

SYSTEMS AND PROCESSES FOR POLYETHYLENE PRODUCTION**CROSS-REFERENCE TO RELATED APPLICATIONS**

[0001] This application claims the benefit of U.S. Provisional Application Serial No. 63/600,889 filed November 20, 2023, the contents of which are incorporated in their entirety herein.

TECHNICAL FIELD

[0002] Embodiments of the present disclosure generally relate to systems and processes by which olefin based polymer compositions are synthesized.

BACKGROUND

[0003] Olefin based polymers, such as polyethylene, are produced with various catalyst. Selection of such catalyst used in the polymerization process of the olefin based polymers is an important factor contributing to the characteristics and properties of such olefin based polymers. Polyethylene is manufactured for a wide variety of articles. The differing catalysts allow polyethylene polymerization process to be varied in a number of respects to produce a wide variety of resultant polyethylene resins having different physical properties that render the various resins suitable for use in different applications.

SUMMARY

[0004] In the recent years, many efforts on the development of molecular catalysts allow for increased reactor temperature and increased polymer concentration, leading to increased rates of production. However, existing reactor technologies are still showing limitations. Solution polymerization is conventionally carried out either in stirred or loop reactors, which each exhibit process constraints. Specifically, stirred reactors operate close to adiabatically, resulting in high solvent usage (such as 15% polymer concentration) and therefore requiring a large solvent removal section for a given production rate. Additionally, stirred tank reactors are well mixed, resulting in minimum concentration gradient and temperature gradient (of less than 2°C) leading to narrow molecular weight distribution (MWD) and short chain branching distribution (SCBD). Loop reactors also have constraints. Commercial loop reactors can operate at conditions that cause

broad SCBD and MWD. Additionally, conventional solution polyethylene production processes still result in high volatile levels of approximately 1000 grams per ton (g/t) in the final product stream as compared to slurry and gas phase technologies.

[0005] Accordingly, a need exists for improved polyethylene production processes that decrease required solvent usage, decrease volatile levels in the product stream, and minimize broadening of MWD and SCBD. The present disclosure addresses these and other needs by using a first reactor where a portion of the heat of reaction can be effectively removed by partly evaporating the reaction solution (solvent, monomer, and comonomer), and cooling/condensing it in an external heat exchanger before feeding the reaction solution back to the reactor. The polymerization takes place in stirred tanks, which regulate the broadening of MWD and SCBD.

[0006] Embodiments of this disclosure include processes for polyethylene production that include introducing a feed stream including ethylene and a comonomer into a first reactor, and heating the feed stream in the first reactor in the presence of a first catalyst to create a preliminary product stream. The process then includes introducing the preliminary product stream into an adiabatic tubular reactor and heating the preliminary product stream in the adiabatic tubular reactor in the presence of a second catalyst to produce a polyethylene product stream.

[0007] Various embodiments of the process include polymerizations that occur in a solution.

[0008] It is to be understood that both the preceding general description and the following detailed description describe various embodiments and are intended to provide an overview or framework for understanding the nature and character of the claimed subject matter. Additional features and advantages of the embodiments will be set forth in the detailed description, and, in part, will be readily apparent to persons of ordinary skill in the art from that description, which includes the accompanying drawings and claims, or recognized by practicing the described embodiments. The drawings are included to provide a further understanding of the embodiments, and together with the detailed description, serve to explain the principles and operations of the claimed subject matter. However, the embodiments are illustrative and exemplary in nature, and not intended to limit the claimed subject matter.

BRIEF DESCRIPTION OF THE FIGURES

[0009] The following detailed description of specific embodiments of the present disclosure can be best understood when read in conjunction with the following drawing, in which:

[0010] FIG. 1 is a schematic view of a process for producing polyethylene, according to the present embodiments; and

[0011] FIG. 2 is a schematic view of a process for producing polyethylene, according to the present embodiments.

DETAILED DESCRIPTION

[0012] Embodiments of the present disclosure are directed to processes for producing polyethylene including heating a feed stream including ethylene and a comonomer in a first reactor in the presence of a first catalyst to create a preliminary product stream and heating the preliminary product stream in an adiabatic tubular reactor in the presence of a second catalyst to produce a polyethylene product stream. It should be understood that the process for producing polyethylene of this disclosure may be embodied in different forms and should not be construed as limited to the specific embodiments set forth in this disclosure. Rather, embodiments are provided so that this disclosure will be thorough and complete, and will fully convey the scope of the subject matter to those skilled in the art.

[0013] As used throughout the disclosure, “polymer” refers to a polymeric compound prepared by polymerizing monomers, whether of the same or a different type. The generic term polymer thus embraces the term “homopolymer,” usually employed to refer to polymers prepared from only one type of monomer as well as “copolymer” which refers to polymers prepared from two or more different monomers.

