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Tuning averaging level controllers

This method makes maximum use of vessel surge capacity and thereby tends to stabilize the whole plant

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A friend of mine showed me the latest and greatest in level control technology. "Look at Fig. 1," he said. Obviously the level movement is an indication of mass imbalance:

$$F_{in} - F_{out} = A * (dL/dt) \tag{1}$$

F_{in} = flow of liquid into the vessel
F_{out} = flow of liquid out of the vessel
A = cross sectional area of the vessel
dt = the time between scans
(dL/dt) = level rate of change.

The level controller operates by manipulating the outlet flow. Every scan it implements flow changes:

$dF_{o1}, dF_{o2}, dF_{o3}, \text{- - - - - - - - - - - -}, dF_{ok}$

dF_{o1} = the change in outlet flow after the first scan from now

dF_{o2} = the change in outlet flow after the second scan from now

dF_{ok} = the change in outlet flow after a certain time horizon of interest: k scans.

We assume that from now on the inlet flow will remain constant. The integral form of Eq. 1 then takes the form:

$$A * (L_k - L_0) = A * dL_0 * k - \sum_{i=1}^{k} ((k - i + 1) * dt * dF_{oi}) \tag{2}$$

L_k = the level that will be reached after the time horizon of k scans

L_0 = the current level reading

dL_0 = the change of level from the previous to the current scan. It indicates the initial mass imbalance.

We want to bring the level to its setpoint after the time horizon, H, while at the same time minimize the movement of the manipulated variable, F_{out}. This desire can be expressed mathematically as:

$$\text{Minimize} \sum_{i=1}^{k} (dF_{oi})^2 \tag{3}$$

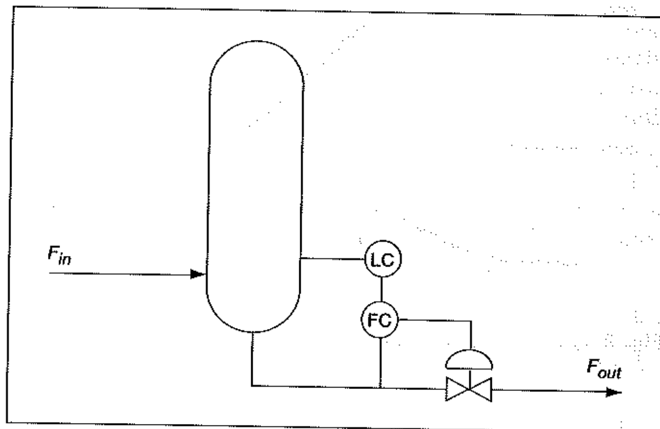

Fig. 1. Simple level cascade control.

This is subject to the constraint of Eq. 2, except we replace Lk by the setpoint L_{SP}, to express the desire to reach the setpoint after k moves:

$$A * (L_{SP} - L_0) = A * dL_0 * k - \sum_{i=1}^{k} ((k - i + 1) * dt * dF_{oi}) \tag{4}$$

L_{SP} = level controller setpoint

With the help of some textbooks, my friend solved the problem as shown in Eq. 5, for the first control action, dF_{o1}. We take interest only in dF_{o1} because each scan the controller has to only come up with the next move. The solution for dF_{o1} is:

$$dF_{o1} = C_1 * dL_0 + C_2 * dt * (L_0 - L_{SP}) \tag{5}$$

$C_1 = 4 * A / H$
$C_2 = 6\, A / (H^2)$
H = the time horizon = $dt * k$

"This is nice," I said. Let's compare it to the famous proportional plus integral (PI) algorithm in its velocity form:

$$dM = K_c * dL_0 + (K_c / T_I) * dt * (L_0 - L_{SP}) \tag{6}$$

dM = change of the manipulated variable = dF_{o1}
K_c = controller gain
T_I = controller reset time.

