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(54) Catalytic cracking with quenching.

(57) Effective quenching is provided in a catalytic cracking process to decrease thermal cracking. For increased product yield, the quench is injected at special locations and a special quench is used. In the illustrated embodiment, the quench is injected into the oil product immediately downstream of a gross cut separator and the quench comprises cycle oil.

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BACKGROUND OF THE INVENTION

This invention relates to catalytic cracking and, more particularly, to a process and system for increasing the yield of valuable liquids in a catalytic cracking unit.

Catalytic cracking of oil is an important refinery process which is used to produce gasoline and other hydrocarbons. During catalytic cracking, the feedstock, which is generally a cut or fraction of crude oil, is cracked in a reactor under catalytic cracking temperatures and pressures in the presence of a catalyst to produce more valuable, lower molecular weight hydrocarbons. Gas oil is usually used as a feedstock in catalytic cracking. Gas oil feedstocks typically contain from 55% to 80% gas oil by volume having a boiling range from about 650° F (343° C) to about 1000° F (538° C) and less than 1% RAMS carbon by weight. Gas oil feedstocks also typically contain less than 5% by volume naphtha and lighter hydrocarbons having a boiling temperature below 430° F, from 10% to 30% by volume diesel and kerosene having a boiling range from about 430° F (221° C) to about 650° F (343° C), and less than 10% by volume resid oil having a boiling temperature above 1000° F. Resid oil is sometimes present in greater concentrations or added to the gas oil feedstock.

In conventional fluid catalytic cracking units (FCCU), the hot products from the riser reactor continue to undergo thermal cracking reactions above 900° F (482° C) downstream of the riser reactor. These thermal cracking reactions degrade the products, reduce yields, and make excess light gases which often unduly limit the production capability of the catalytic cracking unit.

Furthermore, while it is often desirable to operate a riser reactor at higher temperatures, such as at 1025° F (552° C) or higher, to increase gasoline octane and oil and resid conversion, such high temperature cracking have substantially increased the production of ethane and lighter fuel gas. This dramatic increase of fuel gas production can create an imbalance in the refinery fuel gas system. It may also limit the capacity of those FCCUs which have insufficient gas compression capability to handle the increased load. Therefore, despite incentives for increased gasoline and octane production, riser temperatures have sometimes been reduced.

Operation at higher cracking temperatures produce naphthas which are less stable and are more prone to undergo undesired oxidation reactions which form gums. Prior methods for maintaining the stability of cracked naphthas and for maintaining the stability of gasolines containing cracked naphthas have included: 1) addition of antioxidant chemicals such as phenylene diamines or hindered phenols; 2) manipulation of the operating variables of the cracking process, such as lowering the cracking temperature and/or limiting the amount of resid; or 3) limiting the amount of cracked naphtha blended into the finished gasoline.

Typifying some of the many prior art catalytic crackers, regenerators, catalysts, equipment and refinery processes are those shown in U.S. Patents: 2,240,160; 2,382,270; 2,382,382; 2,398,739; 2,398,759; 2,414,002; 2,422,501; 2,425,849; 2,436,927; 2,458,862; 2,669,591; 2,827,422; 2,884,303; 2,901,418; 2,981,676; 2,985,584; 3,004,926; 3,039,953; 3,290,405; 3,338,821; 3,351,548; 3,364,136; 3,513,087; 3,563,911; 3,593,968; 3,661,800; 3,676,519; 3,692,667; 3,838,036; 3,844,973; 3,850,742; 3,886,060; 3,907,661; 3,909,392; 4,043,899; 4,218,300; 4,325,817; 4,331,533; 4,332,674; 4,341,623; 4,341,660; 4,375,021; 4,446,009; 4,478,708; 4,552,645; 4,695,370; 4,764,268; 4,814,067; 4,824,557; 4,859,310; and European Patent Application Nos. 83307095.6 (publication no. EPO 113 180 A2), 85307242.9 (publication no. EPO 180 355 A2), and 88309278.5 (publication no. EPO 311 375 A1). These prior art catalytic crackers, regenerators, catalysts, equipment, and refinery processes have met with varying degrees of success.

It is, therefore, desirable to provide an improved process and system to increase the yield of gasoline (naphtha) in catalytic cracking units and which improves the stability of gasoline (petrol) which contain these naphthas.

SUMMARY OF THE INVENTION

An improved catalytic cracking process and unit are provided which are effective, efficient, and economically attractive.

The novel catalytic cracking process and unit comprises catalytically cracking feed oil, such as gas oil, hydrotreated oil, and/or resid oil, in a reactor of a catalytic cracking unit (FCCU) in the presence of a cracking catalyst to produce a catalytically cracked, effluent product stream of upgraded oil and, after catalytic cracking is substantially completed, quenching the product stream externally and downstream of the reactor with a quench line or injector after the catalytically cracked oil has exited and been discharged from the reactor, to increase the yield of naphtha and gasoline (petrol) produce more stable gasoline. Rapid quenching also attains a desirable shift in coke make and selectivity.

Preferably, the quench has a volumetric expansion on vaporization substantially less than water and steam. In the preferred form, the quench comprises a hydrocarbon stream which has been previously cracked or otherwise processed to remove the most reactive species. Desirably, the quench should have low thermal reactivity. Previously cracked hydrocarbons are very desirable because they are less reactive to thermal quenching than fresh unprocessed virgin feedstocks and hydrotreated stocks.

To this end, the quench can comprise kerosene, light coker gas oil, coke still (coker) distillates (CSD), hydrotreated distillate, or fresh unprocessed virgin feedstocks, such as virgin gas oil, heavy virgin naphtha, light virgin naphtha, but preferably comprises light catalytic cycle oil (LCCO or LCO), heavy catalytic cycle oil (HCCO or HCO), or heavy catalytic naphtha (HCN), or any combination thereof. LCCO boils at a lower temperature than HCCO but they have about the same heat of vaporization. For best results, the quench comprises LCCO which has a greater molecular weight than water. HCCO, however, is also very useful as a quench and less expensive than LCCO.

Steam and water are generally not desirable as a quench, because they: expand a lot on vaporization, take up a lot of reactor volume, expand in overhead lines, cause pressure disruption, disturb catalyst circulation, adversely affect cyclone operation, and produce substantial quantities of polluted water which have to be purified. Excessive quantities of steam are also required in steam quenching.

Light naphtha (light virgin naphtha, light catalytic naphtha, light coker naphtha, etc.) is also not generally desirable as a quench because it occupies too much volume in the reactor. Furthermore, light naphtha is a gasoline blending product and it is not desirable to crack the light naphtha into less valuable hydrocarbons.

Decanted oil (DCO) is not generally desirable as a quench because it has a tendency to coke. Catalyst in the DCO can also erode the interior reactor walls and lines.

Resid is further not desirable as a quench because it has a tendency to coke and plug up lines.

Liquid hydrocarbon quenches are preferred over gas quenches to attain the benefit of the heat of vaporization of the liquid quench. Desirably, the liquid quench is injected into the product stream in an amount ranging from 2% to 20%, and preferably from 5% to 15% of the volume flow rate of feed oil for best results. Advantageously, quenching decreases the temperature of the product stream and minimizes thermal cracking. Quenching can also increase the conversion of feed oil to upgraded oil and can increase the octane of the gasoline.

Kerosene, coker gas oil, and hydrotreated distillates are less advantageous as a quench than are LCCO and HCCO. Liquid nitrogen can be useful as a quench but is very expensive and has an undesirable volumetric expansion.

LCCO and HCCO have a high capacity to absorb heat, enhance operations, and do not materially increase operating utility, maintenance, and waste treatment costs. LCCO and HCCO provide excellent quenches because they are readily available in refineries, economical, stable, have low volume expansion, provide recoverable heat removal and have a low tendency to form coke. Quenching with cycle oil can decrease the amount of coke produced. Cycle oil quenching also permits high temperature cracking without loss of more valuable hydrocarbons, and without damaging internal cyclones, plenum, or refractory walls. Desirably, cycle oil quenching, substantially decreases fuel gas production.

In the preferred process, the coked catalyst is separated from the upgraded oil by gross separation in a vapor catalyst separator, such as in a rough cut cyclone, and the upgraded oil is immediately quenched to decrease thermal cracking of the upgraded oil to less valuable hydrocarbon products and light hydrocarbon gases. Desirably, the quenching occurs downstream of a riser reactor and the vapor product outlet (exit) of the rough cut cyclone of the catalytic cracking unit. It is more efficient and economical to add the quench to the catalytic cracked oil after gross separation of the catalyst from the oil. Required quench volumes and pumping costs are also decreased.

In one of the illustrated embodiments, quenching occurs upstream of the disengaging and stripping vessel. In one preferred form of this application, the catalytic cracking unit has an external rough cut cyclone positioned between the riser reactor and the disengaging and stripper vessel and the quench is injected immediately downstream of the vapor (product) exit of the external rough cut cyclone.

In other illustrated embodiments, the catalytic cracking unit has a disengaging vessel (disengager) with an internal rough cut separator and the quench is injected into the disengager immediately downstream and in proximity to the vapor (product) exit(s) of the internal rough cut separator. The internal rough cut separator can comprise an internal cyclone or an inverted can separator. Ballistic separator and other inertia separators can also be used.

Advantageously, with quenching, the selectivity of coke can be decreased and less coke can be produced in the dilute phase portion of the disengaging and stripping vessel. Spent coked catalyst is regenerated in a regenerator and is recycled to the riser reactor. Desirably, during the novel quenching process the regeneration temperature of the regenerator is decreased. In the preferred mode, the regenerator is operated in full CO (carbon monoxide) combustion whereby the coked catalyst is regenerated in the presence of a combustion-supporting gas, such as air, comprising excess molecular oxygen in an amount greater than the stoichiometric amount required to completely combust the coke on the coked catalyst to carbon dioxide. The regenerator can also be operated in partial CO burn.

As used in this patent application, the term "conversion" means the relative disappearance of the amount of feed which boils above 430° F (221° C).

As used in this application, the term "coke selectivity" means the ratio of coke yield to conversion.

A more detailed explanation is provided in the following description and appended claims taken in conjunction with the accompanying drawings.

BRIEF DESCRIPTION OF THE DRAWINGS

Figure 1 is a schematic flow diagram of a catalytic cracking unit with an external cyclone;
Figure 2 is a schematic flow diagram of another catalytic cracking unit with an external cyclone;
Figure 3 is a schematic flow diagram of part of an oil refinery;
Figure 4 is a schematic flow diagram of another part of the oil refinery;
Figure 5 is a schematic flow diagram of a coker unit;
Figure 6 is a schematic flow diagram of a catalytic cracking unit; and
Figures 7 and 8 are charts of product temperature for various amounts of quenches;
Figure 9 is a chart of quench volume to product volume;
Figures 10 and 11 are charts of the effects of initial quench at different catalytic cracking units;
Figure 12 is a schematic flow diagram of a catalytic cracking unit with an internal rough cut separator;
Figure 13 is a cross-sectional view of the disengager of Figure 12 taken substantially along lines 13-13 of Figure 12;
Figure 14 is an enlarged fragmentary cross-sectional view of a disengager with an inverted can and quench lines, taken substantially along lines 14-14 of Figure 15;
Figure 15 is a schematic flow diagram of a catalytic cracking unit with a center riser reactor and an internal rough cut separator comprising an inverted can; and
Figure 16 is a schematic flow diagram of another catalytic cracking unit with an internal rough cut separator.

DETAILED DESCRIPTION OF THE PREFERRED EMBODIMENTS

In refining, unrefined, raw, whole crude oil (petroleum) is withdrawn from an aboveground storage tank 10 (Figure 3) by a pump 12 and pumped through feed line 14 into one or more desalters 16 to remove particulates, such as sand, salt, and metals, from the oil. The desalted oil is fed through furnace inlet line 18 into a pipestill furnace 20 where it is heated to a temperature, such as to 750°F (399°C) at a pressure ranging from 125 to 200 psi (863 to 1380 kPa). The heated oil is removed from the furnace through exit line 22 by a pump 24 and pumped through a feed line 25 to a primary distillation tower 26.

The heated oil enters the flash zone of the primary atmospheric distillation tower, pipestill, or crude oil unit 26 before proceeding to its upper rectifier section or the lower stripper section. The primary tower is preferably operated at a pressure less than 60 psi (414 kPa). In the primary tower, the heated oil is separated into fractions of wet gas, light naphtha, intermediate naphtha, heavy naphtha, kerosene, virgin gas oil, and primary reduced crude. A portion of the wet gas, naphtha, and kerosene is preferably refluxed (recycled) back to the primary tower to enhance fractionation and efficiency. Wet gas is withdrawn from the primary tower 26 through overhead wet gas line 28. Light naphtha is removed from the primary tower through light naphtha line 29. Intermediate naphtha is removed from the primary tower through intermediate naphtha line 30. Heavy naphtha is withdrawn from the primary tower 26 through heavy naphtha line 31. Kerosene and oil for producing jet fuel and furnace oil are removed from the primary tower through kerosene line 32. Part of the kerosene and/or heavy naphtha can be fed to the quench line 186 (Figure 1) for use as part of the quench, if desired. Primary virgin, atmospheric gas oil is removed from the primary tower through primary gas oil line 33 and pumped to the fluid catalytic cracking unit (FCCU) 34 (Figure 4), sometimes via a catalytic feed hydrotreating unit.

Primary reduced crude is discharged from the bottom of the primary tower 26 (Figure 3) through the primary reduced crude line 35. The primary reduced crude in line 35 is pumped by pump 36 into a furnace 38 where it is heated, such as to a temperature from about 520°F (271°C) to about 750°F (399°C). The heated primary reduced crude is conveyed through a furnace discharge line 40 into the flash zone of a pipestill vacuum tower 42 or directly to the FCU reactor.

The pipestill vacuum tower 42 (Figure 3) is preferably operated at a pressure ranging from 35 to 50 mm of mercury. Steam can be injected into the bottom portion of the vacuum tower through steam line 44. In the vacuum tower, wet gas or vacuum condensate is withdrawn from the top of the tower through overhead wet gas line 46. Heavy and/or light vacuum gas oil are removed from the middle portion of the vacuum tower through gas oil line 48 and can be fed to a catalytic feed hydrotreating unit (CFHU) 49 (Figure 4) or to the riser reactor. Vacuum-reduced crude is removed from the bottom of the vacuum tower 42 (Figure 3) through a vacuum-reduced crude line 50. The vacuum-reduced crude, also referred to as resid or resid oil, typically has an initially boiling point near about 1000°F (538°C).

Some of the resid can be pumped and fed to FCCU 34 (Figure 4) via FCCU resid line 52 or upgraded in a resid hydrotreating unit (RHU) comprising a series of ebullated, expanded bed reactors. Light gas oil (LGO) from the RHU can also be fed to the FCCU 34 via an RHU LGO line 54. Some of the resid can be pumped to a coker unit 56 via a coker resid line 58.

The coker unit 56 (Figure 5) comprises a coker or coke drum 62 and a combined tower 64. In the coker 62, the vacuum tower bottoms are coked at a coking temperature of about 895°F (479°C) to about 915°F (491°C) at a pressure of about 10 psig (69 kPa) to about 50 psig (345 kPa). Coke is withdrawn from the coker 62 a through chute, conduit, or line 66 and transported to a coke storage area for use as solid fuel. Coker product vapors can be withdrawn from the coker 62 through coker vapor line 68 and passed (fed) to a combined coker tower 64. In the combined coker tower 64, the coker product vapor can be separated into fractions of coker gas, coker naphtha, light coker gas oil, coke still distillate (coker distillate) and heavy coker gas oil. Coker gas can be withdrawn from the combined tower 64 through coker gas line 70. Coker naphtha can be withdrawn from the combined tower 64 through coker naphtha line 72. Coke still distillate (coker distillate) can be withdrawn from the combined tower 64 through coke still distillate CSD line 73. Light coker gas oil can be withdrawn from the combined tower 64 through light coker gas line 74 and fed to the FCCU 34 (Figure 4) or the catalytic feed hydro-treater (CFHU) 49. Part of the coke still distillate (coker distillate), light coker gas oil, and/or coker gas can be fed to the quench line 186 for use as part of the quench, if desired. Heavy coker gas oil can be withdrawn from the combined tower 64 (Figure 5) through heavy coker gas oil line 76 and hydrotreated in the catalytic feed hydrotreater (CFHU) 49 (Figure 4) before being catalytically cracked in the catalytic cracker 34 (FCCU).

Heavy coker gas oil from heavy coker gas oil line 76 (Figure 5) and light vacuum gas oil and/or heaving vacuum gas oil from vacuum gas oil line 48 (Figure 3) are conveyed to the riser reactor 100, or alternatively, to the catalytic feed hydrotreater or catalytic feed hydrotreating unit (CFHU) 49 (Figure 4) where they are hydrotreated with hydrogen from hydrogen feed line 78 at a pressure ranging from atmospheric pressure to 2000 psia (13.8 MPa) preferably from 1000 psia (6.9 MPa) to 1800 psia (12.4 MPa) at a temperature ranging from 650°F (343°C) to 750°F (399°C) in the presence of a hydro-treating catalyst. The hydrotreated gas oil is discharged through a catalytic feed hydrotreater discharge line 80 and fed to the catalytic cracker 34 (FCCU).The catalytic cracking reactor 34 of Figure 1 has an upright elongated vertical riser rector 100 with an upper portion 102 and a lower portion 104. Cracking catalyst and feed oil are mixed in the bottom of the riser reactor 100. The catalytic cracker (riser reactor) 100 catalytically cracks feed oil in the presence of a cracking catalyst under catalytic cracking conditions to produce an upgraded effluent product. Stream of catalytically cracked oil containing particulates of spent coked cracking catalyst.

A gross cut inertia separator comprising an external rough cut cyclone 106 (Figure 1) is connected to and communicates with the upper portion of the riser reactor 100 via a cyclone inlet line 105. The external rough cut cyclone 106 is positioned about and at a similar elevation as the upper portion 102 of the riser reactor 100. The rough cut cyclone makes a gross separation of the coked catalyst from the catalytically cracked oil. Preferably, at least 92% to 98% of the coked catalyst in the oil is removed by the rough cut cyclone 106. Positioned downstream of the external cyclone 106 is an upright disengaging vessel or disengager 108.

The disengaging vessel 108 (Figure 1) disengages and separates a substantial amount of the remaining coked catalyst from the catalytically cracked oil. The disengaging vessel 108 operates at a temperature of 900° F (482° C) to 975° F (524° C). The disengaging vessel 108 has an upper dilute phase portion 110 with at least one internal cyclone 112, an effluent product outlet line 113, a lower dense phase portion 114, and a stripping section 116 providing a stripper in which volatile hydrocarbons are stripped from the coked catalyst. The stripping section can have baffles or internals 115. Stripping steam lines and injectors 117 can be connected to the stripper 116.

Extending from the upper portion of the external cyclone 106 (Figure 1) is a cyclone outlet line 118 providing part of the product stream line 119. The product stream line 118 has an upper horizontal section 118, a vertical intermediate section 120, an intermediate horizontal section 122, and an elongated vertical section 124 providing a product stream dipleg which extends downwardly through the upper dilute phase portion 110 of the disengaging vessel 108 to the upper section of the dense phase portion 114. The product stream dipleg 124 with an internal inertia separator providing an outlet 126 located in and communicating with the intermediate section of the upper dilute phase portion of the disengaging vessel 108. The product stream line 118 provides a disengaging vessel input line which extends between, connects and communicates with the external cyclone 106 and the upper dilute phase portion 110 of the disengaging vessel 108.

A cyclone outlet spent catalyst line, conduit, and chute provides a catalyst dipleg 128 which extends into the lower dense phase portion 114 adjacent the stripping section 116 of the disengaging vessel 108. The catalyst dipleg 128 has an upper vertical section 130, an intermediate angle section 132, a lower angle section 134, and a vertical dipleg end section 136 with an outlet opening 137. An aeration steam line 138 can be connected to the upper vertical section 130. A fluidizing steam line 139 can be connected to the lower angle section 134.

