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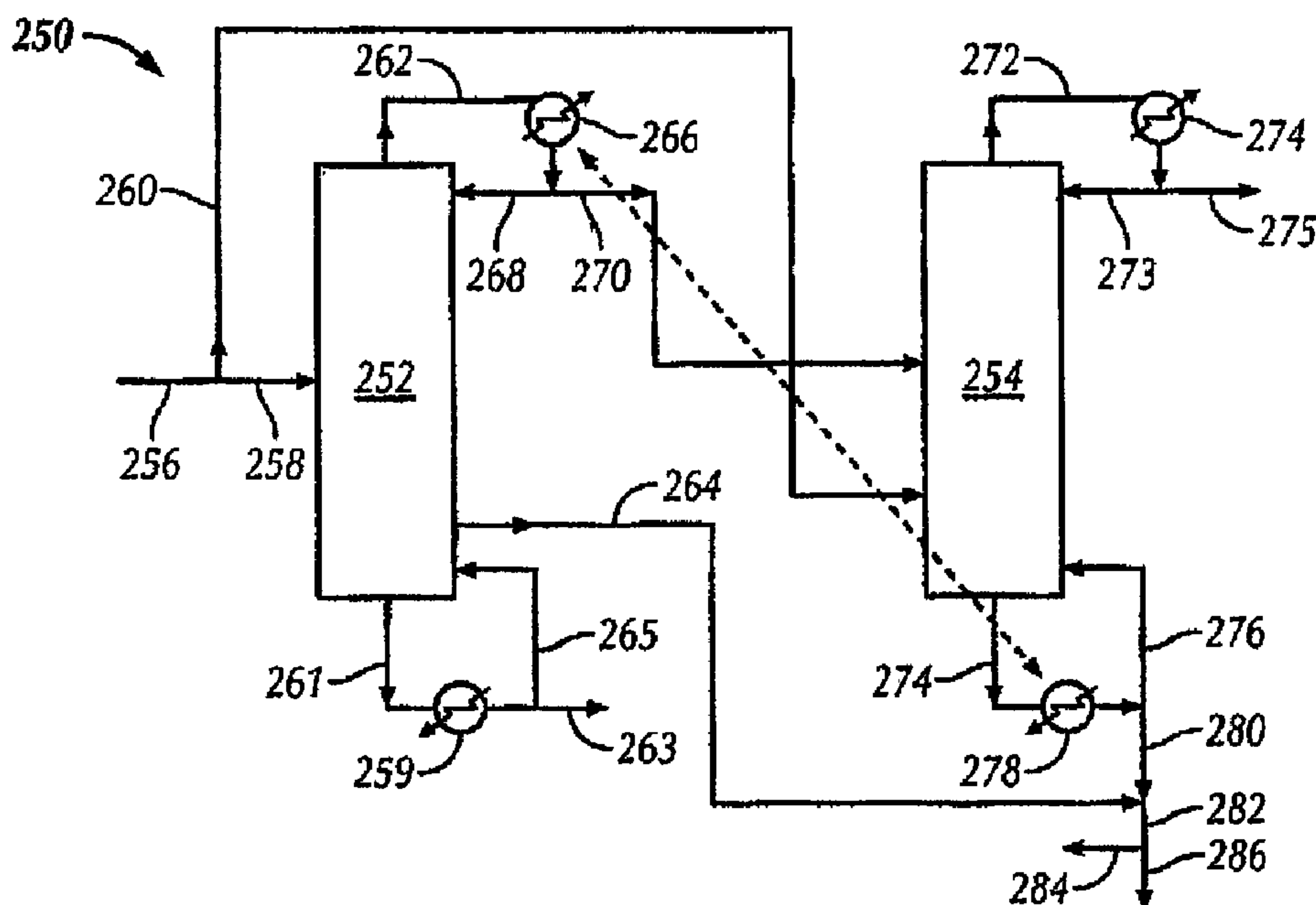
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(54) Title: SYSTEM AND PROCESS FOR PRODUCING LINEAR ALPHA OLEFINS



(57) Abrégé/Abstract:

Processes and systems for producing linear alpha olefins are described herein. One embodiment is a process comprising: a) separating a mixed butene stream comprising 1-butene and 2-butene into an overhead 1-butene stream and a bottoms 2-butene stream in a butene distillation column, a portion of the bottoms 2-butene stream being separated to form a butene reboiler stream that is heated and vaporized in a reboiler and returned to the butene distillation column, (b) subjecting at least a portion of the overhead 1-butene stream from (a) to catalytic metathesis to produce an effluent including 3-hexene, (c) isomerizing 3-hexene from (b) to produce a mixed hexene stream comprising 1-hexene, 2-hexene and 3-hexene, (d) separating the mixed hexene stream in a hexene fractionation tower to form a 1-hexene vapor overhead stream that is condensed in a cooler and a bottoms stream comprising 2-hexene and 3-hexene, and (e) using heat obtained by condensing the 1-hexene vapor overhead stream of (d) to heat the butene reboiler stream of (a). Another embodiment is a superfractionation process. Corresponding systems are also disclosed.



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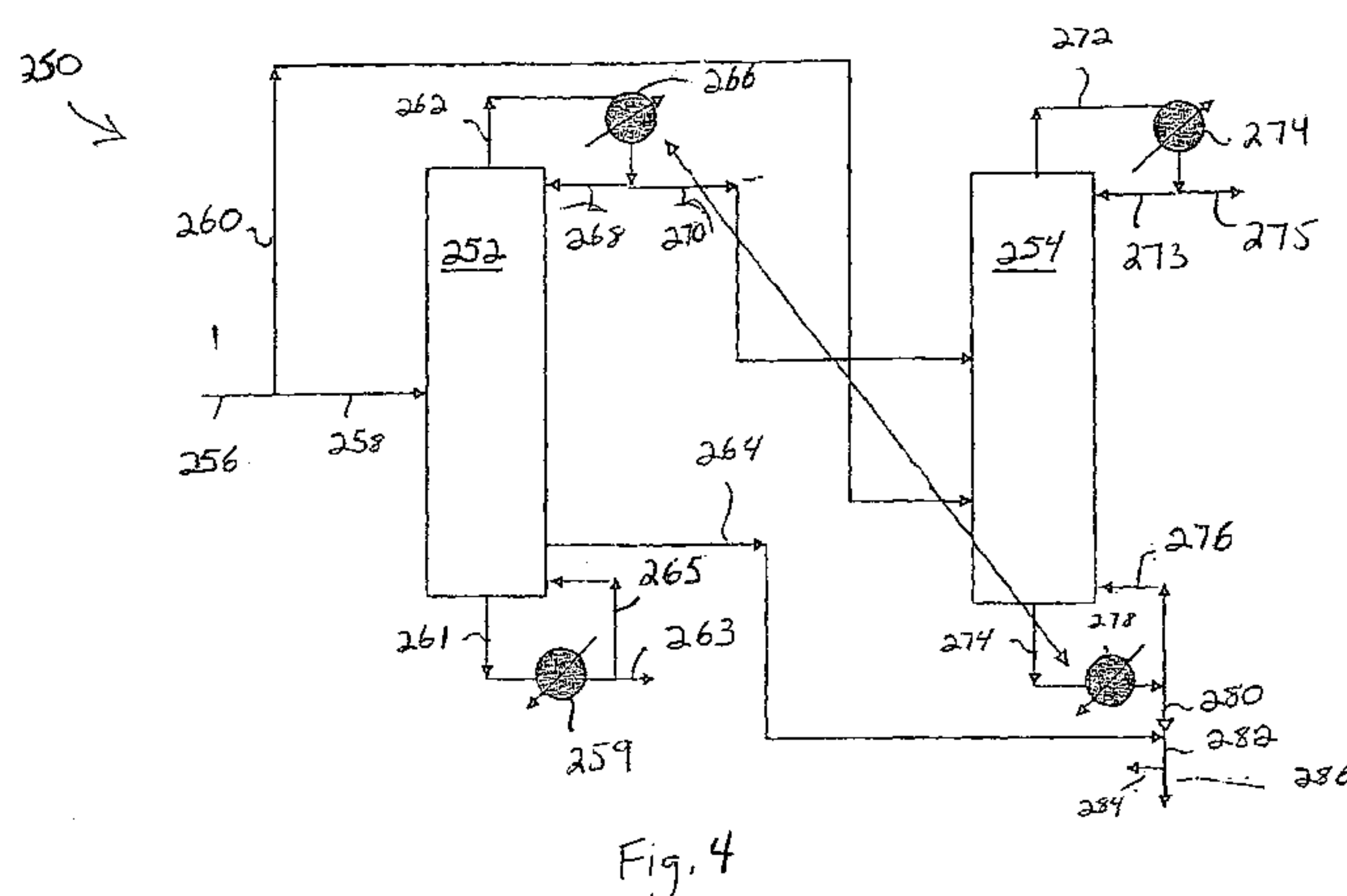


Fig. 4

(57) Abstract: Processes and systems for producing linear alpha olefins are described herein. One embodiment is a process comprising: a) separating a mixed butene stream comprising 1 -butene and 2-butene into an overhead 1 -butene stream and a bottoms 2-butene stream in a butene distillation column, a portion of the bottoms 2-butene stream being separated to form a butene reboiler stream that is heated and vaporized in a reboiler and returned to the butene distillation column, (b) subjecting at least a portion of the overhead 1 -butene stream from (a) to catalytic metathesis to produce an effluent including 3-hexene, (c) isomerizing 3-hexene from (b) to produce a mixed hexene stream comprising 1 -hexene, 2-hexene and 3-hexene, (d) separating the mixed hexene stream in a hexene fractionation tower to form a 1 -hexene vapor overhead stream that is condensed in a cooler and a bottoms stream comprising 2-hexene and 3-hexene, and (e) using heat obtained by condensing the 1 -hexene vapor overhead stream of (d) to heat the butene reboiler stream of (a). Another embodiment is a superfractionation process. Corresponding systems are also disclosed.

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SYSTEM AND PROCESS FOR PRODUCING LINEAR ALPHA OLEFINS

Related Applications

This application claims priority from U.S. Provisional Application 61/072,993 filed April 4, 2008.

Background

The embodiments described herein relate to the production of linear alpha olefins, and more particularly to linear alpha olefin production processes having decreased energy consumption.

Linear alpha olefins can be produced by the selective oligomerization of ethylene. In general, the oligomerization is conducted in the presence of a catalyst, such as an alkylated metal catalyst. Long residence times are used to produce hydrocarbon chains of varying lengths. The olefin double bond remains at the alpha position as each ethylene molecule is added. Ethylene oligomerization produces a wide spectrum of linear alpha olefin products. Extensive fractionation is required to separate the alpha olefins having different carbon numbers.

U.S. Patent No. 3,776,974 is directed to a process for producing detergent range linear monoolefins by disproportionating linear alpha olefins with 2-butene to produce linear beta olefins of both lower and higher carbon numbers. The resulting lower beta olefins fraction is subjected to disproportionation to produce 2-butene and linear internal monoolefins.

U.S. Patent No. 6,727,396, assigned to Lummus Technology Inc., describes a process for producing linear alpha olefins in which a linear alpha olefin having a first number of carbon atoms is subjected to a metathesis reaction under conditions and in the presence of a catalyst that minimizes or eliminates skeletal and/or double bond isomerization. The reaction product includes ethylene and a linear internal olefin (LIO) that has a number of carbon atoms greater than the first number of carbon atoms. Specifically, the feed linear alpha olefin with carbon number n is metathesized with a second feed linear alpha olefin to form ethylene and a linear internal olefin with carbon number $2n-2$. The resulting linear internal olefin may then be isomerized to produce a linear alpha olefin. In one embodiment described therein, 1-butene is subjected to an autometathesis reaction under conditions and in the presence of a metathesis catalyst that minimizes or eliminates skeletal and double bond isomerization to produce a reaction product that includes ethylene and 3-hexene. The 3-hexene is then isomerized to 1-hexene. In the autometathesis reaction, the catalyst and reaction conditions are such as to minimize isomerization of the 1-butene starting material. The catalyst used for the autometathesis reaction may be a supported or unsupported catalyst and the overall catalyst is one that has a minimized amount of both acidic and basic sites. Typical isomerization catalysts include basic metal oxides or promoted zeolites.

