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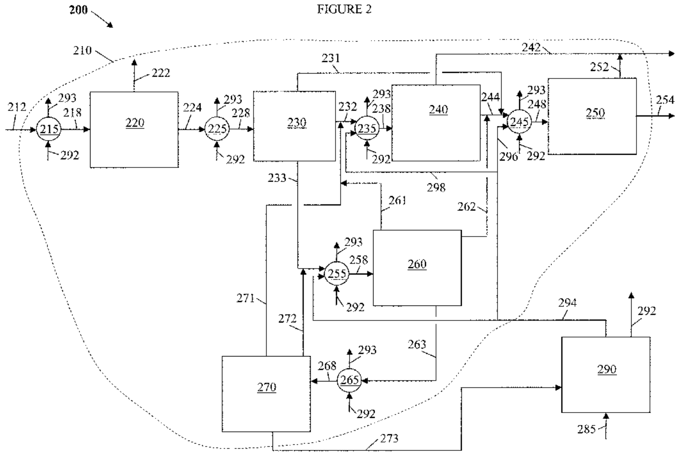
FIGURE 2



(57) Abstract: Disclosed is an integrated hydroprocessing and gasification system and a method tor operating the integrated hydroprocessing and gasification system. The integrated hydroprocessing and gasification system comprises an oil processing facility integrated with a gasification facility. The gasification facility uses cheaper coal/coke feed to produce a hydrogen stream and a high pressure steam stream. The hydrogen stream provides sufficient hydrogen for the oil processing facility, while the high pressure steam stream is used to pre-heat the streams entering the equipment of the oil processing facility in a steam heater. These steam heaters substantially eliminate emissions. Additionally, there is no heavy crude bottoms product formed in the oil processing facility because the bottoms product is fed into the gasification facility, according to some embodiments, the reactors used within the oil processing facility may be ebulating bed reactors, which use the heat of reaction to supply the required heat for operation.

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INTEGRATION OF GASIFICATION AND HYDROPROCESSING FOR LOW EMISSIONS REFINING

CROSS REFERENCE TO RELATED APPLICATIONS

[0001] The present application claims priority from U.S. Provisional Patent Application No. 61/111,263, entitled "Integration Of Gasification And Hydroprocessing For Low Emissions Refining" and filed on November 4, 2008.

TECHNICAL FIELD

[0002] This invention relates generally to methods and systems for integrating two or more industrial processes, and, more particularly, to methods and systems for integrating hydroprocessing and gasification processes to reduce emissions, reduce by-products, and increase refining efficiency.

BACKGROUND

[0003] Refineries currently use fired heaters, fired boilers, methane reformers, and catalytic crackers to refine crude oil. Each of these pieces of equipment burns natural gas or fuel oils to produce the necessary heat. This combustion typically consumes high cost natural gas and crude oil and produces emissions, including CO_2, SO_x, NO_x, VOCs, and particulate emissions. Also, refineries produce an asphalt stream, or coke stream, as a by-product to their oil refining process. Asphalt and coke streams are solid carbonaceous materials, at atmospheric conditions, that are produced in plentiful quantities, which thereby make its disposal costly and difficult for the refineries. Additionally, refineries require hydrogen during the refining process. This hydrogen is either purchased or manufactured separately in a hydrogen plant. In the hydrogen plant, high cost natural gas and steam are used to make hydrogen for use in the refining process. Since higher costs are involved with producing or purchasing hydrogen, there is typically not an abundance of available hydrogen to produce a high standard, cleaner burning fuel. Thus, less cleaner burning fuel is typically produced in the refineries.

[0004] Figure 1 shows a schematic depicting a typical refinery process 100 according to a prior art. As shown, the typical refinery process 100 consists of a crude unit 110, a hydrocracker unit 120, a coker unit 130, a hydrotreater unit 140, a hydrogen plant 150, and a sulfur plant 160. Each of the crude unit 110, the hydrocracker unit 120, the coker unit 130, and the hydrotreater unit 140 includes a fired heater for heating the fluid entering the respective unit. In the typical refinery process 100, a diluted crude stream 105 enters the crude unit 110. In the crude unit 110, the diluted crude stream 105 flows to a desalter (not shown) for removing contaminants such as water, salt, clay, and sand. Once the contaminants have been removed, the diluted crude stream 105 is further heated in a fired heater (not shown) and then flows into a crude column (not shown) for fractionation. Within the crude column (not shown), three streams, or fractions, are generated, a diluent stream 112, a light oil stream 114, and a heavy oil stream 116.

[0005] The diluent stream 112 consists of diluent, which is the lightest component. The diluent is added to the oil to make it easier to pump through the pipeline. Once the diluent has been removed from the oil, the diluent is sent back to the field, via the diluent stream 112, so that it may be added to more oil. The light oil stream 114, which makes up about one-third barrel of oil, consists of jet kerosene and gas oil and is sent to the hydrotreater unit 140, while the heavy oil stream 116, which makes up about two-thirds barrel of oil, consists of residuum and is sent to the hydrocracker unit 120.

[0006] Also, a natural gas stream 107 and a steam stream 109 enter the hydrogen plant 150, wherein hydrogen streams 152, 154 are generated. Within the hydrogen plant 150, the natural gas stream 107 and the steam stream 109 are mixed together and passed through a reformer furnace (not shown) so that the natural gas stream 107 reacts with the steam stream 109 to produce a reformer furnace outlet stream containing hydrogen, carbon monoxide, and carbon dioxide. The reformer furnace outlet stream then enters a shift converter (not shown) to remove the carbon monoxide from the reacted stream, thereby producing a shift converter outlet stream. The shift converter outlet stream then enters a steam methane reformer pressure swing absorption unit (not shown) to remove the remaining carbon monoxide, carbon dioxide, methane, and nitrogen, thereby producing a steam methane reformer pressure swing absorption unit outlet stream which is substantially pure hydrogen. The steam methane reformer pressure swing absorption unit outlet stream may then enter a hydrogen compressor, which compresses the pure hydrogen and sends it to the hydrocracker unit 120 and the hydrotreater unit 140 via hydrogen streams 152, 154. Approximately 1000 cubic feet of natural gas is consumed to produce one barrel of synthetic crude.

[0007] As shown in Figure 1, the heavy oil stream 116 is sent to the hydrocracker unit 120 where the hydrogen stream 152 is added at high temperatures and at high pressure. Within the hydrocracker unit 120, the heavy oil stream 116 is initially sent through a fired heater (not shown) where the heavy oil stream 116 is pre-heated. The pre-heated heavy oil stream is then combined with the hydrogen stream 152 and both are sent to a catalytic reactor (not shown) where cracking of the pre-heated heavy oil stream occurs. A catalytic reactor outlet stream is generated and flows into a series of separators (not shown) to separate the vapors from the liquids. The separated liquids are routed to a fractionator (not shown) for splitting the separated liquids into naphtha, jet kerosene, gas oil, and heavy residuum. The naphtha, jet kerosene, and gas oil are sent to the hydrotreater unit 140 for further cleaning via the hydrocracker unit light stream 122, while the heavy residuum is sent to the coker unit 130 for further cracking via the hydrocracker unit heavy stream 124.

[0008] The hydrocracker unit heavy stream 124 enters the coker unit 130 where it undergoes partial vaporization and mild cracking as it passes through a coking furnace (not shown). A coking furnace outlet stream, containing liquid and vapors, exits the coking furnace (not shown) and enters a coke drum (not shown). The vapors in the coke drum (not shown) undergo cracking as they pass through the coke drum (not shown) and exit the coke drum (not shown) via a coke drum vapor stream. The coke drum vapor stream is sent to a fractionation facility (not shown) where the coke drum vapor stream is separated into gas, naphtha, jet fuel, and gas oil and is sent to the hydrotreater unit 140 via a coker unit light stream 135. The petroleum coke within the coking furnace outlet stream remains in the coke drum (not shown). The heavy hydrocarbon liquid trapped in the coke drum (not shown) is subjected to successive cracking and polymerization until it is converted to vapors and coke. Typically, this coke may be sold as an end product via a coke stream 132.

[0009] The light oil stream 114 from the crude unit 110, the hydrocracker unit light stream 122 from the hydrocracker unit 120, and the coker unit light stream 135 from the coker unit 130 enter the hydrotreater unit 140 for further upgrading. Here, the hydrogen stream 154 is added at high temperatures and at high pressure to the light oil stream 114, the hydrocracker unit light stream 122, and the coker unit light stream 135. The hydrotreater unit 140 may contain a naphtha/jet hydrotreater (not shown) and a gas oil hydrotreater (not shown) for producing a product stream 145. The naphtha/jet hydrotreater (not shown) produces a jet fuel as its product stream 145, while the gas oil hydrotreater (not shown) produces a gas oil as its product stream 145. Each of the naphtha/jet hydrotreater (not shown) and the gas oil hydrotreater (not shown) have a fired heater (not shown) for receiving at least one of the light oil stream 114, the hydrocracker unit light stream 122, and the coker unit light stream 135. Upon being heated, streams 114, 122, 135 enter a catalytic reactor (not shown) where hydrogen sulfide and ammonia are produced. The catalytic reactor (not shown) produces a catalytic reactor outlet stream which enters a separator (not shown) where the hydrogen and other gases are separated from the liquid. The separator (not shown) produces a separator liquid outlet stream which then enters a stripper (not shown) to produce the product stream 145. The separator (not shown) also produces a separator vapor outlet stream 142, which contains hydrogen sulfide, that is sent to the sulfur plant 160. In the sulfur plant 160, hydrogen sulfide is converted to elemental sulfur using methods and processes known to those of ordinary skill in the art. The elemental sulfur may be removed from the typical refinery process 100 via a sulfur stream 162.

[0010] As illustrated above by the use of fired heaters, a large amount of heat is required for crude fractionation and pre-heating of the crude upstream of the reactors. This is especially true for heavier crudes which require significant processing to convert them into lighter material required for liquid fuels production. In addition, heavier crudes produce a high carbon, high sulfur, high metals residuum material that is typically combusted in a boiler. This combustion produces high CO_2, SO_x, NO_x, VOCs, and particulates. Further, the typical refinery process consumes expensive natural gas to produce the hydrogen required for the hydroprocessing.

Moreover, expensive equipment and costly maintenance are required to treat the by-products from the hydroprocessing.

[0011] In view of the foregoing discussion, need is apparent in the art for reducing pollutant emissions, including but not limited to NO_x, CO, and/or VOCs. Additionally, a need is apparent for providing alternative heating methods in lieu of using fired heaters, fired boilers, methane reformers, and catalytic crackers for the refining of heavy crude oil. Further, there exists the need for reducing the consumption of natural gas fuel during the refining of heavy crude oil. Moreover, there exists a need for eliminating the production of heavy crude bottoms products which cause secondary emissions from the combustion of these bottoms products. Furthermore, there exists a need for increasing refining efficiency thereby producing more barrels of fuel for each barrel of heavy crude. A technology addressing one or more such needs, or some other related shortcoming in the field, would benefit oil refining processes. This technology is included within the current invention.

SUMMARY

[0012] One exemplary embodiment of the invention is an integrated hydroprocessing and gasification system that includes an oil processing system and a gasification facility. The oil processing system receives an oil processing feed stream and produces a fuel/diesel outlet stream and a heavy product bottoms outlet stream. The gasification facility receives a gasification feed stream and produces a hydrogen outlet stream and a high pressure steam outlet stream. The gasification feed stream includes the heavy product bottoms outlet stream. The hydrogen outlet stream and the high pressure steam outlet stream are utilized within the oil processing system.

[0013] Another exemplary embodiment of the invention is a method for operating an integrated hydroprocessing and gasification system. The method includes providing an oil processing system and coupling a gasification facility to the oil processing system. The oil processing system receives an oil processing feed stream and produces a fuel/diesel outlet stream and a heavy product bottoms outlet stream. The gasification facility receives a gasification feed stream and produces a hydrogen outlet stream and a high pressure steam outlet stream. The gasification feed stream includes the heavy product bottoms outlet stream. The hydrogen outlet stream and the high pressure steam outlet stream are utilized within the oil processing system.

