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Simple Control Structure for a Compression Purification Process in an Oxy-Combustion Power Plant

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The composition of the stack gas from an oxy-combustion power plant is about 75 mol % carbon dioxide. The stream must be purified to about 95 mol % by removing light inerts such as nitrogen, oxygen, and argon. The product stream must be compressed to 110 bar for sequestration. This article presents a simple control structure for the double-flash compression and purification process to achieve these objectives. *AIChE J*, 61: 1581–1588, 2015
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Introduction

The use of oxygen instead of air in power plant has the advantage of producing a stack gas that has a much higher carbon dioxide concentration because there is very little nitrogen fed to the combustion zone. It is, therefore, much easier to concentrate the stack gas to the required CO_2 purity for sequestration. Many steady-state design studies have been reported. Much less investigation of the dynamics and control of the process has occurred. Several papers[1–3] have discussed dynamic models. However, the development and testing of detailed control structures have been rarely studied.

An interesting recent paper by Jin et al.[4] discussed several alternative processes and studied in detail the partial condensation method with two flash separators. An optimum steady-state economic design is developed and the dynamic control of the process is explored. The control structure proposed by Jin et al.[4] is quite complex with the online measurements of three process streams required to control carbon dioxide recovery and product purity. The existence of a minimum temperature limitation (the freezing point of $CO_2 = -56.6°C$ as shown in Figure 1) imposes a hard constraint on the cryogenic temperature permitted. Jin et al.[4] achieve this by sacrificing drum level control, which could lead to the flash drum running dry or overfilling.

This article proposes a more simple control structure that holds the process at or above the control specifications. Only conventional temperature, level, flow, and pressure controller are required. The process is simulated using Aspen Plus and Aspen Dynamics. Peng–Robinson physical properties are used.

The computer model used in Aspen was polytropic using ASTM method with an efficiency of 80%. The heuristic of keeping the compression ratios the same in all stages is used. The issue of compressor surge control is assumed to be handled by conventional antisurge control systems (prevent flow dropping below the surge limit by spill back).

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Process Studied

Jin et al.[4] studied the process proposed by the International Energy Agency Greenhouse Gas R&D program[5] with a stack gas flow rate of 717,190 kg/h. In this article, the conditions shown in PFD 11 of the report[5] are used with a stack gas flow rate of 596,069 kmol/h (14,616 kmol/h) as shown in Figure 2.

The stack gas at 1.1 bar is compressed to 30 bar in a 3-stage compressor train with intermediate cooling down to 50°C using cooling water. Three stages are required to keep compressor discharge temperatures below the 200°C maximum temperature limitation recommended by Walas.[6] Total compressor work in the feed compressors is 51.7 MW.

The feed is cooled to cryogenic temperature in three multistream heat exchangers using the cold product streams. Pressure drops in the heat exchangers are assumed to be 0.1 bar for all streams. The feed temperature leaving the first heat exchanger is 20.3°C and leaving the second is −22.5°C. At this temperature and 29.6 bar, some liquid is formed (4440 kmol/h) in the first flash drum. Liquid from the bottom of the drum is flashed through a control valve to 19.6 bar, which drops its temperature to −29.2°C, and is fed back upstream through the second and first heat exchangers. It enters the second product compressor at 19.4 bar and 42°C.

The vapor stream from the first flash drum is fed to the third heat exchanger and is cooled to −52.8°C. At this temperature and 29.5 bar, some liquid is formed (6057 kmol/h) in the second flash drum. Liquid from the bottom of the drum is fed back to the third heat exchanger and warmed to −44.8°C. It is then flashed through a control valve to 10.2 bar, which drops its temperature to −55°C. This is the coldest temperature in the process and it must be greater than the freezing temperature of carbon dioxide (−56.6°C).

