Process Control for the Process Industries

PART 1: DYNAMIC CHARACTERISTICS

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Several aspects of chemical processing make control and automation in the process industries different from other types of machine and equipment control. This article focuses on the factors related to process dynamics.

Control theory and practice are generally considered in the realm of electrical engineering. Many of the advances in control theory, such as the frequency response concepts (developed by Nyquist, Bode, and others, principally at Bell Laboratories) and state space and optimal control (developed mainly with funding from the National Aeronautics and Space Administration [NASA]), were made by electrical engineers, And, control theory remains part of the undergraduate and graduate electrical engineering curricula.

However, the technologies and techniques developed and studied by electrical engineers do not provide effective solutions to the control problems encountered in the process industries. Five aspects of chemical processing contribute to the unique nature of process control:

- *disturbances.* Every loop in a process plant must contend with these. Some are measurable; many are not.
- *transportation lag, or dead time.* Material flowing through a 100-m-long pipe at 2 m/sec has a transportation lag of 50 sec. This type of behavior is detrimental to the performance of the controls.
- *process dynamics.* Many processes are commissioned without being analyzed by dynamic modeling.
- *economics.* To be economically beneficial, control improvements must enable the process to operate more efficiently. The approach differs between continuous and batch processes, but the common denominator is the need for a high degree of automatic control.
- *multivariable nature of industrial processes.* For single-loop control configurations relying on proportional-integral-derivative (PID) control logic *(1–3)*, a correct pairing of the controlled and manipulated variables is required. In important applications such as control of distillation columns, the interaction between the loops must be addressed as well.

This two-part article explains these factors and why they are relevant to the practice of process control. Part 1 covers the first three — disturbances, transportation lag, and process dynamics — which pertain to the dynamic behavior of the process. But contrary to common belief, the practice of process control involves more than process dynamics. The last two factors relate largely to the steady-state behavior of the process and are covered in Part 2, which is scheduled to appear in the April issue. (Part 2 can be viewed online with Part 1 at www.aiche.org/cep).

Disturbances

The process depicted in Figure 1 produces hot water by injecting steam into the tank. The primary purpose of the controls for this simple process is to respond to disturbances such as changes in the hot water demand; changes in the setpoint for the temperature controller are rare. The control configuration consists of a measurement device for the hot water temperature (TT), a temperature controller (TC), and a control valve on the steam supply. The controller compares the measured value of the hot water temperature to its target or setpoint; the difference is the control error. The controller adjusts the control valve opening to drive the control error to zero, thereby driving the hot water temperature to its setpoint.

By driving the control error to zero, the hot water temperature controller can respond to changes in two distinct inputs to the loop:

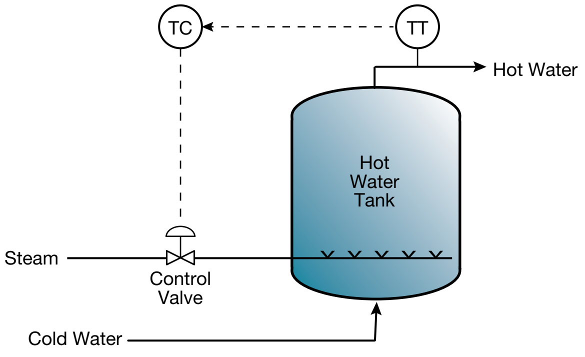
- *setpoint.* The value of the setpoint is the desired hot water temperature. Upon a change in the setpoint, the controller changes the control valve opening so as to drive the hot water temperature to the new value of the setpoint.
- *hot water demand.* The demand or flow of hot water is determined by the amount of hot water being consumed. Upon an increase in the flow, the hot water temperature initially drops, introducing a control error. To return the control error to zero, the controller must increase the control valve opening to return the hot water temperature to its setpoint.

Sometimes called the load, the hot water demand is the process throughput and is a member of a class of inputs known as disturbances. Other disturbances include the cold water inlet temperature, the steam pressure, and the steam enthalpy. All industrial processes are subjected to a variety of disturbances; some disturbances can be measured, but for others, measurement is either impossible or impractical.