[0014] As used throughout the disclosure, “polyethylene” or “ethylene-based polymer” shall mean polymers comprising greater than 50% by weight of units which have been derived from ethylene monomer. This includes polyethylene homopolymers or copolymers (meaning units derived from two or more comonomers). Common forms of polyethylene known in the art include Low Density Polyethylene (LDPE); Linear Low Density Polyethylene (LLDPE); Ultra Low Density Polyethylene (ULDPE); Very Low Density Polyethylene (VLDPE); single-site catalyzed

Linear Low Density Polyethylene, including both linear and substantially linear low density resins (m-LLDPE); Medium Density Polyethylene (MDPE); and High Density Polyethylene (HDPE).

[0015] Specific embodiments will now be described with references to the figure.

[0016] FIG. 1 schematically depicts a process 100 for producing polyethylene from a feed stream 105, according to embodiments described herein.

[0017] A polyethylene production process of embodiments includes introducing a feed stream 105 into a first reactor 110. In some embodiments, the first reactor 110 may be a continuous stirred-tank reactor (CSTR). In other embodiments, the first reactor 110 may be a boiling reactor.

[0018] The feed stream 105 may include a C₂ monomer and a comonomer. The C₂ monomer may include ethylene. In embodiments, the feed stream 105 may include an ethylene concentration of greater than 25 wt.%, greater than 35 wt.%, from 25 to 50 wt.%, from 30 to 50 wt.%, from 35 to 50 wt.%, from 40 to 50 wt.%, from 45 to 50 wt.%, from 25 to 45 wt.%, from 30 to 45 wt.%, from 35 to 45 wt.%, from 40 to 45 wt.%, from 25 to 40 wt.%, from 30 to 40 wt.%, from 35 to 40 wt.%, from 25 to 35 wt.%, from 30 to 35 wt.%, or from 25 to 30 wt.%.

[0019] As stated above, the feed stream 105 includes a comonomer. In embodiments, the comonomer may include a C₄-C₈ α -olefin comonomer. For example, the α -olefin co-monomer may include, but is not limited to, propylene, 1-butene, 1-pentene, 1-hexene, 1-heptene, 1-octene, 1-nonene, 1-decene, 4-methyl-1-pentene, or combinations thereof. In embodiments, the feed stream 105 may include a comonomer concentration of from 0 to 50 wt.%, from 0 to 40 wt.%, from 0 to 30 wt.%, from 0 to 20 wt.%, from 0 to 50 wt.%, from 0 to 40 wt.%, from 0 to 30 wt.%, from 0 to 50 wt.%, from 0 to 40 wt.%, or from 0 to 50 wt.%.

[0020] In embodiments, the feed stream 105 may further include a solvent. In embodiments, the solvent includes an isoparaffinic solvent. In embodiments, the solvent includes cyclohexane, cycloheptane, cyclooctane, or combinations thereof. In embodiments, the feed stream 105 may include a solvent concentration of from 30 to 70 wt.%, from 30 to 60 wt.%, from 30 to 50 wt.%, from 30 to 40 wt.%, from 40 to 70 wt.%, from 40 to 60 wt.%, from 40 to 50 wt.%, from 50 to 70 wt.%, from 50 to 60 wt.%, or from 60 to 70 wt.%. It is contemplated that selecting the solvent and comonomer types significantly improve the volatile (VOC) level in the polyethylene product stream 122 while still maintaining the single liquid phase in the reactor.

[0021] As stated previously, the process includes introducing a feed stream 105 into a first reactor 110. In embodiments, the feed stream 105 may be entirely in liquid phase in the first reactor 110. According to embodiments, the process may include a single first reactor 110 (as shown in FIG. 1) or multiple reactors 110 (not shown) connected in series or in parallel. The first reactor(s) 110 may have dimensions defined by the equation L/D , where L is a length of the first reactor 110 and D is the diameter of the first reactor 110. In one or more embodiments, the L/D value of the first reactor 110 may be sufficient to achieve a superficial velocity of fluid less than 0.2 meter (m)/second (s), or an L/D value sufficient to achieve a superficial velocity of fluid between 1 m/min and 5 m/min. This relatively high superficial velocity of fluid is desired to attain full turbulence of the internal fluid. The desired Reynolds number (a measurement of fluid flow) is greater than 5000, in embodiments. Reynolds number is given by the relationship: $Re = \frac{uD}{\nu}$ where u is the superficial velocity, D is the diameter of the reactor, and ν is the kinematic viscosity. If that equation is rewritten as $u = \frac{\nu Re}{D}$, it can be observed that by decreasing the reactor diameter (D) the superficial velocity (u) is increased (because u and D are inversely proportional to each other ($u \propto \frac{1}{D}$)). For a fixed reactor length at a reference case, decreasing the reactor diameter (D) will increase the ratio (L/D). Furthermore, by increasing the superficial velocity (u), Reynolds Number (Re) is increased (because u and Re are directly proportional to each other ($u \propto Re$)). Therefore, from the above rationale, in order to maintain the flow in high flow turbulence regime ($Re > 5000$), it is required to increase the superficial velocity, and/or decrease the reactor's diameter, and by decreasing the reactor's diameter, the ratio (L/D) is also increased.