BRIEF DESCRIPTION OF THE DRAWINGS

[0014] The foregoing and other features and aspects of the invention may be best understood with reference to the following description of certain exemplary embodiments of the invention, when read in conjunction with the accompanying drawings, wherein:

[0015] Figure 1 shows a schematic depicting a typical refinery process according to a prior art;

[0016] Figure 2 shows a flowchart of an integrated hydroprocessing and gasification system that illustrates the equipment utilized in an oil processing facility in accordance with an exemplary embodiment;

[0017] Figure 3 shows a flowchart of an integrated hydroprocessing and gasification system that illustrates the equipment utilized in the gasification facility in accordance with an exemplary embodiment;

[0018] Figure 4A shows a bitumen yield graph comparing a prior art oil refining system bitumen crude yield versus an integrated hydroprocessing and gasification system bitumen crude yield in accordance with an exemplary embodiment;

[0019] Figure 4B shows a bitumen margin graph comparing a prior art oil refining system cost versus an integrated hydroprocessing and gasification system cost and comparing a prior art oil refining system bitumen margin versus an integrated hydroprocessing and gasification system bitumen margin in accordance with an exemplary embodiment;

[0020] Figure 5A shows a comparison between a prior art oil refining system Arab heavy crude yield flowchart versus an integrated hydroprocessing and gasification system Arab heavy crude yield flowchart in accordance with an exemplary embodiment;

[0021] Figure 5B shows an Arab heavy margin graph comparing a prior art oil refining system cost versus an integrated hydroprocessing and gasification system cost and comparing a prior art oil refining system Arab heavy gross margin versus an integrated hydroprocessing and gasification system Arab heavy gross margin in accordance with an exemplary embodiment;

[0022] Figure 6A shows a comparison between a prior art oil refining system UAE crude yield flowchart versus an integrated hydroprocessing and gasification system UAE crude yield flowchart in accordance with an exemplary embodiment; and

[0023] Figure 6B shows an UAE medium margin graph comparing a prior art oil refining system cost versus an integrated hydroprocessing and gasification system cost and comparing a prior art oil refining system UAE medium gross margin versus an integrated hydroprocessing and gasification system UAE medium gross margin in accordance with an exemplary embodiment.

[0024] The drawings illustrate only exemplary embodiments of the invention and are therefore not to be considered limiting of its scope, as the invention may admit to other equally effective embodiments.

BRIEF DESCRIPTION OF EXEMPLARY EMBODIMENTS

[0025] The application is directed to methods and systems for integrating two or more industrial processes. In particular, the application is directed to methods and systems for integrating hydroprocessing and gasification processes to reduce emissions, reduce by-products, and increase refining efficiency. Although the description of an exemplary embodiment of the invention is provided below in conjunction with a gasification process, alternative embodiments of the invention may be applicable to any industrial process capable of producing hydrogen and steam and utilizing by-products formed from hydroprocessing.

[0026] The invention may be better understood by reading the following description of non-limiting, exemplary embodiments with reference to the attached drawings, wherein like parts of each of the figures are identified by the same reference characters, and which are briefly described as follows.

[0027] Figure 2 shows a flowchart of an integrated hydroprocessing and gasification system 200 that illustrates the equipment utilized in an oil processing facility 210 in accordance with an exemplary embodiment. The integrated hydroprocessing and gasification system 200 comprises an oil processing facility 210 and a gasification facility 290. As shown, the oil processing facility 210 comprises a first distillation unit steam heater 215, a first distillation unit 220, a second distillation unit steam heater 225, a second distillation unit 230, a distillate upgrader steam heater 235, a distillate upgrader 240, a distillate finisher steam heater 245, a distillate finisher 250, a bottoms upgrader steam heater 255, a bottoms upgrader 260, a deasphalter steam heater 265, and a deasphalter 270. Although a steam heater is illustrated with respect to each of the equipment in the oil processing facility, other types of no emission heaters may be used without departing from the scope and spirit of the exemplary embodiment. Additionally, although a steam heater is illustrated with respect to each of the equipment in the oil processing facility, some equipment may not require a steam heater when the temperature of the incoming stream is adequate for proper fluid flow, for proper separation of the fluid components, and/or for assisting in the reaction without departing from the scope and spirit of the exemplary embodiment. Moreover, although specific equipment have been described as being part of the oil processing facility, some equipment may not be required or additional equipment may be added without departing from the scope and spirit of the exemplary embodiment. Although one having ordinary skill in the art would realize that there are many equipment and process and non-process lines in the oil processing facility, only certain key equipment and process and non-process lines relevant to the exemplary embodiment of the invention are illustrated in Figure 2. It is envisioned that any oil processing facility may be used in conjunction with the present embodiment.

[0028] As shown in Figure 2, a heavy feed and diluent stream 212 comprising heavy crude and diluent flows into the first distillation unit steam heater 215 and is heated via high pressure steam to a desired temperature to facilitate separation of the diluent from the heavy crude. The desired temperature may vary and is known to those of ordinary skill in the art. The heavy feed and diluent stream 212 may be sent from a storage tank (not shown) or directly from a pipeline (not shown). Some non-limiting examples of heavy crude include, but is not limited to, Bitumen crude, Arab heavy crude, and United Arab Emirates crude. The high pressure steam enters the first distillation unit steam heater 215 via a high pressure steam inlet stream 292 and exits the first distillation unit steam heater 215 via a high pressure steam outlet stream 293. The high pressure steam inlet stream 292 is produced in the gasification facility 290, which will be described below. A first distillation unit steam heater outlet stream 218 exits the first distillation unit steam heater 215 and enters the first distillation unit 220.

[0029] The first distillation unit 220 comprises a distillation column (not shown) and may include various other equipment (not shown) such as separators, drums, pumps, condensers, and reboilers. Once the first distillation unit steam heater outlet stream 218 enters the first distillation unit 220, the diluent is separated from the rest of the components within the first distillation unit steam heater outlet stream 218 and is recycled so that it may be added to additional incoming heavy crude. The diluent exits the first distillation unit 220 via a diluent outlet stream 222. The remaining components of the first distillation unit steam heater outlet stream 218 include light components and heavy components which exit the first distillation unit 220 and enter the second distillation unit steam heater 225 via a first distillation unit bottoms stream 224.

[0030] The first distillation unit bottoms stream 224 flows into the second distillation unit steam heater 225 and is heated via high pressure steam to a desired temperature to facilitate separation of the first distillation unit bottoms stream 224 into a light fraction, a heavy fraction, and a vac bottoms. The desired temperature may vary and is known to those of ordinary skill in the art. The high pressure steam enters the second distillation unit steam heater 225 via the high pressure steam inlet stream 292 and exits the second distillation unit steam heater 225 via the high pressure steam outlet stream 293. The high pressure steam inlet stream 292 is produced in the gasification facility 290, which will be described below. A second distillation unit steam heater outlet stream 228 exits the second distillation unit steam heater 225 and enters the second distillation unit 230.

[0031] The second distillation unit 230 comprises a distillation column (not shown) and may include various other equipment (not shown) such as separators, drums, pumps, condensers, and reboilers. Once the second distillation unit steam heater outlet stream 228 enters the second distillation unit 230, the second distillation unit steam heater outlet stream 228 is separated into various components, including a light fraction, a heavy fraction, and a vac bottoms. The light fraction exits the second distillation unit 230 via a second distillation unit light fraction outlet stream 231. The heavy fraction exits the second distillation unit 230 via a second distillation unit heavy fraction outlet stream 232. The vac bottoms exits the second distillation unit 230 via a second distillation unit vac bottoms outlet stream 233. The second distillation unit heavy fraction outlet stream 232 combines with a deasphalter light oil outlet stream 271 and a bottoms upgrader heavy oil outlet stream 261 and then enters the distillate upgrader steam heater 235. The deasphalter light oil outlet stream 271 and the bottoms upgrader heavy oil outlet stream 261 will be further described below. Although a first distillation unit 220 and a second distillation unit 230 are both used in the described embodiment, these units may be combined into one unit or separated into multiple units without departing from the scope and spirit of the exemplary embodiment.

[0032] The second distillation unit heavy fraction outlet stream 232, the deasphalter light oil outlet stream 271, and the bottoms upgrader heavy oil outlet stream 261 flow into the distillate upgrader steam heater 235 and is heated via high pressure steam to a desired temperature to facilitate separation of the streams 232, 271, 261 into light products and heavy products. Additionally, a distillate upgrader hydrogen inlet stream 298 also enters the distillate upgrader steam heater 235 for heating and mixes with streams 232, 271, 261 prior to entering the distillate upgrader 240. The desired temperature may vary and is known to those of ordinary skill in the art. The high pressure steam enters the distillate upgrader steam heater 235 via the high pressure steam inlet stream 292 and exits the distillate upgrader steam heater 235 via the high pressure steam outlet stream 293. The high pressure steam inlet stream 292 is produced in the gasification facility 290, which will be described below. A distillate upgrader steam heater outlet stream 238 exits the distillate upgrader steam heater 235 and enters the distillate upgrader 240.

[0033] The distillate upgrader 240 comprises a reactor/reactors (not shown) and may include various other equipment (not shown) such as distillation columns, separators, drums, pumps, condensers, and reboilers. According to some embodiments, the reactor may be combined with the distillation column. Once the distillate upgrader steam heater outlet stream 238 enters the distillate upgrader 240, the distillate upgrader steam heater outlet stream 238 reacts with the hydrogen and catalyst present within the reactor and is separated into a light product and a heavy product. During the reaction, many of the long hydrocarbon chains are broken into smaller hydrocarbon chains, many of the double and triple bonds between carbon molecules are broken into single bonded carbon molecules, and much of the sulfur in the hydrocarbon chains are removed by forming hydrogen sulfide. The light products exit the distillate upgrader 240 via a distillate upgrader light product outlet stream 242 and may be sent to an acid gas removal system (not shown) located within the gasification facility 290. The heavy products exit the distillate upgrader 240 via a distillate upgrader heavy product outlet stream 244. The distillate upgrader heavy product outlet stream 244 combines with the second distillation unit light fraction outlet stream 231 and a bottoms upgrader light oil outlet stream 262 and then enters the distillate finisher steam heater 245. The bottoms upgrader light oil outlet stream 262 will be further described below. According to one exemplary embodiment, the distillate upgrader 240 may comprise at least two reactors coupled together in series and an inter-stage vapor separation. These two reactors may be ebullated bed reactors. The use of ebulating bed hydroprocessing allows the heat of reaction to supply the required heat for the operation. This reduces the feed preheat requirements to a temperature level achievable by high pressure steam heaters. Although this embodiment shows that the reactors are ebullated bed reactors, the reactors may be any type of reactor, including but not limited to ebullated bed reactors, fixed bed reactors, or a combination of these types of reactors. Additionally, the catalyst used in these reactors may comprise hydrotrating extrudates or trilobes type which includes, but is not limited to, Criterion HDS-424, Criterion DN-200, or Axens HTS-358. Although some non-limiting examples have been provided for the catalysts, any type of catalyst used for treating crude oil may be used without departing from the scope and spirit of the exemplary embodiment.

[0034] The distillate upgrader heavy product outlet stream 244, the second distillation unit light fraction outlet stream 231, and the bottoms upgrader light oil outlet stream 262 flow into the distillate finisher steam heater 245 and is heated via high pressure steam to a desired temperature to facilitate separation of the streams 244, 231, 262 into light products and diesel/jet fuel. Additionally, a distillate finisher hydrogen inlet stream 296 also enters the distillate finisher steam heater 245 for heating and mixes with streams 244, 231, 262 prior to entering the distillate finisher 250. The desired temperature may vary and is known to those of ordinary skill in the art. The high pressure steam enters the distillate finisher steam heater 245 via the high pressure steam inlet stream 292 and exits the distillate finisher steam heater 245 via the high pressure steam outlet stream 293. The high pressure steam inlet stream 292 is produced in the gasification facility 290, which will be described below. A distillate finisher steam heater outlet stream 248 exits the distillate finisher steam heater 245 and enters the distillate finisher 250.