Superfractionation is known for use in separating two components having similar boiling points. In general, superfractionation involves a significant number of theoretical plates and high reflux ratios in a distillation column and thus large amounts of energy input are required to provide the reboiling and condensing duties.

It would be useful to provide systems and methods for reducing energy consumption in the production of linear alpha olefins involving superfractionation.

Summary

One embodiment is a process comprising (a) separating a mixed butene stream comprising 1-butene and 2-butene into an overhead 1-butene stream and a bottoms 2-butene stream in a butene distillation column, a portion of the bottoms 2-butene stream being separated to form a butene reboiler stream that is heated and vaporized in a reboiler and returned to the butene distillation column, (b) subjecting at least a portion of the overhead 1-butene stream from (a) to catalytic metathesis to produce an effluent including 3-hexene, (c) isomerizing 3-hexene from (b) to produce a mixed hexene stream comprising 1-hexene, 2-hexene and 3-hexene, (d) separating the mixed hexene stream in a hexene fractionation tower to form a 1-hexene vapor overhead stream that is condensed in a cooler and a bottoms stream comprising 2-hexene and 3-hexene, and (e) using heat obtained by condensing the 1-hexene vapor overhead stream of (d) to heat the butene reboiler stream of (a).

Another embodiment is a system for producing 1-hexene from butenes, comprising a butene distillation column, a metathesis reactor, a metathesis recovery zone, a hexene isomerization reactor, a hexene fractionation tower, and a condenser. The butene distillation column is configured to separate 1-butene from 2-butene, and has a butene reboiler associated therewith including a first heat source. The metathesis reactor converts 1-butene to 3-hexene and ethylene. The metathesis recovery zone separates 3-hexene from ethylene. The hexene isomerization reactor is configured to isomerize 3-hexene to 1-hexene and 2-hexene, and the hexene distillation column is configured to separate 1-hexene vapor from 2-hexene and 3-hexene. The condenser condenses the 1-hexene vapor. The first heat source for the butene reboiler is at least a portion of the heat of condensation of the 1-hexene vapor.

A further embodiment is a separation process comprising obtaining a hydrocarbon mixture comprising a first hydrocarbon, a second hydrocarbon, and higher boiling hydrocarbons, the second hydrocarbon having a boiling point at atmospheric pressure that is higher than the boiling point of the first hydrocarbon at atmospheric pressure by 0.3 to 10° C. The hydrocarbon mixture is divided into a first feed stream and a second feed stream. The first feed stream is fed to a first intermediate stage of a first distillation column and the second feed stream is fed to a second intermediate stage of a second distillation column. The first distillation column operates at a higher overhead pressure than the second distillation column. A first distillate containing most of the first hydrocarbon is removed from the top of the first distillation column, at least

partially condensed, and fed to the second distillation column at a third intermediate stage above the second intermediate stage. Higher boiling hydrocarbons are removed from the bottom of the first distillation column. The first hydrocarbon is removed from the top of the second distillation column and the second hydrocarbon is removed from the bottom of the second distillation column. At least a portion of the second hydrocarbon that is removed from the bottom of the second distillation column is heated using the first distillate and is returned to the second distillation column.

Another embodiment is a superfractionation system comprising a first distillation zone and a second distillation zone. The first distillation zone includes a first distillation column, a first condenser with a first coolant source, and a first reboiler with a first heat source. The first distillation column has a first feed inlet configured to receive a first portion of fresh feed, a first distillate outlet and a first bottoms outlet. The second distillation zone includes a second distillation column configured to operate at a lower pressure than the first distillation column, a second condenser, and a second reboiler with a second heat source. The second distillation column has a second feed inlet configured to receive a second portion of fresh feed, a third feed inlet disposed above the second feed inlet and configured to receive at least a portion of the distillate from the first distillation zone, a second distillate outlet, and a second bottoms outlet. The second heat source for the second reboiler includes the first coolant source for the first condenser.

Brief Description of the Drawings

Fig. 1 is a schematic drawing of a portion of a system for producing 1-hexene according to one embodiment.

Fig. 2 is a schematic drawing of a conventional dual pressure superfractionation system.

Fig. 3 is a graph showing energy consumption for certain disclosed embodiments as compared to conventional systems.

Fig. 4 is a schematic drawing of a dual pressure superfractionation system of an embodiment described herein.

Fig. 5 shows a variation of the system of Fig. 4 in which an open loop heat pump is employed.

Fig. 6 schematically shows a conventional fractionation-isomerization system for obtaining a linear alpha olefin.

Fig. 7 schematically shows a new fractionation-isomerization system for obtaining a linear alpha olefin.

Fig. 8 is a process flow diagram showing an overall process for producing 1-hexene according to certain embodiments.

Detailed Description

The embodiments described herein provide processes for linear alpha olefin production that result in decreased energy use in production. In one embodiment, the overhead condenser from the hexene separation tower is used as the reboiler for the butene separation tower. This provides for a reduced overall heating requirement to the system. In another embodiment, which can be used separately or in conjunction with the first embodiment, a two tower fractionation system is used for either or both the butene and hexene purification towers. The feed is divided and fed to two separate but integrated towers operating at two different pressures, and heat exchange takes place in a manner that results in lower cost production.

As used herein, the term "superfractionation" refers to the separation by distillation of two or more components having difference in boiling points between 0.3 and 10 Deg. C. "Catalytic metathesis" as used herein refers to a disproportionation reaction between two olefins to produce two different olefins.

In the production of hexenes from a metathesis reaction involving butene autometathesis, there are three general process stages. First, 1-butene is produced from a C4 stream such as a raffinate 2 stream. Next, the 1-butene undergoes self disproportionation or autometathesis to form ethylene and 3-hexene. Finally, 3-hexene is isomerized to 1-hexene, which is then purified. Each of these process stages is described below in further detail.

In the production of 1-butene, raffinate 2 is fed to a combination distillation/ butene isomerization system. Raffinate 2 is a C4 stream comprising butanes, butene-1 and butene-2 with significantly reduced quantities of both isobutylene and butadiene. The product of the distillation typically contains an overhead stream containing 1-butene of at least 90 wt % purity. Due to the low relative volatility between the isomers of normal butene, a very large number of distillation trays with a very high reflux ratio are required to achieve the separation. Given these requirements, the distillation tower requires significant energy for reboiling and condensing duty.

In the self-disproportionation or autometathesis of 1-butene to form ethylene and 3-hexene, 1-butene is reacted over metathesis catalyst and the products of that reaction are separated in a fractionation system. There are a number of fractionation options in terms of fractionator sequence since the products include carbon number products from C2 through C7. The requirements of the fractionation system are (1) separation of the C2/C3 products for recovery as valuable olefins, (2) the recycle of the C4/C5 olefins to achieve high process yields, and (3) the separation of the predominately 3-hexene stream as the feed to the final process stage. Non-limiting examples of suitable autometathesis catalysts are Group VI B or Group VII B metal oxides such as tungsten oxide, molybdenum oxide, or rhenium oxide.

In the final step of 1-hexene production, i.e. the isomerization of 3-hexene to 1-hexene and the purification of the 1-hexene, purification take place in a second "superfractionator". This distillation is even more difficult than that for the 1-butene fractionation. A large number of trays

and very high reflux ratios are required, resulting in considerable energy consumption. Together the energy required for the butene separation and the hexene separation represent a considerable energy consumption for the overall process. Several processes and systems described herein reduce energy use in the production of linear alpha olefins such as 1-hexene.

One process for reducing energy consumption in the production of 1-hexene from butenes is to use the hexene-splitter top tray vapor to heat the butene-splitter bottom tray liquid. This eliminates both the heating energy consumption of the n-butene separation system and the overhead cooling requirement for the hexene tower separation system.

Both the butene-splitter and the hexene-splitter require high concentrations of alpha-olefin in the overhead condenser liquid as the product. In many cases, the butene splitter requires greater than 90% 1-butene in the product stream, and sometimes greater than 95%. The hexene splitter often requires greater than 98% 1-hexene in the product stream. The relative volatilities of the internal olefins (2-butene, 2-hexene and 3-hexene) and the alpha olefins (1-butene and 1-hexene) in the butene and hexene towers are very low such that fractionation requires a significant number of stages and reflux in both towers. Relative volatility is the ratio of the vapor pressure of the individual components at identical temperatures and represents the relative tendency to separate by fractionation. The higher the relative volatility, the easier the separation.

In a conventional system, the butene splitter is designed to operate at the lowest possible pressure in order to increase the relative volatility between 1-butene and 2-butene and thus enhance the required separation. Increasing the relative volatility reduces the amount of reflux required and the level of heating/cooling duties. However, the temperature necessary to cool the condenser liquid restricts the butene tower pressure. Very low butene tower pressures result in overhead temperatures that are outside the cooling water temperature range and thus require refrigeration and additional process energy to provide that refrigeration. For this application, the butene tower(s) typically would be designed for approximately 70 psig.

The hexene splitter in a conventional system is designed in the same manner as the butene splitter(s) for the same reasons described above, i.e. use of a low tower pressure enhances the separation by increasing the relative volatility. As hexenes boil at a higher temperature than butenes, the hexene towers can run at atmospheric pressure or below and still use cooling water in the condenser. This tower typically is run at pressures ranging from -10 to 5 psig.

In one of the embodiments described herein, the hexene splitter is operated at a higher pressure in order to enable the top vapor stream to be used to heat the butene splitter reboiler stream. Operating the hexene splitter(s) at a higher pressure results in more theoretical stages at a constant reflux ratio, as the relative volatilities are reduced as the absolute pressure increases. As indicated by the computer simulation described below in Table 1, increasing the overhead

pressure of the hexene splitter from 0 psig to 15 psig results in an increase from 103 theoretical separation stages to 125 stages. This constitutes an increase of 21%. The higher pressure tower would thus have a higher capital cost due to the increased number of stages.

Table 1: Hexene Tower Pressure Comparison (at the same reflux ratio)

1-HEXENE/2-, 3-HEXENE SPLITTER

	LOW TOWER PRESSURE	HIGH TOWER PRESSURE
OVERHEAD PRESSURE (PSIG)	0	15
REFLUX RATIO	105	105
RELATIVE VOLATILITY (1-HEX/CIS 2-HEX)	1.220	1.193
RELATIVE VOLATILITY (1-HEX/TR 2-HEX)	0.872	0.874
RELATIVE VOLATILITY (1-HEX/CIS 3-HEX)	1.124	1.109
RELATIVE VOLATILITY (1-HEX/TR 3-HEX)	1.144	1.124
STAGES	103	125

* SAME TOWER OVERHEAD & BOTTOM COMPONENT SPECS.
USED IN BOTH CASES

Alternatively, if the identical number of stages were used at an overhead pressure of 15 psig, the higher pressure tower would require a higher reflux ratio to achieve the same degree of separation. This would increase both the tower diameter, resulting in an increase in capital cost, and also energy usage resulting from increased reboiling and condensing duty.

In the process in which the hexene tower is operated to provide reboiler duty for the butene tower, the butene tower pressure should preferably be greater than 70 psig, or in the range of about 70 psig to about 100 psig, in order to avoid the use of refrigeration in the condenser. This results in a reboiler temperature in the C4 tower of approximately 139° F. By operating the hexene tower at elevated pressures, the overhead condenser (coldest temperature in the hexene column) can be raised to a point where it is possible to use the hexene tower condenser as the reboiler for the C4 column. Because this increases the fractionation requirements (stages, reflux or both) for the hexene tower, it is unexpected that raising the pressure of the hexene tower to allow for the use of the condenser to provide the reboiling duty of the butene tower will result in a net energy savings.

To heat exchange the 1-butene/2-butene splitter bottom tray liquid with the 1-hexene / 2-hexene and 3-hexene top tray vapor, at a minimum, the hexene tower pressure must be increased to the "high-pressure" range described in Table 1 (15 psig). Operation at a lower pressure in the hexene tower reduces the heat-exchanger minimum approach as shown in Table 2 below:

Table 2: Stream Temperatures & Pressures for Integrated Process

	TEMP. (F)	PRESS. (PSIG)
BUTENE TOWER BOTTOM TRAY LIQUID	139	75
HEXENE TOWER OVERHEAD TRAY VAPOR	143	0
	187	15

Operating the hexene tower at atmospheric pressure would result in a top tray vapor temperature that could not effectively heat the butene bottom tray liquid although that would be optimal for hexene fractionation. Use of the high hexene tower pressure is counter-intuitive as it results in more trays to achieve the required separation. However, by using the high hexene tower pressure, the hexene vapor can be used to heater the butene reboiler stream. This saves energy for the overall system.