The temperature controller in Figure 1 is expected to expeditiously implement each change in the setpoint, as well as respond to disturbances such as changes in the hot water demand. This expectation applies to most loops in process plants. Thus, the focus here is on loops for which an expeditious response to any change is desired.

Exceptions exist — for example, the level loop in a surge vessel, whose purpose is to smooth changes in the input flows while maintaining a discharge flow that changes as little and as slowly as possible. The surge vessel level controller does not respond to each individual change in input flow, but instead makes small adjustments to the discharge flow in response to changes in the long-term mean of the input flows. Such loops are the exceptions and beyond the scope of this article.

For utilities, such as steam boilers, compressed air, or cooling water systems, the major disturbances are usually throughput changes. Changes in the setpoint are rare. The hot water process in Figure 1 is a utility process. The consumers expect the temperature of the hot water to be close to a certain value. Likewise, the consumers of steam expect the steam pressure to be close to a certain value. The expected values change infrequently, making changes in setpoints infrequent and small. The primary purpose of the controller is to respond to changes in the demand for the utility.



▲ **Figure 1.** Steam is injected into a tank to produce hot water.

Continuous processes are operated at reasonably constant throughputs, but otherwise are similar to utility processes. After startup, setpoint changes are small and infrequent. The primary purpose of the controls is to respond to the myriad disturbances — most of which are minor, but occasionally a major one occurs. In batch processes, setpoint changes are common, but the controls must also contend with disturbances.

The PID controllers commonly applied in the process industries attempt to drive the control error to zero. PID controllers respond to both setpoint changes and disturbances, and the manner in which they respond to a setpoint change and to a disturbance is largely the same. However, the desired behavior of the controller for a setpoint change is not necessarily the same as the desired behavior for a disturbance.

Consider how the hot water temperature approaches its setpoint after a change in the hot water temperature setpoint. Figure 2a illustrates the response of a controller tuned to two different performance objectives:

- Response A has a quarter decay ratio. The first peak's overshoot is 2.0°C, and the second peak's overshoot is 0.5°C — the overshoot for the second peak is one-quarter the overshoot of the first peak. Such behavior is sometimes referred to as "big-hump/little-hump" behavior: A significant first overshoot is followed by a small second overshoot, but thereafter the oscillation is inconsequential.
- Response B exhibits minimal overshoot. This behavior is sometimes called critically damped.

The controller tuned for a quarter decay ratio more aggressively implements the setpoint change than the controller tuned for minimal overshoot. The more rapid initial rise to the setpoint is followed by noticeable overshoot and oscillations. Most engineers working in production perceive no benefit from overshoot and oscillations. With the possible exception of applications such as material blending, control errors of one sign do not offset control errors of the opposite sign.

Figure 2b presents the corresponding responses to an increase in the hot water demand. The hot water temperature initially decreases, causing the controller to increase the control valve opening to restore the temperature to its setpoint. In this case:

- Response A clearly cycles. The amplitude is small and the cycle is not centered about the setpoint, making it impossible to accurately compute a decay ratio. However, the first peak in the cycle is followed by a small second peak with little cycling evident thereafter. This is consistent with big-hump/little-hump behavior, or quarter decay ratio.

- Response B does not cycle. After the initial drop in temperature, the response slowly returns to the setpoint. This behavior is consistent with the minimal overshoot criterion.

These responses illustrate that a controller tuned to exhibit a particular behavior for a change in the setpoint exhibits that same type of behavior for a disturbance.

Hot water is a utility. The hot water temperature must be above an agreed-upon temperature at all times. Analogous statements can be made for steam pressure, compressed air pressure, cooling water supply temperature, and the like. However, the controller cannot be operated with the agreed-upon value as its setpoint, because any increase in hot water demand would cause the temperature to drop below the agreed-upon value.

A conservative value must be specified for the target. How conservative depends on two factors: the nature of the day-to-day disturbances to which the process is subjected, which can only be assessed from operational experience; and how effectively the controls respond to the disturbances. The responses in Figure 2b are to the same disturbance; all differences pertain to the performance of the controls.