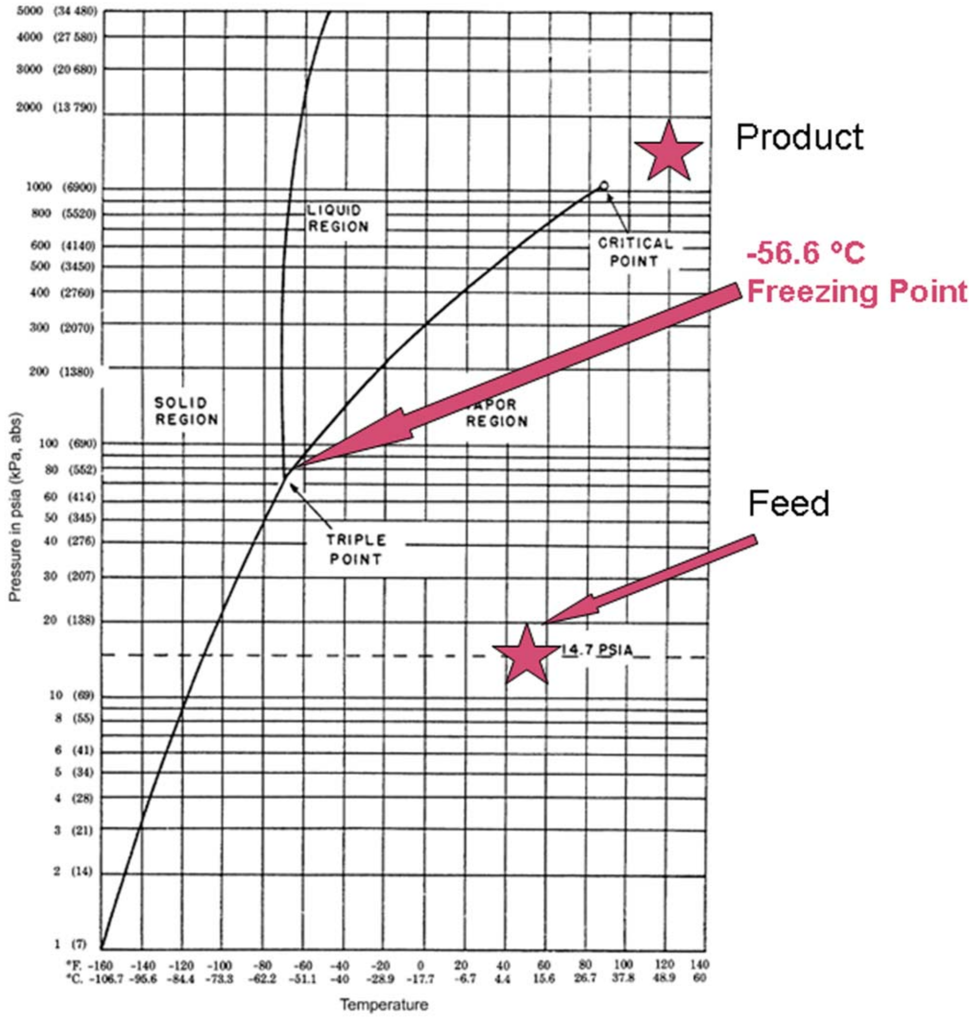

Figure 1. Carbon dioxide phase diagram.

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

The stream flows upstream through the three heat exchangers and enters the first product compressor at 38.2°C and 9.75 bar, which is the lowest pressure in the process unit.

The gas stream from the second flash drum, which is the vent stream containing the inerts to be removed, flows upstream through the three heat exchangers. The flow rate is 4120 kmol/h with a composition of 26.03 mol % CO_2. Since it is at high pressure (29.2 bar) it can be used to generate power in an expander turbine. To increase the power recovered, the stream is heated to 301°C by hot process streams. Expansion to 1.1 bar produces 8.673 MW.

The final compression of the product gas is achieved in two compressors with interstage cooling. The total work in the three product compressors is 20.47 MW. The fluid product leaves at 110 bar and 50°C, which puts it above the critical pressure.

The purity of the 10,496 kmol/h of product is 95.17 mol % CO_2. The recovery of carbon dioxide is 90.3%.

Dynamic Models

The steady-state simulation in Aspen Plus is exported into Aspen Dynamics after the sizes of the process units with dynamic characteristics have been determined. The two flash drum obviously must be considered. In addition, the three plate heat exchangers are very large and have significant volume.

Flash drum sizing

There are two criteria for determining the size of a flash drum. The first is to provide 5 min of liquid holdup when half full. The second is to have a cross-section area sufficiently large to achieve a vapor velocity that prevents liquid entrainment. A design heuristic is to have an *F*-Factor of 0.6

$$F\mathrm{-Factor}=0.6=V_{\max}\,(\mathrm{m/s})\sqrt{\rho_V}(\mathrm{kg/m^3})$$

In the first drum, the vapor density is 68.8 kg/m^3 and the vapor volumetric flow rate is 5184 m^3/h, which leads to a diameter of 5 m. The five-minute liquid holdup time criterion leads to a diameter of 3 m (using an aspect ratio $L/D=2$). So the vertical vessel is sized with a 5 m diameter and a length of 10 m having a total volume of 196 m^3.

