Title: REACTION PROCESS CONTROL

Abstract: A reaction process constraint override control method for optimizing production rates, the method includes automatically and continuously controlling catalyst feed rate to a reactor to control a reaction in response to available heat transfer and at least one additional process constraint. A reaction process constraint override control system for optimizing production rates, the system includes a reactor, a catalyst feeder that is structured to feed catalyst to the reactor, a heat transfer apparatus that is structured to transfer heat associated with the reactor, a heat transfer availability sensor that is structured to determine available heat transfer in the heat transfer apparatus, and at least one controller that is structured to automatically and continuously control the catalyst feeder in response to available heat transfer and at least one additional process constraint.
Reaction Process Control

FIELD OF THE INVENTION

[0001] The present invention relates to methods and systems for controlling catalyst feed rate to control reaction rates. As a preferred example, the present invention is directed to methods and systems for controlling continuous exothermic processes, such as for optimizing production rate during gas phase polymerization in a fluidized bed reactor.

BACKGROUND

[0002] Various methods for controlling reactions are known. Reaction control is often accomplished by controlling catalyst feed rate. In many chemical reactions, such as the polymerization of hydrocarbon monomers, a catalyst is used to promote the chemical reaction. It is known that most reactions are sensitive to the quantity of catalyst utilized so that the production rate of the reaction can be controlled by controlling the rate of catalyst feed to the reactor.

[0003] For example, U.S. Patent No. 4,162,894 to TRIBBLE discloses an apparatus and method for controlling catalyst feed. TRIBBLE discloses monitoring a reaction condition in a reactor and generating a signal to energize a motor to supply a measured amount of catalyst to the reactor. TRIBBLE discloses that catalyst may be fed on demand of the reactor coolant difference (ΔT) controller.

[0004] U.S. Patent No. 4,668,473 to AGARWAL discloses a control system for an ethylene polymerization reactor. AGARWAL discloses that the amount of catalyst supplied to a feed line is regulated as a function of raw material concentration of the feed line or flow of raw material in the feed line. AGARWAL discloses that the control apparatus optimizes the process by maximizing reactor temperature, catalyst flow rate, coolant flow rate, reactor exit temperature, and maximum polymerization in the reactor. AGARWAL further discloses that a catalyst supply means is controlled on the basis of total feed flow to the reactor, maximum reactor temperature, and desired product quality. AGARWAL discloses controlling the addition of catalyst at multiple locations.
[0005] U.S. Patent No. 4,818,372 to MAULEON et al. discloses an apparatus for the catalytic cracking of hydrocarbon feedstocks with reaction temperature control. MAULEON et al. discloses that a temperature sensor controls a valve which manipulates the rate of injection of regenerated catalyst.

[0006] U.S. Patent Nos. 4,093,537 and 4,211,636 to GROSS et al. disclose a fluid catalytic control section. These GROSS et al. patents disclose adjustment in catalyst flow rate in response to reactor temperature. These GROSS et al. patents also disclose control of the recycle ratio of regenerated catalyst to spent catalyst in response to regenerator temperature.

[0007] U.S. Patent No. 4,282,084 to GROSS et al. discloses a catalytic cracking process. GROSS et al. discloses that a computation is made for the control of catalyst inventory, e.g., by catalyst regeneration control, in the process so as to provide oxidative activity at the level required to produce the required CO₂/CO ratio. GROSS et al. discloses that data on feedstock and catalyst activity are fed to an executive optimizer which optimizes the process. GROSS et al. discloses that a temperature controller, using data of reactant and reactor temperature, controls catalyst circulation rate.

[0008] U.S. Patent Nos. 5,883,292 and 5,939,582 to DASSEL et al. disclose reaction control by regulating internal condensation inside a reactor. The DASSEL et al. patents disclose controlling conversion and first droplet temperature by controlling catalyst concentration in response to several variables including reaction temperature.

[0009] U.S. Patent No. 6,183,698 to VASSILIOU et al. discloses devices for controlling the reaction rate of a hydrocarbon to an acid or other intermediate oxidation product by pressure drop rate adjustments. VASSILIOU et al. discloses adjusting feed rates of several components, including catalyst, in response to several variables, such as the composition of the recycle and flow rate of recycle stream, to adjust pressure drop rate and to control the reaction rate. VASSILIOU et al. discloses monitoring reaction temperature and adjusting heat exchangers to achieve the desired reaction temperature.

[0010] U.S. Patent No. 4,619,901 to WEBB et al. discloses that a desired unreacted monomer concentration is maintained in a reaction effluent removed from a polymerization reactor by manipulating catalyst feed rate in response to a non-linear
control signal derived from a comparison of actual and desired unreacted monomer concentration.

[0011] U.S. Patent No. 4,620,049 to SCHMIDT et al. discloses controlling the molecular weight of polybutene. SCHMIDT et al. discloses adjusting variables such as the feed rate of catalyst into the reactor in response to the concentration of isobutylene in the reactor.

[0012] U.S. Patent No. 5,116,915 to MAMEDOV et al. discloses using a computer to calculate Mooney viscosity to adjust process parameters, such as catalyst feed rate. MAMEDOV et al. also discloses that the computer is used to maintain a temperature difference between sections of the reactor to obtain a polymer with desired molecular weight distribution.

[0013] U.S. Patent No. 4,251,503 to SWINDELLS et al. discloses an automatic continuous monitoring system for a chlorine dioxide generating process. SWINDELLS et al. discloses that catalyst feed may be varied in response to the ratio of chlorine dioxide and chlorine in an off-gas stream. Other background references include EP 0906782 and U.S. Patent Nos. 6,063,877 and 4,543,399.

[0014] In a gas-phase polymerization reactor, the reaction is maintained in a fluidized bed of product, e.g., granular polyethylene, and gaseous reactants. Catalyst is injected into the fluidized bed. Heat of reaction is transferred to the circulating gas stream. This gas stream is compressed and cooled in the external recycle line. Make-up feed streams are added to maintain the desired reactant concentrations.

[0015] In a fluidized bed reaction system, as distinguished from stirred or paddle-type reaction systems, uniform distribution of monomer and catalyst in the upwardly moving gas stream is essential to avoid hot spots and resulting polymer chunks. See, for example, U.S. Patent No. 3,790,036. A further requirement of a fluidized bed reactor system is that the velocity of gas flowing through the reactor be adequate to maintain the bed in a fluidized state. The gas velocity required to keep the bed in a fluidized suspension cannot be achieved under normal conditions by mere injection of liquid at the bottom of the bed.

[0016] Fig. 1 is a schematic of a known gas-phase reactor system. The polymerization reactions are highly exothermic in nature, generating heat of reaction as the monomers
combine to form large complex molecular chains (the polymer). The heat of reaction needs to be convected out of the reaction site, as otherwise the temperature rise locally will create hot spots and lead to agglomeration or melting of the polymer. Thus, control of temperature is critical to stable reactor operation. Furthermore, the primary limitation on reaction rate in a fluidized bed reactor is the rate at which heat can be removed from the polymerization zone.

[0017] Polymerization reactors employ a gaseous medium for dispersing the catalyst which serves the dual purpose of providing a reaction matrix, while also acting as a cooling medium to remove heat of reaction from the active sites of the growing polymer. A common means of heat removal employed in conventional fluidized bed reactor processes is by compression and cooling of the recycle gas stream at a point external to the reactor. Depending upon the level of cooling, the reactor system is operated in condensed mode operation or dry mode operation. In condensed mode operation, to increase heat removal, recycle gases may be cooled to a temperature below the dew point of the recycle gas such that a two phase fluid mixture is returned to the reactor. In dry mode operation, recycle gases are cooled to a temperature above the dew point of the recycle gas such that a single phase, i.e., gas phase, is returned to the reactor.

[0018] Referring again to Fig. 1, a known mode of operation involves a cascaded temperature control for the reactor bed to the feed temperature that manipulates the cooling water return (CWR) valve in a slave loop to adjust the water temperature entering the heat exchanger. Variations on this temperature control may involve a single-loop bed temperature controller that directly manipulates the CWR valve. Independent of this the operator adjusts the catalyst feed rate to maintain production at the maximum rates possible.

[0019] In commercial scale fluidized bed reaction systems for producing polymers such as polyethylene, the amount of fluid which must be circulated to remove the heat of polymerization is greater than the amount of fluid required for support of the fluidized bed and for adequate solids mixing in the fluidized bed. The fluid velocity in the reactor is limited to prevent excessive entrainment of solids. A constant bed temperature will result if the heat generated by the polymerization reaction (which is proportional to the
polymer production rate) is equal to the heat absorbed by the fluidizing stream as it passes through the bed, plus any heat removed or lost by other means. An example of a commercial gas phase reactor system is the Unipol™ system developed by Union Carbide.

[0020] Heat convected out of the reaction site is dispersed via heat exchangers that employ water as the cooling medium. The maximum production rate in the reactor depends on the rate of heat removal. The heat generation rate is given by:

$$\text{Heat Rate} = Q_r \text{Prate} = M_g C_p (T_r - T_f) = M_w C_p_w (T_w - T_w_o)$$

where $Q_r$ is the heat of reaction rate (approx. 1525 Btu/lb. (847 kcal/kg) for production of polyethylene); Prate is the polymer production rate; $M_g$ is the mass flow rate of reactor gas (reactants plus any inert components); $C_p$ is the average mass heat capacity of the reaction media; $T_r$ is the reaction temperature; $T_f$ is the recirculating gas feed temperature; $M_w$ is the mass flow rate of water coolant; $C_p_w$ is the average mass heat capacity of cooling water; $T_w$ is the water temperature at the inlet of the exchanger; and $T_w_o$ is the water temperature at the outlet of the exchanger.

[0021] The reactor systems incorporate regulatory control systems for temperature, pressure, feed level, and composition. Typically, the operator will attempt to maximize production rates for a reactor by manually adjusting the catalyst feeder (RPM) to the limit of cooling.

[0022] The objective of maximizing the reactor production rates to the cooling limit is a daily challenge for operators. The cooling water supply temperature varies cyclically, due to day/night temperature variations, as well as seasonal temperature variations. The vagaries of catalyst feeder (CCF) operation, with the possibility of having a pick-up block down at any time as well as the need to bring the CCF down for catalyst refills creates another challenge in maximizing production rates. Changes in heat capacity of the reaction gas (by using isopentane for example) also change the load requirement for cooling water in the heat exchanger.

[0023] Different catalysts are used in the polymerization process, including, Ziegler-Natta, chrome and metallocene catalysts. The reaction catalyst can be either homogeneous or heterogeneous, such as supported Ziegler-Natta or chrome catalyst.
These catalysts have varying response time to catalyst feed level changes since the kinetics of the reactions are vastly different. Thus, it is a complex optimization problem for operators to stay at maximum safe production rates all the time. The operator needs to anticipate the effect of increased catalyst feed rate and ensure the reactor does not run out of cooling within the constraints explained earlier. There is a demonstrated tendency on the part of operators to sit in a “comfort zone”, with excessive reserves of cooling capacity, just in case extra cooling is required.

[0024] As noted above, fluidized bed reactors may be operated in dry mode or condensed mode. In dry mode operation, the recycle gas stream is cooled to a level which is above the dew point such that only a gas phase is present in the recycle stream. In condensed mode operation, the recycle stream is intentionally cooled to a temperature below the dew point of the recycle stream to produce a two-phase gas-liquid mixture under conditions such that the liquid phase of the mixture will remain entrained in the gas phase of the mixture at least from the point of entry into the fluidized bed reactor until volatilized or until passage into the fluidized bed. A substantial increase in space time yield may result from condensed mode operation with little or no change in product properties or quality. The degree of condensation is achieved by maintaining the outlet temperature from the cycle heat exchange zone so as to effect the required degree of cooling below the dew point of the mixture. It will be apparent that if desired, it is possible to form a two-phase fluid stream within the reactor at the point of injection by separately injecting gas and liquid under conditions which will produce a two-phase stream.

[0025] In accordance with condensed mode operation, the space time yield of polymer production in a fluidized bed reactor is increased by cooling the recycle stream to below its dew point and returning the resultant two-phase fluid stream to the reactor to maintain the fluidized bed at a desired temperature above the dew point of the recycle stream. The cooling capacity of the recycle stream is increased both due to the greater temperature differential between the entering recycle stream and the reactor and by the vaporization of the condensed liquids entrained in the recycle stream.

[0026] In view of the above, control of reactions, such as fluidized bed reactions, is often a manual operation. Accordingly, the control of reactions is often cumbersome,
imprecise, and subjective. Although automated control of reactions is generally known, a need exists for improved automated control of reactions.

SUMMARY

[0027] In one embodiment, the present invention is directed to methods and systems for optimizing production rates in polymerization reactions conducted in fluidized bed reactors. The invention is also directed to methods and systems for identification of important variables which are useful in the automatic control of reactions.

[0028] The invention is further directed to methods and systems for control of exothermic or endothermic reactions by controlling catalyst feed rate in response to a number of process variables that may arise and change during the operation of the reactor. As is commonly understood in the art, these variables are commonly referred to as process constraints. It is desirable to monitor and control the process constraints to provide for the safe and optimal operation of the chemical processes in the reactor. For example, in the exothermic reactor, variables that need to be monitored and controlled include: reactor production rate, available heat transfer (measured as a function of cooler delta temperature), dew point of the reactor feed inlet and the %weight condensed material.

[0029] Attempts to monitor and control process constraints manifest themselves in an overall constraint override control strategy. In a constraint override control strategy as described herein, there is only one primary variable whose precise control is the foremost objective. Every constraint override control strategy must have one or more constraint override variables, also referred to as secondary variables. Under normal operation and for most of the time, the primary controlled variable is the one variable being controlled. The secondary variables are typically not at their constraint limits and hence are not contributing to the control action.

[0030] In one embodiment, such a strategy is referred to as a constraint override control. The constraint override control uses the production rate as the primary controlled variable in the control scheme. This embodiment provides for the control of the production rate to a fixed value based on considerations such as market/sales demand.
The cooling water delta temperature, dew point or % weight condensed and purge bin level are the secondary variables or the constraint override variables. In this embodiment, monitoring and controlling the primary controlled variable in continuous coordination of the choice, control, and location of the secondary controlled variables provides for the effective implementation of the control scheme.

[0031] In another embodiment, the constraint override control provides for a dual variable or a two-tier system coupled with the constraint override. The two tiered system looks both at the cooling water valve position and the cooling water delta temperature limit. As catalyst type is changed, as ambient conditions change, or as unmeasured disturbances affect the reactor performance (the latter is frequent in a typical gas phase polymerization process), the two tiered system helps to improve the stability of the control scheme. Improved control stability aids in the increase of production rates.