Figure 1 shows a system having the above-described heat integration. The overall system is designated as 100. The system includes a 1-butene/2-butene splitter 102 with a reboiler 104, and a 1-hexene/2- and 3- hexene splitter 106 with a reflux drum 108. The 1-hexene vapor stream 110 is removed from the top of hexene splitter 106 and is used as the heating stream in reboiler 104, resulting in a reduction in temperature of stream 110 through the reboiler 104 and usually some degree of condensation of stream 110. Stream 112 is removed from the bottom of butene splitter 102 and is divided into reboiling stream 114 and a butene bottoms stream 116. Stream 114 is heated in the reboiler 104 and is then returned to the lower end of the butene splitter 102.

After providing heat in the heat exchanger 104, if necessary, the 1-hexene vapor stream 110 is further cooled in a heat exchanger 118 using a cooling water stream 120 until it is fully condensed. The cooled 1-hexene stream 110 is sent to a reflux drum 108. A portion of the effluent from drum 108 is returned to splitter 106 as reflux stream 122 and the remainder is removed as a 1-hexene product stream 124. In the case described above, the required condenser duty for the hexene tower 106 is greater than the reboiler duty for the butene tower 102. In alternate designs, where the composition of the butene is more dilute or alternately where a product 1-butene stream is taken in addition to a 1-butene stream that would go to metathesis, the duty of the butene splitter reboiler would be greater than the duty of the hexene splitter condenser. In those cases, instead of an additional cooler on stream 110, there would be an additional reboiler on stream 114 to provide the balance of the reboiler duty for the butene splitter.

The extent of cooling water/steam reduction in the hexene and butene splitters depends on the tower specifications and production capacity. The heat integration described in connection

with Figure 1 can yield a significant reduction in both cooling water (CW) and steam for heating (LPS). Tables 3 and 4 below compare the energy for the process before and after the integration of the hexene tower and butene tower for a capacity of 50 KMTA 1 hexene production.

Table 3: Energy balance summary for the high selectivity autometathesis case – before energy exchange

	C4 Isomerization	Autometathesis	C6 Isomerization	Total
Feed vaporizer (LPS), KW	4141	2447	9747	
Feed Heater (fuel), KW	971	1088	1816	
Tower 1 Condenser duty (CW), KW		CW 2500, Refrig 80	24920 Net = 0 (a)	
Tower 1 Re-boiler duty (LPS), KW		3756	19810	
Tower 2 Condenser duty (CW), KW	35110	7551	31070 (b)	
Tower 2 Re-boiler duty (LPS), KW	28400 (b)	3682	23980 Net = 0 (a)	
Pump, power, KW	104	50	165	
<u>Total Utility</u>				
FUEL, KW				3,875
-5 REF, KW				80
CW, KW				76231
LPS, KW				68301
HPS, KW				3,682
POWER, KW				319

Table 4: Energy balance summary for the high selectivity autometathesis case – after energy exchange

	C4 Isomerization	Autometathesis	C6 Isomerization	Total
Feed vaporizer (LPS), KW	4141	2447	9747	
Feed Heater (fuel), KW	971	1088	1816	
Tower 1 Condenser duty (CW), KW		CW 2500 Refrig 80 *	940 (a)	
Tower 1 Re-boiler duty (LPS), KW		3756	19810	
Tower 2 Condenser duty (CW), KW	35110	7551	2670 (b)	
Tower 2 Re-boiler duty (LPS), KW	0 (b)	3682 **	0 (a)	
Pump, power, KW	104	50	165	
<u>Total Utility</u>				
FUEL, KW				3,875
-5 REF, KW				80
CW, KW				48,771
LPS, KW				39,901
HPS, KW				3,682
POWER, KW				319

Notes:

- In autometathesis, tower1 is a depropylenizer (DEC3) and tower 2 is a depentenizer (DEC5).
 - * - DEC3 condenser is -5 deg refrigerant. (ii) ** - DEC5 re-boiler is high pressure steam.
 - (iii) 2500 KW CW interchange on DEC3 to reduce the refrigerant duty. Values shown separately
- In C6 isomerization, tower 1 is higher pressure, and tower 2 is lower pressure than tower 1.
- The energy integration-
 - exchange between tower 1 condenser and tower 2 re-boiler in C6 isom system. Shown as net=0 for both cases. Part of two tower system
 - exchange between tower 2 condenser in C6 isom and re-boiler in C4 Isom. (Note - In first table – values reflect duties before exchange.)

As can be seen from these tables, by exchanging the butene reboiler with the hexene condenser, the CW requirements are reduced from 76231 KW to 48771 KW or 36%, and the low pressure steam heating duty (LPS) is reduced from 68301 KW to 39901 KW or 42%.

For high production capacities, multiple 1-butene/2-butene splitters as well as multiple 1-hexene/ 2-& 3-hexene splitters may be employed. Under this scenario, the combined flows shown as stream 112 of the butene-splitter may be exchanged against the combined flows shown as stream 110 of the hexene-splitters. In some cases, an additional trim cooling water exchanger may be required.

Another technique to increase process efficiency is to divide the feed to the butene splitter and/or the feed to the hexene splitter into two separate towers and run the two towers at different pressures in superfractionation processes. When used in hexene separation, this feature of the disclosed embodiments reduces energy consumption of the hexene tower system by about 33 % at equivalent number of trays and operating pressure. These savings generally are applicable to a superfractionation system in which hydrocarbons with low relative volatilities are separated. For a hexene system, the advantages achieved by this embodiment are added to the advantages that can be achieved by the integrated heat exchange system described above.

Superfractionators consume high quantities of energy in both reboiling and condensing service. Published studies show various options for reducing the energy cost for superfractionation towers. Known systems include:

(a) dual pressure towers in which the tower is split into two distinct towers. The first tower is operated at a higher pressure than the second tower and the condenser of the first tower is used to reboil the second tower, and

(b) use of a heat pump. In this case, the overhead vapor is compressed to a higher pressure. The compression energy heats the vapor to a higher temperature. The higher pressure/higher temperature vapor is then condensed (at the higher temperature) providing reboiling duty for the tower. This option does not save energy per se but allows for operation under conditions that would save energy. By reducing the pressure of the towers, the relative volatility of the hydrocarbons is increased. This allows for fewer stages, or less reflux, or both to achieve a given separation. However, the lower pressure operation will reduce the temperature of the tower and may require refrigeration for the condenser duty as opposed to cooling water. This will increase energy consumption. By operating a heat pump, both the lower temperature duty and the reboiling duty can be supplied by compressive power that does not result in an increase in energy due to added refrigeration.

System (a), a dual tower system, has been used in various separation tower sequences, such as for separating lower olefins from their paraffin analogs. If a single tower is designed for the service, 100 % of the condenser duty is in the overhead and 100% of the reboiling duty is in the bottom. This represents the maximum temperature difference over the tower. By fractionating using two towers at two different pressures, each tower can be operated at a different reflux ratio and the duties can be optimized both in terms of total energy and energy level. The energy level can be defined in many cases as the temperature of refrigeration with the colder levels requiring more energy in refrigeration systems. This integration is not unlike the use of side reboilers or side intercoolers to adjust reflux and temperature profiles within a single tower to optimize utilities.

According to the embodiments described herein, when 1-hexene is separated from 2- and 3-hexene in a two-tower system with split feed, the difference in the pressure at the top of each of the two distillation columns usually is 4- 40 psi, or 6-20 psi, or 8-15 psi. When other closely boiling components are separated using the two tower system described herein, the pressure difference at the top of the two distillation columns usually is in the range of 4-40 psi, or 6-20 psi, or 8-15 psi. The optimal choice of differential pressure is set by the desired temperature difference for heat exchange and the resulting size (and cost) of the exchanger.

The key to efficiently using system (a) is balancing the operation of each of the two towers such that the condensing duty of the first tower matches with the required reboiling duty of the second tower. In some cases, an imbalance is allowed to exist and the difference is made up either by installing a small externally heated reboiler, or alternatively a small cooled condenser, to make up for the difference. This is often used to provide control of the system. Another possibility is to adjust fractionation duty in each tower to achieve a balance. The conventional ways of balancing the operation is to shift duty from one tower to the other by having either some portion of the vapor from the first tower to be allowed to bypass to the second tower, or alternately allowing some portion of the liquid from the second tower to be allowed to pass to the first tower. In the first conventional option, the tower feed is sent to the first or higher pressure column. The amount of condensing duty for the first tower is controlled by operation of a vapor bypass from the first tower to the second tower. This bypass shifts duty from the condensing portion of one tower to the reboiling duty of other thus allowing the condenser of tower 1 to match the duty of the reboiler of tower 2. The savings realized from this type of system are principally in energy levels (temperatures) rather than total energy saving. Furthermore, for towers with large numbers of stages, splitting the tower into two distinct towers reduces the cost compared to a single very tall tower

A comparison between a conventional single tower system and a conventional two tower system with vapor bypass from the first tower to the second tower is provided below. Cases A and B represent single tower systems with different capacities and number of trays. Case E uses lower pressure drop trays (tower dP = 20 kPa versus 35 kPa for Cases A and B). Cases C and D represent two tower systems with similar capacities but differing number of trays. The single tower hexene system is a conventional distillation unit that typically has 100+ (usually 150 +) stages. The stages are numbered from top. The feed enters at stage number 75. The product 1-hexene is removed as the distillate. The 2-hexene and 3-hexene in the bottoms are recycled to the hexene isomerization reactor (not shown) where they are isomerized to result in a mixture of 1-, 2- and 3-hexene and the fluent returned to the tower. In this manner, 3-hexene is converted to 1-hexene. A small purge is removed from the system. The tower specifications are given in Table 5. Cases A and B operate at different flows and pressures and different number of stages and generating slightly different purities of 1-hexene.

Table 5: Summary of specifications for conventional single hexene tower cases

Parameter	Single Tower Case A	Single Tower Case B	Case E Single Tower
Feed to Hexene System	76.3 kmol/H (51296 KTA)	40.7 kmol/H (27464 KTA)	36.5 kmol/h (24640 KTA)
1-Hex Prod rate	73.4 kmol/H 49360 KTA 98.3 wt% 1-hex	37.8 kmol/H 25461 KTA 98.44 wt% 1-hex	33.8 Kmol/H 22760 KTA 98.5 wt% 1-Hex
Number of Stages	125	150	150
Feed Stage	75	75	75
Condenser Pressure	207 Kpa	120 Kpa	120 Kpa
Re-boiler Pressure	241 Kpa	140 Kpa	140 Kpa
Top Spec	0.8 mol% (2- + 3-) hex	1.5 mol% (2- + 3-) hex	1.4 mol% (2- +3-) hex
Bottom Spec	1 mol% 1-hex	0.75 mol% 1-hex	1.6 mol% 1-Hex
Other Spec			C7 at Bottom=45 mol%
Side-Draw Liquid Stage			145
Recycle/Fresh Feed			13

A conventional dual pressure distillation separation of 1-hexene is carried out in two towers with 100 stages each, providing a total of 200 stages. A schematic drawing of the two-tower system is shown as Fig. 2 and is generally designated as 200. The system includes a first distillation tower 202 and a second distillation tower 204. The feed containing approximately 8% 1 hexene with the balance comprising mostly 2 and 3 hexene and a small amount of C7+ material, enters the first tower 202 at stage 40 in stream 206. The bottom product of this tower in stream 211 is mainly C7+, which is purged in stream 213, with the remainder being returned to the bottom of tower 202 in stream 215. The distillate product from the first tower is 35% 1-hexene in stream 212. A vapor side-draw is taken from stage 81 of first tower 202 in stream 217 to balance the interchange duties. If the first tower 202 is integrated with an isomerization reaction system, a liquid side-draw is taken from stage 87 in stream 214 for feeding to the isomerization reaction system. This recycle stream comprising mostly 2 and 3 hexene, is admixed with fresh 3 hexene feed (not shown), passed through the isomerization reactor (not shown) to produce some 1 hexene from the 2 and 3 hexenes, and is returned to the first tower as feed 206.

