2.36 Tuning Level Control Loops

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INTRODUCTION

Liquid level control loops, while among the most common control loops, have some unique and very distinctive characteristics:

- Liquid level is usually not a self-regulating process, but an integrating one.
- Intuitive general rules of thumb used for tuning, such as "If it's cycling, reduce the gain." do not apply and will often produce the opposite effect.
- Liquid level control loops, if once properly tuned, do not usually need retuning, although they may *appear* to go out of tune due to the onset of valve sticking.

This section first presents an approach to level control tuning, which is based on an idealized process model. Many actual level control loops can be approximated by this idealized process model. Later in the section, some of the characteristics of nonidealized level control systems are considered.

Most processes require extensive testing to obtain even their approximate process models. Most liquid level control loops, however, readily yield to an analytical approach because their process models can be formulated, desired performance parameters established, and from this their controller tuning parameters can be calculated. Once this is done, other attributes of the control loop such as the period of oscillation can be predicted.

The counterintuitive nature of level controllers makes their tuning by trial-and-error techniques difficult. On the other hand, the determination of their tuning parameters by analytical means is a natural choice and for that reason liquid level control loops should be engineered, not tuned.

In the following paragraphs, the discussion will start with an ideal model. After that, some of the nonideal characteristics of real installations and worst-case conditions (even if they are very unlikely to occur) will be discussed.

IDEALIZED MODEL

An idealized liquid level control system is shown in Figure 2.36a. Attributes of this idealized system are:

- The tank has constant cross-sectional area.
- The level controller is cascaded to a flow controller.
- A valve positioner is installed on the flow control valve.
- All inflow goes to outflow; the tank is merely a buffer storage tank.
- The maximum outflow is the same as the maximum inflow.
- The tank is of significant size relative to the flow rates.
- There is no thermal effect such as in the case of boiler drum level control.
- The level controller set point is constant.

The consequences of the above attributes are:

- The level process is linear.
- Up- or downstream pressure, line loss, or pump curve have no effect on loop behavior.

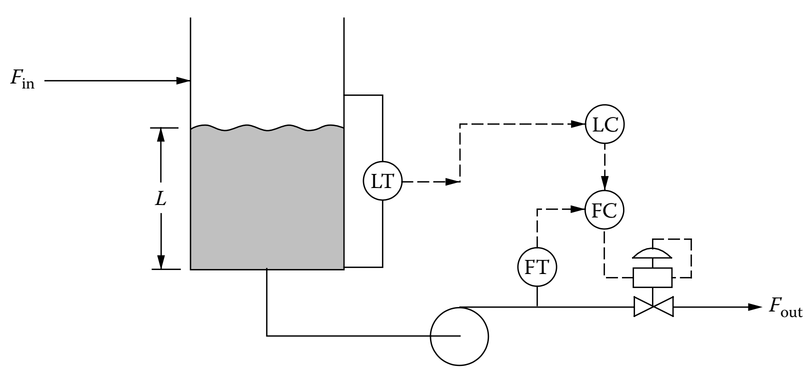

FIG. 2.36a
Liquid level control of an ideal process.

- The control loop performance is not affected by valve size.
- There is no dead time in the loop.
- The speed of response of the flow loop is significantly faster than that of the level in the tank; therefore, the dynamics can be ignored.
- Response to set-point change need not be considered because set-point changes are rarely made. The critical consideration is the response of the loop to load disturbances.

Figure 2.36a describes a common installation where the level controller manipulates the outflow from the tank in response to changes in inflow. The level controller can also manipulate the inflow in response to varying demands for the outflow. The discussion in the following paragraphs is applicable to both cases.

Time Constant of the Tank

A key parameter for the analysis of this process is the tank holdup time, also called the *tank time constant.* If the tank geometry (diameter and distance between the level taps) and maximum outflow rate (flow rate corresponding to 100% of the level measurement span) are known, then the tank time constant can be calculated as

$$T_L = \frac{Q}{F} \tag{2.36(1)}$$

where T_L = tank time constant; Q = tank hold up quantity, between upper- and lower-level sensor taps; and F = maximum flow rate. Q and F should be in compatible units, such as "gallons" and "gallons per minute," in which case the time constant will be in minutes.

A block diagram of the control loop with a PI controller is shown in Figure 2.36b. If the loop operates at constant set point, then the response to a load disturbance (i.e., to a change in the inflow F_{in}) is of more interest, and the set-point response can be neglected. However, in addition to the response of the level to a change in F_{in}, there is also interest in the response of the outflow, F_{out}, to a change in. F_{in}. The transfer functions of these two responses can be derived from Figure 2.36b:

$$\frac{L(s)}{F_{in}(s)} = \frac{\dfrac{s}{T_L}}{s^2 + \dfrac{K_C}{T_L}s + \dfrac{K_C}{T_I T_L}} \tag{2.36(2)}$$

$$\frac{F_{out}(s)}{F_{in}(s)} = \frac{\dfrac{K_C}{T_L}s + \dfrac{K_C}{T_I T_L}}{s^2 + \dfrac{K_C}{T_L}s + \dfrac{K_C}{T_I T_L}} \tag{2.36(3)}$$

where K_C is the controller gain; T_I is the integral time of the controller in minutes per repeat; and T_L is the time constant of the tank in minutes.