Specifying a conservative value for the target usually incurs costs. For instance, heating water to a temperature above the agreed-upon value consumes additional steam.

For a given occurrence of a disturbance, the fundamental question is simple: Did the hot water temperature drop below its agreed-upon value, and if so, by how much? This suggests that only one point in the response to a disturbance is significant, specifically, the maximum departure from the setpoint. For the responses in Figure 2b:

Response	Performance Objective	Max Departure
A	Quarter Decay Ratio	1.1°C
B	Minimal Overshoot	2.2°C

A smaller maximum departure from the setpoint indicates that the controller is more effectively responding to the disturbance. The controller tuned to the quarter decay ratio performance objective provides a more effective response to a change in the hot water demand.

To be clear, the controller is delivering consistent performance for both setpoint changes and disturbance changes. A slightly shorter reset time (maybe by about 25%) would cause the response to approach the setpoint more rapidly, but with little effect on the maximum departure from setpoint. Since departure from setpoint is the important aspect, little incentive exists to shorten the reset time.

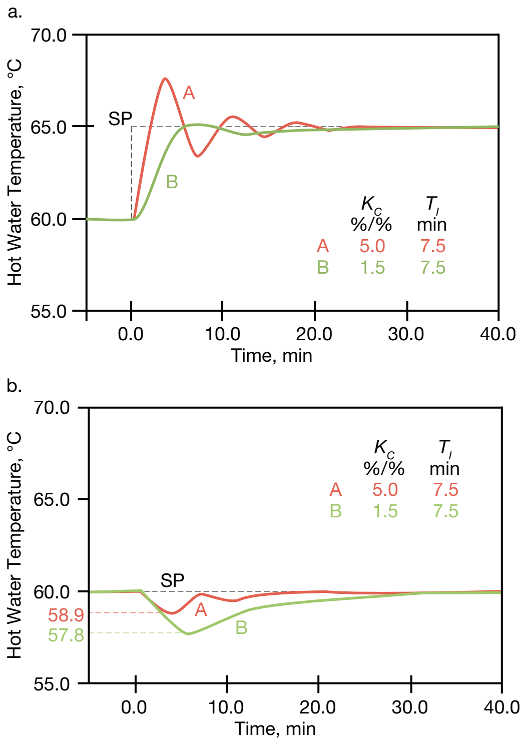
From the perspective of how control loop behavior affects process performance, a disturbance warrants a more aggressive response from the controller:

- *response to setpoint change (Figure 2a).* When tuned to quarter decay ratio, setpoint responses exhibit significant overshoot (40% for Response A in Figure 2a) followed by noticeable oscillations. Production personnel are more comfortable with the performance of controllers tuned for minimal overshoot.
- *response to disturbance (Figure 2b).* When tuned to quarter decay ratio, the oscillations are small and not objectionable. The small oscillation that rapidly decays after the initial peak is not a concern. The key performance metric is the maximum departure from setpoint, which is smaller for the controller tuned to a quarter decay ratio than one tuned for minimal overshoot.

A controller tuned to a quarter decay ratio is more aggressive in both cases. The consequences of aggressive controller behavior are undesirable for setpoint changes, but are beneficial for disturbance changes.

How does one assess the performance of the hot water temperature loop in Figure 1? With few exceptions, by making a small setpoint change and observing the response. How would one manually tune this loop? By making alternate increases and decreases in the setpoint, observing how changes in the tuning parameters affect the response, and adjusting the controller accordingly.

But in the plant, changes in the hot water temperature setpoint are infrequent and small. So, does it make sense



▲ **Figure 2.** (a) In response to a setpoint increase of 5°C, Controller A, which is tuned for a quarter decay ratio, experiences more overshoot than Controller B. (b) Similarly, when hot water demand increases from 230 kg/min to 280 kg/min, Controller A experiences cycling, whereas Controller B does not.

to tune based on setpoint changes? Loop performance for a disturbance is consistent with the performance for a setpoint change; in this regard, the answer is definitely yes, it does make sense to tune based on setpoint changes.