The distillate from the first tower 202 in stream 212 is condensed in a condenser 216 and is then divided into a reflux stream 218, which enters the top of the first tower 202 and a distillate stream 220. The distillate stream 220 enters the second tower 204 at stage 54. The vapor side-

draw in stream 217 enters the second tower 204 at stage 90. Stream 217 enters at a lower stage than stream 220 since it was withdrawn at a lower stage and thus contains less 1-hexene. The product 1-hexene is removed as the distillate from the second tower 204 in stream 222, condensed, and divided into a 1-hexene product stream 223 and a reflux stream 225. The bottom product from the second tower 204 in stream 224 is reboiled in reboiler 228 and a portion is returned to second tower 204 in stream 226. Another portion is mixed with the liquid side-draw from the first tower in stream 214 to form stream 232, which is recycled to the hexene isomerization reactor. A small purge can be removed in stream 234 if required.

The second tower 204 operates at a lower pressure than the first tower 202. This enables the condenser 216 of the first tower 202 to be operated at a higher temperature than the reboiler 228 of the second tower 204, allowing heat exchange between them. Thus, the heat integration is achieved by linking the condenser from the first tower 202 with the reboiler of the second tower 204. The tower specifications are given in Table 6. Two different designs are presented in which the total number of fractionation stages is varied. The first case uses 100 trays in each of the two towers (200 total) while the second uses 150 trays in each of the two towers (300 total).

Table 6 - Summary of specifications for conventional hexene two-tower vapor bypass cases

Parameter	Case C Tower 1	Case C Tower 2	Case D Tower 1	Case D Tower 2
Fresh Feed to Hexene System	34.2 kmol/h 23088 KTA		34.2 kmol/h 23096 KTA	
1-Hex Production Rate		33.4 kmol/h 22472 KTA 98.5 wt% 1-Hex		33.4 kmol/h 22544 KTA 98.5 wt% 1-Hex
Number of Stages	100	100	150	150
Feed Stage	40	54, 90	32	58, 105
Condenser Pressure	207 Kpa	117 Kpa	207 Kpa	117 Kpa
Re-boiler Pressure	241 Kpa	138 Kpa	241 Kpa	138 Kpa
Top Spec	65 mol% (2- + 3-) hex	1.15 mol% (2- + 3-) hhex	65 mol% (2- + 3-) hex	0.6 mol% (2- + 3-) hex
Bottom Spec	95 mol% C7	1 mol% 1-hex	95 mol% C7	1 mol% 1-hex
Side-Draw Vapor Stage	81	NA	90	NA
Side-Draw Vapor Rate	185 kmol/hr	NA	208 kmol/hr	NA
Side-Draw Liquid Stage	87	NA	129	
Side-Draw Liquid Rate	118 kmol/hr	NA	118 kmol/hr	

The energy consumption for single tower Cases A-B and E and two-tower cases C-D is shown below on Tables 7 and 8.

Table 7 - Summary of energy use for conventional single tower cases A-B and E

Parameter	Case A	Case B	Case E
Feed to Hexene System	76.3 kmol/hr (51296 KTA)	40.7 kmol/hr (27464 KTA)	36.5 kmol/h (24640 KTA)
1-Hex Prod rate	73.4 kmol/hr 49360 KTA 98.3 wt% 1-hex	37.8 kmol/hr 25461 KTA 98.44 wt% 1-hex	33.8 Kmole/H 22760 KTA 98.5 wt% 1-Hex
Q,condenser, KW	60,030	30,300	24,560
P overhead Kpa	206		
T condenser C	86		68
Q,reboiler, KW	47,480	24,100	19,030
<i>T reboiler C</i>	99		90
Q,condenser KW/KTA of 1-Hex Prod	1.216	1.19	1.08
Q,reboiler KW/KTA of 1-Hex Prod	0.962	0.9465	0.836
(Qc+Qr) KW /KTA 1-Hex	2.178	2.1325	1.91
Recycle/Fresh Feed	12.755	13.05	13
Hexene recovery (prod/feed) %	96.2	92.7	92.3

Table 8 - Summary of energy use for two-tower cases C-D using a conventional vapor side draw to control balance

Parameter	Case C Tower 1	Case C Tower 2	Case D Tower 1	Case D Tower 2
Feed to hexene System	34.2 kmol/h 23088 KTA		34.2 kmol/h 23096 KTA	
1-Hex Prod Rate		33.4 kmol/h 22472 KTA 98.5 wt% 1-Hex		33.4 kmol/h 22544 KTA 98.5 wt% 1-Hex
Q,c, KW	24,640	26,290	15910	17720
P Overhead KPa	207	117		
Condenser T, deg C	91	68	91	68
Q,r, KW	21,210	21,490	12730	15890
Re-boiler T, deg C	124	78	124	78
Net Q,c, KW	26,290		17720	
Net Q,r, KW	21,210		12730	
Net Q,c KW/KTA Prod	1.17		0.786	
Net Q,r KW/KTA Prod	0.944		0.5646	
(Qc+Qr) KW/KTA Prod	2.114 *		1.35 **	
Recycle/Fresh Feed	10.5		10.3	
Hexene recovery	97.3		97.6	

* 100 TRAYS PER TOWER (=200) ** 150 TRAYS PER TOWER (=300)

When the results are tabulated in terms of energy use it is evident that the total energy usage for the two systems is similar at an equivalent number of distillation stages. This is shown in the graph of Figure 3. The energy use of both the single tower and two tower vapor bypass fall along a similar curve. There are slight differences depending upon operational pressure with the lower pressure single tower case having about 10% lower specific energy. As can be seen, Case A has a high recovery (tight specifications) compared to case B. This does not have a dramatic impact on energy. However case B and case E have similar recovery but case E has much lower tray dP. This impacts relative volatility.

The advantages of a two tower system with a vapor bypass are not in the energy per ton of product but in the ability to increase the reboiler temperature of the first tower. In this case, the reboiler temperature is increased from 99° C for the single tower case to 124° C for the two-tower case. Two-tower vapor bypass systems are most applicable for refrigeration systems where the

reboiler requires refrigeration. In many cases this increase can shift the duty from a low temperature requiring refrigeration compression to a higher temperature level that requires either less refrigeration or potentially can be reboiled by steam or a lower cost energy source. While the exemplified case does not use refrigeration, the shift in reboiler temperature illustrates the principle. It is noted that in many cases with very large numbers of trays (>100), towers are designed as two towers in series to reduce costs by reducing foundations and structures. If the tower is to be built as two columns in series, the dual pressure arrangement is preferred.

Other embodiments disclosed herein are dual pressure two-tower systems with feed splitting and heat integration, along with corresponding methods. These systems have reduced energy consumption as compared to the conventional dual pressure two tower system described above. The system and method are schematically shown in Fig. 4. The overall system is designated as 250. While the drawing and description are directed to the separation of 1-hexene from 2-hexene and 3-hexene, the scope of the embodiment includes the production of other types of closely boiling components.

The distillation separation of 1-hexene from 2- and 3-hexene is carried out in towers 252 and 254. In the embodiment that is illustrated, tower 252 has 80 stages and tower 254 has 70 stages. Thus, the total number of stages (150) is the same as in one of the single-tower case described above. It is noted that in practice other numbers of stages can be used.

The feed to the separation system is split, with one portion of feed stream 256 entering the first tower 252 as stream 258 at stage 25. The other portion enters the second tower 254 as stream 260 at stage 60. The bottom product of tower 252 in stream 261 is mainly C7 and most of this is purged in stream 263. A reboiler 259 heats reboiler stream 265, which is returned to the first tower 252. The distillate from the first tower 252 in stream 262 is 40% 1-hexene. A liquid side-draw containing 2-hexene and 3-hexene is taken from stage 74 of the first tower 252 as stream 264, for feeding to the isomerization reaction system (not shown). This recycle stream comprising mostly 2- and 3-hexene, is admixed with fresh 3-hexene feed (not shown), passed through the isomerization reactor (not shown) to produce some 1-hexene from the 2- and 3-hexenes, and is returned to the first tower as feed 256.

The distillate 262 from the first tower 252 is condensed in condenser 266 and the condensate is separated into a reflux stream 268, which returns to the top of the first tower 252, and a second tower feed stream 270, which enters the second tower 254 at stage 40, which is above the feed point for the second portion of the fresh feed that enters at stage 60. It is noted that side draw and feed locations other than those described herein can be used.

The product 1-hexene is removed from the second tower 254 as the distillate in stream 272. The distillate 272 is condensed in a condenser 274 and separated into a reflux stream 273, and a 1-hexene product stream 275. The bottom product from the second tower 254 in stream 274 is partially reboiled in a reboiler 278. The reboiled stream is separated into stream 276,

which is returned to the bottom of the second tower 254, and a remaining bottoms stream 280. The bottoms stream 280 is mixed with the liquid side-draw in stream 264 from the first tower as stream 282. A portion of stream 282 is separated as a C6 purge stream 284 and the remaining stream 286 is recycled to the hexene isomerization reactor (not shown).

The second tower 254 operates at a lower pressure compared to the first tower 252. This enables the condenser of the first tower to be at a higher temperature than the re-boiler of the second tower, allowing heat exchange between them. These duties are balanced by the splits of the feed to the towers. By sending more or less fresh feed to tower 2, the relative duties of reboiler 278 and condenser 266 can be balanced. Thus the heat integration is achieved by linking the condenser from the first tower with the reboiler of the second tower.

Further advantages to the process of Fig. 4 can be realized if a heat pump is incorporated that circulates a heat and cooling fluid between the reboiler 259 of the second tower 252 and the condenser 274 for the first tower 254. In one embodiment, an open loop heat pump employing distillate from the first tower 254 is used, as is shown in Fig. 5. In this embodiment, many of the system components are the same, with the overall system being designated at 250', with the feed 256' being split into first tower feed 258' entering first tower 252' and second tower feed 260' entering second tower 254'. Distillate 262' is removed from first tower 252', condensed in condenser 266' and separated into reflux stream 268' and second tower feed stream 270'. The differences are that the distillate 272' from second tower 254' is compressed in a compressor 290 and used as the heating fluid in reboiler 259'. Stream 272' is then separated in a separator 292 into a vapor stream 275', which is removed as product, and a liquid stream 273', which is used as a reflux stream to the fractionator 254'. In this embodiment, the compressor pressurizes the hot vapor to the pressure that will result in the desired reboiling temperature.

The tower specifications for a computerized simulation are given in table 9. In the Table, Case F represents a two-tower feed split tower. In this example, the pressure of the first split feed tower is raised considerably compared to the single tower system. The overhead pressure of the second tower is roughly equivalent to the overhead pressure of the single tower. This allows for the equivalent interchange between the hexene condenser and the butene tower reboiler as discussed above if that is used in combination with this embodiment.

Table 9: Summary of Specifications for Case F

Parameter	Case F Tower 1	Case F Tower 2
Feed to hexene System	36.5 kmol/h 24640KTA	
1-Hex Prod Rate		33.8 Kmol/H 22736 KTA 98.5 wt% 1-Hex
Number of Stages	80	70
Feed Stage	25	40,60
Condenser Pressure	207Kpa	117Kpa
Re-boiler Pressure	241 Kpa	138 Kpa
Top Spec	60 mol% (2+3) Hex	1.4 mol% (2+3) Hex
Bottom Spec	1.0 mol% 1-Hex	2.5 mol% 1-Hex
Other Spec	C7 at Bottom=45 mol%	
Side-Draw Liquid Stage	74	
Recycle/ Fresh Feed	13	

Table 10 - Summary of Results for Case F

Parameter	Case F -Tower 1	Case F -Tower 2
Feed to hexene System	36.5 kmol/h 24640KTA	
1-Hex Prod Rate		33.8 Kmol/H 22736 KTA 98.5 wt% 1-Hex
Q,c, KW	17,000 Note 1	18,800
Condenser T, deg C	90	67
Q,r, KW	13,830	16,740 Note 1
Re-boiler T, deg C	110	78
Net Energy	Case F	
Net Q,c, KW	18,800	
Net Q,r, KW	13,830	
Net Q,c/KTA Prod	0.83	
Net Q,r/KTA Prod	0.61	
(Qc+Qr)/KTA Prod	1.44	
Recycle/Fresh Feed	13	
Hexene recovery	92.3	

Note 1 – tower 1 condenser is exchanged with tower 2 reboiler

Table 10 presents the summary of the results of the simulation. The energy usage of the two tower feed split system has dropped to 1.44 KW/KTA product hexene-1. This is also shown on Figure 3 as "Two Tower Feed Split." This value can be compared to 2.13 for the single tower system with identical number of trays and equivalent tray pressure drop and specifications. This is a reduction of 32%. The single tower operated with the same number of lower pressure drop trays (higher cost trays) and lower 1-hexene recovery is still 23% higher energy (1.91 vs 1.44). Further, the energy is comparable to the conventional dual pressure two tower system requiring twice the number of trays. This represents a substantial capital advantage for the system of the disclosed embodiments. The first tower of the two tower system operates at a higher pressure than the single tower system (or the second tower). Thus a large portion of the fractionation occurs at a higher pressure, leading to a smaller tower diameter for that portion and thus a savings in capital costs.