According to these equations, the loop acts as a second-order system. These transfer functions can also be written using the traditional servo-mechanism terminology as

$$\frac{L(s)}{F_{in}(s)} = \frac{\dfrac{s}{T_L}}{s^2 + 2\zeta\omega_n s + \omega_n^2} \tag{2.36(4)}$$

$$\frac{F_{out}(s)}{F_{in}(s)} = \frac{2\zeta\omega_n s + \omega_n^2}{s^2 + 2\zeta\omega_n s + \omega_n^2} \tag{2.36(5)}$$

where ζ, the damping factor, and ω_n, the natural frequency, are given by

$$\zeta = \frac{1}{2}\sqrt{\frac{K_C T_I}{T_L}} \tag{2.36(6)}$$

$$\omega_n = \sqrt{\frac{K_C}{T_I T_L}} \tag{2.36(7)}$$

The damping factor is a dimensionless number; the natural frequency is in radians per minute, if T_I is in minutes per repeat and T_L is in minutes.

Many practicing engineers may be more familiar with the term decay ratio (DR), rather than the damping

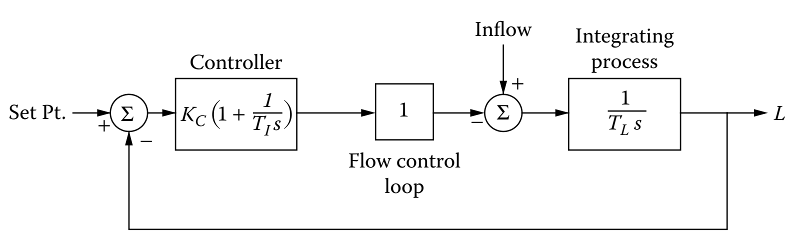

FIG. 2.36b
Block diagram of a liquid level control loop for an ideal process.

TABLE 2.36c
Equations for Calculating the Tuning Parameters of a PI Level Controller for an Ideal Level Process

Tuning Parameter	Underdamped $\zeta < 1$ — Rigorous	Underdamped $\zeta < 1$ — Simplified	Critically Damped $\zeta = 1$
K_C	$2\zeta e^{-\zeta f(\zeta)}\left(\frac{\Delta F_{in}}{\Delta L_{max}}\right)$	$2\zeta e^{-\zeta f(\zeta)}\left(\frac{\Delta F_{in}}{\Delta L_{max}}\right)$	$\frac{2}{e}\left(\frac{\Delta F_{in}}{\Delta L_{max}}\right)$ $(e = 2.71828\cdots)$
T_I	$2\zeta e^{\zeta f(\zeta)}\left(\frac{T_L \Delta L_{max}}{\Delta F_{in}}\right)$	$4\zeta^2\left(\frac{T_L}{K_C}\right)$	$4\left(\frac{T_L}{K_C}\right)$

factor, ζ. The damping factor and decay ratio are related as follows:

$$\zeta = \frac{-\ln(\mathrm{DR})}{\sqrt{4\pi^2 + (\ln(\mathrm{DR}))^2}} \qquad \textbf{2.36(8)}$$

$$DR = \exp\left(\frac{-2\pi\zeta}{\sqrt{1-\zeta^2}}\right) \qquad \textbf{2.36(9)}$$

For example, the familiar quarter-amplitude decay ratio in terms of damping factor is $\zeta = 0.215$.

Determining Tuning Parameters

In addition to knowing the tank hold-up time, T_L (see Equation 2.36[1]), the analytical determination of tuning parameters requires choosing values for three design parameters:

1. ΔF_{in}—The maximum anticipated step change in disturbance (inflow) that can be expected, in percent of full scale measurement of the inflow
2. ΔL_{max}—The maximum allowable deviation from set point, in percent of full scale level measurement, resulting from a step disturbance of size ΔF_{in}
3. DR—the desired decay ratio after such a step disturbance

Once the above three values have been determined, a value for T_L can be obtained and Equation 2.36(8) can be used to convert decay ratio to damping factor. This will then permit one to derive the equations tabulated in Tables 2.36c, 2.36d, and 2.36e.

