separated from the gaseous product and at least in part recycled to the oxygenate to olefins conversion reactor as part of the oxygenate feed. Preferably, part or all of the xylenes in the fraction comprising aromatics are withdrawn from the process as a product prior to recycling the fraction comprising aromatics to the oxygenate to olefins conversion reactor.

The C4+ olefins and paraffins formed in the oxygenate to olefins conversion reactor may be further reacted in an additional reactor containing the same or a different molecular sieve catalyst. In this additional reactor, the C4+ feed is converted over the molecular sieve catalyst at a temperature in the range of from 500 to 700 °C. The additional reactor is also referred to as an OCP reactor and the process that takes place in this reactor is referred to as an olefin cracking process. In contact with the molecular sieve catalyst, at least part of the olefins in the C4+ feed is converted to a product, which includes at least ethylene and/or propylene and preferably both. In addition to ethylene and/or propylene, the gaseous product may

comprise higher olefins, i.e. C4+ olefins, and paraffins. The gaseous product is retrieved from the second reactor as part of a second reactor effluent stream.

The olefin feed is contacted with the catalyst at a temperature in the range of from 500 to 700 °C, preferably of from 550 to 650°C, more preferably of from 550 to 620°C, even more preferably of from 580 to 610°C; and a pressure in the range of from 0.1 kPa (1 mbara) to 5 MPa (50 bara), preferably of from 100 kPa (1 bara) to 1.5 MPa (15 bara), more preferably of from 100 kPa (1 bara) to 300 kPa (3 bara). Reference herein to pressures is to absolute pressures.

In one embodiment, the C4+ olefins are separated into at least two fractions: a C4 olefin fraction and a C5+ olefin fraction. In this embodiment, the C4 olefins are recycled to the oxygenate to olefins conversion reactor and the C5+ olefins are fed to the OCP reactor. The cracking behavior of C4 olefins and C5 olefins is believed to be different when contacted with a molecular sieve catalyst, in particular above 500 °C.

The cracking of C4 olefins is an indirect process which involves a primary oligomerisation process to a C8, C12 or higher olefin followed by cracking of the oligomers to lower molecular weight hydrocarbons including ethylene and propylene, but also, amongst other things, to C5 to C7 olefins, and by-products such as C2 to C6 paraffins, cyclic hydrocarbons and aromatics. In addition, the cracking of C4 olefins is prone to coke formation, which places a restriction on the obtainable conversion of the C4 olefins. Generally, paraffins, cyclics and aromatics are not formed by cracking. They are formed by hydrogen transfer reactions and cyclisation reactions. This is more likely in larger molecules. Hence the C4 olefin cracking process, which as mentioned above includes intermediate oligomerisation, is more prone to by-product formation than direct cracking of C5 olefins. The conversion of the C4 olefins is typically a function of the temperature and space time (often expressed as the weight hourly space velocity). With increasing temperature and decreasing weight hourly space velocity (WHSV) conversion of the C4 olefins in the feed to the OCP increases. Initially, the ethylene and propylene yields increase, but, at higher conversions, yield decreases at the cost of a higher by-product make and, in particular, a higher coke make, limiting significantly the maximum yield obtainable.

Contrary to C4 olefins, C5 olefin cracking is ideally a relatively straight forward-process whereby the C5 olefin cracks into a C2 and a C3 olefin, in particular above 500°C.

This cracking reaction can be run at high conversions, up to 100%, while maintaining, at least compared to C4 olefins, high ethylene and propylene yields with a significantly lower by-product and coke make. Although, C5+ olefins can also oligomerise, this process competes with the more beneficial cracking to ethylene and propylene.

In a preferred embodiment of the process according to the present invention, instead of cracking the C4 olefins in the OCP reactor, the C4 olefins are recycled to the oxygenate to olefins conversion reactor. Again without wishing to be bound by any particular theory, it is believed that in the oxygenate to olefins conversion reactor the C4 olefins are alkylated with, for instance, methanol to C5 and/or C6 olefins. These C5 and/or C6 olefins may subsequently be converted into at least ethylene and/or propylene. The main by-products from this oxygenate to olefins conversion reaction are again C4 and C5 olefins, which can be recycled to the oxygenate to olefins conversion reactor and olefin cracking reactor, respectively.

Therefore, preferably, where the gaseous products further include C4 olefins, at least part of the C4 olefins are provided to (i) the oxygenate to olefins conversion reactor together with or as part of the oxygenate feed, and/or (ii) the olefin cracking reactor as part of the olefin feed, more preferably at least part of the C4 olefins is provided to the oxygenate to olefins conversion reactor together with or as part of the oxygenate feed.

Preferably, where the gaseous products further include C5 olefins, at least part of the C5 olefins are provided to the olefin cracking reactor as part of the olefin feed. Preferably, the olefin feed to the olefin cracking reactor comprises C4+ olefins, preferably C5+ olefins, more preferably C5 olefins.