A regenerator 140 (Figure 1) comprising a regenerator vessel 142 is positioned above the disengaging vessel 108. The regenerator 140 substantially combusts and regenerates the spent coke catalyst in the presence of a combustion sustaining oxygen-containing gas, such as air. An upright vertical elongated lift pipe 144 provides a spent catalyst riser and line, which extends downwardly from the lower portion of the regeneration vessel 142 through the middle section of the dense phase portion 114 of the disengaging vessel 108 for transporting coked catalyst from the disengaging vessel 108 to the overhead regenerator vessel 142. A lift air injector 146 is positioned near the bottom of the lift pipe 144 for injecting air, lifting and transporting the spent catalyst to the regenerator vessel 142 and facilitating combustion of the coked catalyst. she regenerator vessel 142 can have internal cyclones 148 and 150, an upper dilute phase steam ring 152, an overhead flue gas line 154 and a lower dense phase fuel gas ring 156 and line 158.

Regenerated catalyst is discharged through a catalyst discharge line, conduit, and chute 160 (Figure 1) to an overhead withdrawal well and vessel 162 with an optional air ring 164 in its lower portion to offset pressure buildup. A vertical regenerated catalyst standpipe 166 extends downwardly from the withdrawal well 162 to a slide valve 168. A horizontal regenerated catalyst line 170 is connected to the lower portion 104 of the riser reactor 100 to convey regenerated catalyst to the riser reactor. A fluidization steam line 171 can be connected to the regenerated catalyst line 170 below the slide valve 168. An aeration air line 172 can be connected to the middle portion of the regenerated catalyst standpipe 166.

An aeration steam line 176 (Figure 1) can also be connected to the lower portion 104 of the riser reactor 100. Injector nozzles 178 (Figure 1) can be positioned in the lower portion 104 of the riser reactor 100 to inject the feed oil into the riser reactor. In the illustrated embodiment, a combined feed oil line 180 is connected to the nozzles 178 and to a fresh feed oil line 33. A recycle oil line 182 can be connected to and communicate with the combined feed oil line 180 to feed heavy catalytic cycle oil (HCCO), decanted oil (DCO) and/or slurry oil to the riser reactor 100, of up to 40%, preferably at a rate of 5% to 10%, by volume of the fresh feed rate in fresh feed oil line 33. The temperature of the regenerator is decreased from about 1°F (0.5°C) to about 20°F (11°C) by cycle oil quenching.

A catalytic cycle oil quench injection line 184, comprising a LCCO injection line and/or an HCCO injection line, with a vertical catalytic cycle oil injector section 186 extends downwardly, connects and communicates with the vertical section 120 of the disengaging vessel input line 118 to inject a light cycle oil (LCCO) quench and/or a heavy catalytic cycle oil (HCCO) quench into the hydrocarbon products after the products have exited the external cyclone 106 downstream of the riser reactor 100 and before the products have entered the disengaging vessel 108. The quench minimizes and inhibits substantial thermal cracking of the product stream of catalytically cracked grossly separated oil to less valuable hydrocarbons, such as fuel gas. Cycle oil quenching stops about 75% to 90% of thermal cracking of the product oil and concurrently enhances the yield of naphtha to increase the production of gasoline. During quenching, the temperature of the product stream of oil being discharged from the rough cut cyclone 106 is decreased from about 30°F (17°C) to about 200°F (111°C), preferably about 50°F (28°C) to about 80°F (44°C), such as to a range of 900°F (482°C) to about 930°F (499°C).

Cycle oil quenching enhances the conversion of feed oil to upgraded oil and increases gasoline octane. The injection rate of the quench by volume ranges from 2% to 20%, preferably from 5% to 15%, of the input rate of feed oil in the riser reactor 100. Advantageously, less coke is produced in the dilute phase portion 110 of the disengaging vessel 108. Less C_2- fuel gas is also produced during cycle oil quenching.

Mixing and vaporization of the quench can be advantageously increased to less than 5 seconds and preferably less than 3 seconds by spraying the quench with one or more atomized quench injectors to provide a quick contact quench and assure rapid mixing. The quench is injected at a downward velocity of 50 to 100 ft/sec (15 to 30 m/sec.) at a residence time of 0.1 to 5 seconds, preferably less than 0.2 seconds. Losses of quench should be avoided.

High boiling quench media improves energy recovery. The quench can be preheated, preferably above 212°F (100°C) to enhance heat recovery and minimize heat loss. Quench is sprayed into the external cyclone vapor exit line 118 to rapidly cool the products before entering the reactor vessel dilute phase.

For best results, the quench is injected as soon as the reaction is completed and preferably immediately after the coked catalyst particulates have been grossly separated from the product stream of catalytically cracked oil. Lesser amounts of quench are required after catalyst separation than before catalyst separation.

It was unexpectedly and surprisingly found that the use of cycle oil quench increases the yield of high value naphtha and can improve coke make and selectivity.

It appears that gas oil conversion beyond the riser reactor is substantially completed in the rough cut cyclone where catalyst is present. Excess fuel gas production has previously been associated with long residence time in the dilute phase portion of the disengaging vessel as a result of thermal cracking before the addition of cycle oil quench.

Regenerated catalytic cracking catalyst can be fed to the riser reactor 100 (Figure 1) through a regenerated catalyst line 170, respectively. Fresh makeup catalyst can be added to the regenerator 140. In the FCC riser reactor, the hydrocarbon feedstock is vaporized upon being mixed with the hot cracking catalyst and the feedstock is catalytically cracked to more valuable, lower molecular weight hydrocarbons. The temperatures in the riser reactor 100 can range from about 900°F (482°C) to about 1200°F (649°C), preferably from about 950°F (510°C) to about 1040°F (560°C), at a pressure from atmospheric pressure to about 50 psig (345 kPa). Weight hourly space velocity in the riser reactor can range from about 5 to about 200 WHSV. The velocity of the oil vapors in the riser reactor can range from about 5ft/sec (1.5 m/sec) to about 100 ft/sec (30 m/sec).

Suitable cracking catalysts include, but are not limited to, those containing silica and/or alumina, including the acidic type. The cracking catalyst may contain other refractory metal oxides such as magnesia or zirconia. Preferred cracking catalysts are those containing crystalline aluminosilicates, zeolites, or molecular sieves in an amount sufficient to materially increase the cracking aotivity of the catalyst, e.g., between about 1 and about 50% by weight. The crystalline aluminosilicates can have silica-to-alumina mole ratios of at least about 2:1, such as from about 2 to 12:1, preferably about 4 to 6:1, for best results. The crystalline aluminosilicates are usually available or made in sodium form, and this component is preferably reduced, for instance, to less than about 4 or even less than about 1% by weight through exchange with hydrogen ions, hydrogen-precursors such as ammonium ions, or polyvalent metal ions. Suitable polyvalent metals include calcium, strontium, barium, and the rare earth metals such as cerium, lanthanum, neodymium, and/or naturally-occurring mixtures of the rare earth metals. Such crystalline materials are able to maintain their pore structure under the high-temperature conditions of catalyst manufacture, hydrocarbon processing, and catalyst regeneration. The crystalline aluminosilicates often have a uniform pore structure of exceedingly small size with the cross-sectional diameter of the pores being in a size range of about 6 to 20 angstroms, preferably about 10 to 15 angstroms. Silica-alumina based cracking catalysts having a major proportion of silica, e.g., about 60% to 90% by weight silica and about 10% to 40% by weight alumina, are suitable for admixture with the crystalline aluminosilicate or for use as the cracking catalyst. Other cracking catalysts and pore sizes can be used. The cracking catalyst can also contain or comprise a carbon monoxide (CO) burning promoter or catalyst, such as a platinum catalyst, to enhance the combustion of carbon monoxide in the dense phase in the regenerator 140.

Spent catalyst containing deactivating deposits of coke is discharged from the disengaging vessel 108 and lifted upward through the spent catalyst riser 144 and fed to the bottom portion of the overhead fluidized catalyst regenerator or combustor 140. The riser reactor and regenerator together provide the primary components of the catalytic cracking unit. Air is injected upwardly into the bottom portion of the regenerator via the air injector line 146 and spent catalyst riser 144. The air is injected at a pressure and flow rate to fluidize the spent catalyst particles generally upwardly within the regenerator. Residual carbon (coke) contained on the catalyst particles is substantially completely combusted in the regenerator 140 leaving regenerated catalyst for use in the reactor. The regenerated catalyst is discharged from the regenerator 140 through regenerated catalyst line 160 and fed to the riser reactor 100 via the regenerated catalyst line 170 and the regenerated catalyst standpipe 172. The combustion off-gases (flue gases) are withdrawn from the top of the combustor 140 through an overhead combustion off-gas line or flue gas line 154.

As shown in Figure 6, the effluent product stream of catalytically cracked hydrocarbons (volatized oil) is withdrawn from the top of disengaging vessel 108 through an effluent product line 113 and conveyed to the FCC main fractionator 190. In the FCC fractionator 190, the catalytically cracked hydrocarbons comprising oil vapors and flashed vapors can be fractionated (separated) into light hydrocarbon gases, naphtha, light catalytic cycle oil (LCCO), heavy catalytic cycle oil (HCCO), and decanted oil (DCO). Light hydrocarbon gases are withdrawn from the FCC fractionator through a light gas line 192. Naphtha is withdrawn from the FCC fractionator through a naphtha line 194, LCCO is withdrawn from the FCC fractionator through a light catalytic cycle oil line 196. HCCO is withdrawn from the FCC fractionator through a heavy catalytic cycle oil line 198. Decanted oil is withdrawn from the bottom of the FCC fractionator through a decanted oil line 199. Part of the LCCO and/or HCCO can be recycled to the cycle oil quench line 184 (Figure 1) for use as the quench.

Alternatively, in the main fractionator the oil vapors and flashed vapors can be fractionated (separated) into: (a) light hydrocarbons having a boiling temperature less than about 430° F (221° C), (b) light catalytic cycle oil (LCCO), and (c) decanted oil (DCO). The light hydrocarbons can be withdrawn from the main fractionator through an overhead line and fed to a separator drum. In the separator drum, the light hydrocarbons can be separated into (1) wet gas and (2) C_3 to 430-° F (221-° C) light hydrocarbon material comprising propane, propylene, butane, butylene, and naphtha. The wet gas can be withdrawn from the separator drum through a wet gas line and further processed in a vapor recovery unit (VRU). The C_3 to 430-° F (221-° C) material can be withdrawn from the separator drum through a discharge line and passed to the vapor recovery unit (VRU) for further processing. LCCO can be withdrawn from the main fractionator through an LCCO line for use as part of the quench or further refining, processing, or marketing. Decanted oil (DCO) can be withdrawn from the main fractionator through one or more DCO lines for further use. Slurry recycle comprising decanted oil (DCO) can be pumped from the DCO line 199 (Figure 6) at the bottom portion of the main fractionator 190 by a pump through a slurry line 182 (Figure 1) for recycle to the riser reactor 100. The remainder of the DCO can be conveyed through for further use in the refinery.

Spent deactivated (used) coked catalyst discharged from the riser reactor 100 (Figure 1) can be stripped of volatilizable hydrocarbons in the stripper section 116 with a stripping gas, such as with light hydrocarbon gases or steam. The stripped, coked catalyst is passed from the stripper 116 through spent catalyst line 144 into the regenerator 140. Air is injected through air injector line 146 to fluidize and carry the spent coked catalyst into the regenerator 140 via the spent catalyst riser 144 at a rate of about 0.2 ft/sec (0.06 m/sec) to about 4 ft/sec (1.22 m/sec). Preferably, excess air is injected in the regenerator 140 to completely convert the coke on the catalyst to carbon dioxide and steam. The excess air can be from about 2.5% to about 25% greater than the stoichiometric amount of air necessary for the complete conversion of coke to carbon dioxide and steam.

In the regenerator 140 (Figure 1), the coke on the catalyst is combusted in the presence of air so that the catalyst contains less than about 0.1% coke by weight. The coked catalyst is contained in the lower dense phase section of the regenerator, below an upper dilute phase section of the regenerator. Carbon monoxide (CO) can be combusted in both the dense phase and the dilute phase, although combustion of carbon monoxide predominantly occurs in the dense phase with promoted burning, i.e., the use of a CO burning promoter. The temperature in the dense phase can range from about 1050° F (566° C) to about 1400° F (760° C). The temperature in dilute phase can range from about 1200° F (649° C) to about 1510° F (821° C). The stack gas (combustion gases) exiting the regenerator 140 through overhead flue line 154 preferably contains less than about 0.2% CO by volume (2000 ppm). The major portion of the heat of combustion of carbon monoxide is preferably absorbed by the catalyst and is transferred with the regenerated catalyst through the regenerated catalyst line 170 and standpipe 166 riser reactor 100.

In a catalytic cracker (riser reactor) 100, some non-volatile carbonaceous material, or coke, is deposited on the catalyst particles. Coke comprises highly condensed aromatic hydrocarbons which generally contain 4-10 wt.% hydrogen. As coke builds up on the catalyst, the activity of the catalyst for cracking and the selectivity of the catalyst for producing gasoline blending stock diminish. The catalyst particles can recover a major proportion of their original capabilities by removal of most of the coke from the catalyst by a suitable regeneration process.

Catalyst regeneration is accomplished by burning the coke deposits from the catalyst surface with an oxygen-containing gas such as air. The burning of coke deposits from the catalyst requires a large volume of oxygen or air. Oxidation of coke may be characterized in a simplified manner as the oxidation of carbon and may be represented by the following chemical equations:

a. $C + O_2 \rightarrow CO_2$

b. $2C + O_2 \rightarrow 2CO$

c. $2CO + O_2 \rightarrow 2CO_2$

Reactions (a) and (b) both occur at typical catalyst regeneration conditions wherein the catalyst temperature may range from about 1050° F (566° C) to about 1300° F (704° C) and are exemplary of gas-solid chemical interactions when regenerating catalyst at temperatures within this range. The effect of any increase in temperature is reflected in an increased rate of combustion of carbon and a more complete removal of carbon, or coke, from the catalyst particles. As the increased rate of combustion is accompanied by an increased evolution of heat whenever sufficient oxygen is present, the gas phase reaction (c) may occur. This latter reaction is initiated and propagated by free radicals. Further combustion of CO to CO_2 is an attractive source of heat energy because reaction (c) is highly exothermic.

The catalytic cracker (catalytic cracking unit) of Figure 2 is generally structurally and functionally similar to the catalytic cracker of Figure 1, except that the light catalytic cycle oil (LCCO) quench line 284 is at an angle of inclination ranging from about 15 degrees to about 45 degrees, preferably about 30 degrees, relative to the vertical to increase the trajectory of the quench and enhance more uniform blending. The regenerator vessel 242 is also positioned laterally away from the disengaging vessel 208. For ease of understanding, the parts, elements, and components of the catalytic cracker of Figure 2 have been given part numbers similar to the corresponding parts, elements, and components of the catalytic cracker of Figure 1, except increased by 100, i.e., in the 200 series, e.g., riser reactor 200, external cyclone 206, disengaging vessel 208, stripper 216, regenerator 240, etc. The catalytic cracking reactor preferably comprises a riser reactor. Some catalytic cracking units can have two riser reactors, two rough cut cyclones, two slide valves, and two standpipes operatively connected to a single regenerator and to a single disengaging vessel.

The catalytic cracker (catalytic cracking unit) of Figures 12 and 13 is generally structurally and functionally similar to the catalytic cracker of Figure 2, except that four internal rough cut inertia separators 306 comprising gross (rough) cut internal cyclones are used in lieu of external cyclones to grossly separate a substantial amount of catalyst from the catalytically cracked oil after the product stream of catalytically cracked oil has been discharged from the riser reactor 300 via horizontal product line 305. Four CCO quench injector lines 384 extend into the interior dilute phase portion (zone) 310 is the disengaging vessel (disengager) 308 to locations just above the vapor product exit 318 of the internal gross cut separators 306 to inject and spray a CCO quench comprising LCCO and/or HCCO into the catalytically cracked oil after most of the coked catalyst has been removed from the oil by the internal gross cut separators 306. The quench injector lines can be positioned at an angle of inclination ranging from about 15 degrees downwardly to about 90 degrees (horizontal) relative to the vertical to minimize backflow of quench.

In Figure 12, a vertical outlet spent catalyst line, conduit, and chute 328 depends downwardly from the internal gross cut separators 306 to discharge separated spent coked catalyst into the lower dense phase portion (zone) 314 and stripping section (stripper) 316 of the disengaging vessel 308. The top portion of the upper dilute phase zone 310 of the disengaging vessel 308 can have five secondary internal cyclones 312. The disengaging vessel 308 and secondary internal cyclones 312 above the rough cut separators 306, cooperate to remove the remaining coked catalyst particles (fines) from the effluent gases and oil vapors. For ease of understanding, the parts, elements, and components of the catalytic cracker of Figures 12 and 13 have been given part numbers similar to the corresponding parts, elements, and components of the catalytic cracker of Figure 2, riser reactor 300, internal rough cut cyclone 306, stripper 316, regenerator 340, etc.

One of the major design changes implemented on FCCU 600 unit which is similar to the catalytic cracker of Figures 12 and 13, was the use of HCCO instead of LCCO to quench the disengager. HCCO was selected instead of LCCO to avoid flooding, i.e. exceeding the capacity of the LCCO section of the fractionator, and to improve overall unit heat recovery, as well as to take advantage of the greater pumping capacity of the HCCO circuit.

The HCCO quench nozzles are positioned to maximize quench efficiency by cooling the reaction gases as soon as they exit the cyclone. HCCO quench can cool the disengager by 30° F (17° C) to 200° F (111° C), preferably at least about 100° F (55° C).

The catalytic cracker (catalytic cracking unit) of Figures 14 and 15 is generally structurally and functionally similar to the catalytic cracker of Figure 12, except the upright center, central riser reactor 400 extends vertically upwardly into the dilute phase portion (zone) 410 of and along the vertical axis of the disengaging vessel (disengager) 408. Coaxially positioned about the upper end 409 of the riser reactor 400 is an internal rough (gross) cut inertia separator 406 comprising an inverted can. The inverted can 406 has: an open bottom end 406a for discharge (egress) of separated coked catalyst into the dense phase portion (zone) 414 and stripper section (stripper) 416 of the disengaging vessel 408; an imperforate solid planar or flat top or ceiling 406b spaced above the upper end 409 of the riser reactor 400 and providing a striker plate upon which the catalyst laden stream of catalytically cracked oil strikes upon exiting the upper end 409 of the riser reactor; an upper cylindrical tubular wall 406c which extends downwardly from the top 406b; an intermediate portion providing a hood 406d extending below the upper wall 406c; and a lower cylindrical tubular wall 406e about the open bottom 406a which extends downwardly below the hood 406d.

The hood 406d (Figures 14 and 15) comprises an outwardly flared skirt. The hood 406d has an elongated downwardly diverging upper frustroconical wall 406f, which extends downwardly from the upper wall 406c, and has an downwardly converging frustroconical lower wall 406g, which extends downwardly from wall 406f. The upper frustroconical wall 406f has a pair of diametrically opposite rectangular discharge openings or windows 406h which provide outlet ports for egress (exiting) of the effluent product stream of catalytically cracked oil after the oil has been grossly separated from the catalyst.

When the catalyst laden stream of catalytically cracked oil exits the upper end 409 (Figures 14 and 15) of the riser reactor 400, it strikes the top 406b of the internal gross cut separator (inverted can) 406 with sufficient momentum and force to grossly separate a substantial amount of spent coked catalyst from the catalytically cracked oil. The separated catalyst is discharged in part by gravity flow through the open bottom 406a of the inverted can 406. The catalytically cracked oil after being grossly separated from the catalyst, is discharged through the windows 406h of the inverted can 406.

A pair of diametrically opposite horizontal quench lines or injectors 484 (Figure 14) extend horizontally into the interior dilute phase portion (zone) 410 of the disengaging vessel 408 at locations in proximity to and in alignment with the windows 406h to inject and spray a quench comprising LCCO and/or HCCO into the catalytically cracked oil. The quench lines 484 can be positioned at an angle of inclination ranging from about 15 degrees downwardly to about 90 degrees (horizontal) relative to the vertical to minimize backflow of quench.

