The following statement is a full description of this invention, including the best method of performing it known to us:

FLUIDIZED CATALYST REGENERATION PROCESS
The field of art to which this invention pertains is hydrocarbon processing in which a fluidized catalyst is continuously recycled between a reaction zone, where it is deactivated by carbonaceous deposits or coke, and a separate regeneration zone where its activity is restored by removal of these deposits. Specifically, the present invention relates to a regeneration process which is applicable for the oxidative removal of carbonaceous deposits from a spent fluidized catalytic cracking catalyst. Optionally, the regeneration may be conducted to completely burn, within the regenerator, CO produced in a regeneration. Optionally, regenerated catalyst temperature may be controlled. Further, the present invention also permits complete burning of CO to CO$_2$ in the regenerator, while producing a catalyst with a controlled amount of coke.

A well-known process is fluid catalytic cracking of relatively high-boiling hydrocarbons to lighter hydrocarbons boiling in the heating oil and gasoline, or lighter, range. Feed contacts, in one or more reaction zones, particulate
cracking catalyst maintained in a fluidized state.

Gaseous effluent from the reaction zone is passed to a product recovery zone while catalyst is generally passed to a stripping zone, for removal of strippable hydrocarbons from the particles. Stripped catalyst is subsequently introduced into a fluidized regeneration zone where non-strippable carbonaceous material is contacted with an oxygen-containing gas, usually air, at high temperature, to burn off most of the carbon on the catalyst.

Generally, prior art regenerations occur in a single dense phase bed located in the bottom of the regenerator. Catalyst entrained in flue gas effluent from the dense bed is generally recovered by passing this flue gas through cyclones located in a rather large disengaging space positioned above the dense bed. Some catalyst is always physically lost from the regenerator.

Superficial velocities in prior art regenerators are generally about 0.45 to 1.8 meters/second (m/s) with 0.45 to 0.9 m/s being the more common range. Residence time of catalyst in the regenerator is generally 2 to 5 minutes, with 2 to 3 minutes preferred, while the residence time of gas is generally about 10 to 20 seconds.

It was known to control air flow to the regenerator
to maintain a predetermined temperature differential between the gas disengaging space and dense bed to minimize excess oxygen to limit afterburning of CO to CO₂ in the disengaging space. Generally only about 0.1 to 1% oxygen was present in the flue gas. The flue gas, which contained much CO, was either vented to the atmosphere, or to a CO boiler and used as fuel.

Typically daily make-up catalyst rates were about 0.5 to 2.0% of total catalyst inventory. Catalyst seems to be lost in large part due to rubbing against other catalyst, so catalyst loss is almost proportional to catalyst inventory. Catalysts are relatively expensive, and strenuous efforts were made to reduce the initial catalyst inventory to reduce catalyst loss.

Further, in prior art units, it was generally impossible to vary regenerated catalyst temperature. Similarly, prior art designs never could provide for very complete CO combustion in the regenerator while simultaneously permitting control of coke level on regenerated catalyst.

Since most catalyst inventory is in the regenerator refiners have tried to reduce regenerator catalyst inventories by operating the regenerator at high carbon-burning rates. Recently, higher pressures and temperatures have been used to obtain higher burning rates.
Regenerator pressures of about 3 to 3.7 atm. and temperatures of about 620 to 675°C, or higher are now common. Some inventory reduction has been achieved, but at the cost of higher equipment costs due to higher pressures and increased catalyst deactivation due to long residence time in the high temperature regenerators.

One prior art process uses countercurrent regeneration where spent catalyst is partially regenerated in a first fluidized bed with partially spent regeneration gas, further regenerated in a transfer line with fresh regeneration gas and then passed to a second fluidized bed where further regeneration may occur with partially spent regeneration gas. Preferred superficial velocities are 0.75 to 0.9 m/s. Preferred temperatures are about 593 to 635°C. No after-burning of CO to CO₂ is mentioned, but "substantially complete" removal of carbon to a level below 0.2% and possible as low as 0.1% is contemplated. Regenerated catalyst may then be stripped of high oxygen-containing gases in a separate stripper zone with steam or flue gas.

In another known process spent catalyst is partially regenerated in a first dense bed with a first oxygen-containing gas, and further regenerated in a second dense bed
with a second oxygen-containing gas. A common dilute phase is above both dense beds. Superficial velocities are preferably about 0.6 to 1.35 m/s and temperatures are preferably about 607 to 732°C. This process controls air flow to the regenerator to insure that only a small amount of CO afterburning occurs.

The process of the present invention employs high velocities and higher oxygen concentrations than prior art schemes. Enough fresh regeneration gas is added to provide for almost complete combustion of CO. This contrasts with the prior art, which limited the amount of fresh regeneration gas to that required for a limited, small amount of controlled afterburning. Thus higher oxygen concentrations and higher temperatures and coke burning rates are obtained. Catalyst regeneration can therefore be completed in a shorter amount of time permitting dramatic catalyst inventory reduction, lower catalyst make-up rates, and improved regeneration and catalyst stability.

The combustion of CO in the regenerator eliminates an air pollution problem without a CO boiler and also reduces feed preheat requirements.

The process of this invention optionally provides for the complete combustion of CO plus additional combustible fluid in the dilute phase transport riser to heat regenerated catalyst to a predetermined temperature.
Further, the present invention, by separating CO oxidation (occurring primarily in the dilute phase transport riser) from coke oxidation (occurring primarily in the first dense bed), permits addition of regeneration gas to control coke oxidation rather than to limit CO afterburning.