[0032] In another embodiment, the invention is directed to a reaction process constraint override control method for optimizing production rates, the method comprising: automatically and continuously controlling catalyst feed rate to a reactor to control a reaction in response to available heat transfer and at least one additional process constraint. This embodiment is advantageous because exothermic reactors need external coolers to remove the heat generated from the reaction. Such reactor system equipment has capacity limits and, as is commonly understood in the art, those limits present many challenges. If the reaction heat is too high, the cooler may be delivering the maximum cooling possible based on its design capacity. It cannot cool the reactor past its maximum cooling limit. At the maximum cooling limit, the available heat transfer is zero. This embodiment enables the reactor to be pushed closer to an available heat transfer point of zero corresponding to maximum production rates.

[0033] In another aspect, the present invention is directed to a reaction process constraint override control system for optimizing production rates, the system comprising: reactor; catalyst feeder that is structured to feed catalyst to the reactor; heat transfer apparatus that is structured to transfer heat associated with the reactor; heat transfer availability sensor that is structured to determine available heat transfer in the heat transfer apparatus; and at least one controller that is structured to automatically and continuously control the
catalyst feeder in response to available heat transfer and at least one additional process constraint.

[0034] In still another aspect, the present invention is directed to a reaction process constraint override control method for optimizing production rates, the method comprising: automatically and continuously controlling catalyst feed to a reactor to control a reaction in response to available heat transfer; and triggering an alarm in response to variation in available heat transfer above a predetermined amount.

[0035] In yet another aspect, the present invention is directed to a reaction process constraint override control system for optimizing production rates, the system comprising: reactor; catalyst feeder that is structured to feed catalyst to the reactor; heat transfer apparatus that is structured to transfer heat associated with the reactor; heat transfer availability sensor that is structured to determine available heat transfer in the heat transfer apparatus; and at least one controller that is structured to automatically and continuously control the catalyst feeder in response to available heat transfer; wherein the controller triggers an alarm in response to variations in available heat transfer above a predetermined amount.

[0036] In a further aspect, the present invention is directed to a reaction process constraint override control method for optimizing production rates, the method comprising: automatically and continuously controlling catalyst feed to a reactor to control a reaction in response to a position of a heat transfer valve.

[0037] In another aspect, the present invention is directed to a reaction process constraint override control system for optimizing production rates, the system comprising: reactor; catalyst feeder that is structured to feed catalyst to the reactor; a heat exchanger associated with the reactor; a heat transfer source; a conduit which is structured to circulate heat transfer medium between the heat exchanger and the heat transfer source; a recirculation conduit structured to allow at least a portion of the heat transfer medium to bypass the heat transfer source such that the at least a portion of the heat transfer medium is re-introduced into the heat exchanger without passing through the heat transfer source; a heat transfer medium valve structured to control the amount of heat transfer medium which bypasses the heat transfer source and the amount of heat transfer medium
which passes through the heat transfer source; a heat transfer medium valve sensor structured to measure a position of the heat transfer medium valve; heat transfer availability sensor that is structured to determine available heat transfer in the heat transfer apparatus; at least one controller that is structured to control the catalyst feeder; wherein the heat transfer medium valve sensor is connected to the at least one controller; and wherein the at least one controller is structured to control the catalyst feeder in response to the position of the heat transfer medium valve.

[0038] In still another aspect, the present invention is directed to a reaction process constraint override control method for optimizing production rates, the method comprising: automatically and continuously controlling catalyst feed rate to a reactor to control a reaction in response to: available heat transfer, current production rate of the reactor, amount of reaction product in a collector of the reactor, and weight percent condensed in a recycle stream of the reactor; wherein the reaction comprises polymerization of ethylene to form polyethylene.

[0039] In one aspect, the additional process constraint comprises current production rate of the reactor.

[0040] In another aspect, the current production rate is compared with a desired production rate to control the catalyst feed rate.

[0041] In still another aspect, the additional process constraint comprises amount of reaction product in a collector of the reactor.

[0042] In still another aspect, the amount of reaction product in the collector is compared with a desired amount of reaction product in the collector to control the catalyst feed rate.

[0043] In yet another aspect, the additional process constraint comprises weight percent condensed in a recycle stream of the reactor.

[0044] In another aspect, the weight percent condensed is compared with a desired weight percent condensed.

[0045] In still another aspect, the additional process constraint comprises dew point of a recycle stream of the reactor.
[0046] In a further aspect, the dew point is compared with a desired dew point to control the catalyst feed rate.

[0047] In another aspect, the additional process constraint comprises carryover of solids into a recycle stream.

[0048] In still another aspect, the carryover of solids is compared with a desired carryover of solids to control the catalyst feed rate.

[0049] In a further aspect, the additional process constraint comprises granular resin transfer system rate.

[0050] In still another aspect, the granular resin transfer system rate is compared with a desired granular resin transfer system rate to control the catalyst feed rate.

[0051] In another aspect, the additional process constraint comprises pelletizer rate.

[0052] In yet another aspect, the pelletizer rate is compared with a desired pelletizer rate to control the catalyst feed rate.

[0053] In still another aspect, the additional process constraint comprises feed flow.

[0054] In a further aspect, the feed flow is compared with a desired feed flow to control the catalyst feed rate.

[0055] In yet another aspect, automatically and continuously controlling catalyst feed to the reactor in response to available heat transfer comprises monitoring a delta temperature of a mix point of heat transfer medium.

[0056] In still another aspect, automatically and continuously controlling catalyst feed to the reactor in response to available heat transfer comprises monitoring a delta temperature of a mix point of heat transfer medium and monitoring a position of a heat transfer medium valve of a heat transfer apparatus.

[0057] In a further aspect, automatically and continuously controlling catalyst feed rate to the reactor in response to available heat transfer comprises monitoring a position of a heat transfer medium valve of a heat transfer apparatus.

[0058] In still another aspect, the reactor comprises a heat transfer apparatus comprising a heat exchanger associated with the reactor and a conduit which is structured to allow circulation of heat transfer medium between the heat exchanger and a heat transfer source.
[0059] In another aspect, the heat transfer apparatus comprises a recirculation conduit structured to allow at least a portion of the heat transfer medium to bypass the heat transfer source such that the at least a portion of the heat transfer medium is re-introduced into the heat exchanger without passing through the heat transfer source.

[0060] In another aspect, the recirculation conduit is structured to, at a mix point, mix the at least a portion of the heat transfer medium with heat transfer medium which passes through the heat transfer source, prior to introduction into the heat exchanger.

[0061] In a further aspect, a first temperature sensor measures a first temperature of transfer medium at a point between the heat transfer source and the mix point; and a second temperature sensor measures a second temperature of the heat transfer medium between the mix point and the heat exchanger.

[0062] In another aspect, a controller subtracts the first temperature from the second temperature to calculate a delta temperature.

[0063] In yet another aspect, the controller triggers an alarm in response to variations in delta temperature above a predetermined amount.

[0064] In a further aspect, the heat transfer apparatus comprises a heat transfer medium valve structured to control the amount of heat transfer medium which bypasses the heat transfer source and the amount of heat transfer medium which passes through the heat transfer source.

[0065] In still another aspect, a heat transfer medium valve sensor measures a position of the heat transfer medium valve.

[0066] In yet another aspect, the position of the heat transfer medium valve is compared with a maximum allowable heat transfer medium valve position to control the catalyst feed rate, and the heat transfer medium valve position is maintained in a range in which a change in valve position effects a change in heat transfer.

[0067] In still another aspect, the method further comprises measuring a pressure differential made by a primary pump of a heat transfer apparatus.

[0068] In another aspect, the method further comprises controlling a secondary pump in response to the pressure differential made by the primary pump.

[0069] In still another aspect, the reactor comprises a fluidized bed.
[0070] In yet another aspect, the reaction comprises polymerization.

[0071] In still another aspect, the reaction comprises polymerization of ethylene to form polyethylene.

[0072] In another aspect, the system further comprises a heat transfer medium valve structured to control the amount of heat transfer medium which bypasses the heat transfer source and the amount of heat transfer medium which passes through the heat transfer source, the system further comprises a heat transfer medium valve sensor structured to measure a position of the heat transfer medium valve, wherein the heat transfer medium valve sensor is connected to the at least one controller, and wherein the at least one controller is structured to compare a signal from the heat transfer medium valve sensor and the a delta temperature to control the catalyst feeder.

[0073] In another aspect, the pressure differential sensor is connected to a secondary pump controller which is structured to control the secondary pump in response to the pressure differential made by the primary pump, and the secondary pump controller is the same as or distinct from the at least one controller.

[0074] In another aspect, the system further comprises a recycle loop.

[0075] In still another aspect, the system further comprises a dew point sensor structured to measure a dew point of gas in the recycle loop.

[0076] In yet another aspect, the dew point sensor is connected to the at least one controller, and wherein the at least one controller is structured to control the catalyst feeder in response to the dew point of gas in the recycle loop.

[0077] In a further aspect, the system further comprises an operator interface which is structured to receive a dew point approach, and the at least one controller compares the dew point with the dew point approach to control the catalyst feeder.

[0078] In yet another aspect, the system further comprises a weight percent condensed sensor structured to measure a weight percent condensed of fluid in the recycle loop.

[0079] In still another aspect, the weight percent condensed sensor is connected to the at least one controller, and wherein the at least one controller is structured to control the catalyst feeder in response to the weight percent condensed in the recycle loop.
[0080] In a further aspect, the system further comprises an operator interface which is structured to receive a maximum allowable weight percent condensed, and the at least one controller compares the weight percent condensed with the maximum allowable weight percent condensed to control the catalyst feeder.

[0081] In still another aspect, the reactor comprises a collector structured to collect reaction product.

[0082] In yet another aspect, the system further comprises a reaction product sensor structured to measure an amount of reaction product in the collector.

[0083] In another aspect, the reaction product sensor is connected to the at least one controller, and the at least one controller is structured to control the catalyst feeder in response to the amount of reaction product in the collector.

[0084] In still another aspect, the system further comprises an operator interface which is structured to receive a desired amount of reaction product in the collector, and wherein the at least one controller compares the amount of reaction product in the controller with the desired amount of reaction product in the controller to control the catalyst feeder.

[0085] In yet another aspect, the system further comprises a production rate sensor structured to measure a production rate of the reactor.

[0086] In still another aspect, the production rate sensor is connected to the at least one controller, and wherein the at least one controller is structured to control the catalyst feeder in response to the production rate.

[0087] In a further aspect, the system further comprises an operator interface which is structured to receive a desired production rate, and wherein the at least one controller compares the production rate with the desired production rate to control the catalyst feeder.

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**BRIEF DESCRIPTION OF THE DRAWINGS**

[0088] The present invention is further described in the detailed description which follows, in reference to the noted plurality of non-limiting drawings, wherein:

Fig. 1 is a schematic of a known gas-phase reactor system;
Fig. 2 is a schematic of a gas-phase reactor system in accordance with the present invention;

Fig. 3 is a perspective view of a catalyst feeder;

Fig. 4 is an exploded and partial cross-sectional view of the catalyst feeder;

Fig. 5 is a logic diagram wherein production is controlled in response to target production rate, cooling water delta temperature, cooling water valve position, and dew point;

Fig. 6 is a logic diagram wherein production is controlled in response to target production rate, target cooling water valve position, and a target dew point;

Fig. 7 is a logic diagram wherein production is controlled in response to desired production rate, desired purge bin level, maximum weight percent condensed, and maximum cooling water valve position;

Fig. 8 is a logic diagram of RPM initializer logic;

Fig. 9 is a logic diagram of RPM splitter logic;

Fig. 10 is a logic diagram of setpoint processor logic;

Fig. 11 is a logic diagram of RPM anti-windup logic;

Fig. 12 illustrates a production rate controller operator interface;

Fig. 13 illustrates a controller screen which is provided for engineer interface;

Fig. 14 shows results from Example 1 in which production rate is maximized to cooling constraint;

Fig. 15 shows results from Example 2 in which the production rate controller optimizes rates;

Fig. 16 shows results from Example 3 in which production rate varies as a function of catalyst in the reactor and cooling water delta temperature mirrors production rate and catalyst in the reactor; and

Fig. 17 shows the results from Comparative Example 1 which shows the potential importance of monitoring dew point.

DETAILED DESCRIPTION
The particulars shown herein are by way of example and for purposes of illustrative discussion of the various embodiments of the present invention only and are presented in the cause of providing what is believed to be the most useful and readily understood description of the principles and conceptual aspects of the invention. In this regard, no attempt is made to show details of the invention in more detail than is necessary for a fundamental understanding of the invention, the description taken with the drawings making apparent to those skilled in the art how the several forms of the invention may be embodied in practice.

All percent measurements in this description, unless otherwise stated, are measured by weight based upon 100% of a given sample weight. Thus, for example, 30% represents 30 weight parts out of every 100 weight parts of the sample.

Unless otherwise stated, a reference to a compound or component, includes the compound or component by itself, as well as in combination with other compounds or components, such as mixtures of compounds.

As an overview, the present invention is generally directed to methods and systems for controlling catalyst feed rate in response to one or more constraint variables. Although the catalyst feed rate may controlled in response to one variable, better results are often obtained when controlling in response to multiple variables. The constraint variable may involve reaction conditions such as desired production rate, heat transfer availability (e.g., cooling water delta temperature, cooling water valve position), dew point limit, weight percent condensed, weight percent condensables, removal of product (e.g., purge bin level), level of fluidization, level of entrainment, pelletizer, granular resin transfer capability, and/or feed supply (e.g., feed supply pressure).

The present invention may be used to control any reaction the rate of which is controlled by catalyst feed rate. Accordingly, almost any type of reactor may be used in the practice of the present invention. The present invention can be practiced in connection with endothermic or exothermic reaction processes. Examples of reactions of the present invention include catalyst cracking and gas phase polymerization in a fluidized bed.
[0094] As a preferred example, however, the present invention is directed to controlling exothermic polymerization reactions in a fluidized bed. In very general terms, a fluidized bed process for producing resins, particularly polymers produced from monomers, is practiced by passing a gaseous stream containing one or more monomers continuously through a fluidized bed reactor under reactive conditions and in the presence of catalyst. The gaseous stream containing unreacted gaseous monomer is withdrawn from the reactor continuously, compressed, cooled and recycled into the reactor. Product is withdrawn from the reactor. Make-up monomer is added, e.g., to the recycle stream.

[0095] The polymer-forming reaction is exothermic, making it necessary to maintain in some fashion the temperature of the gas stream inside the reactor at a temperature not only below the resin and catalyst degradation temperatures, but also below the fusion or sticking temperature of resin particles produced during the polymerization reaction. This temperature control is necessary to prevent plugging of the reactor due to rapid growth of polymer chunks which cannot be removed in a continuous fashion as product. It will be understood, therefore, that the amount of polymer that can be produced in a fluidized bed reactor of a given size in a specified time period is often directly related to the amount of heat which can be withdrawn from the fluidized bed.

[0096] As discussed in more detail below, a number of constraints limit production rates. The present invention is based on the realization that these constraints include, but are not limited to, desired production rate, available heat transfer, dew point limit, weight percent condensed, weight percent condensables, removal of product (e.g., purge bin level), level of fluidization, level of entrainment, pelletizer, granular resin transfer capability, and/or feed supply (e.g., feed supply pressure). As used herein, the term "available heat transfer" refers to the amount of cooling capacity available at a given time, as determined by, for example, the cooling water delta temperature (greater delta temperature indicating greater available heat transfer) and/or the cooling water valve position.