In the second drum, the vapor density is 62.3 kg/m^3 and the vapor volumetric flow rate is 2155 m^3/h, which leads to a diameter of 3 m. The five-minute liquid holdup time criterion also leads to a diameter of 3 m (using an aspect ratio $L/D=2$). So the vertical vessel is sized with a 3 m diameter and a length of 6 m having a total volume of 42 m^3.

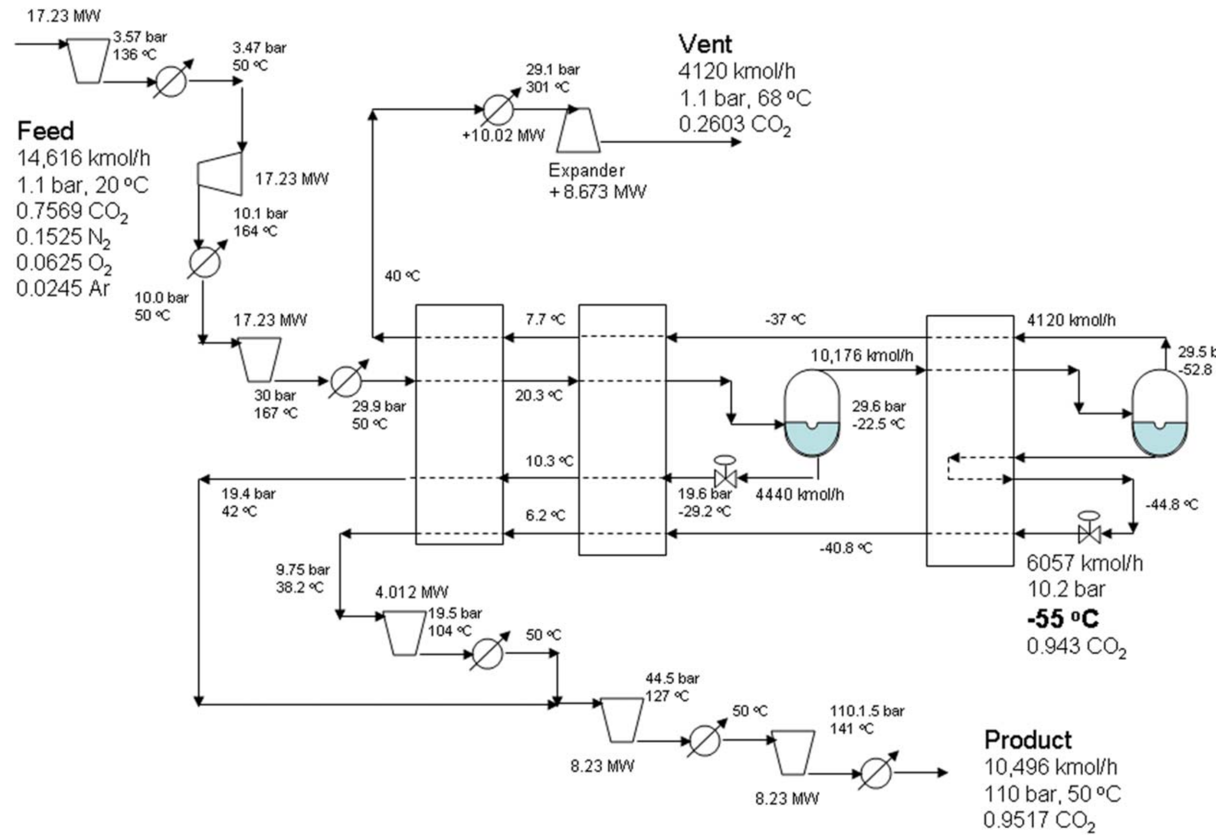

Figure 2. Compression purification process.

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

Heat exchanger sizing

Plate heat exchangers are used in this clean service. Most of the heat transfer occurs between gas phases, so a small overall heat-transfer coefficient $U = 0.17$ kW m^{-2} K^{-1} is assumed as suggested by Turton et al.[7]

In the first heat exchanger, there is one hot and three cool streams flowing countercurrently. The total heat duty is 5.108 MW, which is transferred to the warm feed stream. The heat duties in the three cold streams are 1.203 MW to the vent stream, 2.097 MW to the stream originating from the liquid leaving the first drum and 1.809 MW to the stream originating from the liquid leaving the second drum. The total heat-transfer area is 2721 m^2. Using a conservative estimate of 78 m^2/m^3 for a plate heat exchanger gives a total volume of 35 m^3, which is prorated to each stream based on its fraction of the total heat duty.

The second and third heat exchangers have duties of 25.91 and 21.4 MW, total heat-transfer areas of 32,320 and 22,000 m^2 and volumes of 414 and 282 m^3, respectively.