Accordingly, the present invention provides a process for regenerating coke-contaminated particle form spent catalyst withdrawn from a hydrocarbon conversion zone comprising the steps of:

(a) introducing said spent catalyst and an oxygen containing regeneration gas into a first dense bed of fluidized particles and therein partially regenerating said catalyst by combustion of said coke and producing partially spent regeneration gas;

(b) passing resulting partially regenerated catalyst and partially spent regeneration gas directly from said dense bed upwardly in dilute phase into and through a dilute phase transport riser and therein effecting further combustion of coke from said partially regenerated catalyst;

(c) separating resulting regenerated catalyst from regeneration gas;

(d) recovering said regenerated catalyst as a second dense bed of particles; and,

(e) withdrawing regenerated catalyst from said second dense bed for return to said conversion zone.
In one preferred embodiment, complete combustion of CO to CO₂ occurs within the regenerator.

In another embodiment, control of regenerated catalyst temperature is possible by adding a combustible fluid to the dilute phase transport riser or first dense bed or both.

In yet another embodiment, control of regenerated catalyst coke level is possible by limiting coke oxidation occurring in the first dense bed while still providing for CO combustion in the dilute phase transport riser.

Other embodiments will be found in the following detailed description.

Figure 1 depicts schematically a specific apparatus suitable for carrying out the process of the present invention.

Figures 2 and 3 depict alternative apparatus also suitable for carrying out this invention. Figure 3 indicates how the present invention might be applied to an existing regeneration vessel.

Figure 1 shows regeneration apparatus 100. First dense bed 1 is connected, via transition region 28, to one end of vertically extending dilute phase riser 2.
Catalyst-gas separator 3 is attached to outlet 7 of transport riser 2. Regeneration gas from separator 3 enters disengaging space 5, then separator 4, and then exits at 26 into plenum chamber 27. Gas leaves chamber 27 and regenerator 100 via outlets 8 and 8'. Separated catalyst, from separators 3 and 4, is directed to second dense bed 6.

Spent catalyst enters first dense bed 1 via line 9. The level in bed 1, shown at 10, is in transition region 28 between bed 1 and riser 2. Fresh regeneration gas enters via line 11 into bed 1 through distributor 12 which disperses gas in bed 1. As air is preferred the following discussion of the invention will refer to use of air, but this term does not exclude use of other mixtures of gas, as the present invention will also work well with air enriched or deficient of oxygen. Distributor 12 may be a metal plate with holes or slots or preferably a pipe grid. Both types of distributors are well known. Controlled oxidation of the carbonaceous deposits occurs in dense bed 1. Gas and catalyst are carried out of bed 1 through transition region 28 into transport riser 2 wherein CO oxidation can occur.

Riser 2 is vertical, with its inlet at the bottom and its outlet 7 near the top. Outlet 7 may be single or multiple openings located at or near the top of riser 2.
to allow catalyst and gas to exit into separator 3. Separator 3, attached to outlet 7, is typically a cyclone separator which substantially separates gas from entrained catalyst. Separator 3 may comprise from 1 to 4 cyclones. Gas, substantially free of catalyst, passes through outlet 13 while catalyst passes via dipleg 14 directed to second dense bed 6. Separator 3 may be omitted, permitting gas and catalyst exiting outlet 7 to enter space 5. Some separation of gas and catalyst is achieved, but not as much as with cyclone 3 as shown.

Separator 4, typically a cyclone, has inlet 16 receiving regeneration gas and entrained catalyst from space 5. Gas and entrained catalyst are then substantially separated. Gas exits the regenerator via outlet 26, plenum chamber 27, and outlets 8 and 8'. Catalyst, separated from the gas, is passed via dipleg 17 down to second dense bed 6. One separator 4 is shown, but more than one could be used.

The diplegs 14 and 17 of cyclones 3 and 4 preferably have catalyst outlets above the dense bed of catalyst. The dipleg outlets may contain a flapper valve on the catalyst outlets to prevent flow of gas and catalyst up the diplegs. The cyclones are designed to operate with unsealed diplegs, because if the diplegs were "sealed" by a dense bed of catalyst, cyclone operation would vary as the level of the dense bed varied. Thus cyclone operation re-
mains constant, with diplegs 14 and 17, even if the dense bed level varies. Diplegs 14 and 17 may enter the dense bed, but this is not preferred.

A combustible fluid, such as fuel gas or a liquid hydrocarbon, may be added to riser 2 via line 20 and distributor 21. Burning of such a fluid aids in "startup" of the process and in starting CO afterburning in riser 2. Optionally, burning of this combustible material increases the temperature of catalyst passing through the riser, permitting control of regenerated catalyst temperature.

A second stream of air may be added to riser 2 via line 18 and distributor 19 to supply oxygen to support burning of added combustible fluid or of CO.

Second dense bed 6 has a level of catalyst 15 sufficiently high to overcome any pressure drop in regenerated catalyst exit line 22, control valve 23, and other downstream equipment.

Dense bed 6 may be above or below or beside dense bed 1. Similarly, the entire regenerator may be above or below or alongside of a reactor using regenerated catalyst.

Dense bed 6 may be a stripper as in Figures 1, 2, and 3, or may merely collect enough regenerated catalyst in dense phase for "head" and sealing purposes without stripping.

Catalyst in bed 6 moves down through conduit 22 and valve 23 which controls withdrawal of catalyst from bed 6.
Typically, valve 23 is a slide valve operated by a reactor temperature controller or level controller.

A stripping medium may be added to second dense bed 6 via line 24 and distributor 25 to strip adsorbed and interstitial regeneration gas from regenerated catalyst. A preferred stripping medium is superheated steam.

A majority of the catalyst in regenerator 2 preferably is in sense bed 1, with a minority in dense bed 6. When steam stripping is used in bed 6, catalyst residence time in this bed should be less than one minute, preferably less than 30 seconds.

Figure 2 is an alternate apparatus 200 for the practice of the present invention.

