FEEDBACK CONTROL

1. Feedback closed-loop system

The simplest and most widely used method for process control is the feedback control loop shown in Figure (1). A measurement of the controlled process variable (CV) is compared to a set point (SP) to create an error. This error is used to derive the correction action of the final control element (FCE) via the controller. The controller output changes the manipulated variable (MV). The action of the controller may be aggressive or sluggish; it depends on the internal control law and tuning that is used. In this lecture we will discuss feedback controller types and tuning them in order to keep the CV at the desired SP in presence of process disturbances. Disturbances typically come in three types: input disturbances, load disturbances, and set point disturbances. An input disturbance is the change in the mass or energy of the supply to the process which could change the conditions of the process variables. A load disturbance is any other upset, except for an input mass or energy changes. In this work we use the letter D for the previous type of disturbances. A set point disturbance (SP) occurs when the desired state of the CV changes, and the process must adjust to a new state.

All elements of the feedback loop can affect control performance. The controller output consists of the feedback signal which has a range usually expressed as 0 to 100%. When the signal is transmitted electronically, it usually converted to a standard signal range of 4 to 20 milliamperes (mA) and can be transmitted long distances (over one mile). When the signal is transmitted pneumatically, it has a range of 3 to 15 psig and can only be transmitted over a shorter distance. Pneumatic transmission is not common with modern equipment since it requires longer time (several seconds) than electronic transmission.

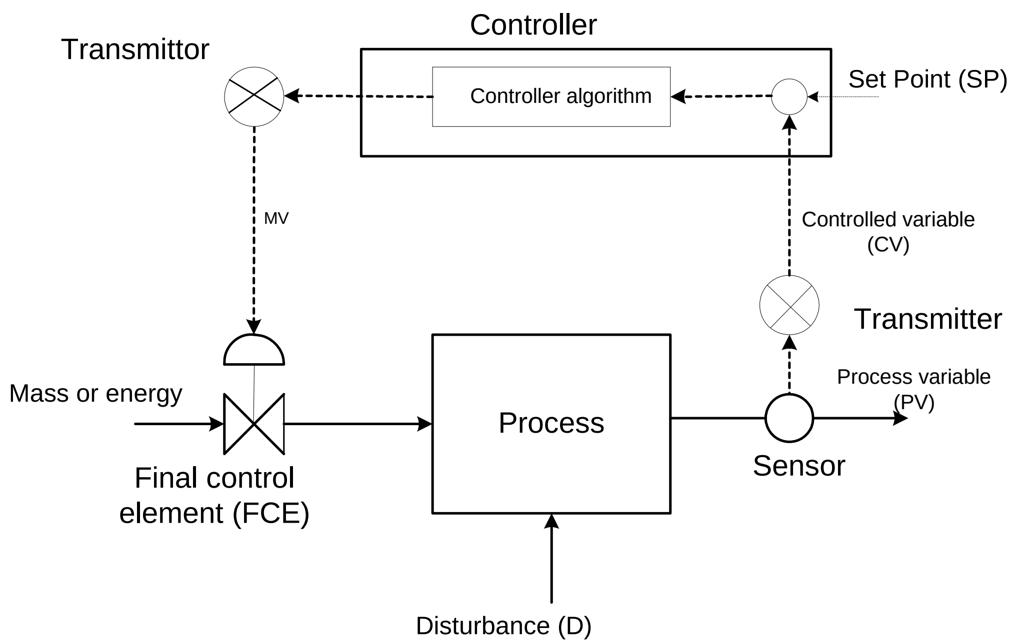
At the process unit, the transmitted signal is used to adjust the final control element. The FCE as in over 90% of process control applications, is a valve. The valve percent opening could be set by an electrical motor, but this is not usually done because of the danger of explosion with high-amperage power supply a motor would require. The alternative power supply typically used is a compressed air. The signal is converted from electrical to pneumatic with 3 to 15 psig range. The pneumatic signal is transmitted a short distance to the control valve. Control valves respond relatively quickly, with typical time constant ranging from 1 to 4 seconds.

After the FCE has been adjusted, the process responds to the change. The process dynamics vary greatly for the wide range of equipment in the process industries, with typically dead times and time constants ranging from a few seconds (or faster) to hours. When the process is by far the slowest element in the control loop, the dynamics of the other elements are negligible. This situation is common in chemical engineering applications.

The sensor responds to the change in the plant conditions. Usually, the sensor is not in direct contact with the potentially corrosive process materials; therefore, the protective equipment or sample system must be included in the dynamic response. For example, a thin thermocouple wire respond quickly to a change in temperature, but the metal sleeve around the thermocouple, the thermowell, can have a time constant of 5 to 20 seconds. Most sensor systems for the flow, pressure, and level have a time constants of a few seconds. Analyzers that perform complex physiochemical analysis can have much slower responses, on the order of 5 to 30 minutes or longer.

The sensor signal is transmitted to the controller, which is considered to be located in a remote control room. The transmission could be pneumatic (3-15 psig) or electrical (4-20 mA). The controller receives the signal and performs its control calculations. The controller can be an analog system (electrical analog controller consists of an electrical circuit that obeys the same equations as the desired controller calculations) or a digital system where the controller is a continuous electronic controller that performs its calculations instantaneously (usually a very fast digital computer is used as in most modern control equipment).

It is worth recalling that the empirical methods for determining the process dynamics presented in the previous lectures involve changes to the manipulated signal and monitoring the response of the sensor signal as reported to the control system. Thus, the resulting model includes all elements in the loop including instrumentation and transmission.



Figure(1): Feedback Control System

2. Feedback Block Diagram

Consider a general block diagram for a feed back control system as shown in Figure (2). The transfer functions for the control system relate the controlled variable to the external disturbance (D) and to the desired set point (SP) can be defined as

$$\frac{CV(s)}{SP(s)} = \frac{G_p G_v G_c}{1 + G_p G_v G_c G_s}$$

$$\frac{CV(s)}{D(s)} = \frac{G_d}{1 + G_p G_v G_c G_s}$$

The block diagram provides a visual "picture of the equations" showing the feedback loop. The general closed-loop transfer function model can be applied to any specific system by substituting the transfer function models for the loop elements.