[0097] The present invention is directed to constraint override control methods and systems for controlling the production rate by changing catalyst feed rate in response to an input based on one or more of these constraints. In one aspect, the present invention
is directed to constraint override control methods and systems for automatic reaction control by controlling catalyst feed rate in response to desired production rate, available heat transfer, dew point limit, weight percent condensed, weight percent condensables, removal of product (e.g., purge bin level), level of fluidization, level of entrainment, pelletizer, granular resin transfer capability, and/or feed supply (e.g., feed supply pressure).

[0098] The present invention is also directed to methods and systems for triggering an alarm if the available heat transfer varies suddenly.

[0099] Furthermore, the present invention is directed to methods and systems for controlling a secondary pump, where appropriate, to ensure the automatic production rate control is optimally engaged, without a run-away reaction.

[0100] To better understand the advantages of the control system and method of the present invention, non-limiting examples directed to gas phase polymerization follow. To better understand the advantages of the control system and method of the present invention as related to gas phase polymerization, it is important to generally understand the fluidized bed gas phase polymerization process.

[0101] Although not limited to any particular kind of reactor, the fluidized bed reactors of the present invention may be operated in dry mode or condensed mode, as discussed above. In general, in dry mode operation, the recycle gas stream is cooled to a level which is above the dew point such that only a gas phase is present in the recycle stream. In condensed mode operation, the recycle stream is intentionally cooled to a temperature below the dew point of the recycle stream to produce a two-phase gas-liquid mixture under conditions such that the liquid phase of said mixture will remain entrained in the gas phase of said mixture at least from the point of entry into the fluidized bed reactor until volatilized or until passage into the fluidized bed.

[0102] The entry point for the recycle stream should be below the fluidized bed (polymerization zone) to ensure uniformity of the upwardly flowing gas stream and to maintain the bed in a suspended condition. The recycle stream is introduced into the reactor at a point in the lower region of the reactor and most preferably at the very bottom
of the reactor to ensure uniformity of the fluid stream passing upwardly through the fluidized bed.

[0103] A baffle or similar means for preventing regions of low gas velocity in the vicinity of the recycle stream entry point may be provided to keep solids and liquids entrained in the upwardly flowing recycle stream.

[0104] The recycle stream can be divided into two or more separate streams, one or more of which can be introduced directly into the polymerization zone provided that sufficient gas velocity below and through the bed is provided to keep the bed suspended. In all cases the composition of the recycle stream is kept essentially uniform and flowing in a manner such that there are no dead spaces in the bed where unremovable solids can form.

[0105] It may be desirable to inject make-up monomer in the form of a liquid and/or gas. The injection of liquid or gaseous make-up monomer at the point of entry of the recycle stream or elsewhere in the reactor or in the recycle stream is contemplated by this invention.

[0106] Fig. 2 involves a non-limiting example of the present invention. Initially, it is noted that the process of the present invention is not limited to any particular apparatus or piping details. For example, each of the equipment items shown in the examples of the present invention may be directed to multiple units which function in series or in parallel.

[0107] Fig. 2 shows a fluidized bed reactor which is particularly suited for the production of polyolefin resin in accordance with the present invention. In Fig. 2, TI = temperature indicator, TC = temperature controller, OP = output, SP = setpoint, SC = speed controller, and PV = process variable. The reactor 10 includes a reaction zone 12 and a velocity reduction zone 14.

[0108] In general, the height to diameter ratio of the reaction zone preferably varies in the range of about 2.7:1 to 5:1. The range, of course, can vary to larger or smaller ratios and depends upon the desired production capacity. The cross-sectional area of the velocity reduction zone 14 is typically within the range of about 2.6 to 2.8 times larger than the cross-sectional area of the reaction zone 12.
[0109] The reaction zone 12 includes a bed of growing polymer particles, formed polymer particles and a minor amount of catalyst particles fluidized by the continuous flow of polymerizable and modifying gaseous components in the form of make-up feed and recycle fluid through the reaction zone. To maintain a viable fluidized bed, the superficial gas velocity through the bed must exceed the minimum flow required for fluidization, and preferably is at least about 0.2 ft/s (0.06 m/s) above the minimum flow required for fluidization. Ordinarily, the superficial gas velocity does not exceed 5.0 ft/s (1.5 m/s) and usually 2.5 ft/s (0.76 m/s) is sufficient.

[0110] It is important that the bed always contain particles to prevent the formation of localized "hot spots" and to entrap and distribute the particulate catalyst throughout the reaction zone. On start up, the reactor is usually charged with a base of particulate polymer particles before gas flow is initiated. Such particles may be identical in nature to the polymer to be formed or different therefrom. When different, they are withdrawn with the desired formed polymer particles as the first product. Eventually, a fluidized bed of desired polymer particles supplants the start-up bed.

[0111] The partially or totally activated precursor composition and/or catalyst used in the fluidized bed is preferably stored for service in a reservoir 16, such as one or more catalyst feeders 16 which are controlled by speed controllers SC, under a blanket of a gas which is inert to the stored material, such as nitrogen or argon. As discussed in more detail below, the reservoir preferably comprises two catalyst feeders.

[0112] Fluidization is achieved by a high rate of fluid recycle to and through the bed, typically in the order of about 20 to 70, more typically about 30 to 60, and most typically about 40, times the rate of feed of make-up fluid. The fluidized bed has the general appearance of a dense mass of individually moving particles as created by the percolation of gas through the bed. The pressure drop through the bed is equal to or slightly greater than the weight of the bed divided by the cross-sectional area. The pressure drop is thus dependent on the geometry of the reactor.

[0113] Make-up fluid is fed to the bed at point 18. The composition of the make-up stream is determined in response to a gas analyzer 21. The gas analyzer 21 determines the composition of the recycle stream, and the composition of the make-up stream is
adjusted accordingly to maintain an essentially steady state gaseous composition within the reaction zone. Gas analyzers are commercially available from a wide variety of sources. Generally, the gas analyzer 21 can be positioned so as to receive gas from a point between the velocity reduction zone 14 and heat exchanger 24.

[0114] To ensure complete fluidization, the recycle stream and, where desired, part of the make-up stream are returned through recycle line 22 to the reactor at point 26 below the bed. There is preferably a gas distributor plate 28 above the point of return to aid in fluidizing the bed. In passing through the bed, the recycle stream absorbs the heat of reaction generated by the polymerization reaction.

[0115] The portion of the fluidizing stream which does not react in the bed constitutes the recycle stream which is removed from the polymerization zone, preferably by passing it into velocity reduction zone 14 above the bed where entrained particles are given an opportunity to drop back into the bed.

[0116] The recycle stream is then compressed in a compressor 30 and then passed through a heat exchange zone wherein the heat of reaction is removed before the recycle stream is returned to the bed. The heat exchange zone is typically a heat exchanger 24, such as the horizontal or vertical type. The recycle stream is then returned to the reactor at its base 26 and to the fluidized bed through gas distributor plate 28. A gas deflector is preferably installed at the inlet to the reactor to prevent contained polymer particles from settling out and agglomerating into a solid mass.

[0117] The temperature of the bed is controlled at an essentially constant temperature under steady state conditions by constantly removing the heat of reaction. No noticeable temperature gradient appears to exist within the upper portion of the bed. A temperature gradient will exist in the bottom of the bed in a layer of typically up to about 48 inches (1.2 m), more typically up to about 24 inches (61 cm), and most typically up to about 12 inches (30 cm), between the temperature of the inlet fluid and the temperature of the remainder of the bed. The temperature gradient is typically up to about 50°C, more typically up to about 30°C, with a typical range of about 10 to 20°C.

[0118] Good gas distribution plays an important role in the operation of the reactor. The fluidized bed contains growing and formed particulate polymer particles, as well as
catalyst particles. As the polymer particles are hot and possibly active, they must be prevented from settling, for if a quiescent mass is allowed to exist, any active catalyst contained therein may continue to react and cause fusion. Diffusing recycle fluid through the bed at a rate sufficient to maintain fluidization throughout the bed is, therefore, important.

[0119] Gas distribution plate 28 is a preferred means for achieving good gas distribution and may be a screen, slotted plate, perforated plate, a plate of the bubble-cap type and the like. The elements of the plate may all be stationary, or the plate may be of the mobile type disclosed in U.S. Patent No. 3,298,792 to Di DRUSCO, the disclosure of which is herein incorporated by reference in its entirety.

[0120] Whatever its design, the gas distribution plate is intended to diffuse the recycle fluid through the particles at the base of the bed to keep the bed in a fluidized condition, and also serve to support a quiescent bed of resin particles when the reactor is not in operation.

[0121] Any fluid inert to the catalyst and reactants can also be present in the recycle stream. Examples include saturated hydrocarbons, such as alkanes, such as isopentane, normal pentane, isohexane, and normal hexane.

[0122] One or more activator compounds, such as one or more co-catalysts, if utilized, is preferably added to the reaction system downstream from heat exchanger 24. Thus, the activator may be fed into the recycle system from dispenser 38 through line 40. Typical co-catalysts are organic aluminum alkyls such as triethyl aluminum (TEAL), trimethyl aluminum (TMA), tributyl aluminum, methylalumoxane (MAO) and the like. Such activators may be used in Ziegler-Natta and metallocene catalyst systems.

[0123] The fluidized bed reactor is preferably operated at a temperature below the sintering temperature of the polymer particles to ensure that sintering will not occur. The sintering temperature is a function of resin density. In general, polyethylene low-density resins, for example, have a low sintering temperature and polyethylene high-density resins, for example, have a higher sintering temperature. For example, temperatures of from about 75 to 95°C are used to prepare ethylene copolymers having a density of from about 0.91 to 0.95 g/cm³, while temperatures of from about 100 to 115°C are used to
prepare ethylene copolymers or homopolymers having a density of from about 0.95 to 0.97 g/cm$^3$.

[0124] The fluidized bed reactor may be operated at pressures of up to about 1000 psi (6.9 MPa). For polyolefin resin production the fluidized bed is preferably operated at a pressure of about 100 to 400 psi (0.69 to 2.8 MPa), more preferably about 200 to 400 psi (1.4 to 2.8 MPa), and most preferably about 250 to 350 psi (1.7 to 2.4 MPa), with operation at the higher pressures in such ranges favoring heat transfer since an increase in pressure increases the unit volume heat capacity of the gas. As noted in the Background of the Invention, the operating pressure may be raised as needed to increase the dew point of the recycle stream.

[0125] The partially or totally activated precursor composition and/or catalyst (hereinafter collectively referred to as catalyst) is injected into conduit 42 and may be injected into the bed at a rate equal to its consumption at a point 44 which is above distributor plate 28. Preferably, the catalyst is injected at a point in the bed where good mixing of polymer particles occurs. Injecting the catalyst at a point above the distribution plate is a preferred feature for satisfactory operation of a fluidized bed polymerization reactor. Since catalysts are highly active, injection of the catalyst into the area below the distributor plate may cause polymerization to begin there and eventually cause plugging of the distributor plate. Injection into the fluidized bed, instead, aids in distributing the catalyst throughout the bed and tends to preclude the formation of localized spots of high catalyst concentration which may result in the formation of "hot spots." Injection of the catalyst into the reactor above the bed may result in excessive catalyst carryover into the recycle line where polymerization may begin and plugging of the line and heat exchanger may eventually occur. Accordingly, the catalyst is preferably fed into the reactor at a point preferably 10 to 50%, more preferably about 15 to 40%, and most preferably about 20 to 35%, of the reactor diameter away from the reactor wall and at a height of preferably about 5 to 30%, more preferably about 5 to 25%, and most preferably about 5 to 15%, of the height of the bed.

[0126] The catalyst can be injected into the reactor by various techniques. It is preferred, however, to continuously feed the catalyst into the reactor utilizing a catalyst feeder
similar to that disclosed in U.S. Patent No. 3,779,712 to CALVERT et al., the disclosure of which is herein incorporated by reference in its entirety.

[0127] Fig. 3 shows a catalyst feeder 16 which may be used in accordance with the present invention. The catalyst feeder 16 is similar to the catalyst feeder disclosed in the CALVERT reference. Unlike the catalyst feeder of CALVERT et al., the catalyst feeder 16 comprises two entrainment chambers or blocks 46.

[0128] Fig. 4 is an exploded and partial cross-sectional view of the catalyst feeder 16. Catalyst is held in a lower casing 48 of the catalyst feeder 16. The lower casing 48 includes two ports 50. Beneath the lower casing 48 is a metering disc 52 having cavities 54. The metering disc 52 may be rotated by rotating rotatable shaft member 56. Beneath the metering disc 52 is a lower flange 58 having two inlets 60. Each of the inlets 60 is connected to one of the blocks 46.

[0129] In operation, the catalyst feeder 16 functions in substantially the same manner as the catalyst feeder of CALVERT et al. In general, as the metering disc 52 rotates, each cavity 54 will be filled with catalyst through ports 50. Subsequently, when the cavity 54 filled with catalyst aligns with an inlet 60, the catalyst is introduced into one of the blocks 46. Thus, the catalyst feed rate can be controlled by controlling the rotational speed of the metering disc 52.

[0130] A gas which is inert to the catalyst, such as nitrogen or argon, is preferably used to carry the catalyst into the bed.

[0131] The rate of polymer production in the bed depends on the rate of catalyst injection and the concentration of monomer(s) in the recycle stream. A number of constraints, however, can dictate the maximum production rates achieved. Examples of constraints include, but are not limited to, cooling water temperature (available heat sink), dew-point of the cycle gas, the maximum allowable inert coolant used in the reactor for good fluidization, and transfer rate out of the reactor. The present invention is equally applicable to each of these constraints. Thus, the present invention is directed to controlling the production rate with respect to one or more of these constraints. Examples of these constraints include desired production rate, heat transfer availability which may vary as a function of ambient conditions (e.g., cooling water delta
temperature, cooling water valve position, flow rate through cooling tower), dew point limit, weight percent condensed, weight percent condensables, removal of product (e.g., purge bin level), level of fluidization, level of entrainment, pelletizer, granular resin transfer capability, and feed supply (e.g., feed supply pressure).

[0132] In gas phase polymerization reactions, one of the constraints often reached is the available cooling. An important measure of the available cooling is the temperature difference between the cooling water entering the heat exchanger and the cooling water supply temperature. The temperature of the water entering the exchanger is referred to as the 'mix point', after the make up cooling water supply is mixed with the re-circulating water of the cycle water loop. For the purpose of tracking the available cooling, the following definition for Cooling water delta temperature (CWD): is used:

\[
\text{CWD} = \text{Mix point' water temperature} - \text{Water supply temperature}.
\]

[0133] The theoretical limit to cooling is achieved when the CWD approaches zero, e.g., when the cooling water valve 70 is fully open and greater than 95% of the water passes through the cooling tower in the cooling system of Fig. 2. The dynamic relationship between the available cooling as measured by CWD and the catalyst feed changes is an important aspect for the design of the production rate controller of the present invention.