Spent catalyst enters first dense bed 201 via inlet line 209. The catalyst level, at 210, is in transition region 228. Air is added via line 211 and distributor 212 into dense bed 201. Controlled oxidation of the carbonaceous deposits takes place in dense bed 201 permitting control of regenerated catalyst coke level. Gas and catalyst exit bed 201 through transition region 228 into transport riser 202 wherein CO oxidation can occur.

Transport riser 202 is vertical with its inlet at the bottom and its outlets 207 and 207' near the top. Outlets
207 and 207' are at or near the top of riser 202 and allow catalyst and gas to exit riser 202. Separators 203 and 203' typically cyclones, are attached to outlets 207 and 207'. These cyclones substantially separate gas from catalyst. Gas exits the cyclones via outlets 213 and 213'. Material in disengaging space 205 is essentially gas with a little entrained catalyst. Catalyst exists the cyclones, via diplegs 214 and 214', down to second dense bed 206 having level 215. Although not shown in Figure 2, diplegs 214 and 214' may pass below level 215.

Cyclones 203 and 203' may be omitted, with some decrease in efficiency of catalyst-gas separation upon exiting riser 202. This would increase catalyst entrainment in space 205 and consequently increase the load on separator 204.

Separator 204 is typically a cyclone. It has inlet 216 which receives gas and entrained catalyst from disengaging space 205. Gas and catalyst are substantially separated. Gas exits cyclone 204 and regenerator 200 via outlet 208. Catalyst exits cyclone 204 via dipleg 217 down to second dense bed 206.

An external combustible fluid may be added to riser 202 via line 220 and distributor 221 for "startup" or to increase the temperature in zone 2 enough to start afterburning or, optionally, to increase the temperature of catalyst passing
through the riser. Air may be added via line 218 and
distributor 219 to supply oxygen for combustion occurring in
riser 202. Alternatively, gas may be added via line 218 to
cool the catalyst to a predetermined temperature, if desired.

The level, 215, in bed 206 is high enough to maintain
a sufficient head of regenerated catalyst to overcome
pressure drop in exit line 222, control valve 223, and
downstream equipment. Level 215 may be controlled to pro-
vide the desired catalyst residence time within the bed.

The catalyst in bed 206 moves down and out of regener-
ator 200 via line 222. Valve 223, which controls rate of
catalyst withdrawal, is a slide valve operated by a reactor
temperature controller or level controller.

A stripping medium, preferably superheated steam, may
be added to bed 206 via lines 224 and 224' and distributors
225 and 225' to strip off adsorbed and interstitial regener-
ation gas.

As in Figure 1, most of the catalyst in regenerator 200
is in the first dense bed. Similarly, if steam stripping
occurs in the second dense bed, catalyst residence time
therein is less than 1 minute, preferably less than 30
seconds.

Figure 3 represents a possible modification of existing
regenerators to produce an apparatus suitable for carrying out
the process of this invention. Basically an existing regen-
erator 300 is modified to provide first dense bed 301, dilute
phase transport riser 302, and second dense bed 306.

Spent catalyst enters via inlet line 309, bed 301 with
a level at 310 in transition region 328 between bed 301
and riser 302. Air is added via line 311 and distributor
312. Oxidation of coke occurs in dense bed 301. Gas and
fluidized catalyst are swept out of bed 301 through transition
region 328 into riser 302 wherein CO oxidation can occur.

A combustible fluid may be added to riser 302 via line
320 and distributor 321 to heat up catalyst in this riser.
More air may be added to riser 302 via line 318 and distribu-
tor 319.

Catalyst and gas exit riser 302 via outlet 307 into
disengaging space 305. Outlet 307 is not a cyclone. Prefer-
ably outlet 307 ejects catalyst and gas downward to
reduce catalyst entrainment in space 305.

Separator 304, typically a cyclone, has inlet 316 and
receives gas and entrained catalyst from space 305. Gas
exits cyclone 304 and regenerator 300 via outlet 308.
Catalyst exits cyclone 304 via dipleg 317 downward to second
dense bed 306. Level 315 defines the boundary between the
second dense bed 306 and the disengaging space 305.

Catalyst in bed 306 moves down and leaves regenerator
300 via conduit 322. Catalyst withdrawal is controlled by
valve 323, operated by a reactor temperature or level con-
troller.
Stripping medium, preferably superheated steam, may be added to bed 306 via line 324 and distributor 325 to strip regeneration gas from regenerated catalyst.

The process of this invention works well with known fluidized catalytic cracking catalysts. High activity zeolitic catalysts are preferred because of their tolerance for high temperatures, exposure to steam, and exposure to metals contained in the feed stock. Amorphous catalysts may also be used, and are specially benefited by the present invention because the shorter catalyst residence times, used in this process extend their useful life.

Charge stocks used in this process may be conventional, or may contain even higher levels of contaminants, such as Conradson carbon and metals such as nickel, iron, and vanadium, than found in conventional charge stocks.

A high carbon content feed stock, as measured by Conradson carbon analysis, reduces the capacity of units limited by regeneration capacity. Further, regeneration temperatures tend to increase with increasing carbon content. Metals in the feed are also deposited on the catalyst and cause it to produce less gasoline and more coke and light gas and deactivate the catalyst. To minimize Conradson carbon and metals, vacuum distillation, deasphalting, vis-breaking and coking are often used to prepare suitable feed stocks for fluid catalytic cracking.
Typical feed stocks for a cracking operation include vacuum gas oils which boil between about 200 to 550°C or higher, depending on Conradson carbon and metals. The feed may contain recycled materials, e.g., gasoline, light or heavy cycle oils or slurry oil. The shorter catalyst residence time, and smaller catalyst inventories of the process of the present invention, as compared to conventional regeneration techniques, permit heavier, more contaminated feeds to be charged to the reaction zone for a given air rate (or coke burning capacity) and a given catalyst make-up rate.