[0035] The distillate finisher 250 comprises a reactor/reactors (not shown) and may include various other equipment (not shown) such as distillation columns, separators, drums, pumps, condensers, and reboilers. According to some embodiments, the reactor may be combined with the distillation column. Once the distillate finisher steam heater outlet stream 248 enters the distillate finisher 250, the distillate finisher steam heater outlet stream 248 reacts with the hydrogen and catalyst present within the reactor and is separated into a light product and a diesel/jet fuel. During the reaction, the remaining longer hydrocarbon chains are broken into smaller hydrocarbon chains, any remaining double and triple bonds between carbon molecules are broken into single bonded carbon molecules, and any remaining sulfur in the hydrocarbon chains is removed by forming hydrogen sulfide. The light products exit the distillate finisher 250 via a distillate finisher light product outlet stream 252 and may be combined with the distillate upgrader light product outlet stream 242 before being sent to the acid gas removal system (not shown) located within the gasification facility 290. The diesel/jet fuel exits the distillate finisher 250 via a diesel/jet fuel stream 254 and may then be collected, stored, and sold. According to one exemplary embodiment, the distillate finisher 250 may comprise at least two reactors coupled together in series and an inter-stage vapor separation. These two reactors may be ebullated bed reactors. The use of ebulating bed hydroprocessing allows the heat of reaction to supply the required heat for the operation. This reduces the feed preheat requirements to a temperature level achievable by high pressure steam heaters. Although this embodiment shows that the reactors are ebullated bed reactors, the reactors may be any type of reactor, including but not limited to ebullated bed reactors, fixed bed reactors, or a combination of these types of reactors. Additionally, the catalyst used in these reactors may comprise hydrotrating extrudates or trilobes type which includes, but is not limited to, Criterion HDS-424, Criterion DN-200, or Axens HTS-358. Although some non-limiting examples have been provided for the catalysts, any type of catalyst used for treating crude oil may be used without departing from the scope and spirit of the exemplary embodiment.

[0036] The second distillation unit vac bottoms outlet stream 233 and a deasphalter heavy oil outlet stream 272 flow into the bottoms upgrader steam heater 255 and is heated via high pressure steam to a desired temperature to facilitate separation of the streams 233, 272 into a light oil, a heavy oil, and a vac bottoms. The deasphalter heavy oil outlet stream 272 will be further described below. Additionally, a bottoms upgrader hydrogen inlet stream 294 also enters the bottoms upgrader steam heater 255 for heating and mixes with streams 233, 272 prior to entering the bottoms upgrader 260. The desired temperature may vary and is known to those of ordinary skill in the art. The high pressure steam enters the bottoms upgrader steam heater 255 via the high pressure steam inlet stream 292 and exits the bottoms upgrader steam heater 255 via the high pressure steam outlet stream 293. The high pressure steam inlet stream 292 is produced in the gasification facility 290, which will be described below. A bottoms upgrader steam heater outlet stream 258 exits the bottoms upgrader steam heater 255 and enters the bottoms upgrader 260.

[0037] The bottoms upgrader 260 comprises a reactor/reactors (not shown) and may include various other equipment (not shown) such as distillation columns, separators, drums, pumps, condensers, and reboilers. According to some embodiments, the reactor may be combined with the distillation column. Once the bottoms upgrader steam heater outlet stream 258 enters the bottoms upgrader 260, the bottoms upgrader steam heater outlet stream 258 reacts with the hydrogen and catalyst present within the reactor and is separated into a light oil, a heavy oil, and a vac bottoms. During the reaction, many of the remaining long hydrocarbon chains are broken into smaller hydrocarbon chains, many of the remaining double and triple bonds between carbon molecules are broken into single bonded carbon molecules, and much of the remaining sulfur in the hydrocarbon chains are removed by forming hydrogen sulfide. The light oil exits the bottoms upgrader 260 via the bottoms upgrader light oil outlet stream 262 and is combined with the distillate upgrader heavy product outlet stream 244 and the second distillation unit light fraction outlet stream 231 prior to entering the distillate finisher steam heater 245, as described above. The heavy oil exits the bottoms upgrader 260 via the bottoms upgrader heavy oil outlet stream 261 and is combined with the second distillation unit heavy fraction outlet stream 232 and the deasphalter light oil outlet stream 271 prior to entering the distillate upgrader steam heater 235, as also described above. The vac bottoms exits the bottoms upgrader 260 via a bottoms upgrader vac bottoms outlet stream 263, which flows to the deasphalter 270 via the deasphalter steam heater 265. According to one exemplary embodiment, the bottoms upgrader 260 may comprise at least two reactors coupled together in series and an inter-stage vapor separation. These two reactors may be ebullated bed reactors. The use of ebulating bed hydroprocessing allows the heat of reaction to supply the required heat for the operation. This reduces the feed preheat requirements to a temperature level achievable by high pressure steam heaters. Although this embodiment shows that the reactors are ebullated bed reactors, the reactors may be any type of reactor, including but not limited to ebullated bed reactors, fixed bed reactors, or a combination of these types of reactors. Additionally, the catalyst used in these reactors may comprise sediment control extrudates type which includes, but is not limited to, Criterion TEX-2710, Criterion TEX-2910, Criterion TEX-2710N, or Criterion TEX-2731. Although some non-limiting examples have been provided for the catalysts, any type of catalyst used for treating crude oil may be used without departing from the scope and spirit of the exemplary embodiment.

[0038] The bottoms upgrader vac bottoms outlet stream 263 flows into the deasphalter steam heater 265 and is heated via high pressure steam to a desired temperature. It is then mixed with an extracting solvent (not shown) in an extractor (not shown). The extractor temperature is controlled to facilitate separation of the bottoms upgrader vac bottoms outlet stream 263 into a light deasphalted oil, a heavy deasphalted oil, and an asphalt. The desired temperature may vary and is known to those of ordinary skill in the art. The high pressure steam enters the deasphalter steam heater 265 via the high pressure steam inlet stream 292 and exits the deasphalter steam heater 265 via the high pressure steam outlet stream 293. The high pressure steam inlet stream 292 is produced in the gasification facility 290, which will be described below. A deasphalter steam heater outlet stream 268 exits the deasphalter steam heater 265 and enters the deasphalter 270.

[0039] The deasphalter 270 comprises an extractor (not shown) and may include various other equipment (not shown) such as distillation columns, separators, drums, pumps, condensers, and reboilers. According to some embodiments, the extractor may be combined with the distillation column. Once the deasphalter steam heater outlet stream 268 contacts with the solvent present within the extractor and is separated into a light deasphalted [DA] oil, a heavy deasphalted [DA] oil, and an asphalt. The light DA oil exits the deasphalter 270 via the deasphalter light DA oil outlet stream 271 and is combined with the second distillation unit heavy fraction outlet stream 232 and the bottoms upgrader heavy oil outlet stream 261 prior to entering the distillate upgrader steam heater 235, as described above. The heavy DA oil exits the deasphalter 270 via the deasphalter heavy DA oil outlet stream 272 and is combined with the second distillation unit vac bottoms outlet stream 233 prior to entering the bottoms upgrader steam heater 255, as also described above. The asphalt exits the deasphalter 270 via an asphalt stream 273, which flows as one of the feed sources to the gasification facility 290. According to one exemplary embodiment, the deasphalter 270 may comprise at least one or two extractors. The extracting solvents may include, but are not limited to, C_3, C_4, C_5, C_6, or C_7.

[0040] The gasification facility 290 may be of any type of gasification facility that is capable of producing hydrogen and high pressure steam. As shown in Figure 2, the asphalt stream 273 and a coke stream 285 is fed into one or more gasifiers (not shown) in the gasification facility 290. The gasification facility 290 processes the asphalt stream 273 and the coke stream 285 to produce at least the distillate upgrader hydrogen inlet stream 298, the distillate finisher hydrogen inlet stream 296, the bottoms upgrader hydrogen inlet stream 294, and the high pressure steam inlet stream 292. The gasification facility 290 also comprises an acid gas removal system (not shown) that processes the distillate upgrader light product outlet stream 242 and the distillate finisher light product outlet stream 252, which both contain hydrogen sulfide. Additionally, the gasification facility may produce synthetic natural gas ("SNG") and CO_2, which may be used for enhanced oil recovery.

[0041] Figure 3 shows a flowchart of an integrated hydroprocessing and gasification system 200 that illustrates the equipment utilized in the gasification facility 290 in accordance with an exemplary embodiment. The integrated hydroprocessing and gasification system 200 comprises an oil processing facility 210 and a gasification facility 290. As shown, the gasification facility 290 comprises an air separation unit 310, at least one gasifier 320, 330, 340, 350, 360, 370, a first stream treating unit 380, and a second stream treating unit 390. Although specific equipment have been described as being part of the gasification facility, some equipment may not be required or additional equipment may be added without departing from the scope and spirit of the exemplary embodiment. Also, although one having ordinary skill in the art would realize that there are many equipment and process and non-process lines in the gasification facility, only certain key equipment and process and non-process lines relevant to the exemplary embodiment of the invention are illustrated in Figure 3. It is envisioned that any gasification facility capable of producing high pressure steam and hydrogen may be used in conjunction with the present embodiment.

[0042] As shown in Figure 3, the air separation unit 310 (3x40% or 2x60%) separates air into at least a nitrogen ("N_2") component and an oxygen ("O_2") component. At least some of the O_2 component exits the air separation unit 310 through an air separation unit outlet stream 311. Upon exiting the air separation unit 310, the air separation unit outlet stream enters an O_2 supply header 312, which serves to distribute the O_2 to each of the at least one gasifier 320, 330, 340, 350, 360, 370. The use of the oversized air separation unit 310 may result in an increase in overall production of SNG, hydrogen, and/or steam. Although exemplary capacities of the air separation unit 310 have been provided, alternative capacities may be used without departing from the scope and spirit of the exemplary embodiment.

[0043] The at least one gasifier 320, 330, 340, 350, 360, 370 comprises a first gasifier 320 (1x25%), a second gasifier 330 (1x25%), a third gasifier 340 (1x25%), a fourth gasifier 350 (1x25%), a fifth gasifier 360 (1x10%), and a sixth gasifier 370 (1x10%). According to the embodiment shown in Figure 3, the coke stream 285 is fed into each of the first gasifier 320, the second gasifier 330, the third gasifier 340, and the fourth gasifier 350, while the asphalt stream 273, 304 from the oil processing facility 210 is fed into each of the fifth gasifier 360 and the sixth gasifier 370. Each of the at least one gasifier 320, 330, 340, 350, 360, 370 produces a slag stream (not shown) and a gasifier outlet stream 322, 332, 342, 352, 362, 372. The slag stream (not shown) may comprise metals naturally occurring in the coke stream 285 and the asphalt stream 273, 304, and added minerals to control the melting point of the slag stream (not shown). The slag stream (not shown) may be utilized as an aggregate in concrete manufacturing and/or the manufacturing of other materials. The gasifier outlet stream 322, 332, 342, 352, 362, 372 comprises about 35% CO, about 15% hydrogen ("H_2"), about 40% water ("H_2O"), and about 10% carbon dioxide ("CO_2"). The conversion of the coke stream 285, the asphalt stream 273, 304, and the air separation unit outlet stream 311 into the slag stream (not shown) and the gasifier outlet stream 322, 332, 342, 352, 362, 372 is an exothermic process and as a result, a high pressure saturated steam stream (not shown) also is produced, which may be used in a methanation unit 383. Although specific compositions have been provided for the gasifier outlet stream 322, 332, 342, 352, 362, 372, alternative compositions may be achieved without departing from the scope and spirit of the exemplary embodiment.

[0044] Although one exemplary configuration of gasifiers and feed streams have been illustrated and described, alternative configurations may be utilized without departing from the scope and spirit of the exemplary embodiment. For example, the coke stream and the asphalt stream may both be fed into the same gasifier without departing from the scope and spirit of the exemplary embodiment.

[0045] The gasifier outlet stream 322, 332, 342, 352, 362, 372 exits the at least one gasifier 320, 330, 340, 350, 360, 370 and flows into a gasifier outlet distribution header 374. A first distribution header outlet stream 376 exits the gasifier outlet distribution header 374 and flows into the first stream treating unit 380 (2x50%), which may be designed to process anyone of the gasifier outlet stream 322, 332, 342, 352, 362, 372 so that SNG is produced . A second distribution header outlet stream 378 exits the gasifier outlet distribution header 374 and flows into the second stream treating unit 390 (1x100%), which may be designed to process anyone of the gasifier outlet stream 322, 332, 342, 352, 362, 372 so that hydrogen for the oil processing facility 210 is produced.

[0046] The first stream treating unit 380 comprises an acid gas removal unit ("AGR") 381, a sulfur recovery unit ("SRU") 382, and a methanation unit ("MU") 383. The first distribution header outlet stream 376 flows into the AGR 381. The AGR 381 may utilize Selexol™ for hydrogen sulfide ("H_2S") removal and CO_2 capture. As a result, a vapor CO_2 stream 389 is produced. The vapor CO_2 stream 389 may be compressed into a liquid CO_2, which may then be utilized for enhanced oil recovery by pumping the liquid CO_2 into the ground to increase the production of oil. The CO_2 may be used in other areas of the facility, for example, during startup processes or even sold.

[0047] The AGR 381 also produces an acid gas stream (not shown), which enters the SRU 382 to produce a sulfur stream 388 and a recycle tail gas stream (not shown). The sulfur stream 388 comprises sulfur and may be sold to fertilizer plants and the like. The recycle tail gas stream (not shown) comprises some sulfur and may be recycled back into the AGR 381.