The first major reason for the feedback system is to maintain the controlled variables at the desired set point which is also refereed as the servo problem. This can be achieved by choosing the suitable controller (G_c) which make the ratio of CV(s)/SP(s) equal one or

$$G_p G_v G_c \approx 1 + G_p G_v G_c G_s$$

The second reason is to reject the effect of the disturbance on the controlled variable. For this case, the controller transfer function (G_c) should be selected to make the ratio of CV(s)/D(s) approaches zero

$$G_d <<< 1 + G_p G_v G_c G_s$$

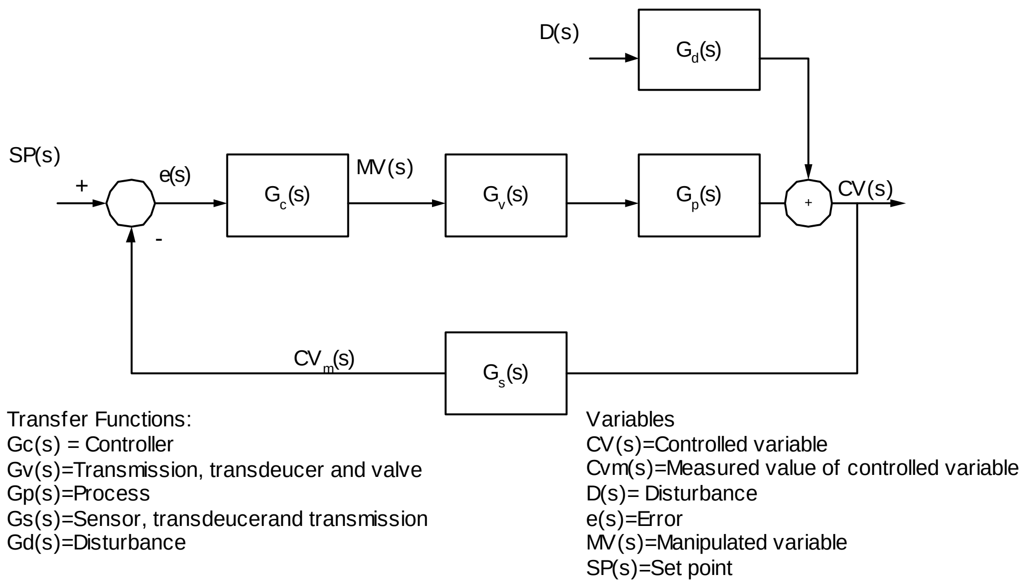


Figure (2) Feedback System Block Diagram

2.1 Closed loop stability

The essential information on stability is in the denominator of the closed loop transfer functions, called the characteristic polynomial:

$$1 + G_p G_v G_c G_s$$

A closed-loop control system is locally stable at the steady state point if all roots of the characteristic polynomial have negative real parts. If one or more roots with positive or zero parts exist, the system is locally unstable. Entries in the overall transfer function denominator demonstrate that only the elements in the feedback loop affect the system stability; neither the disturbance nor the set point change affects stability. Essentially all real process control systems can be made unstable simply by using unsuitable controllers. The most commonly used stability analysis methods are summarized in Table (1).

Table (1) Stability analysis methods

Method	Plant model	Stability results	Results display
Root locus	Polynomial in *s*	Relative	Graphical
Routh	Polynomial in *s*	Yes or No	Tabular
Bode	Open loop stable Monotonic decreasing amplitude ratio and phase angle as frequency increases	Relative	Graphical
Nyquist	Linear	Relative	Graphical

3. Controllers Modes

3.1 On-off control

The simplest form of control law is on-off control. This type of controller is primarily intended for use with final control elements (FCE) that are non-throttling in nature. An example is a home heating system, whenever a temperature goes above the set point, the heating process shuts off. For this example, the behavior of this control is given by:

$$MV = 0\% \quad for \quad CV > SP$$
$$MV = 100\% \quad for \quad CV < SP$$

This type of control can be used successfully for processes with large capacitance where tight control is not important (capacitance represents system ability to absorb or store mass or energy and it could be defined as the resistance of a system to the change of mass or energy stored in it). A good example of this process is a surge tank. One disadvantage of this type is the continual opening and closing of the controller, the FCE quickly becomes worn and must be replaced. The primary characteristic of on-off controller is that the process variable is always cycling about the set point. The rate of CV cycling and its deviation from the set point are a function of the dead time and capacitance in the system. Most on-off controllers are built with an adjustable differential gap, inside which no control action takes place. This is to minimize the controller cycling and extend the operational life of the FCE.

3.2 Proportional Control (P)

Proportional control is the simplest continuous control mode that can damp out oscillations in the feedback control loop. This mode stops the CV from cycling but does not necessarily return it to the set point. The output of the P controller is proportional to the error:

$$MV = K_c e$$

the transfer function for the proportional controller has the form

$$G_c(s) = K_c$$

where K_c is the controller gain and e is the error defined by the difference between the controlled variable and the set point.

$$e = SP - CV \text{ for reverse acting}$$
$$e = CV - SP \text{ for direct acting}$$

Increasing K_c can decrease the error, but we should be aware not to increase K_c such that it makes the closed loop unstable. Because of this loop gain limit, there is another approach to reducing the error to zero. Another term can be added to the proportional controller:

$$mv = K_c e + b$$

this additional term is called the bias, and is simply defined as the output of the controller when the error is zero.

The controller gain is the ratio of the change of controller output to the change in error. Since there is a one-to-one relationship between CV and e the controller gain can be written as:

$$K_c = \frac{\Delta MV}{\Delta CV}$$

The proportional band is defined as the change in CV that will cause the output of the controller to change by 100%. The controller gain can be related to the proportional band by:

$$K_c = \frac{100\%}{PB\%}$$

Virtually all modern controllers use a gain adjustment, however a few older controllers exist that still use a proportional band adjustment.