In a typical prior art process, catalyst from a regenerator contacts feed in a riser reactor. As the mixture passes up the riser, conversion of feed to lighter products, and to coke deposited on the catalyst, occurs. The riser effluent is discharged into a disengaging space where additional conversion may occur. The hydrocarbon vapors, containing entrained catalyst, then enter one or more cyclones to separate spent catalyst from hydrocarbon vapor. Hydrocarbon vapor is passed into a fractionator and separated into fractions such as light gases, gasoline, light cycle oil, heavy cycle oil and slurry oil. Various fractions may be recovered as products or recycled along with fresh feed, to the riser reactor. Light gases and gasoline are usually further processed in a gas concentration unit.
spent catalyst passes into the lower portion of the disen-gaging space and passes through a stripping zone where stripping gas, usually steam, countercurrently contacts spent catalyst to purge it of hydrocarbons. Catalyst leaves the stripping zone and enters a regenerator. Air is added to burn coke and produce a regenerated catalyst with reduced coke and flue gas of CO, CO₂, H₂O, N₂, and perhaps a little O₂. Typical spent catalyst contains 0.5 to 1.0 wt % coke, while regenerated catalyst contains about 0.2 to 0.4 wt % coke or in some recent prior art units, as little as 0.01 to 0.05 wt % coke. Flue gas is usually passed to a CO boiler and burned. Regenerated catalyst then contacts the feed in the reaction zone.

Most of the catalyst inventory of a fluid catalytic cracking (FCC) unit is in the regenerator. The trend is to short contact time reactors, so an even larger percentage of the catalyst inventory is in the regenerator. Catalyst inventory in a typical regenerator is a function of feed rate to the FCC unit, or more specifically, of expected coke yield. Coke yield fixes the fresh regeneration gas requirement of the regenerator.

A maximum superficial vapor velocity is picked for the dense bed, typically less than about 1 m/sec., in prior art units, to prevent excessive entrainment of catalyst in flue gas. With gas flow and superficial vapor velocity fixed, the minimum cross sectional area of the regenerator
is fixed. The height of the dense bed, for a fixed cross-sectional area, will determine the catalyst residence time, for a fixed catalyst flow rate, assuming the dense bed density is reasonably constant. Catalyst residence time must be enough, for a given regenerator's geometry, temperature and oxygen concentration, to ensure removal of the desired amount of coke from the catalyst. Catalyst residence times for prior art units are generally about 2 to 5 minutes.

In contrast, the present invention provides a process in which spent catalyst containing coke is regenerated in a regenerator comprising a first dense bed, a dilute phase transport riser and a second dense bed. Spent catalyst and fresh regeneration gas enter a first dense bed wherein coke is oxidized to produce CO containing gas. This gas and entrained catalyst pass from the first dense bed to a dilute phase transport riser wherein further oxidation of coke occurs. Catalyst and regeneration gas leaving the riser are separated and regenerated catalyst is passed to a second dense bed. Depending upon operating conditions, carbon monoxide can be oxidized in the first dense bed or the dilute phase riser or in both. In a preferred embodiment coke oxidation occurs primarily in the first dense bed and CO oxidation primarily occurs in the dilute phase transport riser.

When CO oxidation occurs in the present invention, this heat of combustion is transferred to the catalyst, increasing its temperature significantly. If higher catalyst tempera-
tures are required, it is a unique feature of the present invention to provide for increasing catalyst temperature by burning a combustible substance in the dilute phase riser. Combustion occurring here results in very efficient heat transfer to catalyst in this zone. Further, the present invention also provides the unique option of burning CO in the regenerator while maintaining a controlled coke level on regenerated catalyst. This may be accomplished by supplying stoichiometrically insufficient air to the first dense bed so that significant quantities of coke remain on the catalyst or by maintaining a low temperature in the first dense bed to limit coke combustion. When this catalyst enters the dilute phase riser, additional air is preferably added resulting in combustion of CO in the regeneration gas from the dense bed. The oxidation of gaseous CO in the riser occurs very quickly, so that even though gas residence time in this riser is short, almost all CO present is burned to CO$_2$. Some combustion of residual coke also occurs in the riser, but the catalyst residence time in this riser is so short that a large fraction of the residual coke on catalyst entering the riser remains on the catalyst as it passes through the riser, despite the high temperatures in the riser.
Some truisms should be mentioned.

Coke burning is apparently never complete, i.e., the catalyst will always contain some coke even after regeneration, though perhaps only 0.01 wt % or even less. If coke and oxygen are in contact, oxidation will occur, very, very slowly at low temperatures and more rapidly at higher temperatures. Similarly, CO combustion will occur whenever oxygen is present. Coke and CO are both reacting in the first dense bed and in the dilute phase transport riser, both are competing for oxygen. Relative rates of reaction are influenced by temperatures and availabilities of reactants. The terms "coke-burning" or "afterburning" should not be considered mutually exclusive, but rather simplified descriptive terms indicating what reactions are proceeding at a substantial rate.

Most refiners want a regenerated catalyst with as little coke as possible, so most refiners practicing the present invention will put almost all of the regeneration
gas into the first dense bed. The dense bed of catalyst will generally act as a heat sink and keep temperatures in this bed low enough so that significant afterburning of CO to CO₂ does not occur in this bed. In the riser, less catalyst is in contact with gas, so CO afterburning occurs. If the catalyst is not hot enough with CO afterburning, the refiner may add a combustible substance to the dilute phase riser to cause increased burning therein and heat up the catalyst. Most refiners will be content with catalyst temperatures produced by complete combustion of coke to CO₂, and will not require addition of combustible fluid in the riser.

Similarly, if a refiner does not want a catalyst with a very low coke level, he can still keep some coke on his catalyst while recovering, in the transport riser of the regenerator, the heat of CO combustion. This lets a refiner "deactivate" his catalyst, to, e.g., shift production from maximum gasoline to maximum fuel oil.