It is unexpected that the arrangement shown in Fig. 4 leads to a significant reduction in energy usage for a two tower system used in separating mixtures containing materials with very

close boiling points. Split feed towers have been utilized in some applications using single towers in which each portion of the split feed is of a different composition. For example, a split feed tower is known for use after a vapor-liquid separation. In this case, the vapor is fed higher in the tower and liquid is fed lower in the tower to take advantage of the equilibrium flash separation. Feeding the identical feed composition and of similar phases (vapor or liquid) into locations that are 25 fractionation trays apart, located in two different towers, and operated at different pressures is contrary to established fractionation practice that searches for an optimum feed location. The optimum feed location is typically defined by the point within the tower at which a similar composition occurs. However by using the concept of the disclosed embodiments as a means of balancing the operation of a two tower – two pressure system, significant energy savings can be realized.

In the split feed embodiment that is described, the pressure levels of the towers can be raised or lowered to optimize energy use. This can lead to additional energy savings in refrigeration systems, such as with conventional two tower systems, in addition to achieving absolute fractionation energy savings.

In a further embodiment, the side draw location of stream 264 in Fig. 4 from first tower 252 can be selected in order to minimize the 1-hexene content of that stream dependent upon the particulars of the system design. As indicated above, due to the close relative volatility between the 1-hexene and the 2- and 3-hexenes, separation of high purity 1-hexene (>99 mol.%) requires a significant number of stages and tower reflux. As a result, the hexene fractionation step is energy intensive. The hexene fractionator removes 1-hexene in the tower overhead and 2- and 3-hexene in the tower bottoms. The 2- and 3-hexene tower bottoms is recycled to the hexene isomerization reactor where it forms more 1-hexene. The hexene isomerization reactor effluent is combined with fresh 3-hexene feed and returned to the hexene fractionator. At any given isomerization reactor outlet temperature, the concentration of 1-hexene is limited by equilibrium. Operation at high temperatures is limited by fouling, and operation at lower temperatures is limited by unfavorable equilibrium. The conversion per pass for the isomerization reactor is set by the inlet concentration of 1-hexene in the recycle stream and the outlet concentration in the reactor outlet set by equilibrium.

It is expected that one who is building a system for producing linear alpha olefins such as 1-hexene will design a fractionation system to achieve a low concentration of 1-hexene in the fractionator bottoms (an example of 1 mol.% is shown in Mode 1 below), thereby increasing the downstream hexene isomerization reactor per-pass conversion. The lowest possible concentration of 1-hexene in the tower system is at the bottom of the tower. A single tower scheme is plausible because a low concentration of 1-hexene in the fractionator bottoms reduces the amount of hexene tower recycle. Fig. 6 shows a conventional fractionator 290. A portion of the bottoms stream 291 is recycled to a hexene isomerization reactor 292 and is then returned to

the fractionator 290. The design of Fig. 6 (Mode 1, shown below, with a low concentration of 1-hexene in the bottoms) also allows for a high recovery of 1-hexene (ratio of 1-hexene produced over total hexenes in the fresh feed) because little 1-hexene is lost through purging.

The scheme of Fig. 7 can be used in conjunction with the embodiment of Fig. 4 and also can be incorporated into the embodiment of Fig. 2. In the embodiment depicted in Fig. 7 (Mode 2), a hexene tower side draw is removed at a higher 1-hexene composition than is present at the bottom of the tower and is fed to the hexene isomerization reactor. More specifically, the fractionator 294 has a side draw 295, which is sent to an isomerization reactor 296. A non-limiting example of the 1-hexene concentration in the side draw is 2.5 mole % 1-hexene. The larger concentration of 1-hexene in the reactor feed reduces the per-pass hexene reactor conversion and increases the hexene isomerization reactor recycle rate by 2.7%. This increase will increase the energy requirement for the isomerization reactor. The hexene purge (tower bottoms) in Fig. 7 (Mode 2) is still removed from the bottom of the tower at the same 1-hexene bottom specification. Mode 2 increases the hexene isomerization reactor recycle rate by 25%, thereby increasing the utilities of the isomerization system including vaporization and preheat, and thus it is not obvious that the hexene tower internal-reflux flow decreases by a larger amount on an absolute basis. As a result, there is a net reduction in tower plus isomerization energy consumption and reflux ratio using Mode 2 for the same 1-hexene recovery rate. This net reduction in tower reflux flow occurs because less energy is required to meet a slightly relaxed hexene isomerization recycle specification.

The reduced energy usage when Mode 2 is used instead of Mode 1 is shown below on Table 11.

Table 11 – Material and energy balances showing effect of side draw

Parameter	MODE1	MODE2
Material balance		
C6 Feed to Hexene System, kg/hr	6751	6751
C6 Recycle Flow, kg/hr	92468	94937
Purge Flow, kg/hr	690	684
1-Hex prod, kg/hr	6061	6067
Other Parameters		
Recycle/ Fresh Feed rate	13.69	14.06
C6 tower reflux ratio	93.3	88.2
1-Hex in bottom purge, mol%	1.0	1.0
1-Hex in recycle flow, mol%	1.0	2.50
Energy Balance		
C6 tower Condenser duty, MKCAL/hr	47.01	44.5
C6 tower Re-boiler duty, MKCAL/hr	36.03	33.29
Rx feed heater fuel, MKCAL/hr	1.26	1.50
Rx Feed Vaporizer LPS, MKCAL/hr	9.29	9.29

C6 Recycle Pump, KW	16.4	16.9
Total Energy Usage (Tower plus isomerization reactor) MM Kcal/hr	93.59	88.58

As shown on Table 11, the total energy usage decreases from 93.59 MM Kcal/hr to 88.58 MM Kcal/hr when Mode 2 is used, resulting in an energy reduction of about 5%.

Certain embodiments of the autometathesis process described in U.S. Patent No. 6,727,396 include a butene isomerization reactor and a hexene isomerization reactor used to convert the interior butene and hexene olefins to the exterior alpha olefins, respectively. In addition the process employs an autometathesis reactor system. Each of these three reaction systems are vapor phase fixed bed reaction systems. Each requires that the feed to the reactor be first vaporized (since the feeds are liquid at the point of feeding the unit). Furthermore, for each reactor, the effluent temperature is significantly higher than the feed temperature and/or the fractionation tower that follows the reactor. These temperatures are required since higher temperatures in the isomerization step favor the desired alpha olefin, and higher temperatures for the autometathesis are required for the reaction kinetics. In a conventional system, the effluent from the reactor would be used to vaporize and preheat the feed to a certain level and the balance of the preheat would be supplied by a fired heater.

Fig. 8 shows a process flow scheme for producing 1-hexene in which additional heat exchangers 330, 332 and 334 are added in order to reduce the capital costs of the furnaces 342, 344 and 346. The system shown in Fig. 8 includes a hexene isomerization and purification zone 302, a butene isomerization and purification zone 304, and an autometathesis zone 306. Butene feed stream 305 is combined with recycle stream 311 to form stream 307. Stream 307 is heated in heat exchanger 330, further heated in heat exchanger 336, sent to furnace 342 and then isomerized in reactor 331. The reactor effluent stream 308 is fractionated in fractionator 309 and the top stream 310 is divided into a reflux stream 341 and a metathesis feed stream 313, which is sent to the metathesis zone 306. The bottoms stream 331 is recycled and can be partially purged in stream 337. In the metathesis zone 306, stream 313 is combined with metathesis recycle stream 312 to form stream 314. Stream 314 is heated in heat exchanger 334, further heated in heat exchanger 340, sent to furnace 346, and subjected to metathesis in reactor 335. The reactor effluent in stream 316 is fractionated in a fractionation section 318. By-products are removed in stream 320 and purge stream 322 is withdrawn. C4's and C5's are recycled in stream 312. The metathesis product stream 324 is sent to the hexene isomerization and purification zone 302.

In zone 302, stream 324 is heated in heat exchanger 332, heated in heat exchanger 338, sent to furnace 344, and then isomerized in reactor 333. The reactor effluent in stream 325 is fractionated in fractionation section 326 to form a 1-hexene product stream 328, a C6 purge

stream 329 and a C7+ purge stream 339. By adding separate heat exchangers 330, 332, 334 to preheat the reactor feeds upstream from the butene isomerization reactor 331, hexene isomerization reactor 333, and autometathesis reactor 335, respectively, energy consumption in the furnaces that also are upstream from each of these reactors is reduced. Most of the required heat duty takes place in the vaporizers 330, 332 and 334, upstream from the process/process exchangers 336, 338 and 340, which are directly upstream from the furnaces 342, 344 and 346, respectively. The flow scheme illustrated in Fig. 8 results in a reduction in furnace capital cost by over 80 % as compared to a prior known system in which the heat exchangers 330, 332 and 334 are not included.

The heat exchangers 330, 332 and 334 often are vaporizers. These heat exchangers optionally can utilize waste heat from another source if available. In an ethylene system, for example, this can be quench water. Furthermore, in any integrated steam system, maximizing the use of "pass-out" steam will increase the overall efficiency of the steam system. A fired heater requires additional heat recovery (after the process duty preheat) in order to not waste energy. A larger fired duty in the heater requires a larger quantity of additional heat, resulting in an additional capital cost for the system. It is counter-intuitive to add equipment to achieve reduced equipment cost. However, in this case, a low cost vaporizer is added to reduce the size of a higher cost fired heater thus saving capital and reducing energy consumption.

While preferred embodiments have been shown and described, various modifications may be made to the processes described above. Accordingly, it is to be understood that the present invention has been described by way of example and not by limitation, and the scope of the claims should not be limited by particular examples set forth herein, but should be construed in a manner consistent with the description as a whole.

CLAIMS:

1. A process comprising:
 - (a) separating a mixed butene stream comprising 1-butene and 2-butene into an overhead 1-butene stream and a bottoms 2-butene stream in a butene distillation column, a portion of the bottoms 2-butene stream being separated to form a butene reboiler stream that is heated and vaporized in a reboiler and returned to the butene distillation column,
 - (b) subjecting at least a portion of the overhead 1-butene stream from (a) to catalytic metathesis to produce an effluent including 3-hexene,
 - (c) isomerizing 3-hexene from (b) to produce a mixed hexene stream comprising 1-hexene, 2-hexene and 3-hexene,
 - (d) separating the mixed hexene stream in a hexene fractionation tower to form a 1-hexene vapor overhead stream that is condensed in a cooler and a bottoms stream comprising 2-hexene and 3-hexene, and
 - (e) using heat obtained by condensing the 1-hexene vapor overhead stream of (d) to heat the butene reboiler stream of (a).
2. The process of claim 1 wherein the 1-hexene is separated from 2- and 3-hexene in (d) at a pressure of from about 5 psig to 40 psig.
3. The process of claim 1 wherein the separation in (d) takes place at an overhead tray temperature from about 120° C to about 220° C.
4. The process of claim 1 wherein the separation in (a) takes place at a pressure from about 40 psig to 100 psig.
5. The process of claim 1 wherein the separation in (a) takes place at a butene distillation column bottoms temperature from about 80° C to about 180° C.
6. The process of claim 1, wherein the use of the 1-hexene vapor overhead stream to heat the butene reboiler stream results in a reduction in steam and cooling water duty of at least 10 % as compared to a process in which the hexene fractionation tower is operated at a

pressure from about -5 to about +5 psig and steam provides all of the heating of the butene reboiler stream at the same 1-hexene production rate.