In a preferred embodiment, the oxygenate to olefins conversion reactor and the optional OCP reactor are operated as riser reactors where the catalyst and feedstock are fed at the base of the riser and an effluent stream with entrained catalyst exits the top of the riser. In this embodiment, gas/solid separators are necessary to separate the entrained catalyst from the reactor effluent. The gas/solid separator may be any separator suitable for separating gases from solids. Preferably, the gas/solid separator comprises one or more centrifugal separation units, preferably cyclone units, optionally combined with a stripper section.

The reactor effluent is preferably cooled in or immediately after the gas/solid separator to terminate the conversion process and prevent the formation of by-products outside the reactors. The cooling may be achieved by use of a water quench.

Once the catalyst is separated from the effluent, the catalyst may be returned to the reaction zone from which it came, another reaction zone, a stripping zone or a regeneration zone. Further, the catalyst that has been separated in the gas/solid separator may be combined with catalyst from other gas/solid separators before it is sent to a reaction zone, a stripping zone or the regeneration zone.

During conversion of the oxygenates to olefins, carbonaceous deposits known as "coke" are formed on the surface of and/or within the molecular sieve catalysts. To avoid a significant reduction in activity of the catalyst, the catalyst must be regenerated by burning off the coke deposits.

In one embodiment, a portion of the coked molecular sieve catalyst is withdrawn from the reactor and introduced into a regeneration system. The regeneration system comprises a regenerator where the coked catalyst is contacted with a regeneration medium, preferably an oxygen-containing gas, under regeneration temperature, pressure and residence time conditions.

Examples of suitable regeneration media include oxygen, O_3, SO_3, N_2O, NO, NO_2, N_2O_5, air, air enriched with oxygen, air diluted with nitrogen or carbon dioxide, oxygen and water, carbon monoxide and/or hydrogen. The regeneration conditions are those capable of burning at least a portion of the coke from the coked catalyst, preferably to a coke level of less than 75% of the coke level on the catalyst entering the regenerator. More preferably the coke level is reduced to less than 50% of the coke level on the catalyst entering the regenerator and most preferably the coke level is reduced to less than 30% of the coke level on the catalyst entering the regenerator. Complete removal of the coke is not necessary as this may result in degradation of the catalyst.

The regeneration temperature is in the range of from 200 °C to 1500 °C, preferably from 300 °C to 1000 °C, more preferably from 450 °C to 700 °C and most preferably from 500 °C to 700 °C. In a preferred embodiment, the catalyst is regenerated at a temperature in the range of from 550 to 650 °C.

The preferred residence time of the coked molecular sieve catalyst in the regenerator is in the range of from 1 minute to several hours, most preferably 1 minute to 100 minutes. The preferred volume of oxygen in the regeneration medium is from 0.01 mole percent to 10 mole percent based on the total volume of the regeneration medium.

In one embodiment, regeneration promoters, typically metal containing compounds such as platinum and palladium are added to the regenerator directly or indirectly, for example with the coked catalyst composition. In another embodiment, a fresh molecular sieve catalyst is added to the regenerator.

In an embodiment, a portion of the regenerated molecular sieve catalyst from the regenerator is returned to the reactor, directly to the reaction zone or indirectly by pre-contacting with the feedstock.

The burning of coke is an exothermic reaction and in certain embodiments, the temperature in the regeneration system is controlled to prevent it from rising too high. Various known techniques for cooling the system and/or the regenerated catalyst may be employed including feeding a cooled gas to the regenerator, or passing the regenerated catalyst through a catalyst cooler. A portion of the cooled regenerated catalyst may be returned to the regenerator while another portion is returned to the reactor.

In certain embodiments, there is not sufficient coke on the catalyst to raise the temperature of the catalyst to desired levels. In one embodiment, a liquid or gaseous fuel may be fed to the regenerator where it will combust and provide additional heat to the catalyst.

Catalysts suitable for use in the conversion of oxygenates to olefins may be made from practically any small or medium pore molecular sieve. One example of a suitable type of molecular sieve is a zeolite. Suitable zeolites include, but are not limited to AEI, AEL, AFT, AFO, APC, ATN, ATT, ATV, AWW, BIK, CAS, CHA, CHI, DAC, DDR, EDI, ERI, EUO, FER, GOO, HEU, KFI, LEV, LOV, LTA, MFI, MEL, MON, MTT, MTW, PAU, PHI, RHO, ROG, THO, TON and substituted forms of these types. Suitable catalysts include those containing a zeolite of the ZSM group, in particular of the MFI type, such as ZSM-5, the MTT type, such as ZSM-23, the TON type, such as ZSM-22, the MEL type, such as ZSM-11, and the FER type. Other suitable zeolites are for example zeolites of the STF-type, such as SSZ-35, the SFF type, such as SSZ-44 and the EU-2 type, such as ZSM-48. Preferred zeolites for this process include ZSM-5, ZSM-22 and ZSM-23.