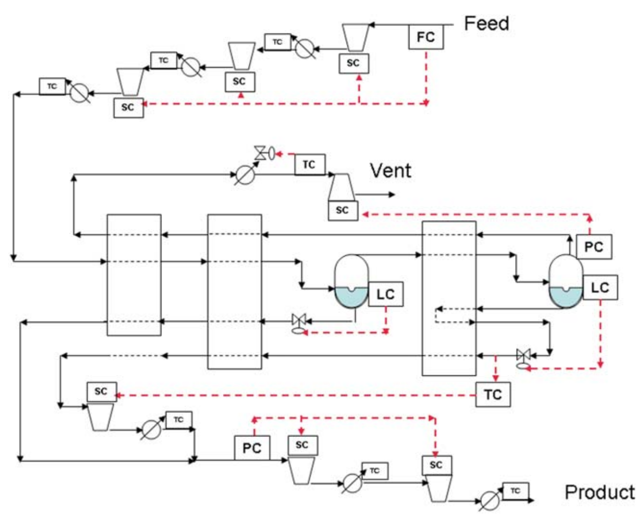

Figure 3. CPU control structure.

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

Control Structure

There are six fundamental control objectives of the compression purification process:

1. The feed flow rate is set by the upstream power plant, so the unit must be able to handle feed flow rate disturbances.
2. The minimum temperature in the process, which occurs when the liquid from the second drum is flashed through a valve, must be higher than the freezing point of carbon dioxide (−56.6°C).
3. The liquid levels in the two drums must be maintained between high and low limits.
4. The carbon dioxide composition of the product gas must be greater than 95 mol % CO_2.
5. The recovery of CO_2 in the product stream must be held near 90%.
6. The product gas must be delivered at 110 bar.

It should be noted that recovery and purity are not explicitly controlled by the proposed control structure. However, results presented below demonstrate that these variables