The disengaging vessel 408 (Figure 15) and the secondary internal cyclones 412 at the top of the disengaging vessel, above the rough cut separator 406, cooperate to remove the remaining coke catalyst particulates (fines) from the effluent gases and oil vapors. For ease of understanding, the parts, elements, and components of the catalytic cracker of Figures 14 and 15 have been given part numbers similar to the corresponding parts, elements, and components of the catalytic cracker of Figure 12, except in the 400 series, e.g., riser reactor 400, internal rough cut separator 406, stripper 416, regenerator 440, etc.

The catalytic cracker (catalytic cracking unit) of Figure 16 is generally structurally and functionally similar to the catalytic cracker of Figure 12, except that the regenerator 540 is positioned below the disengaging vessel (disengager) 508. For ease of understanding, the parts, elements, and components of the catalytic cracker of Figure 16 have been given part numbers similar to the corresponding parts, elements, and components of the catalytic cracker of Figure 12, except in the 500 series, e.g., riser reactor 506, stripper 516, regenerator 540, etc.

In some circumstances, it may be desirable to use a fluid bed reactor or a fluidized catalytic cracking reactor instead of or with a riser reactor.

EXAMPLES

The following examples serve to give specific illustration of the practice of this invention but are not intended in any way to limit the scope of this invention.

Examples 1 and 2

Experimental tests were conducted in a catalytic cracking unit (Unit Y) similar to that shown in Figure 1. The test of Example 1 provided the base case. Catalytic cracking in Example 1 proceeded without a LCCO quench. Catalytic cracking in the test of Example 2 was conducted with an LCCO quench with a temporary gerry-rig quench line. The operating conditions and test results are shown below. The LCCO quenching test produced unexpected, surprisingly good results since naphtha octanes increased by 0.2 RM/2, conversion increased by 0.64 volume %, naphtha yield increased by 0.5 volume %, heavy catalytic naphtha stability improved, C_2-gas yield decreased by 23% by weight, and coke selectivity (e.g. coke yield/conversion) improved. The extent, amount, and quality of the products produced during catalytic cracking with LCCO quench were unexpected. Such increase due to LCCO quenching has produced a substantial increase in product value.

	Example 1	Example 2	Difference Delta
LCCO Quench, MB/D	0.0	1.5	+1.5
Fresh Feed, MB/D	24.6	24.7	+0.1
Riser, °F (°C)	955.(512.7)	956.(513.3)	+1.(+0.5)
Rough-Cut Cyclone Overhead Line, °F (°C)	951.(511)	903.(484)	-48.(-27)
Vapor Line from Disengaging Vessel to Main Fractionator, °F (°C)	940.(504)	903.(484)	-34.(-20)
Temperature Difference Between Riser Reactor and Disengaging Vessel, °F (°C)	15.(8.3)	53.(29.4)	+38.(+21.1
Preheated Temperature, °F (°C)	437.(225)	435.(224)	-2.(-1)
HCCO, Recycle, B/D	710.	790.	+80.
Slurry Recycle, B/D	700.	700.	0.
Regen. Bed. Temp., °F (°C)	1312.(711)	1305.(707)	-7.(-4)
Conversion, Vol %	69.15	69.79	+0.64
Fresh Feed Properties			
API Gravity	23.9	24.0	+0.1
Nitrogen, Wt %	0.112	0.111	-0.001
Sulfur, Wt %	1.23	1.22	-0.01
Gas Oil, Vol %	97.8	97.6	+0.2

	Example 1	Example 2	Difference Delta
Conversion, Vol%	69.15	69.79	+0.64
TC2-, Wt%	3.34	3.11	-0.23
TC3, Vol%	10.80	10.90	+0.10
C3=/TC3	0.689	0.693	+0.004
TC4, Vol%	13.19	13.52	+0.33
C4=/TC4	0.524	0.513	-0.009
iC4/C4 saturates	0.789	0.785	-0.004
C5/430, Vol% gasoline blending material, e.g. pentane, pentene	51.83	52.33	+0.50
LCCO, Vol%	25.54	24.91	-0.63
DCO, Vol%	4.62	4.57	-0.05
Coke, Wt%	5.90	6.16	+0.26
Volume Recovery, Vol%	105.99	106.24	+0.24
C5/430			
MCL Octane	93.7	93.9	+0.2
RCL Octane	81.4	81.6	+0.2

	Example 1	Example 2	Difference Delta
LCN			
induction period, min	25	25	0
RON	94.5	94.5	0
MON	80.2	80.2	0
HCN			
induction period	395	615	+220
RON	92.9	93.4	+1.6
MON	80.9	81.4	+0.5

Examples 3 and 4

Bench study tests were performed on kerosene to simulate catalytically cracked oil after the coke catalyst particles have been removed. In the tests of Examples 3 and 4, the quench rate was 60 grams/hr and the oil product rate was 125 grams/hour. The quench of Example 3 was HCCO. The quench of Example 4 was LCCO. Quench results of HCCO and LCCO were very similar and are reported below

	Example 3	Example 4
Reactor, °F (°C)	1095 (591)	1100 (593)
C2- fuel gas (wt%)		
Isothermal at quenching	14	16+
20 minutes of cooling	10	9
60 minutes of cooling	8	6

Examples 5 and 6

Experimental tests were conducted in another catalytic cracking unit (FCCU 500) similar to that shown in Figure 2. The test of Example 5 provided a base case without the use of a LCCO. Catalytic cracking in the test of Example 6 was performed with a LCCO quench. The oil feed rate was 79 MBD. Riser reactor temperature was 1020° F (549° C). Without LCCO quench, the reactor temperature at the top of the disengaging vessel was 12° F (7° C) below the riser reactor. At 5.6 MBD of LCCO quench, the riser reactor temperature decreased 53° F (30° C). LCCO quench yielded a desirable decrease in drying gas production by about 16.7% from 1140 MSCFH to 980 MSCFH, significantly increased gasoline production 4.4% from 39.5 MBD to 41.2 MBD, and increased volume recovery by about 1%. LCCO quenching also decreased the production of propane, propylene, and isobutane. The operating conditions and test results are:

	Example 5 No quench	Example 6 with LCCO quench
LCCO quench rate	0 MBD	5.6 MBD
Riser temperature	1020°F (549°C)	1020°F (549°C)
Disengaging Vessel top temp.	1008°F (542°C)	967°F (519°C)
Temperature diff. between riser and disengager	+12°F (7°C)	+53°F (30°C)
C_2- dry gas	1140 MSCFH	980 MSCFH
C_5+ gasoline	39.5 MBD	41.2 MBD
Volume % recovery	108.9%	109.8%

	Example 5 No quench (wt.%)	Example 6 With quench (wt.%)	Difference Delta (wt.%)
C_2- dry gas	4.8	4.0	- .8
Propane	2.1	1.8	- .3
Propylene	5.7	5.3	- .4
Isobutane	3.7	3.5	- .2
N-butane	1.2	1.2	-
Butenes	6.8	6.7	-
C_5 gasoline	41.2	43.0	+ 1.8
LCO/DCO	29.4	29.3	-
Coke	5.1	5.2	-

Examples 7-9

Further experimental tests were conducted at catalytic cracking units with cycle oil quenches. In Example 7, LCCO quench was injected immediately after the product exit of the external rough cut cyclone in a catalytic cracking unit (Unit Y) similar to that shown in Figure 1 with a temporary gerry-rig quench line. Example 8, LCCO quench was injected immediately after the product exit of two external rough cut cyclones in another catalytic cracking unit (FCCU 500) similar to that shown in Figure 2. In Example 9, HCCO quench was injected immediately after the product exited four internal rough cut cyclones in a disengager in a catalytic cracking unit similar to that shown in Figures 12 and 13. Experimental test conditions and results are shown below and in the charts of Figures 10 and 11.

Example 7

Feed Rate	24,700 B/D
Riser Outlet Temp.	951°F (511°C)
Quench Media	LCCO
Quench Rate	1500 B/D (6.1%)
Vapor Res Time in Disengager	16 sec
Fuel Gas Reduction	635 M SCFD

$$K = \frac{635 \text{ M SCFD}}{24.7 \text{ M B/D (16 sec)}} = 1.61 \text{ SCF/BBL-Sec}$$

Example 8

Feed Rate	77,000 B/D
Riser Outlet Temp.	1017° F (547° C)
Quench Media	LCCO
Quench Rate	5500 B/D (7.1%)
Vapor Res Time in Disengager	9 sec
Fuel Gas Reduction	5 MM SCFD

$$K = \frac{5{,}000 \text{ M SCFD}}{77 \text{ M B/D (9 seconds)}} = 7.21 \text{ SCF/BBL-Sec}$$

Example 9

Feed Rate	37,000 B/D
Riser Outlet Temp.	980° F (527° C)
Quench Media	HCO
Quench Rate	~3000 B/D (8.1%)
Vapor Res Time in Disengager	13 sec
Fuel Gas Reduction	1.5 MM SCFD

$$K = \frac{1{,}500 \text{ M SCFD}}{37 \text{ M B/D (13 sec)}} = 3.12 \text{ SCF/BBL-Sec}$$

Examples 10-18

Increased reactor temperature at or above 940° F (504° C), but especially above 1000° F (538° C) diminishes the oxidation stability of the naphtha product and gasoline. Also, active matrix octane catalysts (cracking catalysts containing ultrastable-Y zeolite with or without rare earth exchanged into the zeolite, supported on a carrier matrix which exhibits cracking activity independent of the zeolite) will produce a less stable naphtha product and gasoline than will rare earth exchanged Y catalysts, which produce larger volumes of lower octane naphtha. Furthermore, inclusion of residual oil in the FCU feedstock mixture will diminish the stability of the naphtha product and gasoline.

Quenching in accordance with this invention can substantially increase the oxidation and storage stability of the naphtha product and gasoline by reducing the temperature in the dilute phase of the disengaging vessel as quickly as possible following the initial gross cut separation of the mixture of oil vapor product and catalyst.

Oxidation stability tests were conducted at catalytic cracking units with and without cycle oil quenches. In Examples 10-13, gas oil feed was catalytically cracked in a catalytic cracking unit (Unit Y) similar to that shown in Figure 1 with a temporary gerry-rig quench line, and LCCO quench, if indicated, was injected immediately after the product exit of the external rough cut cyclone. In Examples 14-16, gas oil feed was catalytically cracked in a catalytic cracking unit (FCCU 500) similar to that shown in Figure 2, and LCCO quench, if indicated, was injected immediately after the product exit of two rough cut cyclones. In Examples 17 and 18, gas oil feed was catalytically cracked in a catalytic cracking unit (FCCU 600) similar to that shown in Figures 12 and 13, and HCCO quench, if indicated, was injected immediately after the product exited two internal rough cut cyclones in the disengager (disengaging vessel). Experimental test conditions and results are shown below:

EFFECT OF DILUTE PHASE FCU QUENCH ON FCCU NAPHTHA OXIDATION STABILITY

Ex.	Unit	Product Stream
10	Y	Heavy Catalytic Naphtha
11	Y	Heavy Catalytic Naphtha
12	Y	Light Catalytic Naphtha
13	Y	Light Catalytic Naphtha
14	FCCU-500	C5-430 Total Catalytic Naphtha
15	FCCU-500	C5-430 Total Catalytic Naphtha
16	FCCU-500	C5-430 Total Catalytic Naphtha
17	FCCU-600	FCU Wild Gasoline -2045 hrs
18	FCCU-600	FCU Wild Gasoline -2000 hrs

EFFECT OF DILUTE PHASE FCU QUENCH ON FCCU NAPHTHA OXIDATION STABILITY

Cont'd

Ex.	Riser Outlet Temperature, °F (°C)	Dilute Phase Temperature, °F (°C)	Quench Fluid
10	940 (504)	940 (504)	None
11	941 (505)	903 (484)	LCCO
12	940 (504)	940 (504)	None
13	941 (505)	903 (484)	LCCO
14	1019 (548)	999 (537)	None
15	1020 (549)	940 (504)	LCCO
16	1019 (548)	939 (504)	LCCO
17	1020 (549)	990 (532)	None
18	1020 (549)	910 (488)	HCCO

EFFECT OF DILUTE PHASE FCU QUENCH ON FCCU NAPHTHA OXIDATION STABILITY

Cont'd

	Feedstock Composition			ASTM D-525
Ex.	% HVGO	% Hydrotreated Gas Oil	% Resid	Stability, Minutes
10	100	0	0	395
11	100	0	0	615
12	100	0	0	25
13	100	0	0	25
14	72	28	0	200
15	78	22	0	225
16	72	28	0	250
17	52	32	16	75
18	52	32	16	125

The preceding Examples 10-18 show the beneficial effects on quench of product stability.

Examples 19-48

Further oxidation stability tests were conducted with cycle oil quenches. LCCO quench was injected immediately after the product exit of two rough cut cyclones in a catalytic cracking unit (FCCU 500) similar to Figure 2. HCCO quench was injected immediately after the product exited two internal rough cut cyclones in the disengager (disengaging vessel) in a catalytic cracking unit (FCCU 600) similar to that shown in Figures 12 and 13. The Catalyst Complex was comprised of FCCU 500 and FCCU 600. Weighted average riser outlet temperature reflects the relative flow rates of feed to each unit (FCCU 500 and FCCU 600) and the cracking temperature of each unit (FCCU 500 and FCCU 600). Stabilities of LCN and HCN were measured as received from a sample point in the rundown line. ULR is blended from LCN and HCN which have been treated with an antioxidant additive. Test conditions and results are shown below.

EFFECT OF QUENCHING ON THE OXIDATION STABILITY OF UNLEADED REGULAR (ULR) GASOLINES OF CONTAINING FCCU PRODUCT NAPHTHA

		Riser Outlet Temperatures F (°C)			Percent Resid in Feed to Catalytic Complex
Ex.		FCCU 500	FCCU 600	Weighted Average	
19		970 (521)	980 (527)	974 (523)	4.4
20		973 (523)	980 (527)	976 (524)	4.0
21		980 (527)	980 (527)	980 (527)	4.2
22		980 (527)	980 (527)	980 (527	4.3
23		980 (527)	980 (527)	980 (527)	4.7
	Average	977 (525)	980 (527)	978 (526)	4.3
	Std. Deviation	5 (3)	-	3 (2)	0.2
24		980 (527)	981 (527)	980 (527)	4.3
25		983 (528)	983 (528)	983 (528)	4.4
26		975 (524)	985 (529)	979 (526)	4.4
27		985 (529)	985 (529)	985 (529)	4.4
28		985 (529)	985 (529)	985 (529)	4.4
29		985 (529)	985 (529)	985 (529)	4.3
30		985 (529)	985 (529)	985 (529)	4.5
	Average	983 (528)	984 (529)	983 (528)	4.4
	Std. Deviation	4 (2)	2 (1)	3 (2)	0.1

EFFECT OF QUENCHING ON THE OXIDATION STABILITY OF UNLEADED REGULAR (ULR) GASOLINES OF CONTAINING FCCU PRODUCT NAPHTHA

Cont'd

Ex.		% HCN in ULR	% LCN in ULR	% Total Catalytic Naphtha in ULR
19		50.3	14.5	64.8
20		49.8	19.2	69.0
21		56.5	14.0	70.5
22		58.0	14.7	72.7
23		56.0	9.2	65.2
	Average	54.1	14.3	68.4
	Std. Deviation	3.8	3.5	3.4
24		55.3	13.6	68.9
25		55.6	11.3	66.9
26		54.4	9.5	63.9
27		59.3	14.4	73.7
28		54.3	14.7	69.0
29		48.5	17.8	66.3
30		59.0	11.7	70.7
	Average	55.2	13.3	68.5
	Std. Deviation	3.6	2.7	3.2

EFFECT OF QUENCHING ON THE OXIDATION STABILITY OF UNLEADED REGULAR (ULR) GASOLINES OF CONTAINING FCCU PRODUCT NAPHTHA

Cont'd

		ASTM D-525 Oxidation Stability			LCCO Quench at	HCCO Quench at
Ex.		ULR	LCN	HCN	FCCU 500	FCCU 600
19		530	120	225	yes	no
20		535	120	195	yes	no
21		465	90	200	yes	no
22		415	-	-	yes	no
23		400	-	-	yes	no
	Average	469	110	207	yes	no
	Std. Deviation	63	17	16	-	-
24		710	110	>240	yes	yes
25		740	135	105	yes	yes
26		730	135	>240	yes	yes
27		770	105	>300	yes	yes
28		710	105	-	yes	yes
29		825	120	>300	yes	yes
30		725	-	-	yes	yes
	Average	744	118	-	yes	yes
	Std. Deviation	41	14	-	-	-

EFFECT OF QUENCHING ON THE OXIDATION STABILITY OF UNLEADED REGULAR (ULR) GASOLINES OF CONTAINING FCCU PRODUCT NAPHTHA

	Riser Outlet Temperatures °F (°C)						
Ex.	**FCCU 500**		**FCCU 600**		**Weighted Average**		**Percent Resid in Feed to Catalytic Complex**
31	1010	(543)	1015	(546)	1012	(544)	6.4
32	1010	(543)	1015	(546)	1012	(544)	6.5
33	1010	(543)	1015	(546)	1012	(544)	6.4
34	1010	(543)	1015	(546)	1012	(544)	6.4
35	1010	(543)	1015	(546)	1012	(544)	5.8
36	1010	(543)	1015	(546)	1012	(544)	5.5
37	1010	(543)	1015	(546)	1012	(544)	5.4
38	1010	(543)	1015	(546)	1012	(544)	5.4
39	1012	(544)	1015	(546)	1013	(545)	5.5
40	1012	(544)	1015	(546)	1013	(545)	5.7
41	1013	(545)	1016	(547)	1014	(546)	5.4
Average	1011	(544)	1015	(546)	1012	(544)	5.8
Std. Deviation	1	(0.5)	0.2	(0.1)	1	(0.5)	0.5
42	1010	(543)	1020	(549)	1014	(546)	5.3
43	1010	(543)	1020	(549)	1014	(546)	5.3
44	1010	(543)	1020	(549)	1014	(546)	5.8
45	1010	(543)	1020	(549)	1014	(546)	6.0
46	1010	(543)	1020	(549)	1014	(546)	6.0

EFFECT OF QUENCHING ON THE OXIDATION STABILITY OF UNLEADED REGULAR (ULR) GASOLINES OF CONTAINING FCCU PRODUCT NAPHTHA

Cont'd

Ex.		% HCN in ULR	% LCN in ULR	% Total Catalytic Naphtha in ULR
31		56.0	12.3	68.3
32		56.0	12.6	68.6
33		50.0	15.2	65.2
34		52.0	15.7	67.7
35		49.0	14.8	63.8
36		56.0	9.0	65.0
37		55.1	12.5	67.6
38		55.1	12.5	67.6
39		54.7	12.7	67.4
40		56.5	10.5	67.0
41		59.3	13.6	72.9
	Average	54.5	12.9	67.3
	Std. Deviation	3.0	2.0	2.4
42		52.8	12.3	65.1
43		52.8	12.3	65.1
44		55.8	8.3	64.1
45		50.0	10.3	60.3
46		50.0	10.3	60.3

EFFECT OF QUENCHING ON THE OXIDATION STABILITY OF UNLEADED REGULAR (ULR) GASOLINES OF CONTAINING FCCU PRODUCT NAPHTHA

Cont'd

Ex.		ASTM D-525 Oxidation Stability ULR	LCN	HCN	LCCO Quench at FCCU 500	HCCO Quench at FCCU 600
31		415	--	--	yes	no
32		420	>240	>240	yes	no
33		395	--	--	yes	no
34		330	--	--	yes	no
35		340	--	--	yes	no
36		355	--	--	yes	no
37		335	--	--	yes	no
38		320	--	--	yes	no
39		340	90	--	yes	no
40		330	105	75	yes	no
41		335	90	135	yes	no
	Average	355	--	--	yes	no
	Std. Deviation	36	--	--	--	--
42		720	120	--	yes	yes
43		720	105	>300	yes	yes
44		825	105	>240	yes	yes
45		710	105	>240	yes	yes
46		710	105	>240	yes	yes

Examples 49-54

Quenching downstream of the rough cut cyclone also reduces the yield of diolefins. Diolefins (molecules containing two unsaturated carbon-carbon bonds) are believed to be the product of thermal rather than catalytic cracking reactions, and are formed in regions of the FCCU where the temperature is high, or where the residence time is long. By reducing the temperature in the disengaging zone, the application of quench will reduce the yield of diolefinic molecules.