7. The process of claim 1, further comprising isomerizing 2-butene to form additional 1-butene before (a).

8. The process of claim 1, wherein catalytic metathesis in (b) also produces ethylene, and the process further comprises separating 3-hexene from ethylene before (c).

9. The process of claim 1, wherein the hexene fractionation tower includes a first hexene distillation column having a first overhead stream and a first bottoms stream and a second hexene distillation column downstream from the first hexene distillation column having a second overhead stream and a second bottoms stream, a first portion of the mixed hexene stream is fed to the first distillation column and a second portion of the mixed hexane stream is fed to the second distillation column, and at least a portion of the second bottoms stream is heated using the first overhead stream, and is returned to the first hexene distillation column.

10. The process of claim 9, wherein at least a portion of the first overhead stream is fed to the second distillation column above the feed point of the second portion of the mixed hexene stream.

11. The process of claim 1, further comprising: (g) removing a side draw from the hexane fractionation tower and isomerizing the removed side draw in (c).

12. A system for producing 1-hexene from butenes, comprising:

a butene distillation column configured to separate 1-butene from 2-butene, the butene distillation column having a butene reboiler associated therewith including a first heat source,

a metathesis reactor configured to convert 1-butene to 3-hexene and ethylene,

a metathesis recovery zone configured to separate 3-hexene from ethylene,

a hexene isomerization reactor configured to isomerize 3-hexene to 1-hexene and 2-hexene,

a hexene distillation column configured to separate 1-hexene vapor from 2-hexene and 3-hexene, and a condenser configured to condense the 1-hexene vapor, the first heat source for the butene reboiler being at least a portion of the heat of condensation of the 1-hexene vapor.

13. The system of claim 12, further comprising a butene isomerization reactor positioned upstream from the butene distillation column configured to isomerize 2-butene to form 1-butene.

14. The system of claim 12, wherein the butene reboiler has a second heat source comprising steam.

15. The system of claim 12, further comprising a vaporizer, a process/process heat exchanger and a furnace upstream from at least one of the metathesis reactor and the hexene isomerization reactor, the vaporizer reducing the heating requirement of the furnace.

16. The system of claim 13, further comprising a vaporizer, a process/process heat exchanger and a furnace upstream from at least one of the metathesis reactor, the hexane isomerization reactor and the butene isomerization reactor, the vaporizer reducing the heating requirement of the furnace.

17. A separation process, comprising:

(a) obtaining a hydrocarbon mixture comprising a first hydrocarbon, a second hydrocarbon, and higher boiling hydrocarbons, the second hydrocarbon having a boiling point at atmospheric pressure that is higher than the boiling point of the first hydrocarbon at atmospheric pressure by 0.3 to 10° C,

(b) dividing the hydrocarbon mixture into a first feed stream and a second feed stream,

(c) feeding the first feed stream to a first intermediate stage of a first distillation column operating at a first overhead pressure,

(d) feeding the second feed stream to a second intermediate stage of a second distillation column operating at a second overhead pressure that is lower than the first overhead pressure,

(e) removing a first distillate containing most of the first hydrocarbon from the top of the first distillation column, at least partially condensing the first distillate, and feeding the at least partially condensed first distillate to the second distillation column at a third intermediate stage above the second intermediate stage,

(f) removing higher boiling hydrocarbons from the bottom of the first distillation column,

(g) removing the first hydrocarbon from the top of the second distillation column,

(h) removing the second hydrocarbon from the bottom of the second distillation column, and

(i) heating at least a portion of the second hydrocarbon removed from the bottom of the second distillation column using the first distillate and returning the heated portion to the second distillation column, wherein the second distillation column is operated at an average pressure that is at least 7 psi lower than the average operating pressure of the first distillation column.

18. The process of claim 17, further comprising (j) removing a first liquid side draw from the first distillation column at a fourth intermediate stage that is below the first intermediate stage.

19. The process of claim 18, wherein a first portion of the first liquid side draw is purged.

20. The process of claim 17, wherein the first hydrocarbon is 1-hexene and the second hydrocarbon is at least one member selected from the group consisting of 2-hexene and 3-hexene.

21. The process of claim 17, wherein the first hydrocarbon is 1-butene and the second hydrocarbon is 2-butene.

22. The process of claim 17, wherein a first portion of the first liquid side draw is purged and a second portion is isomerized.

23. The process of claim 18, wherein energy usage in the isomerization reactor can be controlled by adjusting the flow rate of the first liquid side draw.

24. The process of claim 17, wherein the third intermediate stage of the second distillation column is above the second intermediate stage.

25. The process of claim 17, wherein heating at least a portion of the second hydrocarbon removed from the bottom of the second distillation column using the first distillate and returning the heated portion to the second distillation column results in an energy saving of at least 10 % as compared to a system in which a single distillation column is used at the same 1-hexene production rate.

26. The process of claim 17, wherein heating at least a portion of the second hydrocarbon removed from the bottom of the second distillation column using the first distillate and returning the heated portion to the second distillation column results in an energy saving of at least 20 % as compared to a system in which a single distillation column is used at the same 1-hexene production rate.

27. A superfractionation system comprising:

a first distillation zone including a first distillation column, a first condenser with a first coolant source, and a first reboiler with a first heat source, the first distillation column having a first feed inlet configured to receive a first portion of fresh feed, a first distillate outlet and a first bottoms outlet,

a second distillation zone including a second distillation column configured to operate at a lower pressure than the first distillation column, a second condenser, and a second reboiler with a second heat source, the second distillation column having a second feed inlet configured to receive a second portion of fresh feed, a third feed inlet disposed above the second feed inlet and configured to receive at least a portion of the distillate from the first distillation zone, a second distillate outlet, and a second bottoms outlet,

wherein the second heat source for the second reboiler is the first coolant source for the first condenser.

28. The superfractionation system of claim 27, wherein the first and second distillation zones are configured to separate 1-hexene from 2-hexene and 3-hexene.

29. The superfractionation system of claim 27, wherein the first and second distillation zones are configured to separate 1-butene from 2-butene.

30. The superfractionation system of claim 27, further including a first side outlet on the first distillation column configured to withdraw a side stream to be fed to an isomerization reactor.

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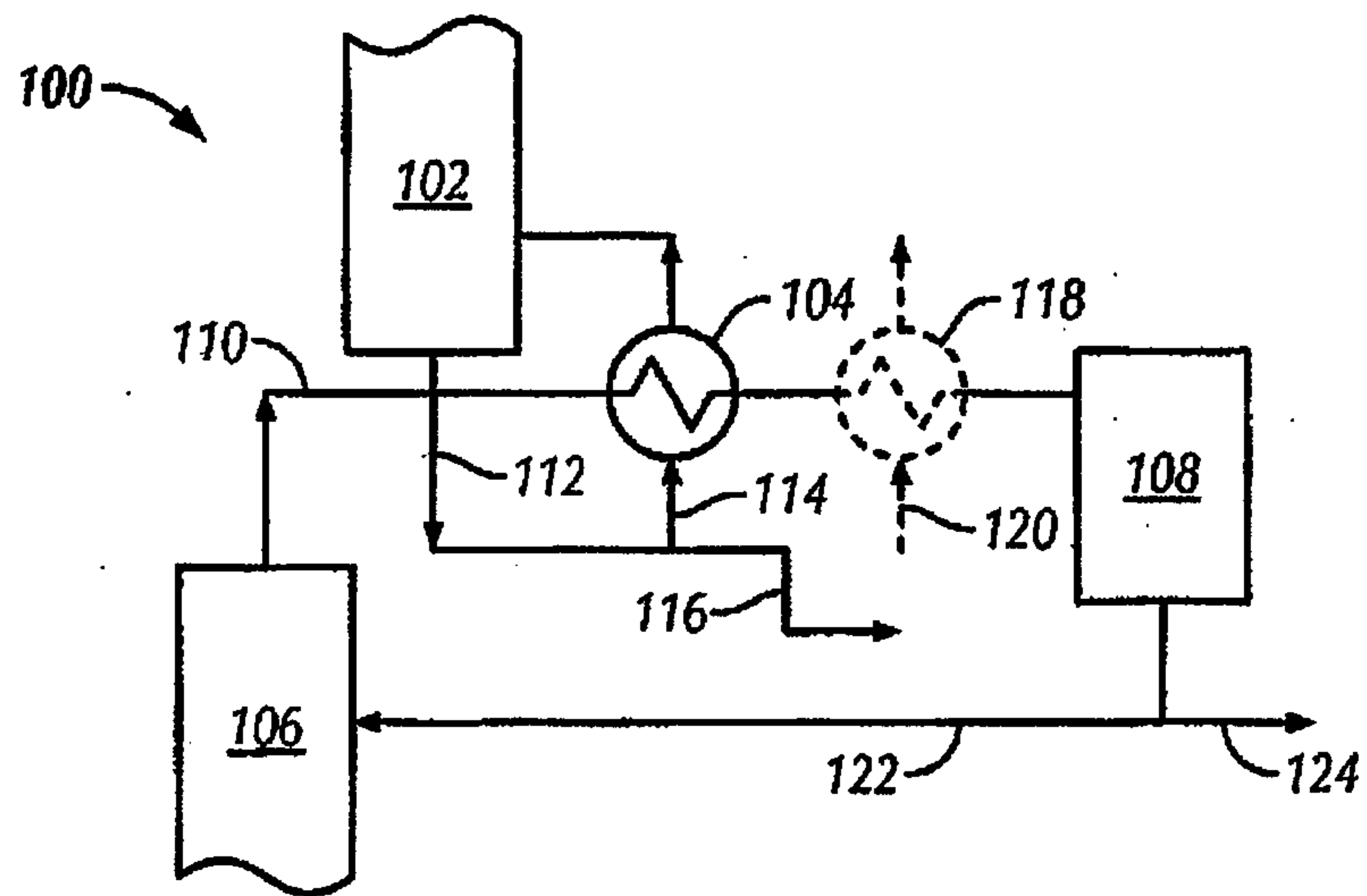