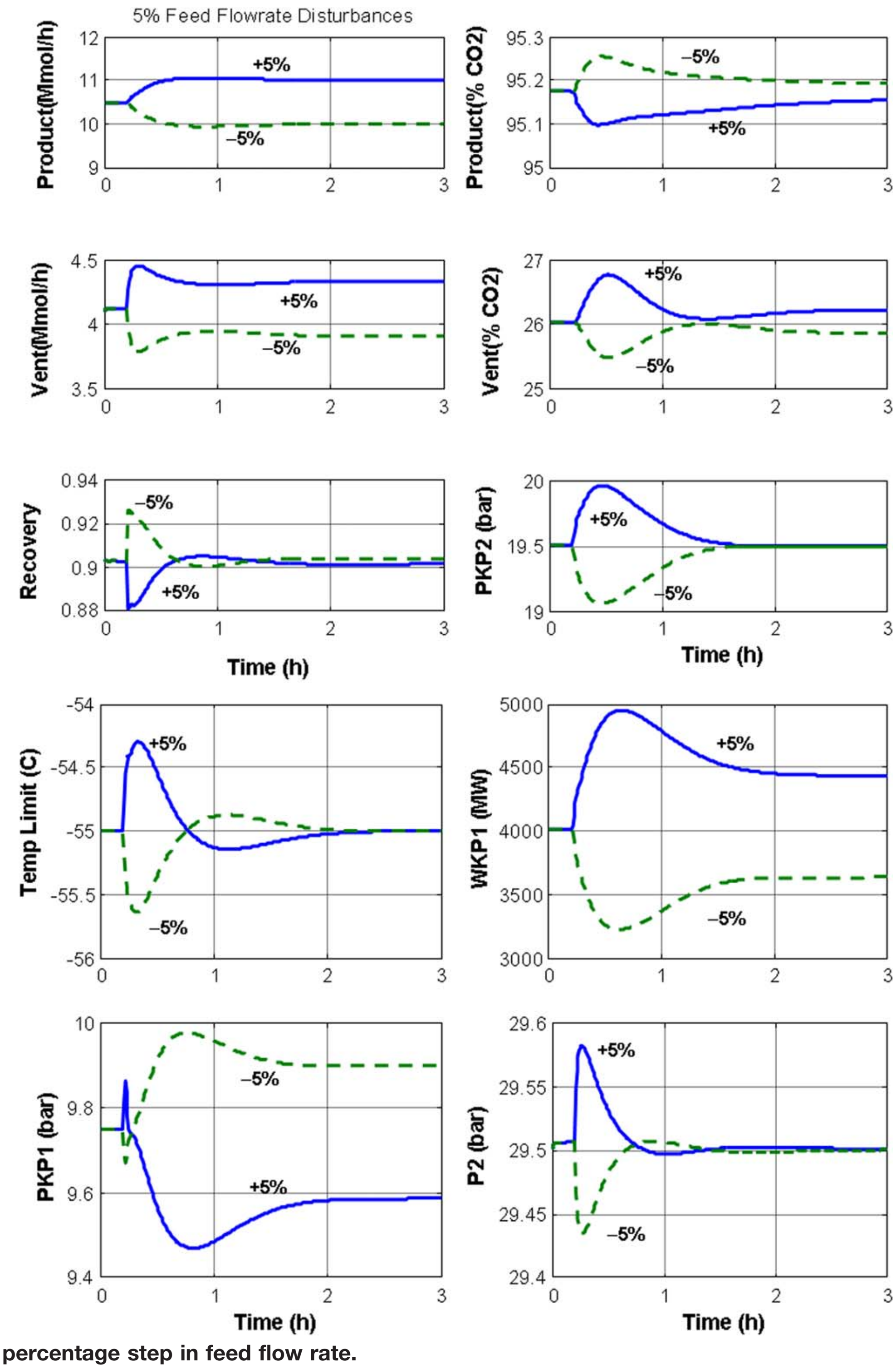

Figure 4. Five percentage step in feed flow rate.

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change only very slightly in the face of large disturbances. The following list defines all of the control loops used to achieve these objectives. The proposed control structure is shown in Figure 3. Conventional PI controllers are used and no composition measurements are required.

1. Fresh feed is flow controlled by manipulating the speed of the three feed compressors. Step and ramp changes in the setpoint of the feed flow controller serve as disturbances to test the dynamic effectiveness of the control structure.

2. The temperatures of the gas streams leaving all of the water-cooled heat exchangers that cool the compressor discharge gas down to 50°C are controlled by manipulating cooling water.

3. The temperature at the exit of the valve on the liquid leaving the second drum is controlled by manipulating the speed of the first product compressor. A one-minute temperature-measurement deadtime is assumed, and the controller is tuned using relay-feedback testing and Tyreus–Luyben tuning