C4 diolefins (butadienes, and in particular 1,3,butadiene) are considered detrimental in subsequent processing of FCCU butylenes in an isobutane alkylation unit; they cause a higher than desired dilution of the acid alkylation catalyst.

C5 diolefins, including, but not limited to isoprene, 1,3-pentadiene, and cyclopentadiene are considered similarly undesirable in an FCCU product stream. If the C5 FCCU product is charged to an isobutane alkylation unit, the C5 diolefins contained in this C5 hydrocarbon stream can cause a high dilution of the acid alkylation catalyst.

Alternatively, FCCU product streams containing C5 and high molecular weight diolefins may be blended into product gasolines. In gasoline, diolefins are suspected to contribute to product instability. The high reactivity of chemical compounds containing two unsaturated bonds will cause the diolefins to rapidly react with oxygen or other substances, forming undesired gums.

Accordingly, a process which produces a lower diolefin yield is to be desired. Quenching of the reactor dilute phase will lower the diolefin yield. The chemical reactions which contribute to instability in gasoline are complex. Diolefins are believed to participate in these reactions, but it is possible that the stability improvements with quenching in Examples 19-48 involve additional molecular compounds other than diolefins as well.

An example of the beneficial effect of quenching in reducing diolefin yields is given below. C5 diolefin yields from the tests are presented. Quenching is expected to change the yield of other diolefins in a similar fashion.

Yield tests were performed in a catalytic cracking unit (FCCU 500) similar to Figure 2. Samples of the total overhead C5-430 naphtha product were obtained from the vapor product line leaving the disengaging vessel.

The samples in Examples 50 and 51 were taken with one riser reactor out of service. Only one riser reactor, discharging through a single external rough cut cyclone into the common disengaging vessel, was operating.

The samples taken in Examples 49 and 52-54 were taken with both riser reactors operating. Gas products from both external rough cut cyclones were quenched immediately downstream of the external rough cut cyclones with LCCO, then both quenched streams entered the common disengaging vessel.

The rates to each riser reactor in in Examples 49, 52, and 54 were identical but were reasonably split, roughly 50/50. For Example 53, the flow rate of quench was 2500 b/d to the A outlet, 4100 b/d to the B outlet, giving a total of 6600 b/d.

The following results were obtained:

Tests Performed Without Quench

Ex.	Cracking Temp	Fresh Feed B/D	LCCO Quench B/D	C5 Diolefin Volume of Fresh Feed
49	1016°F (547°C)	73,500	0	0.21
50	1027°F (553°C)	48,900	0	0.22
51	1027°F (553°C)	48,900	0	0.17

Tests Performed With LCCO Quench

Ex.	Cracking Temp	Fresh B/D	LCCO Quench B/D	C5 Diolefin Volume of Fresh Feed
52	1009°F (543°C)	80,600	7,000	0.08
53	1019°F (548°C)	76,500	6,600	0.11
54	1020°F (549°C)	74,600	6,300	0.11

At substantially the same cracking temperature, C5 diolefin yields were reduced approximately 35-50% by the application of LCCO quenching.

Quench Selection

In general, the quench should have a boiling point of 125° F (52° C), preferably at least 430° F (221° C) in order to have a sufficient heat capacity to effectively cool the catalytically cracked oil product to minimize thermal cracking of the oil product as well as to allow heat recovery at the bottom rather than the top of the fractionator. Desirably, the quench should have a molecular weight over 90 to limit the total volumetric expansion of the quench and oil product upon vaporization to 100% to 120%, preferably 103% to 105% or less, of the volume of the oil products without the quench, i.e., the volumetric expansion of the quench should be from 0 to 20%, preferably 3% to 5% or less of the volume of the catalytically cracked oil. Furthermore, the quenoh should be inactive and inert to thermal cracking at 900° F (482° C) to 1100° F (593° C) for a residence time of 1-30 seconds in the dilute phase zone of the disengaging vessel. Previously cracked hydrocarbons, such as LCCO, HCCO, HCN, coker gas oil and coker distillates, are very desirable as quenches since they are less reactive to thermal cracking than fresh unprocessed virgin stocks, such as virgin gas oil and virgin naphtha, and hydrotreated stocks, such as hydrotreated gas oil and hydrotreated distillates. Moreover, the quench preferably has a boiling point under 900° F (482° C) to completely vaporize in the dilute phase of the disengager in order provide effective cooling of the catalytically cracked oil product and avoid coking of the walls and lines of the refinery equipment.

It is also desirable that the quench decrease C_2 fuel gas production in order to allow higher operating temperatures at the catalytic cracking unit.

The properties of various quenches are shown in Table A. LCCO in this patent application also includes intermediate reflux on tower pump arounds with a boiling range, API gravity, and molecular weight similar to that shown for LCCO in Table A.

Table A

Properties of Quenches

	Nominal Boiling Point	Nominal API Gravity	Average Molecular Wt.
Light Catalytic Cycle Oil (LCCO)	430-650°F (221-343°C)	11-30 20 avg	200-300 215 avg
Heavy Catalytic Cycle Oil (HCCO)	650-850°F (343-454°C)	2-25 10 avg	250-350
Heavy Catalytic Naphtha (HCN)	300-430°F (149-221°C)	20-50 35-40 avg	100-250
Light Coker Gas Oil (LCGO)	500-650°F 526°F avg (260-343°C 274°C avg)	30-35 34 avg	210-300
Coker Still Distillates	430-650°F 504°F avg (221-343°C 262°C avg)	30-40 35	200-250 215 avg
Kerosene	320-530°F 400-450°F avg (160-277°C 204-232°C avg)	35-45 41 avg	150-200 175

Table A (Cont'd.)

	Nominal Boiling Point	Nominal API Gravity	Average Molecular Wt.
Hydrotreated Distillate	430-650°F 560°F avg (221-343°C 293°C avg)	35-45 39.5 avg	225-265 230
Virgin Gas Oil (VGO)	650-1000°F (343-538)°C	21-32 25 avg	300-400 350 avg
Heavy Virgin Naphtha (HVN)	300-350°F 325°F avg (149-177°C 163°C avg)	40-60 50	100-250 150
Light Virgin Naphtha (LVN)	125-175°F 150°F avg (52-79°C 66°C avg)	60-80 70	70-150 125
Hydrotreated Gas Oil (HGO)	650-1000°F (343-538°C)	24-35 29 avg	300-400 350 avg
Decanted Oil (DCO)	700-1200°F 900°F avg (371-649°C 482°C avg)	-4 to +10 +2 avg	400-600 500 avg
Resid	1000-1600°F (538-871°C)	0-20 12 avg	300-1000 700 avg
Water (H_2O)	212°F (100°C)	10 avg	18

Quenching involves injecting a fluid, preferably a liquid, into the catalytic cracking unit, preferably immediate downstream of the gross cut separator (cyclone), to stop the reactions.

Generally, a superior quench process:

1) Will provide maximum economic benefits by effectively reducing the loss of valuable products to the thermal reactions that occur after catalytic cracking is substantially completed.

2. Will have minimum adverse effects on operations.
3. Will minimally affect utility costs.

Although it is quite clear that a number of fluids could be used as quench, because the requirements of a quenching process are complex, the selection of a quench material and implementation of quenching are neither simple nor obvious. A fluid that is outstanding in one aspect may be unacceptable in another.

The quench fluid cools and dilutes the FCC riser products and so reduces the yield of thermal products. Figures 7 and 8 show, i.e., the ability of various quenches to cool the product stream, i.e., show the relative cooling capacities of different fluids. Quenched product temperature is plotted as a function of the amount of quench addition. The LCCO/CAT in Figure 7 means that LCCO quench was injected into the oil product before the catalyst was grossly separated from the oil product. The quench addition, expressed as a percentage, is the ratio of the weight of quench fluid to the weight of the product stream. The heat capacity of the quench fluid and its heat of vaporization (if a liquid) influence the cooling capacity. Water is very effective and cools at 20°F (11°C) per 1 wt% addition. Hydrocarbons are also effective and provide cooling at approximately 7°F (4°C) per 1 wt% addition. Less effective is steam (4°F (2°C) per 1 wt%) because it is already vaporized. Cooling the products before removing catalyst requires tremendous amounts of quench fluid because the catalyst holds large quantities of heat and there is so much catalyst present (typically 6 times the weight of oil). Although water provides good cooling, it has drawbacks that offset this advantage.

Table B

	Quench Molecular Wt.	Relative Thermal Crack-ability	Product Con-centration	Quench Con-centration	C2- Fuel Gas Reduction Wt.%
Water	18	0.00	0.737	0.263	78.0
Hydro-carbons					
"	106	0.36	0.846	0.154	69.4
"	114	0.59	0.856	0.144	68.0
"	130	1.37	0.871	0.129	64.4
"	201	0.65	0.913	0.087	65.7
"	216	1.06	0.918	0.082	64.0
"	245	2.46	0.927	0.073	60.8
"	339	1.05	0.946	0.054	63.1
"	365	1.71	0.950	0.050	62.1
"	414	3.96	0.956	0.044	59.1

Adding a quench fluid reduces the fuel gas by decreasing the temperature of the product diluting the concentration of riser products. The rate of thermal degradation of the riser products (and also the hydrocarbon quench) depends upon the temperature, the residence time in the system, the concentration of vapor, and the inherent reactivity (thermal crackability) of the material. Reducing the concentration of riser products slows the rate of degradation provided that the quench fluid itself has a lower thermal crackability than the riser product. Table B gives the relative molar concentrations of riser product initially at 1000° F (538° C) and quench fluid for various quench fluids of different molecular weights injected at a ratio of about 15% by weight of the product. In Table B and the following tables the C2-fuel gas reduction is relative to the instantaneous cooling of the hydrocarbon products from 1000° F (538° C) to 900° F (482° C) with a residence time of about 13 seconds. The quench fluids (injected as liquids) expand to different volumes depending on the molecular weights. The lowest molecular weights provide the maximum expansion and, therefore, the maximum dilution of the riser product. Table B also provides an estimate of the reduction in C2-fuel gas production based on laboratory tests and includes the relative thermal reactivity of the quench fluids. Quench fluids that have low molecular weights give the maximum reduction in C2-fuel gas production since C2-fuel provided measures the extent of thermal degradation, provided that the quench fluid itself has a low susceptibility to thermal cracking.

Stability of the quench is important. A quench material that is unstable will require excessive replacement and will itself contribute to the C2-yield. Table B includes the thermal stability of the various fluids. The thermal stability (crackability) was determined from laboratory tests of various quench fluids. The values in the table are relative to the thermal stability of heavy catalytic naphtha, which will have properties similar to riser products. Of course, the non-hydrocarbon, water, does not crack, so its performance establishes a target for the hydrocarbons. Hydrocarbons with low crackability give satisfactory performance.

Mixing time is also an important factor in quenching. When the quench fluid is injected into the hot product stream, the quench and product streams must mix as quickly as possible in order to get the maximum rate of cooling. Inefficient mixing of the two streams allows extra time for the thermal reactions to proceed. By using atomizing nozzles to inject the quench fluid, very small droplets are formed that disperse and vaporize quickly.

The effect of mixing time on the reduction in thermal products is indicated in Table C, based on laboratory results for LCCO quench:

Table C

Effects of Cooling Time on C2 Fuel Gas Reduction for LCCO Quench

Product Temperature	Cooling Time sec	Wt% of Quench to Product	C2-Fuel Gas Reduction %
1000°F (538°C)	1	15	92.4
1000°F (538°C)	5	15	64.0
1200°F (649°C)	1	60	92.4
1200°F (649°C)	5	60	68.5

Vapor expansion is an important factor in selecting the proper quench. Vaporized quench enters the product recovery system and must be compatible with the process equipment and control. Improper selection of the quench fluid can lead to upsets in the riser discharge flow, in the separation of catalyst from the product vapors, and can cause interference with the efficient operation of the product fractionator. In order to minimize these disruptions, the quench fluid should give the minimum expansion to the vapor so that erratic and extreme pressure levels are avoided. Figure 9 shows the ratio of the volume of the quenched product stream to the product stream alone as a function of temperature drop upon quenching for various quench fluids. The legend LCCO/CAT in Figure 9 means that LCCO quench was injected into the oil product before the catalyst was grossly separated from the oil product. The gases, steam and propane, have the largest increases because substantial quantities must be added to cool the stream, and the low molecular weight gives large volumes of gas. Water also has a substantial vapor expansion. A water-quenched stream will have almost 20% more volume than the product stream alone. This magnitude of expansion can affect operations adversely and should be avoided. On the other hand, the liquid hydrocarbons exhibit a nearly neutral volume change. For the liquid hydrocarbons, the molecular weight is typically high enough so that the volume of gas is much less than for water. Also, the expansion of the hydrocarbon is offset by the contraction of the cooled product so that a nearly constant volumetric flow rate is achieved. This criterion is in contrast to the benefit of low molecular weight diluting the product vapor.

There are practical limits on the amount of quench that is used. The benefits diminish as the amount of quench increases. Also, the benefits are greatest the higher the riser product temperature. Table D illustrates this. Each pair of conditions in the table correspond to two levels of quench addition. At 1000° F (538° C) doubling the amount of quench reduces the C2-yield by only 45%. At 1200° F (649° C) increasing quench by a factor of 4 brings only a 30% improvement.

Table D

	Pre-Quench Time, seconds	Pre-Quench Temp.	Post-Quench Time, seconds	Post Quench Temp.	Ratio of Quench to Feed %	C2-Fuel Gas Reduction, %
Quench:						
Water	1	1000°F (538°C)	12	950°F (510°C)	2.33	71.2
Water	1	1000°F (538°C)	12	900°F (482°C)	5	99.3+
LCCO	1	1000°F (538°C)	12	950°F (510°C)	7	64.0
LCCO	1	1000°F (538°C)	12	900°F (482°C)	15	92.4
LCCO	5	1000°F (538°C)	8	950°F (510°C)	7	43.9
LCCO	5	1000°F (538°C)	8	900°F (482°C)	15	64.0

Table D (Cont'd.)

	Pre-Quench Time, seconds	Pre-Quench Temp.	Post-Quench Time, seconds	Post Quench Temp.	Ratio of Quench to Feed %	C2- Fuel Gas Reduction, %
Quench:						
Water	1	1200°F (649°C)	12	1100°F (593°C)	5	74.3
Water	1	1200°F (649°C)	12	900°F (482°C)	20	97.5
LCCO	1	1200°F (649°C)	12	1100°F (593°C)	15	67.2
LCCO	1	1200°F (649°C)	12	900°F (482°C)	60	92.4
LCCO	5	1200°F (649°C)	8	1100°F (593°C)	15	45.8
LCCO	5	1200°F (649°C)	8	900°F (482°C)	60	68.5

Coking is another important criteria in determining the proper quench. A high tendency to form coke is detrimental to a quench fluid. Coke deposits can restrict process flows that could force a shutdown. Excessive coke in the regenerator could adversely affect the unit's heat balance and economics. On the other hand, a quench fluid that reduces coke by interaction with catalyst in the dilute zone of the disengager vessel improves the unit's coke selectivity and economics.

The use of quench increase utilities costs. A superior quench fluid minimizes those costs. Costs that are associated with the following: replacement of lost quench fluid; pumping the quench fluid; incomplete heat recovery and losses; water requirements for cooling and as boiler feed; and treatment of dirty process water.

Some hydrocarbon quench materials can thermally degrade. C2-fuel gas is produced by the degradation. Table E presents computer model predictions on the effects of various quench medium properties on the gross reduction in C2-. A quench fluid that degrades the products shows a lower C2-fuel gas reduction.

Table E

Effects of Quench Material Properties on Predicted Performance

Spec Gravity	Avg Boiling Pt ABP °F (°C)	Molecular wt	Relative Thermal Crack-ability	Concen-tration of Quench	C2- Fuel Gas Reduc-tion, %
0.825	300 (149)	130	1.38	0.129	91.3
0.825	575 (301)	245	2.46	0.073	89.5
0.825	800 (426)	414	4.00	0.044	88.6
0.93	300 (149)	114	0.59	0.144	94.5
0.93	575 (301)	216	1.06	0.082	92.4
0.93	800 (426)	365	1.72	0.050	91.4
0.99	300 (149)	106	0.37	0.154	95.6
0.99	575 (301)	201	0.66	0.087	93.4
0.99	800 (426)	339	1.06	0.054	92.4

There are not any or very little additional process water cost associated with the use of hydrocarbon fluids as quench material. Process water must be obtained when water is the quench material. The use of process water has additional cost. Water becomes contaminated when it goes through the process and must be treated to meet pollution control regulations.

Heat recovery is another important factor in selecting the proper quench, Substantial quantities of heat are absorbed by the quench material. This heat must be recoverable in a usable form if the quench process is to be practical. Generally, the higher the temperature at which heat is available, the more easily it can be recovered. Therefore, quench fluids that boil at higher temperatures will enable better heat recovery. In the FCC catalytic cracking unit, the heat recovery is integrated into the product fractionator system. Low temperature energy in the fractionator system is typically lost to cooling water. Energy in streams below approximately 212°F (100°C) to 350°F (177°C) is not recovered. Therefore, water is a poor quench medium from an energy recovery standpoint since it condenses at 212°F (100°C) at atmospheric pressure and since most of its energy is released when it condenses. A fluid that boils just below the target quench temperature will provide the maximum heat recovery.

In Table F, the enthalpies of some candidate quench fluids (LCCO, HCCO, HVGO Gas Oil, Water) are given that correspond to the temperatures in the table. The heats, Q1, Q2, Q3, Q4, are shown which are the heats absorbable above (a) 625°F (329°C), (b) between 625°F (329°C) and 475°F (246°C), (c) between 475°F (246°C) and 325°F (163°C), (d) and between 325°F (163°C) and 60°F (16°C), respectively. Materials that absorb large amounts of heat at high temperatures (e.g., high Q1) are preferred, and those that absorb heat at low temperature (e.g., high Q4) are not preferred. For the materials in Table F, the order of preference as a quench medium is (1) HCCO, (2) LCCO, (3) Gas Oil, and lastly Water. The quenched product temperature and Q1 upper limit for each quench was at 900°F (482°C). The enthalpies were determined at a pressure of 20 psig (238 kPa).