Since catalyst need not remain in the first dense bed, the superficial velocity of fresh regeneration gas is not limited to the low velocities of prior art processes. Thus superficial velocities of about 0.9 to 3 m/s are tolerable so that catalyst can be carried from the first dense bed into the dilute phase transport riser. Velocities in the dilute phase transport riser will be about 3 to 7.5 m/s.
Higher velocities used in the regenerator of the present invention produce more turbulent flow, better mixing and more efficient regeneration. This improved gas-solid contact, higher oxygen partial pressures and higher temperatures increase the rate of coke combustion. Since coke is removed sooner, catalyst residence time can be reduced, from the 2 to 5 minutes of the prior art to less than 2 minutes for catalyst and from about 20 seconds to less than 10 seconds for gas, as compared to prior art schemes. Shorter catalyst exposure time to high temperatures also prolongs catalyst activity to reduce catalyst addition. Shorter catalyst residence time also decreases the catalyst inventory.

Catalyst inventories using the process of this invention will be about 40 to 60 percent of those of present single or multistage regeneration processes. A moderately sized FCC may contain about 140 metric tons of catalyst. By using the process of this invention, a refiner can save the initial cost of about 70 metric tons of catalyst. Catalyst addition rates will also be reduced because losses tend to be a percentage of catalyst inventory. Conversely, heavier, more contaminated feeds may be used without requiring a catalyst addition rate in excess of that presently accepted on an FCC process using conventional regeneration techniques. Feed stocks no longer are limited
to relatively clean vacuum gas oils containing limited quantities of Conradson carbon, metals, and nitrogen compounds. Higher molecular weight feeds containing more contaminants can now be tolerated.

Because of the unique flow of catalyst and regeneration gas of the present invention, stripping, especially steam stripping, of regenerated catalyst is now feasible. Although steam deactivates catalyst, steam is preferred because of ease of separation of steam from hydrocarbon products in subsequent processing steps. Steam stripping of deactivated catalyst is possible because the small amounts used, and short catalyst residence time and low temperatures the reactor minimize any deactivation. Similarly steam stripping of flue gas from regenerated catalyst prior to entry of catalyst into the reactor is desirable to prevent entry of flue gas into the hydrocarbon product. Steam stripping of regenerated catalyst has generally not been practiced because of the longer catalyst residence time in the regenerator. Exposure of this catalyst to steam for long periods of time decreases catalyst activity. The present invention, by minimizing catalyst residence time in the second dense bed of the regenerator permits steam stripping of regenerated catalyst. Further, the vapors from steam stripping do not contact catalyst in either the first dense bed or in dilute phase transport riser, so steam deactivation in these zones is eliminated.
Steam stripping of regenerated catalyst is desirable because regenerated catalyst contains about 1 to 3 weights of flue gas components per thousand weights of catalyst. In a FCC unit of modest size about 1,350,000 Kg/hr. of catalyst is circulated which entrains about 2000 Kg/hr. of flue gas into the reactor. Reactor effluent thus contains about 1,600 m³/hr., at s.t.p., of flue gas components. Reaction zone effluent is eventually separated into gas and unstabilized gasoline by distillation. To recover and separate light hydrocarbons from gasoline, these materials are sent to a gas concentration unit comprising a compressor, absorber and fractionation columns. A typical analysis of gas from the last absorber follows:

Table No. 1

<table>
<thead>
<tr>
<th>Composition</th>
<th>Mole %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon dioxide</td>
<td>3.8</td>
</tr>
<tr>
<td>Oxygen + argon</td>
<td>0.3</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>22.8</td>
</tr>
<tr>
<td>Carbon monoxide</td>
<td>4.5</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>8.9</td>
</tr>
<tr>
<td>Methane</td>
<td>23.6</td>
</tr>
<tr>
<td>Ethylene</td>
<td>11.8</td>
</tr>
<tr>
<td>Ethane</td>
<td>12.5</td>
</tr>
<tr>
<td>Propylene</td>
<td>4.8</td>
</tr>
<tr>
<td>Propane</td>
<td>1.4</td>
</tr>
<tr>
<td>Isobutylene + 1-butene</td>
<td>0.7</td>
</tr>
<tr>
<td>cis - 2 - Butene</td>
<td>0.4</td>
</tr>
<tr>
<td>Trans - 2 - Butene</td>
<td>0.5</td>
</tr>
<tr>
<td>Isobutane</td>
<td>1.7</td>
</tr>
<tr>
<td>Normal butane</td>
<td>0.5</td>
</tr>
<tr>
<td>Isopentane</td>
<td>0.7</td>
</tr>
<tr>
<td>Normal pentane</td>
<td>0.2</td>
</tr>
<tr>
<td>Total C₆+ and C₅ olefins</td>
<td>0.9</td>
</tr>
</tbody>
</table>

100.0
Entrained flue gas N₂, CO, and CO₂ comprise 31.1 mole % of this product stream. Steam stripping of regenerated catalyst can remove most of this material in the regenerator. Removal of flue gas from the reactor effluent results in considerable savings, i.e., smaller compressors and absorbers may be used, and the absorber off gas will have a higher heat content.

In the process of the present invention, oxidation of coke can occur in the first dense bed or in the dilute phase riser or in both. CO oxidation may also occur in either the first dense bed or in the transport riser or in both. Preferably, most coke oxidation occurs in the first dense bed and most of CO oxidation occurs in the transport riser. This minimizes the time catalyst is exposed to very high temperatures.

Temperatures associated with CO oxidation are higher than those required for simple coke oxidation. It is therefore preferred that high temperatures due to CO oxidation occur in the transport riser where the catalyst residence time is very short. To minimize dense bed oxidation of CO, temperature should be about 620 to 675°C. With temperatures of about 675 to 695°C, some afterburning of CO in the presence of oxygen occurs, but for complete afterburning temperatures of about 695 to 760°C are preferred. When the first dense bed temperature is 620 to 760°C, the phase transport riser temperature will correspondingly be 650 to 800°C.