In general, a proportional controller provides a fast response compared to other controllers but a sustained error occurs where the CV does not return to the set point even when steady-state is reached. This sustained error is called offset and is undesirable in most cases.

3.3 Integral control (I-only)

The action of integral control is to remove any error that may exist. As long as there is an error present, the output of this controller mode continues to move the FCE in a direction to eliminate the error. The equation for this control mode is:

$$MV = \frac{1}{T_i}\int edt + mv_o$$

mv_o is defined as either the controller output before integration, or the initial condition at time zero. The integral time T_i, is defined as the amount of time it takes the controller output to change by an amount equal to the error. Thus, it is measured in minutes per repeat. As a result of the reciprocal relationship of the integral time, some manufactures adjust their controllers in repeat per minute ($1/T_i$). Then increasing such adjustment gives less integral action.

Although an integral only controller provides the advantage of eliminating the offset, there is a significant difference in its response time when compared to a P only controller. The output of the P controller changes as quickly as the measurement changes, so if the error measurement change as a step, the controller output also changes as a step by an amount depending on the controller gain. For a step input to an integral controller, the output does not change instantaneously but rather by a rate that is affected by T_i and e.

Hence, an integral only control, due to the additional lag introduced by this mode, has an overall response that is much slower that that for proportional controller. If no offset is required then a slower period of response must be tolerated. If the requirement is a return to the set point with no offset and a faster response time is necessary, then the controller must be composed of both P and I actions.

3.4 PI Control

The majority (>90%) of controllers found in plants are PI controllers. The equation for PI is given as:

$$MV = K_c(e + \frac{1}{T_i}\int edt)$$

and the PI controller transfer function is:

$$G_c(s) = K_c(1 + \frac{1}{T_i s})$$

The PI controller gain has an effect not only on the error, but also on the integral action. The bias term for PI control when compared with the *P* controller equation is:

$$b = K_c \frac{1}{T_i} \int edt$$

Therefore, the integral action provides a bias hat is automatically adjusted to eliminate any error. The PI control is faster in response than the I-only controller because of the addition of the proportional action. The gain of the PI controller, K_{PI} can be defined as:

$$K_{PI} = K_c + \frac{K_c}{T_i}$$

The PB and T_i are used to adjust the PI controller to give the loop a desired response. When setting T_i to a very large number in min/rep, the integral action is minimized and the controller response would be very close to the P-controller. While, if we set T_i to a very small value, the controller gain would approach that of integral-only controller and the controller action will return the CV to the set point but a with a long response period.

The response of the PI controller will be slower than the P controller. Thus, the response period of the a loop under PI control is 50% longer than that for a loop under P-only control. In order to increase the speed of the response it may be necessary to add an additional control mode.

3.5 Derivative control

The purpose of the derivative controller is to provide lead to overcome lags in the loop. It anticipates where the process is going by looking at the rate of change of error. The output of the controller is defined as:

$$output = T_d \frac{de}{dt}$$

Figure (3) shows how the output from a derivative block would vary for different inputs given a fixed value of T_d. As the rate of change of the input gets larger, the output gets larger. When the slop of the input approaches infinity (very close to a step function), the output would be a pulse of infinite amplitude and zero time length. This output is unrealizable since a perfect step with zero rise time is physically impossible, but signals that have short rise and fall time do occur and these are referred as noise. Thus, the output from the derivative block would be a series of positive and negative pulses, which would try to derive the FCE either full open or full close. This would result in accelerated wear on the FCE and no useful control.

For processes, where the measured variable has high frequency noise (even with small amplitude), the derivative control will notice them and the control outputs would be a series of large amplitude pulses. For such case, the noise must be filtered or eliminated by modifying the installation of the primary sensor.

It is important to note that derivative control would never be the sole control mode used in a controller. The derivative action does not know what the set actually is and hence can not control to a desired set point.

The minimum controller configuration containing derivative action is a combination of proportional plus derivative action:

$$MV = K_c(e + T_d \frac{de}{dt}) + b$$

The addition of the derivative action results in a faster response for the measurement process variable and a smaller offset than the loop under P-only control. In the previous PI case, T_i can be set to a large number to eliminate the integral action, while in the PD controller, even by setting T_d to a very small value, there is still the possibility of a sizable derivative action if there is a noisy input (if dCV/dt is large). In electronic controllers and distributed control systems (DCS) the derivative action can be eliminated by setting T_d to zero. In a pneumatic controller the derivative action can not be eliminated but can be reduced to a minimum value of approximately 0.01 min. If a PD controller is installed on a flow loop there will still be considerable derivative action due to the noisy flow measurement. It is therefore important when applying a pneumatic controller to a noisy loop to make certain the controller does not contain a derivative block.

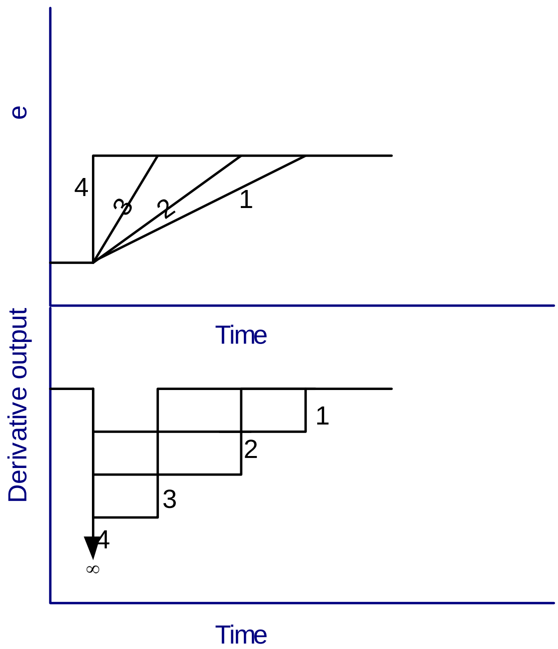


Figure (3) Derivative action

3.6 Proportional integral derivative (PID) control

The primary purpose of a PID is to provide a fast response that is much the same as with P-only controller but which has no offset. The derivative action adds the additional response speed required to overcome the lag in the response from the integral action. The control algorithm for the PID controller is given by:

$$MV = \frac{100}{PB}(e + \frac{1}{T_i}\int edt + T_d \frac{de}{dt})$$

$$MV = \frac{100}{PB}(e + \frac{1}{T_i}\int edt - T_d \frac{dCV}{dt})$$

The PID transfer function is given by:

$$G_c(s) = K_c(1 + \frac{1}{T_i s} + T_d s)$$

Figure (4) presents a comparison of the responses for P, PI and PID controllers to a step change in the load. Therefore, a PID controller provides a tight dynamic response, but since it contains a derivative term, it cannot be used in any process which noise is anticipated.