TABLE F

ENTHALPIES OF FCCU PRODUCT QUENCH FLUIDS AND AVAILABLE HEATS

LCCO			HCCO		
Temp. °F (°C)	**BTU/LB (J/grm)**	**Available Heat BTU/LB (J/grm)**	**Temp. °F (°C)**	**BTU/LB (J/grm)**	**Available Heat BTU/LB (J/grm)**
1200 (649)	866.2 (2015)	Maximum	1200 (649)	811.5 (1888)	Maximum
1174 (634)	846.9 (1970)	Product	1174 (634)	793.3 (1845)	Product
1125 (607)	810.5 (1885)	Temp.	1125 (607)	758.9 (1765)	Temp.
1050 (566)	756.1 (1759)		1050 (566)	707.7 (1646)	
975 (524)	703.3 (1636)		975 (524)	657.8 (1530)	
900 (482)	652.1 (1517)		900 (482)	609.5 (1418)	
825 (441)	602.7 (1402)		825 (441)	562.9 (1309)	
750 (399)	555.1 (1291)		750 (399)	433.0 (1007)	Liquid
675 (357)	508.4 (1183)		675 (357)	384.0 (893)	
625 (329)	417.9 (972)	Q1= 234.2 (545)	625 (329)	353.7 (823)	Q1=255.8 (595)
600 (316)	372.6 (867)	Liquid	600 (316)	338.6 (788)	
525 (274)	325.9 (758)		525 (274)	295.5 (687)	
475 (246)	296.4 (689)	Q2= 121.4 (282)	475 (246)	268.3 (624)	Q2=85.4 (199)
450 (232)	281.7 (655)		450 (232)	254.7 (592)	
375 (191)	239.9 (558)		375 (191)	216.4 (503)	
325 (163)	213.8 (497)	Q3= 82.7 (192)	325 (163)	192.7 (448)	Q3=75.6 (176)
300 (149)	200.7 (467)		300 (149)	180.9 (421)	
225 (107)	164.5 (383)		225 (107)	148.4 (345)	
150 (66)	131.3 (305)		150 (66)	119.3 (277)	
75 (24)	101.8 (237)		75 (24)	94.0 (219)	
60 (16)	95.9 (225)	Q4= 117.9 (274)	60 (16)	88.9 (207)	Q4=103.8 (240)
32 (0)	84.9 (197)		32 (0)	79.5 (185)	

TABLE F (cont'd.)

ENTHALPIES OF FCCU PRODUCT QUENCH FLUIDS AND AVAILABLE HEATS

	Feed Oil (HVGO) Gas Oil			**Water**	
Temp. °F (°C)	**LCCO BTU/LB (J/grm)**	**Available Heat BTU/LB (J/grm)**	**Temp. °F (°C)**	**BTU/LB (J/grm)**	**Available Heat BTU/LB (J/grm)**
1200 (649)	925.6 (2153)	Maximum	1200 (649)	1639 (3812)	Maximum
1174 (634)	904.5 (2103)	Product	1174 (634)	1626 (3782)	Product
1125 (607)	864.8 (2011)	Temp.	1125 (607)	1600 (3721)	Temp.
1050 (566)	805.3 (1873)		1050 (566)	1560 (3629)	
975 (524)	747.2 (1738)		975 (524)	1522 (3540)	
900 (482)	678.6 (1578)		900 (482)	1483 (3449)	
825 (441)	566.8 (1318)	Liquid	825 (441)	1445 (3361)	
750 (399)	509.6 (1185)		750 (399)	1408 (3275)	
675 (357)	454.5 (1057)		675 (357)	1371 (3189)	
625 (329)	419.2 (975)	Q1=259.4 (603)	625 (329)	1347 (3133)	Q1=136.15(317
600 (316)	401.5 (934)		600 (316)	1335 (3105)	
525 (274)	350.7 (816)		525 (274)	1299 (3021)	
475 (246)	318.4 (741)	Q2=100.8 (234)	475 (246)	1275 (2966)	Q2=72 (167)
450 (232)	302.2 (702)		450 (232)	1263 (2938)	
375 (191)	256.3 (596)		375 (191)	1227 (2854)	
325 (163)	227.5 (529)	Q3=90.9 (211)	325 (163)	1203 (2798)	Q3=72 (167)
300 (149)	213.1 (496)		300 (149)	1191 (2770)	
225 (107)	173.0 (402)		225 (107)	193 (449)	
150 (66)	136.3 (317)		150 (66)	118 (274)	
75 (24)	103.5 (241)		75 (24)	45 (105)	Liquid
60 (16)	96.9 (225)	Q4=130.6 (304)	60 (16)	20 (47)	Q4=1175 (2733
32 (0)	84.7 (197)		32 (0)	0 (0)	

Quench Material Selection:

Some quench fluids are evaluated in Table G. Different refineries may use different quench materials to meet specific requirements or to take advantage of special opportunities. Among the fluids examined below, LCCO is best and HCCO is second best. Water has some serious shortcomings. The remaining materials have certain characteristics that can reduce their attractiveness as a quench fluid.

Table G

Evaluation of Candidate Quench Materials

Quench	Water	Steam	Propane	HCN	LCCO	HCCO
Cooling Capacity	E	P	P	A	A	A
Product Dilution	E	E	E	A	G	G
Stability	E	E	E	G	G	G
Volume Expansion	P	P	P	A	E	E
Coking Tendency	E	E	E	E	G	F
Pumping/ Transporting	G	F	F	A	A	A
Heat Recovery	P	P	P	F	G	E
Water Use	P	P	E	E	E	E
Waste Disposal	P	P	E	E	E	E

Key:

P: Poor F: Fair A: Average G: Good E: Excellent

Table G (cont'd.)

Evaluation of Candidate Quench Materials

Quench	DCO	(HVGO) GasOil	Kerosene	Hydro-treated GasOil	Hydro-treated LCCO
Cooling Capacity	A	A	A	A	A
Product Dilution	F	F	A	F	A
Stability	G	A	A	A	A
Volume Expansion	E	E	E	E	E
Coking Tendency	P	G	E	E	E
Pumping/ Transporting	A	A	A	A	A
Heat Recovery	E	G	G	G	G
Water Use	E	E	E	E	E
Waste Disposal	E	E	E	E	E

Key:

P: Poor F: Fair A: Average G: Good E: Excellent

Among the many advantages of the novel catalytic cracking and quenching process and system are:

1. Enhanced product values and quality.
2. Greater yield of more valuable hydrocarbons.
3. Production of more naphtha and finished gasoline.
4. Higher throughput.
5. Better throughput and oxidation stability of product naphtha.
6. Decreased thermal cracking and product degradation thereby minimizing overcracking of gasoline into ethane and light fuel gas.
7. Lower pentadiene content in the naphtha product.
8. Less low value fuel gas production.
9. Increased octane number of naphtha and finished gasoline.
10. Economical
11. Efficient
12. Effective.

Although embodiments of the invention have been shown and described, it is to be understood that various modifications and substitutions, as well as rearrangements of process steps, can be made by those skilled in the art without departing from the novel spirit and scope of the invention.

Claims

1. A process for catalytically cracking feed oil, in which the feed oil is catalytically cracked in the presence of a cracking catalyst and at an elevated cracking temperature to produce a product stream comprising cracked feed oil, and the product stream is cooled to a temperature below said cracking temperature by contact with a quench fluid to deter further cracking of the cracked feed oil to light hydrocarbon gases, characterised in that directly after separation of catalyst from the product stream following said catalytic cracking, the quench fluid is delivered into said product stream to mix intimately therewith and thereby rapidly cool the product stream.

2. A catalytic cracking process, comprising the steps of:
   catalytically cracking feed oil in a reactor of a catalytic cracking unit in the presence of a cracking catalyst to produce a catalytically cracked effluent product stream of upgraded oil; and
   quenching said product stream externally and downstream of said reactor with a quench comprising at least one member selected from the group consisting of light catalytic cycle oil, heavy catalytic cycle oil, heavy catalytic naphtha, kerosene, coker distillates, light coker gas oil, hydrotreated distillate, fresh unprocessed virgin gas oil, and fresh unprocessed virgin naphtha.

3. A catalytic cracking process in accordance with claim 2 wherein:
   said feed oil comprises gas oil; and
   said quenching further includes decreasing the temperature of said product stream and
   minimising thermal cracking of said product stream; and
   said quench is injected into said product stream in an amount ranging from about 2% to about 20% per barrel of feed oil.

4. A catalytic cracking process in accordance with claim 2 or 3, wherein:
   said reactor comprises a riser reactor;
   said product stream is quenched after said catalytic cracking is substantially completed; and
   said quench contacts said product stream in an amount ranging from about 5% to about 15% per barrel of feed oil.

5. A catalytic cracking process, comprising the steps of:
   catalytically cracking feed oil in the presence of a cracking catalyst; and
   cooling said catalytically cracked feed oil after said catalytic cracking is substantially completed to substantially minimise thermal cracking of said catalytically cracked oil to fuel gas by contacting said catalytically cracked oil with a hydrocarbon liquid quench having a boiling point greater than water, a molecular weight over 90, and a volumetric expansion less than about 20% of the volume of said catalytically cracked oil.

6. A catalytic cracking process in accordance with claim 5 wherein:
   said quench is substantially inert to thermal cracking at about 482°C to about 593°C;
   said feed oil is catalytically cracked in a reactor;
   said quench is selected from the group consisting of previously cracked hydrocarbons, fresh unprocessed virgin feedstock, and hydrotreated hydrocarbons; and
   said catalytically cracked oil is cooled by said quench by an amount ranging from about 17°C to about 111°C.

7. A catalytic cracking process in accordance with claim 5 or 6, wherein:
   said volumetric expansion of said quench is less than about 5%;
   said quench improves the oxidation stability of the naphtha product and gasoline;
   a substantial portion of said quench has a boiling point of at least about 221°C and below about 482°C and substantially completely vaporises in the dilute phase.

8. A catalytic cracking process in accordance with claim 5, 6 or 7, wherein said quench is selected from the group consisting of light catalytic cycle oil, heavy catalytic cycle oil, heavy catalytic naphtha, coker gas oil, and coker distillates.

9. A catalytic cracking process in accordance with any one of claims 5 to 8, including substantially separating said catalyst from said catalytically cracked oil in a gross cut separator having a vapor exit providing a product outlet and injecting said quench into said catalytically cracked oil after said

separating in proximity to said product outlet of said gross cut separator.

10. A catalytic cracking process, comprising the steps of:
catalytically cracking feed oil in a catalytic cracking unit comprising a regenerator and at lest one catalytic cracking reactor selected from the group consisting of a riser reactor and a fluidised bed rector, in the presence of a cracking catalyst to produce upgraded oil leaving coked catalyst;
making a gross-cut separation of said coked catalyst from said upgraded oil and substantially immediately thereafter;
quenching said upgraded oil to substantially decrease thermal cracking of said upgraded oil to less valuable hydrocarbon products and light hydrocarbon gases;
regenerating said coked catalyst in a regenerator; and
recycling said regenerated catalyst to said catalytic cracking reactor.

11. A catalytic cracking process in accordance with claim 10 wherein;
said upgraded oil is quenched with about 2% to about 15% of said quench per volume of said feed oil;
said quench is selected from the group consisting of light cycle oil, heavy cycle oil, heavy catalytic naphtha, and combinations thereof; and
said quenching occurs downstream of said riser reactor.

12. A catalytic cracking process, comprising the steps of:
catalytically cracking feed oil in a reactor of a catalytic cracking unit in the presence of a cracking catalyst to produce a catalytically cracked effluent stream of upgraded oil containing catalyst;
substantially separating said catalyst from said upgraded oil in an external gross cut separator and in a disengaging vessel;
quenching said upgraded oil downstream of said external gross cut separator and upstream of said disengaging vessel with a quench comprising at least one member selected from the group consisting of light catalytic cycle oil, heavy catalytic cycle oil, heavy catalytic naphtha, kerosene, coker distillates, light coker gas oil, hydrotreated distillate, fresh unprocessed virgin gas oil, and fresh unprocessed virgin naphtha; and
said quench is injected into said stream in an amount ranging from about 2% to about 20% by volume per barrel of feed oil.

13. A catalytic cracking process, comprising the steps of:
substantially desalting petroleum comprising crude oil;
heating said desalted crude oil in a furnace;
pumping said heated crude oil to a primary distillation tower;
separating said heated crude oil in said primary distillation tower into streams of naphtha, primary gas oil, and primary reduced crude;
pumping said primary reduced crude oil to a pipestill vacuum tower;
separating said primary reduced crude oil in said pipestill vacuum tower into streams of wet gas, heavy gas oil, and vacuum reduced crude oil providing resid oil;
conveying a feed oil comprising said primary gas oil from said primary distillation tower to an upright elongated riser reactor of a catalytic cracking unit;
feeding fresh and regenerated crystalline cracking catalyst to said riser reactor;
catalytically cracking said feed oil in said riser reactor in the presence of said cracking catalyst under catalytic cracking conditions to produce an upgraded effluent product stream of catalytically cracked oil containing spent coked catalyst;
separating a substantial amount of said spent coked catalyst from said product stream in an external rough cut separator downstream of said riser reactor to make a gross separation of said coked catalyst from said product stream;
injecting and quenching said product stream soon after said product stream exits said external separator with a cycle oil quench for substantially minimising thermal cracking of said product stream to less valuable hydrocarbons and concurrently enhancing the yield of naphtha to substantially increase the production of gasoline, said quench comprising a cycle oil selected from the group consisting of light catalytic cycle oil and heavy catalytic cycle oil, said quench being injected into said product stream in an amount ranging from about 5% to about 15% by volume per barrel of feed oil;
conveying said quenched product stream into an upper dilute phase portion of a disengaging

vessel;

disengaging and separating a substantial amount of the remaining spent coked catalyst fines from said quenched product stream in at least one internal cyclone in said dilute phase portion of said disengaging vessel;

stripping volatile hydrocarbons from said coked catalyst in a stripping section of said disengaging vessel;

passing said stripped coked catalyst to a regenerator of said catalytic cracking unit;

injecting air into said regenerator;

regenerating said spent catalyst by substantially combusting coke on said spent catalyst in the presence of air in said regenerator;

recycling said regenerated catalyst to said riser reactor;

separating said cracking oil in a fractionator into streams of light hydrocarbon gases, catalytic naphtha,

catalytic cycle oil including light catalytic cycle oil, and decanted oil; and

recycling said light catalytic cycle oil from said fractionator to a quench injection line located between said external cyclone and said fluidised bed reactor for use as said quench.

14. A catalytic cracking process, comprising the steps of:

catalytically cracking feed oil in a rector of a catalytic cracking unit in the presence of a cracking catalyst to produce a catalytically cracked effluent stream of upgraded oil containing catalyst;

substantially separating said catalyst from said upgraded oil in an internal gross cut separator in a disengaging vessel, said internal gross cut separator comprising an inertia separator selected from the group consisting of an internal cyclone and an inverted can separator; and

quenching said upgraded oil downstream of said internal gross cut separator in said disengaging vessel with a quench comprising at least one member selected from the group consisting of light catalytic cycle oil, heavy catalytic cycle oil, heavy catalytic naphtha, kerosene, coker distillates, light coker gas oil, hydrotreated distillate, fresh unprocessed virgin gas oil, and fresh unprocessed virgin naphtha, said quench being injected into said stream in an amount raging from about 2% to about 20% by volume per barrel of feed oil.

15. A catalytic cracking process, comprising the steps of:

substantially desalting petroleum comprising crude oil;

heating said desalted crude oil in a furnace;

pumping said heated crude oil to a primary distillation tower;

separating said heated crude oil in said primary distillation tower into streams of naphtha, primary gas oil, and primary reduced crude;

pumping said primary reduced crude oil to a pipestill vacuum tower;

separating said primary reduced crude oil in said pipestill vacuum tower into streams of wet gas, heavy gas oil, and vacuum reduced crude oil providing resid oil;

conveying a feed oil comprising said primary gas oil from said primary distillation tower to an upright elongated riser reactor of a catalytic cracking unit;

feeding fresh and regenerated crystalline cracking catalyst to said riser reactor;

catalytically cracking said feed oil in said riser reactor in the presence of said cracking catalyst under catalytic cracking conditions to produce an upgraded effluent product stream of catalytically cracked oil containing spent coked catalyst;

separating a substantial amount of said spent coked catalyst from said product stream in an internal rough cut separator in the dilute phase portion of a disengaging vessel located downstream of said riser reactor to make a gross separation of said coked catalyst from said product stream;

injecting and quenching said product stream soon after said product stream exits said internal separator with a cycle oil quench for substantially minimising thermal cracking of said product stream to less valuable hydrocarbons and concurrently enhancing the yield of naphtha to substantially increase the production of gasoline, said quench comprising a cycle oil selected from the group consisting of light catalytic cycle oil and heavy catalytic cycle oil, said quench being injected into said product stream in an amount ranging from about 5% to about 15% by volume per barrel of feed oil;

disengaging and separating a substantial amount of the remaining spent coked catalyst fines from said quenched product stream in at lease one internal cyclone in said dilute faze portion of said disengaging vessel;

stripping volatile hydrocarbons from said coked catalyst in a stripping section of said disengaging

vessel;
passing said stripped coked catalyst to a regenerator of said catalytic cracking unit;
injecting air into said regenerator;
regenerating said spent catalyst by substantially combusting coke on said spent catalyst in the presence of air in said regenerator;
recycling said regenerated catalyst to said riser reactor;
separating said quenched catalytically cracked oil in a fractionator into streams of light hydrocarbon gases, catalytic naphtha, catalytic cycle oil including light catalytic cycle oil, and decanted oil; and
recycling said light catalytic cycle oil from said fractionator to said disengaging vessel for use as said quench.

16. A catalytic cracking process in accordance with claims 13 or 15 including feeding some of said resid oil to said riser reactor for use as part of said feed oil and said quench comprises heavy catalytic cycle oil.

17. A catalytic cracking process in accordance with claims 13 or 15 wherein light catalytic cycle oil is injected into said feed oil as at least part of said quench.

18. A catalytic cracking process in accordance with claim 15 wherein:
said separating includes impinging said effluent product stream upon exiting said riser reactor against a striker plate of an inverted can separator, discharging said separated catalyst through the open bottom of said inverted can separator, and passing said separated catalytically cracked oil through at least one window comprising a product outlet of said inverted can separator; and
said quenching comprises spraying said cycle oil in proximity to said window of said inverted can separator.

19. A catalytic cracking unit comprising:
an upright riser reactor (100; 200; 300; 400; 500) for catalytically cracking feed oil in the presence of a cracking catalyst to produce an upgraded effluent product stream of catalytically cracked oil, and separating means (106; 206; 306; 406; 506) connected to the riser reactor for separating said product stream from catalyst and having an outlet for catalyst and another outlet for the separated product stream, characterised in that means (184; 284; 384; 484; 584) are arranged to deliver into said separated product stream issuing through said other outlet a fluid quench to reduce the temperature of the stream to deter thermal cracking of the catalytically cracked oil.

20. A catalytic cracking unit, comprising:
an upright elongated riser reactor (100; 200) for or catalytically cracking feed oil in the presence of a cracking catalyst to produce an upgraded effluent product stream of catalytically cracked oil leaving coked catalyst, said riser reactor having an upper portion and a lower portion;
an external rough cut separator (106; 206) connected to and communicating with said upper portion of said riser reactor and being spaced from, positioned about and at a substantially similar elevation as said upper portion of said riser reactor for making a gross separation of said coked catalyst from said catalytically cracked oil;
an upright disengaging vessel (108; 208) for substantially disengaging and separating a substantial amount of remaining coked catalyst from said catalytically cracked oil, said disengaging vessel having an upper dilute phase zone with at least one internal separator, a lower dense phase zone, and a stripping section providing a stripper;
product line (120; 220) extending between and connecting said external separator and said upper diluate phase portion of said vessel;
a spent catalyst line (128; 228) extending between and connecting said external separator and said dense phase portion of said disengaging vessel;
a regenerator (140; 240) comprising a vessel (142, 242) an upright elongated lift pipe (144; 244) for transporting coked catalyst to regenerator, an air injector (146; 246) communicating with said lift pipe for injecting air and facilitating combustion of said coked catalyst, and a regenerated catalyst line (166; 266) connected to said riser reactor for conveying regenerated catalyst to said riser reactor; and a cycle oil quench injection line (184; 284) connected to and communicating with said input line for injecting a quench comprising cycle oil selected from the group consisting of light catalytic cycle oil and heavy catalytic cycle oil, into said catalytically cracked oil after said catalytically cracked oil has exited said external separator downstream of said riser reactor and has been grossly separated from

said catalyst and before said catalytically cracked oil enters said disengaging vessel for enhancing the yield of naphtha and substantially decreasing thermal cracking of said product stream of oil, said cycle oil quench line including a substantial vertical light cycle oil injector for injecting said quench substantially vertically downwardly into said input line or being at an angle of inclination ranging from about 15 degrees to about 45 degrees relative to a vertical reference line for increasing mixing of said quench with said products.