A suitable molecular sieve catalyst may have a silica-to-alumina ratio (SAR) of less than 280, preferably less than 200 and more preferably less than 100. The SAR may be in the range of from 10 to 280, preferably from 15 to 200 and more preferably from 20 to 100.

A preferred MFI-type zeolite for the oxygenate to olefins conversion catalyst has a silica-to-alumina ratio, SAR, of at least 60, preferably at least 80. More preferred MFI-type zeolite has a silica-to-alumina ratio, SAR, in the range of 60 to 150, preferably in the range of 80 to 100.

The zeolite-comprising catalyst may comprise more than one zeolite. In that case it is preferred that the catalyst comprises at least a more-dimensional zeolite, in particular of the MFI type, more in particular ZSM-5, or of the MEL type, such as zeolite ZSM-11, and a one-dimensional zeolite having 10-membered ring channels, such as of the MTT and/or TON type.

It is preferred that zeolites in the hydrogen form are used in the zeolite-comprising catalyst, e.g., HZSM-5, HZSM-11, and HZSM-22, HZSM-23. Preferably at least 50wt%, more preferably at least 90wt%, still more preferably at least 95wt% and most preferably 100wt% of the total amount of zeolite used is in the hydrogen form. It is well known in the art how to produce such zeolites in the hydrogen form.

Another example of suitable molecular sieves are siliocoaluminophosphates (SAPOs). SAPOs have a three dimensional microporous crystal framework of PO_2^+, AlO_2^-, and SiO_2 tetrahedral units. Suitable SAPOs include SAPO-17, -18, 34, -35, -44, but also SAPO-5, -8, -11, -20, -31, -36, 37, -40, -41, -42, -47 and –56; aluminophosphates (AlPO) and metal substituted (silico)aluminophosphates (MeAlPO), wherein the Me in MeAlPO refers to a substituted metal atom, including metal selected from one of Group IA, IIA, IB, IIIB, IVB, VB, VIB, VIIB, VIIIB and lanthanides of the Periodic Table of Elements. Preferred SAPOs for this process include SAPO-34, SAPO-17 and SAPO-18. Preferred substituent metals for the MeAlPO include Co, Cr, Cu, Fe, Ga, Ge, Mg, Mn, Ni, Sn, Ti, Zn and Zr.

The molecular sieves described above are formulated into molecular sieve catalyst compositions for use in the oxygenates to olefins conversion reaction and the olefin cracking step. The molecular sieves are formulated into catalysts by combining the molecular sieve with a binder and/or matrix material and/or filler and forming the composition into particles by techniques such as spray-drying, pelletizing, or extrusion. The molecular sieve may be further processed before being combined with the binder and/or matrix. For example, the molecular sieve may be milled and/or calcined.

Suitable binders for use in these molecular sieve catalyst compositions include various types of aluminas, aluminophosphates, silicas and/or other inorganic oxide sol. The binder

acts like glue binding the molecular sieves and other materials together, particularly after thermal treatment. Various compounds may be added to stabilize the binder to allow processing.

Matrix materials are usually effective at among other benefits, increasing the density of the catalyst composition and increasing catalyst strength (crush strength and/or attrition resistance). Suitable matrix materials include one or more of the following: rare earth metals, metal oxides including titania, zirconia, magnesia, thoria, beryllia, quartz, silica or sols, and mixtures thereof, for example, silica-magnesia, silica-zirconia, silica-titania, and silica-alumina. In one embodiment, matrix materials are natural clays, for example, kaolin. A preferred matrix material is kaolin.

In one embodiment, the molecular sieve, binder and matrix material are combined in the presence of a liquid to form a molecular sieve catalyst slurry. The amount of binder is in the range of from 2 to 40 wt%, preferably in the range of from 10 to 35 wt%, more preferably in the range of from 15 to 30 wt%, based on the total weight of the molecular sieve, binder and matrix material, excluding liquid (after calcination).

After forming the slurry, the slurry may be mixed, preferably with rigorous mixing to form a substantially homogeneous mixture. Suitable liquids include one or more of water, alcohols, ketones, aldehydes and/or esters. Water is the preferred liquid. In one embodiment, the mixture is colloid-milled for a period of time sufficient to produce the desired texture, particle size or particle size distribution.

The molecular sieve, matrix and optional binder can be in the same or different liquids and are combined in any order together, simultaneously, sequentially or a combination thereof. In a preferred embodiment, water is the only liquid used.

In a preferred embodiment, the slurry is mixed or milled to achieve a uniform slurry of sub-particles that is then fed to a forming unit. A slurry of the zeolite may be prepared and then milled before combining with the binder and/or matrix. In a preferred embodiment, the forming unit is a spray dryer. The forming unit is typically operated at a temperature high enough to remove most of the liquid from the slurry and from the resulting molecular sieve catalyst composition. In a preferred embodiment, the particles are then exposed to ion-exchange using an ammonium nitrate or other appropriate solution.