Overshoot and cycling are undesirable in setpoint responses, so conservative tuning is often viewed as desirable. But for responses to disturbances, conservative tuning produces larger maximum departures from setpoint, which is not desirable.

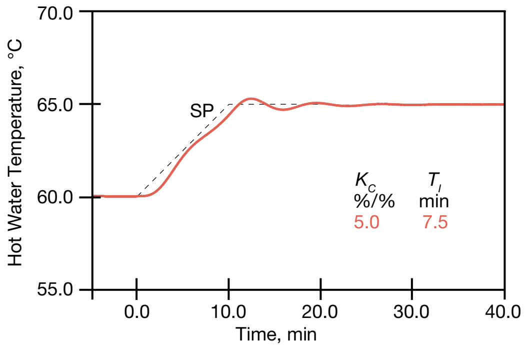
The basic issue is how aggressively the controller should respond. Aggressive response is not essential for setpoint changes, but is essential for disturbances.

How can one have aggressive response to a disturbance while avoiding noticeable overshoot and cycling in the response to a setpoint change? Digital controls offer two possibilities:

- *proportional based on the process variable (PV).* Most digital implementations of the PID equation provide the option to base the proportional mode on the measured variable instead of the control error. Selecting this option makes the controller implement setpoint changes more slowly, but has no effect on the response to disturbances.
- *ramp the setpoint.* Process operators are sometimes instructed to implement setpoint changes in small increments. A 10°C change in the setpoint could be implemented by making 2°C changes at some interval of time. This is a crude form of ramping the setpoint; digital systems provide a smooth ramp that frees the operator to attend to other duties — the operator simply changes the setpoint by 10°C, and all ramping is done “under the hood.”

Most find setpoint ramping easier to understand. Setpoint ramping only affects the response to a setpoint change; it has no effect on the response to a disturbance.

For a controller tuned to a quarter decay ratio, Figure 3 illustrates the results of implementing a 5°C change in the hot water temperature setpoint by ramping over a period of 10 min (the ramp rate is 0.5°C/min). The controller tuned for minimal overshoot would not benefit from setpoint ramping.



▲ **Figure 3.** A 5°C change in setpoint of a controller tuned for a quarter decay ratio is accomplished by ramping at 0.5°C/min.

Transportation lags

Process facilities require material transport — a belt conveyor is a material transport system for solids, a pipe is a material transport system for fluids. Each involves two parameters:

- distance over which the material is transported
- velocity at which the material is transported.

The transport time is simply distance divided by velocity. In process terminology, the transport time is the transportation lag.

Transportation lags, also called dead times, were rarely, if ever, discussed in traditional control courses offered by electrical engineering departments. Electrons move at the speed of light, so the transport time is insignificant, unless:

- distance is large — when bouncing radio waves off the moon, the distance is large enough for the transport time to be noticeable
- time frame is short — in the design of high-speed digital circuits, where the time frame is measured in picoseconds, the transport time cannot be ignored, even though the distances are short.

In process facilities, materials move at much slower rates. On the fast end are tissue paper machines in which the sheet moves at speeds approaching 2,000 m/min. Most materials within a process facility move far more slowly.

Processes differ from electronics in another very significant manner. The speed of light is a universal constant. Distances from earth to moon and distances between components of a circuit can be accurately determined, and either do not change or change in a manner that can be predicted or measured. The values of these transportation lags can be accurately computed.

This is occasionally true for process systems. The speed of a paper machine is accurately measured. The length of the paper path through the machine is also accurately known. The transportation lag is the length of the paper path divided by the machine speed. This approach applies to all industry sectors that rely on sheet processing.

For fluid flow systems, the residence time of a vessel or length of pipe is the volume of that entity divided by the volumetric flowrate through it. Volumetric flowrates and volumes are known. But the degree of mixing within the vessel or pipe is not known, and will fall somewhere between the two extremes of plug flow and perfect mixing.

Plug flow is the complete absence of mixing. The transportation lag is the residence time. Any change in variables, such as temperature or composition of the fluid entering the pipe, will not appear in the discharge until one transportation time has elapsed. Fluid flowing through a pipe approaches plug flow, but some mixing occurs, especially for turbulent flow. Consequently, the transportation lag is smaller than the residence time, but usually by an insignificant amount.