21. A catalytic cracking unit, comprising:

an upright elongated riser reactor (300; 400; 500) for catalytically cracking feed oil in the presence of a cracking catalyst to produce an upgraded effluent product stream of catalytically cracked oil leaving coked catalyst, said riser reactor having an upper portion and a lower portion;

an upright disengaging vessel (308; 408; 508) communicating with said riser reactor for substantially disengaging and separating a substantial amount of coked catalyst from said catalytically cracked oil, said disengaging vessel having an upper dilute phase zone with at least one internal cyclone, a lower dense phase zone, and a stripping section providing a stripper;

an internal gross cut separator (306; 406; 506) positioned in said dilute phase zone of said disengaging vessel for making a gross separation of said coked catalyst from said catalytically cracked oil, said internal gross cut separator defining a vapor port (318; 406h; 518) providing an oil outlet and having a lower portion (328; 406e; 528) providing a catalyst outlet;

a regenerator (340; 440; 540) comprising a vessel, an upright elongated lift pipe for transporting coked catalyst from said disengaging vessel to said regenerator, an air injector communicating with said lift pipe for injecting air and facilitating combustion of said coked catalyst, and a regenerated catalyst line connected to said riser reactor for conveying regenerated catalyst to said riser reactor; and

at least one cycle oil quench injection line (384; 484; 584) extending into the interior of said disengaging vessel and substantially aligned in registration with said oil outlet of said internal gross cut separator for injecting a quench comprising cycle oil selected from the group consisting of light catalytic cycle oil and heavy catalytic cycle oil, into said catalytically cracked oil after said catalytically cracked oil has exited said oil outlet of said internal gross cut separator and has been grossly separated from said catalyst for substantially enhancing the yield of naphtha and substantially decreasing thermal cracking of said product stream of oil.

22. A catalytic cracking unit in accordance with claim 21 wherein said cycle oil quench line (384; 484; 584) is at an angle of inclination ranging from about 15 degrees to about 90 degrees relative to a vertical reference line extending through said disengager for substantially minimising backflow of said quench.

23. A catalytic cracking unit in accordance with claim 21 wherein said regenerator (540) is located below said disengaging vessel (508) and said catalytic cracking unit includes a substantially horizontal conduit (505) extending between and connecting the top portion of said riser reactor (500) to said internal gross cut cyclone (506).

Fig. 1

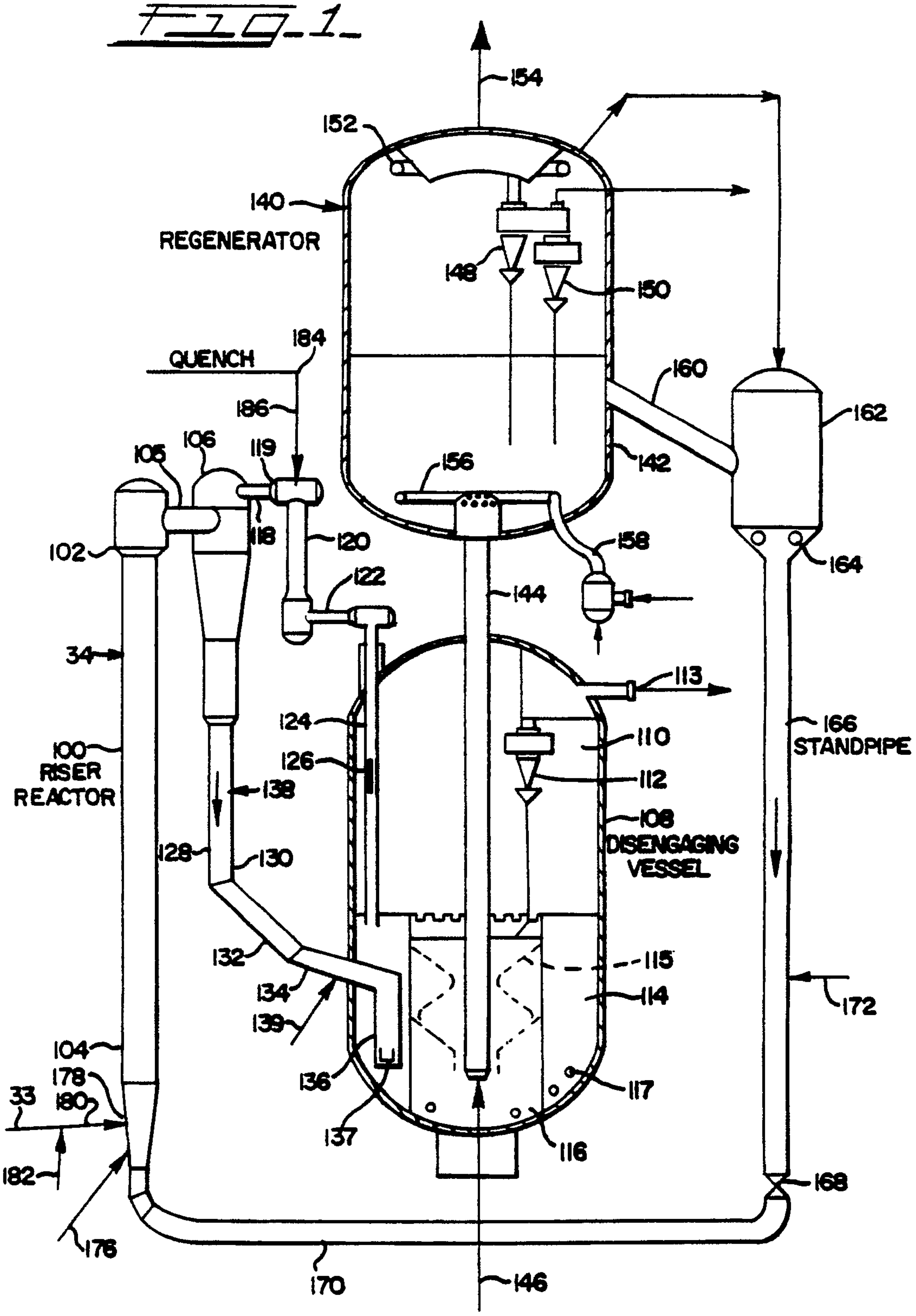
154
152
140
REGENERATOR
148
150
QUENCH
184
160
186
162
142
156
105
106
119
118
102
120
158
164
122
144
34
113
124
110
166
STANDPIPE
126
112
100
RISER
REACTOR
138
108
DISENGAGING
VESSEL
128
130
132
115
134
114
139
172
104
178
136
117
33
180
137
116
182
168
176
170
146

FIG. 2

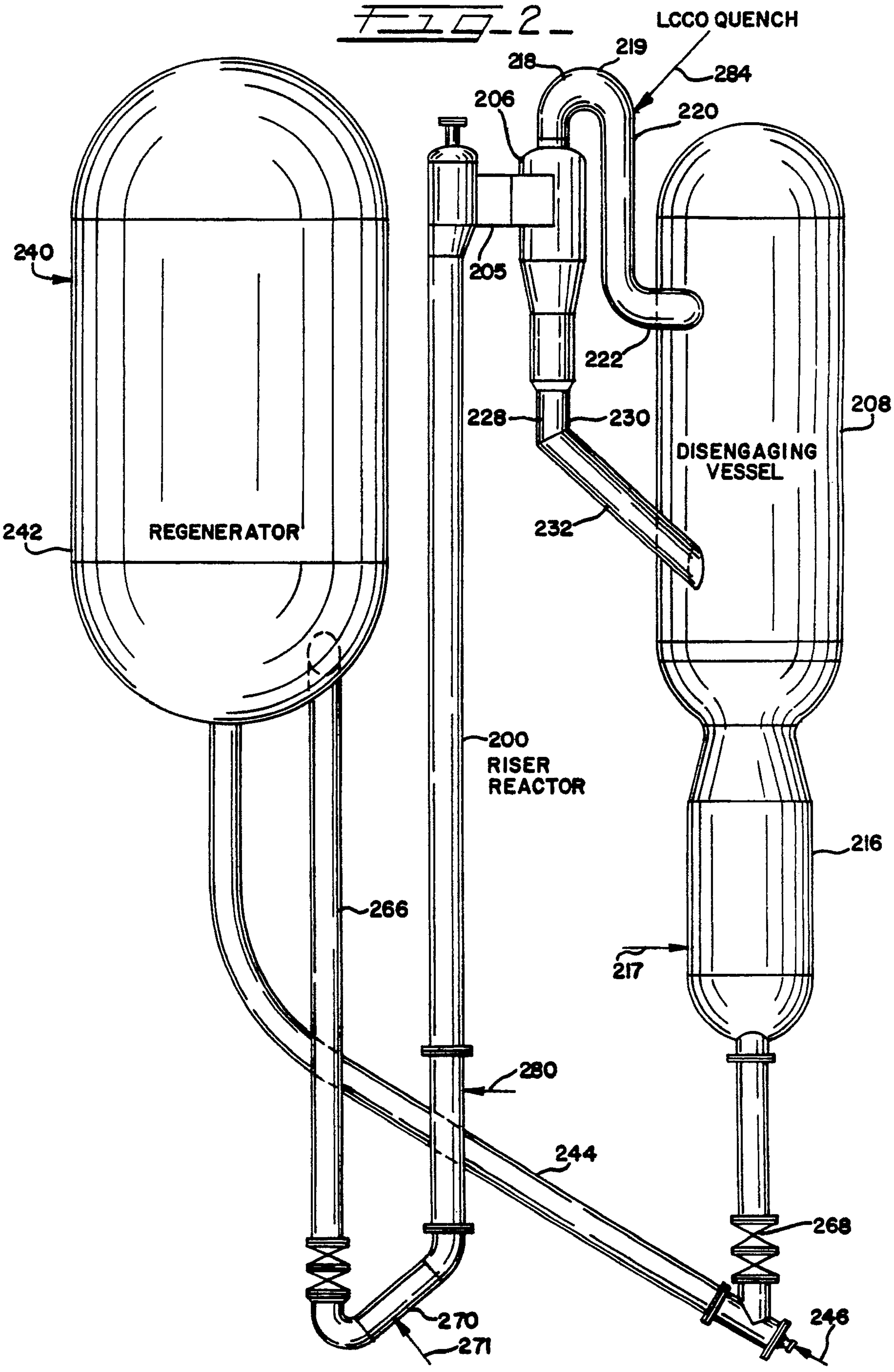
LCCO QUENCH
218
219
284
206
220
240
205
222
228
230
208
DISENGAGING VESSEL
232
242
REGENERATOR
200
RISER REACTOR
216
266
217
280
244
268
270
271
246

FIG_3

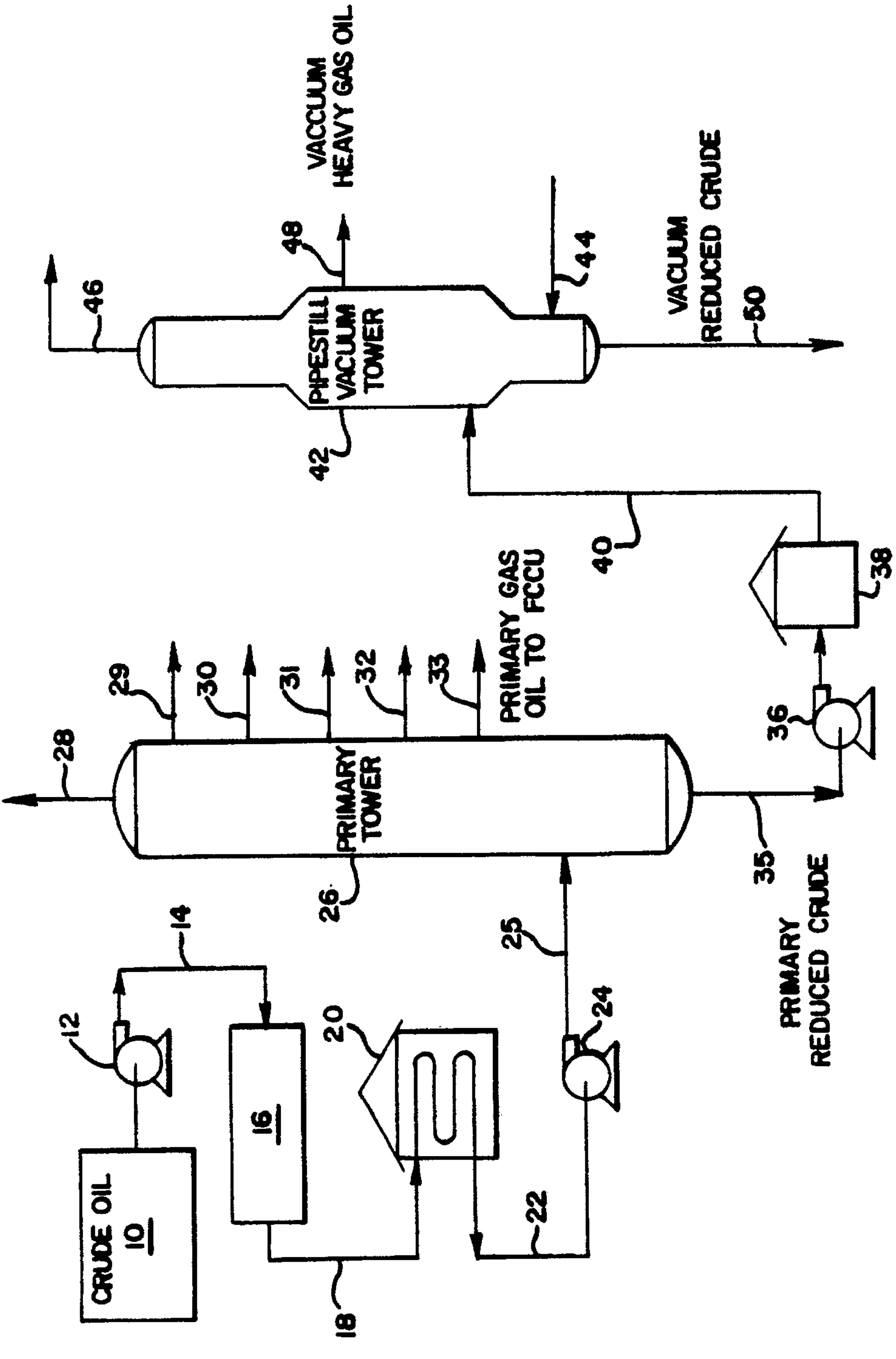
CRUDE OIL
10
12
14
16
18
20
22
24
25
26
PRIMARY TOWER
28
29
30
31
32
33
PRIMARY GAS OIL TO FCCU
35
PRIMARY REDUCED CRUDE
36
38
40
42
PIPESTILL VACUUM TOWER
44
46
48
VACCUUM HEAVY GAS OIL
50
VACUUM REDUCED CRUDE

Fig-4-

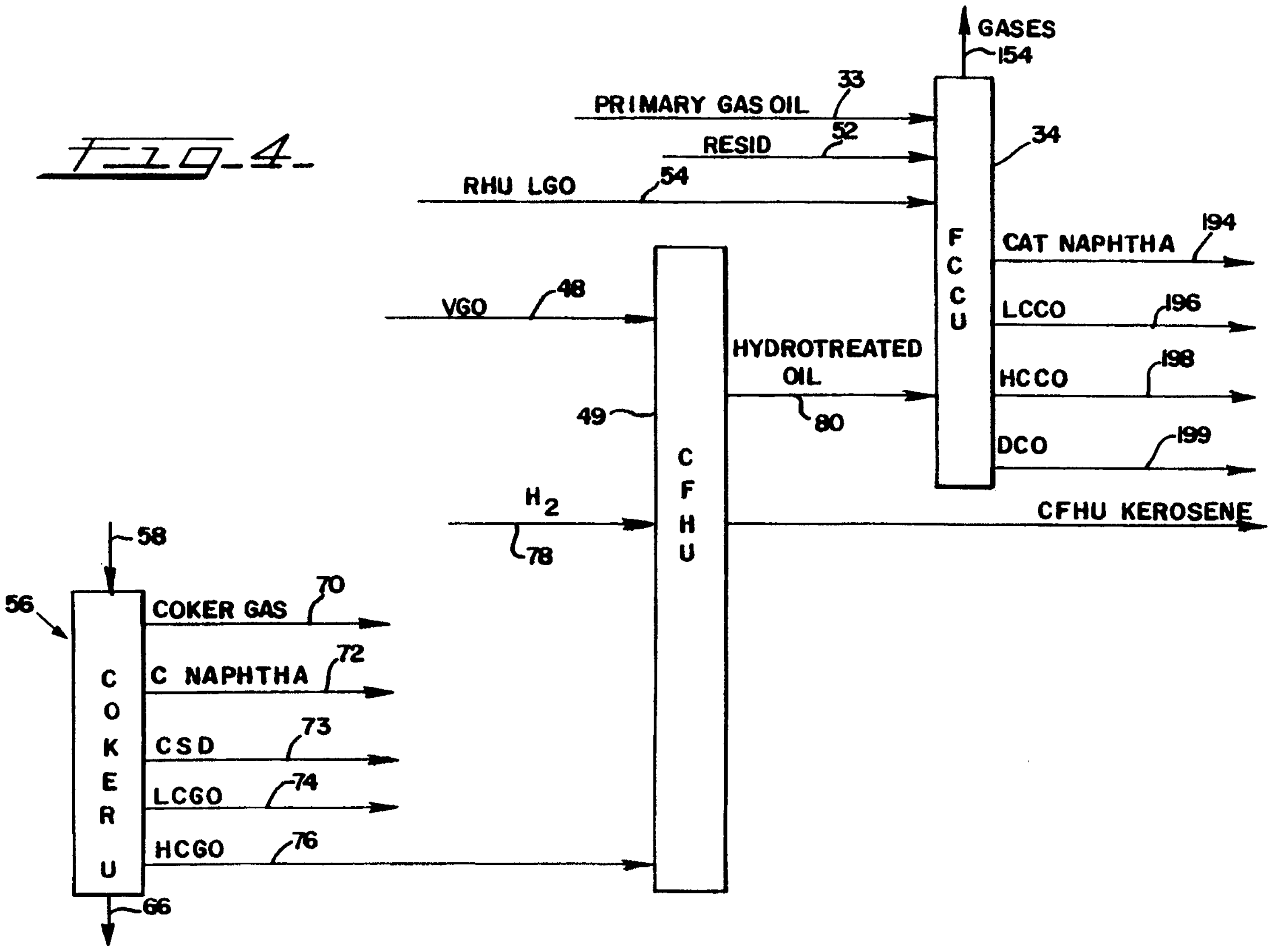
GASES
154
33
PRIMARY GAS OIL
52
RESID
54
RHU LGO
34
F C C U
194
CAT NAPHTHA
196
LCCO
198
HCCO
199
DCO
48
VGO
HYDROTREATED OIL
80
49
C F H U
H$_2$
78
CFHU KEROSENE
58
56
C O K E R U
70
COKER GAS
72
C NAPHTHA
73
CSD
74
LCGO
76
HCGO
66