"You have discovered the PI algorithm," I told my friend, with settings of:

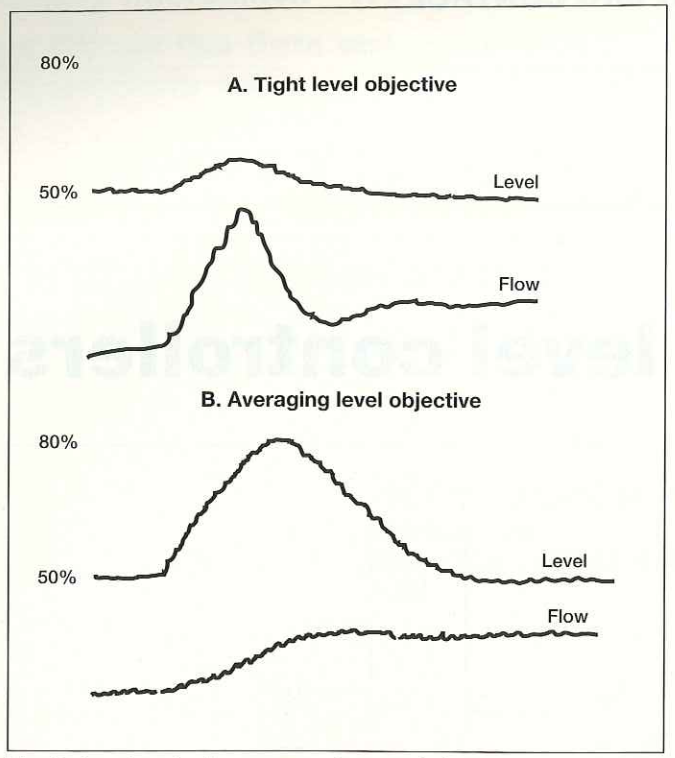

Fig. 2. Level controller response to mass balance disturbance.

$$K_c = 4 * A / H \qquad (7)$$

$$T_I = 0.67 * H$$

There was a moment of silence. It turned out that my friend's company offers a package which provides the controller of Eq. 5 on a host computer, not realizing that any DCS would be quite capable of providing the same, or even better PI controller.

I am writing this not to embarrass the company in question, but rather to show again here that the PI algorithm is mighty good for level control, and usually no alternatives are needed. I would also like to suggest a way of tuning for the averaging level controller based not on some arbitrary time horizon, but by relating the tuning constants to physical characteristics of the vessel in question.

The averaging level control objective. Fig. 2 shows two level applications with entirely different objectives. The objective of Fig. 2A is to keep the level always at its setpoint, while moving the manipulated variable as much as it takes to steady the level quickly. In contrast, the objective of Fig. 2B is to minimize the movement of the manipulated variable while keeping the level within its alarm limits. This second application is called averaging level control, and it is the subject of this article.

Before attempting to tune the controller, it would be useful to quantify the control objective. Say the controller is at steady state and the level is at setpoint. At some point a disturbance comes in such that the previous inlet flow changes by FD. Eq. 1 tells us that the level will then start changing at a rate of:

$$(dL/dt) = FD / A \qquad (8)$$

The objective is to tune the controller such that it can handle a flow disturbance of magnitude FD without exceeding the alarm limits. It is assumed that before the disturbance occurred the vessel was at steady state with its level reading being equal to the setpoint. After reaching a level peak, which would almost equal the alarm setting, the objective is to gradually bring the level back to its setpoint.

But how do we choose the magnitude of FD? One obvious way is by observing the plant operation and selecting one of the largest disturbances. Another way, less precise but often more practical, is to set FD as 20% of the typical outlet flow. A 20% step change is a fairly large flow disturbance, and thus the resulting control settings will not often trigger an alarm.

Setting the initial controller gain. Consider next a proportional only controller, not that one is recommended, but in the initial part of the response the integral action plays only a minor role and can be ignored. As Eq.8 starts working and the level trends upward, the proportional action would increase the outlet flow by:

$$F_{out} = F_{out0} + K_c * (L - L_{SP}) \qquad (9)$$

F_{out0} = the initial outlet flow before the disturbance occurred.

The manipulated variable, F_{out}, will continue to increase as long as the level is rising, until finally, when the outlet flow equals the inlet flow, the level would stabilize at a new reading:

$$F_{out} = F_{out0} + FD = F_{out0} + K_c * (L - L_{SP}) \qquad (10)$$

or:

$$FD = K_c * (L - L_{SP}) \qquad (11)$$

Our tuning objective specifies that at that time the level will be just equal to the alarm limit, and that gives us a formula for calculating the controller gain:

$$K_c = FD / (LIMIT - L_{SP}) \qquad (12)$$

$LIMIT$ = the alarm limit of the controller.