Table 2.36c provides equations to calculate tuning parameters, while Tables 2.36d and 2.36e give equations for the calculation of various characteristics of the level in the tank and of the response to a step change in inflow. In Tables 2.36c, 2.36d, and 2.36e:

$$f(\zeta) = \frac{1}{\sqrt{1-\zeta^2}} \tan^{-1}\frac{\sqrt{1-\zeta^2}}{\zeta} \qquad \textbf{2.36(10)}$$

TABLE 2.36d
Equations for Calculating Some of the Predicted Behavior Attributes of Control on an Ideal Level Process

Behavior Attribute	Underdamped $\zeta < 1$ — Rigorous	Underdamped $\zeta < 1$ — Simplified	Critically Damped $\zeta = 1$
Arrest time—T_{aL}	$f(\zeta)e^{\zeta f(\zeta)}\left(\frac{T_L \Delta L_{max}}{\Delta F_{in}}\right)$	$\frac{f(\zeta)}{2\zeta}T_I$	$\frac{T_I}{2}$
Period—P	$\frac{2\pi}{\sqrt{1-\zeta^2}}e^{\zeta f(\zeta)}\left(\frac{T_L \Delta L_{max}}{\Delta F_{in}}\right)$	$\frac{\pi}{\zeta\sqrt{1-\zeta^2}}T_I$	N/A
IAE	$\left(\frac{1+e^{\frac{-\zeta\pi}{\sqrt{1-\zeta^2}}}}{1-e^{\frac{-\zeta\pi}{\sqrt{1-\zeta^2}}}}\right)e^{2\zeta f(\zeta)}\left(\frac{T_L(\Delta L_{max})^2}{\Delta F_{in}}\right)$	Same as ←	$e^2\left(\frac{T_L(\Delta L_{max})^2}{\Delta F_{in}}\right)$

TABLE 2.36e
Equations for Calculating the Predicted Behavior Attributes of Level Controls

Behavior Attribute	*Underdamped* $\zeta < 1$ — *Rigorous*	*Underdamped* $\zeta < 1$ — *Simplified*	*Critically Damped* $\zeta = 1$
Maximum change in outflow $\Delta F_{\text{out-max}}$	$(1+e^{-2\zeta f(\zeta)})\Delta F_{\text{in}}$	$(1+e^{-2\zeta f(\zeta)})\Delta F_{\text{in}}$	$(1+e^{-2})\Delta F_{\text{in}}$
Arrest time—T_{aF}	$2\,f(\zeta)e^{\zeta f(\zeta)}\left(\dfrac{T_L\,\Delta L_{\max}}{\Delta F_{\text{in}}}\right)$	$2T_{\text{aL}}$	$2T_{\text{aL}}$
Max rate of change of outflow $\left(\dfrac{dF_{\text{out}}}{dt}\right)_{\max}$	$\zeta \le \frac{1}{2}$: $\left[\exp\left(-\zeta f(\zeta) - \dfrac{1}{\sqrt{1-\zeta^2}}\tan^{-1}\left(\dfrac{(1-4\zeta^2)\sqrt{1-\zeta^2}}{\zeta(3-4\zeta^2)}\right)\right)\right]\left(\dfrac{(\Delta F_{\text{in}})^2}{T_L\,\Delta L_{\max}}\right)$	$\zeta \le \frac{1}{2}$: Same as ←	
	$\frac{1}{2} < \zeta < 1$: $2\zeta e^{-\zeta f(\zeta)}\left(\dfrac{(\Delta F_{\text{in}})^2}{T_L\,\Delta L_{\max}}\right)$	$\frac{1}{2} < \zeta < 1$: Same as ←	$\dfrac{2}{e}\left(\dfrac{(\Delta F_{\text{in}})^2}{T_L\,\Delta L_{\max}}\right)$

In the three tables, the columns labeled "rigorous" provide equations that are entirely a function of the following four fixed or chosen parameters—T_L, ζ (as determined from the chosen decay ratio), ΔF_{in}, and $\Delta L_{\max}$. The column labeled "simplified" produces the same results, calculated in terms of a previously calculated quantity.

Once the tuning parameters have been calculated, the predicted behavior of the level and the outflow can be calculated from Tables 2.36d and 2.36e. In Table 2.36d, the level arrest time, T_{aL}, is the time period beginning at the disturbance and ending when the maximum deviation from set point is reached.

In Table 2.36e, the outflow arrest time, T_{aF}, is the time period that begins with the disturbance and ends when the maximum change in outflow is reached. The maximum rate of change of outflow is provided because it is this quantity, rather than the size of the outflow change itself, that can be the maximum disturbance to a downstream process unit.

The equations given in Tables 2.36c, 2.36d, and 2.36e describe this tuning technique, but they are not very useful due to the large amount of computation required. For three specific decay ratios, Tables 2.36g and 2.36h provide equations for calculating tuning settings, level and outflow parameters. The three decay ratios chosen are: 1) critically damped, 2) quarter-decay ratio and 3) 1/20 decay ratio.

The critically damped decay ratio is chosen because it is a recognized basis for tuning. Quarter amplitude damping is chosen because of its familiarity. The third response, although less familiar, is chosen because it provides both the minimum IAE and the lowest maximum rate of change in outflow. Figure 2.36f depicts the level responses when the level controllers are tuned on the basis of these three forms of responses with equal values of maximum deviation.

