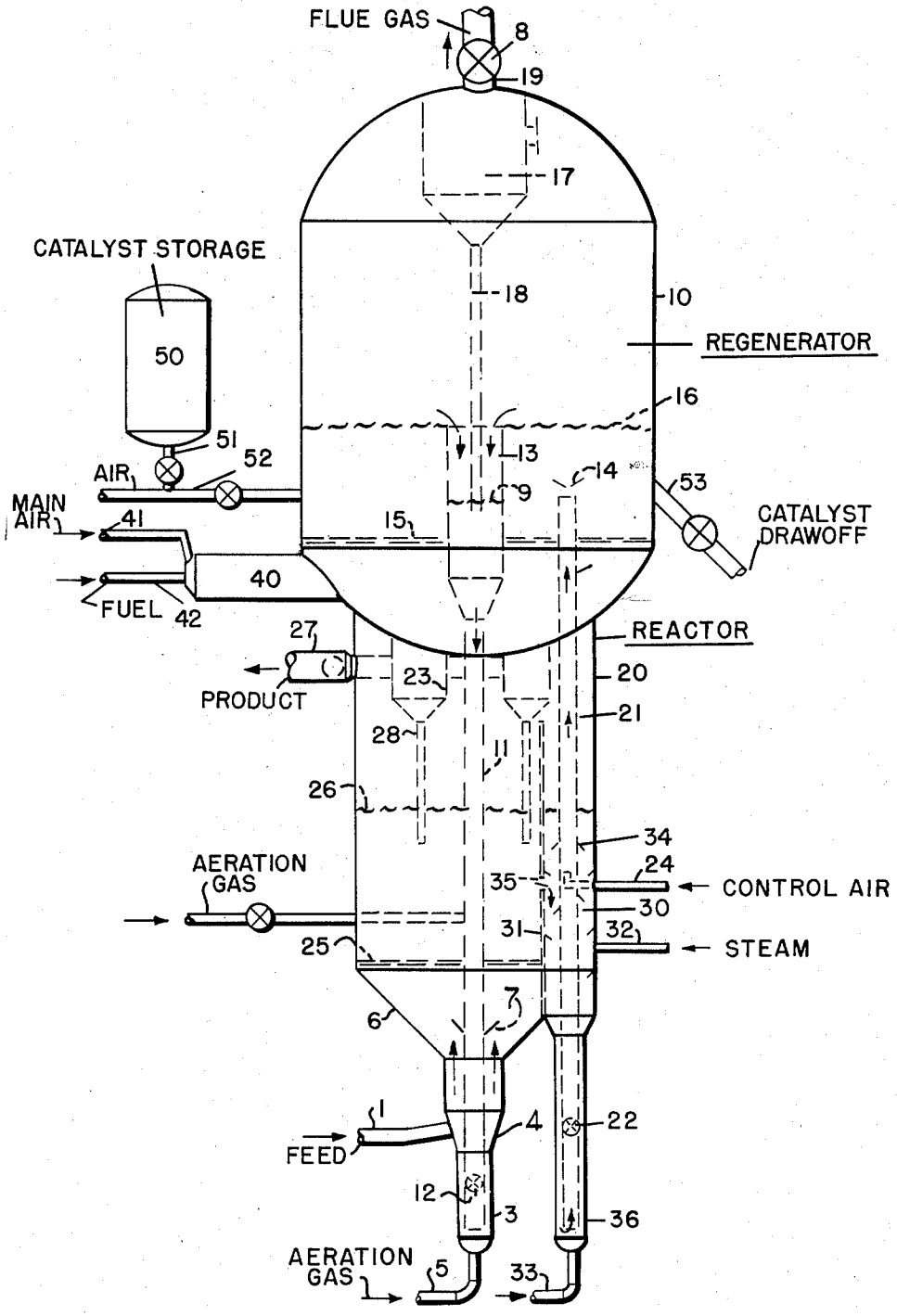
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CONTACTING OF GASES WITH FLUIDIZED SOLIDS

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CONTACTING OF GASES WITH FLUIDIZED SOLIDS

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This invention relates to the contacting of finely divided solids with gases. More particularly it relates to the conversion of hydrocarbon oils in the presence of finely divided solids which continuously circulate back and forth through internal ducts between the conversion zone and a burning or regenerating zone arranged in vertically spaced relation one with the other. Still more specifically the invention involves systems of the foregoing type wherein the solids circulation is obtained by fluistatic pressure differentials, and wherein an effective gas seal is maintained between the two zones.