[0022] In embodiments, the residence time of the internal fluid in the first reactor 110 may be longer than 5 seconds, such as longer than 1 minute. In some embodiments, the residence time of the internal fluid in the first reactor 110 may be from 1 to 30 minutes, from 1 to 20 minutes, from 1 to 15 minutes, from 1 to 12 minutes, from 1 to 10 minutes, from 1 to 8 minutes, from 1 to 5 minutes, from 1 to 2 minutes, from 2 to 30 minutes, from 2 to 20 minutes, from 2 to 15 minutes, from 2 to 12 minutes, from 2 to 10 minutes, from 2 to 8 minutes, from 2 to 5 minutes, from 5 to 30 minutes, from 5 to 20 minutes, from 5 to 15 minutes, from 5 to 12 minutes, from 5 to 10 minutes, from 5 to 8 minutes, from 8 to 30 minutes, from 8 to 20 minutes, from 8 to 15 minutes, from 8 to 12 minutes, from 8 to 10 minutes, from 10 to 30 minutes, from 10 to 20 minutes, from 10 to 15 minutes, from 10 to 12 minutes, from 12 to 30 minutes, from 12 to 20 minutes, from 12

to 15 minutes, from 15 to 30 minutes, from 15 to 20 minutes, or from 20 to 30 minutes. In embodiments, the residence time may be no greater than 15 minutes and no less than 2 minutes.

[0023] After introducing the feed stream 105 into the first reactor 110, the process then includes heating the feed stream 105 in the presence of a first catalyst to create a preliminary product stream 112. The first catalyst may include a chromium-based catalyst, a Ziegler–Natta catalyst, a molecular (either metallocene or non-metallocene) catalyst, or combinations thereof. In embodiments, heating the feed stream 105 in the first reactor 110 may include heating the feed stream 105 to from 150°C to 225°C, from 150°C to 200°C, from 150°C to 175°C, from 175°C to 225°C, from 175°C to 200°C, or from 200°C to 225°C. The first reactor 110 operates at a pressure of greater than 40 bar, greater than 45 bar, greater than 50 bar, from 40 bar to 150 bar, from 40 to 125 bar, from 40 to 100 bar, from 40 to 75 bar, from 40 to 50 bar, from 40 to 45 bar, from 45 bar to 150 bar, from 45 to 125 bar, from 45 to 100 bar, from 45 to 75 bar, from 45 to 50 bar, from 50 bar to 150 bar, from 50 to 125 bar, from 50 to 100 bar, or from 50 to 75 bar.

[0024] It is contemplated that the high ethylene concentration in the first reactor 110 suppresses/reduces the viscosity of the feed stream 105, allowing for higher polymer concentration in the first while still having a viscosity that allows good mixing and evaporation. In embodiments, the feed stream 105 may have a viscosity of less than 20 Pa·s, 15 Pa·s, less than 10 Pa·s, less than 5 Pa·s, from 1 to 15 Pa·s, from 1 to 10 Pa·s, from 1 to 5 Pa·s, from 5 to 15 Pa·s, from 5 to 10 Pa·s, or from 10 to 15 Pa·s in the first reactor 110.

[0025] Additionally, it is contemplated that the high ethylene concentration in the first reactor 110 ensures high catalyst efficiency in the reaction. In embodiments with a single first reactor 110, the first reactor 110 may have an ethylene conversion of the feed stream 105 of from 60% to 90%, from 65% to 90%, from 70% to 90%, from 75% to 90%, from 80% to 90%, from 85% to 90%, from 60% to 85%, from 65% to 85%, from 70% to 85%, from 75% to 85%, from 80% to 85%, from 60% to 80%, from 65% to 80%, from 70% to 80%, from 75% to 80%, from 60% to 75%, from 65% to 75%, from 70% to 75%, from 60% to 70%, from 65% to 70%, or from 60% to 65% as a percentage of the final polyethylene content of the polyethylene product stream 122. In embodiments with dual reactors 110 in series (FIG. 2), the ethylene conversion of the feed stream 105 may be of from 40% to 90%, from 40% to 45%, from 45% to 50%, from 50% to 55%, from 55% to 60%, from 65% to 70%, from 70% to 75%, from 75% to 80%, from 80% to 85%, or from 85% to 90% as a percentage of polyethylene product stream 122. The second reactor 110b in series

may have an ethylene conversion of stream 107 of from 40% to 90%, from 40% to 45%, from 45% to 50%, from 50% to 55%, from 55% to 60%, from 65% to 70%, from 70% to 75%, from 75% to 80%, from 80% to 85%, or from 85% to 90% as a percentage of the final polyethylene content of the polyethylene product stream 122.