In one embodiment, the ion exchange is carried out before the phosphorous impregnation. The ammonium nitrate is used to ion exchange the zeolite to remove alkali ions. The zeolite can be impregnated with phosphorous using phosphoric acid followed by a thermal treatment to H+ form. In another embodiment, the ion exchange is carried out after the phosphorous impregnation. In this embodiment, alkali phosphates or phosphoric acid may be used to impregnate the zeolite with phosphorous, and then the ammonium nitrate and heat treatment are used to ion exchange and convert the zeolite to the H+ form.

Alternatively to spray drying the catalyst may be formed into spheres, tablets, rings, extrudates or any other shape known to one of ordinary skill in the art. The catalyst may be extruded into various shapes, including cylinders and trilobes.

The average particle size is in the range of from 1-200 μm, preferably from 50-100 μm. If extrudates are formed, then the average size is in the range of from 1 mm to 10 mm, preferably from 2 mm to 7 mm.

The catalyst may further comprise phosphorus as such or in a compound, i.e. phosphorus other than any phosphorus included in the framework of the molecular sieve. It is preferred that a MEL or MFI-type zeolite comprising catalyst additionally comprises phosphorus.

The molecular sieve catalyst is prepared by first forming a molecular sieve catalyst precursor as described above, optionally impregnating the catalyst with a phosphorous containing compound and then calcining the catalyst precursor to form the catalyst. The phosphorous impregnation may be carried out by any method known to one of skill in the art.

The phosphorus-containing compound preferably comprises a phosphorus species such as PO_4^{3-}, P-$(OCH_3)_3$, or P_2O_5, especially PO_4^{3-}. Preferably the phosphorus-containing compound comprises a compound selected from the group consisting of ammonium phosphate, ammonium dihydrogen phosphate, dimethylphosphate, metaphosphoric acid and trimethyl phosphite and phosphoric acid, especially phosphoric acid. The phosphorus containing compound is preferably not a Group II metal phosphate. Group II metal species include magnesium, calcium, strontium and barium; especially calcium.

In one embodiment, phosphorus can be deposited on the catalyst by impregnation using acidic solutions containing phosphoric acid (H_3PO_4). The concentration of the solution

can be adjusted to impregnate the desired amount of phosphorus on the precursor. The catalyst precursor may then be dried.

The catalyst precursor, containing phosphorous (either in the framework or impregnated) is calcined to form the catalyst. The calcination of the catalyst is important to determining the performance of the catalyst in the oxygenate to olefins process.

The calcination may be carried out in any type of calciner known to one of ordinary skill in the art. The calcination may be carried out in a tray calciner, a rotary calciner, or a batch oven optionally in the presence of an inert gas and/or oxygen and/or steam

The calcination may be carried out at a temperature in the range of from 400 °C to 1000 °C, preferably in a range of from 450 °C to 800 °C, more preferably in a range of from 500 °C to 700 ° C. Calcination time is typically dependent on the degree of hardening of the molecular sieve catalyst composition and the temperature and ranges from about 15 minutes to about 2 hours.

The calcination temperatures described above are temperatures that are reached for at least a portion of the calcination time. For example, in a rotary calciner, there may be separate temperature zones that the catalyst passes through. For example, the first zone may be at a temperature in the range of from 100 to 300 °C. At least one of the zones is at the temperatures specified above. In a stationary calciner, the temperature is increased from ambient to the calcination temperatures above and so the temperature is not at the calcination temperature for the entire time.

In a preferred embodiment, the calcination is carried out in air at a temperature of from 500 °C to 600 °C. The calcination is carried out for a period of time from 30 minutes to 15 hours, preferably from 1 hour to 10 hours, more preferably from 1 hour to 5 hours.

The calcination is carried out on a bed of catalyst. For example, if the calcination is carried out in a tray calciner, then the catalyst precursor added to the tray forms a bed which is typically kept stationary during the calcination. If the calcination is carried out in a rotary calciner, then the catalyst added to the rotary drum forms a bed that although not stationary does maintain some form and shape as it passes through the calciner.

The process of the invention provides for feeding the oxygenate containing stream into the reactor via a feed introduction system that comprises feed nozzles located at or near the bottom of the reactor with corresponding caps above each nozzle. The feed introduction

system works in a riser reactor, turbulent fluidized bed reactor, fast fluidized bed reactor, and/or any combination thereof.

The oxygenate containing stream is preferably methanol or dimethyl ether. The oxygenate containing stream is preferably fed into the reactor as a vapor.