Tables 2.36g and 2.36h also provide correction factors to account for two real-world phenomena:

1. The use of non-cascade control
2. The presence of dead time in the control loop

In the equations, θ is the ratio,

$$\frac{\text{Dead time}}{T_L}$$

and K_V is the valve gain. The correction factors were determined as a "best fit" to simulation results, for the values of θ between 0.0 and 0.5.

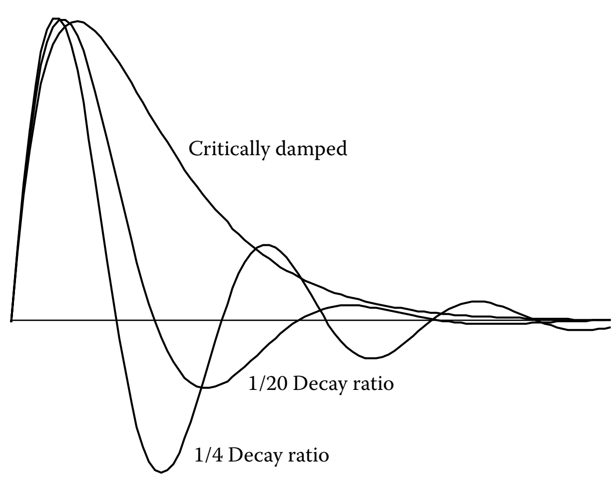

FIG. 2.36f
Response of the levels to equal maximum deviation when the controllers are tuned for the noted three decay ratios.

TABLE 2.36g
Working Equations for Calculating Tuning Parameters for a PI Level Controller for a Process with Dead Time and Non-Cascade Control

ΔF_{in} = max. step change in disturbance
L_{max} = max. allowable deviation of level from set point
T_L = hold up time, minutes

$\theta = \frac{\text{Date Time}}{T_L}$ $(0 \le \theta \le 0.5)$
K_V = Valve Gain, if non-cascade
= 1.0 if level is cascaded to flow

Decay Ratio *DR*	*Damping Factor* ζ	*Gain* K_C	*Integral Time* T_I
Critically damped	1.0	$\frac{0.74\,\Delta F_{in}}{(1-\theta)^{0.5}\,K_V\,\Delta L_{max}}$	$\frac{4.0\,T_L}{(1-\theta)^{0.5}\,K_V K_C}$
0.05	0.430	$\frac{0.5\,\Delta F_{in}}{(1-\theta)\,K_V\,\Delta L_{max}}$	$\frac{0.74\,T_L}{(1-\theta)^{1.25}\,K_V K_C}$
0.25	0.215	$\frac{0.32\,\Delta F_{in}}{(1-\theta)^{1.5}\,K_V\,\Delta L_{max}}$	$\frac{0.19\,T_L}{(1-\theta)^{2.4}\,K_V K_C}$

Example

Assume that a tank has the following specifications:

Tank diameter 5.0 feet
Distance between level transmitter taps 8.0 feet
Maximum outflow (upper end of outflow transmitter span) 250 gpm

Assume also that the level controller is the cascade master of a flow controller (valve gain K_V = 1.0), and that there is no dead time in the loop (the dead-time/T_L ratio = θ = 0). The tank holdup time is calculated by:

$$\text{Surge volume} = \frac{\pi}{4} \times 5^2 \times 8 = 157.3\ \text{ft}^3$$

$$\text{Surge quantity} = Q = 157.3\,\text{ft}^3 \times 7.48\,\frac{\text{gal}}{\text{ft}^3} = 1176.6\ \text{gal}$$

$$\text{Hold up time} = T_L = \frac{1176.6}{250} = 4.7\ \text{min}$$

Also, assume that the worst-case disturbance is anticipated to be a step change in inflow of 10%. In the event of this disturbance, the maximum level deviation should not exceed 5% (about 5 inches) and the system should settle out rapidly, so a decay ratio of 0.05 is chosen.

$$\Delta F_{in} = 10\%$$

$$\Delta L_{max} = 5\%$$

With these data, Table 2.36g can be used to calculate tuning parameters:

$$K_C = \frac{0.50 \times 10}{5} = 1.0$$

$$T_I = \frac{0.74 \times 4.7}{1.0} = 3.45\ \text{min/repeat}$$

Tables 2.36h and 2.36i can be used to predict the level and outflow response:

TABLE 2.36h
Working Equations for Calculating Predicted Behavior Attributes of Level Response to a Step Change in Inflow, Level Control Loop with Dead Time and Non-Cascade Control

Decay Ratio *DR*	*Level Arrest Time* T_{aL}	*Period* *P*	*IAE*
Critically Damped	$0.5(1-\theta)^{0.96}T_I$	Not Applicable	$7.39(1-\theta)^{0.52} \times T_L \frac{\Delta L^2_{max}}{\Delta F_{in}}$
0.05	$1.45(1-\theta)^{0.70}T_I$	$8.09(1-\theta)^{1.13}T_I$	$4.61(1-\theta) \times T_L \frac{\Delta L^2_{max}}{\Delta F_{in}}$
0.25	$3.22(1-\theta)^{0.47}T_I$	$14.93(1-\theta)^{1.56}T_I$	$5.45(1-\theta)^{0.54} \times T_L \frac{\Delta L^2_{max}}{\Delta F_{in}}$