[0026] In embodiments, heating the feed stream 105 in the first reactor 110 may include forming a vapor stream 116. In embodiments, the vapor stream 116 may be formed by the evaporation of at least a portion of the solvent present in the feed stream 105. The process may then include introducing the vapor stream 116 to a heat exchanger 117, and cooling the vapor stream 116 in the heat exchanger 117 to form a condensed light stream 118. Any remaining vapor in the heat exchanger 117 may be recycled to the first reactor 110 by gas recycle stream 119. There may be one or more heat exchangers 117 within the system 100, which may be parallel and/or in series. It has been found that by operating the first reactor 110 at the conditions described above (where the feed stream 105 includes a relatively high ethylene concentration (*e.g.* an ethylene concentration of at least 25 g/l). The vapor stream 116 is cooled to from 25°C to 40°C, from 25°C to 35°C, from 25°C to 30°C, from 30°C to 40°C, from 30°C to 35°C, or from 35°C to 40°C to form a condensed light stream 118 by using cooling tower water (CTW), chilled water, or both. In embodiments, the vapor stream 116 may be cooled to from 30°C to 40°C with only CTW. In embodiments where the vapor stream 116 is cooled to form the condensed light stream 118, the process may further include introducing the condensed light stream 118 to the first reactor 110 to mix with the feed stream 105. This allows introducing the feed stream 105 and the condensed light stream 118 to the first reactor 110 in the liquid phase, avoiding gas bubble-to-liquid mass transfer limitations of the monomers inside the first reactor 110. In the absence of gas-to-liquid mass transfer limitations, the polymer mass concentration in the first reactor 110 may be greater than 20%, greater than 30 %, or greater than 40 % with melt index (I_2) of at least 0.5 g/min, as will be described in more detail below.

[0027] As stated previously, the process includes heating the feed stream 105 in the presence of the first catalyst to create the preliminary product stream 112. In embodiments, the preliminary product stream 112 may include an ethylene concentration of from 11 to 45 g/l, from 11 to 40 g/l, from 11 to 35 g/l, from 11 to 30 g/l, from 11 to 25 g/l, from 11 to 20 g/l, from 11 to 15 g/l, from 15 to 45 g/l, from 15 to 40 g/l, from 15 to 35 g/l, from 15 to 30 g/l, from 15 to 25 g/l, from 15 to 20 g/l, from 20 to 25 g/l, from 20 to 30 g/l, from 20 to 35 g/l, from 20 to 40 g/l, from 25 to 30 g/l,

from 25 to 35 g/l, from 25 to 40 g/l, from 25 to 45 g/l, from 30 to 35 g/l, from 30 to 40 g/l, from 30 to 45 g/l, from 35 to 40 g/l, from 35 to 45 g/l, or from 40 to 45 g/l. In embodiments, the preliminary product stream 112 comprises a polymer mass concentration of from 24 to 36 wt.%, from 24 to 33 wt.%, from 24 to 30 wt.%, from 24 to 27 wt.%, from 27 to 36 wt.%, from 27 to 33 wt.%, from 27 to 30 wt.%, from 30 to 36 wt.%, from 30 to 33 wt.%, or from 33 to 36 wt.%. In embodiments, preliminary product stream 112 may have a melt index greater than 0.3 g/min, greater than 0.5 g/min, from 0.3 to 1 g/min, from 0.5 to 1 g/min, from 0.7 to 1 g/min, from 0.3 to 0.7 g/min, from 0.3 to 0.5 g/min, from 0.5 to 1 g/min, from 0.5 to 0.7 g/min, or from 0.7 to 1 g/min in which melt index I_2 is measured according to ASTM D1238 (incorporated herein by reference in its entirety) at 190°C and 2.16 kg load.

[0028] In some embodiments, the preliminary product stream 112 may have a molecular-weight distribution (MWD) from 1 to 10, where MWD is defined as M_w/M_n with M_w being a weight average molecular weight and M_n being a number average molecular weight, where M_w and M_n are determined by a conventional gel permeation chromatography. In embodiments, preliminary product stream may have a MWD from 1 to 10, from 1 to 5, from 1 to 3, from 1.5 to 10, from 1.5 to 5, from 1.5 to 3, from 2 to 10, from 2 to 5, from 2 to 3, or approximately 2.5.

[0029] The process further includes introducing the preliminary product stream 112 into an adiabatic tubular reactor 120. In embodiments, the preliminary product stream 112 may be entirely in liquid phase in the adiabatic tubular reactor 120. Due to the relatively high ethylene concentration present in the preliminary product stream 112, the adiabatic tubular reactor 120 follows the first reactor 110 (in series) for further polymerization of the product stream 112 under adiabatic conditions. In embodiments, a second catalyst 113 is fed and mixed with the preliminary product stream 112 before the preliminary product stream 112 is introduced to the adiabatic tubular reactor 120 to boost ethylene conversion. In embodiments, the preliminary product stream 112 and the second catalyst 113 may be combined in a feed mixer 114. The feed mixer 114 can be any type of mixing device capable of mixing the preliminary product stream 112 and the second catalyst 113. In embodiments, the feed mixer 114 may be a mixing tee. The feed mixer 114 may be an ultrasonic device, a small continuous stir tank reactor (CSTR), or any suitable mixer.