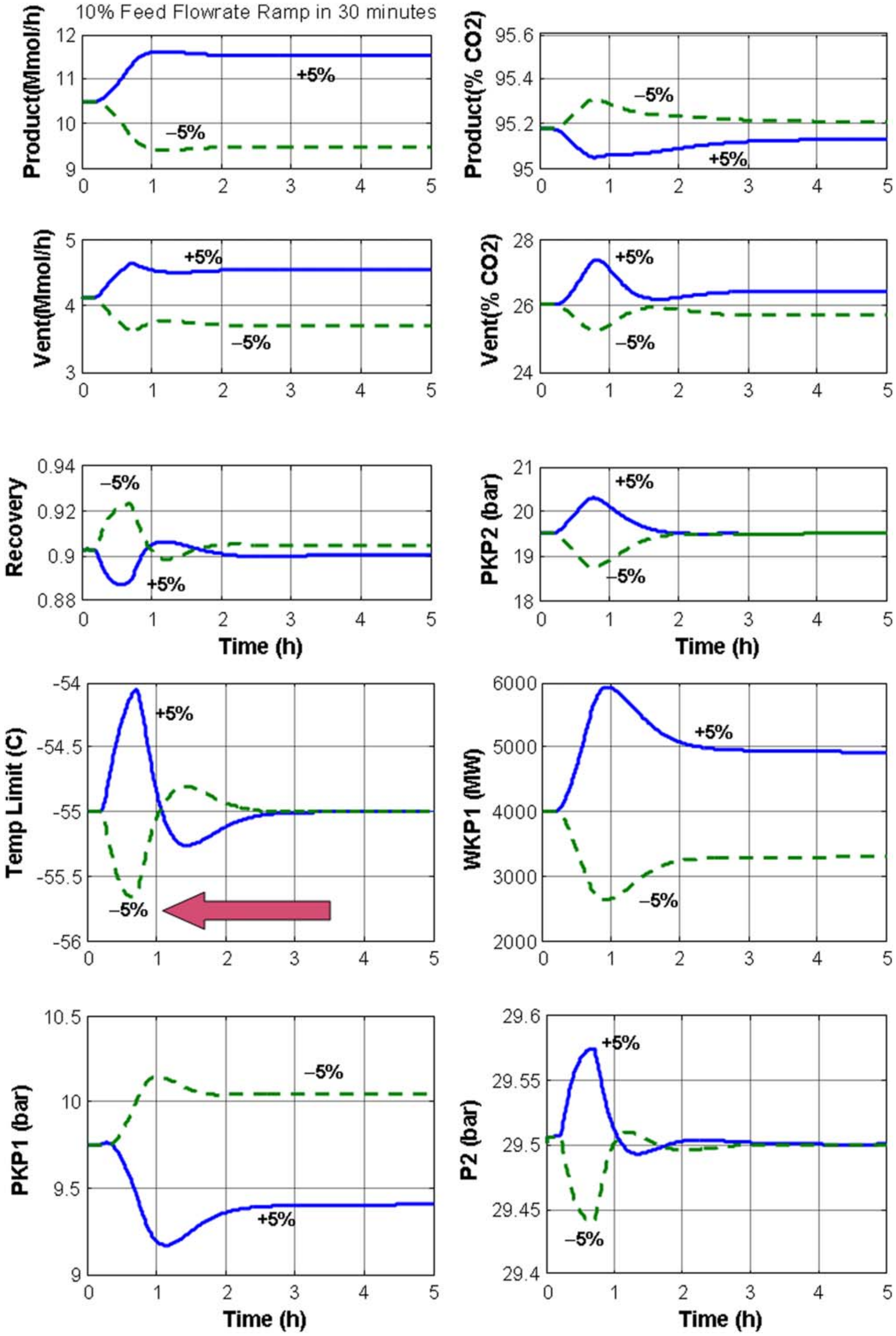

Figure 5. Thirty-minute ramp to 10% feed flow rate.

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rules. Using a 100°C temperature transmitter span and a controller output span of 13.45 MW, the resulting tuning constants are $K_C = 2.7$ and $\tau_I = 5.3$ min.

4. The liquid levels in the two drums are controlled by manipulating the valves in the liquid lines. Proportional controllers are used with $K_C = 2$.

5. The pressure in the second drum is controlled by manipulating the speed of the expander on the vent stream. Controller constants are $K_C = 1$ and $\tau_I = 20$ min.

6. The product gas discharges into a pipeline that is at a pressure of 110 bar. The suction pressure of the second product compressor is controlled at 19.4 bar by manipulating the speed of the two final product compressors. Controller constants are $K_C = 1$ and $\tau_I = 20$ min. This loop keeps the pressure downstream of the liquid valve on the first drum at 19.6 bar, which affects the temperature downstream of the valve. This influences the cooling in the second heat exchanger and the temperature in the first drum.