TABLE 2.36i
Working Equations for Calculating Predicted Behavior Attributes of Outflow Response to a Step Change in Inflow; Level Control Loop with Dead Time and Non-Cascade Control

Decay Ratio *DR*	*Outflow* $\Delta F_{out\text{-}max}$	*Outflow Arrest Time* T_{aF}	*Outflow* *Max Rate of Change*
Critically Damped	$\frac{1.14\Delta F_{in}}{(1-\theta)^{0.1}}$	$1.0\,(1-\theta)^{1.26}T_I$	$\frac{0.74}{(1-\theta)^{0.70}}\times\frac{\Delta F_{in}^2}{T_L\Delta L_{max}}$
0.05	$\frac{1.34\Delta F_{in}}{(1-\theta)^{0.22}}$	$2.50(1-\theta)^{1.36}T_I$	$\frac{0.52}{(1-\theta)^{1.37}}\times\frac{\Delta F_{in}^2}{T_L\Delta L_{max}}$
0.25	$\frac{1.55\Delta F_{in}}{(1-\theta)^{0.25}}$	$6.43(1-\theta)^{1.75}T_I$	$\frac{0.61}{(1-\theta)^{1.16}}\times\frac{\Delta F_{in}^2}{T_L\Delta L_{max}}$

Level arrest time (from time of disturbance to time when maximum deviation is reached):

$$T_{aL} = 1.45 \times 3.45 = 5.0 \text{ minutes}$$

$$\text{Period } P = 8.09 \times 3.45 = 27.9 \text{ minutes}$$

The period may seem to be excessive; however, since a fast settling behavior was selected (decay ratio of 0.05), the maximum deviation during the second half-cycle will be about 1.1 inches, during the third half-cycle about 0.25 inches, and so on.

NONIDEAL PROCESSES

The discussion so far has been based on an idealized process model. Many real applications will deviate from the idealized model. In the following paragraphs some commonly encountered situations will be described, along with suggested procedures for coping with them.

Irregular Vessel Shapes

For irregularly shaped vessels, such as horizontal or spherical tanks, the direct use of the level measurement signal as the input to the level controller may result in highly nonlinear process characteristics and undesirable control loop behavior. In this case, the loop can be linearized by converting the level measurement into the volumetric holdup in the vessel, which can be computed from the vessel geometry and the actual level. In this case, the converted measurement signal should be scaled in 0 to 100% of maximum volumetric holdup.

No Cascade Loop

If the slave flow control loop shown in Figure 2.36a is not provided, then the holdup time cannot be calculated from Equation 2.36(1), because the maximum outflow cannot be related to the maximum setting of a secondary controller.

To attempt to determine the maximum outflow rate when the outlet valve is wide open would probably be futile because of the unknown variables such as line loss, pump curve effects, and hydrostatic head effects in the tank. In addition, because the process response is nonlinear, the maximum outflow rate with a wide-open valve will vary with the level in the tank.

In this case, what needs to be determined is the apparent holdup time at the nominal operating point and the valve gain. The corresponding block diagram is shown in Figure 2.36j, and the required test is illustrated in Figure 2.36k.

To determine the apparent holdup time and the valve gain at the actual operating point, the operation must be stable, the inflow and the level at the normal set point must be constant, and the controller must be in the automatic mode. In addition, the controller output should be within its extreme limits. It is necessary for inflow to remain constant during the test. The control valve modulating the outflow should have a positioner or at least be as “sticktion free” as possible.

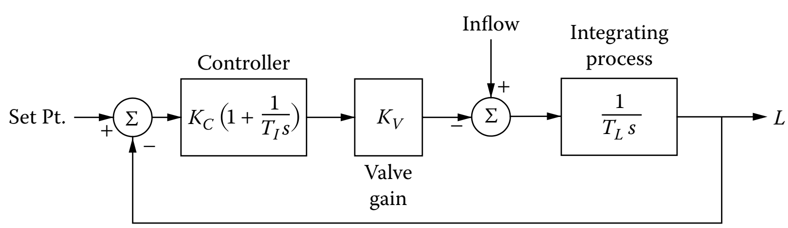

FIG. 2.36j
Block diagram of a conventional liquid level control loop, where there is no cascade slave controller and the level controller directly throttles the valve of an ideal process.