[0030] Similar to the first catalyst, the second catalyst may include a chromium-based catalyst, a Ziegler-Natta catalyst, a molecular (either metallocene or non-metallocene) catalyst,

or combinations thereof. The second catalyst may be the same as the first catalyst or may be a different catalyst. The second catalyst may tune the properties of the polyethylene product stream 122 created in the adiabatic tubular reactor 120. In embodiments, hydrogen or hydrogenation catalysts may be added to the preliminary product stream 112 to further tune the molecular weight of the polyethylene product stream 122.

[0031] In embodiments, a third catalyst may be added to the adiabatic tubular reactor 120. The third catalyst may include a chromium-based catalyst, a Ziegler–Natta catalyst, a molecular (either metallocene or non-metallocene) catalyst, or combinations thereof. The third catalyst may be the same as the first catalyst or may be a different catalyst. The third catalyst may be the same as the second catalyst or may be a different catalyst. The third catalyst may tune the properties of the polyethylene product stream 122 created in the adiabatic tubular reactor 120.

[0032] Coupling the first reactor 110 with the adiabatic tubular reactor 120 allows for relatively high overall ethylene conversion, such that only a relatively small portion of the polyethylene product stream 122 needs to be recycled to the first reactor 110. Specifically, the polyethylene product stream 122, when removed from the adiabatic tubular reactor 120, may be processed to remove the polyethylene product from the solvent and the unreacted reactants, with the solvent and unreacted reactants being recycled back into the first reactor 110 (not shown). The adiabatic tubular reactor 120 improves overall polymer mass concentration (>30%) and provides high exit temperature (> 240°C) allowing the polyethylene product stream 122 exiting the adiabatic tubular reactor 120 to directly enter the phase separator 130 operating under adiabatic conditions ($\geq 180^{\circ}\text{C}$) and low pressure. This allows eliminating the post reactor heater (conventionally present downstream from the reactor in polyethylene production processes) which typically requires large energy consumption to heat up the polyethylene product stream 122 to a high temperature, such as greater than 150°C, or greater than 200°C, before the adiabatic flash and which often faces with the fouling issues when using the Ziegler Natta catalysts.

[0033] After introducing the preliminary product stream 112 into the adiabatic tubular reactor 120 the process then includes heating the preliminary product stream 112 in the adiabatic tubular reactor 120 in the presence of the second catalyst to produce a polyethylene product stream 122. Heating the preliminary product stream 112 in the adiabatic tubular reactor 120 includes heating the preliminary product stream 112 to greater than 220°C, greater than 230°C, greater than 240°C,

greater than 245°C, greater than 250°C, greater than 275°C, from 220°C to 300°C, from 220°C to 275°C, from 220°C to 250°C, from 220°C to 245°C, from 220°C to 240°C, from 220°C to 230°C, from 230°C to 300°C, from 230°C to 275°C, from 230°C to 250°C, from 230°C to 245°C, from 230°C to 240°C, from 240°C to 300°C, from 240°C to 275°C, from 240°C to 250°C, from 240°C to 245°C, from 245°C to 300°C, from 245°C to 275°C, from 245°C to 250°C, from 250°C to 300°C, from 275°C to 300°C, from 230°C to 275°C, from 250°C to 275°C, or from 230°C to 250°C. The adiabatic tubular reactor 120 may have an ethylene conversion of the preliminary product stream 112 of less than 90%.

[0034] The polyethylene product stream 122 may include a polymer mass concentration of from 26 to 44 wt.%, from 30 to 44 wt.%, from 35 to 44 wt.%, from 40 to 44 wt.%, from 26 to 40 wt.%, from 30 to 40 wt.%, from 35 to 40 wt.%, from 26 to 35 wt.%, from 30 to 35 wt.%, or from 26 to 30 wt.%. In embodiments, the polyethylene product stream 122 comprises an ethylene concentration of less than 10 g/l, less than 8 g/l, less than 5 g/l, from 1 to 10 g/l, from 1 to 8 g/l, from 1 to 5 g/l, from 1 to 3 g/l, from 3 to 10 g/l, from 3 to 8 g/l, from 3 to 5 g/l, from 5 to 10 g/l, from 5 to 8 g/l, or from 8 to 10 g/l. According to embodiments, the polyethylene product stream 122 may include greater than 30%, greater than 35%, greater than 38%, from 30% to 50%, from 30% to 45%, from 30% to 40%, from 30% to 35%, from 35% to 50%, from 35% to 45%, from 35% to 40%, from 40% to 50%, from 40% to 45%, or from 45% to 50% solids content. In embodiments, the polyethylene product stream 122 has a density of from 0.855 to 0.970 g/cc, from 0.855 to 0.950 g/cc, from 0.855 to 0.925 g/cc, from 0.855 to 0.9 g/cc, from 0.855 to 0.875 g/cc, from 0.875 to 0.970 g/cc, from 0.875 to 0.950 g/cc, from 0.875 to 0.925 g/cc, from 0.875 to 0.9 g/cc, from 0.9 to 0.970 g/cc, from 0.9 to 0.950 g/cc, from 0.9 to 0.925 g/cc, from 0.925 to 0.970 g/cc, from 0.925 to 0.950 g/cc, or from 0.950 to 0.970 g/cc.

