Closing the control loop (why we need process control)

Julian Morris FREng FIChemE
Centre for Process Analytics and Control Technology, School of Chemical Engineering and Advanced Materials, University of Newcastle, Newcastle upon Tyne NE1 7RU, UK

Column Editors: John Andrews and Paul Dallin
Clairet Scientific Ltd, 17 Scirocco Close, Moulton Park Industrial Estate, Northampton NN3 6AP, UK.
E-mail: john.andrews@clairet.co.uk, paul.dallin@clairet.co.uk

In previous articles we've looked at process analysis using spectroscopic techniques, but not covered the important area of Process Control. It is seen as increasingly important that process analysis is used to "close the loop" in real processes and indeed it has been said that PAT applications are not true PAT applications without the implementation of process control. We are therefore extremely grateful to Professor Julian Morris for the following article explaining the basics of Control to our spectroscopist readers.

Introduction

Many systems make control decisions based on the feedback of information (a signal) from the plant or process under control. Control systems that use feedback are called closed-loop control systems. The feedback is used to make decisions about changes to the control signal that drives the plant. An open-loop control system doesn't have or doesn't use feedback. A basic feedback (closed-loop) control system is shown in Figure 1.

Improved closed loop control is one of the most cost-effective ways of improving profit. Remarkable benefits often result from simple, inexpensive improvements and it is virtually the only technology where major improvements can be made between shutdowns. If the automatic control systems on a plant are ineffective, operational decisions must be left to operators. Good as they may be, they all have different perceptions of how best to operate the plant and most prefer a stress-free life. The result is that they will run the plant conservatively, tend to over purify products and maximise their "comfort margin". Operators prefer to keep the plant well away from operating constraints, but this is usually at the cost of inefficient operation. On batch plants it is remarkable how many batches last for the length of a shift!

Benefits of improved control have been quantified at between 2% and 6% of energy and material costs, AND can usually be recovered by applying the correct level of improved control. Most plants will benefit from one or more of: improved material efficiency; reduced energy consumption; increased capacity; improved product quality, consistency and reduced product "give-away"; and better response times. Poor temperature control is a common cause of extended batch times.

These benefits do not come free, but control improvement projects are one of the most cost-effective ways of improving profitability. Simple and inexpensive improvements can often yield remarkable benefits. Despite excellent returns on investments, control projects are often perceived as expensive and risky, usually because the benefits have been inadequately estimated. Indeed the importance of closed loop control has still not been fully recognised in the now widely publicised Process Analytical Technologies (PAT) initiative.

So what is closed loop process control? It is the control of production processes through a rational arrangement of process equipment (unit operations) with measuring/analytical devices, control valves, actuators, process control systems and computers. These together help satisfy the need for the continuous monitoring and control of process operations to provide for:

Safety—the safe operation of a chemical process is the primary requirement for the well-being of the people in and around the plant, the environment and the plant itself.

Production Specifications—the maintenance of product and quality (e.g. composition, purity, colour, yield, chemical effects etc.) to meet changing customer demands at minimum cost and minimum environmental impact.

Environmental Regulations—to meet environmental constraints and regulations by ensuring that the production process stays within operating limits on key process variables and concentrations of chemicals and effluents.

Operational Constraints—to ensure that the various types of equipment used in a chemical or biochemical process operate within their physical, chemical and biological limits.

Economics—to ensure that the manufacturing process conforms to market conditions in terms of availability of raw

PROCESS COLUMN

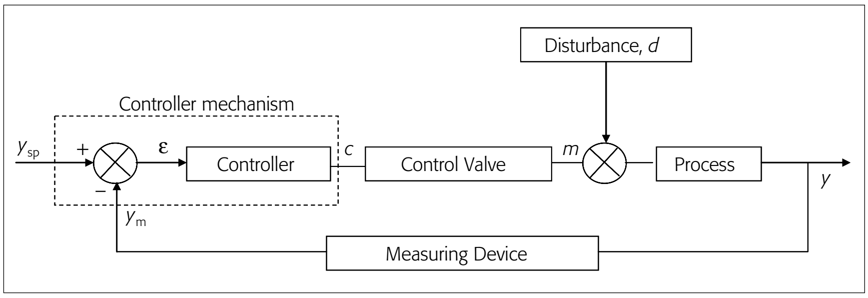

Figure 1. Feedback control.