It is already known to crack hydrocarbon oils in a fluid cracking unit in which the cracking and regeneration vessels are arranged in vertically spaced relation to each other and wherein the catalyst circulates between the two zones through internal vertical ducts. However, these previous systems generally control the flow of solids through the internal ducts by the throttling effect of mechanical valves of complicated design through which the oil to be cracked or the air for regeneration is injected. These designs depend upon the valve to maintain an effective seal between the zones, which involves a wasteful pressure drop across the valve as well as erosion of the valve and undesirable attrition of the circulating solids. Since the cracking zone contains vaporized hydrocarbons while the regeneration zone contains oxygen, an effective seal between the two is imperative to prevent backflow of gases from one such zone to the other.

It is the object of the present invention to provide an economic and reliable reaction system having two vertically superimposed reaction zones connected by straight internal ducts, and adapted for regulating the circulation of solids pneumatically, without throttling valves. Another object is to provide an effective seal against backflow of reactants without incurring a pressure drop across a mechanical valve. Another object is to minimize erosion and attrition, and to avoid curved conduits. These and other objects, as well as the manner in which they may be attained, will become more clearly apparent from the following description and attached drawing.

The drawing is a schematic illustration of apparatus embodying the present invention, particularly as applied to catalytic cracking of hydrocarbons.

Referring to the drawing, the illustrated system includes a regenerator vessel 10 supported on or above a reactor vessel 20, the two vessels being connected by straight ducts 11 and 21 and having auxiliary lines and other equipment as hereafter described.

In operating the illustrated system the hydrocarbon feed stock, such as gas oil, is injected through line 1 into annular feed zone 4 formed by standpipe 11 and the expanded wall portion of reactor extension or seal well 3. This feed zone 4 is preferably of larger cross-section than the seal well proper, to provide space for the sudden evaporation of the injected feed. The feed injection line 1 is quite a substantial distance above the foot of standpipe 11, such as from 10 to 20 feet and such that the hydrostatic pressure at the bottom of the seal well is from about 2 to 5 lbs./sq. in. greater than the pressure at the point of feed injection. This pressure differential forms an effective seal against reversal of flow. Still longer seal legs may be employed where unusually large pressure fluctations are anticipated and particularly when the unit is operated at elevated pressures.

Upon injection into zone 4 the hydrocarbon feed is vaporized by contact with hot catalyst. This catalyst comes from the regenerator, passing first as a dense downward column through standpipe 11 and then in dense phase upward through annular seal 3 to feed zone 4. The resulting dilute suspension of vaporized hydrocarbon and catalyst passes from feed zone 4 upwardly into reactor cone 6 at a superficial linear gas velocity of about 25 to 30 feet per second. In cone 6 the suspension is preferably spread out by deflecting baffles 7 and finally passed through distributing grid 25 into the dense turbulent bed where the main conversion takes place. The cracked hydrocarbon vapors pass through the upper level 26 of the dense bed into the superposed dilute phase and finally through cyclones 23 and product outlet line 27 into a recovery system including conventional fractionation towers and the like. Entrained catalyst particles are separated from the product vapors in cyclones 23 and returned through dip legs 28 into the dense catalyst bed at a point below bed level 26.

Spent catalyst is withdrawn from the reaction zone by flowing through orifice 35 into an eccentrically located stripping well 30 which is formed between vertical partition 31, the outside shell 20 and duct 21. Instead of passing through an orifice, the catalyst may of course overflow into well 30 over the upper edge of partition 31 if bed level 26 is maintained near the upper end of partition 31.