[0035] As stated previously, it is contemplated that selecting the solvent and comonomer types significantly improve the volatile organic carbon (VOC) level in the polyethylene product stream 122 while still maintaining the single liquid phase in the reactor. Specifically, in embodiments where the comonomer includes hexane-1 and the solvent includes cyclohexane, the polyethylene product stream 122 includes less than 355 ppm, less than 350 ppm, less than 340 ppm, less than 330 ppm, less than 325 ppm, or less than 320 ppm VOC.

[0036] The method may further include introducing the polyethylene product stream 122 to a phase separator 130. The phase separator 130 may separate the polyethylene product stream 122 into a vapor product fraction 132 and a liquid product fraction 134. The phase separator 130 may be any separator known in the industry. While the phase separator 130 may separate the polyethylene product stream 122 into at least a vapor product fraction 132 and a liquid product fraction 134, it should be appreciated that additional fractions may also be produced.

[0037] Now referring to FIG. 2, in one or more embodiments, the process 200 producing polyethylene from a feed stream 105 may be substantially similar to the process 100 depicted in FIG. 1. The difference between the process 200 depicted in FIG. 2 and the process 100 depicted in FIG. 1 relates to the utilization of two reactors 110a and 110b for the conversion of feed stream 105 into polyethylene. The two reactors 110a and 110b may be in parallel or in series.

[0038] In embodiments, after introducing the feed stream 105 into a first reactor 110a, the process then includes heating the feed stream 105 in the presence of a first catalyst to create a first effluent 107. The first effluent 107 is then introduced into a second reactor 110b, and the process then may include heating the first effluent 107 in the presence of the same first catalyst or a different catalyst to create a preliminary product stream 112.

[0039] In embodiments, heating the feed stream 105 in the first reactor 110a may include forming a first vapor stream 116a. In embodiments, the first vapor stream 116a may be formed by the evaporation of at least a portion of the solvent present in the feed stream 105. The process may then include introducing the first vapor stream 116a to a first heat exchanger 117a, and cooling the first vapor stream 116a in the first heat exchanger 117a to form a first condensed light stream 118a. Any gas remaining in the first heat exchanger 117a may be recycled to the first reactor 110a by a first gas recycle stream 119a.

[0040] In embodiments, heating the first effluent 107 in the second reactor 110b may include forming a second vapor stream 116b. In embodiments, the first effluent 107 may be mixed with a second feed 106 prior to entering the second reactor 110b. In embodiments, the second vapor stream 116b may be formed by the evaporation of at least a portion of the solvent present in the first effluent 107. The process may then include introducing the second vapor stream 116b to a second heat exchanger 117b, and cooling the second vapor stream 116b in the second heat

exchanger 117b to form a second condensed light stream 118a. Any gas remaining in the second heat exchanger 117b may be recycled to the second reactor 110b by a second gas recycle stream 119b. The heat exchangers 117a and 117b within the system 200 may be parallel and/or in series. There may be more than one heat exchanger in the system 200, such as 2, 3, 4, or more heat exchangers.

EXAMPLES

[0041] A polyethylene production process in accordance with the embodiments described above was calculated using the operating conditions, components, and concentrations outlined below in Table 1.

Table 1

<i>Example</i>			A	B	C	D
PE Properties	Melt Index (I ₂)	dg/min	1.0	2.3	0.5	1.0
	Density	g/cm ³	0.9200	0.9170	0.9175	0.9160
CSTR	Polymer Concentration	wt. %	31.8%	35.9%	31.5%	32.3%
	Ethylene Concentration (in liquid phase)	g/l	35.9	40.3	39.1	37.8
	Catalyst type	-	Ziegler Natta	Ziegler Natta	Ziegler Natta	Molecular
	Temperature	°C	190.0	180.0	180.0	190.0
	Pressure (absolute)	MPa	6.6	6.6	6.6	6.7
	Polymer split	%	85.5	84.9	83.2	86.2
	Comonomer type	-	1-hexene	1-hexene	1-hexene	1-hexene
	Solvent type	-	cyclohexane	cyclohexane	cyclohexane	cyclohexane
	<i>CSTR Inlet composition (mass fraction)</i>					
	Hydrogen	wt. %	0.003 %	0.003 %	0.003 %	0.003 %
	Ethylene	wt. %	36.1%	39.8%	35.5%	35.5%
	Comonomer	wt. %	18.9%	26.4%	25.6%	10.7%