[0048] The AGR 381 also produces an acid gas removal system outlet stream (not shown) comprising mainly of CO and H_2. Since the CO_2 has been removed within the AGR 381, the acid gas removal system outlet stream (not shown) comprises about 25% CO and about 75% H_2. In certain embodiments, the acid gas removal system outlet stream (not shown) may be sold as syngas to market or consumed by other systems requiring the syngas (not shown). The syngas may be used as ammonia, methanol, or hydrogen, or be utilized in the production of power or chemicals. Although specific compositions have been provided for the acid gas removal system outlet stream (not shown), alternative compositions may be achieved without departing from the scope and spirit of the exemplary embodiment.

[0049] The acid gas removal system outlet stream (not shown) may then enter a MU 383, which includes one or more methanation reactors (not shown). The MU 383 converts the acid gas removal system outlet stream (not shown) into a methanation unit outlet stream 387. The methanation unit outlet stream 387 comprises SNG and may be sold to market or consumed by other systems requiring the syngas. In certain alternative embodiments, a portion of the methanation unit outlet stream 387 may enter combustion turbines (not shown) to produce power to be sold to market.

[0050] As previously mentioned, the high pressure saturated steam stream (not shown) from the at least one gasifier 320, 330, 340, 350, 360, 370 may enter the MU 383. The conversion of the acid gas removal system outlet stream (not shown) into the methanation unit outlet stream 387 in the MU 383 is an exothermic reaction and as a result, the high pressure saturated steam stream (not shown) is converted to a high pressure superheated steam stream 292, which may be utilized in at least the steam heaters of the oil processing facility 210 and also in a steam turbine (not shown) to produce power to be sold to market or consumed by other systems requiring the power.

[0051] The second stream treating unit 390 comprises an acid gas removal unit ("AGR") 391 and a pressure swing absorber ("PSA") 392. The second distribution header outlet stream 378 flows into the AGR 391. Also, the distillate upgrader light product outlet stream 242, which includes the distillate finisher light product outlet stream 252, exits the oil processing facility 210 and enters the AGR 391 for treating. The AGR 391 may utilize Selexol™ for hydrogen sulfide ("H_2S") removal and CO_2 capture. As a result, a vapor CO_2 stream 394 is produced. The vapor CO_2 stream 394 may be compressed into a liquid CO_2, which may then be utilized for enhanced oil recovery by pumping the liquid CO_2 into the ground to increase the production of oil. The CO_2 may be used in other areas of the facility, for example, during startup processes or even sold.

[0052] The AGR 391 also produces an acid gas stream 398, which enters the SRU 382 of the first stream treating unit 380 to contribute to the production of the sulfur stream 388, as described above. The AGR 391 also produces an acid gas removal system outlet stream (not shown) comprising mainly of CO and H_2. The acid gas removal system outlet stream (not shown) may then enter the PSA 392, which separates the gases and produces a high purity hydrogen stream that exits the PSA 392 via the bottoms upgrader hydrogen inlet stream 294, which includes the distillate upgrader hydrogen inlet stream 298 and the distillate finisher hydrogen inlet stream

296 and is utilized in at least the distillate upgrader 240 (Figure 2), the distillate finisher 250 (Figure 2), and the bottoms upgrader 260 (Figure 2) of the oil processing facility 210. The PSA 392 also produces a tail gas stream 396, which may be recycled back into a shift reactor system (not shown) that may be located between the at least one gasifier 320, 330, 340, 350, 360, 370 and the first stream treating unit 380. This tail gas stream 396 is eventually converted to SNG, which thereby eliminates the need to combust this tail gas stream 396.

[0053] The oil processing facility 210 may be of any type of oil processing facility that is capable of utilizing at least hydrogen and/or high pressure steam produced by the gasification facility 290. As shown in Figure 3, the heavy feed and diluent stream 212, the high pressure superheated steam stream 292, and the bottoms upgrader hydrogen inlet stream 294, which includes the distillate upgrader hydrogen inlet stream 298 and the distillate finisher hydrogen inlet stream 296, are fed into the oil processing facility 210. The oil processing facility 210 processes the heavy feed and diluent stream 212 to produce at least the asphalt stream 273, the distillate upgrader light product outlet stream 242, which includes the distillate finisher light product outlet stream 252, the diluent outlet stream 222, and the diesel/jet fuel stream 254.

[0054] Although the embodiment described above illustrates a gasification facility having both the first stream treating unit and the second stream treating unit, the gasification facility may have an alternate steam treating unit having at least an AGR, a SRU, and a PSA without departing from the scope and spirit of the exemplary embodiment. Additionally, although a separate AGR is described for each of the first stream treating unit and the second stream treating unit, there may be only one AGR for both the first stream treating unit and the second stream treating unit without departing from the scope and spirit of the exemplary embodiment. Further, although the embodiment described above illustrates a gasification facility having at least one gasifier capable of being fed both the coke stream and the asphalt stream, the gasification facility may have gasifiers designed to be fed only the asphalt stream without departing from the scope and spirit of the exemplary embodiment. Moreover, although the above embodiment illustrates that the high pressure superheated steam stream generated from the gasification facility is used in the steam heaters of the oil processing facility, any type of high pressure steam, including saturated steam, generated from the gasification facility may be used in the steam heaters of the oil processing facility without departing from the scope and spirit of the exemplary embodiment.

[0055] According to another exemplary embodiment of the oil processing facility 210, the distillate upgrader 240 may comprise at least two reactors coupled together in series and an inter-stage vapor separation. These two reactors may be ebullated bed reactors. The use of ebulating bed hydroprocessing allows the heat of reaction to supply the required heat for the operation. This reduces the feed preheat requirements to a temperature level achievable by high pressure steam heaters. Although this embodiment shows that the reactors are ebullated bed reactors, the reactors may be any type of reactor, including but not limited to ebullated bed reactors, fixed bed reactors, or a combination of these types of reactors. Additionally, the catalyst used in the first stage reactor may comprise hydroconversion extrudates type which includes, but is not limited to, Criterion HDS-2443B or TEX-2910. The catalyst used in the second stage reactor may comprise hydrotrating extrudates or trilobes type which includes, but is not limited to, Criterion HDS-424, Criterion DN-200, or Axens HTS-358. Although some non-limiting examples have been provided for the catalysts for each of the reactors, any type of catalyst used for treating crude oil may be used without departing from the scope and spirit of the exemplary embodiment. Also, according to this exemplary embodiment, the bottoms upgrader 260 may comprise at least two reactors coupled together in series and an inter-stage vapor separation. These two reactors may be ebullated bed reactors. The use of ebulating bed hydroprocessing allows the heat of reaction to supply the required heat for the operation. This reduces the feed preheat requirements to a temperature level achievable by high pressure steam heaters. Although this embodiment shows that the reactors are ebullated bed reactors, the reactors may be any type of reactor, including but not limited to ebullated bed reactors, fixed bed reactors, or a combination of these types of reactors. Additionally, the catalyst used in these reactors may comprise sediment control extrudates type which includes, but is not limited to, Criterion TEX-2710, Criterion TEX-2910, Criterion TEX-2710N, or Criterion TEX-2731. Although some non-limiting examples have been provided for the catalysts of these reactors, any type of catalyst used for treating crude oil may be used without departing from the scope and spirit of the exemplary embodiment. The reactors of the distillate finisher 250 may be configured similarly to the reactors of the distillate upgrader 240. Similarly, the extractors of the deasphalter 270 may be configured similarly to the reactors of the bottoms upgrader 260. In the stream configuration for this exemplary embodiment, the deasphalter light DA oil outlet stream 271 and the deasphalter heavy DA oil outlet stream 272 are now fed into the first stage reactor of the distillate upgrader 240 and a bottoms stream (not shown) from the distillate upgrader 240 may now be fed into the reactor of the bottoms upgrader 260. This configuration may increase the total crude rate to the refinery.

[0056] Figure 4A shows a bitumen yield graph 400 comparing a prior art oil refining system bitumen crude yield 402 versus an integrated hydroprocessing and gasification system bitumen crude yield 404 in accordance with an exemplary embodiment. The bitumen yield graph 400 illustrates the prior art oil refining system bitumen crude yield 402 and the integrated hydroprocessing and gasification system bitumen crude yield 404 on the x-axis and a volumetric yield percentage 406 on the y-axis.

[0057] According to the prior art oil refining system bitumen crude yield 402, the volumetric yield percentage, which comprises a distillate yield 410 and a naphtha and light products gas yield 420, is about 85%. The distillate yield 410 comprises diesel fuel, jet fuel, and other valuable fuels and has about 40% volumetric yield. The naphtha and light products gas yield 420 has about 45% volumetric yield. However, according to the integrated hydroprocessing and gasification system bitumen crude yield 404 in accordance with an exemplary embodiment, the volumetric yield percentage, which comprises a distillate yield 430 and a naphtha and light products gas yield 440, is about 120%. The distillate yield 430 comprises diesel fuel, jet fuel, and other valuable fuels and has about 85% volumetric yield. The naphtha and light products gas yield 440 has about 35% volumetric yield. The distillate yield 410, 430 has a margin of about $15-$25/barrel, while the naphtha and light products gas yield 420, 440 has a margin of about $0-$5/barrel. Thus, the profitability margin substantially increases because greater volumetric yield percentages of higher value fuels are produced using the integrated hydroprocessing and gasification system in accordance with an exemplary embodiment versus the prior art oil refining system.

[0058] Figure 4B shows a bitumen margin graph 450 comparing a prior art oil refining system cost 451 versus an integrated hydroprocessing and gasification system cost 453 and comparing a prior art oil refining system bitumen margin 452 versus an integrated hydroprocessing and gasification system bitumen margin 454 in accordance with an exemplary embodiment. The information illustrated within the bitumen margin graph 450 is for refining only. The bitumen margin graph 450 illustrates the prior art oil refining system cost 451, the prior art oil refining system bitumen gross margin 452, the integrated hydroprocessing and gasification system cost 453, and the integrated hydroprocessing and gasification system bitumen gross margin 454 on the x-axis and a dollar/barrel value 456 on the y-axis.

[0059] According to the prior art oil refining system cost 451, the total cost, which comprises a capital recovery cost 460, an operating and maintenance cost 462, a fuel cost 464, and a transportation cost 466, is about $28 per barrel. The capital recovery cost 460 is about $7 per barrel. The operating and maintenance cost 462 is about $2 per barrel. The fuel cost 464 is about $1 per barrel. The transportation cost 466 is about $18 per barrel. However, according to the integrated hydroprocessing and gasification system cost 453 in accordance with an exemplary embodiment, the total cost, which comprises a capital recovery cost 480, an operating and maintenance cost 482, a fuel cost 484, and a transportation cost 486, is about $26 per barrel. The capital recovery cost 480 is about $14 per barrel. The operating and maintenance cost 482 is about $3 per barrel. The fuel cost 484 is about $1 per barrel. The transportation cost 486 is about $8 per barrel. Thus, the total cost for refining oil is slightly less when utilizing the integrated hydroprocessing and gasification system in accordance with an exemplary embodiment versus the prior art oil refining system.

[0060] Also shown in Figure 4B, according to the prior art oil refining system bitumen gross margin 452, the bitumen crude margin 470 is about $30 per barrel. However, according to the integrated hydroprocessing and gasification system bitumen gross margin 454 in accordance with an exemplary embodiment, the bitumen crude margin 490 is about $72 per barrel. Thus, the gross margin for refining oil is substantially higher when utilizing the integrated hydroprocessing and gasification system in accordance with an exemplary embodiment versus the prior art oil refining system.

[0061] Figure 5A shows a comparison between a prior art oil refining system Arab heavy crude yield flowchart 500 versus an integrated hydroprocessing and gasification system Arab heavy crude yield flowchart 520 in accordance with an exemplary embodiment. The prior art oil refining system Arab heavy crude yield flowchart 500 illustrates an Arab heavy crude stream 504 and a natural gas stream 506 entering a prior art oil refining system 510. The prior art oil refining system 510 processes the Arab heavy crude stream 504 and the natural gas stream 506 and produces a coke stream 512, a distillates stream 514, and a light product gas and gasoline stream 516. According to one embodiment, the flowrate of the Arab heavy crude stream 504 may be about 400 million barrels per day ("MBD") and the flowrate of the natural gas stream 506 may be about 130 million standard cubic feet per day ("MMSCFD"). These feed flowrates produce the coke stream 512 having a flowrate of about 5000 standard tons per day (sTPD), the distillates stream 514 having a flowrate of about 190 MBD, and the light product gas and gasoline stream 516 having a flowrate of about 190 MBD. This prior art refining system 510 produces about a 95% yield when having the Arab heavy crude stream 504 enter the prior art refining system 510. Although an exemplary flowrate has been provided for each of these streams, alternative stream flowrates may be produced without departing from the scope and spirit of the exemplary embodiment.