Fig. 5

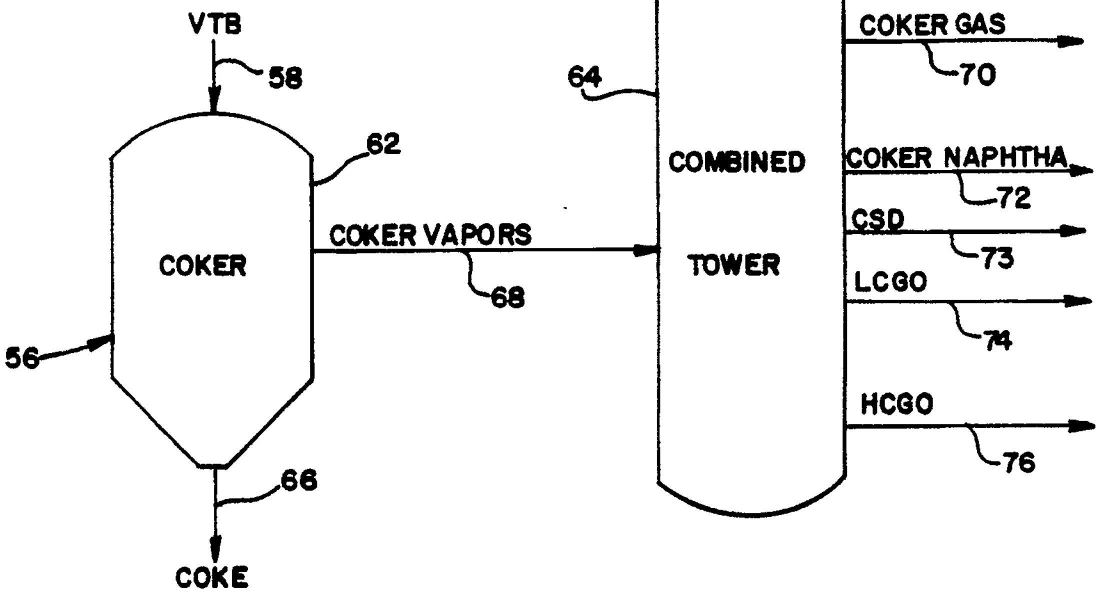
VTB
58
62
COKER
COKER VAPORS
68
56
66
COKE
64
COMBINED
TOWER
COKER GAS
70
COKER NAPHTHA
72
CSD
73
LCGO
74
HCGO
76

Fig. 6

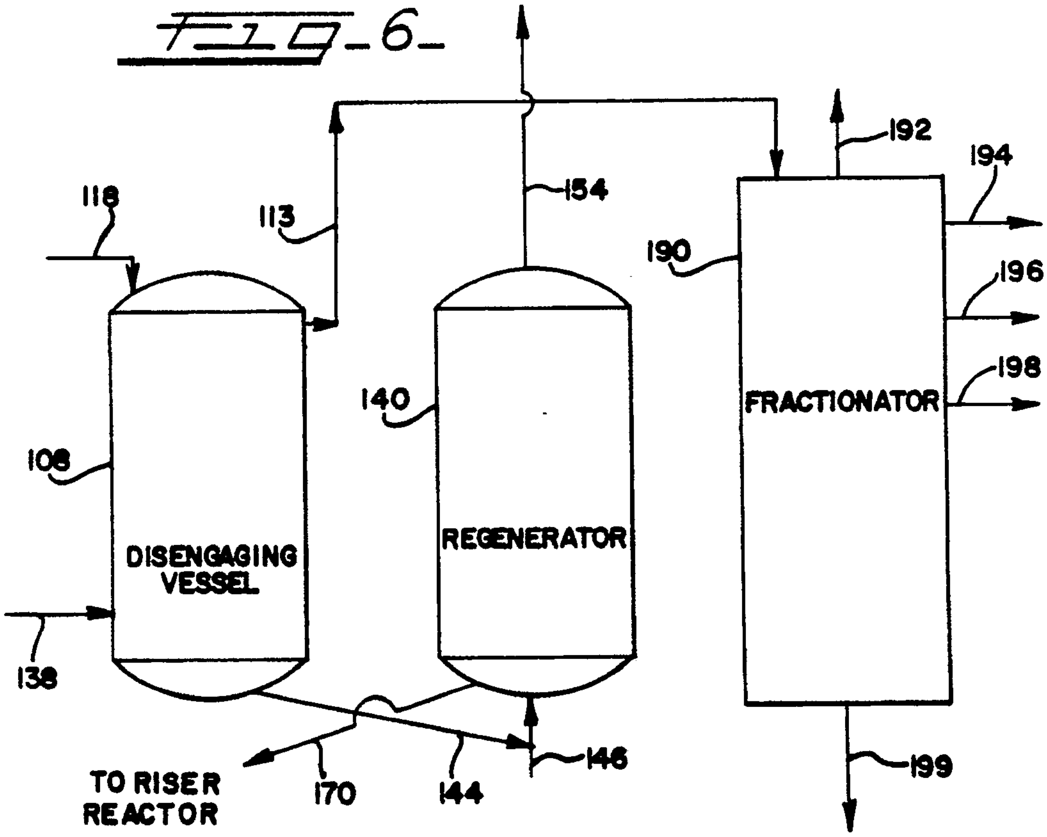
118
113
154
192
194
190
196
198
140
108
DISENGAGING VESSEL
REGENERATOR
FRACTIONATOR
138
TO RISER REACTOR
170
144
146
199

FIG-7-

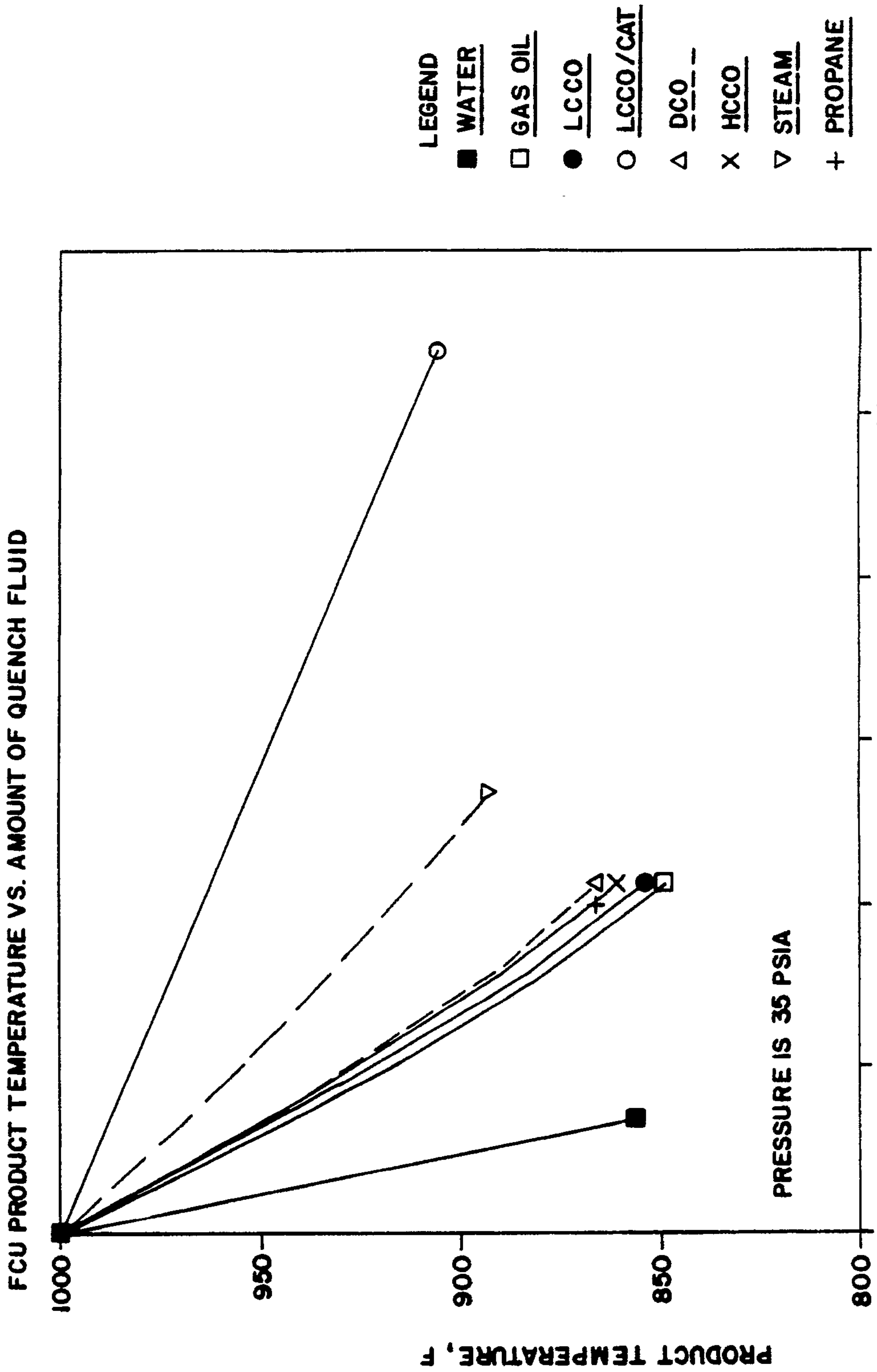
FCU PRODUCT TEMPERATURE VS. AMOUNT OF QUENCH FLUID
PRODUCT TEMPERATURE, F
1000
950
900
850
800
QUENCH ADDITION, WT % OF TOTAL PRODUCT
0
10
20
30
40
50
60
PRESSURE IS 35 PSIA
LEGEND
■ WATER
□ GAS OIL
● LCCO
○ LCCO/CAT
△ DCO
× HCCO
▽ STEAM
+ PROPANE

Fig-8-

FCU PRODUCT TEMPERATURE VS. AMOUNT OF QUENCH FLUID

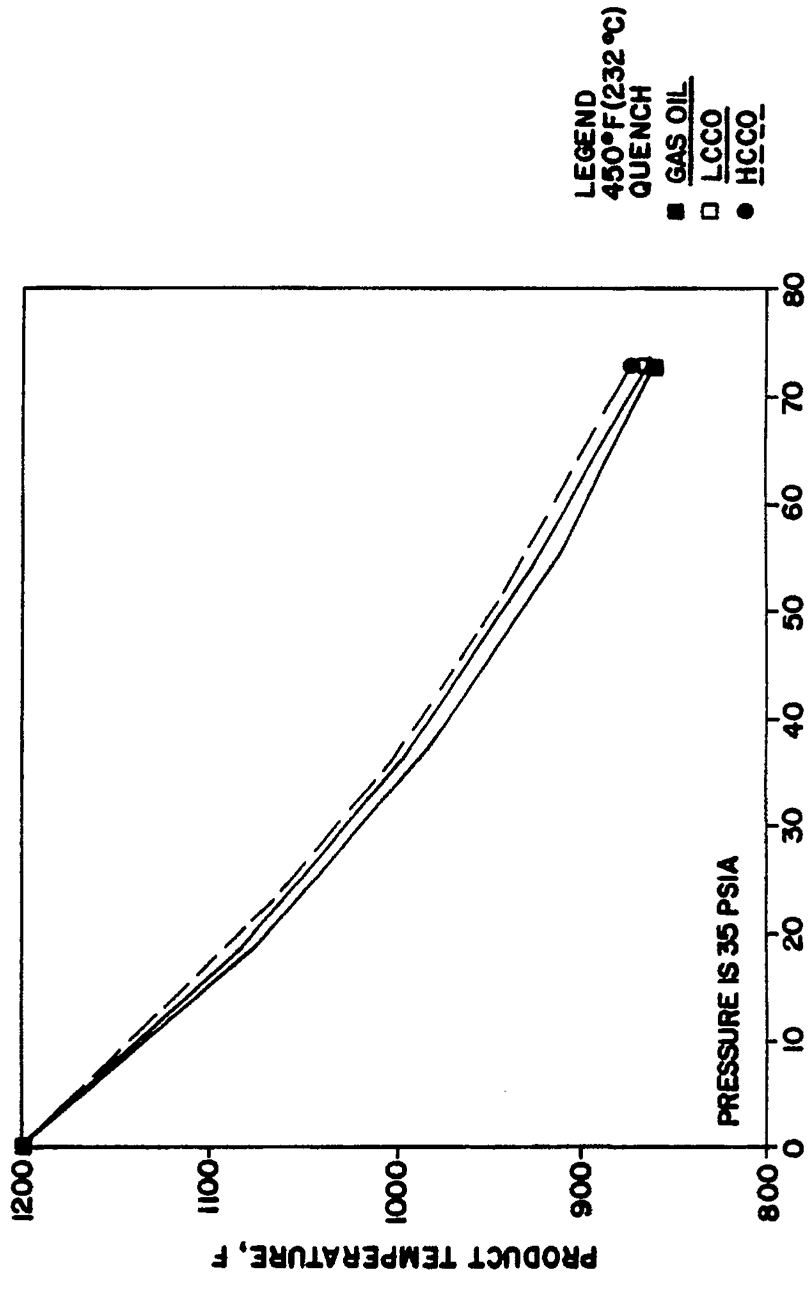

Fig. 9

QUENCHED VOLUME TO PRODUCT VOLUME RATIO

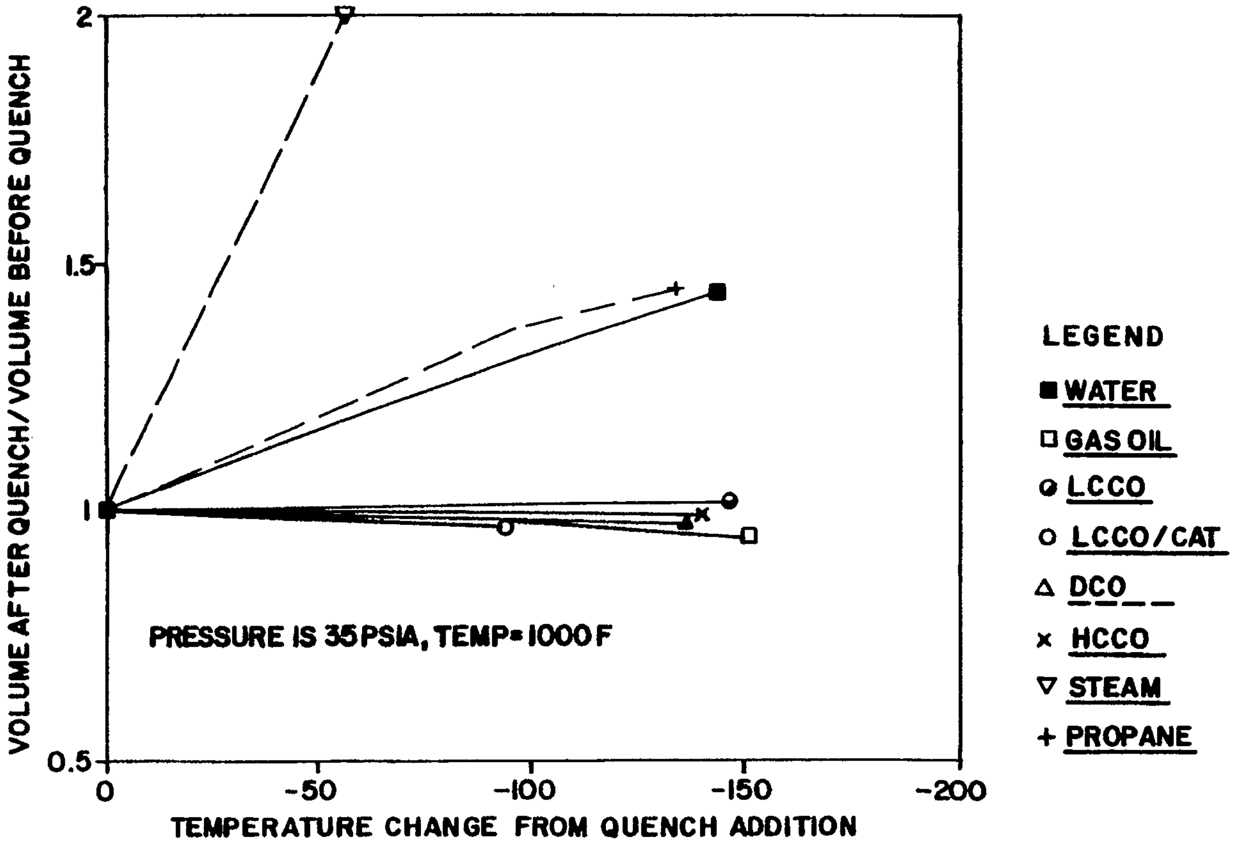

Fig-10

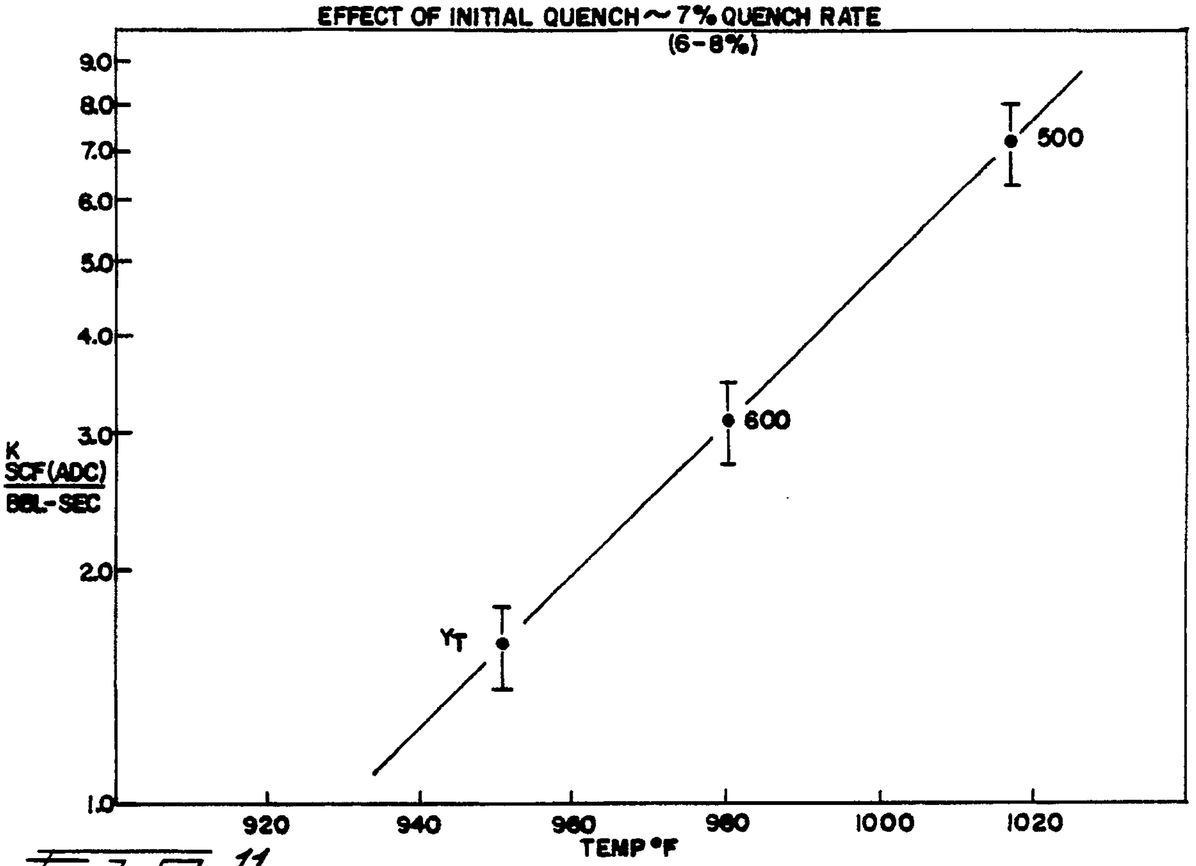
EFFECT OF INITIAL QUENCH ~ 7% QUENCH RATE
(6-8%)
K
SCF(ADC) / BBL-SEC
9.0
8.0
7.0
6.0
5.0
4.0
3.0
2.0
1.0
920
940
960
980
1000
1020
TEMP °F
500
600
Y_T