The stripper-sealing well 30 and pipe 21 preferably extend below the reactor, for which purpose well extension 36 may be constructed so as to extend substantially to ground level. This allows development of particularly high back pressures, and hence particularly effective sealing, without requiring additional height for the unit. Such downward extension 36 also allows more precise and more flexible control over catalyst transfer rate in riser 21, since in this fashion it is possible to provide both a deep seal and a greater height for the relatively less dense catalyst column above air inlet 24. Alternatively, however, seal well 30 may be entirely within reactor 20, since a dense phase seal of sufficient depth may frequently be obtained within the reactor proper, without any external extension 36. The fact that seal well 30, with or without extension 36, also doubles as a stripper results in especially economical construction.

Stripping steam or other inert gas may be introduced into the stripper through one or more steam lines 32, and a small amount of aeration gas is preferably also injected at the very bottom of the sealing well through line 33.

As the catalyst passes down through stripper 30, which is preferably provided with baffles 34, occluded hydrocarbons are stripped out by the counterflowing steam. Downwardly flowing stripped catalyst finally reverses its direction of flow at the bottom of the well where it flows into the straight line riser 21 and flows upward in dense phase for a substantial distance. Thereafter air or other regenerating gas is injected into the riser through line 24, at a point a substantial distance above the bottom of the riser. In commercial units this distance may be about 10 to 20 or more feet, that is, it is desirably large enough to give a hydrostatic seal pressure drop of at least 2 to 5 lbs./sq. in. between the control air inlet

and the foot of the riser. The height of the air injection point above the bottom of the riser is a measure of the seal preventing reversal flow of the air, and naturally is of great importance since no throttling valves are used in the present invention.

The rate of catalyst transfer from reactor 20 to regenerator 10 is controlled by regulating the quantity of control air injected into riser 21 through line 24. The height and diameter of the riser above the point of air injection are preferably designed so that gas velocities of about 5 to 20 feet a second are sufficient to move the catalyst at proper rate from one vessel to the other. This desired catalyst transfer rate normally equals about 6 to 18 times the weight of hydrocarbon feed injected into the reaction zone per unit time. The stated relatively low gas velocities are important in that they result in having apparent densities of about 8 to 25 lbs./cu. ft. in the riser even above the control air inlet. Higher velocities, and corresponding lower densities, are preferably avoided since they reduce the sensitivity of the mechanism for controlling catalyst transfer rate, and also lead to increasing attrition of catalyst as well as erosion of the equipment.

According to the present invention the catalyst flows mostly at relatively low velocity, or in relatively dense phase. Since the density as well as the height of the catalyst in the annular stripper 30 remains essentially constant, variation of the density in riser 21 above control air inlet 24 increases or decreases the density differential between these two communicating legs and thus provides more or less driving force for the catalyst to rise through the riser. Alternatively, control over catalyst circulation rate may also be had at constant air flow by increasing the pressure differential existing between the dilute phases of the reactor and regenerator, respectively; or both control air rate and the pressure differential may be varied simultaneously.

Upon discharge from riser 21 into the fluid bed in regenerator 10, the spent catalyst stream is preferably deflected by baffle 14 so as to become more evenly distributed throughout the bed.

The main portion of the required regeneration air is introduced into regenerator 10 through air line 41 which may feed the air through auxiliary burner 40 and distributing grid 15. Burner 40 may be used to heat the system to operating temperature when starting up a run, in which event a liquid or gaseous fuel may be injected through line 42, in addition to the air being fed through line 41. However, once the unit is in proper operation, the supply of extraneous fuel through line 42 is cut off, since the heat liberated by combustion of the coke deposited on the catalyst in the course of the hydrocarbon conversion is usually sufficient for supplying heat for the cracking reaction.

The fact that the novel unit permits feeding most of the regeneration air directly into the regenerator, without having to be injected at relatively high pressure as a lift gas into the lower part of the riser, is of course one of the important advantages, since it results in a considerable saving on compression.

Regenerated catalyst from regenerator 10 overflows into a central withdrawal well 13 wherein it forms a free level 9 some 2 to 20 feet below the main bed level 16. The well 13 advantageously has a larger cross section than the standpipe 11 of which it forms a part, and this enlarged cross-section provides surge capacity to accommodate small fluctuations in the rate at which the catalyst overflows and also provides adequate space for gas disengaging so that the catalyst may flow as a dense phase down the standpipe. In this manner normal fluctuations in pressure or circulation rate are taken care of by a corresponding automatic rise or fall in free level 9, without throwing the system out of balance. Conversely, as suggested earlier, by varying the pressure differential between the vessels, level 9 may be caused to ride up or down so as to compensate for changes in hydraulic balance of the system. The difference between levels 9 and 16 is a measure of the reserve driving force for catalyst circulation which permits the desired adjustments.