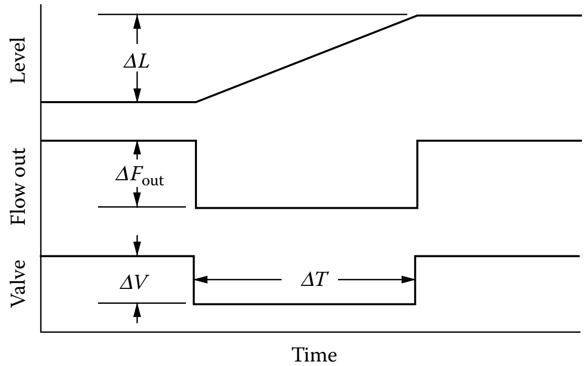

FIG. 2.36k
Illustration of a test to estimate the apparent holdup time and valve gain for noncascade level control.

The testing is done in the following sequence: First, switch the controller to manual and change the previously constant output signal by a small amount, say, $\Delta V\%$ (see Figure 2.36k). In response, the outflow will change by an amount ΔF and the level will start changing. After a certain period of time, say Δt, move the controller output back to where it was before the test. As a consequence, the level should stop changing. Once the level has stabilized, determine the change in level, ΔL, that occurred during the test.

Convert the readings for ΔL, ΔF_{out}, and ΔV into percent of full scale. The equation for estimating the apparent holdup time is

$$T_L = \frac{\Delta F_{out}\ \Delta t}{\Delta L} \qquad \textbf{2.36(11)}$$

and the valve gain is

$$K_V = \frac{\Delta F_{out}}{\Delta V} \qquad \textbf{2.36(12)}$$

Dead Time

The ideal level process has neither dead time nor lag time, but real processes can have either or both. For example, in case of the level-flow cascade loop in Figure 2.36a, the flow control loop may have a finite response time. If there is a dead time in the loop, θ, the ratio of dead time to holdup time should be calculated and used in the equations listed in Tables 2.36g, 2.36h, and 2.36i. Similarly, the actual valve gain (K_V) should be used in Table 2.36g (for cascade loops, $K_V = 1.0$).

Unequal In- and Outflows

When determining step changes in the inflow to a vessel, it should always be the actual flow that enters the tank. Hence the change in feed rate, ΔF_{in}, can be due to any cause, such as actual change in vessel feed rate, change in the liquid/vapor ration of the feed, change in reboiler heat, or change in liquid load on the trays in the distillation tower.

Flashing Liquids

There are cases where flashing liquids result in a false indication of level. An example is the "shrink and swell" effect in a boiler drum. The shrink and swell effect can be approximated as dead time; hence, the dead time correction factors in Tables 2.36g, 2.36h, and 2.36i may be used.

Sinusoidal Disturbance

If a sinusoidal variation in inflow is anticipated (for instance, due to cycling of a process controller at an upstream process unit), then the maximum variation in level and the amplitude of variation of outflow should be investigated. If the frequency of the sinusoidal variation is not known in advance, then a worst case condition could be assumed for the investigation, where the frequency of inflow disturbance in the same as the natural frequency of the level control loop.

An oscillating input will cause both the level and the outflow to oscillate with the same period. If the process can be approximated by the ideal model, which was defined at the beginning of this section, and if the inflow oscillates with a known amplitude and period (P_{in}), then Figures 2.36l and 2.36m can be used to determine the amplitude of both the level and of the outflow oscillations.

The undamped natural frequency, ω_n, is calculated from the damping factor, ζ, which is obtained from Table 2.36g, based on the chosen response (decay ratio) to a step disturbance and from the tuning parameters. If the process fits the ideal process model, Equation 2.36(7) can be used to calculate ω_n. Otherwise, the period, P, given by Table 2.36h, should be used to calculate ω_n from the equation below:

$$\begin{aligned} \omega_n &= \frac{2\pi}{P\sqrt{1-\zeta^2}} && \text{if DR} = 1/20 \text{ or } 1/4 \\ &= \frac{2}{T_I} && \text{if Critically Damped} \end{aligned} \qquad \textbf{2.36(13)}$$

From the period of oscillation of the input, P_{in}, the frequency in radians per minute can be calculated from:

$$\omega = \frac{2\pi}{P_{in}} \qquad \textbf{2.36(14)}$$

The frequency ratio $\frac{\omega}{\omega_n}$ in Figure 2.36l should be used to calculate $K_C \frac{L(\omega)}{F_{in}(\omega)}$. (If the frequency of the disturbance is unknown, then for a worst case analysis, use a frequency ratio of 1.0.) From this and from the amplitude of the disturbance oscillation, $F_{in}(\omega)$, one can calculate the predicted amplitude of oscillation of level about the set point, $L(\omega)$.