The flow chart given in Figure (5) summarizes a procedure for controller selection.

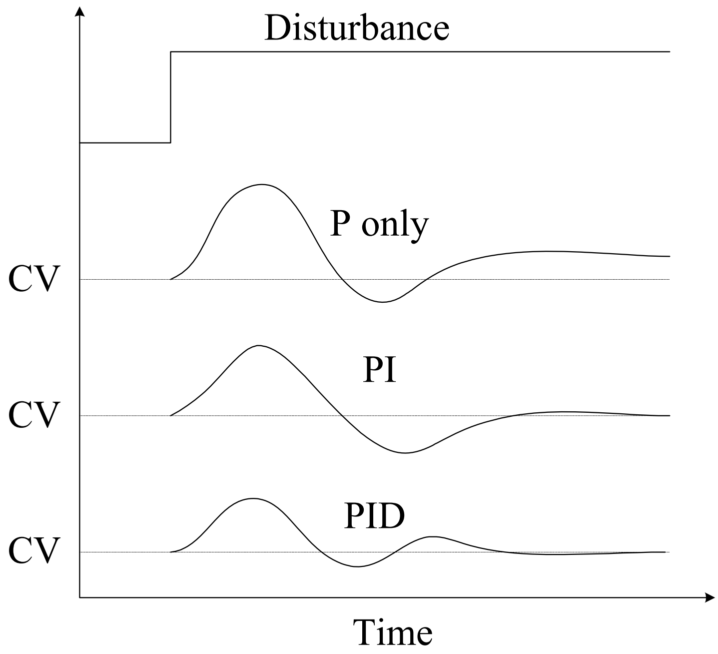


Figure 4: P-only, PI and PID responses to a load disturbance

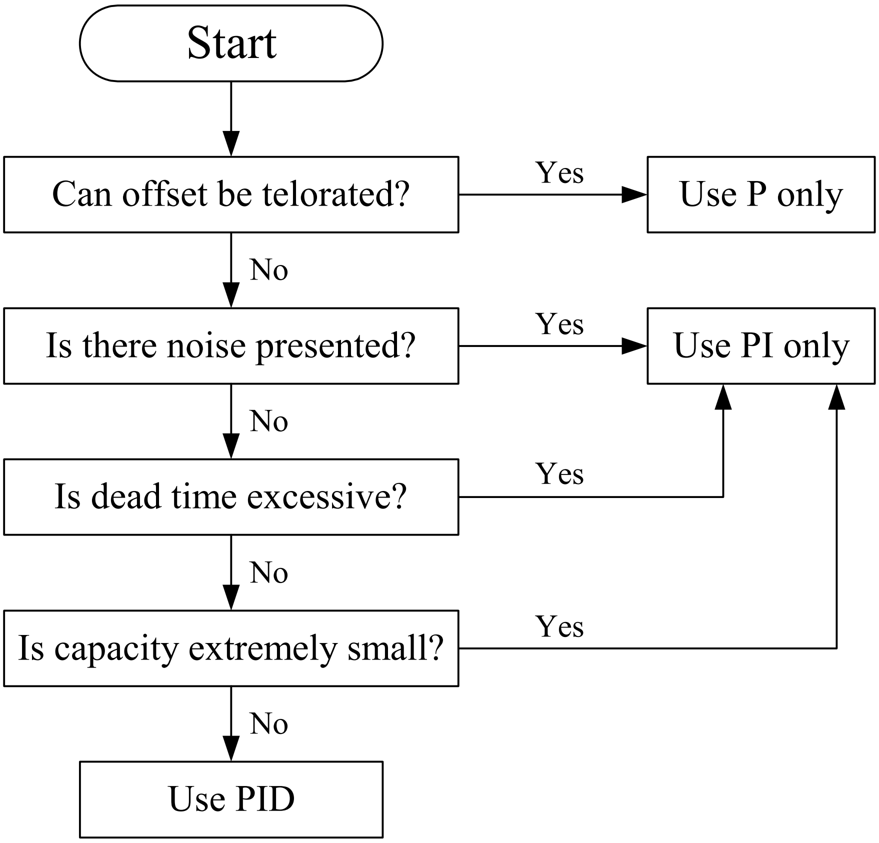


Figure (5) Flowchart for Controller selection

4. Tuning Feedback Controllers

Controller tuning can be defined as an optimization process that involves a performance criterion to the form of the controller response and to the error between the process variable and the set point. Depending on the process to be controlled, the first consideration is to decide what type of response is optimal. Typical process responses to load changes are shown in Figure (6). There are three general possible responses:

1. Over-damped - slow response with no oscillations
2. Critically damped – fastest response without oscillations
3. Under-damped – fast return to set point with considerable oscillations.

The selection of good control is a trade off between the speed of the response and the deviation from the set point. A highly tuned controller may become unstable if large disturbances occur, whereas a sluggish tuned controller provides poor performance but is very robust. What is typically required for most process control loops is a compromise between performance and robustness.