The nozzles are preferably located in the bottom of the reactor and allow for the oxygenate containing stream to be fed into the bottom of the reactor. The nozzles may pass through the bottom of the reactor or there may be a tube sheet above the bottom of the reactor with the nozzles passing through the tube sheet into a single feed inlet chamber. The interior surface of the nozzles may be coated with an erosion resistant coating. The erosion resistant coating is preferably not refractory. The bottom wall or bottom tubesheet of the reactor may be insulated to prevent heat conduction from the reactor to the bottom wall or bottom tubesheet. The nozzles may be manufactured of carbon steel or any metal or metal alloy known to one of ordinary skill in the art. Since the nozzles do not reach the higher temperatures found in the reactor, the materials of construction can be cheaper and less resistant to temperature than those that would be required for typical nozzles that were in contact with the reactor temperatures.

The nozzles may have an orifice located at the entrance to the nozzle to set a pressure drop and prevent catalyst backflow. The downward turn in the nozzle also helps prevent catalyst intrusion.

The caps are preferably located above each and every nozzle although it can be envisioned that the caps may only be located above one or more of the nozzles. The caps are preferably located and designed in such a manner that at least a portion of the flow exiting the nozzle is directed in a direction at or below horizontal. In another embodiment, at least a portion of the flow exiting the nozzle is directed towards the bottom of the reactor or bottom tubesheet.

The caps are preferably rounded caps that are positioned above each nozzle. The caps may be hemispherical, elliptical, conical or cylindrical in shape with the opening of the cap located above the nozzle. The flow is directed from the nozzle into the cap and then directed downwards and under the edge of the cap. The caps may be coated with an erosion resistant coating. The erosion resistant coating can be any coating known to one of ordinary skill in the art. The coating is preferably selected from the group consisting of ceramics, fire brick, high

temperature calcium silicate, alumina, silica-alumina ceramics, diatomaceous silica brick, carbide, cement or refractory.

One embodiment of a nozzle and cap configuration is depicted in Figure 1. The figure depicts a nozzle 18 that extends from below the reactor bottom sheet 16 to above the reactor bottom sheet. The nozzle provides a flow path for feed from a feed system through the nozzle 18 and to the outlet of the nozzle 12. The flow path extends from the outlet 12 to the outlet of the nozzle/cap outlet 14. The cap 10 is positioned above the nozzle 18.

An embodiment of a nozzle layout is depicted in Figure 2. The figure shows a plurality of nozzles 20 located on the reactor bottom sheet 22.

Figure 3 depicts an embodiment of a reactor having a reaction zone 30 that has a feed zone 34. The feed passes into the feed zone 34 and then passes via the nozzle/cap assemblies 32. The reactor bottom sheet 36 separates the feed zone 34 from the reaction zone 30.

CLAIMS

1. A process for converting oxygenates to olefins comprising:
   a. providing an oxygenate containing stream to an oxygenate to olefins conversion reactor;
   b. passing the oxygenate containing stream through a feed introduction system comprising one or more nozzles and one or more corresponding caps;
   c. contacting the oxygenate containing stream with a molecular sieve catalyst in the oxygenate to olefins conversion reactor to form an olefin containing product stream; and
   d. removing the product stream from the reactor.
2. The process of claim 1 wherein the reactor is a riser reactor, turbulent fluidized bed reactor, fast fluidized bed reactor and/or a combination thereof.
3. The process of any of claims 1-2 wherein the oxygenate containing stream comprises methanol and/or dimethylether.
4. The process of any of claims 1-3 wherein the oxygenate containing stream comprises an olefin co-feed and/or a diluent stream.
5. The process of claim 4 wherein the diluent stream comprises steam.
6. The process of any of claims 1-5 wherein the oxygenate containing stream is a vapor when introduced into the reactor.
7. The process of any of claims 1-6 wherein the inner surface of the nozzles has an erosion resistant coating.
8. The process of claim 7 wherein the coating is not refractory.
9. The process of any of claims 1-8 wherein the nozzles are located in and allow feed to pass through the bottom wall of the reactor.
10. The process of claim 9 wherein the bottom wall of the reactor is insulated to prevent the reactor heat from heating the metal of the bottom wall.
11. The process of any of claims 1-10 wherein the caps are such that at least a portion of the oxygenate containing stream that passes through the nozzles is directed in a direction at or below horizontal.