Flue gas passes from the regenerator 10 through one or more cyclones 17 and outlet 19, while separated catalyst particles are returned through dip leg 18. This dip leg preferably extends below the catalyst level 9 in well 13, since this provides extra length for the dip leg, without increasing total vessel height. Also, since the catalyst in well 13 is quiescent in comparison with the turbulence of the main bed in the regenerator, discharging dip leg 18 into well 13 further increases the effectiveness of cyclones 17. This arrangement greatly reduces the tendency for gas to back up through the dip leg, as in conventional units where dip legs are immersed in the turbulent bed which is fluidized with upflowing gas. Such upflow in dip legs seriously damages cyclone performance.

From well 13 the hot regenerated catalyst passes in dense phase down through standpipe 11 which terminates near the bottom of reactor extension or well 3. Catalyst then flows from the standpipe and, aerated with steam or other inert gas injected at 5 in a sufficient amount to insure mobility, flows in dense phase upward through the annular space of well 3 toward the oil injectors 1. As described before, here oil is injected preferably in an enlarged section or feed zone 4 and the resulting relatively less dense mixture of vaporized feed and catalyst returns to reactor 20, thus beginning another conversion cycle.

Regenerator bed level 16 is held constant by overflow of catalyst into well 13. At the same time the sealing well 3 which contains the catalyst in relatively dense phase and extends a substantial distance below feed injectors 1 prevents vapors of oil feed from backing up into regenerator 10.

Since standpipe 11 and riser pipe 21 are straight pipes requiring no throttling valves at the ends thereof, these pipes can be supported at one point only, e. g., where they pass through the common head which separates regenerator 10 from reactor 20, and this mode of support allows free longitudinal expansion of the pipe in either direction, thus avoiding the need for troublesome expansion joints or other special construction.

Of course, shut-off valves 12 and 22 may be provided on ducts 12 and 22 for use in starting up the unit or in an emergency. It will be understood, however, that these valves will always be either wide open or closed tight, since they are not required for actual regulation of catalyst circulation when the unit is in operation. These valves may be located essentially anywhere along the length of the ducts, the valves being operated by a stem which extends through the outer wall of wells 3 and 30. However, as an alternative, regenerator 10 may be spaced some distance above the upper header of reactor 20, in which event the shut-off valves may be located on ducts 11 and 21 on the exposed portions thereof between the vessels.

As is well known, the catalyst employed for this process may be a silica base catalyst prepared by the acid activation of bentonitic clays or a synthetic catalyst derived from silica gel or other forms of silicic acid. The catalyst may be of the silica-alumina or silica-magnesia type, with suitable additions of other active constituents such as zirconia, boria or the like. This catalyst may be in the form of a finely divided powder prepared by grinding or in the form of small spheres prepared by suitable drying procedures in the case of the synthetic catalysts. The catalyst contains preferably particles having a range of particle sizes, including particles within the size range from 0 to about 200 microns in diameter. The top pressure within the reactor may be about 10 to 20 lbs./sq. in. as determined by the product recovery system, that is,

the pressure drop which the product vapors must overcome. The pressure in the top of the regenerator, which may be controlled by throttle valve **8** in flue line **19**, may be about 4 to 10 lbs./sq. in., and the relationship between the reactor pressure and the regenerator pressure is an important operating variable. For instance, an increase in regenerator pressure, unless compensated by a similar increase in reactor pressure, will of course increase the distance between regenerator bed level **16** and overflow well level 9. An uncompensated increase of 1 lb./sq. in. in regenerator pressure thus will depress level 9 by about 3 feet. At the same time, this increase in regenerator pressure will offer more resistance to catalyst flow from the reactor to the regenerator. This will correspondingly reduce the back pressure or driving force at control air inlet **24**, and thus result in a decrease in catalyst circulation rate. Naturally, both reactor pressure and the regenerator pressure may be increased considerably beyond the values given above, e. g. to 100 lbs./sq. in. or higher, provided that a proper pressure differential, e. g. about 5 to 15 lbs./sq. in., is maintained between the two vessels.