There are several common performance criteria that can be used for controller tuning, based on the closed loop response. In the following we review the most important criteria:

Cyclic radian frequency

$$\omega = 2\pi f$$

And

$$f = \frac{1}{period}$$

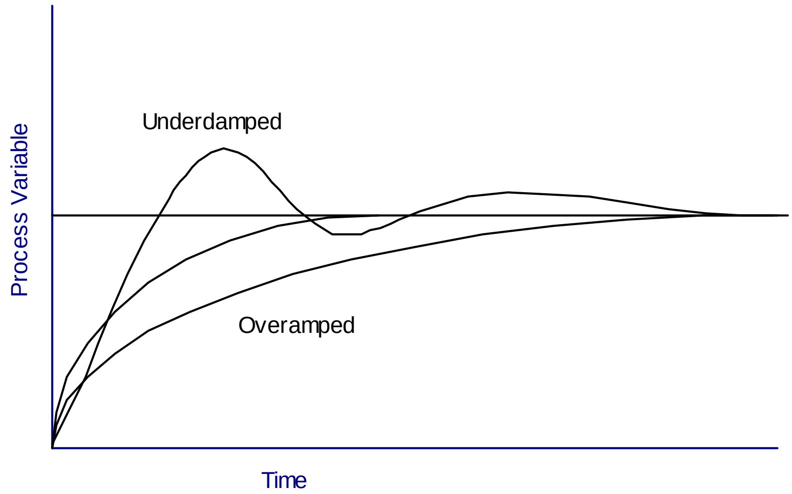


Figure (6): Typical responses to a set point change

The cyclic radian frequency can be related to the under-damped natural frequency ω_n and the damping coefficient ζ as follows:

$$\omega = \omega_n \sqrt{1 - \xi^2}$$

Overshoot

Overshoot is the amount by which the response exceeds the steady-state final value. Referring to Figure (7), the overshoot is defined as:

$$\frac{B}{A} = e^{-\pi\xi(1-\xi^2)}$$

Decay ratio

Decay ratio is the amplitude of an oscillation to the amplitude of the preceding oscillation, (C/B in Figure 7). The quarter decay ratio (QDR) which lies between critical damping and under-damping:

$$QDR = \frac{C}{B} = \frac{1}{4}$$

The QDR is often used to establish whether the controller is providing a satisfactory response. It has been shown through experience that the QDR provides a good trade off between minimum deviation from the set point and fastest return to the set point. For a second order system it can be shown that:

$$\frac{C}{B} = \sqrt{\mathbf{exp}(-2\pi(1-\xi^2))}$$

Rise Time

The rise time is the time required by the transient response to reach the final steady-state value.

Response time

Is the time required for the response to settle within the specified arbitrary limits. These limits are typically set as $\pm 3-5\%$ of the process variable steady state value.

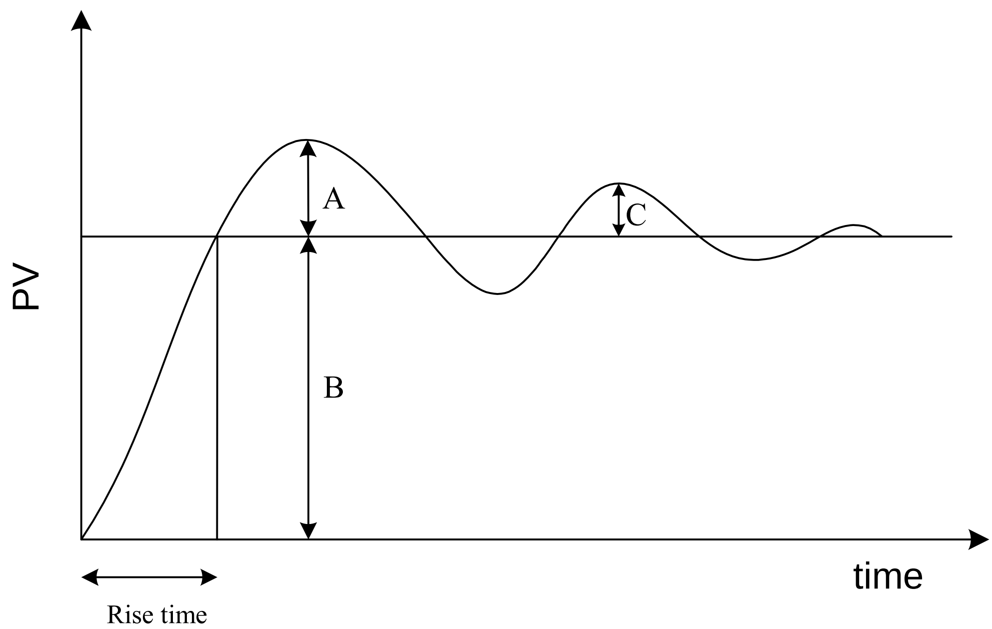


Figure (7) Second or higher order typical response to a set point change

4.1 Error Performance criteria

The previous simple criteria use only few points in the response and therefore are simple to use. The following criteria are based on the entire response of the process.

Integrated error (IE)

The integrated error is defined as:

$$IE = \int_0^{\infty} e dt$$

The controller parameters are selected to minimize the IE criteria. But, the designer should bare certain situations where the positive and negative deviations from the set point cancel each other.

Integrated absolute error (IAE)
Both negative and positive errors will be considered. The IAE is defined as:

$$IAE = \int_0^\infty |e|\,dt$$

Integrated square error (ISE)
This error uses the square of the error, thereby penalizing large errors more than small errors. This gives more conservative response (faster return to set point).

$$ISE = \int_0^\infty e^2\,dt$$

Integrated time absolute error (ITAE)
This criterion is based on the integral of the absolute value of the error multiplied by time. It results in errors existing over time being penalized even though may be small, which results in a more heavily damped response.

$$ITAE = \int_0^\infty t\,|e|\,dt$$

Figure (8) shows the various responses of a loop that is tuned to the above criteria.