[0134] Fig. 5 is an overview of a non-limiting example showing production rate controller logic for controlling gas phase polymerization reactions, wherein production is constrained by a target production rate, cooling water delta temperature, cooling water valve 70 position, and dew point. The present invention, however, generally extends to control system configurations which accomplish the purpose of the present invention.

[0135] In Fig. 5, a rate controller 100 preferably comprises a computer program. Accordingly, the present invention preferably includes a digital computer to calculate the required control signals based on the measured process parameters as well as the setpoints supplied to the computer. The present invention may be implemented as part of a Distributed Control System (DCS). Commercially available examples of control systems in accordance with the present invention include Foxboro IA (Intelligent
Automation) and Honeywell TDC-3000. Other types of control systems could also be used in the present invention.

[0136] The rate controller 100 is programmed with a target production rate 102, a target cooling water valve position (CW Valve Position) 104, and a target dew point (DewPt) 106. Thus, the operator specifies the targets for the approach limits for each of the production rate, cooling water valve 70 position and dew point constraints. Based on historical data, the rate controller 100 automatically translates the target cooling water valve position 104 into the appropriate cooling water delta temperature (CWDT) approach limit.

[0137] The rate controller 100 compares these target values with the measured production rate 110, the measured cooling water valve position (CW Valve Position) 112 which is compared with the measured cooling water delta temperature (CWDT) 114, and the measured dew point (DewPt) 116. The comparison of the cooling water valve 70 position with CWDT involves cascade logic which is discussed in more detail below, with reference to Fig. 7.

[0138] The rate controller 100 interacts with an operator interface 120 via signals 122. The operator interface may comprise, e.g., a monitor and a keyboard. The operator uses the operator interface 120 to enter the desired setpoints.

[0139] The controller 100 also interacts with a catalyst feeder balancing controller 130 via signals 128. Based on plant dynamic response data, the control logic of the catalyst feeder balancing controller 130 computes the optimized target for the total catalyst RPM which is then balanced and split between the running feeder blocks supplying the catalyst to the reactor. For example, historical data of the relationship between the CWDT and catalyst feed rate are analyzed to extract the dynamic response. Commercially available dynamic identification software such as "PITOPS™" software, available from Artcon Incorporated, Cincinnati, OH, can be used to extract the transfer function dynamics for all the constraints. Analysis has shown that there is sufficient information in the dynamics of the changes in cooling water delta temperature to define the process dynamics required for the controller.
The catalyst feeder balancing controller 130 interacts with a block and feeder startup and initialization control logic 140 via signals 132. The block and feeder startup and initialization control logic 140 are intended to smooth the transition from manual to automatic control.

The block and feeder startup and initialization control logic 140 sends an RPM signal 142 to RPM PID (proportional-integral-derivative) controllers 150. The RPM PID controllers 150 compare the current feeder RPM 152 to the RPM signal 142 from the block and feeder startup and initialization control logic 140 and computes an RPM signal 154 which is sent to the motors of the catalyst feeders to control the catalyst feed rate.

Using the controller logic of Fig. 5, if the production rate is to be maximized, the operator will provide a high set point for the target production rate 104. The controller 100 then responds by ramping the catalyst feeder RPM—until one of the other set point limits is reached and will maintain the RPM at that value. In other words, if production rate is to be maximized, then a high, unachievable setpoint is specified. In this case, the catalyst feeders will be ramped up to achieve the production setpoint. As the rates are increased, another constraint may be approached. The output from the constraining PID will be lower than the production rate PID; this in turn will reduce the catalyst feed rate into the reactor. For instance, the limiting constraint is often available cooling. The dynamics of available cooling is continuously monitored using cooling water valve 70 position and CWDT.

Fig. 5 also shows loss of cooling potential alarm 160. Preferably, the high priority alarm 160 alerts the operator if the CWDT rises quickly for any reason. A sudden rise in CWDT can be because of an unknown poison in the feeds, bad catalyst, or other reasons that cause the reaction to subside. In such cases, it may be undesirable to increase catalyst flow into the reactor. The alarm alerts the operator of imminent problems warranting checking of reactor conditions followed up with appropriate action which may include de-activating the controller.

Fig. 5 further shows a CWDT deviation high alarm 170. This high priority alarm 170 is preferably provided to alert the operators of a potential runaway. If the reaction
fires up suddenly, manifested by quick rise in the rate of loss of CWDT, or quick opening of the CW valve 70, this alarm is generated.

[0145] As shown in Fig. 6, the reactor cooling system may include a primary cooling water pump 80 and a secondary cooling water pump 82. Fig. 5 also shows CW pump control logic (auto start of standby pump) 180.

[0146] In the event of failure of the primary cooling water pump 80, control logic to automatically start the secondary (standby) pump 82 is recommended. The logic may be triggered based on several inputs-delta pressure across the running pump 80, motor on/off status, pump discharge pressure and other potential inputs. If any of the inputs indicate pump failure, the standby pump 82 will be started automatically. Thus, the second pump 82 is started if the reactor is running and the running pump 80 shows either low delta pressure across the pump, low motor kW or amps, ON status changes to OFF, feedback indication is motor "not running", shutdown request detected from Distributed Control System alarm. Alternatively, the secondary pump 82 may normally be run at a low speed, with the speed being increased in the event of failure of the primary cooling pump 80. Furthermore, if both pumps fail, the logic should shut down the reactor.

[0147] Fig. 7 is an overview of another non-limiting example showing production rate controller logic for controlling a gas phase polymerization reaction, wherein production is constrained by desired production rate, desired purge bin level, maximum weight percent condensed, and maximum cooling water valve 70 position. In this regard, the purge bin is a temporary storage area for product prior to further processing.

[0148] In Fig. 7, a low signal selector 200 selects the lowest signal of several constraints as the constraint that limits production. The low signal selector 200 together with the PID controllers above the low signal selector 200, discussed below, perform the function of the rate controller of Fig. 5.

[0149] The setpoint signals of Fig. 7 include a desired production rate 202, a desired purge bin level 204, a maximum weight percent condensed 206, and an allowable maximum cooling water (CW) valve position 208.

[0150] The desired production rate 202 and a measured current production rate 210 are compared in a PID production rate controller (PRATC) 212 which sends a production
rate signal 214 to the low signal selector 200. The production rate controller (PRATC) 212 is reverse acting (if production rate (PV) increases, output must decrease). The output signal 214 is the total RPM fed to the reactor, as indicated by production rate.

[0151] The desired purge bin level 204 and a measured purge bin level 220 are compared in a PID purge bin level controller (PBNLC) 222 which sends a purge bin level signal 224 to the low signal selector 200. The purge bin level controller (PBNLC) 222 is reverse acting (if level (PV) increases, output must decrease). The output signal 224 is the total RPM fed to the reactor, as indicated by purge bin level.

[0152] The maximum weight percent condensed 206 and a measured weight percent condensed 230 are compared in a PID weight percent condensed controller (WCONC) 232 which sends a weight percent condensed signal 234 to the low signal selector 200. The weight percent condensed controller (WCONC) 232 is reverse acting (if the measured weight percent condensed 230 increases, output must decrease). The output signal 234 is the total RPM fed to the reactor, as indicated by weight percent condensed.

[0153] In this regard, the operators set the maximum condensation limit as the setpoint of the controller. As the production rates are raised or other process conditions change, if the limit is approached, then the control logic automatically adjusts the catalyst RPM to maintain the condensation limit at or below the setpoint. While not wishing to be limited by experience, most of the time, one of the other constraints is normally selected.

[0154] The allowable maximum cooling water valve position 208 and a measured cooling water valve position 240, which may be an averaged cooling water valve 70 position, are compared in a PID valve position controller (VALVC) 242 which sends a valve position signal 244 to a PID cooling water delta temperature controller (CWDTC) 246 which also receives a measured cooling water delta temperature signal 248. The valve position controller (VALVC) functions as a master PID and is direct acting (if the measured CW valve position (PV) increases, output must increase). The cooling water delta temperature controller (CWDTC) 246 functions as a slave PID and is direct acting (if the measured CWDTC increases, output must increase). The valve position signal 244 is the total RPM fed to the reactor, as indicated by the cooling water valve position. The PID cooling water delta temperature controller (CWDTC) 246 sends an available cooling
signal 250, which is the total RPM fed to the reactor based on CWDT, to the low signal selector 200.

[0155] Thus, the cascade control loop formed by VALVC and CWDTC is for gauging and control of the cooling constraint. The cooling constraint is an important constraint in exothermic reaction systems, typically the most important constraint. Any design elements that improve the control of the cooling constraint directly help to maximize the production rate.

[0156] The cascade control scheme linearizes the nonlinear relation caused by the cooling water valve 70 characteristics. Based on observation and experience, it is possible to determine the CW valve 70 position limit beyond which negligible additional cooling is available. This position can vary from a CW valve position of about 65% to 80% open and needs to be determined experimentally for each reactor and each valve. As the CW valve 70 approaches this maximum limit, the master-slave PID combination works effectively to cut back on the catalyst feed rate. The same effectiveness cannot be achieved by using the standalone CWDT signal or any other variable indicative of the cooling constraint, such as flow rate through the cooling tower.

[0157] There is generally a one-to-one relationship between the CW valve 70 position and the CWDT. When the reactor is out of cooling (as manifested by almost fully open CW valve 70 position), the CWDT is at its minimum value (pinch point corresponding to no additional cooling). This relationship is capitalized in the cascade control scheme consisting of the VALVC and the CWDTC cascade points. The CWDTC is a linear signal and works well in the relatively fast-tuned slave (inside) loop. As the available cooling drops, the CWDTC slave effectively cuts back on its output, thus accurately taking the appropriate action. While the CWDTC takes the right action in the short-term horizon, it does not have the ability to truly identify the cooling limit in the long term. This can be more accurately determined by the average CW valve 70 position. This is because the valve position irrefutably shows the accurate availability (or lack of) of cooling. Using a filtered value of the valve position in the VALVC controller and using it to reset the setpoint of the CWDT controller marries the strengths of both signals. The combined cascade knows the accurate cooling limit and can respond with appropriate
speed. The VALVC controller cannot be used alone without the CWDTC because the valve position signal is too nonlinear, especially when the valve is open more than about 75% (the regime with most interest from the point of view of maximizing rates). By using the two controllers in a cascade structure, both the short-term and long-term control objectives are achieved.

[0158] Accordingly, the output from each PID, i.e., 212, 222, 232, and 246, is in units of catalyst feeder speed. Accordingly, the lowest output from the four PIDs (the constraining output) is selected by the low signal selector 200.

[0159] The low signal selector 200 interacts with a custom calculation logic controller 260 via signals 262, which include the lowest signal of the four PIDs. The custom calculation logic controller 260 performs the function of both the catalyst feeder balancing controller 130 and the block and feeder startup and initialization control logic 140 of Fig. 5.

[0160] Regarding the custom calculation logic controller 260, the presence of two catalyst feeders 16 with two independent blocks 46 on each catalyst feeder 16 poses certain challenges for the control configuration and commissioning. Special control logic is required for proper initialization, bump-less transfer from Local to Remote (Auto-PCAS (i.e., program cascade)) modes. For maintaining clarity and ease of maintenance, the complete control logic is split into four main logic blocks, i.e., RPM initializer logic, RPM splitter logic, setpoint processor logic, and RPM anti-windup logic. The logic may be implemented differently depending on the controller platform type. For instance, the logic blocks shown herein may be combined or divided without departing from the scope of the present invention.

[0161] As shown in Fig. 7, the custom calculation logic controller 260 receives a feeder 1 on/off signal 264, a feeder 2 on/off signal 266, a block 11 on/off signal 268, a block 12 on/off signal 270, a block 21 on/off signal 272, and a block 22 on/off signal 274. The feeder 1 on/off signal 264 and the feeder 2 on/off signal 266 indicate whether feeders 1 and 2, respectively, are on or off. The block 11 on/off signal 268 and the block 12 on/off signal 270 indicate whether the first and second, respectively, blocks 46 (see Fig. 3) of the first catalyst feeder are operational or, e.g., plugged. Similarly, the block 21 on/off
signal 272 and the block 22 on/off signal 274 indicate whether the first and second, respectively, blocks 46 (see Fig. 3) of the second catalyst feeder are operational or, e.g., plugged.

[0162] As disclosed in more detail with respect to Fig. 8, discussed below, the RPM initializer logic of the logic controller 260 is intended to provide a smooth transition from operator control to automatic control. Reference is also made to the Appendix of the present invention which includes a nomenclature list for the exemplary control logics of the present application.

[0163] A non-limiting example of the RPM initializer logic is shown in Fig. 8. The logic block is preferably run at scan time of preferably between about 1 to 60 seconds, more preferably about 5 to 30 seconds, and most preferably about 15 seconds.

[0164] In general, the logic checks whether catalyst feeder #1 and #2 are ready to accept a remote setpoint from the advanced control strategy. A catalyst feeder is considered to be able to accept a remote setpoint if one of the blocks is open, feeder motor is running and the respective RPM PID is in Auto and Remote states and is not initialized. If all conditions are satisfied, then the binary output flags (BO01 and BO02) are set true or false (0 or 1).

[0165] The logic then determines the total number of blocks feeding the reactor. For a block to be feeding the reactor, it should be “Open” and its respective feeder motor must be “Running”. The total number of blocks feeding the reactor is then calculated.

[0166] The total RPM from all “Open” blocks and “Running” feeders is calculated and written to RO02. If the control strategy is not in “Ready” state (cannot be turned ON) because none of the blocks are in the appropriate modes, then the RO02 value (total RPM) is written to RO01. RO01 is used to initialize the outputs of the upstream PIDs. If the strategy is in “Ready” mode, then the RO02 value is not copied to RO01.

[0167] Referring to Fig. 8 in more detail, the logic is begun at start 300. At 302, the logic determines whether block 11 or block 12 of the first catalyst feeder is open. If neither block is open, BO01 is set to be false at 304. If at least one of the blocks is open, at 306, the logic determines whether the motor of the first catalyst feeder is on. If the motor of the first catalyst feeder is not on, then BO01 is set to be false at 304. If the
motor of the first catalyst feeder is on at 306, the logic determines whether the RPM PID of the first catalyst feeder is in the Auto and Remote states and is not initialized at 308. If the RPM PID of the first catalyst feeder is not in a condition wherein it is in the Auto and Remote states and is not initialized, then BO01 is set to be false at 304. If the RPM PID of the first catalyst feeder is in a condition wherein it is in the Auto and Remote states and is not initialized at 308, then BO01 is set to be true at 310.