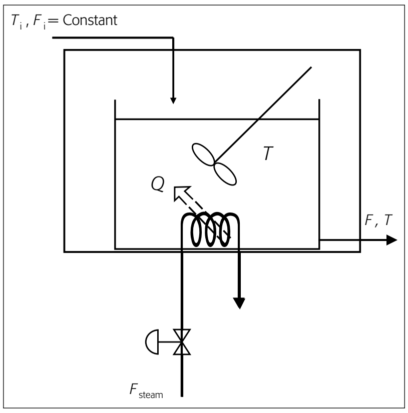

Figure 2. CSTR open loop.

materials, energy, labour and customer needs.

Feedback control

Feedback control goes back to very ingenious control systems for water-level control used by the Greeks as early as 250 BC. The mode of operation was very similar to that of the water level regulator in the modern flush toilet. The "fly-ball" governor, which was first applied by James Watt to his new steam engine in 1788 played a key role in the development of steam power. Feedback control became essential for the development in the 1930s of high-gain operational amplifiers that are widely used in electronic equipment and in guidance systems for guns, aircraft and missiles. During the 1930s the three-mode pneumatic controllers with Proportional, Integral and Derivative (PID) feedback control actions became commercially available with the first theoretical papers on process control being published in 1933/34. The 1940s saw their widespread industrial acceptance with electronic versions entering the market in the 1950s and computer control being seen in the late 1950s and early 1960s. Today the use of the PI(D) controller is ubiquitous along with its multiloop counterparts. However, more and more model-based predictive controller approaches are being seen, especially in large scale continuous processing but rarely, as yet, in batch processing.

Process control systems, Figure 1, are used to control or regulate dynamic (time varying) process operations whilst ensuring that steady state (equilibrium) operations are also satisfied. They achieve this by minimising process variability on important process outputs, or product quality measures, and moving variability to other process inputs where variability can be better tolerated. Thus, process control transfers variability from one part of a production process to another—it does not reduce overall process variability.

Regulatory and servo control

Process Variables are classified as being: measured input variables (e.g. temperature, flow and composition, viscosity etc. of feed streams); measured output variables (e.g. temperature, flow, viscosity and composition etc. of outlet streams); manipulated or adjustable variables by a human operator or control system; measured disturbances; unmeasured process Inputs, Outputs and Disturbances.

Consider the simple stirred tank heating system:

Consider the open loop continuous stirred tank reactor (CSTR) in Figure 2 under steady state operating conditions. The energy balance is:

$$0 = F\rho c_p(T_{i,s} - T_s) + Q_s$$

where the subscript, s, indicates steady state; T_s is the steady state temperature in the tank and Q_s is the corresponding steady state heat input. For a step change in the feed temperature from $T_{i,s}$ to T_i, the dynamic relationship for the temperature, T, is:

$$V\rho c_p \frac{dT}{dt} = F\rho c_p(T_i - T) + Q$$

The simplest control law is to require that the heat input, Q, changes proportionally to the error $(T - T_s)$, this is termed ***Proportional (P)*** control action:

$$Q = K_P(T_s - T) + Q$$

where K_P (sometimes written as K_C) is known as the proportional gain. A proportional controller reduces error but does not eliminate it (unless the process has naturally integrating properties). Thus P action alone will not make the tank temperature return to the steady state value of T_s following the change in inlet temperature T_i and a steady state ***offset*** (a fixed discrepancy between the desired process output and the actual process output) will result.

Consider now the improvement in the quality of the control performance that can be achieved using ***Integral (I)*** control action. Here Q is proportional to the integral of $(T - T_s)$:

$$Q = K_I \int_0^t (T_s - T)dt + Q$$

where K_I is the Integral gain. This control action drives the error $(T - T_s)$ to zero.

Reset is often used to describe the integral mode. Reset is the time it takes for the integral action to produce the same change in manipulated variable as the Proportional mode initial (static) change.

Combining ***Proportional action*** with ***Integral action*** gives the ***Proportional + Integral Controller*** (***PI controller***):

$$Q = K_P(T_s - T) + K_I \int_0^t (T_s - T)dt + Q_s$$

In a similar manner, ***Derivative action (D)*** can be added by taking the rate of change of the error.

$$Q = K_D \frac{d(T_s - T)}{dt} + Q_s$$

where K_D is the Derivative gain.