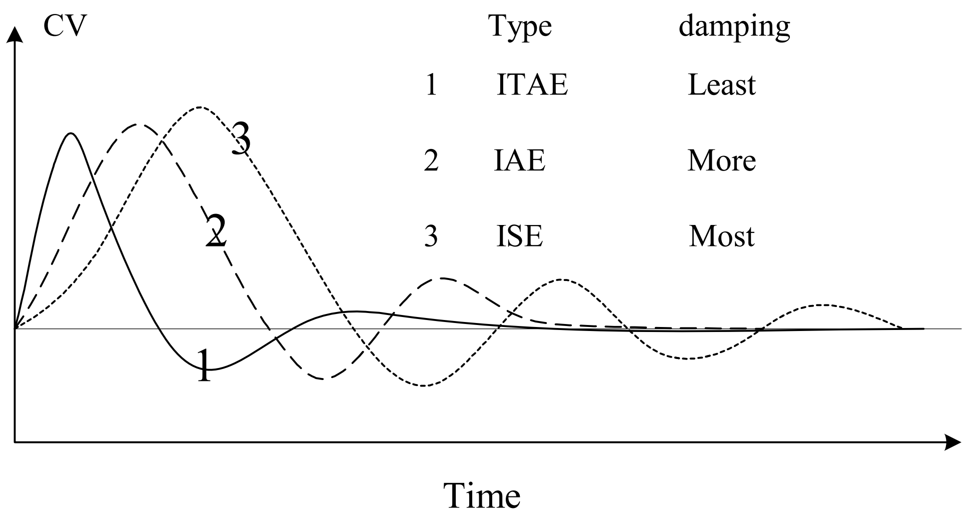


Figure (8) Responses to various error criteria

5. Tuning methods

5.1 Trial and error method

Guidelines:

1. Proportional action is the main control. Integral and derivative actions are used to trim the response.
2. The starting point is always with the controller gain, integral action and derivative action all at a minimum.
3. Make adjustments in the controller gain by using a factor of two.
4. Use the QDR criteria to determine the optimal response.
5. When in trouble (for example, unstable behavior) decrease the integral and derivative actions to a minimum and adjust the controller gain for stability.

It is important to know that controller parameters are strongly dependent on the individual process. When dealing with a flow control loop, P-only control can be used with a low control gain. For accuracy, PI controller can be used with a low gain and high integral action. Derivative actions can not be used for flow loops because such process has a very fast dynamics and flow measure met is inherently noisy.

Levels represent material inventory that can be used as a surge capacity to dampen disturbances. Hence, loosely tuned P-only control is sometimes used. However, some operators don not like offset, so PI controllers are typically used. Temperature dynamic responses are usually fairly slow, so PID control is used.

5.2 Process reaction curve method

In this method a process reaction curve is generated in a response to a disturbance. This curve is then used to calculate the controller gain, integral time and derivative time. These methods are performed in open loop so no control action occurs. To generate this a curve, the process is allowed to reach steady state. Then, a small disturbance (step input) is introduced and the reaction of the process variable is recorded. Figure (9) shows a typical process reaction curve for a generic self-regulating process. The process parameters than may be obtained from this curve are as follows:

θ = lag time (min)
τ = time constant estimate (min)
P = initial step disturbance (%)
ΔCp = change in process variable (PV) to step disturbance
N = $\Delta Cp/\tau$ = reaction rate (%)
R = $(N\ \theta)/\ \Delta Cp$ = lag ratio

Based on the process reaction curve parameters, three tuning methods are presented:

Ziegler-Nichols (ZN) open loop rules
The ZN open loop recommended controller parameters for quarter decay ratio are as follows:

P-only $$k_c = \frac{P}{N\theta}$$

PI $$k_c = 0.9\left(\frac{P}{N\theta}\right), \qquad \tau_I = 3.33\theta$$

PID $$k_c = 1.2\left(\frac{P}{N\theta}\right), \qquad \tau_I = 2\theta, \qquad \tau_d = 0.5\theta$$

These setting should be taken as recommendations only and tested thoroughly in closed loop. Note that the ratio $\Delta Cp/P$ is the process gain, i.e. k_p. Note also that the integral time T_i and derivative time T_d is replaced (in the above and following tables) by τ_I and τ_d, respectively.

Cohen-Coon open loop
The Cohen-Coon (C-C) recommendations correct for one deficiency in the ZN open loop rules. This is the sluggish closed loop response given by the ZN in the relatively rare occasions when process with dead time is large relative to the dominant time constant. The C-C settings are as follows:

P-only $$k_c = \frac{P}{N\theta}(1 + \frac{R}{3})$$

PI $$k_c = \frac{P}{N\theta}(0.9 + \frac{R}{12}) \qquad \tau_I = \theta\left(\frac{30 + 3R}{9 + 20R}\right)$$

PID $$k_c = \frac{P}{N\theta}(0.33 + \frac{R}{4}) \qquad \tau_I = \theta\left(\frac{32 + 6R}{13 + 8R}\right) \qquad \tau_d = \theta\left(\frac{4}{11 + 2R}\right)$$

As with the Z-N open loop method, the C-C recommendations should be tested in closed loops and adjust accordingly to the QDR.

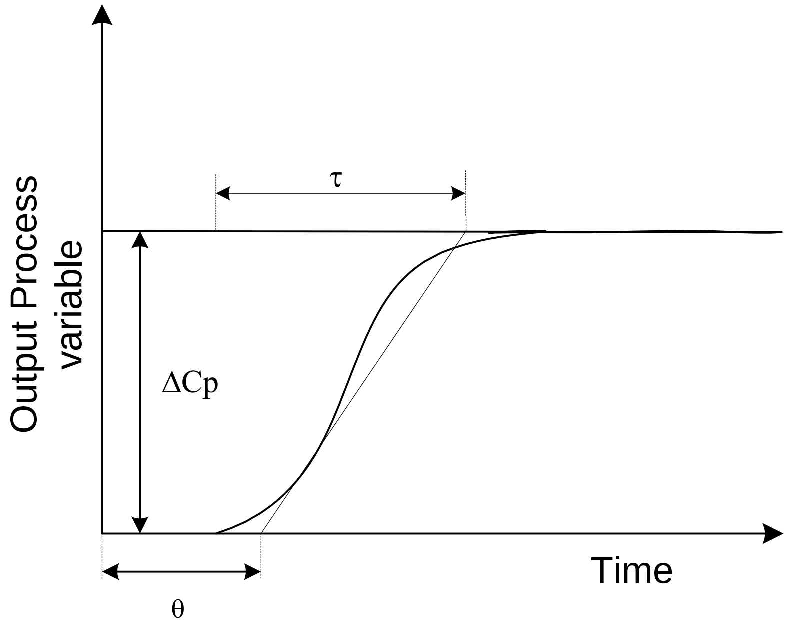


Figure (9) Process reaction curve

Internal model control (IMC) tuning rules

Many practitioners have found that the Z-N open loop and C-C rules are too aggressive for most chemical processes since they give a large controller gain and short integral time. Tuning rules based on the internal model of the process were developed with robustness in mind. These rules were related directly to the closed loop time constant and the robustness of the control loop. As a sequence, the closed loop step load response exhibits no oscillations or overshoot. The following simplified IMC rules were developed by Fruehauf et. al. (1993) for PID controller tuning.