<i>Example</i>			A	B	C	D
	Solvent	wt. %	44.3%	33.2%	38.3%	53.3%
	Inerts	wt. %	0.6%	0.6%	0.7%	0.5%
Tubular reactor	Polymer concentration	wt. %	38.18 %	42.40 %	37.87 %	37.51 %
	ethylene concentration (in liquid phase)	g/L	7.9	7.8	7.6	6.1
	Catalyst type	-	Ziegler Natta	Ziegler Natta	Ziegler Natta	Molecular
	Temperature	°C	249.0	249.0	247.0	251.0
	Pressure drop	MPa	0.3	0.3	0.3	0.3
	Comonomer type		1-hexene	1-hexene	1-hexene	1-hexene
	Solvent type		cyclohexane	cyclohexane	cyclohexane	cyclohexane
Condenser	Temperature	°C	40.0	40.0	40.0	40.0
	Pressure	MPa	6.4	6.4	61.8	6.4
Recycle	liquid/ gas mass flow ratio	-	63.5	78.0	35.4	35.7
Polymerization energy removal	Total recycle/ Production rate ratio	-	6.2	4.6	4.2	3.8

[0042] A polyethylene production process in accordance with another embodiment described above was calculated using the operating conditions, components, and concentrations outlined below in Table 2.

Table 2

<i>Example</i>			F		G		H	
Unit			<i>CSTR 1</i>	<i>CSTR 2</i>	<i>CSTR 1</i>	<i>CSTR 2</i>	<i>CSTR 1</i>	<i>CSTR 2</i>
PE Properties	Melt Index (I ₂)	dg/min	0.15	0.85	4.2	3.7	15	21.5
	Density	g/cm ³	0.8990	0.9200	0.9070	0.9190	0.8980	0.9070
CSTR	Polymer Concentration	wt. %	18.5%	32.8%	28.4%	35.3%	27.3%	36.8%
	Ethylene Concentration	g/l	18.9	38.7	23.0	38.5	36.3	36.4

<i>Example</i>			F		G		H	
Unit			<i>CSTR 1</i>	<i>CSTR 2</i>	<i>CSTR 1</i>	<i>CSTR 2</i>	<i>CSTR 1</i>	<i>CSTR 2</i>
	n (in liquid phase)							
	Catalyst type	-	Molec- ular	Ziegler Natta	Molec- ular	Ziegler Natta	Molec- ular	Ziegler Natta
	Temperature	°C	180	190	180	190	180	190
	Pressure (absolute)	MPa	3.8	7.0	4.4	7.1	6.0	7.1
	Polymer split	%	37.5%	47.8%	14.7%	60.0%	25.4%	56.0%
	Comonomer type	-	1- hexene	1- hexene	1- hexene	1- hexene	1- hexene	1- hexene
	Solvent type	-	cycloh -exane	cycloh -exane	cycloh -exane	cycloh -exane	cycloh -exane	cycloh -exane
	<i>CSTR Inlet composition (mass fraction)</i>							
	Hydrogen	wt.%	0.007 %	0.005 %	0.004 %	0.004 %	0.004 %	0.004 %
	Ethylene	wt.%	18.2%	24.1%	27.6%	16.1%	27.6%	17.7%
	Comonomer	wt.%	14.9%	9.0%	16.4%	10.3%	34.9%	25.4%
	Solvent	wt.%	66.2%	51.9%	55.4%	48.3%	36.8%	32.4%
	Polyethylene	wt.%	--	14.5%	--	24.8%	--	24.0%
	Inerts	wt.%	0.6%	0.5%	0.6%	0.5%	0.6%	0.5%
Tubular reactor	Pressure	Mpa	6.83		6.93		6.86	
	Temperature	°C	254		254		252	
	Catalyst type	-	Ziegler Natta		Ziegler Natta		Ziegler Natta	
	ethylene concentration (in liquid phase)	g/l	7.49		7.58		7.16	
	Polymer concentration (exit tube)	wt.%	33.30%		41.30%		42.90%	
Condenser	Temperature	°C	40	40	40	40	40	40
	Pressure	MPa	3.6	6.8	4.2	6.9	5.8	6.9
Recycle	Liquid/gas Mass Ratio	-	3.1	6.0	3.6	Liquid only	17.9	Liquid only
	Total Recycle / Production rate Ratio	-	2.29	3.12	3.62	1.39	3.15	1.71

[0043] It will be apparent to persons of ordinary skill in the art that various modifications and variations can be made without departing from the scope disclosed herein. Since modifications, combinations, sub-combinations, and variations of the disclosed embodiments, which incorporate the spirit and substance disclosed herein, may occur to persons of ordinary skill in the art, the scope disclosed herein should be construed to include everything within the scope of the appended claims and their equivalents.