Almost complete oxidation of CO in the regenerator alleviates an air pollution problem without a CO boiler. Further the process of the present invention permits transfer of heat of combustion
of CO to the catalyst in the transport riser to produce hotter regenerated catalyst. This reduces the feed preheat requirement.

Regenerator pressures used in the present invention may be about atmospheric to 4.5 atm, preferably 2 to 3.5 atm. Because of the improved gas-solid contact, higher temperatures and higher oxygen concentrations, lower pressures may be used with no penalty in coke burning capacity. Less expensive vessels and compressors may now be used.

"Spent catalyst" means catalyst withdrawn from a reaction vessel because of reduced activity caused by coke deposits. Spent catalyst entering the first dense bed may contain a few tenths to about 5 wt % coke, but usually contains about 0.5 to 1.5 wt % coke. Regenerated catalyst is catalyst from which no more coke will be removed. Regenerated catalyst will typically contain less than about 0.5 wt % coke and perhaps less than 0.05 wt % coke depending on the operating conditions in the regenerator and the conversion required in the reactor. Partially regenerated catalyst is catalyst which has had a portion of removable coke removed.

Regeneration gas means any gas which is to contact catalyst in the regenerator, such as air, or air enriched or deficient in oxygen. Fresh regeneration gas enters the first dense bed of the regenerator to burn coke from spent catalyst. Partially spent regeneration gas is gas which has
contacted catalyst in the first dense bed and which contains a reduced quantity of free oxygen. Partially spent regeneration gas typically contains H₂O, N₂, O₂, CO and CO₂. Spent regeneration gas is gas which contains almost no CO, from a few tenths to 15 mole % O₂, and also CO₂, N₂ and H₂O.

Gas-solid separators may be used to separate regeneration gas from entrained regenerated catalyst. Preferred separators are the well-known cyclone separators. A single cyclone, or more than one, in parallel or in series, may be used.

The following example compares the process of the present invention with a prior art method. A commercially available molecular sieve catalyst, after contacting a vacuum gas oil in a reactor and after steam stripping, was regenerated with air. Spent catalyst contained about 0.9 wt % coke and the coke contained 10.1 wt % hydrogen. The conventional regenerator comprised a single dense bed with a large dilute phase disengaging space above the dense bed.
Table 2
Regeneration Process Comparison

<table>
<thead>
<tr>
<th></th>
<th>Prior Art</th>
<th>Present Invention</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Temperatures, °C.</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Dense bed</td>
<td>643</td>
<td>677</td>
</tr>
<tr>
<td>Dilute phase</td>
<td>641</td>
<td>---</td>
</tr>
<tr>
<td>Transport riser</td>
<td>---</td>
<td>741</td>
</tr>
<tr>
<td>Flue gas</td>
<td>674</td>
<td>738</td>
</tr>
<tr>
<td>Second dense bed</td>
<td>---</td>
<td>727</td>
</tr>
<tr>
<td>Pressure, atm.</td>
<td>2.67</td>
<td>2.36</td>
</tr>
<tr>
<td><strong>Net dry air to regenerator, Kg/hr.</strong></td>
<td>105,600</td>
<td>105,100</td>
</tr>
<tr>
<td><strong>Dry air/coke, Kg/Kg.</strong></td>
<td>11.13</td>
<td>14.51</td>
</tr>
<tr>
<td><strong>Coke yield, wt. % of fresh feed (at 75% reaction zone conversion)</strong></td>
<td>6.61</td>
<td>5.05</td>
</tr>
<tr>
<td><strong>Coke on regenerated catalyst, wt. %</strong></td>
<td>0.2</td>
<td>0.02</td>
</tr>
<tr>
<td>Vessel size, diameter, m.</td>
<td>7.5</td>
<td>4.8</td>
</tr>
<tr>
<td><strong>Catalyst residence time, min.</strong></td>
<td>3</td>
<td>0.9</td>
</tr>
<tr>
<td><strong>Gas residence time, sec.</strong></td>
<td>15.5</td>
<td>5.5</td>
</tr>
<tr>
<td><strong>Superficial velocity, m/sec.</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Dense bed</td>
<td>0.72</td>
<td>1.5</td>
</tr>
<tr>
<td>Dilute phase</td>
<td>0.72</td>
<td>6.0</td>
</tr>
<tr>
<td><strong>Flue gas analysis, mole %</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CO₂</td>
<td>9.2</td>
<td>14.9</td>
</tr>
<tr>
<td>Argon</td>
<td>1.1</td>
<td>1.0</td>
</tr>
<tr>
<td>N₂</td>
<td>79.2</td>
<td>82.3</td>
</tr>
<tr>
<td>CO</td>
<td>10.2</td>
<td>(270 ppm. vol.)</td>
</tr>
<tr>
<td>O₂</td>
<td>0.2</td>
<td>1.8</td>
</tr>
<tr>
<td>CH₄</td>
<td>0.1</td>
<td>0</td>
</tr>
<tr>
<td>Regenerated catalyst stripping</td>
<td>no</td>
<td>yes</td>
</tr>
<tr>
<td>Feed preheated to °C.</td>
<td>211</td>
<td>136</td>
</tr>
<tr>
<td>Catalyst inventory, metric tons</td>
<td>54</td>
<td>32</td>
</tr>
</tbody>
</table>
Air is the regeneration gas for both processes, however, the process of the present invention maintains a higher oxygen concentration in the regenerator as evidenced by the higher outlet concentration, 1.8 mole % $O_2$, as compared to 0.2 mole % $O_2$ for the prior art scheme. Regenerated catalyst of the present invention is hotter, permitting reduced catalyst circulation and consequently reduced coke yield, from 6.61 to 5.05 wt. % coke. This also increases product yield. The novel process used less air at a lower pressure, but produced regenerated catalyst having less coke and produced flue gas containing almost no CO. Thus, burning CO in the regenerator eliminated an air pollution problem and reduced the feed preheat required by about 75°C. Catalyst inventory is about 40% less than in a conventional process, so the size of the regenerator is smaller. Although not shown, it is anticipated that the reduced inventory and shorter residence time of catalyst will result in reduced catalyst addition rates. Regenerated catalyst of the present invention is stripped, which will reduce equipment sizes required in the gas concentration facilities.