In practice P, I and D control actions are not used alone but in combinations, the most usual being alone, PI-action (the most widely seen) and sometimes PID-action—the well know being ***Proportional–Integral (PI) and Proportional–Integral–Derivative*** (***PID***) controllers:

$$Q = K_P(T_s - T) + K_I \int_0^t (T_s - T)dt + Q_s$$

$$Q = K_P(T_s - T) + K_I \int_0^t (T_s - T)dt + \frac{d(T_s - T)}{dt} + Q_s$$

The most general, sometimes referred to as the "ideal" or "text book" form of the PID controller is written in terms of its output to the control valve or actuator termed $c(t)$ $m(t)$ or $u(t)$:

$$c(t) = K_P \varepsilon(t) + \frac{1}{\tau_I} \int_0^t \varepsilon(t)dt + \tau_D \frac{d\varepsilon(t)}{dt}$$

where τ_I is the integral time constant or reset time, τ_D is the derivative time and $\varepsilon = (T_s - T)$. The resulting closed loop control system is as shown in Figure 3.

Unfortunately, there is no "standard" or correct PID algorithm.[a] A major disadvantage of this ideal "textbook" configuration, above, is that a sudden change in set point (and hence ε) will cause the derivative term to become very large and thus provide a "derivative kick" to the final control element—this is undesirable. In the text-book controller form it can be seen that the derivative action acts on the set point change and hence any sudden change of the desired process output by a set point change will result in a sudden change of the manipulated variable which is not ideal in a real process situation. There are a number of ways of avoiding this situation. For example, avoid differentiating the set point error (ε) and differentiate to process output, filter the derivative term etc.

In addition, in industrial applications, PI and PID controllers need special provisions to stop the controller becoming saturated during plant start-up, shutdown, change-overs, manual operations etc. where the error can persist for long periods of time—***integral saturation or integral (reset) windup***.

As an example of closed loop control, consider now that a liquid enters the

[a]Different control manufacturers use different definitions for the integral mode of a controller. It can be defined as minutes, minutes/repeat or repeats per minute. The difference is very important so as to ensure problems do not occur during controller tuning. τ_I is the integral time (minutes), if specified as repeats/minute then it is $1/T_i$ that must be entered into the controller, while minutes/repeat is again τ_I. This can be confusing and is compounded by the fact that manufacturers are not consistent.

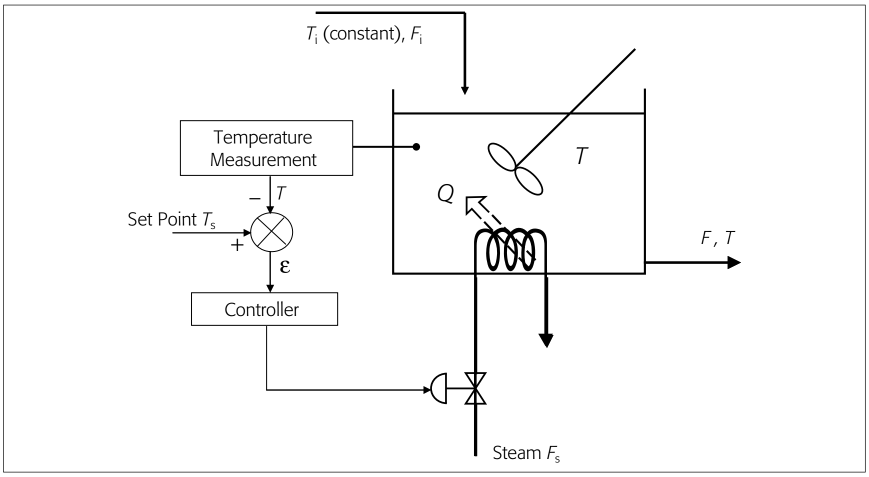

Figure 3. CSTR closed loop control.