12. The process of any of claims 1-11 wherein the caps are rounded caps that are positioned above each of the nozzles.
13. The process of claim 12 wherein the caps are hemispherical, elliptical and/or cylindrical caps that are positioned above each of the nozzles.
14. The process of any of claims 1-13 wherein the caps are coated with an erosion resistant coating.
15. The process of claim 14 wherein the caps are coated with a coating selected from the group consisting of ceramics, fire brick, high temperature calcium silicate, alumina, silica-alumina ceramics, diatomaceous silica brick, carbide, cement or refractory.
16. The process of any of claims 1-15 wherein the molecular sieve catalyst is a zeolite catalyst.
17. The process of claim 16 wherein the molecular sieve catalyst comprises ZSM-5.
18. The process of any of claims 1-17 further comprising passing the product stream through downstream processing equipment to separate the products from other components in the product stream.
19. The process of any of claims 1-18 wherein the nozzle is made of carbon steel.
20. The process of any of claims 1-19 wherein the caps are made of carbon steel.
21. A system for converting oxygenates to olefins comprising:
    a. an oxygenate to olefins conversion reactor;
    b. one or more catalyst inlets for introducing catalyst into the reactor;
    c. one or more feed inlet nozzles located at the bottom of the reactor for introducing an oxygenate containing feed into the reactor; and
    d. one or more protective caps located above each of the feed inlet nozzles.
22. The system of claim 21 wherein the catalyst is fresh catalyst, recycled catalyst, regenerated catalyst or a combination thereof.
23. The system of any of claims 21-22 wherein the inner surface of the nozzles has an erosion resistant coating.
24. The system of claim 23 wherein the coating is not refractory.
25. The system of any of claims 21-24 wherein the caps are rounded caps that are positioned above each of the nozzles.

26. The system of claim 25 wherein the caps are hemispherical, elliptical and/or cylindrical caps that are positioned above each of the nozzles.
27. The system of any of claims 21-26 wherein the caps are coated with an erosion resistant coating.
28. The system of claim 27 wherein the caps are coated with a coating selected from the group consisting of ceramics, fire brick, high temperature calcium silicate, alumina, silica-alumina ceramics, diatomaceous silica brick, carbide, cement or refractory.
29. The system of any of claims 21-28 wherein the nozzles are made of carbon steel.
30. The system of any of claims 21-29 wherein the caps are made of carbon steel.
31. A feed introduction system for an oxygenates to olefins conversion system comprising:
    a. one or more feed inlet nozzles located in the bottom of an oxygenate to olefins conversion reactor; and
    b. one or more protective caps located above each of the feed inlet nozzles.
32. The system of claim 31 wherein the inner surface of the nozzles has an erosion resistant coating.
33. The system of any of claims 31-32 wherein the coating is not refractory.
34. The system of any of claims 31-33 wherein the caps are rounded caps that are positioned above each of the nozzles.
35. The system of claim 34 wherein the caps are hemispherical, elliptical and/or cylindrical caps that are positioned above each of the nozzles.
36. The system of any of claims 31-35 wherein the caps are coated with an erosion resistant coating.
37. The system of claim 36 wherein the caps are coated with a coating selected from the group consisting of ceramics, fire brick, high temperature calcium silicate, alumina, silica-alumina ceramics, diatomaceous silica brick, cement or refractory.
38. The system of any of claims 31-37 wherein the nozzles are made of carbon steel.
39. The system of any of claims 31-38 wherein the caps are made of carbon steel.