FIG. 1

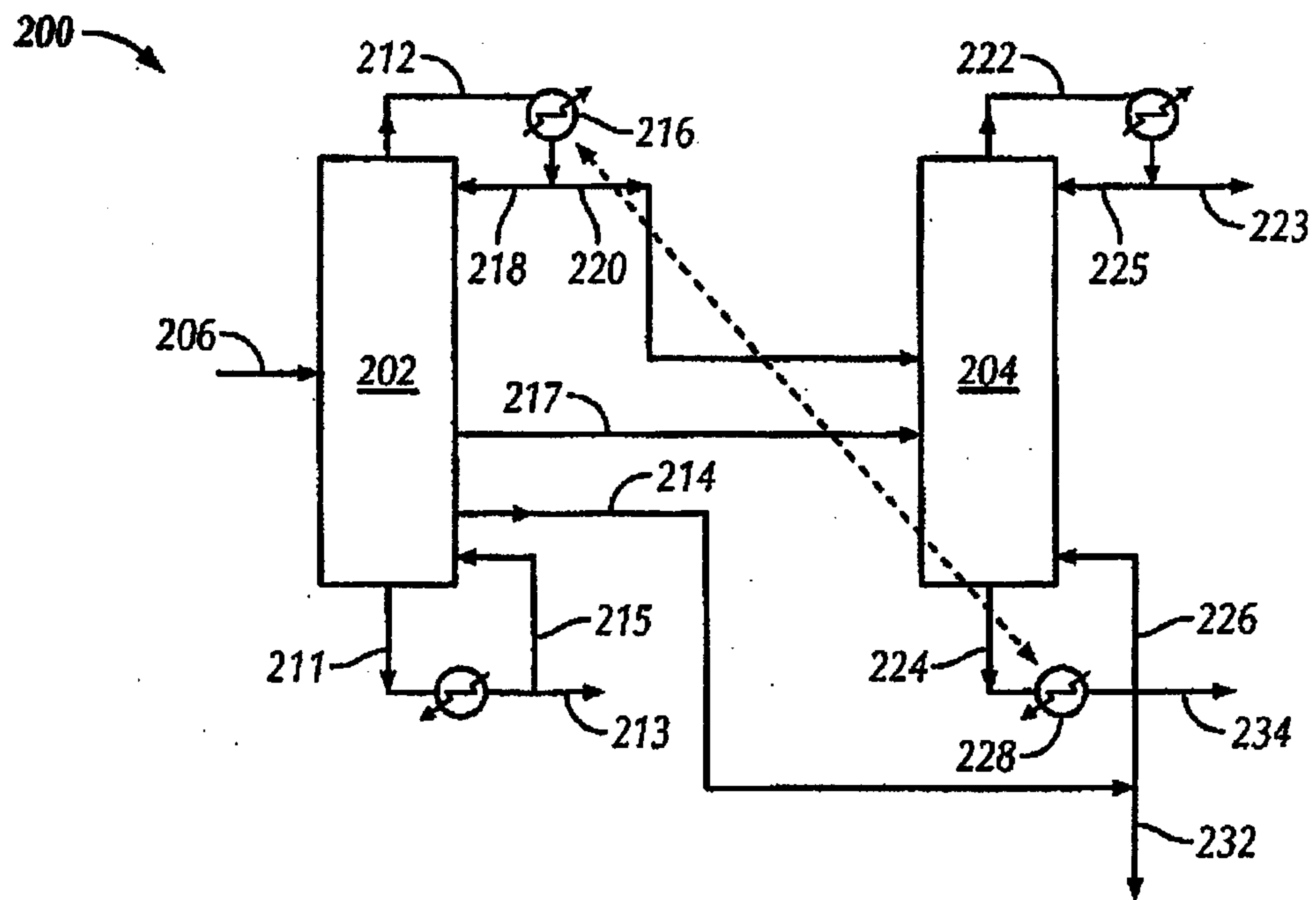


FIG. 2

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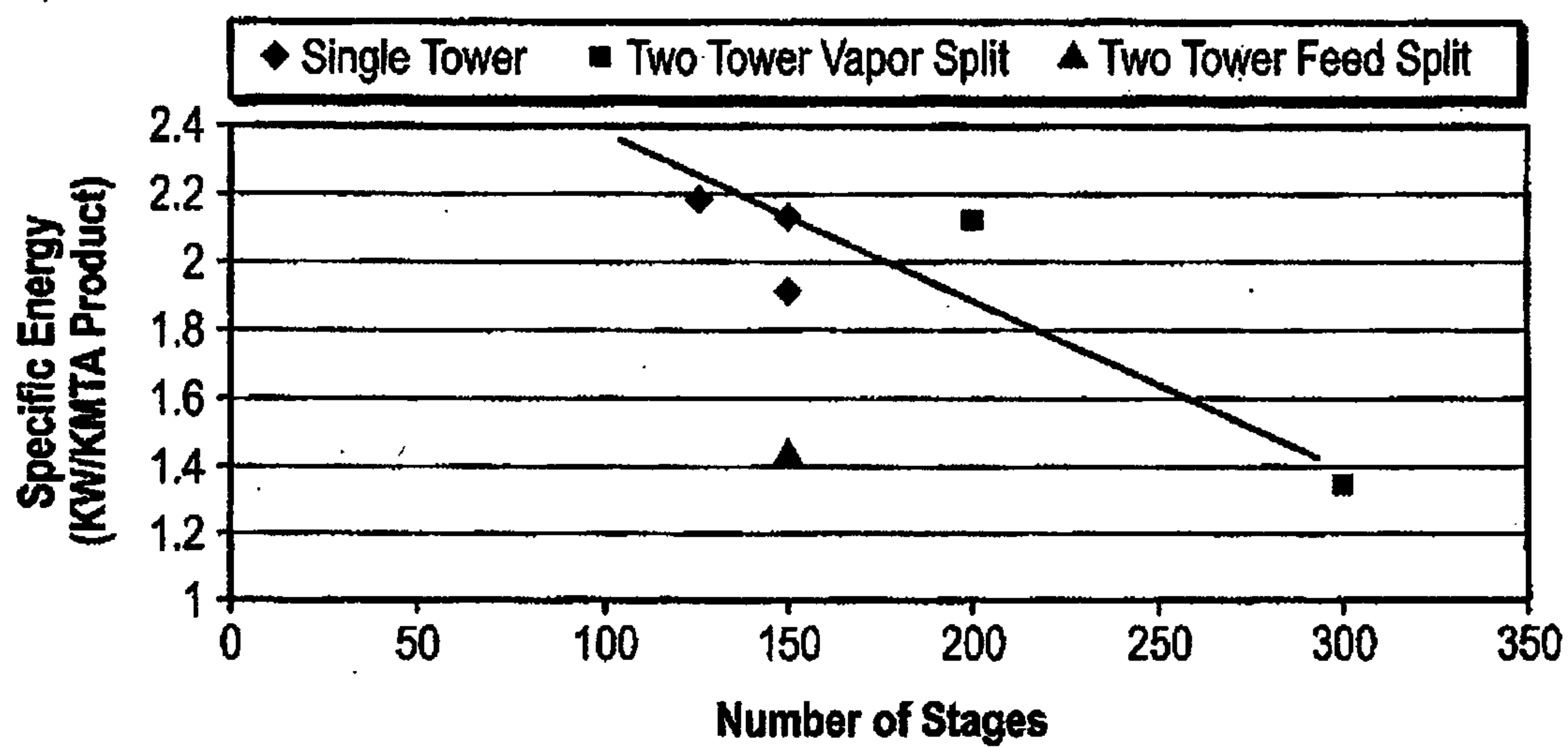