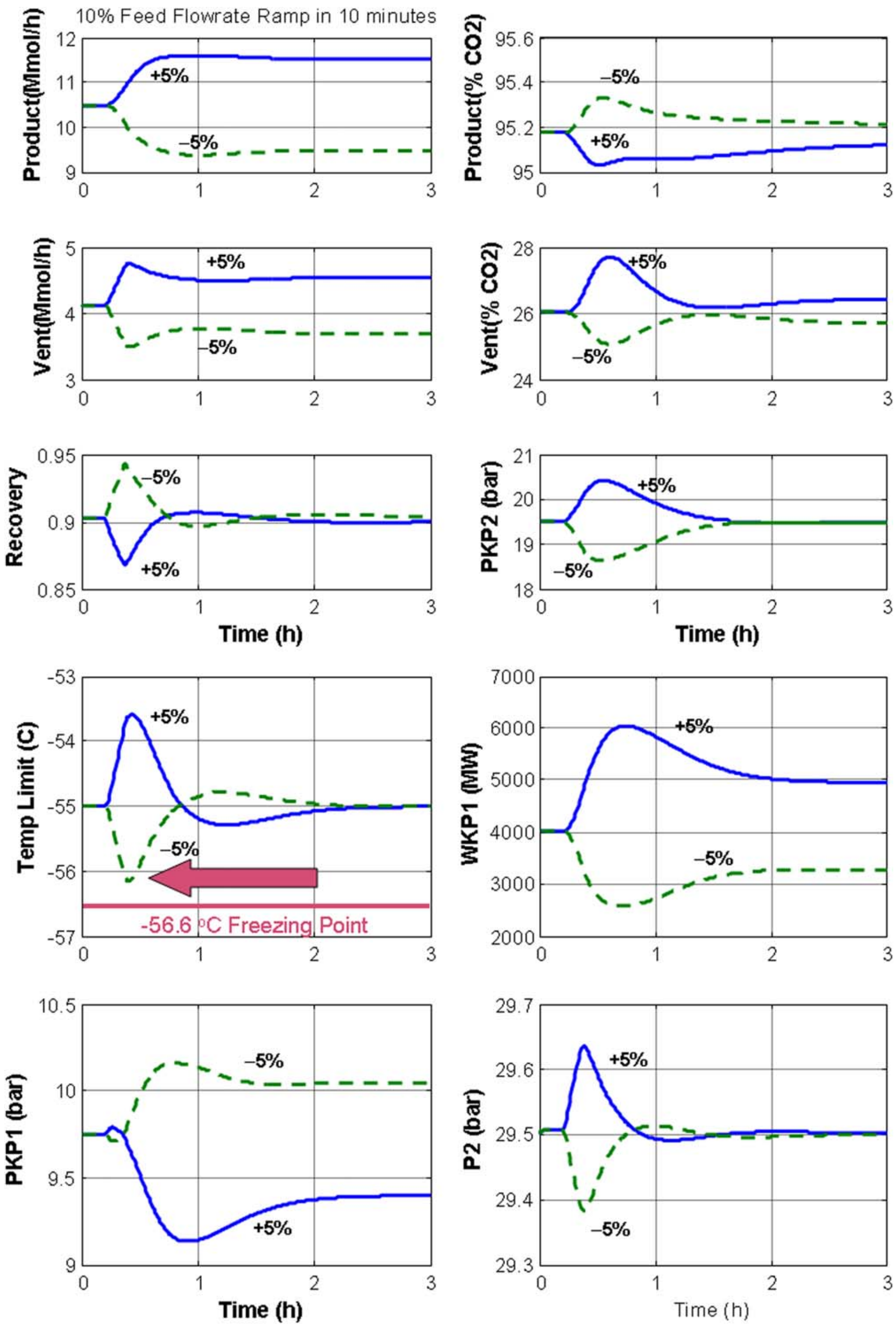

Figure 6. Ten-minute ramp to 10% feed flow rate.

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7. The temperature of the vent stream leaving the heater is controlled by manipulating the flow rate of a hot process stream.

Controlling the temperature of the stream downstream of the expansion valve on the liquid from the second drum is critical because freezing of the CO_2 must be avoided. Figures 4–7 show the dynamic responses of the process for changes in throughput and feed composition.

The dynamic simulation is quite sensitive to disturbances. Only small changes could be made without producing integration errors. The Gear integration algorithm was found to be more robust than the default Implicit Euler. The Hybrid method was used in the nonlinear solver.

Dynamic Results

Figure 4 shows how the process responses to step changes of 5% in the setpoint of the feed flow controller. Solid lines are for 5% increases and dashed lines are for 5% decreases in throughput. Stable closed-loop regulatory control is achieved. Product and vent flows increase or decrease as expected. Pressure in the second drum (P2) and suction