From a control point of view it is best to select the setpoint to be at the middle of the range, i.e., 50% of the level, and have alarm limits at 20% and 80%. Though of course the setting of level setpoint and limits depends not only on control considerations, but also on vessel geometry and process considerations.

Eq. 12 gives the gain in engineering units of say bpd/%. Since most DCS systems work in % of range, not in engineering units, the gain of Eq. 12 needs to be modified to:

$$K_c\% = (FD / F_{range}) * (L_{range} / (LIMIT - L_{SP})) \qquad (13)$$

F_{range} = range of the manipulated variable flow measurement.

L_{range} = range of the level measurement, normally 100%.

Setting initial reset time. Once a new steady level is reached after a disturbance, the whole plant has stabilized. Is there a need to apply a reset action which would disturb the newly accomplished steady state, only to bring the level back to its setpoint? The answer is "yes," if you wish to regain the freedom of movement in both directions. But how quickly do we need to regain that freedom? The simple answer is that the level should be brought back to its setpoint in time to meet the next major disturbance. However, we do not precisely know when the next disturbance would come.

A reasonable way to address this uncertainty is to relate the reset time to the disturbance residence time in the vessel. The disturbance residence time is defined as:

$$R_{TIME} = VS / FD \qquad (14)$$

VS = surge volume. This is the volume bound between the level setpoint, L_{SP}, and the level alarm limit, $LIMIT$. The units of VS must be such that R_{TIME} is calculated in minutes.

R_{TIME} = disturbance residence time in minutes.

A starting point which is likely to give a good result is a reset time of:

$$T_I = 4 * R_{TIME} \qquad (15)$$

By "good result" it is meant that there would be a reasonable tradeoff between using the surge volume of the vessel, and between recovering quickly from an upset situation.

Tuning test. Most level loops will have no dynamic difficulties with the settings of Eqs. 13 and 15. Loops without dead time would certainly perform well. Loops with some dead time will likely still perform well because averaging level tuning calls for slow, stable responses. It happens however, that due to improper vessel design, interactions with other loops or excessive dead time, the controller cannot support the initial tuning objective. The response then becomes oscillatory. This section suggests a set of rules for tuning averaging level loops which exhibit such oscillatory behavior.

Since averaging level tuning is normally stable, it would be a good idea, in case of sustained oscillations, to first check whether these oscillations are the result of unstable level control or just simply an oscillatory disturbance. When in doubt, the level controller can be turned to manual. If the oscillations were caused by level instability they would then die down.

If it has been established that the level controller is responsible for the oscillations, the presence of these oscillations indicates unusual dynamic problems as stated previously. It is advisable then to have discussions with process engineers and operators to understand why the loop is oscillatory. Attention should be paid to the heat exchange system which often provides the mechanism for slow loop to loop interactions.

When there are oscillations, either damped or unstable, the reset time ought to be considered first. For a satisfactory response the reset time should be at least twice as long as the oscillation period. The period of oscillation is defined in Fig. 3 as twice the time between the first high and first low peaks of a response.

If T_I is sufficiently long and the response is still too oscillatory, it is an indication that this loop cannot fulfill the objective of handling a disturbance of magnitude

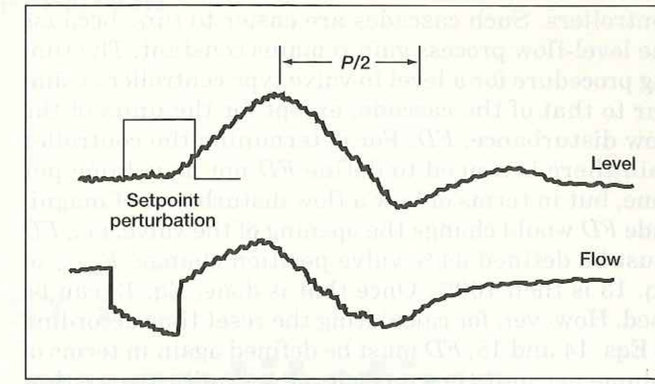

Fig. 3. Oscillatory level tuning.

FD. There is a need then to reduce the gain until the loop response is satisfactory. For level control loops, it would be reasonable to allow no more than three peaks in the response, one high and one low, followed by another small high or low peak. As shown in Fig. 3, the loop then will have no more than one overshoot and one undershoot.