Perfect mixing is the gold standard for mixing. Variables such as temperature and composition are uniform throughout the mixed volume. The transportation lag is zero. Immediately after any change in temperature or composition of the inlet stream(s), the conditions throughout the vessel, as well as in the discharge stream(s), begin to change. Real mixers fall short of perfectly mixed systems, especially for large vessels mixing viscous materials, slurries, and the like. This can introduce a true transportation lag, but more likely produces several small mixed volumes in series that together exhibit behavior much like a transportation lag.

Assuming plug flow for a pipe is usually acceptable, but assuming all mixers are perfectly mixed is a dubious practice. In reacting vessels, pockets of poor mixing can create products other than the one(s) desired, thereby degrading the product quality. Experiments with tracers provide the data needed to assess the performance of the mixing equipment, and those data can be translated into the response behavior with which a control system must contend. However, the cost of conducting such tests solely to address process control issues is difficult to justify. A simpler approach is to obtain the response of the variable of interest to known changes in the inputs to the process. Response testing is much simpler, but still challenging to conduct in a production facility.

As a result, the transportation lag within a pipe, vessel, or any process equipment is rarely accurately known. At best, response testing provides an estimate of the transportation lag under the conditions at which the test was conducted. Changes in throughput certainly affect the transportation lag, but not necessarily proportionally. That is, a throughput increase of 10% does not necessarily reduce the transportation lag by 10%. Changes in fluid properties, especially viscosity, also affect the nature of the mixing and consequently the transportation lag.

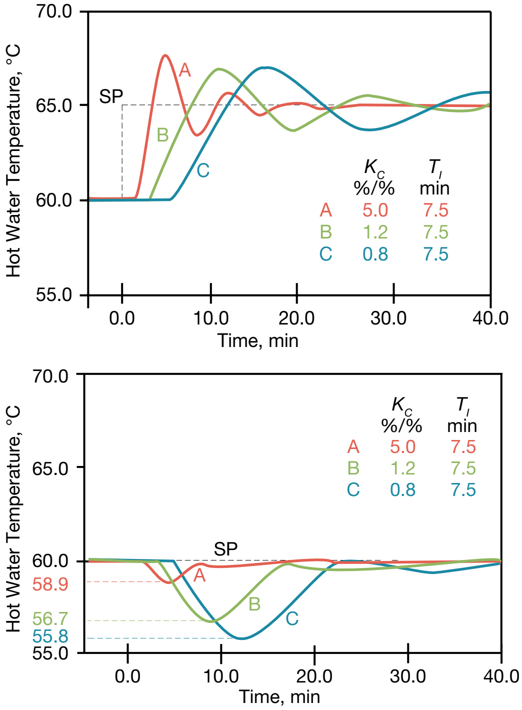
In processes for which the transportation lag can be computed (*e.g.,* paper machines), applying a control technique known as dead-time compensation significantly improves control system performance. This technique, in essence, remembers all control actions taken within the last transportation lag, uses a model to predict the effect of those actions on the controlled variable, adjusts the control error for this effect, and takes control action on what remains. The objective is quite simple: The results of control actions taken within the last transportation lag have not appeared in the controlled variable, so remember what control actions were taken and do not correct for the same control error more than once.

This model predictive control strategy has been very successful in paper machine control systems. However, its performance for applications outside the sheet processing industries is spotty — sometimes it works, but often it does not.

Transportation lag seriously degrades the performance of PID control loops, causing such loops to respond very slowly. For large transportation lags, the derivative mode is ineffective, and the controller gain must be reduced substantially to avoid excessive cycling in the loop. In effect, the integral mode becomes the primary contributor to the control action. Integral-only controllers respond very slowly.

Although not recommended in practice, transportation lag can be introduced into the temperature loop in Figure 1 by relocating the temperature transmitter at increasing distances downstream of the vessel. The responses presented in Figure 2 were produced by a controller whose temperature transmitter was located at the vessel discharge, so the only transportation lag within the loop was due to imperfect mixing within the vessel. The residence time of the vessel (volume divided by volumetric flowrate) is approximately 4 min.