[0168] At 320, the logic determines whether block 21 or block 22 of the second catalyst feeder is open. If neither block is open, BO02 is set to be false at 322. If at least one of the blocks is open, at 324, the logic determines whether the motor of the second catalyst feeder is on. If the motor of the first catalyst feeder is not on, then BO02 is set to be false at 322. If the motor of the second catalyst feeder is on at 324, the logic determines whether the RPM PID of the second catalyst feeder is in the Auto and Remote states and is not initialized at 326. If the RPM PID of the first catalyst feeder is not in a condition wherein it is in the Auto and Remote states and is not initialized, then BO02 is set to be false at 322. If the RPM PID of the second catalyst feeder is in a condition wherein it is in the Auto and Remote states and is not initialized at 326, then BO02 is set to be true at 328.

[0169] At 330, the logic determines whether BO01 or BO02 is true. If neither BO01 nor BO02 is true, then BO03 is set to be false at 332. If either BO01 or BO02 is true, then BO03 is set to be true at 334.

[0170] At 340, the logic calculates the number of running blocks in the first catalyst feeder. The logic also calculates the number of running blocks in the second catalyst feeder. From these calculations, the logic calculates the total number of running blocks. The logic also calculates the total RPM feeding the reactor. The logic further sets BO04 to be the inverse of BO03.

[0171] At 350, the logic determines whether BO04 is true. If BO04 is not true, the logic exits at 360. If BO04 is true, then RO01 is set to equal RO02 at 352 and then the logic exits at 360.

[0172] Referring again to Fig. 7, the RPM splitter logic of the logic controller 260 is intended to significantly increase the automation and control quality in the production
rate controller. In this regard, the catalyst in fine powder form is fed to the reactor by two motor driven catalyst feeders 16. As discussed above, each catalyst feeder 16 has two injection points in the form of blocks 46 (see Fig. 3). Thus, there are a total of four blocks driven by the two feeders. The total catalyst injected into the reactor is a function of injection points, i.e., the number of catalyst feeders 16 running multiplied the number of open (feeding) blocks 46.

[0173] Based on the input signals shown in Fig. 7, the control logic checks the control status of each catalyst feeder 16. If the feeder is not ready to accept a supervisory setpoint, e.g., because the feeder has been shutdown for a refill or if the logic tripped the catalyst feeder due to a blockage of a downstream injection point, the total RPM is calculated and saved in memory as local or operator-set RPM.

[0174] The operator-set RPM is subtracted from the total RPM (output of the Low Selector tag) to calculate the RPM in "Auto" mode (RPM that can be manipulated by the Rate Control strategy).

[0175] The logic calculates the total number of blocks that are ready to accept a remote setpoint from the Rate Control strategy. The number of ready blocks is the number of Auto blocks.

[0176] The total Auto RPM is divided by the number of Auto blocks to calculate the setpoints for the RPM controllers.

[0177] If the feeder is not ready to accept a remote setpoint (e.g., because the feeder is in the Manual, Local or Initialized state), then the setpoints are not downloaded. Instead, the value of the setpoints are matched with the current RPM PID setpoint set by the operator. If both feeders are in local mode (not remote mode), then the logic adds total RPM from all the blocks and writes the sum to the output of the low signal selector.

[0178] A non-limiting example of the RPM splitter logic is shown in Fig. 9. The logic block is preferably run at scan time of preferably between about 1 to 60 seconds, more preferably about 5 to 30 seconds, and most preferably about 15 seconds. In general, the logic of Fig. 9 involves a check of RPM PID status of each feeder.

[0179] Looking at Fig. 9 in more detail, the logic is started at start 400. At 402, if the feeder RPM PID is in either in Local or Manual mode, then the RPM PID's setpoint is
multiplied by the number of blocks running and this value is saved as the Local RPM (RPM that is under control of the operator). The feeder #1 value is saved as M01 and the feeder #2 value is saved as M02. These two values are added to calculate the total RPM in Local/Manual state and this value is saved as RO02. This value is called Operator RPM. The Operator RPM is subtracted from the total RPM (output of the Low Selector tag) to calculate the RPM in Auto mode (RPM that can be manipulated by the Rate Control strategy). This value is saved as RO01, which is called Auto RPM. The total number of blocks in Auto and Remote modes (blocks that are ready to accept a remote setpoint from the Rate Control strategy) is calculated. This value is saved as IO01, which is the number of Auto blocks.

[0180] At 404, the logic determines whether IO01 is not zero. If IO01 is not zero, i.e., if there is at least one remotely controlled block, then, at 406, the total Auto RPM is divided by the number of Auto blocks to calculate the setpoints for the #1 and #2 rate-of-change RPM limiter blocks 280, which setpoints are sent via signals 278 to the rate-of-change limiter blocks 280 (see Fig. 6). The outputs 282 from these blocks 280 constitute the setpoints for the slave RPM PDs 290.

[0181] At 408, if the first catalyst feeder is not ready to accept a remote setpoint because of any reason (Manual, Local or Initialized state), then the setpoint is not downloaded to the rate of change blocks 280. Instead, at 410, the value of the setpoint for downloading is set equal to the current RPM PID setpoint set by the operator.

[0182] Similarly, at 412, if the first catalyst feeder is not ready to accept a remote setpoint because of any reason (Manual, Local or Initialized state), then the setpoint is not downloaded to the rate of change blocks 280. Instead, at 414, the value of the setpoint for downloading is set equal to the current RPM PID setpoint set by the operator.

The logic then exits at 416.

[0183] The RPM splitter logic significantly improves the automation level and convenience for control room operators. When the production rate controller is started-up, the upper level controllers generate the RPMs based on their setpoints and process conditions. The RPM splitter logic carries further by splitting the total RPM into RPM
required in each block. Control stability is noticeably improved because of tighter maintenance of the total RPMs generated at any time instance.

[0184] Referring again to Fig. 7, the setpoint processor logic of the logic controller 260 is intended to smooth the transition from manual to automatic control by ensuring that the setpoints are varied smoothly.

[0185] In this regard, if the upstream PIDs- VALVC 242, WCONC 232, PRATC 212, and PBNLC 222 are configured with Setpoint Tracking active, then when these PIDs are in initialized state (when slave RPM PIDs are not ready to accept remote setpoint), then the setpoints of the upstream PIDs will match their respective current measurements.

This condition is not desirable since all the setpoints of these PIDs need to be somehow saved and restored when the control strategy is active.

[0186] If the configuration is set for no Setpoint Tracking, then the setpoints will be saved based on the last entry, but this can be undesirable if the difference between the setpoint and current measurement is high.

[0187] The five upstream PIDs VALVC, PRATC, WCONC, PBNLC and CWDTC should all be configured with Setpoint Tracking active. The four PIDs VALVC, PRATC, WCONC and PBNLC should be configured also with a limit on rate of setpoint change.

[0188] The drawbacks in both cases above are overcome by providing setpoint processor logic. A non-limiting example of the setpoint processor logic is shown in Fig. 10. The logic block is preferably run at scan time of preferably between about 1 to 60 seconds, more preferably about 5 to 30 seconds, and most preferably about 15 seconds.

[0189] The logic starts at 500. At 502, the logic determines whether the cooling water valve position PID (VALC) 242 is in manual or auto status, whether VALVC 242 is in local or remote status, and whether VALVC 242 is initialized. If the status of VALVC 242 is auto, remote, and noninitialized, then, at 504, the setpoint for VALVC 242 is set to be the maximum allowable limit for the valve position. If the status of VALVC 242 is either manual, local, or initialized, then, at 506, the setpoint for VALVC 242 is set to be the current cooling water valve position.

[0190] At 508, the logic determines whether the weight percent condensed PID (WCONC) 232 is in manual or auto status, whether WCONC 232 is in local or remote
status, and whether WCONC 232 is initialized. If the status of WCONC 232 is auto, remote, and noninitialized, then, at 510, the setpoint for WCONC 232 is set to be the maximum allowable limit for weight percent condensed. If the status of WCONC 232 is either manual, local, or initialized, then, at 512, the setpoint for WCONC 232 is set to be the current weight percent condensed.

[0191] At 514, the logic determines whether the production rate PID (PRATC) 212 is in manual or auto status, whether PRATC 212 is in local or remote status, and whether PRATC 232 is initialized. If the status of PRATC 212 is auto, remote, and noninitialized, then, at 516, the setpoint for PRATC 212 is set to be the desired production rate. If the status of PRATC 212 is either manual, local, or initialized, then, at 518, the setpoint for PRATC 212 is set to be the current production rate.

[0192] At 520, the logic determines whether the purge bin level PID (PBNLC) 222 is in manual or auto status, whether PBNLC 222 is in local or remote status, and whether PBNLC 222 is initialized. If the status of PBNLC 222 is auto, remote, and noninitialized, then, at 522, the setpoint for PBNLC 222 is set to be the purge bin level setpoint. If the status of PBNLC 222 is either manual, local, or initialized, then, at 524, the setpoint for PBNLC 222 is set to be the current purge bin level. The logic exits at 526.

[0193] With this logic and the standard setpoint-tracking feature, when the rate control strategy is activated, all the upstream setpoints start with their initial measurements and then ramp toward the desired value. In this manner, all setpoints are always saved at their last respective values. At the same time, at the time of activation of the strategy, a large bump in the catalyst feeder RPM because of a potentially large deviation between the setpoint and measurement of the upstream PIDs is avoided.

[0194] Referring again to Fig. 7, the RPM anti-windup logic of the logic controller 260 is intended to ensure that the automatic control stays within the mechanical limits of the catalyst feeder, e.g., maximum catalyst feed rate.

[0195] The RPM anti-windup logic provides protection against windup in case too few blocks are running. For example, if all four blocks are running with each block at 500 RPM, total RPM to the reactor is 2000. If three blocks are shut down, the single running block will ramp up to the max RPM limit of, e.g., about 1600 RPM. In this case, the rate
controller is asking for 2000 RPM, but only 1600 RPM are available. Loss in production rate is unavoidable here. The control logic freezes the integral windup of the upstream PIDs in such a scenario. Otherwise, the upstream PIDs will slowly keep ramping up their outputs and the controller will be unaware that the extra RPM cannot be achieved.

[0196] A non-limiting example of the RPM anti-windup logic is shown in Fig. 11. The logic block is preferably run at scan time of preferably between about 1 to 60 seconds, more preferably about 5 to 30 seconds, and most preferably about 15 seconds.

[0197] At 600, the logic starts. At 602, the RO02 value is the total RPM from all running blocks and feeders, plus 10 is added as a tolerance for comparison against the output from the low selector. At 604, the logic determines whether RO02 is less than the output from the low signal selector 262. If the RO02 is not less than the output from the low signal selector 262, then, at 606, BO01 is set to false, at 608, the RPM value used for inputting to the feedback parameter for upstream PIDs (to prevent wind-up) is set equal to the output from the low signal selector 262, and the logic exits at 610. If RO02 is less than the output from the low signal selector 262, then, at 612, the logic determines whether BO01 is true. If BO01 is true, the logic exits at 610. If BO01 is false, BO01 is set to be true at 614, the RPM value used for inputting to the feedback parameter for upstream PIDs (to prevent wind-up) is set equal to the output from the low signal selector 262 at 608, and the logic exits at 610.

[0198] Since the value of 10 is added at 602, if the difference between the total RPM being fed to the reactor and the low selector output is more than 10 RPM (tolerance), the windup protection is activated.

[0199] Normally, the output from the low selector will be the same as the total RPM to reactor and flag BO01 (used internally and for display purposes only) is set to false (cleared). In this normal state, the feedback value for the upstream PIDs is set equal to the output from the low selector.

[0200] When the windup condition is true, the flag BO01 is set true. As long as the condition remains true, the feedback signal is set equal to the last saved low selector output. Windup is thus prevented.
[0201] The RPM initializer logic, RPM splitter logic, setpoint processor logic, and RPM anti-windup logic may be performed in series or in parallel. Preferably, this logic is performed in parallel.

[0202] Referring again to Fig. 7, as discussed above, signals 278 from the custom calculation logic 260 are sent to the rate-of-change limiter blocks 280. The rate-of-change limiter blocks 280 account for mechanical limitations of equipment which cannot tolerate abrupt changes. The rate of change limiter blocks 280 send signals 282 to the catalyst RPM controllers 290. The RPM controllers compare desired speed and actual speed to adjust the amperage to the catalyst feeder motor to adjust the speed of the catalyst feeders 16 accordingly.

[0203] Fig. 12 is a non-limiting example of a production rate controller operator interface. There are only three action items for the operator:

1. Enter desired production rate setpoint. If rates need to be maximized, then a high unachievable number should be set, e.g., 50 ton/hr (907 kg/hr).
2. Specify purge bin level setpoint. Normally, there is not much need to change this setpoint.
3. To activate the rate controller, set one or both feeders to Auto and Remote. If one of the feeders is in Auto and Remote, then the rate controller begins manipulating its RPM setpoint while the other feeder is at fixed RPM.

[0204] To disable the rate controller, both feeders should be set to Local. Once the setpoints are appropriately set, turning the rate controller from off to on and vice versa, just involves setting one or both feeders from Local to Remote. There is no other action needed on any of the other tags.

[0205] Fig. 13 is a non-limiting example of a separate controller screen which is provided for engineer interface. The engineer screen has more information for checking control performance and ensuring that all features of the controller are working well. All detailed information is provided on the engineer screen. This screen allows keeping the operator interface screen simple with just the required information for the operator. This
arrangement helps to avoid confusion while training new operators and makes the overall operation and understanding easy and convenient.

[0206] At any given instance, the low signal selector selects the lowest of the four outputs. The selected tag is highlighted with a box, as shown in the Fig. 13. In this regard, the average cooling water valve position and the average CWD are important diagnostic values that help the engineer to make sure that production rate is being fully maximized. Preferably, a 20 or 30 minute average is displayed.

[0207] The five PID controllers- PRATC, PBNLC, WCONC, CWDTC, VALVC should be precisely tuned using an industrial process control software package. For instance, as discussed above, PITOPS™ software, available from Artcon Incorporated, Cincinnati, OH, for dynamics identification and determining optimal PID tuning parameters by simulation.

[0208] Since the rate controller benefits can be achieved only at a constraint limit, tight control action at the constraint limit is important. Sub-optimal or slightly conservative tuning can easily cause a constraint to be violated followed by slower-than-required control action. This action will cause an upset or result in the operator disabling the rate controller.

[0209] For determining optimal tuning parameters, data is studied in PITOPS™ software and the transfer function is identified. Five transfer functions are identified, one for each CV-MV pair, i.e., controlled variable and manipulated variable pair, e.g., constraint variable and catalyst feed rate.