Fig-11

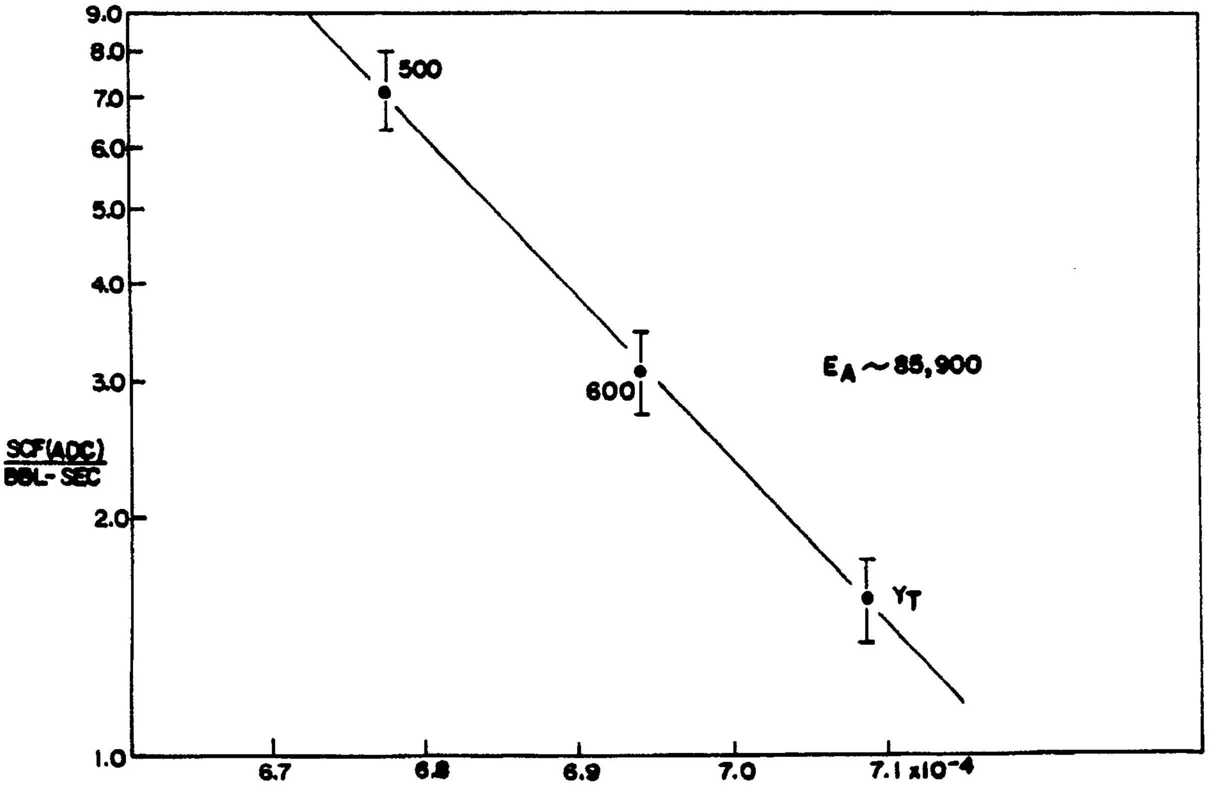
9.0
8.0
7.0
6.0
5.0
4.0
3.0
2.0
1.0
SCF(ADC) / BBL-SEC
6.7
6.8
6.9
7.0
7.1 x10^{-4}
500
600
Y_T
E_A ~ 85,900

FIG. 12

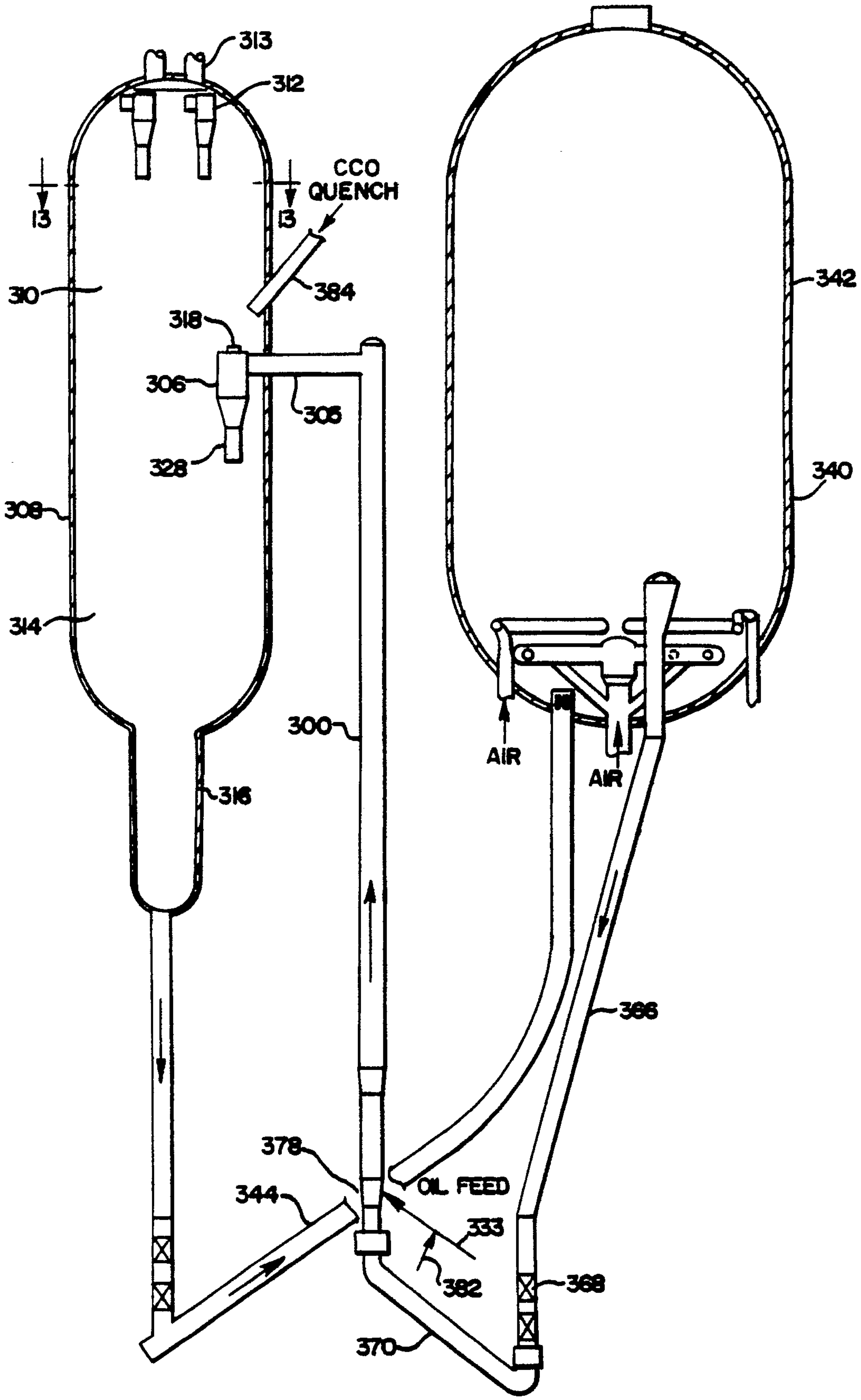
313
312
13
13
CCO
QUENCH
342
310
384
318
306
305
328
340
308
314
300
AIR
AIR
316
386
378
OIL FEED
344
333
382
368
370

FIG. 13

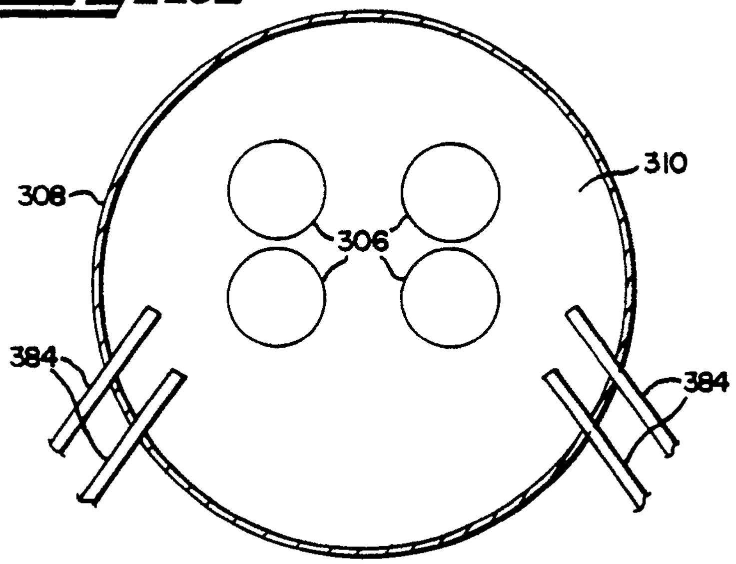
308
310
306
384
384

FIG. 14

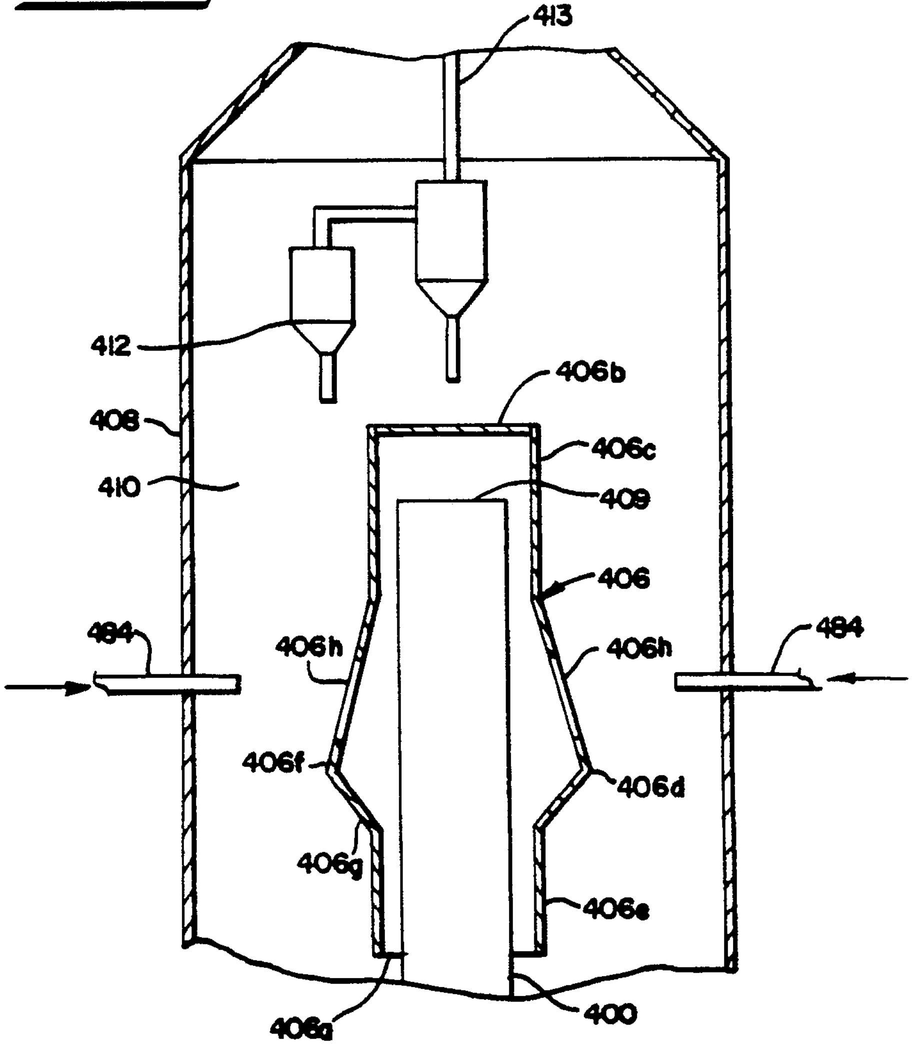
413
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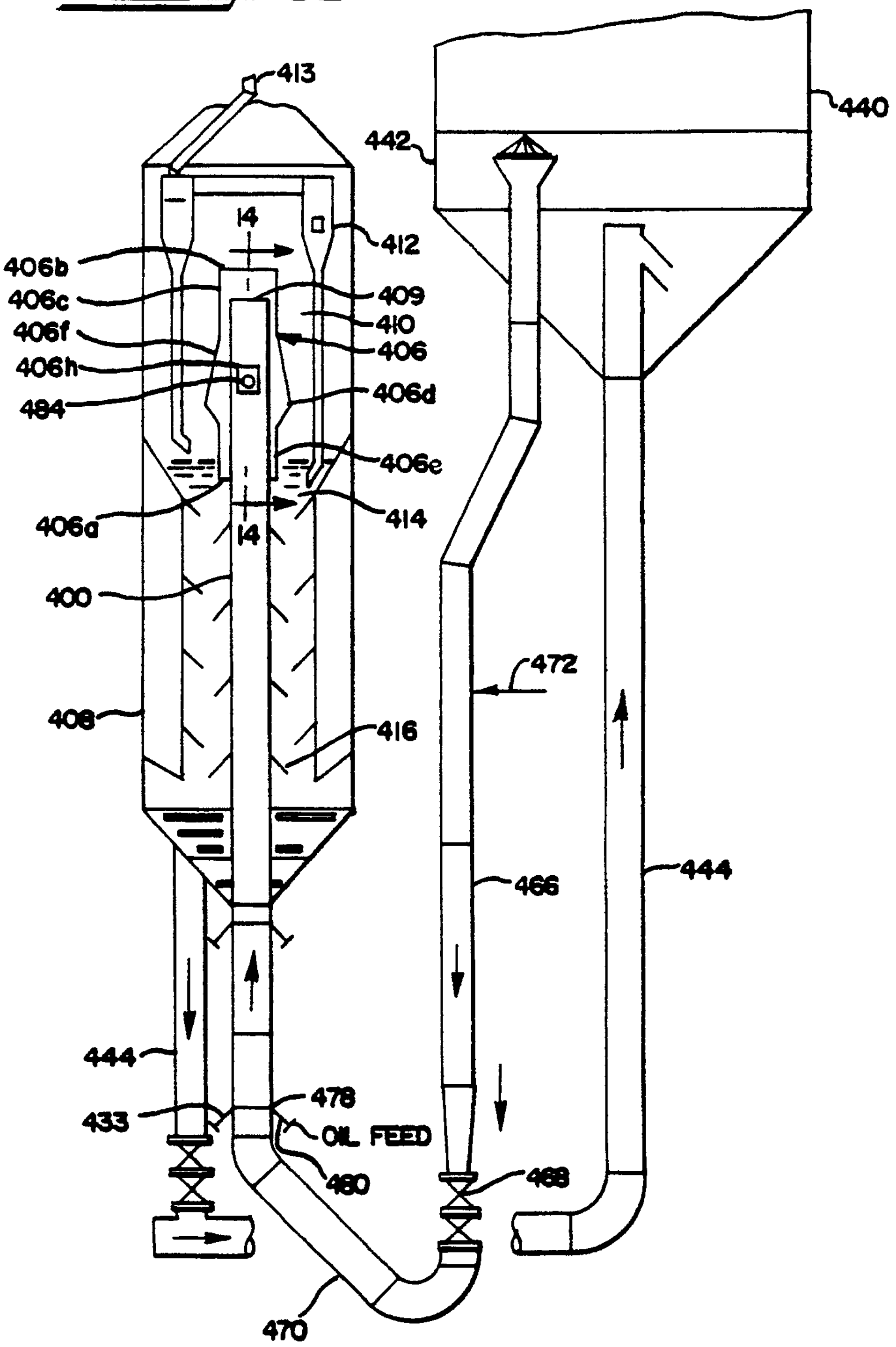
Fig. 15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OIL FEED
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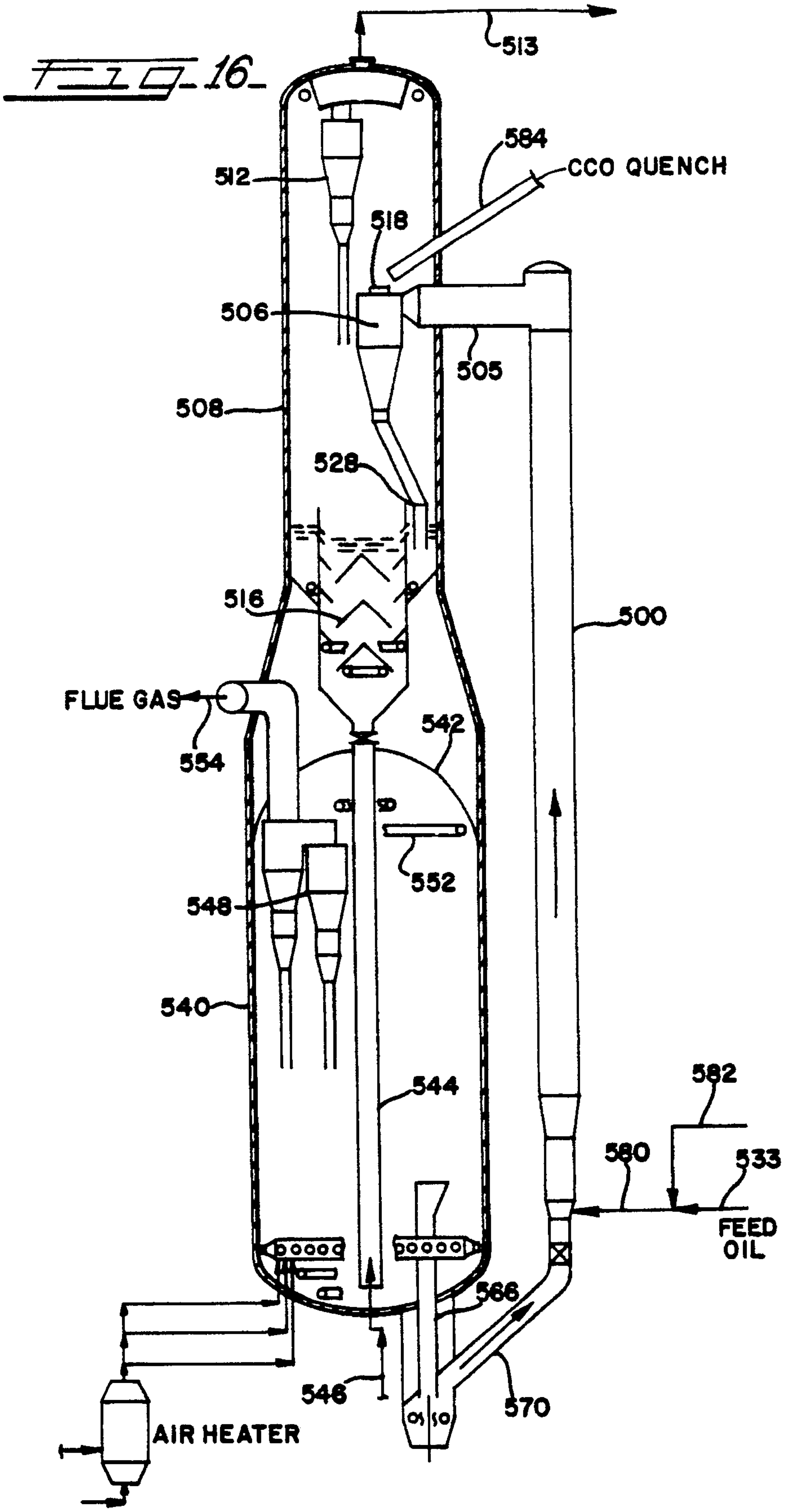
Fig. 16
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CCO QUENCH
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FLUE GAS
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FEED OIL
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AIR HEATER

European
Patent Office

EUROPEAN SEARCH REPORT

Application Number

EP 90 30 7175

DOCUMENTS CONSIDERED TO BE RELEVANT

Category	Citation of document with indication, where appropriate, of relevant passages	Relevant to claim	CLASSIFICATION OF THE APPLICATION (Int. Cl.5)
P,X	EP-A-0 381 870 (STONE & WEBSTER) * Figure 1; claims 1,2,14; page 6, lines 21-22 *	1-12,14, 16-23	C 10 G 11/18
X	US-A-2 906 695 (BOSTON) * Claim 1; figure 1; column 3, lines 52-56 *	1-12,14, 16-23	
X	EP-A-0 100 182 (MOBIL OIL) * Figures 1-3; claims; page 16, lines 14-20 *	1-12,14, 16-23	
Y		13,15	
A	US-A-4 764 268 (LANE) * Claims 1-11 *	1,2	
Y	EP-A-0 334 665 (AMOCO) * Claims; fig. *	13,15	

TECHNICAL FIELDS SEARCHED (Int. Cl.5)

G 10 G

The present search report has been drawn up for all claims

Place of search	Date of completion of search	Examiner
The Hague	05 June 91	MICHIELS P.

CATEGORY OF CITED DOCUMENTS

X : particularly relevant if taken alone
Y : particularly relevant if combined with another document of the same catagory
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