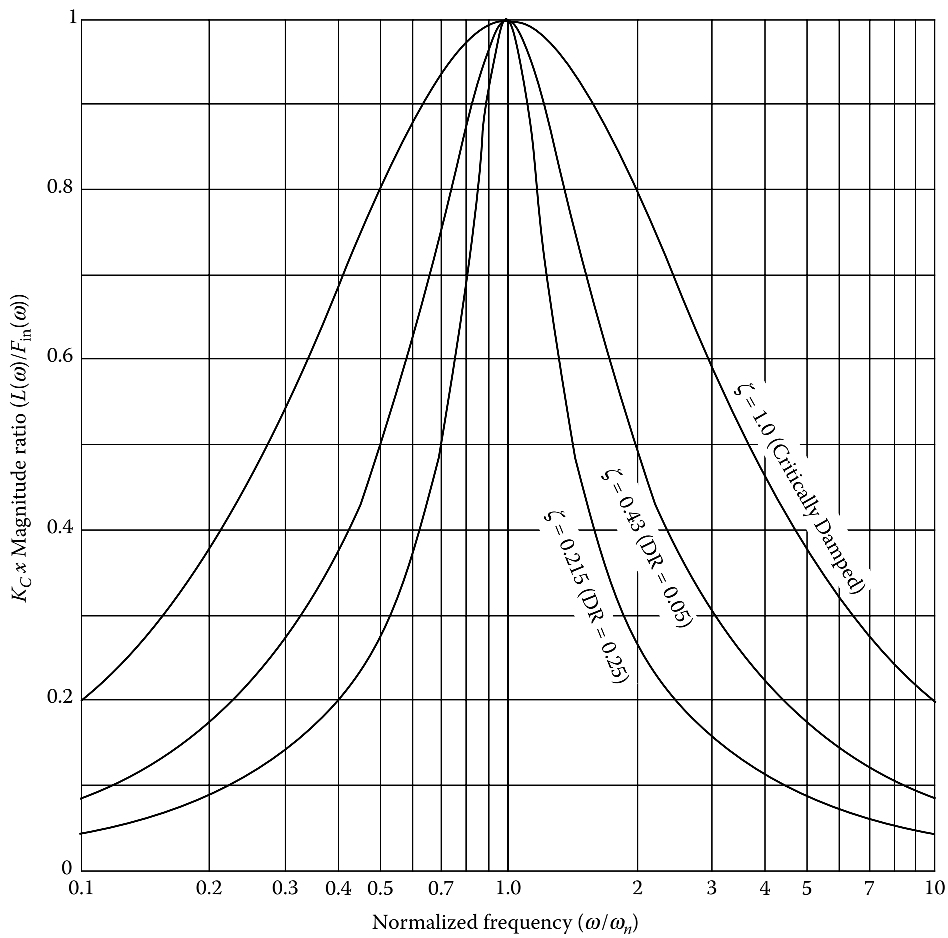

FIG. 2.36l
Magnitude ratio of changes in level to sinusoidal changes to inflow.

If the peak deviation from set point (one half the amplitude of oscillation) exceeds the allowable maximum deviation, then K_C should be increased by the relation:

$$K_{C\text{-new}} = \frac{0.5\,L(\omega)}{\Delta L_{\max}} \qquad \mathbf{2.36(15)}$$

After this, one can return to Tables 2.36g, 2.36h, and 2.36i and calculate T_I and the predicted attributes of the response.

If the magnitude and period of oscillation of the input are assumed or known, use Figure 2.36m to calculate the magnitude ratio of oscillations of outflow to inflow,

$$\frac{F_{\text{out}}(\omega)}{F_{\text{in}}(\omega)}$$

then calculate the amplitude of oscillation of the outflow.

As a final check, if there are level control loops in series, such as in case of a train of distillation towers, it is necessary to check the natural frequency of each tower. Ideally, the natural frequency of each tower should be no more than half the natural frequency of the preceding tower. Since the size of the towers (holdup time) cannot be changed, the best "handle" for adjusting the ratio of the natural frequencies is to increase $\Delta L_{\max}$ for the downstream tower or to decrease $\Delta L_{\max}$ for the tower upstream.

OTHER APPROACHES TO TUNING

Averaging Level Control

Many liquid level loops are not critical and one can tolerate fluctuation, even offset in levels, if it smooths out the flow to a downstream process unit. This can be accomplished by using a proportional-only controller. This technique is called "averaging liquid level control."[1] Assuming that the allowable excursions above and below set point are equal, then the controller gain should be set according to the equation

$$K_C = \frac{100}{2\,\Delta L_{\max}} \qquad \mathbf{2.36(16)}$$

or the proportional band should equal $2 \times \Delta L_{\max}$ and the bias (manual reset) should be set to 50%. With this arrangement, if the disturbance is such that a controller output of 50% is