[0062] The integrated hydroprocessing and gasification system Arab heavy crude yield flowchart 520 illustrates an Arab heavy crude stream 524 and a coke stream 526 entering an integrated hydroprocessing and gasification system 530 in accordance with an exemplary embodiment. The integrated hydroprocessing and gasification system 530 processes the Arab heavy crude stream 524 and the coke stream 526 and produces a distillates stream 534 and a light product gas and gasoline stream 536. According to one embodiment, the flowrate of the Arab heavy crude stream 524 may be about 400 MBD and the flowrate of the coke stream 526 may be about 10,000 sTPD. These feed flowrates produce the distillates stream 534 having a flowrate of about 275 MBD and the light product gas and gasoline stream 536 having a flowrate of about 170 MBD. This integrated hydroprocessing and gasification system 530 produces about a 110% yield when having the Arab heavy crude stream 524 enter the integrated hydroprocessing and gasification system 530. Although an exemplary flowrate has been provided for each of these streams, alternative stream flowrates may be produced without departing from the scope and spirit of the exemplary embodiment. As seen in Figure 5A, the percent yield substantially increases when utilizing the integrated hydroprocessing and gasification system 530 in accordance with an exemplary embodiment than when utilizing the prior art oil refining system 510. This yield increases because byproducts are not formed in the integrated hydroprocessing and gasification system 530 and hydrogen is not produced in the integrated hydroprocessing and gasification system 530 by using valuable natural gas feed streams.

[0063] Figure 5B shows an Arab heavy margin graph 550 comparing a prior art oil refining system cost 551 versus an integrated hydroprocessing and gasification system cost 553 and comparing a prior art oil refining system Arab heavy gross margin 552 versus an integrated hydroprocessing and gasification system Arab heavy gross margin 554 in accordance with an exemplary embodiment. The Arab heavy margin graph 450 illustrates the prior art oil refining system cost 551, the prior art oil refining system Arab heavy gross margin 552, the integrated hydroprocessing and gasification system cost 553, and the integrated hydroprocessing and gasification system Arab heavy gross margin 554 on the x-axis and a dollar/barrel value 556 on the y-axis.

[0064] According to the prior art oil refining system cost 551, the total cost, which comprises a capital recovery cost 560, an operating and maintenance cost 562, and a fuel cost 564, is about $19 per barrel. The capital recovery cost 560 is about $12 per barrel. The operating and maintenance cost 562 is about $2 per barrel. The fuel cost 564 is about $5 per barrel. However, according to the integrated hydroprocessing and gasification system cost 553 in accordance with an exemplary embodiment, the total cost, which comprises a capital recovery cost 580, an operating and maintenance cost 582, and a fuel cost 584, is about $16 per barrel. The capital recovery cost 580 is about $12 per barrel. The operating and maintenance cost 582 is about $3 per barrel. The fuel cost 584 is about $1 per barrel. Thus, the total cost for refining oil is slightly less when utilizing the integrated hydroprocessing and gasification system in accordance with an exemplary embodiment versus the prior art oil refining system.

[0065] Also shown in Figure 5B, according to the prior art oil refining system Arab heavy gross margin 552, the total margin, which comprises an Arab heavy margin 570, is about $22 per barrel. However, according to the integrated hydroprocessing and gasification system Arab heavy gross margin 554 in accordance with an exemplary embodiment, the total margin, which comprises an Arab heavy margin 590 and a CO_2 and N_2 margin 592, is about $46 per barrel. The Arab heavy margin 590 is about $43 per barrel. The CO_2 and N_2 margin 592 is about $3 per barrel. Thus, the gross margin for refining oil is substantially higher when utilizing the integrated hydroprocessing and gasification system in accordance with an exemplary embodiment versus the prior art oil refining system.

[0066] Figure 6A shows a comparison between a prior art oil refining system UAE crude yield flowchart 600 versus an integrated hydroprocessing and gasification system UAE crude yield flowchart 620 in accordance with an exemplary embodiment. The prior art oil refining system Arab heavy crude yield flowchart 600 illustrates an UAE crude stream 604 and a natural gas stream 606 entering a prior art oil refining system 610. The prior art oil refining system 610 processes the UAE crude stream 604 and the natural gas stream 606 and produces a coke stream 612, a distillates stream 614, and a light product gas and gasoline stream 616. According to one embodiment, the flowrate of the UAE crude stream 604 may be about 500 MBD and the flowrate of the natural gas stream 606 may be about 150 MMSCFD. These feed flowrates produce the coke stream 612 having a flowrate of about 5000 sTPD, the distillates stream 614 having a flowrate of about 260 MBD, and the light product gas and gasoline stream 616 having a flowrate of about 240 MBD. This prior art refining system 610 produces about a 98% yield when having the UAE crude stream 604 enter the prior art refining system 610. Although an exemplary flowrate has been provided for each of these streams, alternative stream flowrates may be produced without departing from the scope and spirit of the exemplary embodiment.

[0067] The integrated hydroprocessing and gasification system UAE crude yield flowchart 620 illustrates an UAE crude stream 624 and a coke stream 626 entering an integrated hydroprocessing and gasification system 630 in accordance with an exemplary embodiment. The integrated hydroprocessing and gasification system 630 processes the UAE crude stream 624 and the coke stream 626 and produces a distillates stream 634 and a light product gas and gasoline stream 636. According to one embodiment, the flowrate of the UAE crude stream 624 may be about 500 MBD and the flowrate of the coke stream 626 may be about 10,000 sTPD. These feed flowrates produce the distillates stream 634 having a flowrate of about 340 MBD and the light product gas and gasoline stream 636 having a flowrate of about 205 MBD. This integrated hydroprocessing and gasification system 630 produces about a 109% yield when having the UAE crude stream 624 enter the integrated hydroprocessing and gasification system 630. Although an exemplary flowrate has been provided for each of these streams, alternative stream flowrates may be produced without departing from the scope and spirit of the exemplary embodiment. As seen in Figure 6A, the percent yield substantially increases when utilizing the integrated hydroprocessing and gasification system 630 in accordance with an exemplary embodiment than when utilizing the prior art oil refining system 610. This yield increases because byproducts are not formed in the integrated hydroprocessing and gasification system 630 and hydrogen is not produced in the integrated hydroprocessing and gasification system 630 by using valuable natural gas feed streams.

[0068] Figure 6B shows an UAE medium margin graph 650 comparing a prior art oil refining system cost 651 versus an integrated hydroprocessing and gasification system cost 653 and comparing a prior art oil refining system UAE medium gross margin 652 versus an integrated hydroprocessing and gasification system UAE medium gross margin 654 in accordance with an exemplary embodiment. The UAE medium margin graph 650 illustrates the prior art oil refining system cost 651, the prior art oil refining system UAE medium gross margin 652, the integrated hydroprocessing and gasification system cost 653, and the integrated hydroprocessing and gasification system UAE medium gross margin 654 on the x-axis and a dollar/barrel value 656 on the y-axis.

[0069] According to the prior art oil refining system cost 651, the total cost, which comprises a capital recovery cost 660, an operating and maintenance cost 662, and a fuel cost 664, is about $18 per barrel. The capital recovery cost 660 is about

$11 per barrel. The operating and maintenance cost 662 is about $2 per barrel. The fuel cost 664 is about $5 per barrel. However, according to the integrated hydroprocessing and gasification system cost 653 in accordance with an exemplary embodiment, the total cost, which comprises a capital recovery cost 680, an operating and maintenance cost 682, and a fuel cost 684, is about $15 per barrel. The capital recovery cost 680 is about $12 per barrel. The operating and maintenance cost 682 is about $2 per barrel. The fuel cost 684 is about $1 per barrel. Thus, the total cost for refining oil is slightly less when utilizing the integrated hydroprocessing and gasification system in accordance with an exemplary embodiment versus the prior art oil refining system.

[0070] Also shown in Figure 6B, according to the prior art oil refining system UAE medium gross margin 652, the total margin, which comprises an UAE medium margin 670, is about $20 per barrel. However, according to the integrated hydroprocessing and gasification system UAE medium gross margin 654 in accordance with an exemplary embodiment, the total margin, which comprises an UAE medium margin 690 and a CO_2 and N_2 margin 692, is about $38 per barrel. The UAE medium margin 690 is about $36 per barrel. The CO_2 and N_2 margin 692 is about $2 per barrel. Thus, the gross margin for refining oil is substantially higher when utilizing the integrated hydroprocessing and gasification system in accordance with an exemplary embodiment versus the prior art oil refining system.

[0071] In summary, when using the integrated hydroprocessing and gasification system, emissions of CO, NO_x, VOCs, and particulates are substantially eliminated. There are still fugitive emissions, but no stack emissions. Also, the consumption of more valuable natural gas fuel is substantially reduced. Instead of hydrogen gas being generated from natural gas in a methane reformer, the hydrogen may be generated from less valuable coal or coke within the gasification facility and used to refine the heavy crude oil. The heavy crude oil, like bitumen, may be entirely hydroprocessed using the hydrogen and steam from a gasification facility, without the need of blending the heavy crude with better crudes. The gasification facility produces much more hydrogen gas in an efficient manner than in a methane reformer. The hydrogen gas is used in the hydroprocessing of the heavy crude oil to replace the carbon double and triple bonds, the carbon rings, and the single carbon bonds when the longer carbon chains are broken down into smaller carbon chains. This process, as described within one of the embodiments of the present invention, thereby increases the cetane number and makes the fuel cleaner burning and more environmentally friendly than the processes used in the prior art. Thus, in the integrated hydroprocessing and gasification system, there are no fired heaters, fired boilers, methane reformers, or catalytic crackers used to refine the heavy crude oil; instead, there are high pressure steam heaters and an integrated gasification facility in accordance with some of the embodiments.

[0072] Also, when using the integrated hydroprocessing and gasification system, there are no heavy crude bottoms products, or asphalt streams, that are produced, which thereby eliminates secondary emissions. The heavy crude bottoms product is sent to the gasification facility as a feed stream. Typically, in prior art processes, these heavy crude bottoms product, or asphalt streams, contain sulfur and metals and the combustion of these heavy crude products typically produce more NO_x, SO_x, VOCs, and particulates than the cleaner distillate fuels. Additionally, these heavy crude products must be consumed in a lower efficiency engine which increases the relative CO_x emissions.

[0073] Further, when using the integrated hydroprocessing and gasification system, higher volumetric refinery yields are achieved, even when processing heavy crudes. These higher volumetric yields are due to the hydrogen being produced from an external low value fuel source, which is added to the crude during refining. This hydrogen addition increases refining efficiency, thereby producing more barrels of fuel for each barrel of heavy crude. Hence, by refining crude oil using the integrated hydroprocessing and gasification system, the crude reserves are effectively increased.

[0074] Although the invention has been described with reference to specific embodiments, these descriptions are not meant to be construed in a limiting sense. Various modifications of the disclosed embodiments, as well as alternative embodiments of the invention will become apparent to persons of ordinary skill in the art upon reference to the description of the invention. It should be appreciated by those of ordinary skill in the art that the conception and the specific embodiments disclosed may be readily utilized as a basis for modifying or designing other structures or methods for carrying out the same purposes of the invention. It should also be realized by those of ordinary skill in the art that such equivalent constructions do not depart from the spirit and scope of the invention as set forth in the appended claims. It is therefore, contemplated that the claims will cover any such modifications or embodiments that fall within the scope of the invention.

WHAT IS CLAIMED IS:

1. An integrated hydroprocessing and gasification system, comprising:

an oil processing system producing a fuel/diesel outlet stream and a heavy product bottoms outlet stream from an oil processing feed stream,

a gasification facility producing a hydrogen outlet stream and a high pressure steam outlet stream from a gasification feed stream,

wherein the gasification feed stream comprises the heavy product bottoms outlet stream, and

wherein the hydrogen outlet stream and the high pressure steam outlet stream are utilized within the oil processing system.

2. The integrated hydroprocessing and gasification system of claim 1, wherein the high pressure steam stream is a superheated steam stream.

3. The integrated hydroprocessing and gasification system of claim 1, wherein the oil processing feed stream is a heavy crude.