[0044] For the purposes of defining the present technology, the transitional phrase “consisting of” may be introduced in the claims as a closed preamble term limiting the scope of the claims to the recited components or steps and any naturally occurring impurities. For the purposes of defining the present technology, the transitional phrase “consisting essentially of” may be introduced in the claims to limit the scope of one or more claims to the recited elements, components, materials, or method steps as well as any non-recited elements, components, materials, or method steps that do not materially affect the novel characteristics of the claimed subject matter. The transitional phrases “consisting of” and “consisting essentially of” may be interpreted to be subsets of the open-ended transitional phrases, such as “comprising” and “including,” such that any use of an open ended phrase to introduce a recitation of a series of elements, components, materials, or steps should be interpreted to also disclose recitation of the series of elements, components, materials, or steps using the closed terms “consisting of” and “consisting essentially of.” For example, the recitation of a composition “comprising” components A, B, and C should be interpreted as also disclosing a composition “consisting of” components A, B, and C as well as a composition “consisting essentially of” components A, B, and C. Any quantitative value expressed in the present application may be considered to include open-ended embodiments consistent with the transitional phrases “comprising” or “including” as well as closed or partially closed embodiments consistent with the transitional phrases “consisting of” and “consisting essentially of.”

[0045] As used in the Specification and appended Claims, the singular forms “a”, “an”, and “the” include plural references unless the context clearly indicates otherwise. The verb “comprises” and its conjugated forms should be interpreted as referring to elements,

components or steps in a non-exclusive manner. The referenced elements, components or steps may be present, utilized or combined with other elements, components or steps not expressly referenced.

[0046] It should be understood that any two quantitative values assigned to a property may constitute a range of that property, and all combinations of ranges formed from all stated quantitative values of a given property are contemplated in this disclosure. The subject matter disclosed herein has been described in detail and by reference to specific embodiments. It should be understood that any detailed description of a component or feature of an embodiment does not necessarily imply that the component or feature is essential to the particular embodiment or to any other embodiment. Further, it should be apparent to those skilled in the art that various modifications and variations can be made to the described embodiments without departing from the spirit and scope of the claimed subject matter.

CLAIMS

1. A polyethylene production process comprising:
 - introducing a feed stream into a first reactor, wherein the feed stream comprises ethylene, hydrocarbon solvent and a comonomer;
 - heating the feed stream in the first reactor in the presence of a first catalyst to create a preliminary product stream;
 - introducing the preliminary product stream into an adiabatic tubular reactor; and
 - heating the preliminary product stream in the adiabatic tubular reactor in the presence of a second catalyst to produce a polyethylene product stream.
2. The polyethylene production process of claim 1, wherein the first reactor is selected from a continuous stir tank reactor or a boiling reactor.
3. The polyethylene production process of claim 1, wherein heating the feed stream in the first reactor comprises heating the feed stream to from 150 °C to 225 °C.
4. The polyethylene production process of claim 3, wherein the first reactor operates at a pressure of from 40 bar to 150 bar.
5. The polyethylene production process of any previous claim, wherein heating the preliminary product stream in the adiabatic tubular reactor comprises heating the preliminary product stream to greater than 230°C to 300°C.
6. The polyethylene production process of any previous claim:
 - wherein the polyethylene product stream comprises polyethylene product, solvent, and unreacted reactants; and
 - further comprising processing the polyethylene product stream to remove polyethylene product from the solvent and unreacted reactants and recycling the solvent and unreacted reactants back into the first reactor.
7. The polyethylene production process of any previous claim:

wherein heating the feed stream in the first reactor comprises forming a vapor stream;
and
further comprising:
introducing the vapor stream to a heat exchanger,
cooling the vapor stream in the heat exchanger to form a condensed light stream;
and
introducing the condensed light stream to the first reactor.

8. The polyethylene production process of any previous claim, wherein the comonomer comprises an alpha olefin having from 4 to 8 carbon atoms.
9. The polyethylene production process of any previous claim, wherein the feed stream further comprises an isoparaffinic solvent.
10. The polyethylene production process of any previous claim, wherein the feed stream further comprises a solvent comprising cyclohexane, cycloheptane, cyclooctane, or combinations thereof. .
11. The polyethylene production process of claim 10, wherein the comonomer comprises 1-hexene, 1-heptene, 1-octene, or combinations thereof.
12. The polyethylene production process of any previous claim, wherein the feed stream comprises an ethylene concentration of from 25 to 50 wt.%.
13. The polyethylene production process of any previous claim, wherein the preliminary product stream comprises a polymer mass concentration of from 24 to 36 wt.%.
14. The polyethylene production process of any previous claim, wherein the preliminary product stream comprises an ethylene concentration of from 11 to 50 g/l.
15. The polyethylene production process of any previous claim, wherein the polyethylene product stream comprises a polymer mass concentration of from 26 to 44 wt.%.