The present invention also permits control of regenerated catalyst temperature. The importance of regenerated catalyst delivery temperature as a reactor variable in fluid catalytic cracking has generally not been recognized. Concern for regenerator temperatures
has primarily been for regenerator metallurgy or to maintain a satisfactory coke burning rate. In short contact time systems where thermal equilibrium between catalyst and hydrocarbon vapors may not be attained, the high temperature and activity of the catalyst promote rapid and efficient cracking of the feedstock. Cracked products, upon separation from hot catalyst, return to the cooler oil vapor environment wherein secondary cracking is greatly reduced. Thus, primary cracking can be controlled by catalyst delivery temperature and secondary cracking by hydrocarbon residence time. From the standpoint of better yield distribution and product quality, it is therefore advantageous to control the temperature of the regenerated catalyst.

The effect of catalyst delivery temperature, at a constant reactor outlet temperature, on product yields is shown in Table 3 which follows:
Table 3

<table>
<thead>
<tr>
<th>Operation</th>
<th>Prior Art</th>
<th>Present Invention</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor outlet temp., °C.</td>
<td>536</td>
<td>536</td>
</tr>
<tr>
<td>Catalyst delivery temp., °C.</td>
<td>677</td>
<td>732</td>
</tr>
<tr>
<td>Feed Preheated to °C.</td>
<td>364</td>
<td>316</td>
</tr>
<tr>
<td>Gasoline Yield, LV% (C5 to 90% distilled at 193°C.)</td>
<td>67.1</td>
<td>67.0</td>
</tr>
<tr>
<td>Alkylation Feed Yield, (C₃-C₄) LV%</td>
<td>25.1</td>
<td>27.1</td>
</tr>
<tr>
<td>Total Cycle Oil Yield, LV%</td>
<td>21.3</td>
<td>23.5</td>
</tr>
<tr>
<td>Coke Yield, wt. %</td>
<td>5.3</td>
<td>3.8</td>
</tr>
<tr>
<td>Conversion, LV%</td>
<td>78.7</td>
<td>76.5</td>
</tr>
<tr>
<td>Gasoline Octane Number, (Research + 0 cc TEL)</td>
<td>93.0</td>
<td>93.3</td>
</tr>
</tbody>
</table>

The example shows that with higher catalyst delivery temperature, conversion is less but product value is higher. The gasoline yield with hotter catalyst is about the same, but the octane is higher. A definite advantage is the 2.0 LV% higher yield of alkylation feed and 2.3 LV% higher yield of total cycle oil with hotter catalyst. This higher alkylation feed yield represents potential high octane alkylate. Cycle oils are not as valuable as alkylation feed or gasoline, but are more valuable than coke. Feed preheat required is reduced with hotter catalyst.

The present invention also permits control of residual coke level on the catalyst. At constant catalyst and reactor temperature an increase in coke on regenerated
catalyst reduces conversion and gasoline yield and increases cycle oil (fuel oil) yield. A decrease in residual coke increases conversion at the expense of fuel oil yield. If regenerated catalyst has very low residual coke, it is possible to increase gas yield (LPG) at the expense of gasoline yield.

An example of these possible operations at a constant catalyst delivery temperature is shown in Table 4.

<table>
<thead>
<tr>
<th>Type of FCC Operation</th>
<th>LPG</th>
<th>Gasoline</th>
<th>Fuel Oil</th>
<th>Effect of residual coke at LPG cond.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor temp, °C.</td>
<td>521</td>
<td>521</td>
<td>521</td>
<td>521</td>
</tr>
<tr>
<td>Feed Preheat Temp, °C.</td>
<td>191</td>
<td>316</td>
<td>316</td>
<td>191</td>
</tr>
<tr>
<td>Regen. Cat. Delivery temp. °C.</td>
<td>732</td>
<td>732</td>
<td>732</td>
<td>732</td>
</tr>
<tr>
<td>Residual Coke on Regen. Cat., wt.%</td>
<td>&lt;0.02</td>
<td>0.11</td>
<td>0.62</td>
<td>0.62</td>
</tr>
</tbody>
</table>