4. The integrated hydroprocessing and gasification system of claim 3, wherein the heavy crude is selected from the group consisting of a bitumen heavy crude, an Arab heavy crude, and an UAE heavy crude.

5. The integrated hydroprocessing and gasification system of claim 1, wherein the oil processing system comprises a first distillation unit, a second distillation unit, a distillate upgrader, a distillate finisher, a bottoms upgrader, and a deasphalter.

6. The integrated hydroprocessing and gasification system of claim 5, wherein the oil processing system further comprises at least one steam heater coupled to each of the first distillation unit, the second distillation unit, the distillate upgrader, the distillate finisher, the bottoms upgrader, and the deasphalter, wherein at least one steam heater utilizes the high pressure steam produced from the gasification facility.

7. The integrated hydroprocessing and gasification system of claim 5, wherein the distillate upgrader, the distillate finisher, and the bottoms upgrader utilize the hydrogen stream produced from the gasification facility.

8. The integrated hydroprocessing and gasification system of claim 5, wherein the bottoms upgrader comprises a plurality of ebullated bed reactors having a catalyst within the plurality of ebullated bed reactors and an interstage vapor separation.

9. The integrated hydroprocessing and gasification system of claim 8, wherein at least two of the plurality of ebullated bed reactors are connected in series.

10. The integrated hydroprocessing and gasification system of claim 8, wherein the catalyst comprises sediment control extrudates type.

11. The integrated hydroprocessing and gasification system of claim 5, wherein the distillate upgrader comprises a plurality of ebullated bed reactors having a catalyst within the plurality of ebullated bed reactors and an interstage vapor separation.

12. The integrated hydroprocessing and gasification system of claim 11, wherein at least two of the plurality of ebullated bed reactors are connected in series.

13. The integrated hydroprocessing and gasification system of claim 11, wherein the catalyst is selected from at least the group consisting of a hydrotrating extrudates type or trilobes type.

14. The integrated hydroprocessing and gasification system of claim 11, wherein the catalyst of one of the plurality of ebullated bed reactors comprises hydroconversion extrudates type and the catalyst of another of the plurality of ebullated bed reactors is selected from at least the group consisting of a hydrotrating extrudates type or trilobes type.

15. The integrated hydroprocessing and gasification system of claim 1, wherein the gasification facility comprises an air separation unit, at least one gasifier, an acid gas removal unit, a sulfur recovery unit, and a pressure swing absorber.

16. The integrated hydroprocessing and gasification system of claim 15, wherein the gasification facility further comprises a methanation unit.

17. The integrated hydroprocessing and gasification system of claim 15, wherein the pressure swing absorber generates a tail gas, the tail gas being converted into a synthetic natural gas.

18. The integrated hydroprocessing and gasification system of claim 1, wherein the volumetric yield percentage is greater than 100%.

19. A method for operating an integrated hydroprocessing and gasification system, comprising:

providing an oil processing system, the oil processing system producing a fuel/diesel outlet stream and a heavy product bottoms outlet stream from an oil processing feed stream,

coupling a gasification facility to the oil processing system, the gasification facility producing a hydrogen outlet stream and a high pressure steam outlet stream from a gasification feed stream,

wherein the gasification feed stream comprises the heavy product bottoms outlet stream, and

wherein the hydrogen outlet stream and the high pressure steam outlet stream are utilized within the oil processing system.

20. The method of claim 19, wherein the high pressure steam stream is a superheated steam stream.

21. The method of claim 19, wherein the oil processing feed stream is a heavy crude.

22. The method of claim 21, wherein the heavy crude is selected from the group consisting of a bitumen heavy crude, an Arab heavy crude, and an UAE heavy crude.

23. The method of claim 19, wherein the oil processing system comprises a first distillation unit, a second distillation unit, a distillate upgrader, a distillate finisher, a bottoms upgrader, and a deasphalter.

24. The method of claim 23, wherein the oil processing system further comprises at least one steam heater coupled to each of the first distillation unit, the

second distillation unit, the distillate upgrader, the distillate finisher, the bottoms upgrader, and the deasphalter, wherein the at least one steam heater utilizes the high pressure steam produced from the gasification facility.

25. The method of claim 23, wherein the distillate upgrader, the distillate finisher, and the bottoms upgrader utilize the hydrogen stream produced from the gasification facility.

26. The method of claim 23, wherein the bottoms upgrader comprises a plurality of ebullated bed reactors having a catalyst within the plurality of ebullated bed reactors and an interstage vapor separation.

27. The method of claim 26, wherein at least two of the plurality of ebullated bed reactors are connected in series.

28. The method of claim 26, wherein the catalyst comprises sediment control extrudates type.

29. The method of claim 23, wherein the distillate upgrader comprises a plurality of ebullated bed reactors having a catalyst within the plurality of ebullated bed reactors and an interstage vapor separation.

30. The method of claim 27, wherein at least two of the plurality of ebullated bed reactors are connected in series.

31. The method of claim 27, wherein the catalyst is selected from at least the group consisting of a hydrotrating extrudates type or trilobes type.

32. The method of claim 27, wherein the catalyst of one of the plurality of ebullated bed reactors comprises hydroconversion extrudates type and the catalyst of another of the plurality of ebullated bed reactors is selected from at least the group consisting of a hydrotrating extrudates type or trilobes type.

33. The method of claim 19, wherein the gasification facility comprises an air separation unit, at least one gasifier, an acid gas removal unit, a sulfur recovery unit, and a pressure swing absorber.

34. The method of claim 33, wherein the gasification facility further comprises a methanation unit.

35. The method of claim 33, wherein the pressure swing absorber generates a tail gas, the tail gas being converted into a synthetic natural gas.

36. The method of claim 19, wherein the volumetric yield percentage is greater than 100%.

FIGURE 1

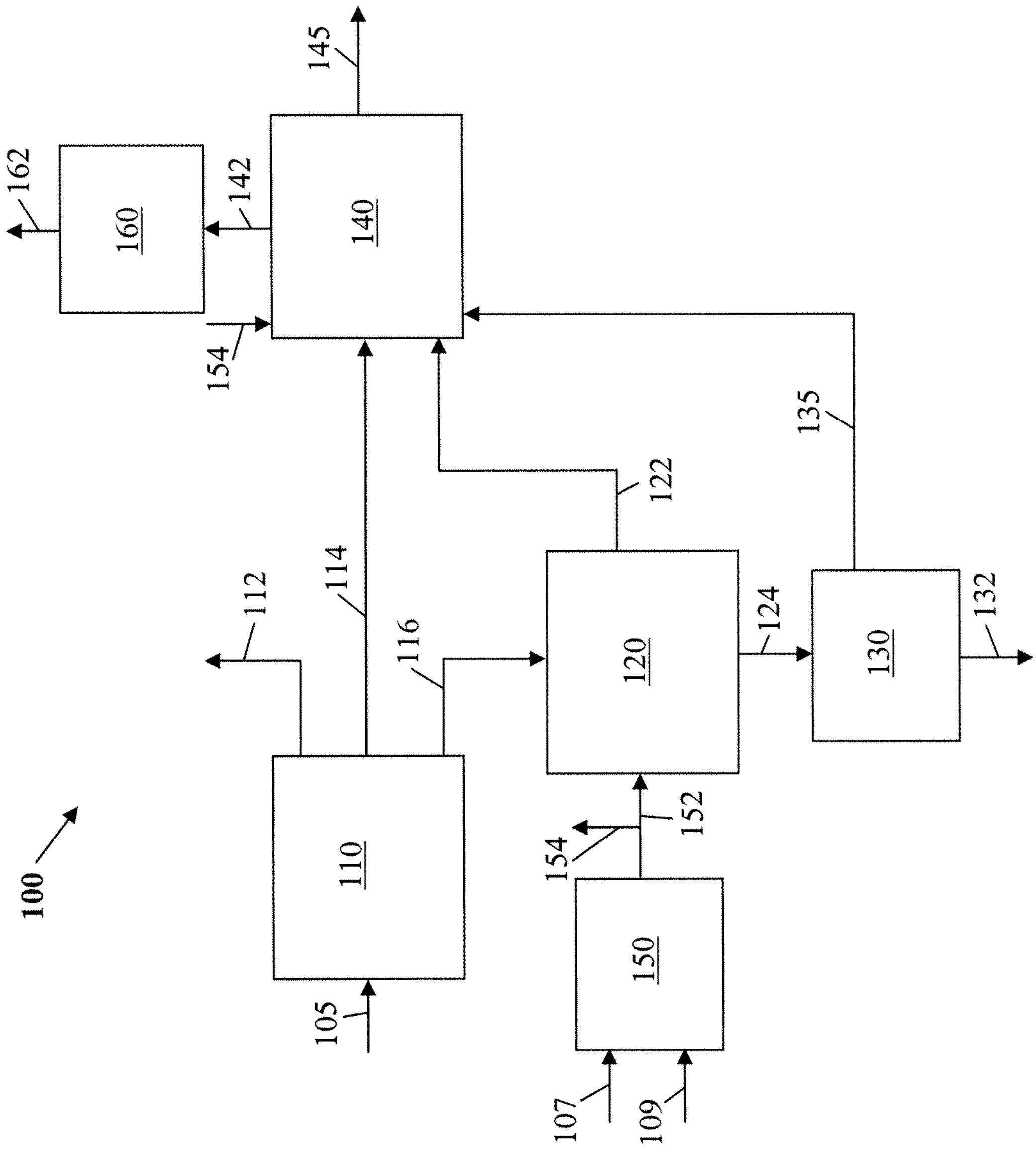
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FIGURE 2