The feed stock for the catalytic cracking process may be a gas oil, naphtha, heavy distillate, topped crude, whole crude, or other fractions of crude oil separately or in combination, or the process may be applied to liquid hydrocarbons or hydrocarbon blends derived in part from sources other than petroleum.

The temperature of the cracking reaction may be within the range of between 700 to 1100° F., preferably 900 to 950° F., and the temperature of the regeneration may be between about 900 to 1200° F., preferably about 1100 to 1150° F. The system is adapted to be run under heat balance conditions, that is, substantially all of the heat liberated during regeneration is transmitted to the oil and utilized for vaporization and cracking thereof. The catalyst-oil ratio required to maintain this heat balance operation will vary with the characteristics of the feed stock, the temperature to which this feed stock is preheated by indirect heat exchange with various effluent streams from the cracking and regeneration processes, and the cracking temperature desired. This catalyst-oil ratio may vary from about 5 to 1 to 30 to 1 parts by weight, and may be about 10 to 1 under preferred conditions.

Illustrative mass densities and gas velocities characteristics of a system operating in accordance with the present invention will now be given. Thus, in a typical case, the densities of the turbulent fluidized catalyst beds in reactor **20** and regenerator **10** may range from about 10 to 45 lbs./cu. ft., these densities being obtained at upward linear gas velocities which may range from about 0.1 to about 15 ft./sec., or preferably about 1 to 5 ft./sec., depending mainly on the true density of the catalyst particles, as is well known. In the aerated standpipe **11** and seal well **3** the catalyst density may be as much as 35 to 50 lbs./cu. ft., and similar densities may also exist in stripping section **30**. The density in the spent catalyst riser may be between about 2 to 20 lbs./cu. ft., but preferably not below about 8 lbs./cu. ft. Still lower densities, e. g. 2 to 8 lbs./cu. ft. will prevail in the annular feed riser **4** above the sealing section **3**, that is, above the level where injected feed is vaporized with a resultant dilution of the catalyst phase.

While the total amount of catalyst hold-up in the fluid bed of the regenerator is substantially constant due to the fixed level **16** at the top of withdrawal well **13**, the catalyst within the withdrawal well finds a free level as previously described, depending on the desired circulation, the catalyst hold-up in the reactor, and the pressure differential maintained between the regenerator and the reactor.

The catalyst hold-up within reactor **20** is controlled by the total inventory of catalyst within the system. Changes in this inventory to produce a change in reactor bed level **26** may be made by the addition of make-up catalyst or withdrawal of old catalyst at any suitable point. For instance, make-up catalyst may be added in conventional manner from storage hopper **50** to regenerator **10** through standpipe **51** and transfer line **52**, while any unwanted catalyst may be drawn off from regenerator **10** through line **53**. It will be understood, of course, that the size of storage hopper **50** shown in the drawing is not represented according to any specific scale but is shown only schematically. In reality, hopper **50** may be as high as the main conversion unit.

While the invention has been described particularly with reference to a catalytic cracking process, it will be understood that it is not limited thereto but is similarly applicable to other hydrocarbon conversion processes as well as to uses outside the petroleum industry. The novel system may be particularly advantageous in processes wherein the upper reactor is small in comparison with the lower reactor, as in hydroforming of naphthas wherein only a comparatively small regenerator is required. This permits supporting the small regenerator directly by the reactor, without special reinforcement of the latter.