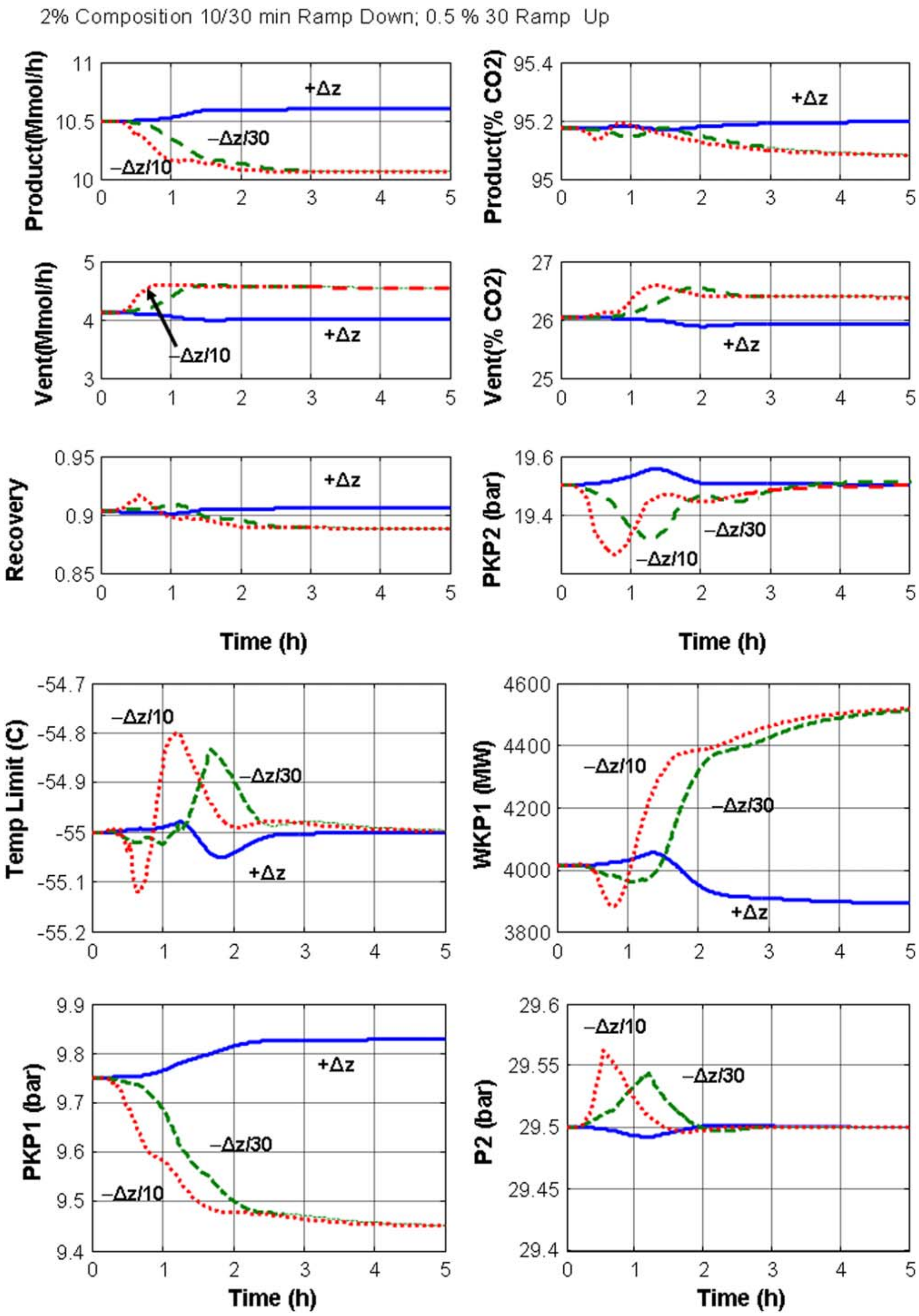

Figure 7. Ramp changes in feed composition.

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

pressure of the second product compressor (PKP2) are returned to their desired values in about 1.5 h. Both product purity and recovery stay very close to their desired values.

For an increase in feed flow rate, the important temperature downstream from the valve on the liquid line from the second drum (TempLimit) increases so the temperature controller increases the work to the first product compressor (WKP1), which lowers the suction pressure in this compressor and brings the temperature back to the setpoint. For a decrease in feed flow rate, the variables move in the opposite direction, with the TempLimit initially dipping closer to the $-56.6^\circ C$ limit by about $0.6^\circ C$.

The simulation would not run for any larger step disturbance, so ramp disturbances were made. Figure 5 gives results for 10% changes in feed flow rate that are achieved by ramping the setpoint of the feed flow controller up or down for 30 min. Product purity, recovery, and the "TempLimit" are well controlled. The important temperature drops to about $1^\circ C$ above the limit for the ramp down.

Figure 6 show the effect of shortening the ramp time from 30 to 10 min. The transient drop in "TempLimit" becomes larger and the temperature gets quite close to the limit. However, the system recovers and pressures, product purity and recovery return to values very close to their specifications.

Finally, feed composition disturbances were imposed on the process. The system is quite sensitive to increases in the CO_2 composition of the feed. Only small changes could be successfully run without leading to integration and nonlinear solver errors. The solid lines in Figure 7 show the response of the process when the CO_2 feed composition is increased from 75.69 to 76.19 mol % CO_2 over a 30-minute ramp with the nitrogen feed composition decreased from 15.25 to 14.75 mol % N_2. Note that this is a 3.3% relative change in N_2 composition.

The simulation could handle larger decreases in CO_2 feed composition. The dashed lines in Figure 7 are when the CO_2 is decreased from 75.69 to 73.69 mol % CO_2 over a 30-minute ramp with a corresponding 2 mol % increase in N_2. The dotted lines are when the CO_2 is decreased from 75.69 to 73.69 mol % CO_2 over a 10-minute ramp.

As expected, the lower CO_2 feed composition leads to smaller product flow rates and larger vent flow rates. Product quality remains above 90 mol %, and recovery drops slightly to about 89%. The faster the ramp, the more quickly the process reaches the new steady-state conditions. The important "TempLimit" remains well above the low limit for all disturbances.

Conclusion

A simple control structure is proposed and tested for a compression purification process to increase CO_2 purity of the combustion gas from an oxy-fired power plant. No explicit control of product composition and CO_2 recovery are required since they change very little when only simple pressure, temperature, and level control loops are used. Violating the CO_2 freezing temperature limitation is avoided by manipulating the speed of the first product compressor. Feed flow rate is controlled by manipulating the speed of the three feed compressors. Drum liquid levels are controlled by their downstream control valves.

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