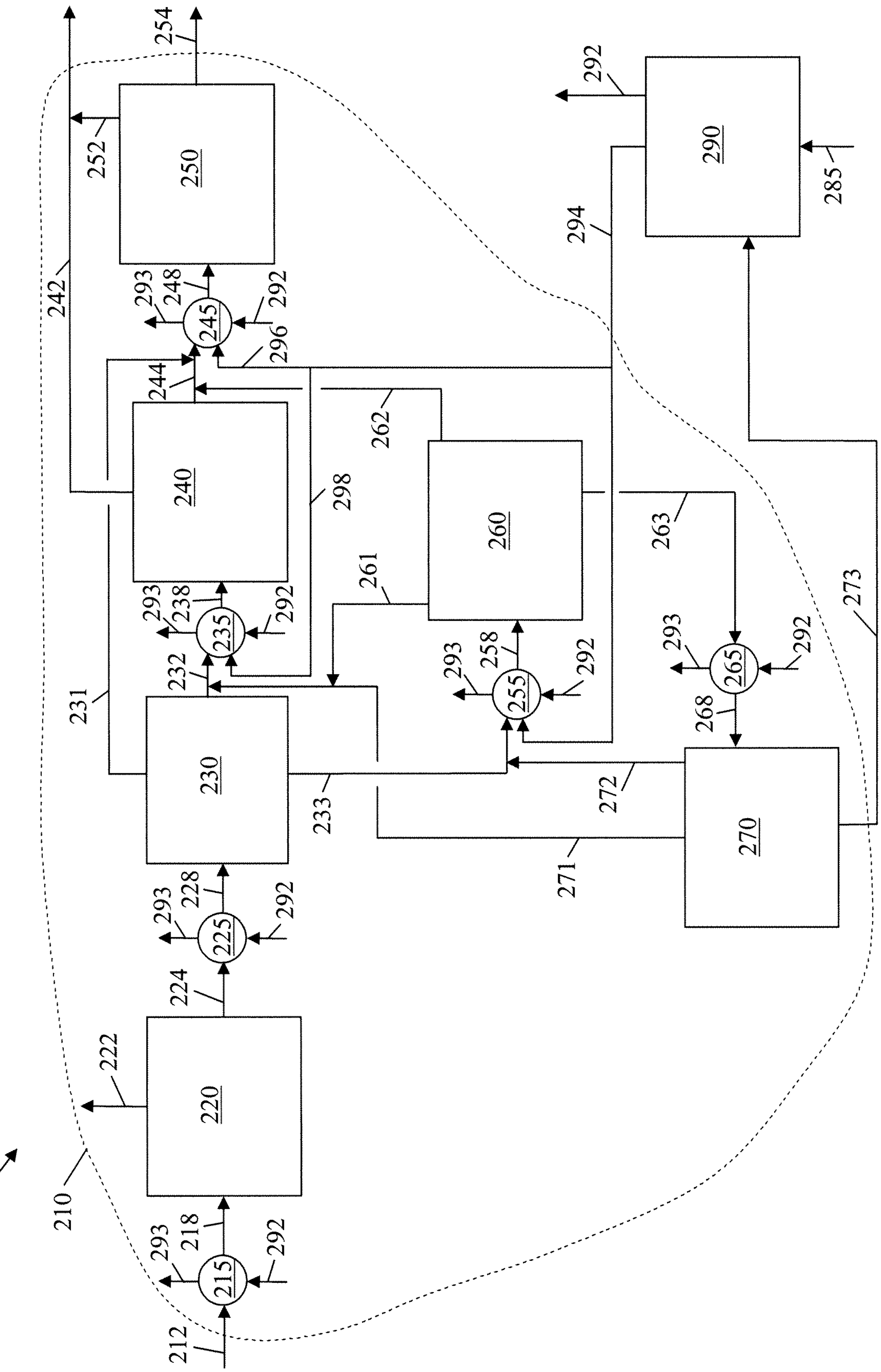
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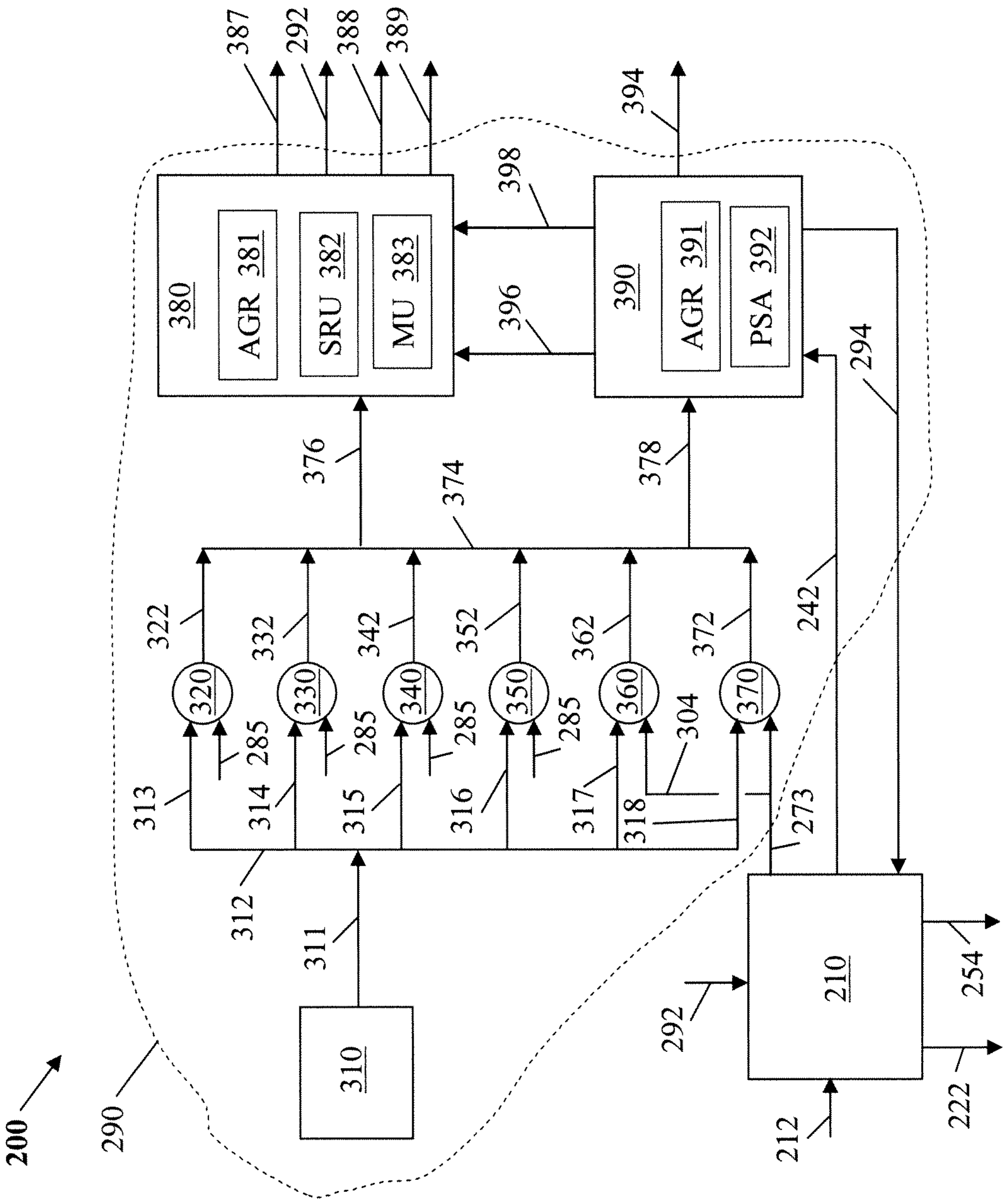
FIGURE 3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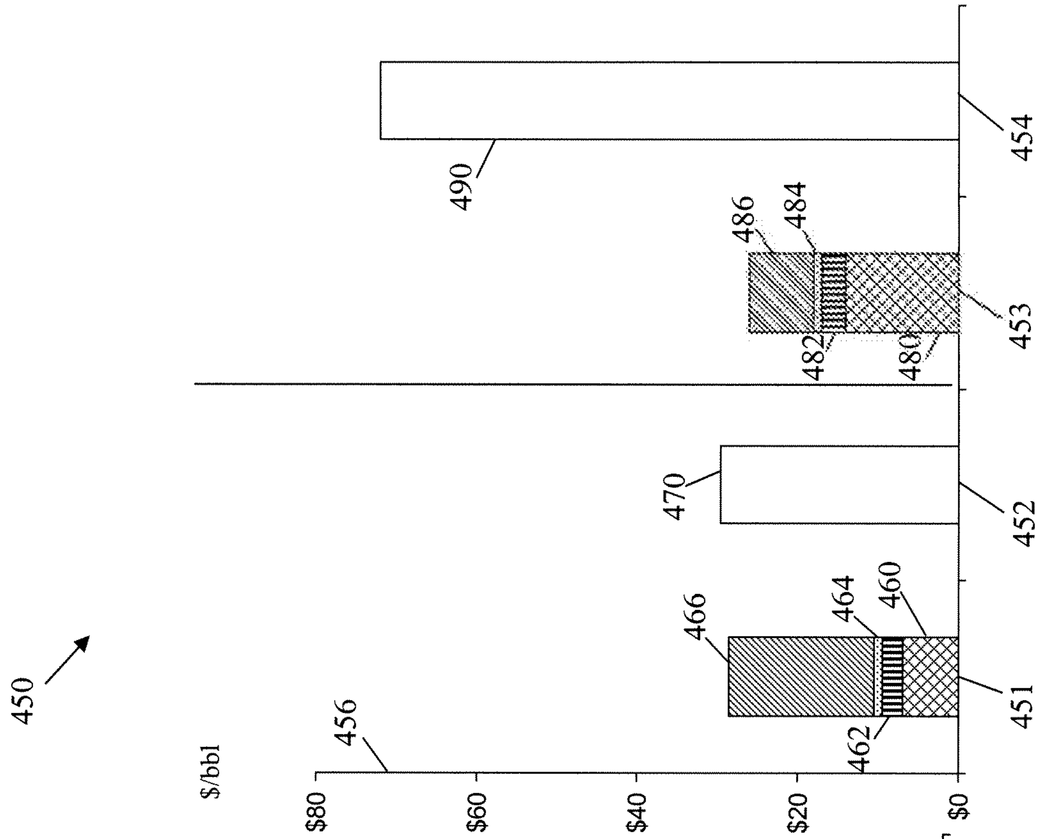