[0210] After the transfer function is identified, a simulation is preferably performed, e.g., by PITOPS™ software. Normal setpoint changes, typical unmeasured disturbances and white noise are configured to make the simulation match the real controller behavior as closely as possible. The software then determines optimal tuning parameters. At this stage, several "what-if" studies are conducted to generate the most aggressive and non-oscillatory control action within limits of how fast the catalyst RPM can be moved. The simulation is helpful in gaining the confidence of the operators before activation of the rate control strategy.
[0211] In a complex process such as polymerization, there appears to be a number of manipulated variables that can affect a single control variable such as production rate. These include, recirculating reaction gas flow rate, the gas composition, change in catalyst productivity, reactor pressure and so on. This suggests that a multivariable approach, using any of the commercial MPC (Model Predictive Controllers) might be more suitable approach for control. However, an important characteristic of PE reactor (and perhaps typical of polymerization reactors) is that often there is a single key manipulated variable that is used to achieve the desired control variable change. Thus, the control problem becomes one of a sparse (diagonal Matrix) of Controlled and Manipulated variables. In other words, the control problem is not linked, as the diagonal matrix has one-to-one relationship. The second feature is that the control variables tend to be open loop unstable, drifting to spiraling region of instability at the slightest perturbation. This second feature is especially the case with the reactor bed temperature control. Thus, the reaction control system lends itself to the present approach of identifying the key dynamics between the controlled and manipulated variables and the use of classical advanced regulatory control techniques, such as PID algorithm, and cascade control.

[0212] Examples presented were related to the particular constraint of cooling limitation. However the methodology employed can be extended to other constraints following the guidance set forth herein, such as desired production rate, dew point limit, weight percent condensed, weight percent condensables, removal of product (e.g., purge bin level), level of fluidization, level of entrainment, pelletizer, granular resin transfer capability, and feed supply (e.g., feed supply pressure).

[0213] Since any change in the rate of catalyst injection will change the reaction rate and hence rate of generation of the heat of reaction, the temperature of the recycle stream entering the reactor is adjusted upwards and downwards to accommodate any change in the rate of heat generation. This ensures the maintenance of an essentially constant temperature in the bed. Complete instrumentation of both the fluidized bed and the recycle stream cooling system is, of course, useful to detect any temperature change in the bed.
[0214] Under a given set of operating conditions, the fluidized bed is maintained at essentially a constant height by withdrawing a portion of the bed as product at the rate of formation of the particulate polymer product. Since the rate of heat generation is directly related to the rate of product formation, a measurement of the temperature rise of the fluid across the reactor (the difference between inlet fluid temperature and exit fluid temperature) is indicative of the rate of particulate polymer formation at a constant fluid velocity if no vaporizable liquid is present in the inlet fluid.

[0215] On discharge of particulate polymer product from reactor 10, it is desirable and preferable to separate fluid from the product and to return the fluid to the recycle line 22. There are numerous ways known to the art to accomplish this, such as described in U.S. Patent No. 4,543,399 to JENKINS, III et al..

[0216] The fluidized-bed reactor can be equipped with an adequate venting system (not shown) to allow venting the bed during start up and shut down. The reactor does not require the use of stirring and/or wall scraping. The recycle line 22 and the elements therein (compressor 30, heat exchanger 24) are preferably smooth surfaced and devoid of unnecessary obstructions so as not to impede the flow of recycle fluid or entrained particles.

[0217] While not limited to any particular type or kind of reaction, this invention is particularly suited to catalyst cracking and polymerization reactions involving the polymerization of one or more of the monomers listed below:

I. Olefins: such as those having 2 to 20, preferably 2 to 15 carbons, such as ethylene, propylene, 1-butene, 1-pentene, 4-methyl-1-pentene, 1-octene, 1-hexene, styrene.

II. Polar vinyl monomers: such as vinyl chloride, vinyl acetate, vinyl acrylate, methyl methacrylate, tetrafluoroethylene, vinyl ether, acrylonitrile.

III. Dienes (conjugated and non-conjugated): such as butadiene, 1,4-hexadiene, isoprene, ethylidene norbornene.

IV. Acetylenes: such as acetylene, substituted acetylene, such as methyl acetylene.

V. Aldehydes: such as formaldehyde.
[0218] The present invention has been found to be especially advantageous in polymerizations of olefin-type monomers.

[0219] The present invention is generally directed to any type of catalyst, e.g., cracking catalysts and polymerization catalysts. For example, catalysts employable in the fluidized bed polymerization of the above monomer types would, respectively, most usually be as follows:

I. Coordinated anionic catalyst

II. Cationic catalyst for copolymers with ethylene only; others of this type require a free-radical catalyst

III. Either a free-radical catalyst or a coordinated anionic catalyst

IV. Coordinated anionic catalyst

V. Anionic catalyst

VI. Metallocenes such as those disclosed in U.S. Patent No. 5,405,922 to DeCHELLIS et al. and U.S. Patent No. 4,937,301 to CHANG, the disclosures of which are herein incorporated by reference in their entireties.

[0220] Among the polymers which may be produced in the process of the present invention are homopolymers of ethylene, propylene, butene or copolymers of a major mole percent of ethylene, propylene or butene and a minor mole percent of one or more C2 to C8 alpha-olefins. The C2 to C8 alpha-olefins preferably should not contain any branching on any of their carbon atoms which is closer than the fourth carbon atom. The preferred C2 to C8 alpha-olefins are ethylene, propylene, 1-butene, 1-pentene, 1-hexene, 4-methyl-1-pentene, and 1-octene.

[0221] The ethylene homopolymers preferably have a density of about 0.958 to 0.972 g/cm³, more preferably about 0.958 to 0.967 g/cm³, and most preferably about 0.958 to 0.965 g/cm³.

[0222] The ethylene copolymers have a density less than about 0.96 g/cm³, with ranges of preferably about 0.890 to 0.960 g/cm³, more preferably about 0.910 to 0.958 g/cm³, and most preferably about 0.910 to 0.955 g/cm³. The density of the ethylene copolymer, at a given melt index level for the copolymer, is primarily regulated by the amount of the C3 to C8 comonomer which is copolymerized with the ethylene. In the absence of the
comonomer, the ethylene would homopolymerize to provide polymers having a density of about 0.96 g/cm$^3$. Thus, the addition of progressively larger amounts of the comonomers to the copolymers results in a progressive lowering of the density of the copolymer. The amount of each of the various C3 to C8 comonomers needed to achieve the same result will vary from monomer to monomer, under the same reaction conditions.

[0223] Thus, to produce binary copolymers of ethylene with the same density and melt index, larger molar amounts of the different comonomers would be needed in the order of C3 < C4 < C5 < C6 < C7 < C8.

[0224] When made in the fluid-bed process described herein, ethylene polymers are granular materials which have a settled bulk density of typically about 15 to 40 lb/m$^3$ (240 to 640 kg/m$^3$), more typically about 15 to 38 lb/m$^3$ (240 to 610 kg/m$^3$), and most typically about 17 to 32 lb/m$^3$ (270 to 510 kg/m$^3$), with an average particle size of the order of typically about 0.005 to 0.06 inches (0.13 to 1.5 mm), more typically about 0.010 to 0.040 inches (0.25 to 1.0 mm), and most typically about 0.012 to 0.030 inches (0.30 to 0.76 mm), as measured with mesh sieves. Particle size is important for the purposes of readily fluidizing the polymer particles in the fluid-bed reactor, as herein described.

[0225] When operating in condensed mode, while the liquid content of the two-phase recycle stream can be quite high, as a general rule, the quantity of condensed liquid contained in the gas phase preferably should not exceed about 50 wt%, more preferably should not exceed about 45 wt%, and most preferably should not exceed about 20 wt%, with ranges of preferably about 2 to 50 wt%, more preferably about 5 to 45 wt%, and most preferably about 5 to 40 wt%, provided that the velocity of the two-phase recycle stream is high enough to keep the liquid phase in suspension in the gas and to support the fluidized bed within the reactor.

[0226] The space time yield of the present invention is typically up to about 30 lb/hr-ft$^3$ (480 kg/hr-m$^3$), with ranges of typically about 5 to 25 lb/hr-ft$^3$ (80 to 400 kg/hr-m$^3$), and most typically about 5 to 20 lb/hr-ft$^3$ (80 to 320 kg/hr-m$^3$). Although dependent on reactor size, the production rate of the present invention is typically about 10 to 100
ton/hr (9100 to 91,000 kg/hr), more typically about 20 to 60 ton/hr (18,000 to 54,000 kg/hr), and most typically about 30 to 50 ton/hr (27,000 to 45,000 kg/hr).

[0227] The target production rate typically ranges from about 10 to 100 tons/hr (9070 to 90,700 kg/hr), more typically about 20 to 60 tons/hr (18,100 to 54,400 kg/hr), and most typically about 30 to 50 tons/hr (27,200 to 45,400 kg/hr). In this regard, the target production rate is determined by factors such as inventory, limits of the reactor, limits of other process constraints, feed availability, and marketing decisions. The target cooling valve position typically ranges from about 20 to 100% open, more typically from about 30 to 80% open, and most typically from about 40 to 70% open. The target dew point approach is typically up to about 20°C, more typically up to about 10°C, with a most typical range of about 3 to 7°C. The desired purge bin level typically ranges from about 10 to 95%, more typically about 20 to 80%, and most typically about 30 to 70%. The target weight percent condensed typically ranges from about 2 to 50 wt%, more typically about 5 to 45 wt%, and most typically about 5 to 40 wt%.

[0228] The present method vastly simplifies the complex interactions that are present in polymerization reactor control to a manageable few key constraints. Commercial controllers such as MPC (Model Predictive Control) systems might have been considered as a potential alternative for this application. As stated previously the sparse nature of the key matrix variables and the open-loop unstable dynamics renders the use of the MPC systems unnecessary.

[0229] Thus, an advantage of the present invention is that stability and control of the reactor is generally improved.

[0230] Since the stability of the reactor is improved, the present invention can be used to increase production. For instance, since bed temperature variations may be reduced, the temperature may be set at a level to optimize catalyst activity. In this regard, some catalyst systems are temperature sensitive, such as high activity catalysts, such as chrome oxide catalysts, Ziegler-Natta catalyst, and metallocene catalysts.

[0231] Moreover, since the stability of the reactor is improved, the present invention increases control of product characteristics, especially in those reactions which are temperature sensitive.
[0232] Another advantage of improved stability is that the amount of waste may be reduced and environmental impact minimized, especially in those reactions which are temperature sensitive.

[0233] The present invention may also be used to increase production by maximizing production with respect to active constraints. In this regard, the present invention preferably continuously monitors the constraint parameters so that close control of the process may be effected. In methods involving manual operator control of catalyst feed rate, operators tend to operate in a safe or comfort zone. With the systems and methods of the present invention, the need to maintain this comfort zone is eliminated, and the operators can safely maintain higher average production rates.

[0234] Accordingly, another advantage of the present invention is improved safety. By automating the catalyst feed, the present invention eliminates the possibility of human error. In addition to eliminating the possibility of human error in controlling the catalyst feed rate, the improved reaction control of the present invention minimizes the risk of temperature run-away, e.g., bed temperature run-away in exothermic polymerization reactions.

[0235] Yet another advantage of the present invention is that the increased automation of the present invention facilitates reactor operation for operators. Instead of constantly adjusting the catalyst feed rate, operators can focus on other aspects of the process.

[0236] Additionally, the control system of the present invention is relatively simple and economical when compared with other control systems.

[0237] In view of all of the above advantages, the present invention reduces overall operating costs.

[0238] The recognition of the present invention of the importance of the dynamics of the cooling water delta temperature (CWDT) represents an important finding.

[0239] The synergy effect in dampening the bed temperature variations (self-quenching action) in an anticipatory manner using the regulatory control scheme of the present invention is another important finding.
[0240] Although the present invention is primarily directed to automation of the reaction control, the automated aspects of the present invention may be combined with manual control of other aspects of the reaction.

[0241] For the sake of brevity and clarity, known equipment has not been described in detail in the above description.

[0242] The present invention will be further illustrated by way of the following Examples. These examples are non-limiting and do not restrict the scope of the invention.

[0243] Unless stated otherwise, all percentages, parts, etc. presented in the examples are by weight.

EXAMPLES AND COMPARATIVE EXAMPLES

[0244] The following Examples and Comparative Examples involve polymerization of ethylene using a “Unipol G-5000” reactor available from Union Carbide.

EXAMPLES 1-3 and COMPARATIVE EXAMPLE 1

[0245] Examples 1-3 and Comparative Example 1 were carried out under the conditions described in Table 1. For Examples 1-3, the control system was a Foxboro IA (Intelligent Automation) control system programmed with the logic shown in Figs. 7-11. In Comparative Example 1, the catalyst feed was manually controlled.

[0246] In Table 1, “Supported Ti” refers to a complex of tetrahydrofuran, magnesium chloride, and titanium chloride reduced with diethyl aluminum chloride (diethyl aluminum chloride to tetrahydrofuran ratio of 0.5) and triethyl aluminum (triethyl aluminum to tetrahydrofuran ratio of 0.25) impregnated on triethyl aluminum treated silicon dioxide. “Chrome(1)” refers to chrome oxide catalyst supported on silicon dioxide. “Chrome(2)” refers to a bis-triphenyl silyl chromate supported on silicon dioxide with 0.25 wt% of chrome.
<table>
<thead>
<tr>
<th>Characteristic</th>
<th>Example 1</th>
<th>Example 2</th>
<th>Example 3</th>
<th>Comparative Example 1</th>
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<tr>
<td>Recycle Ethylene (bar (kg/cm³))</td>
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<td>9 (9)</td>
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<td>Recycle H2/C2 (molar ratio)</td>
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<td>5 (5)</td>
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<td>Recycle stream dew point (dry mode) (°C)</td>
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<td></td>
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<td>Recycle Isopentane (bar)</td>
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<td>Wt% condensed in recycle stream (condensing mode)</td>
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<td>Catalyst</td>
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<td>Superficial gas velocity in fluidized bed (m/s)</td>
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<td>Space time yield (lb/hr/ft³ (kg/hr/m³))</td>
<td>11 (180)</td>
<td>11 (180)</td>
<td>10 (160)</td>
<td>7.6 (120)</td>
</tr>
</tbody>
</table>
EXAMPLE 1

[0247] In this Example, the benefit of the production rate controller in its ability to consistently maintain reactor rates close to the cooling constraint is illustrated. The production rates vary slightly, always ensuring that the cooling constraint (CWDT set point) is maintained.

[0248] Fig. 14 shows that production rates can be maximized by adjusting catalyst feed rate and minimizing cooling water delta temperature (CWDT). In Fig. 14, the production rate is in ton/hr, the purge bin level is in meters, the weight percent condensed is in percentage, the cooling water valve position is in percentage open, and the cooling water delta temperature (CWDT) is in °C. The CWDT setpoint was 1°C and the production rate setpoint was 32 ton/hr.

EXAMPLE 2

[0249] In Fig. 15 the ability of the production rate controller to move from one constraint to another to optimize rates continuously is illustrated. In Fig. 15, the production rate is in ton/hr, the purge bin level is in meters, the weight percent condensed is in percentage, the cooling water valve position is in percentage open, and the cooling water delta temperature (CWDT) is in °C. The initial setpoint for the production rate was 19 ton/hr and the final setpoint for the production rate was 40 ton/hr. The CWDT setpoint was 1°C.