Figure 4 illustrates the effect on loop performance after relocating the temperature transmitter so as to introduce transportation lags of 2.0 min (Case B) and 4.0 min (Case C). (Case A shows the responses depicted in Figure 2 and is included for comparison.) The direct effect of the transportation lag is to introduce more cycling into the response, which is undesirable in process applications. With PID control, maintaining the same degree of cycling requires reducing the controller gain, which causes the loop to respond more slowly (Figure 4a). A loop that responds more slowly to a disturbance has a larger maximum departure from the setpoint. Figure 4b shows that the maximum



▲ **Figure 4.** A transportation lag introduces more cycling into a loop in response to either (a) a setpoint change or (b) a disturbance. Reducing the controller gain maintains approximately a quarter decay ratio in each response.

departure from the setpoint increases from 1.1°C for the transmitter at the tank (no transportation lag) to 3.3°C for a transportation lag of 2 min to 4.2°C for a transportation lag of 4 min. When transportation lag dominates the dynamic characteristics of the process, a PID controller responds so slowly as to be of no value. This was the case for paper machines before the application of dead time compensation.

Process dynamics

Modern designs for continuous processes rely on steady-state models. Although not routine practice, the capability exists to extend these models to provide a dynamic simulation that accurately represents the behavior of the process. Dynamic simulations are often developed for sections of a process about which specific concerns have been voiced, but detailed dynamic simulations of complete plants are the exception.

Batch process designers rarely make use of models. It takes more effort to develop a simulation for a batch process, in large part because most batch plants make multiple products. Given a projected production requirement for each product to be manufactured, it is possible to optimize the design of a batch process. But the production requirements of a batch process plant continually change, and sometimes in big ways.

Dynamic simulations are effective for developing a startup strategy, testing the response to failures in critical components, and verifying that control configurations will perform as expected. Such purposes require a dynamic simulation that accurately represents the behavior of the process.

Not all undertakings require such accuracy. Simulations for operator training require only a crude representation of the dynamic behavior of the process. The systems integration effort for the process controls often includes a loopback simulation that involves connecting the outputs of the control loops to the inputs for the measured variables to demonstrate that the basic hardware and software are functional.

For good reasons, dynamic simulation practices in the process industries are very different from those of the aerospace industries. Before an airplane's first test flight, a detailed simulation is developed, because the basic controls must be operational when the vehicle lifts off the ground. Imagine doing this using the practices common in the process industries: Fuel up the vehicle, fire up the engines, and the vehicle lifts off with instrument technicians frantically tuning loops. That's not likely to have a good outcome.

Process facilities have an option not available to the aerospace industry. Process plants can be operated with the controls completely on manual (the possible exception being the occasional very fast loop). Manual operations have become unacceptable for normal production — additional personnel are required and plant performance suffers. However, manual operations are acceptable for startup. A few loops can be tuned prior to startup while blowing air or pumping water through the plant equipment. Otherwise, process operations initially commence under manual control, with the objective of achieving near-normal operating conditions. Only then can most of the important loops be tuned.

Manual control is not limited to the initial startup. Most loops are on manual during subsequent startups. Loops are tuned to the process behavior at normal operating conditions. During the early phases of a plant startup, the process conditions are so different from the normal operating conditions that most loops perform poorly, if at all. Manual control is continued until the process is sufficiently close to normal operating conditions that a loop will function properly.

What would provide the incentive to develop a complete dynamic simulation for all new chemical processing plants? When the controls must be largely (if not totally) on automatic during the initial plant startup. This is the case in the aerospace industry, but is not expected in the process industries for the foreseeable future. Short of this, dynamic simulations will be undertaken only for those parts of the process for which the cost of the simulation effort can be justified by the benefits.

Steady-state characteristics

The last two unique aspects of chemical process control — economics and the multivariable nature of industrial processes — pertain to steady-state characteristics of the process. These will be explained in Part 2, which is scheduled to appear in the April issue but can be viewed online together with Part 1 at www.aiche.org/cep. **CEP**

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