FIGURE 4B

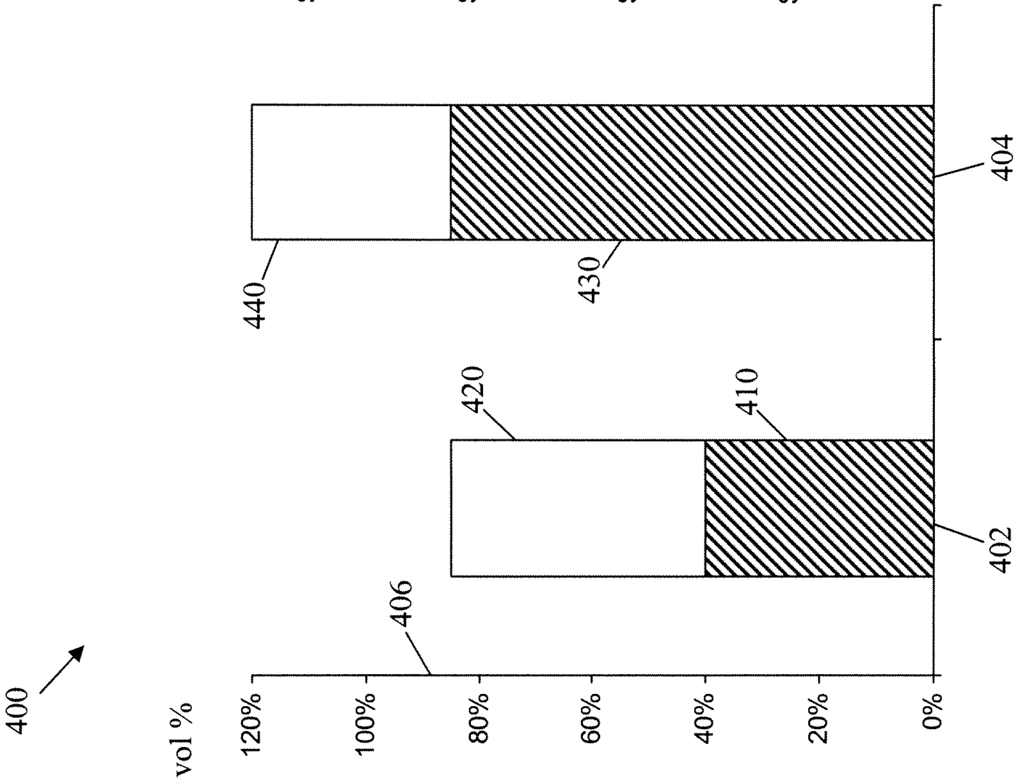

FIGURE 4A

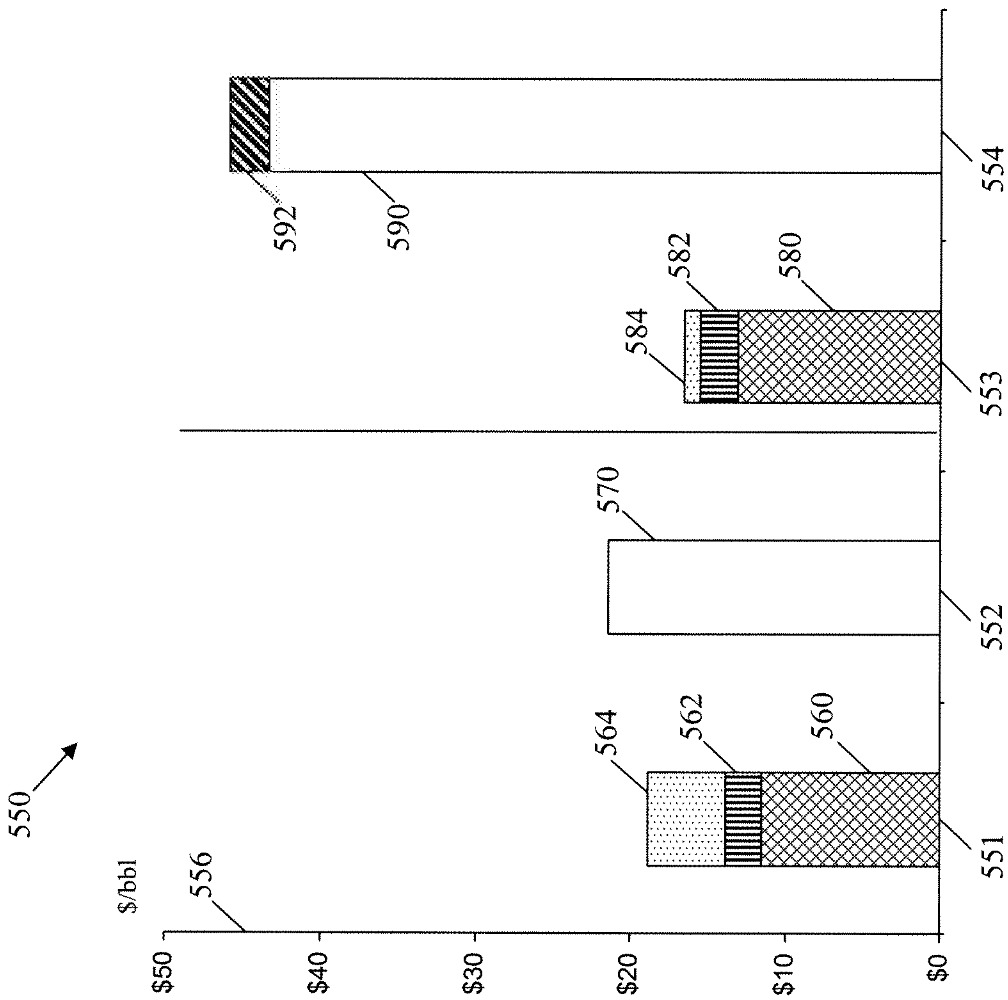

FIGURE 5B

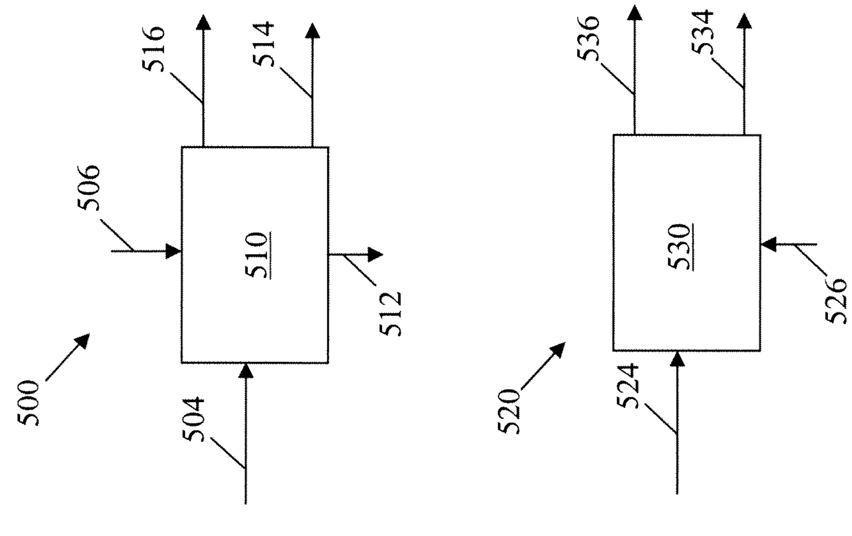

FIGURE 5A

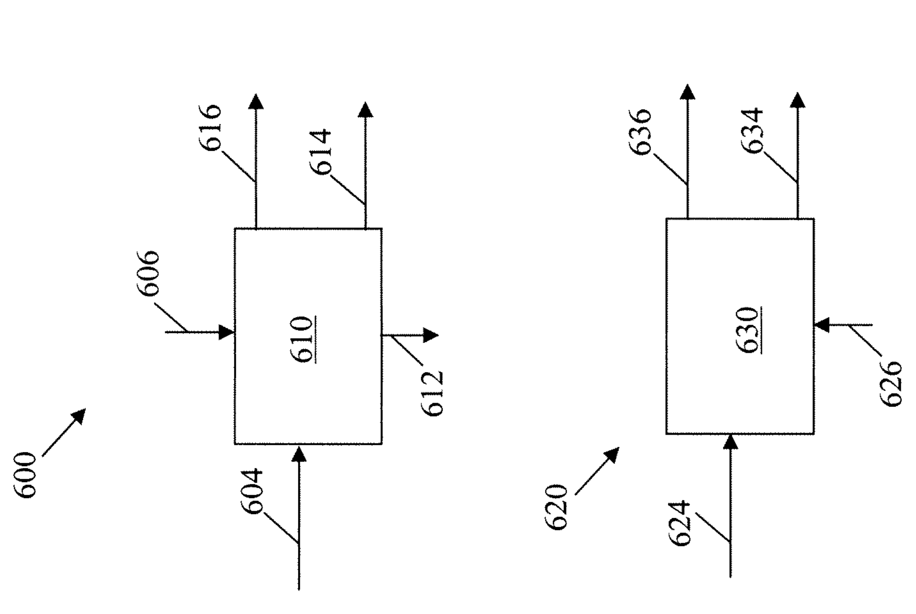

FIGURE 6A

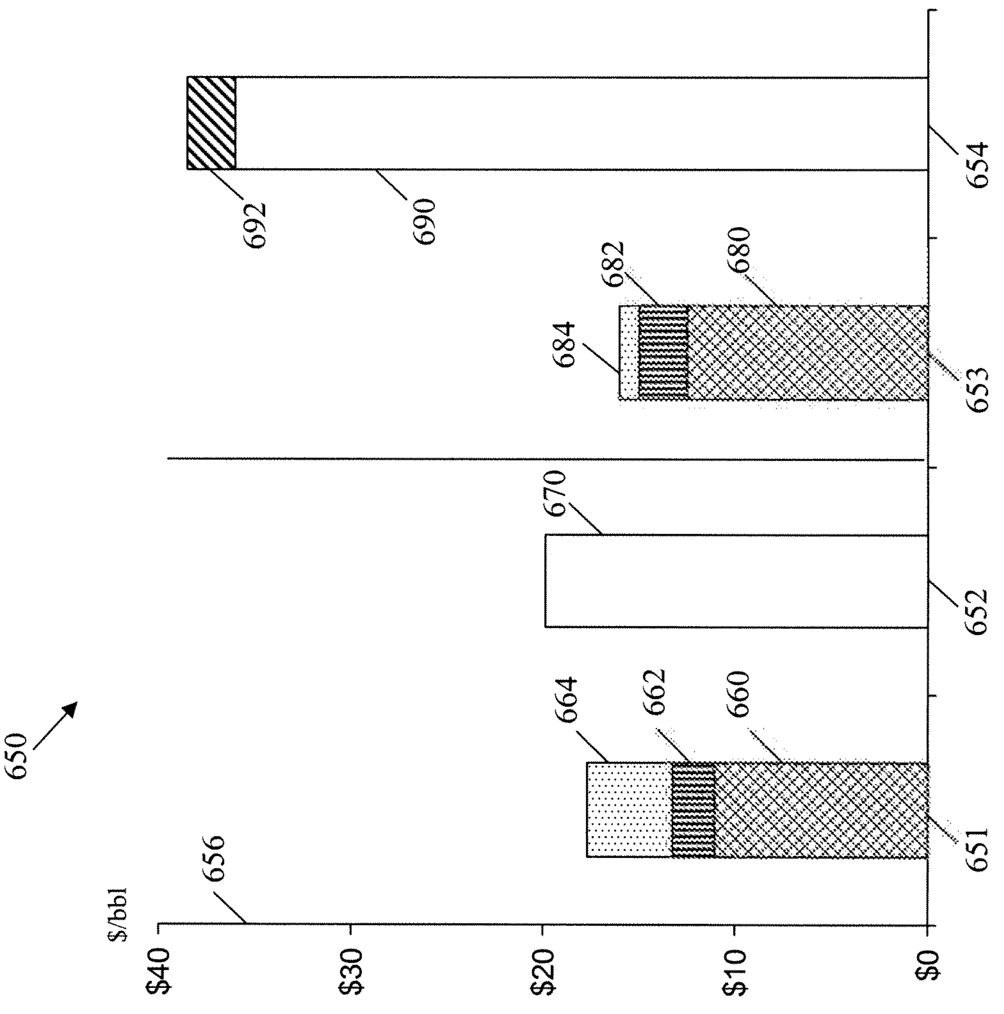

FIGURE 6B

INTERNATIONAL SEARCH REPORT

International application No.
PCT/US 09/62964

A. CLASSIFICATION OF SUBJECT MATTER
IPC(8) - C01B 3/22; C10G 45/00 (2009.01)
USPC - 208/49; 208/89; 423/650
According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)
IPC(8) -- C01B 3/22; C10G 45/00 (2009.01) and USPC -- 208/49; 208/89; 423/650

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched
USPC -- 208/86

Electronic data base consulted during the international search (name of data base and, where practicable, search terms used)
PubWest (PGPB,USPT,USOC,EPAB,JPAB); USPTO; Espacenet; Google Patents; Google Scholar; Google -- please see extra sheet for Search Terms Used

C. DOCUMENTS CONSIDERED TO BE RELEVANT

Category*	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
Y	US 5,958,365 A (Liu) 28 September 1999 (28.09.1999) Fig 2; col 2, ln 10-13 ; col 3, ln 26-32; col 3, ln 46-61; col 5, ln 29-34; col 5, ln 40-43; col 5, ln 50-53; col 6, ln 10-15 ; col 6, ln 41-50	1-36
Y	US 3,854,895 A (Muller) 17 December 1974 (17.12.1974) Figs 1-3; col 5, ln 30-37; col 5, ln 48-50; col 5, ln 66 to col 6, ln 1; col 7, ln 39-41	1-36
Y	US 4,235,044 A (Cheung) 25 November 1980 (25.11.1980) Fig 1; col 6, ln 62 to col 6, ln 6	2,20
Y	US 2004/0118745 A1 (Rettger et al.) 24 June 2004 (24.06.2004) Fig 7; para [0094]; para [0095]	5,7-17,23,25-35
Y	US 4,678,480 A (Heinrich et al.) 07 July 1987 (07.07.1987) Fig 1; col 4, ln 10-13	15-17,33-35
Y	US 2007/0209966 A1 (Abhari) 13 September 2007 (13.09.2007) Fig 3; para [0004]; [0007]; [0025]; [0029]; [0032]	6,24

☒ Further documents are listed in the continuation of Box C. ☐

* Special categories of cited documents:
"A" document defining the general state of the art which is not considered to be of particular relevance
"E" earlier application or patent but published on or after the international filing date
"L" document which may throw doubts on priority claim(s) or which is cited to establish the publication date of another citation or other special reason (as specified)
"O" document referring to an oral disclosure, use, exhibition or other means
"P" document published prior to the international filing date but later than the priority date claimed
"T" later document published after the international filing date or priority date and not in conflict with the application but cited to understand the principle or theory underlying the invention
"X" document of particular relevance; the claimed invention cannot be considered novel or cannot be considered to involve an inventive step when the document is taken alone
"Y" document of particular relevance; the claimed invention cannot be considered to involve an inventive step when the document is combined with one or more other such documents, such combination being obvious to a person skilled in the art
"&" document member of the same patent family

Date of the actual completion of the international search	Date of mailing of the international search report
22 DECEMBER 2009 (22.12.2009)	12 JAN 2010

Name and mailing address of the ISA/US	Authorized officer:
Mail Stop PCT, Attn: ISA/US, Commissioner for Patents P.O. Box 1450, Alexandria, Virginia 22313-1450 Facsimile No. 571-273-3201	Lee W. Young PCT Helpdesk: 571-272-4300 PCT OSP: 571-272-7774

INTERNATIONAL SEARCH REPORT

International application No.
PCT/US 09/62964

C (Continuation). DOCUMENTS CONSIDERED TO BE RELEVANT

Category*	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
Y	US 2006/0118463 A1 (Colyar et al.) 08 June 2006 (08.06.2006) Fig 1; para [0063]; [0065]; [0066]	7-10,13,14,25-28,31,32
Y	US 2008/0193345 A1 (Lott et al.) 14 August 2008 (14.08.2008) Fig 2C; para [0021]; [0093]; [0173]	8-13,26-31
Y	US 4,676,885 A (Bush) 30 June 1987 (30.06.1987) col 9, ln 25-30	14,32
Y	US 4,895,639 A (Bellinger et al.) 23 January 1990 (23.01.1990) col 2, ln 25-47	10,28

INTERNATIONAL SEARCH REPORT

International application No.

PCT/US 09/62964

Search Terms Used:

ASPHALT$ COKE CRUDE OIL DEASPHALT$ DE-ASPHALT$ DIESEL DISTILL$ EBULLAT$ EXTRUDATE$ FUEL/DIESEL FUEL-DIESEL GASIF$ HEAT-EXCHANG$ HEAVY CRUDE HIGH PRESSURE STEAM HYDROGEN HYDROTREAT$ HYDRO-TREAT$ JET FUEL KEROSINE KEROSENE METHANATION OIL PROCESSING PREHEAT$ PRE-HEAT$ PRESSURE SWING PRESSURE-SWING PSA SEDIMENT SEDIMENT CONTROL SLURRY REACTOR STEAM STEAM HEAT$ SUPERHEATED STEAM TRILOBE$ TRI-LOBE$ UPGRAD$ UP-GRAD$