	$\frac{\tau}{\theta} > 3$	$\frac{\tau}{\theta} < 3$	$\frac{\tau}{\theta} < 0.5$
k_c	$\frac{P}{2N\theta}$	$\frac{P}{2N\theta}$	$\frac{P}{N\theta}$
τ_I	5θ	τ	4
τ_d	$\leq 0.5\theta$	$\leq 0.5\theta$	$\leq 0.5\theta$

5.3 Constant Cycling Method

Ziegler-Nichols closed loop method

The closed-loop technique of Ziegler and Nichols is a technique that is commonly used to determine the two important system constants: ultimate period and ultimate gain. These are determined by disturbing the closed loop system and using disturbance response to extract the values of these constants. The following is a step-by-step approach to using Ziegler-Nichols closed loop method for the controller tuning:

1. Attach a P-only controller with low gain.

2. Increase the controller gain until a constant amplitude limit cycles occurs (Figure 10).
3. Determine the following parameters from the constant amplitude limit cycle:

   T_u = ultimate period = period taken from the limit cycle
   k_u = ultimate gain = controller gain that produces the limit cycle

4. calculate the tuning parameters using the following equations:

P-only	$k_c = k_u/2$		
PI:	$k_c = k_u/2.2$,	$\tau_I = T_u/1.2$	
PID	$k_c = k_u/1.7$,	$\tau_I = T_u/2$,	$\tau_d = T_u/8$

5. Fine-tune by adjusting the controller parameters.

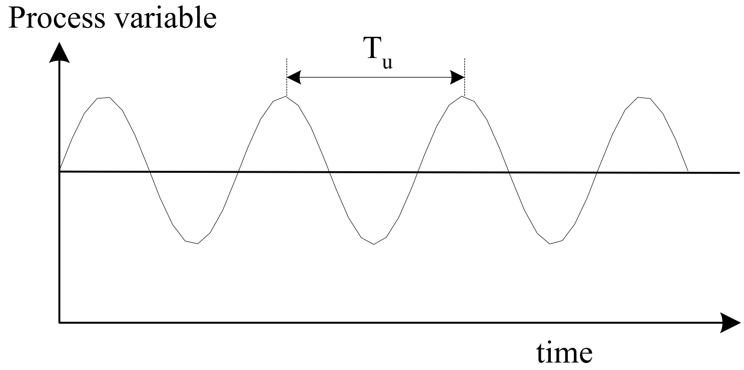


Figure 10: Continuous cycling response

6. **Guidelines for common control loops**

Flow Control

Flow and liquid pressure control loops (Figure 11) are characterized by fast responses (seconds), with essentially no time delay. The process dynamics are due to compressibility (in a gas stream) or inertial effects (in a liquid). The sensors and signal transmission line may introduce sufficient dynamic lag if pneumatic instruments are used. Disturbances in flow-control system tend to be frequent but generally not of large magnitude. Most of the disturbances are of high-frequency noise due to stream turbulence, valve change, and pump vibration. PI flow controllers are generally used with intermediate values of the controller gain Kc. The presence of high frequency noise rules out the use of derivative action.

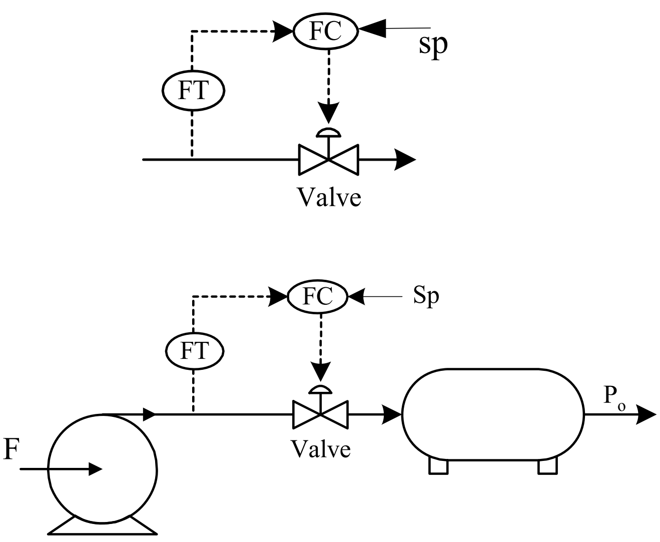


Figure (11) Flow and liquid pressure control loops

Liquid Level

Figure (12) shows two control loops used to control the liquid level. Because of the liquid level process integrating nature, a relatively high-gain controller can be used with little concern about instability of the control system. Integral action is normally used but is not necessary if small offsets in the liquid level can be tolerated. Derivative action is not normally used, since the level measurements often contain noise due to splashing and turbulence of the liquid entering the tank.

In many level control problems, the liquid storage tank is used as a surge tank to damp out fluctuations in its inlet streams. If the exit flow rate from the tank is used as the manipulated variable, then conservative controller setting should be applied to avoid large, rapid fluctuations in the exit flow rate (this strategy is called as averaging control). If the level also involves heat transfer such as in an evaporator, the process model and controller design become much more complicated.