Figure 1

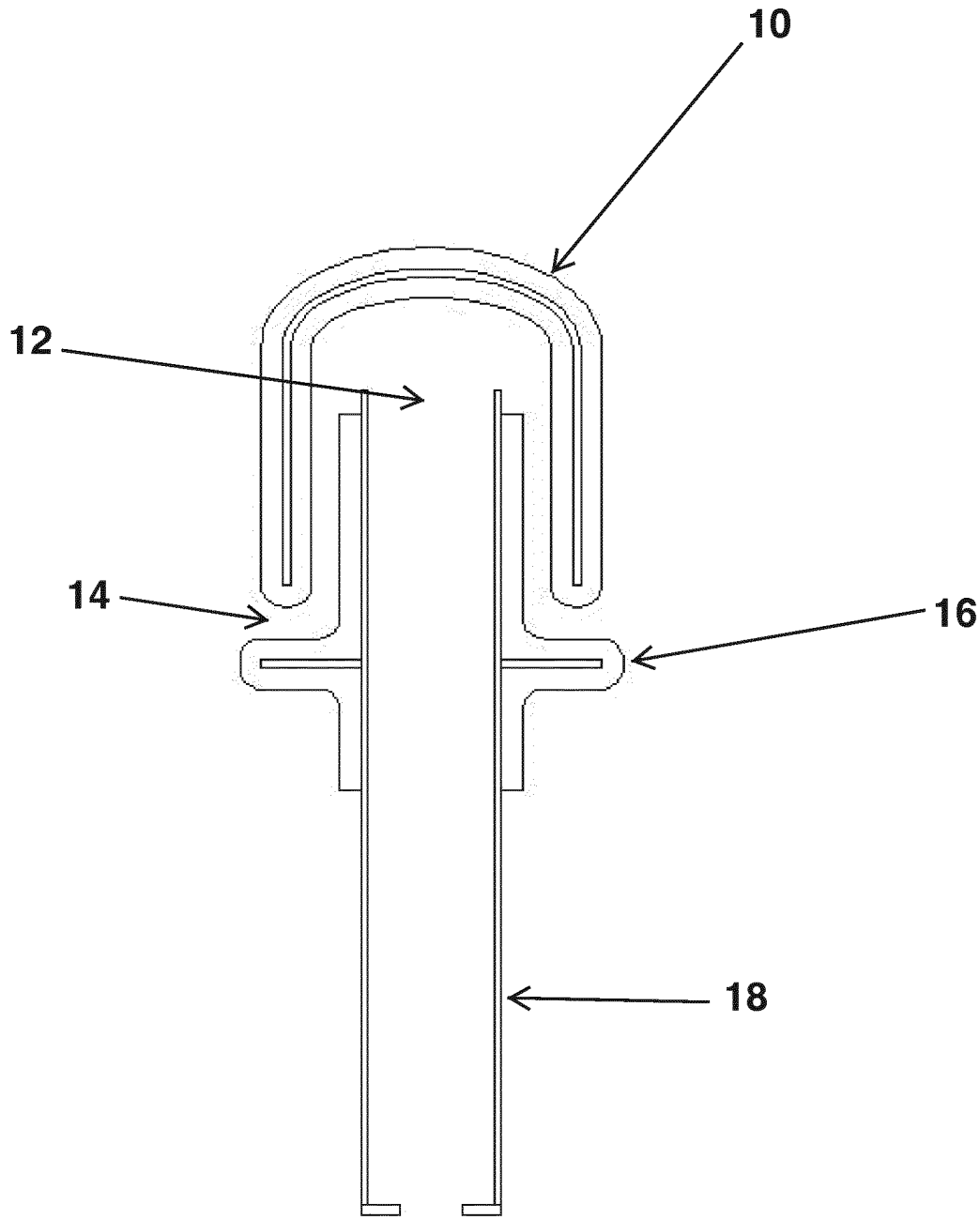


Figure 2

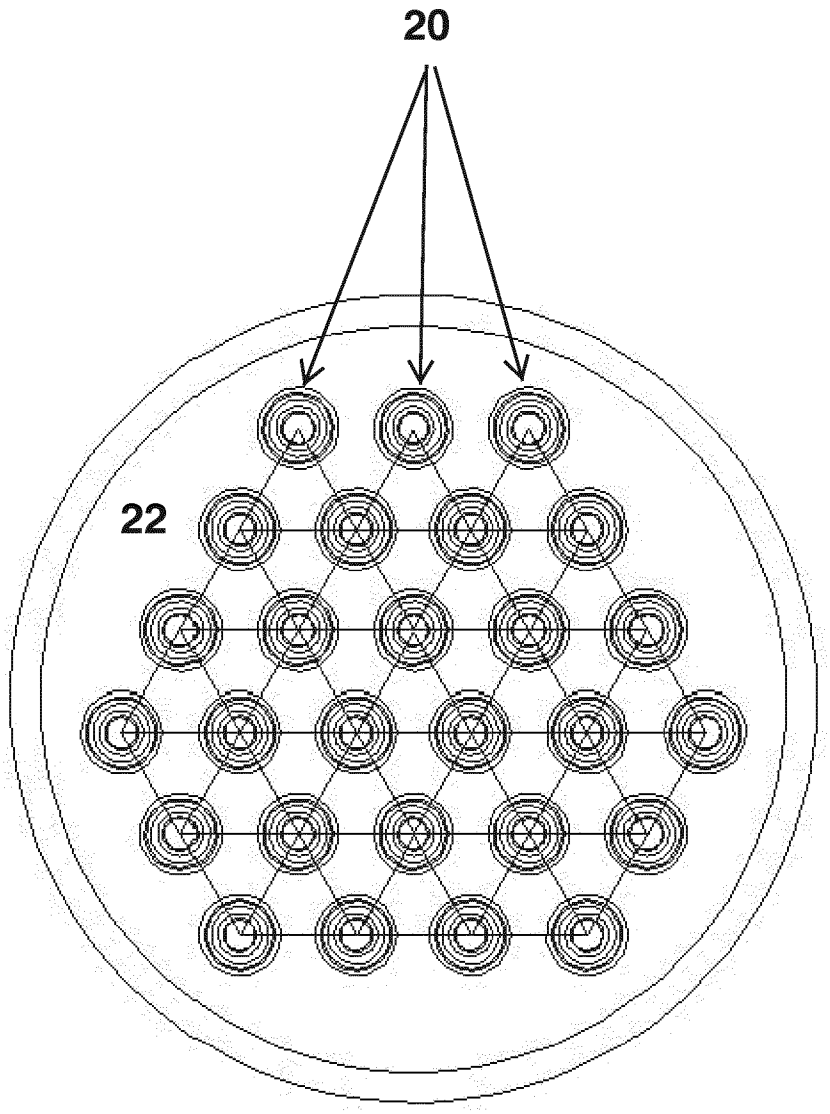
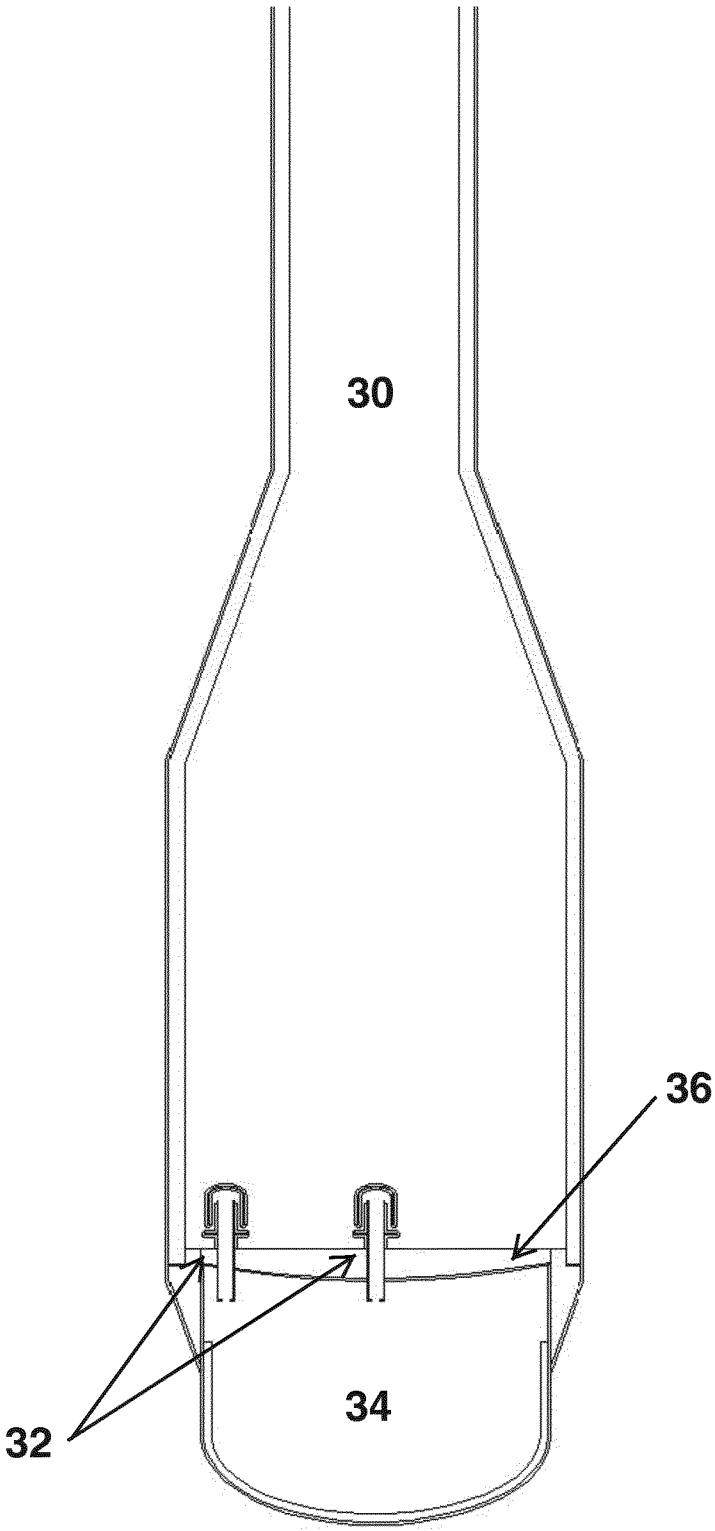


Figure 3



INTERNATIONAL SEARCH REPORT

International application No
PCT/EP2014/073424

A. CLASSIFICATION OF SUBJECT MATTER
INV. C07C1/20 B01J4/00 B01J8/44
ADD.

According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)
C07C B01J

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the international search (name of data base and, where practicable, search terms used)
EPO-Internal, WPI Data

C. DOCUMENTS CONSIDERED TO BE RELEVANT

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A	US 7 034 196 B2 (CLEM KENNETH RAY [US] ET AL) 25 April 2006 (2006-04-25) cited in the application the whole document	1-20
A	US 2010/152513 A1 (VAUGHN STEPHEN N [US] ET AL) 17 June 2010 (2010-06-17) the whole document	1-20
X	GB 1 436 123 A (TONGEREN UK LTD VAN) 19 May 1976 (1976-05-19) page 1, line 84 - page 2, line 9; figure 1	21-39

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[X] Further documents are listed in the continuation of Box C.

[X] See patent family annex.

* Special categories of cited documents :

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"O" document referring to an oral disclosure, use, exhibition or other means

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"&" document member of the same patent family

Date of the actual completion of the international search	Date of mailing of the international search report
26 January 2015	03/02/2015

Name and mailing address of the ISA/	Authorized officer
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INTERNATIONAL SEARCH REPORT

International application No
PCT/EP2014/073424

C(Continuation). DOCUMENTS CONSIDERED TO BE RELEVANT

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