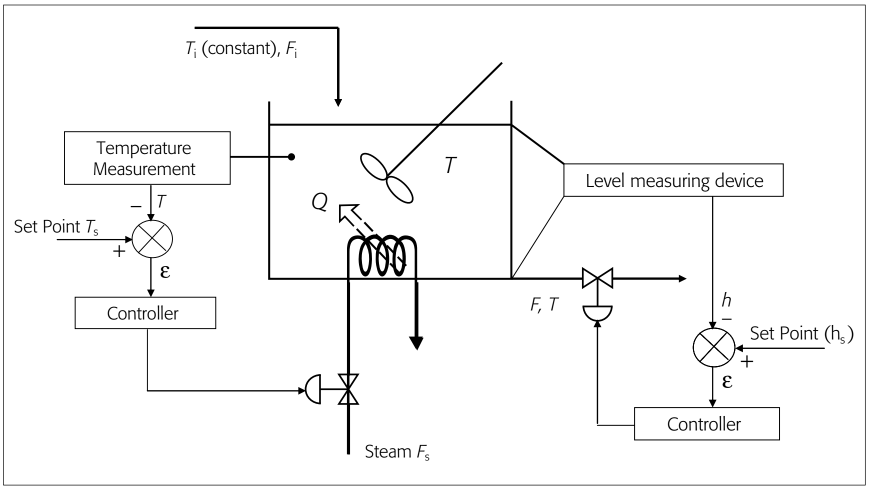

Figure 4. Closed loop temperature and level control.

open loop tank heater system shown in Figure 2 with flowrate F_i and temperature T_i, where it is heated by steam having a flowrate F_{st}. The tank is assumed to be well stirred and the effluent stream has the same temperature as the liquid in the tank. The operational requirements are:

- to keep the effluent temperature T at a desired value T_s (target value or set point)
- to keep the volume of the liquid in the tank (i.e. liquid level) at the desired value V_s (level control, h_s).

The operation of the heater is disturbed by external factors such as changing feed flowrate and feed temperature. If nothing changed than once a steady state has been established ($T=T_s$ and $h=h_s$) then the system could be left without supervision and control.

This is not the case in real life and disturbances will enter the system. Some form of control is required.

Figure 3 shows a feedback control scheme to keep $T=T_s$ when T_i or F_i changes. A thermocouple, or resistance thermometer, measures the tank temperature T and compares it to the desired temperature, or set point value T_s, giving a deviation $\varepsilon=(T-T_s)$.

The value of the error or deviation is passed to the controller which applies an algorithm to make the effluent temperature return towards T_s.

Controller actions

If $\varepsilon>0$, implying that $T<T_s$, the controller opens the steam valve to increase the temperature and return T towards T_s.

If $\varepsilon<0$ or $T>T_s$ the controller will close the steam valve.

It is clear that when $T=T_s$ (i.e. $\varepsilon=0$), the controller does nothing.

A similar configuration can be used to keep the tank volume V, or the liquid level h, at its set point h_s when F_i changes. Figure 4 shows one possible scheme.

Although feedback control is the most common control strategy in the process industries (~95% of all control loops); in some processes it may not provide the required control performance and more advanced control strategies are required. For example, some of these disadvantages can be overcome to a greater or lesser extent by other control methods such as: ratio control, cascade control, combined feed forward–feedback control, selective, override and auctioneering control.

Advantages of feedback control

Corrective action occurs as soon as the controlled variable deviates from set point, regardless of the source or type of disturbance.

Requires minimal knowledge about the process—a mathematical model is ***not*** required although they are very useful in control systems design.

The Proportional + Integral + Derivative (PID) controller (regulator) is versatile and robust. When process conditions change, re-tuning the controller settings usually produces satisfactory control.

Disadvantages of feedback control

No corrective action is taken until a deviation in the controlled variable is observed. Thus perfect control is impossible.

It does not provide predictive control action to compensate for the effects of known or measurable disturbances. It may not be satisfactory for processes with large time constants and/or long time delays (transportation lags). If large and frequent disturbances occur, the process may operate continually in a transient state and never attain the desired steady-state.

To illustrate the impact of a time delay, consider a distillation composition measurement example shown in Figure 4, has required the use of an on-line analyser.