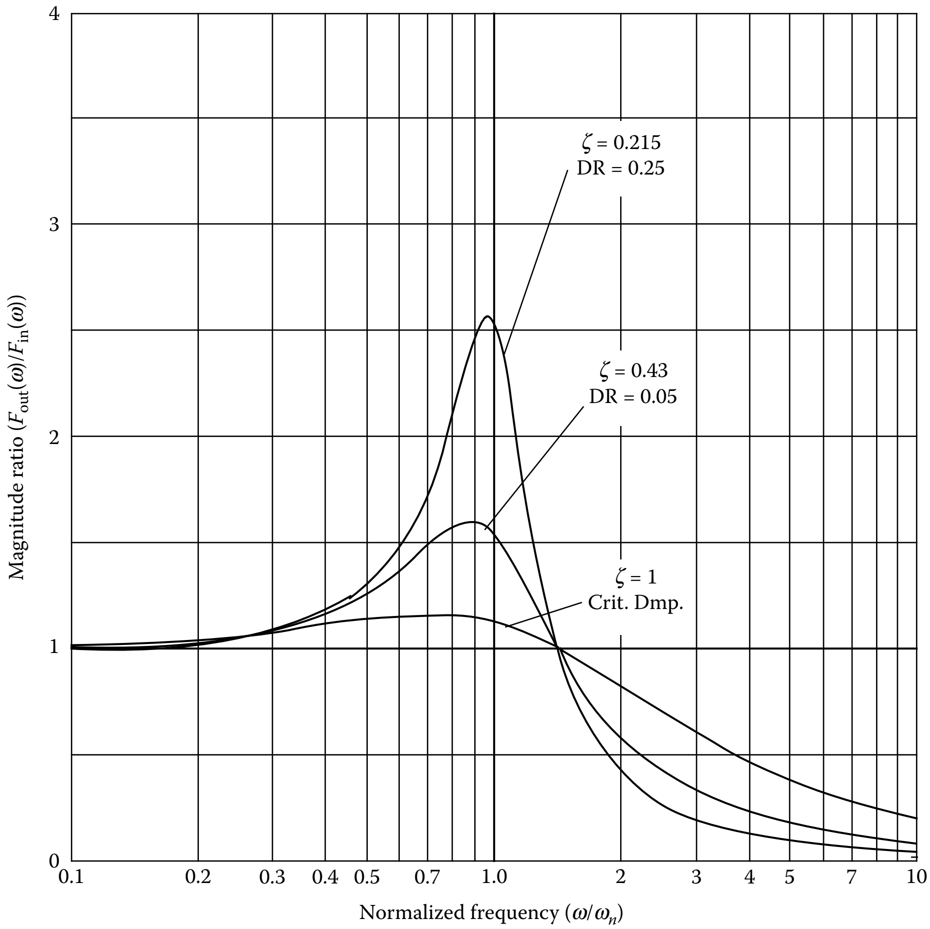

FIG. 2.36m
Magnitude ratio of changes in outflow to sinusoidal changes in inflow.

required, the level will be on set point. Otherwise, there will be a steady-state offset between set point and level measurement. When a step change in inflow occurs, the level will respond as a first-order lag with a time constant equal to the holdup time divided by the controller gain.

This technique ensures that the level never exceeds the limits because at the limits, the outflow is either at 0 or 100%. This technique also provides the lowest rate of change of the outflow, hence the minimum disturbance to a downstream processing unit. The disadvantage of this method is the fact that the level is rarely at set point. This is probably more of a disadvantage from the operator's acceptance point of view than any other.

It might appear that one could achieve the advantages of averaging level control and still maintain the level at the set point by using Equation 2.36(16) to determine the controller gain and then by using a slow reset action. Equation 2.36(6) shows, however, that a large value for T_I will produce a larger damping factor or even an overdamped response. If the response is underdamped, the loop will have a very long period; if overdamped, the return to set point will be excessive. Furthermore, with the reset action causing an effective shift of the proportional band, the positive protection of knowing exactly where the level will be when the controller output is at maximum or minimum is lost.

Controller Gain and Resonance

If it is desired to stay closer to the set point, the controller gain can be increased. Although there is no theoretical upper limit for the gain of a proportional-only controller used on an integrating process, in practice the gain will be limited by resonance that may occur within the loop.

If the level sensor is of the external cage type, then there may be a manometer effect between the liquid in the vessel and the liquid within the level sensor cage. This will appear as an oscillation within the control loop even when the total mass holdup is unchanging. If there is a large surface area of the liquid, this may result in a resonant sloshing, with a period proportional to the cross-sectional areas.

For a probe or a point-source sensor, this will also show up as an oscillation within the loop. Furthermore, splashing, such as from upper trays in a distillation tower, may result in the appearance of noise on the level measurement. Thus there will be a practical limit to the controller gain. With a high gain, any measurement noise present will cause excessive valve action. Therefore, the gain may be reduced, and some integral action is added within the controller.

Nonlinear Gain

Some manufacturers provide a nonlinear control algorithm that has the effect of increasing the controller gain as the measurement gets further away from set point. An example is the "error-squared algorithm," in which a modified error, ê, is calculated as:

$$\hat{e} = e|e|. \qquad \textbf{2.36(17)}$$

When the level is on set point, this algorithm gives a very low gain, which increases as the measurement gets further away from set point. Other manufacturers accomplish a similar operation by linear characterization of the error signal. Sometimes the nonlinear behavior is applied to only one of the controller modes, such as to the proportional mode, with the other controller mode seeing the normal error signal.

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