Yield Distribution, LV% of Fresh Feed

<table>
<thead>
<tr>
<th></th>
<th>Conversion</th>
<th>C&lt;sub&gt;3&lt;/sub&gt;-C&lt;sub&gt;4&lt;/sub&gt; (LPG)</th>
<th>Gasoline (90% distilled at 193°C)</th>
<th>Light Cycle Oil</th>
<th>Clarified Slurry Oil</th>
<th>Coke, wt. %</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>85.7</td>
<td>38.2</td>
<td>52.0</td>
<td>11.4</td>
<td>5.0</td>
<td>6.8</td>
</tr>
<tr>
<td></td>
<td>75.0</td>
<td>26.3</td>
<td>62.3</td>
<td>20.0</td>
<td>5.0</td>
<td>3.8</td>
</tr>
<tr>
<td></td>
<td>64.4</td>
<td>20.4</td>
<td>48.3</td>
<td>30.6</td>
<td>5.0</td>
<td>5.3</td>
</tr>
<tr>
<td></td>
<td>77.0</td>
<td>31.4</td>
<td>58.3</td>
<td>18.0</td>
<td>5.0</td>
<td>6.8</td>
</tr>
</tbody>
</table>
Maximum LPG yield is produced at these operating conditions, using regenerated catalyst having very low wt. % residual coke. Catalyst with 0.02 wt % coke results in higher conversion and even cracks gasoline to produce LPG. Catalyst with 0.11 wt % coke is less active, so conversion and LPG yield are reduced, but gasoline yield is increased. As the residual coke level is further increased, as shown in the LCO operation, to about 0.6 wt % coke, the catalyst becomes less active so conversion and yields of both LPG and gasoline decrease, but yield of light cycle oil (fuel oil) increases. The effect of residual coke on product yields is best illustrated by comparing the first and last columns of Table 4. Operating conditions are the same, except for the coke levels. Conversion and LPG yield are less with catalyst containing 0.62 wt % coke because the catalyst is less active with the higher residual coke level.

The process of the present invention allows refiners to directly and conveniently control and optimize product yields and qualities by controlling catalyst delivery temperature or by controlling the residual coke level or both. The refiner therefore can maximize fuel oil production, or maximize gasoline yield or maximize LPG production.
THE CLAIMS DEFINING THE INVENTION ARE AS FOLLOWS:

1. A process for regenerating coke-contaminated particle
form spent catalyst withdrawn from a hydrocarbon conversion
zone comprising the steps of:

   (a) introducing said spent catalyst and an oxygen-
       containing regeneration gas into a first dense bed of
       fluidized particles and therein partially regenerating said
       catalyst by combustion of said coke and producing partially
       spent regeneration gas;

   (b) passing resulting partially regenerated catalyst and
       partially spent regeneration gas directly from said dense
       bed upwardly in dilute phase into and through a dilute phase
       transport riser and therein effecting further combustion of
       coke from said partially regenerated catalyst;

   (c) separating resulting regenerated catalyst from
       regeneration gas;

   (d) recovering said regenerated catalyst as a second
       dense bed of particles; and

   (e) withdrawing regenerated catalyst from said second
       dense bed for return to said conversion zone.

2. Process of Claim 1 wherein coke is oxidized in said
first dense bed to produce a partially spent regeneration
gas containing CO and at least a portion of said CO is
oxidized to CO₂ in said dilute phase transport riser.
3. Process of Claim 1 or 2 wherein the temperature of regenerated catalyst is controlled by adding a regulated amount of a combustible fluid to said dilute phase transport riser.

4. Process of any of Claims 1 to 3 wherein the temperature of regenerated catalyst is controlled by adding a regulated amount of a combustible fluid to said first dense bed.

5. Process of Claim 3 or 4 wherein said combustible fluid is selected from the group of fuel gas and liquid hydrocarbons.

6. Process of any of Claims 1 to 5 wherein the coke level on regenerated catalyst is controlled by limiting the flow of said fresh regeneration gas to said first dense bed.

7. Process of any of Claims 1 to 6 wherein the coke level on regenerated catalyst is controlled by limiting the temperature in said first dense bed.

8. Process of any of Claims 1 to 7 wherein additional regeneration gas is added to said dilute phase transport riser.

9. Process of any of Claims 1 to 8 wherein residence time of said spent catalyst in said first dense bed is less than two minutes.
10. Process of any of claims 1 to 9 wherein said regenerated catalyst is stripped of regeneration gas in said second dense bed.

11. Process of Claim 1 wherein regenerated catalyst is stripped with superheated steam.

12. Process of Claim 12 wherein residence time of said spent catalyst in said second dense bed is less than about one minute.

13. Process of any of claims 1 to 12 wherein total residence time of regeneration gas in said first dense bed and said dilute phase transport riser is less than ten seconds.

14. Process of any of Claims 1 to 13 wherein said first dense bed temperature is about 620 to 760°C., and the superficial gas velocity is about 0.9 to 3.0 meters per second.

15. Process of any of Claims 1 to 14 wherein said dilute phase transport riser temperature is about 650 to 800°C. and the superficial gas velocity is about 3.0 to 7.5 meters per second.

16. Process of any of Claims 1 to 15 wherein said first dense bed of fluidized particles is in a lower portion of a regeneration vessel and regenerated catalyst is separated from regeneration gas in an upper portion of said regeneration vessel.
Process for regenerating a coke-contaminated particle-form catalyst withdrawn from a hydrocarbon conversion reaction zone, which comprises introducing said catalyst into the lower portion of a vertically elongated confined regeneration chamber, having three sections comprising a lower section of a predetermined relatively smaller cross-sectional area and an upper section of a predetermined relatively smaller cross-sectional area and an intermediate transitional section, introducing into the bottom portion and withdrawing from the top portion of said chamber a stream of an oxygen-containing regenerating gas at a rate which maintains the accumulated catalyst in said lower section in a dense phase fluidized bed condition and which provides in said transitional section and in said upper section gas velocities which lift said catalyst from said transitional section and transports said catalyst in a dilute phase condition through said upper section, maintaining the oxygen content of said introduced regenerating gas at a level which (1) effects combustion of a substantial portion of said coke from said catalyst in said lower section while forming CO$_2$ and CO in said regenerating gas (2) effects combustion of a lesser portion of coke from said catalyst during transit through said upper section, and (3) effects conversion of a substantial portion of said CO into CO$_2$ during passage of regenerating gas through said upper section; and separating resulting substantially regenerated catalyst from regenerating gas withdrawn from said upper section of said regeneration chamber, and introducing said regenerated catalyst into said reaction zone.
18. A process for regenerating coke contaminated spent catalyst substantially as hereinbefore described.

Dated the 18th Day of March, 1973

UNIVERSAL OIL PRODUCTS COMPANY

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