Once the new gain is set, Eq. 12 can be rearranged to assess how much of a disturbance this controller can now handle without sounding an alarm:

$$FD = K_c * (LIMIT - L_{SP}) \qquad (16)$$

If FD is lower than 10% of the normal outlet flow, the loop will likely remain a problem loop.

Nonlinear control algorithms. Some DCSs have the facility of scheduling the controller gain, K_c, so that it would increase with control deviation as follows:

$$K_c = K_l * (1 + K_{nl} * \text{ABS}(L - L_{SP}) / 100) \qquad (17)$$

K_l = linear control gain

K_{nl} = a nonlinear coefficient.

Such a control action provides the ability to reduce the manipulated variable activity at small disturbances, at the expense of large adjustments when there are large disturbances.

In the author's opinion, the usefulness of nonlinear level algorithms is overrated. If the mass imbalance disturbance is large, then a weak initial action will cause the level to stray further away from setpoint. The level will eventually reach the region of higher gain and force the strong action required for the disturbance. The changes in the manipulated variable will then be even more abrupt than those of the linear controller. As a rule, vessels whose residence time is too short perform better with linear control. Vessels whose residence time is adequate give a somewhat better performance with nonlinear control.

For tuning the nonlinear algorithm, it is recommended to first find K_{cmax} as per Eq. 13:

$$K_{cmax} = (FD / F_{range}) * (L_{range} / (LIMIT - L_{SP}) \qquad (18)$$

Then set K_l and K_{nl} so that the control action would double when approaching the level limit:

$$K_l = K_{cmax} / 2 \qquad (19)$$

$$K_{nl} = 100 / \text{ABS}\ (LIMIT - L_{SP}) \qquad (20)$$

Direct level to valve control. Modern units are likely to have most of their level controllers cascading to flow controllers. Such cascades are easier to tune because the level-flow process gain remains constant. The tuning procedure for a level to valve type controller is similar to that of the cascade, except for the units of the flow disturbance, *FD*. For determining the controller gain there is a need to define *FD* not as volume per time, but in terms of how a flow disturbance of magnitude *FD* would change the opening of the valve, i.e., *FD* must be defined as % valve position change. F_{range} of Eq. 13 is then 100%. Once that is done, Eq. 13 can be used. However, for calculating the reset time according to Eqs. 14 and 15, *FD* must be defined again in terms of volume per unit time, or % level per unit time, so that the reset time is in minutes.

Level reading correction with irregular vessel shape. Many DCSs have the ability to correct the level reading so that it represents real volume, rather than height. To determine whether the additional complication of correcting that reading is worth it, there is a need to discriminate between two cases. First, consider the special, but common case of a horizontal vessel, where the level setpoint is at the middle of the vessel, and the high and low limits are at 20% and 80% of the vessel diameter respectively. Second, consider a more general case of irregular vessels.

To begin with the special case of the level setpoint being at the middle of a horizontal vessel, the Pythagorean theorem shows that at 80% or 20% level reading, the cross sectional area is reduced to 80% of the area in the middle of the cylinder. The effect of that reading error would be similar to the nonlinear controller of Eq. 17, with the control gain being 25% stronger at high deviations ($1/0.8 = 1.25$). Hence, we conclude that it is in fact desirable to have such an error, and no level linearization is required. About the same conclusion can be reached for a spherical vessel, except there the area at the limits is 64% of the area in the middle, and the nonlinear effect is then so that the gain at the limits is 56% stronger than in the middle. This increase is perfectly acceptable, as evidenced by the recommendation of Eqs. 18 through 20.

For the general case of irregular shape, or even for cylinders where the setpoint is not in the middle, it is recommended that level correction be performed, if that is possible. Otherwise the controller becomes difficult to tune and the tuning objective looses its meaning. ■

The author

Y. Zak Friedman *is a consultant in the field of process control and on-line optimization. He has pursued research in the area of using process models for real-time closed loop control and optimization of refinery units, and in standardization of control technology. His experience spans 25 years in the refining industry, working for such employers as Exxon Research and Engineering, and KBC Advanced Technology. Dr. Friedman holds a BS degree from the Israel Institute of Technology and a PhD degree from Purdue University.*