[0250] Referring to Fig. 15, the reactor had early in the day a downstream constraint (due to granular resin handling and compounding) and this required the production rate to be maintained at a lower value. Upon alleviating this constraint, the rates were automatically raised by adjusting the set point of the production rate controller till the second constraint (cooling water delta temperature) was achieved, thus optimizing rates.

EXAMPLE 3

[0251] This Example illustrates the improved stability achieved by a closed loop catalyst feeder control by adjusting the feed rate. In this regard, it is important to
maintain the catalyst inventory in the reactor. Fig. 16 shows that the production rate variations are directly related to the catalyst inventory changes. At constant productivity of catalyst, maintaining stable catalyst inventory is important to achieve higher average production rates.

[0252] In Fig. 16, the production rate is in ton/hr, the purge bin level is in meters, the weight percent condensed is in percentage, the cooling water valve position is in percentage open, and the cooling water delta temperature (CWDT) is in °C. The initial production rate setpoint was 32 ton/hr and the final production rate setpoint was 32 ton/hr. The CWDT setpoint was 11.5°C.

COMPARATIVE EXAMPLE 1

[0253] This Comparative Example shows the importance of dew point controller function in preventing plate fouling that will result in a reactor shutdown. Fig. 17 shows that dew point varied in the reactor over a two day period, drifting lower.

[0254] A very low dew point resulted in rapid fouling of the distributor plate as indicated by a rising pressure drop. This pressure drop resulted in a reactor shut down. This shut down could have been avoided by the present invention which continuously monitors the dew point and controls the reaction to ensure operation does not transgress into a low dew point region while operating close to condensing.

EXAMPLES 4-6 AND COMPARATIVE EXAMPLES 2-4

[0255] In Example 4 and Comparative Example 2, the reaction conditions were the same as Example 3, as shown in Table 1. In Example 5 and Comparative Example 3, the reaction conditions were the same as Comparative Example 1, as shown in Table 1. In Example 6 and Comparative Example 4, the reaction conditions were the same as Examples 1 and 2, as shown in Table 1.

[0256] In the Examples, the control scheme was the same as in Example 1. In the Comparative Examples, the catalyst feed was manually controlled.
[0257] Table 2 summarizes the production credit achieved over many days of operation at close to constraint rates:

**TABLE 2**
Production Rate Improvement Demonstrated in Commercial Reactors
(averaged over number of days)

<table>
<thead>
<tr>
<th>Example</th>
<th>Catalyst</th>
<th>CWDT, minimum (°C)</th>
<th>Rate (normalized)</th>
</tr>
</thead>
<tbody>
<tr>
<td>4</td>
<td>Chrome(1)</td>
<td>11.5</td>
<td>106</td>
</tr>
<tr>
<td>5</td>
<td>Chrome(2)</td>
<td>8.9</td>
<td>108</td>
</tr>
<tr>
<td>6</td>
<td>Supported Ti</td>
<td>2.5</td>
<td>105</td>
</tr>
<tr>
<td>Comp. Ex. 2</td>
<td>Chrome(1)</td>
<td>14.9</td>
<td>100</td>
</tr>
<tr>
<td>Comp. Ex. 3</td>
<td>Chrome(2)</td>
<td>12.6</td>
<td>100</td>
</tr>
<tr>
<td>Comp. Ex. 4</td>
<td>Supported Ti</td>
<td>3.5</td>
<td>100</td>
</tr>
</tbody>
</table>

[0258] Thus, an incremental capacity of 5% at least is demonstrated.

EXAMPLES 7-8 AND COMPARATIVE EXAMPLES 5-6

[0259] The following Examples and Comparative Examples show that an important benefit of controlling based on CWDT is overall production rate maximization, by running to the true cooling water limit consistently. A secondary benefit arises by this cascade control scheme in dampening reactor bed temperature excursions. The CWDT controller action will ensure that by cutting back on the catalyst RPM in anticipation of approach to the CWDT limit, consistent temperature deviation from the set-point is not sustained. The CWDT controller also automatically ramps up the catalyst RPM if one of the catalyst pick-up blocks go down, thus taking up the slack to ensure target CWDT is maintained.

[0260] In Example 7 and Comparative Example 5, the reaction conditions were the same as Example 1, as shown in Table 1. In Example 8 and Comparative Example 6, the reaction conditions were the same as Examples 1 and 2, as shown in Table 1.

[0261] In the Examples, the control scheme was the same as in Example 1. In the Comparative Examples, the catalyst feed was manually controlled.
[0262] The following data shows that significant improvement is achieved, especially for the Chrome(1) catalyst due to the production rate control and its synergistic effect on temperature.

<table>
<thead>
<tr>
<th>Table 3</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bed Temperature Standard Deviations With and Without the Rate Controller</td>
</tr>
<tr>
<td>Example</td>
</tr>
<tr>
<td>---------</td>
</tr>
<tr>
<td>7</td>
</tr>
<tr>
<td>8</td>
</tr>
<tr>
<td>Comp. Ex. 5</td>
</tr>
<tr>
<td>Comp. Ex. 6</td>
</tr>
</tbody>
</table>

[0263] While the invention has been described in connection with certain preferred embodiments so that aspects thereof may be more fully understood and appreciated, it is not intended to limit the invention to these particular embodiments. On the contrary, it is intended to cover all alternatives, modifications and equivalents as may be included within the scope of the invention as defined by the appended claims.

APPENDIX OF NOMENCLATURE FOR EXEMPLARY CONTROL SYSTEM OVERVIEW

BI0n indicates a binary input signal fed into the control logic block. BO0n indicates a binary output signal generated by the control logic block. II0n indicates an integer input signal fed into the control logic block. IO0n indicates an integer output signal generated by the control logic block. RI0n indicates a real (floating point) input signal fed into the control logic block. RO0n indicates a real (floating point) output signal generated by the control logic block. Mnn indicates an internal variable whose value needs to be stored in memory between successive executions of the control logic block. The "~" symbol indicates an inverse. ~(True) = False and ~(False) = True.

RPM INITIALIZATION LOGIC

Real Inputs: RI01 = Current RPM from Feeder #1; RI02 = Current RPM from Feeder #2.
Binary Inputs: BI01 = Manual/Auto status of Feeder #1 RPM PID (Manual = 0, Auto = 1); BI02 = Local/Remote (Auto/PCAS) status of Feeder #1 RPM PID (Local = 0, Remote = 1); BI03 = Initialization Status (Tracking) of Feeder #1 RPM PID (INIT = 1, NOINIT = 0); BI04 = Manual/Auto status of Feeder #2 RPM PID (Manual = 0, Auto = 1); BI05 = Local/Remote (Auto/PCAS) status of Feeder #2 RPM PID (Local = 0, Remote = 1); BI06 = Initialization Status (Tracking) of Feeder #2 RPM PID (INIT = 1, NOINIT = 0); BI07 = Feeder #1 Block #1 Status (0 = Closed, 1 = Open); BI08 = Feeder #1 Block #2 Status (0 = Closed, 1 = Open); BI09 = Feeder #1 Motor Status (0 = Stopped, 1 = Running); BI10 = Feeder #2 Block #1 Status (0 = Closed, 1 = Open); BI11 = Feeder #2 Block #2 Status (0 = Closed, 1 = Open); BI12 = Feeder #2 Motor Status (0 = Stopped, 1 = Running).

Real Outputs: RO01 = Total RPM for initializing upstream PIDs (INITI parameter) when rate controller is OFF; RO02 = Total RPM for initializing upstream PIDs- for display only, when rate controller is ON.

Integer Outputs: IO01 = Number of Blocks from Feeder #1 feeding the reactor; IO02 = Number of Blocks from Feeder #2 feeding the reactor; IO03 = Total number of blocks feeding the reactor.

Binary Outputs: BO01 = If “zero” means Feeder #1 not ready to accept remote setpoint; if “one”, then Feeder #1 can accept a remote setpoint. BO02 = If “zero” means Feeder #2 not ready to accept remote setpoint; if “one”, then Feeder #2 can accept a remote setpoint. BO03 = “Zero” means “Initializing”, cannot turn advanced control strategy ON; “One” means at least one block can accept a remote setpoint, advanced control strategy is ready to turn ON. BO04 = Inverse of BO03.

RPM SPLITTER LOGIC

Real Inputs: RI01 = Output of Low Signal Selector (LOSEL shown in Fig. 6); RI02 = Current RPM setpoint to Feeder #1; RI03 = Current RPM setpoint to Feeder #2.

Binary Inputs: BI01 = BO01 from DCS control logic block (If “zero” means Feeder #1 not ready to accept remote setpoint. If “one”, then Feeder #1 can accept a remote setpoint.) BI02 = BO02 from DCS control logic block (If “zero” means Feeder #2 not
ready to accept remote setpoint. If “one”, then Feeder #2 can accept a remote setpoint.) BI03 = BO03 from DCS control logic block (“Zero” means “Initializing”, cannot turn advanced control strategy ON. “One” means at least one block can accept a remote setpoint, advanced control strategy is ready to turn ON.)

**Integer Inputs:** II01 = IO01 from DCS control logic block (Number of Blocks from Feeder #1 feeding the reactor); II02 = IO01 from DCS control logic block (Number of Blocks from Feeder #2 feeding the reactor).

**Real Outputs:** RO01 = Total RPM in Auto and Remote Mode (RPM that can be adjusted by the Rate Controller- remote setpoint can be downloaded); RO02 = Total RPM in Manual or Local Mode (RPM that is under control of the operator and cannot be adjusted by the Rate Controller- remote setpoint cannot be downloaded); RO03 = RPM setpoint to Feeder #1 Rate-of-Change block; RO04 = RPM setpoint to Feeder #2 Rate-of-Change block.

**Internal Variables:** M01 = Feeder #1 RPM under control of operator (RPM PID in Manual or Local Mode); M02 = Feeder #2 RPM under control of operator (RPM PID in Manual or Local Mode).

**SETPOINT PROCESSOR LOGIC**

**Real Inputs:** RI01 = CW Valve Maximum Allowable Limit; RI02 = % Weight Condensed Maximum Limit; RI03 = Current CW Valve Position; RI04 = Current % Weight Condensed Limit; RI05 = Desired Production Rate Setpoint; RI06 = Current Production Rate; RI07 = Purge Bin Level Setpoint; RI08 = Current Purge Bin Level.

**Binary Inputs:** BI01 = Manual/Auto status of CW Valve Position PID-tagname VALVC (Man =0, Auto =1); BI02 = Local/Remote (Auto/PCAS) status of VALVC (Local = 0, Remote = 1); BI03 = Initialization Status (Tracking) of VALVC (INIT = 1, NOINIT = 0); BI04 = Manual/Auto status of %Wt. Condensed PID- tagname WCONC (Man =0, Aut =1); BI05 = Local/Remote (Auto/PCAS) status of WCONC (Local = 0, Remote = 1); BI06 = Initialization Status (Tracking) of WCONC (INIT = 1, NOINIT = 0); BI07 = Manual/Auto status of Rate Control PID- tagname PRATC (Manual = 0, Auto = 1); BI08 = Local/Remote (Auto/PCAS) status of PRATC (Local
= 0, Remote = 1); BI09 = Initialization Status (Tracking) of PRATC (INIT = 1,
NOINIT = 0); BI10 = Manual/Auto status of Purge Bin Level PID-tagname
PBNLC (Manual = 0, Auto = 1); BI11 = Local/Remote (Auto/PCAS) status of PBNLC
(Local = 0, Remote = 1); BI12 = Initialization Status (Tracking) of PBNLC (INIT =
1, NOINIT = 0).

Real Outputs: RO01 = Setpoint for CW Valve Position PID- tagname VALVC;
RO02 = Setpoint for %Wt. Condensed PID- tagname WCONC; RO03 = Setpoint for
Rate Control PID- tagname PRATC; RO04 = Setpoint for Purge Bin Level PID-
tagname PBNLC.

RPM ANTI-WINDUP LOGIC

Real Inputs: RI01 = Current RPM setpoint of Feeder #1; RI02 = Current RPM
setpoint of Feeder #2; RI03 = Output from Low Signal Selector tagname LOSEL (see
Fig. 7).

Integer Inputs: II01 = IO01 from control logic block (Number of Blocks from Feeder
#1 feeding the reactor); II02 = IO02 from control logic block (Number of Blocks
from Feeder #2 feeding the reactor).

Real Outputs: RO01 = RPM value used for inputting to the FBK (feedback)
parameter for upstream PIDs (to prevent wind-up).

All patents, test procedures, and other documents cited herein, including
priority documents, are fully incorporated by reference to the extent such disclosure
is not inconsistent with this invention and for all jurisdictions in which such
incorporation is permitted.

What is claimed is:
CLAIMS:

1. A reaction process constraint override control method for optimizing production rates, the method comprising:
   determining available heat transfer and at least one additional process constraint; and
   automatically and continuously controlling catalyst feed rate to a reactor to control a reaction in response to available heat transfer and the at least one additional process constraint.

2. The method of claim 1, wherein the additional process constraint comprises current production rate of the reactor.

3. The method of claim 2, wherein the current production rate is compared with a desired production rate to control the catalyst feed rate.

4. The method of claim 1, wherein the additional process constraint comprises amount of reaction product in a collector of the reactor.

5. The method of claim 4, wherein the amount of reaction product in the collector is compared with a desired amount of reaction product in the collector to control the catalyst feed rate.

6. The method of claim 1, wherein the additional process constraint comprises weight percent condensed in a recycle stream of the reactor.

7. The method of claim 6, wherein the weight percent condensed is compared with a desired weight percent condensed.
8. The method of claim 1, wherein the additional process constraint comprises dew point of a recycle stream of the reactor.

9. The method of claim 8, wherein the dew point is compared with a desired dew point to control the catalyst feed rate.

10. The method of claim 1, wherein the additional process constraint comprises carryover of solids into a recycle stream.

11. The method of claim 10, wherein the carryover of solids is compared with a desired carryover of solids to control the catalyst feed rate.

12. The method of claim 1, wherein the additional process constraint comprises granular resin transfer system rate.

13. The method of claim 12, wherein the granular resin transfer system rate is compared with a desired granular resin transfer system rate to control the catalyst feed rate.

14. The method of claim 1, wherein the additional process constraint comprises pelletizer rate.

15. The method of claim 14, wherein the pelletizer rate is compared with a desired pelletizer rate to control the catalyst feed rate.

16. The method of claim 1, wherein the additional process constraint comprises feed flow.

17. The method of claim 16, wherein the feed flow is compared with a desired feed flow to control the catalyst feed rate.

18. The method of claim 1, wherein automatically and continuously controlling catalyst feed to the reactor in response to available heat transfer comprises monitoring a delta temperature of a mix point of heat transfer medium.
19. The method of claim 1, wherein automatically and continuously controlling catalyst feed to the reactor in response to available heat transfer comprises monitoring a delta temperature of a mix point of heat transfer medium and monitoring a position of a heat transfer medium valve of a heat transfer apparatus.