For the same reason the invention, with minor modifications, is also well suited for non-catalytic coking of heavy hydrocarbons such as reduced crudes or petroleum residues. For instance, in such a case steam or other inert gas may be injected into the vessel through line **1** in order to fluidize the solids in the annual riser **3** and in cone **6**, while the residuum or other suitable heavy hydrocarbon feed may be injected through multiple nozzles directly into the fluidized bed in reactor **20**, as by nozzles which may extend across the entire bed. Instead of a catalyst, such a process normally employs a finely divided non-catalytic solid such as coke or other inert materials such as sand, spent clays, pumice and the like. The temperature in the coking zone **20** will normally be kept between about 850 and 1100° F., so as to produce the maximum amount of naphtha and particularly of gas oil suitable for further catalytic cracking. The required heat of coking normally may be supplied in regenerator **10** through which the inert solids circulate and where excess coke produced in the process may thus be burned, while the remaining reheated solids carry sensible heat back to the coking reactor **20**. Otherwise, the operation of such a non-catalytic coking process is essentially similar to the illustrative catalytic cracking process described in detail earlier.

In any event, when using the novel scheme for conversion of hydrocarbons, it is most desirable to keep the regenerator above the reactor, as otherwise it becomes necessary to pass hydrocarbon feed through a riser passing through the entire height of the regenerator. This has been found to be quite dangerous, since line failure due to unforeseen high regenerator temperatures may result in disastrous explosions.

The invention, of course, is not limited to the particular embodiments and illustrations specifically described herein, but its eventual scope is particularly pointed out in the appended claims.

We claim:

1. In an apparatus for contacting gases with finely divided solids wherein the solids are circulated through two vertically spaced axially aligned reaction chambers, maintained under different gaseous pressures and wherein the solids are circulated from one reaction chamber to the other through internal vertical conduits, the improvement which comprises means for maintaining an effective seal between the two chambers during such circulation, said means comprising a first well section of smaller diameter depending downwardly from the bottom of the lower reaction chamber, an internal vertical conduit forming a standpipe for transferring solids from the upper chamber to the lower chamber having its lower end in open communication with the lower portion of said well section and its upper end in open communication with said upper

chamber at a point spaced from the bottom thereof, means for introducing an aerating gas into the bottom portion of said well and into said standpipe in amounts controlled to maintain a dense fluidized body of solids in said well and standpipe capable of generating a hydrostatic head of pressure at the bottom of said well, means for introducing a reactant gas into the lower reaction chamber at a point above the bottom of said well section for a distance such that the hydrostatic pressure seal between the point of introduction of the reactant gas and the base of said well section is at least 2 pounds per square inch, a second well section depending from the bottom of said lower chamber, a second internal vertical conduit forming a riser for transferring solids from the lower reaction chamber to said upper reaction chamber having its lower end in open communication with the lower portion of said second well section and its upper end communicating with the upper reaction chamber, means for introducing an aerating gas into said second well in amounts controlled to maintain a dense fluidized body of solids in said well capable of generating a hydrostatic pressure at the base thereof, means for introducing a gas into said riser at a point above the bottom of said second well for a distance such that the hydrostatic pressure seal between such point and the bottom of said second well is at least 2 pounds per square inch, means for introducing

said last named gas into said riser in an amount sufficient to reduce the density of said fluidized body in the upper portion of the riser substantially below the density in said standpipe to thereby effect circulation of the solids in the direction aforesaid and means for controlling the amount of gas introduced into said riser to regulate the rate of flow of the solids between said chambers.

2. In an apparatus defined in claim 1 the further improvement which comprises a vertical baffle in said lower chamber between said first named well section and said second named well section dividing the lower portion of said lower chamber into two compartments, said baffle having passageways permitting the flow of solids from one of said compartments to the other

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CERTIFICATE OF CORRECTION

Patent No. 2,843,460 July 15, 1958

Daniel S. Borey et al.

It is hereby certified that error appears in the above numbered patent requiring correction and that the said Letters Patent should read as corrected below.

In the grant, lines 2 and 3, for "assignors to Standard Oil Development Company, a corporation of Delaware," read -- assignors to Esso Research and Engineering Company, a corporation of Delaware, -- line 12, for "Standard Oil Development Company, its successors" read -- Esso Research and Engineering Company, its successors --; in the heading to the printed specification, lines 5 and 6, for "assignors to Standard Oil Development Company, a corporation of Delaware" read -- assignors to Esso Research and Engineering Company, a corporation of Delaware --.

Signed and sealed this 4th day of November 1958.

(SEAL)
Attest:
KARL H. AXLINE
Attesting Officer

ROBERT C. WATSON
Commissioner of Patents