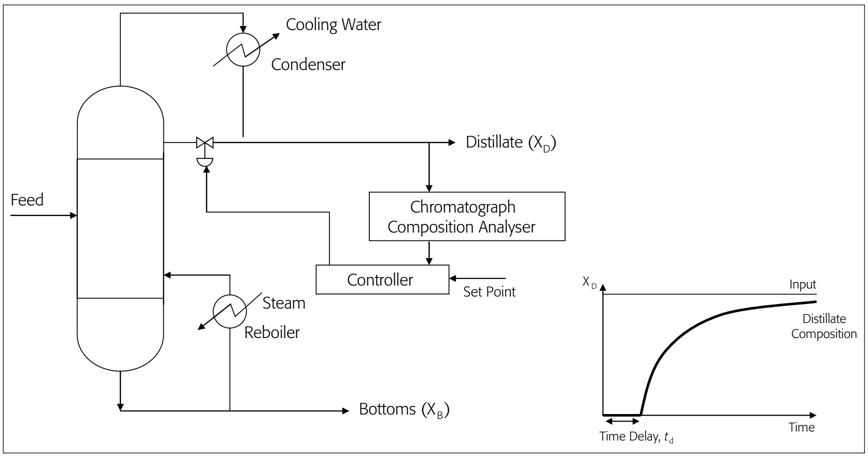

Figure 5. Distillation Column Analyser Control. (Inset: effect of the chromatograph cycle time–time delay.)

Analysers are relatively delicate and expensive instruments and may need to be remotely mounted in a protective environment. A slipstream from the product provides a sample to this analyser. The combination of the flow delay with the analyser sampling interval and analysis time add a significant delay in measuring product composition.

Disturbance variables (say in the column feed, utility supplies such as steam etc., also affect control loop performance. These are variables that can't be controlled or manipulated. Sometimes they can be measured, providing an opportunity for feedforward control. When this is not possible, the variation in disturbance variables is only recognised by their affect on the controlled variables.

In process control, the time delay or dead time is the time it takes since the moment any change in the control signal until a reaction is seen in the output variable. Possible sources of time delays are: (i) the process may involve the transportation of materials or fluids over long distances, (ii) the measuring device may be subject to long delays to provide a measurement (e.g. operator sampling, long cycle time chromatographs, or in some cases the final control element may need some time to develop the actuating signal). In contrast spectroscopic analysers may provide faster analysis times which reduces the time delay. The presence of time delays causes the following difficulties in process control: (i) a disturbance entering the process will not be detected until after a significant period of time, (ii) the control action will be inadequate since its effects on a current error will affect the process variable only after a long delay; and long time delays may originate instability in the system. The relationship between such characteristic process parameters and the difficulty of control has been the focus of countless technical articles and textbooks, described by mathematics and practical experience.

Key generic characteristics that describe a process response to input changes and its approach to steady state include: ***Time delay or Dead time***—the time between a change in the control signal and the ***beginning*** of the process variable response; ***Time Lag***—the time constant of the process variable response, once it begins; and ***Steady state gain***—the ratio of the size of process variable change to the size of control signal change. In terms of closed loop control:

Time delay is the process characteristic that makes control difficult. During this delay, a controller sees no response to its control action. A feedback controller has to be detuned so that it will not overreact during this delay and thereby overcorrect for an error condition. But too much detuning can cause sluggish control.

Controlling processes dominated by capacity lags (large values of τ) is much easier. A large lag makes a process variable change more slowly and filters noise from the measurement signal. However, many small lags can combine and look like dead time.

Process gains dictate how tightly the controller can be tuned. High gain processes require low gain controllers, and vice versa. Variable gains in the process or valve characteristics can be a problem for any controller, since loop stability is variable. Controllers have to be tuned for stability when the process gain is highest. Then, and at other times, the controller can be too sluggish. Adaptive tuning can address this problem.

Conclusions

Finally some observations are important:

A control system can be ***too perfect***—it must be justified on a ***cost–benefit*** basis.

Design the process and the process controls as an integrated task.

The use of dynamic process simulations can ensure the success of the final control scheme. These can be based both on known physical and chemical relationships such as dynamic mass and energy balances and known kinetic relationships and "'transfer functions" to represent the "shape" of any process dynamics not known.

Knowledge of multivariate statistics or process chemometrics is desirable.

The job of the Process Control Engineer is to understand the system—the chemistry or biochemistry, the chemical or biochemical engineering, the mathematics of the control theory, and how to design control schemes and tune them.

With the increasing interest now being shown in the FDA Process Analytical Technologies (PAT) initiative, it is becoming clear from most all of the conferences addressing PAT that "closing the control loop" is not receiving the attention it warrants and again risks being left out of the system design and being thought about later. Closing the loop with on-line real-time spectroscopy opens up immense potential for product and process scale-up, faster time-to-market and production cost savings. Not considering control technologies from the start of product and process innovation risks repeating the mistakes so often made in the past.