20. The method of claim 1, wherein automatically and continuously controlling catalyst feed rate to the reactor in response to available heat transfer comprises monitoring a position of a heat transfer medium valve of a heat transfer apparatus.

21. The method of claim 1, wherein the reactor comprises a heat transfer apparatus comprising a heat exchanger associated with the reactor and a conduit which is structured to allow circulation of heat transfer medium between the heat exchanger and a heat transfer source.

22. The method of claim 21, wherein the heat transfer apparatus comprises a recirculation conduit structured to allow at least a portion of the heat transfer medium to bypass the heat transfer source such that the at least a portion of the heat transfer medium is re-introduced into the heat exchanger without passing through the heat transfer source.

23. The method of claim 22, wherein the recirculation conduit is structured to, at a mix point, mix the at least a portion of the heat transfer medium with heat transfer medium which passes through the heat transfer source, prior to introduction into the heat exchanger.

24. The method of claim 23, wherein a first temperature sensor measures a first temperature of transfer medium at a point between the heat transfer source and the mix point; and a second temperature sensor measures a second
temperature of the heat transfer medium between the mix point and the heat exchanger.

25. The method of claim 24, wherein a controller subtracts the first temperature from the second temperature to calculate a delta temperature.

26. The method of claim 25, wherein the controller triggers an alarm in response to variations in delta temperature above a predetermined amount.

27. The method of claim 21, wherein the heat transfer apparatus comprises a heat transfer medium valve structured to control the amount of heat transfer medium which bypasses the heat transfer source and the amount of heat transfer medium which passes through the heat transfer source.

28. The method of claim 27, wherein a heat transfer medium valve sensor measures a position of the heat transfer medium valve.

29. The method of claim 28, wherein the position of the heat transfer medium valve is compared with a maximum allowable heat transfer medium valve position to control the catalyst feed rate, and wherein the heat transfer medium valve position is maintained in a range in which a change in valve position effects a change in heat transfer.

30. The method of claim 1, further comprising measuring a pressure differential made by a primary pump of a heat transfer apparatus.

31. The method of claim 30, further comprising controlling a secondary pump in response to the pressure differential made by the primary pump.

32. The method of claim 1, wherein the reactor comprises a fluidized bed.

33. The method of claim 1, wherein the reaction comprises polymerization.
34. The method of claim 1, wherein the reaction comprises polymerization of ethylene to form polyethylene.

35. A reaction process constraint override control system for optimizing production rates, the system comprising:
   reactor;
catalyst feeder that is structured to feed catalyst to the reactor;
heat transfer apparatus that is structured to transfer heat associated with the reactor;
heat transfer availability sensor that is structured to determine available heat transfer in the heat transfer apparatus; and
at least one controller that is structured to continuously control the catalyst feeder in response to available heat transfer and at least one additional process constraint.

36. The system of claim 35, wherein the additional process constraint comprises desired production rate of the reactor.

37. The system of claim 35, wherein the additional process constraint comprises amount of reaction product in a collector of the reactor.

38. The system of claim 35, wherein the additional process constraint comprises weight percent condensed in a recycle stream of the reactor.

39. The system of claim 35, wherein the additional process constraint comprises dew point of a recycle stream of the reactor.

40. The system of claim 35, wherein the heat transfer apparatus comprises a heat exchanger.

41. The system of claim 40, wherein the heat exchanger is structured to remove heat associated with the reactor.
42. The system of claim 40, wherein the heat exchanger is structured to add heat to the reactor.

43. The system of claim 35, wherein the heat transfer apparatus comprises a heat exchanger associated with the reactor, a heat transfer source, and a conduit which is structured to circulate heat transfer medium between the heat exchanger and the heat transfer source.

44. The system of claim 43, further comprising a recirculation conduit structured to allow at least a portion of the heat transfer medium to bypass the heat transfer source such that the at least a portion of the heat transfer medium is re-introduced into the heat exchanger without passing through the heat transfer source.

45. The system of claim 44, wherein the recirculation conduit is structured to, at a mix point, mix the at least a portion of the heat transfer medium with heat transfer medium which passes through the heat transfer source, prior to introduction into the heat exchanger.

46. The system of claim 45, wherein the heat transfer availability sensor comprises: a first temperature sensor which measures a first temperature of transfer medium at a point between the heat transfer source and the mix point; and a second temperature sensor which measures a second temperature of the heat transfer medium between the mix point and the heat exchanger.

47. The system of claim 46, wherein the at least one controller subtracts the first temperature from the second temperature to calculate a delta temperature.

48. The system of claim 47, wherein the delta temperature represents the heat transfer availability.
49. The system of claim 48, wherein the controller triggers an alarm in response to variations in delta temperature above a predetermined amount.

50. The system of claim 43, further comprising a heat transfer medium valve structured to control the amount of heat transfer medium which bypasses the heat transfer source and the amount of heat transfer medium which passes through the heat transfer source.

51. The system of claim 50, further comprising a heat transfer medium valve sensor structured to measure a position of the heat transfer medium valve.

52. The system of claim 51, further comprising an operator interface which is structured to receive a maximum allowable heat transfer medium valve position, and wherein the at least one controller compares the position of the heat transfer medium valve with the maximum allowable heat transfer medium valve position to control the catalyst feeder.

53. The system of claim 52, wherein the heat transfer medium valve sensor is connected to the at least one controller, and wherein the at least one controller is structured to control the catalyst feeder in response to the position of the heat transfer medium valve.

54. The system of claim 52, further comprising a heat transfer medium valve structured to control the amount of heat transfer medium which bypasses the heat transfer source and the amount of heat transfer medium which passes through the heat transfer source, further comprising a heat transfer medium valve sensor structured to measure a position of the heat transfer medium valve, wherein the heat transfer medium valve sensor is connected to the at least one controller, and wherein the at least one controller is structured to compare a
signal from the heat transfer medium valve sensor and the a delta temperature
to control the catalyst feeder.

55. The system of claim 35, further comprising an alarm which is structured to
be triggered in response to a predetermined amount of fluctuation in available
heat transfer.

56. The system of claim 35, further comprising a primary pump associated with
the heat transfer apparatus.

57. The system of claim 56, further comprising a secondary pump associated with
the heat transfer apparatus.

58. The system of claim 57, further comprising a pressure differential sensor
structured to detect a pressure differential made by the primary pump.

59. The system of claim 58, wherein the pressure differential sensor is connected
to a secondary pump controller which is structured to control the secondary
pump in response to the pressure differential made by the primary pump, and
wherein the secondary pump controller is the same as or distinct from the at
least one controller.

60. The system of claim 35, further comprising a recycle loop.

61. The system of claim 60, further comprising a dew point sensor structured to
measure a dew point of gas in the recycle loop.

62. The system of claim 61, wherein the dew point sensor is connected to the at
least one controller, and wherein the at least one controller is structured to
control the catalyst feeder in response to the dew point of gas in the recycle
loop.
63. The system of claim 62, further comprising an operator interface which is structured to receive a dew point approach, and wherein the at least one controller compares the dew point with the dew point approach to control the catalyst feeder.

64. The system of claim 60, further comprising a weight percent condensed sensor structured to measure a weight percent condensed of fluid in the recycle loop.

65. The system of claim 64, wherein the weight percent condensed sensor is connected to the at least one controller, and wherein the at least one controller is structured to control the catalyst feeder in response to the weight percent condensed in the recycle loop.

66. The system of claim 65, further comprising an operator interface which is structured to receive a maximum allowable weight percent condensed, and wherein the at least one controller compares the weight percent condensed with the maximum allowable weight percent condensed to control the catalyst feeder.

67. The system of claim 35, wherein the reactor comprises a collector structured to collect reaction product.

68. The system of claim 67, further comprising a reaction product sensor structured to measure an amount of reaction product in the collector.

69. The system of claim 68, wherein the reaction product sensor is connected to the at least one controller, and wherein the at least one controller is structured to control the catalyst feeder in response to the amount of reaction product in the collector.
70. The system of claim 69, further comprising an operator interface which is structured to receive a desired amount of reaction product in the collector, and wherein the at least one controller compares the amount of reaction product in the controller with the desired amount of reaction product in the controller to control the catalyst feeder.

71. The system of claim 35, further comprising a production rate sensor structured to measure a production rate of the reactor.

72. The system of claim 71, wherein the production rate sensor is connected to the at least one controller, and wherein the at least one controller is structured to control the catalyst feeder in response to the production rate.

73. The system of claim 72, further comprising an operator interface which is structured to receive a desired production rate, and wherein the at least one controller compares the production rate with the desired production rate to control the catalyst feeder.

74. The system of claim 35, wherein the reactor comprises a fluidized bed.

75. A reaction process constraint override control method for optimizing production rate of polyethylene homopolymers or copolymers in a gas phase fluidized bed reactor, the method comprising automatically and continuously controlling catalyst feed rate to a reactor to control a reaction in response to available heat transfer, current production rate of the reactor, amount of reaction product in a collector of the reactor, and weight percent condensed in a recycle stream of the reactor.
Fig. 10

500 Start

502 BIO1 AND BIO2 AND -BIO3

504 RO01 = RI01

506 Y

508 BIO4 AND BIO5 AND -BIO6

510 RO02 = RI02

512 N

508 RO01 = RI03

514 Y

516 BIO7 AND BIO8 AND -BIO9

518 RO03 = RI05

520 N

522 BIO10 AND BIO11 AND -BIO12

524 RO04 = RI07

526 N

526 Exit

RO02 = RI01 * II01 + RI02 * II02 + 10
(Add RPM setpoints of both feeders, 10 is a small number used for equality comparison)

Fig. 11

500 Start

602

604 RO02 - RI03

606 Y

612 Set BO01 False

608 N

610 Exit

612 Y

614 Set BO01 True

608 RO01 = RI03

610 Exit
<table>
<thead>
<tr>
<th></th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
</tr>
</thead>
<tbody>
<tr>
<td>FEEDER</td>
<td>ON</td>
<td>ON</td>
<td>ON</td>
<td>AUTO REMOTE</td>
</tr>
<tr>
<td>BLOCK 1</td>
<td>ON</td>
<td>ON</td>
<td>ON</td>
<td>AUTO REMOTE</td>
</tr>
<tr>
<td>BLOCK 2</td>
<td>ON</td>
<td>ON</td>
<td></td>
<td>AUTO REMOTE</td>
</tr>
<tr>
<td>AUTO STATUS</td>
<td></td>
<td></td>
<td></td>
<td>AUTO REMOTE</td>
</tr>
<tr>
<td>TOTAL RPM FROM RATE CONTROLLER</td>
<td>2010</td>
<td>2012</td>
<td>2173</td>
<td></td>
</tr>
<tr>
<td>CURRENT ACTUAL RPM FROM ALL FEEDERS</td>
<td>77%</td>
<td>69%</td>
<td></td>
<td></td>
</tr>
<tr>
<td>AVERAGE CW VALVE POSITION</td>
<td>3.6 C</td>
<td>3.5 C</td>
<td></td>
<td></td>
</tr>
<tr>
<td>CURRENT COOLER CWDT</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>AVERAGE COOLER CWDT</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>NO. OF LOCAL BLOCKS</td>
<td></td>
<td></td>
<td></td>
<td>4</td>
</tr>
<tr>
<td>NO. OF REMOTE BLOCKS</td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
</tbody>
</table>
Fig. 14

Production rates maximized to the cooling constraint by adjusting catalyst feed rates to minimize CWDT (target 1°C)

Production rate set point and actual values shows rates maximized to the cooling constraint.

Catalyst feeder is continuously adjusted (manipulated variable) to optimize control variables to target (CWDT = 1°C and rate = 32 TPH)

Cooling water delta temperature (CWDT) setpoint and actual illustrating operation running to maximum cooling (CWDT = 1°C)
Fig. 15

Illustrates the ability of the Production rate controller to optimize rates (initial constraint set by downstream limit - operator set; later rates run to cooling constraint).

- Production rate set point ramped up as downstream constraint is relieved.
- Target production rate maintained (downstream constraint).
- Excess cooling available.
- Catalyst feed ramped up with downstream constraint relieved to the limit of cooling.
- Cooling limit reached (CWDT = 0).

Legend:
- 00
- 0 0.1 0.2 0.3 0.4 0.5 0.6 0.7 0.8 0.9 1.0
- 4000 4500 5000 5500 6000 6500 7000 7500 8000 8500 9000 9500 10000
- -0.0 -0.1 -0.2 -0.3 -0.4 -0.5 -0.6 -0.7 -0.8 -0.9 -1.0
Production rate is a function of catalyst inventory in the reactor with CWDT mirroring both

Production rate fluctuations - match catalyst inventory

Catalyst Inventory in the reactor

Cooling water delta temperature (CWDT) - mirrors rate changes
A. CLASSIFICATION OF SUBJECT MATTER
   IPC(7): Please See Extra Sheet.
   US CL: Please See Extra Sheet.
   According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED
   Minimum documentation searched (classification system followed by classification symbols)
   U.S.: 486/55

   Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

   Electronic data base consulted during the international search (name of data base and, where practicable, search terms used)
   EAST, DERWENT

C. DOCUMENTS CONSIDERED TO BE RELEVANT

<table>
<thead>
<tr>
<th>Category</th>
<th>Citation of document, with indication, where appropriate, of the relevant passages</th>
<th>Relevant to claim No.</th>
</tr>
</thead>
<tbody>
<tr>
<td>X</td>
<td>US 3,790,036 A (MILLER) 05 FEBRUARY 1974, see entire reference.</td>
<td>1-75</td>
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<td>X</td>
<td>US 4,668,473 A (AGARWAL) 26 MAY 1987. see entire reference.</td>
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<td>A</td>
<td>US 6,063,877 A (KOCIAN et al.) 16 MAY 2000.</td>
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<td>US 4,162,894 A (TRIBBLE) 31 JULY 1979. see entire reference.</td>
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<td>US 6,183,698 B1 (VASSILIOU et al.) 06 FEBRUARY 2001. see entire reference.</td>
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</table>

X Further documents are listed in the continuation of Box C.  □ See patent family annex.

* Special categories of cited documents:
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  *O* document referring to an oral disclosure, use, exhibition or other means
  *P* document published prior to the international filing date but later than the priority date claimed

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"X" document of particular relevance; the claimed invention cannot be considered novel or cannot be considered to involve an inventive step when the document is taken alone

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Date of the actual completion of the international search: 22 MAY 2002

Date of mailing of the international search report: 01 JUL 2002

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Commissioner of Patents and Trademarks
Box PCT
Washington, D.C. 20231
Facsimile No. (703) 805-3920

Authorized officer
BRIAN SINES

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<tr>
<td>X</td>
<td>US 4,620,049 A (SCHMIDT et al.) 28 OCTOBER 1986, see entire reference.</td>
<td>1-75</td>
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A. CLASSIFICATION OF SUBJECT MATTER:
IPC (7):
G01N 37/00

A. CLASSIFICATION OF SUBJECT MATTER:
US CL:
436/55