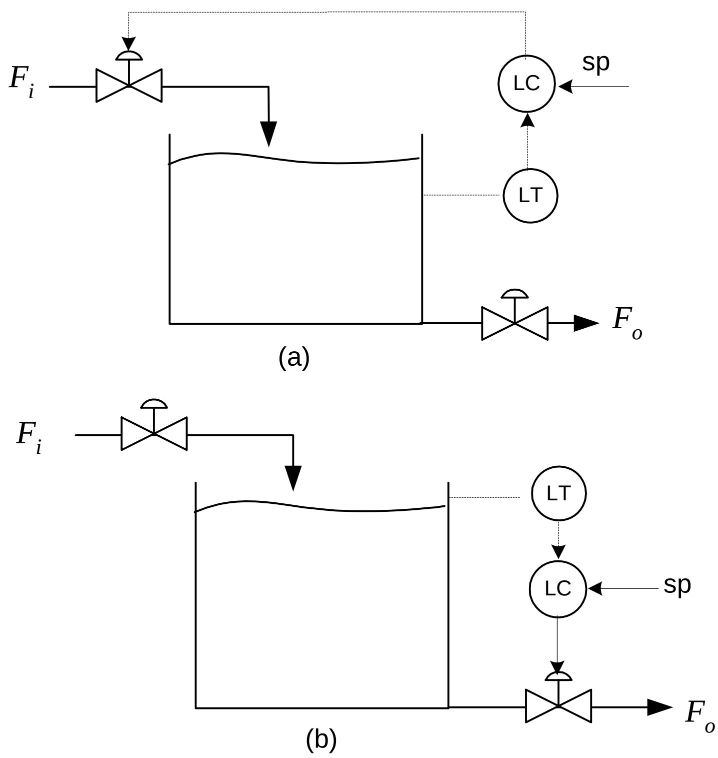


Figure (12) Level Control

Gas Pressure

Gas pressure is relatively easy to control, except when the gas is in equilibrium with a liquid. A gas pressure process is self-regulating: the vessel or pipeline admits more feed when the pressure is too low, and reduce the intake when the pressure becomes too high. PI controllers are normally used with only a small amount of integral action.

Temperature

General guidelines for temperature control loops are difficult because of the wide variety of processes and equipment involving heat transfer. For example, the temperature control problems are quite different for heat exchangers, distillation columns, chemical reactors, and evaporators. Due to the presence of time delays and/or multiple thermal capacitances, there will be usually a stability limit in the controller gain. PID controllers are commonly employed to provide more rapid responses than can be obtained with PI controllers.

Temperature control loops can be divided into two main categories:

1.Endothermic – require heat energy
2.Exothermic – generating heat energy.

Both of these processes have similar characteristics in that they are typically comprised of one large and many small capacities (valve, transmitter,…). The net result is a response of a process with a dominant capacitance plus dead time. It is very important to select the measuring device which adds a minimum lag to the process lag.

The exothermic reactor is perhaps the most difficult process to control due to its instability and extreme nonlinear response. A control scheme for an exothermic chemical reactor is shown in Figure (13). The degree of stability that can be achieved in this temperature control loop depends on the rate at which the heat can be removed from the reactor. The reactor can be stabilized if the reaction temperature changes fairly slowly when compared to the rate at which the jacket temperature changes. The idea of the control loop shown in Figure 13 is that once the feedstock and catalyst are added, hot water in the jacket is used to initiate the reaction. As the reaction temperature increases, the controller output decreases, closing the hot water valve, and opening the cold water valve. Multiple water inputs to the jackets can be used to minimize dead time and to change the jacket temperature as quickly as possible (to minimize the time constant of the jacket). Typically a PID controller is used, but using a proportional only controller may stabilize the reactor provided that the reactor is the dominant single capacitance in the loop and there is no appreciable dead time.

A good example of an endothermic process is a heat exchanger being used to heat a fluid from the inlet temperature Ti to an outlet temperature To. This heat exchanger response will be that of single large dominant capacitance with dead time. Typically either a PI or PID controller is used. Derivative action can be used since the temperature measurement is not noisy.

Composition

Composition loops generally have characteristics similar to temperature loops, but with several differences: (1) measurement noise is a more significant problem in composition loops, and (2) time delays due to analyzers are significant factors. These factors can limit the effectiveness of the derivative action. Due to their importance and the difficulty to control, composition and temperature loops often are prime candidates for the advanced control strategies discussed in the following lectures.

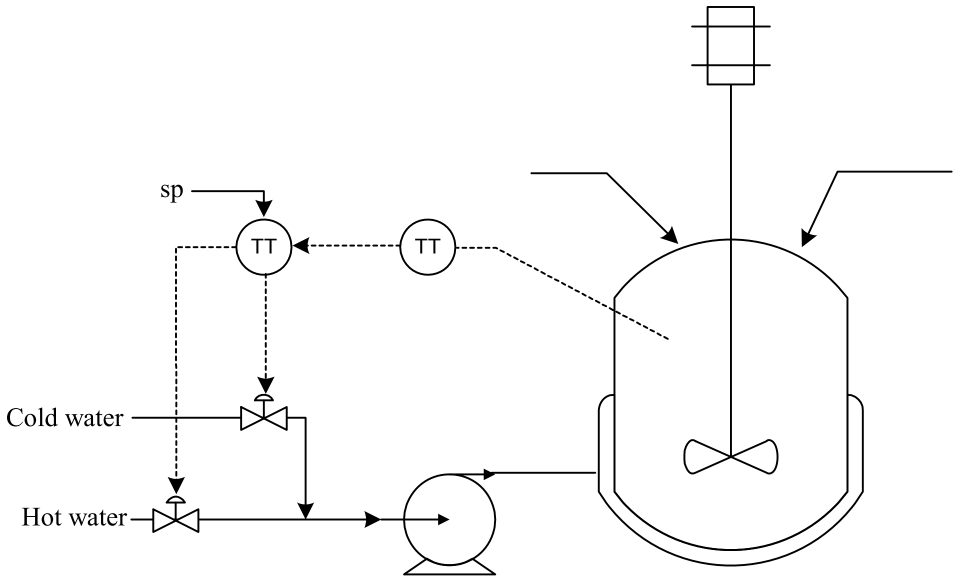


Figure (13) Control scheme for an exothermic reactor

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