FIG. 3

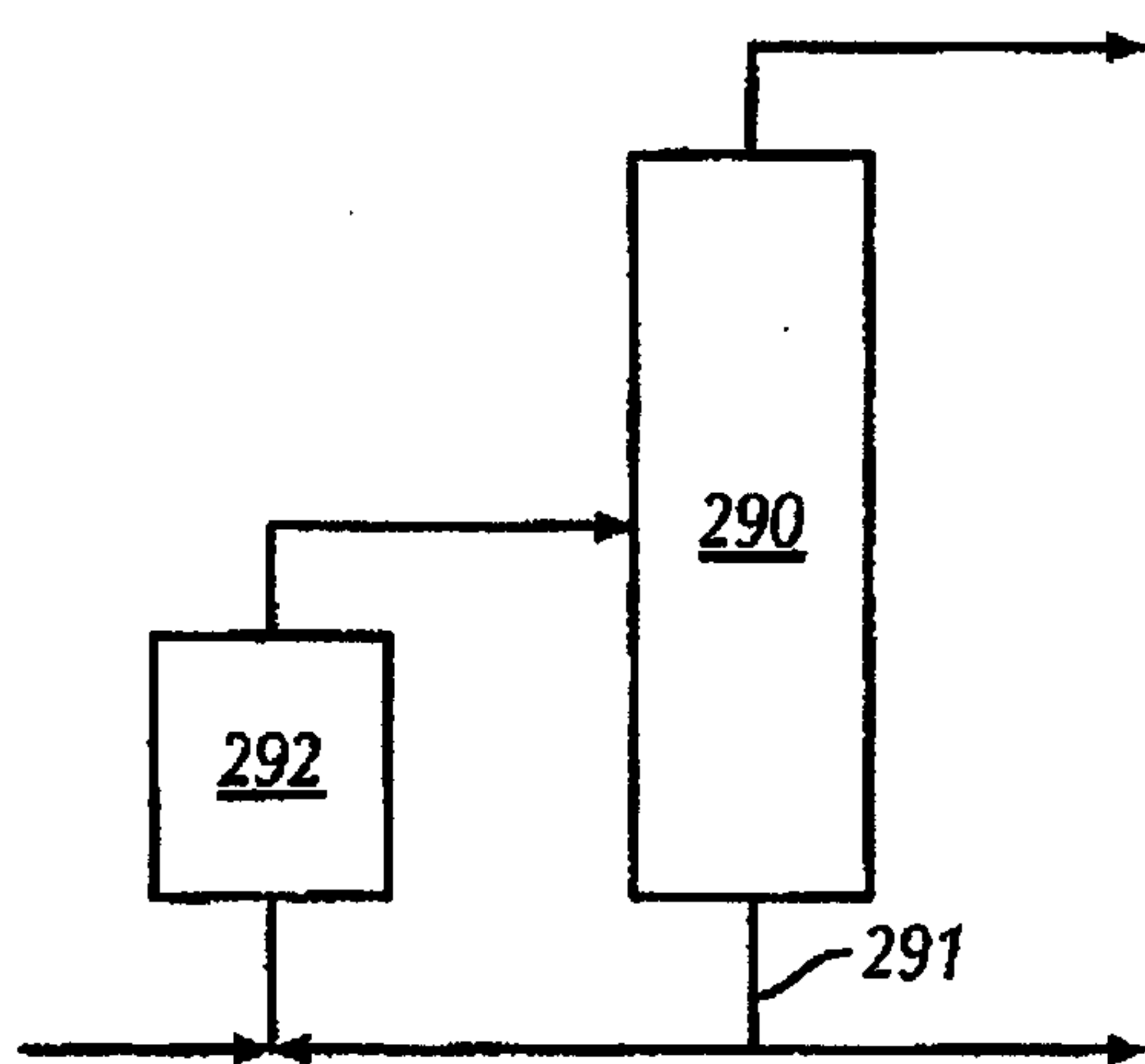


FIG. 6

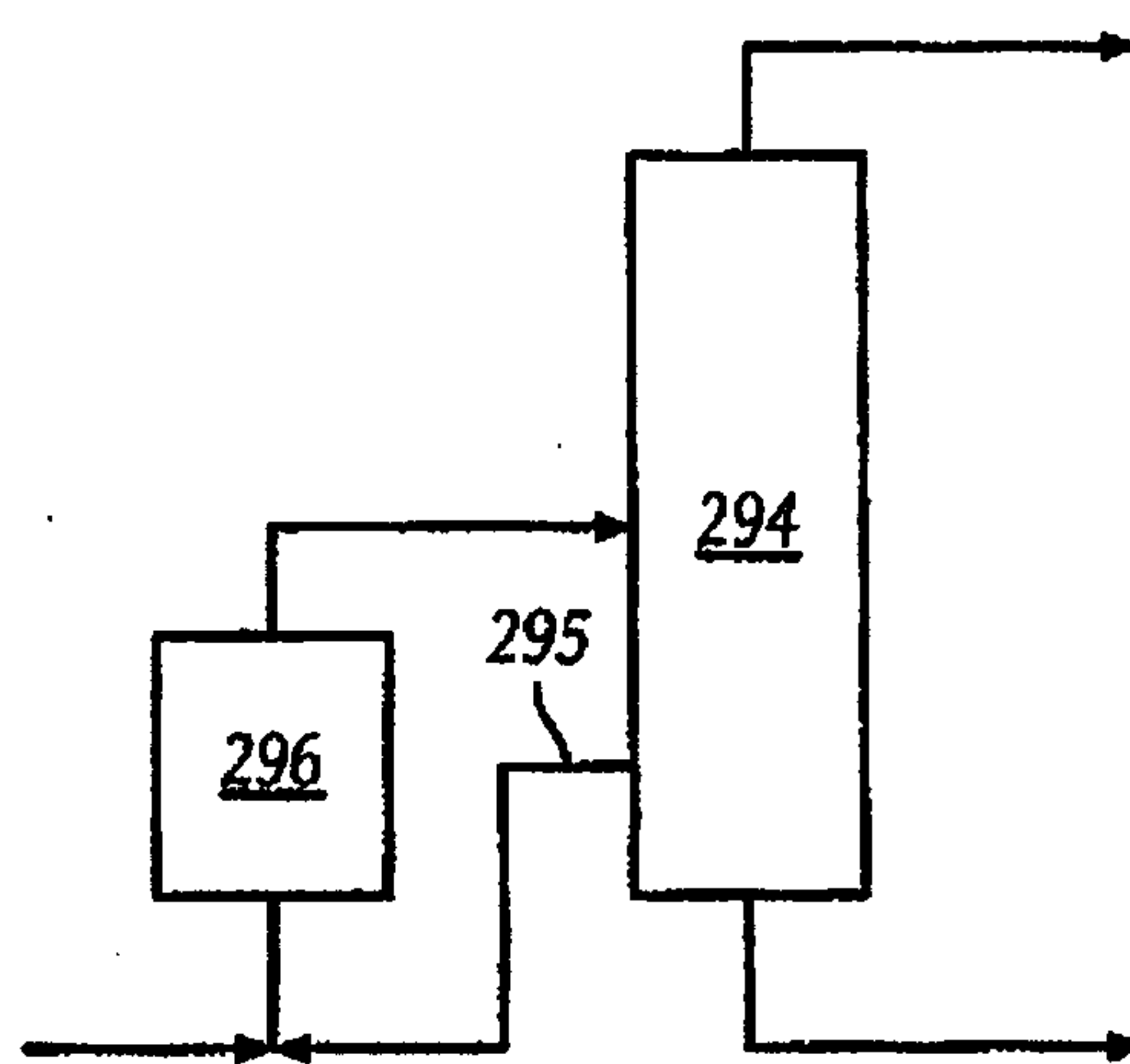
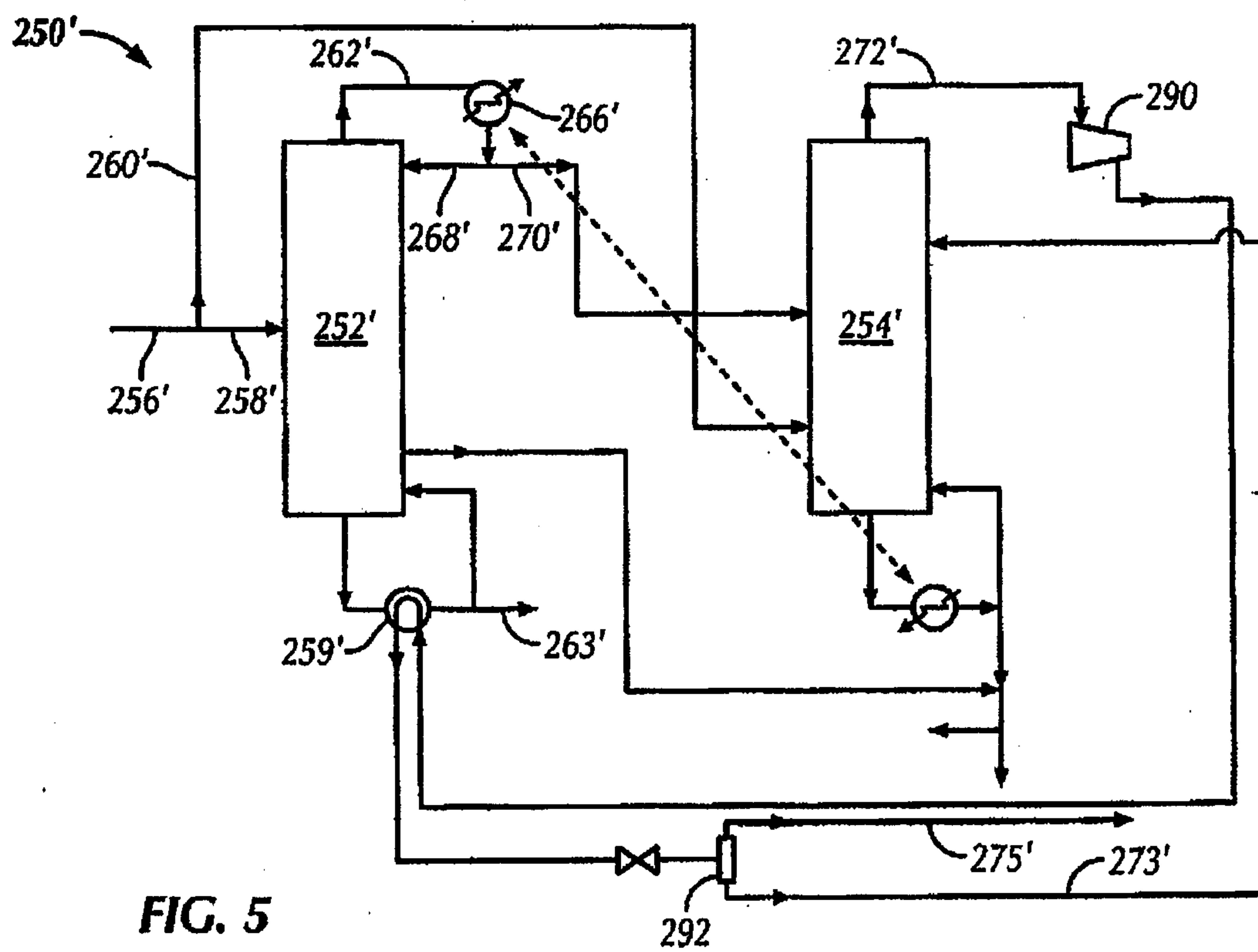
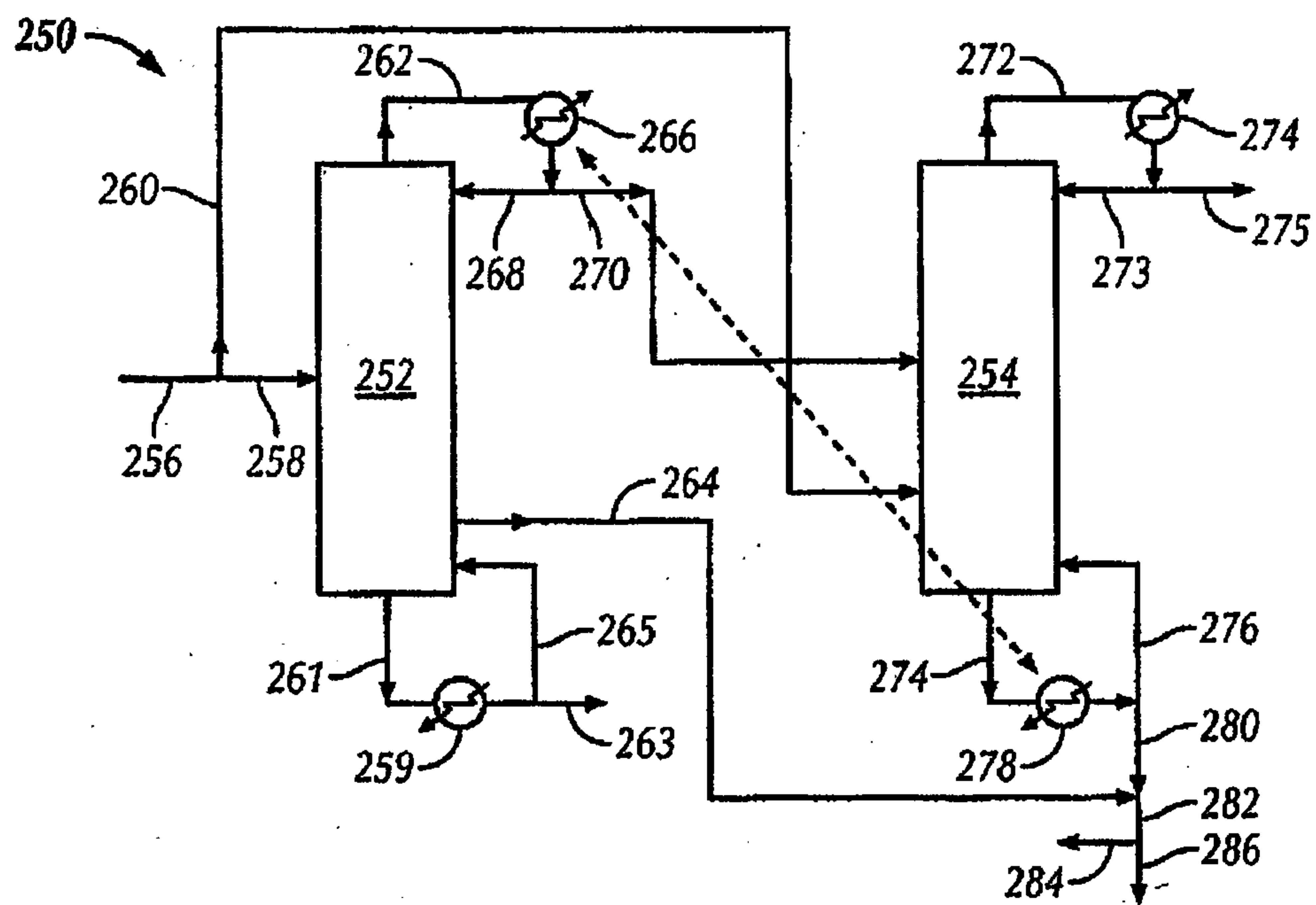


FIG. 7

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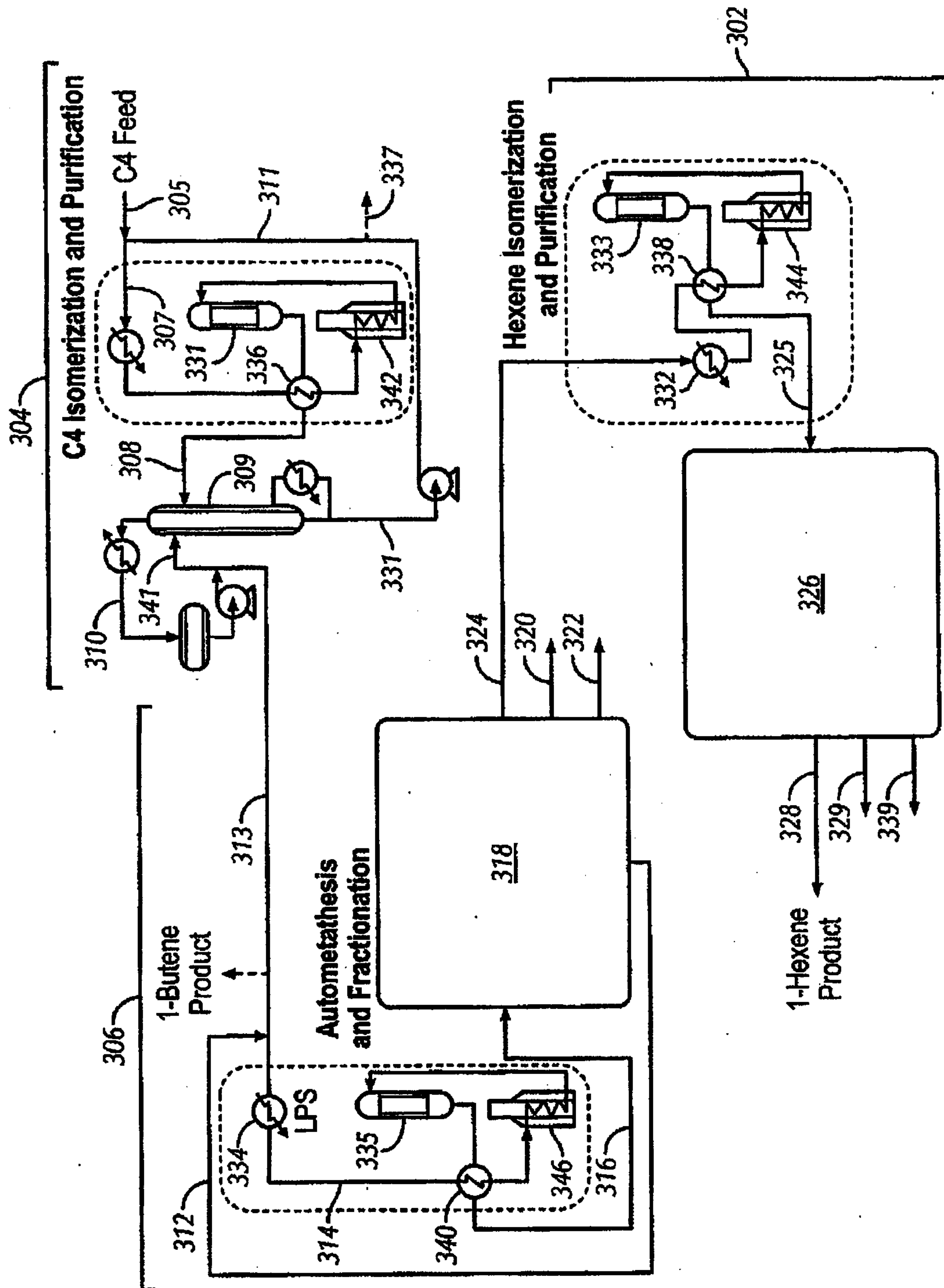


FIG. 8

