2: Determining Process Dynamics

Myke King continues his detailed series on process control, seeking to inspire chemical engineers to exploit untapped opportunities for improvement

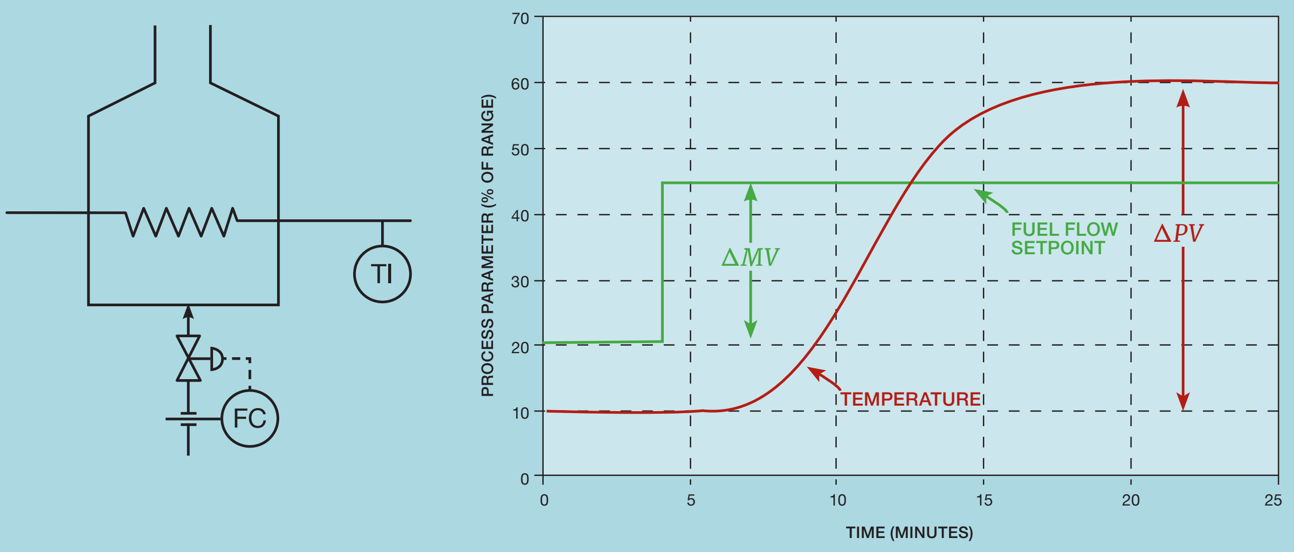
IDENTIFYING process dynamics is an essential first step to achieving good control design. Many control engineers will be guilty of spending many hours tuning controllers by trial and error. This can largely be avoided. Using the process dynamics, obtained from straightforward plant testing, simple calculations can then be applied to identify optimum tuning for all types of controller.

So, imagine that you're the first into the bathroom in the morning and you turn on the hot tap. Initially it produces cold water and does so for some time. We anticipate this behaviour; the water, held up between the source of hot water and the tap, has had time to cool to room temperature. This water first has to be displaced. It has to be transported to the tap and causes what we describe as a *transport delay*. Also known as *deadtime* (θ), it is the first of the dynamic constants that we use to describe process behaviour. We also notice that, when the cold water has been displaced, the water temperature doesn't immediately increase to that of the hot water; it ramps up to it. Part of the energy in the water is required to warm up the pipework leading to the tap. The process is said to have *capacity*, in this case thermal capacity. It is responsible for the second of our dynamic parameters which we describe as *lag* (τ). The water temperature has changed from cold to hot, from one *steady state* to another. We describe this difference as the *steady state change*.

PROCESS GAIN

Now we must leave the analogy because we can't use the opening of the tap to actually control the temperature. We'll consider instead the fired heater shown in Figure 1. It has a flow controller (FC) installed on the fuel supply and temperature indicator (TI) on the product outlet. Our intent is to install a temperature controller that will adjust the fuel flow. To design this controller we first determine the process dynamics. We make a step change to the FC set-point. The temperature shows much the same response as our bathroom tap. The TI is located downstream of the firebox and so exhibits transport delay. The heating coil, and the fluid it contains, have significant capacity to absorb heat; so we also see lag. The final, and most important parameter, is the *process gain* (K_p). This is defined as the

Figure 1: Process response



steady state change in temperature, our process variable (*PV*), divided by the steady state change in fuel flow set-point, our manipulated variable (*MV*).

$$K_p = \frac{\Delta PV}{\Delta MV}$$

We need to be careful with units. The process gain, as we'll see in a future article on tuning, is largely used to determine the *controller gain* (K_c). We'll see that controllers installed in programmable logic controllers (PLC) or distributed control systems (DCS) work in dimensionless form. K_c must be dimensionless and so, therefore, must K_p. To make both *PV* and *MV* dimensionless; we divide each by its corresponding range (or *span*). The instrument ranges for the TI and FC will have been assigned by the instrument engineer as part of the control system configuration. They usually remain constant. (The need to retune the controller, if the instrument ranges are ever changed, is frequently overlooked.)

$$K_p = \frac{\left(\frac{\Delta PV}{PV_{range}}\right)}{\left(\frac{\Delta MV}{MV_{range}}\right)}$$

Note that control applications, like multivariable predictive controllers (MPC), residing at the computer level generally operate in engineering units. For these applications, K_p should not be converted to its dimensionless form. Numerically K_p can be positive or negative, even zero in very unusual circumstances. In our example of the fired heater, increasing fuel increases the temperature and so the process gain is positive. If we were planning to control the temperature of a cooler by manipulating the flow of cooling water, the process gain would be negative. The choice of instrument ranges will often result in K_p being close to 1 but there can be occasions where it differs by several orders of magnitude.

ESTIMATING THE TIME CONSTANTS

There are numerous published methods for determining θ and τ. Most are based on the assumption that the response curve can be represented by

$$\Delta PV = K_p \Delta MV_{t-\theta}\left(1 - e^{-t/\tau}\right)$$

To illustrate the principle behind some of these methods, we set the time elapsed since expiry of the deadtime (*t*) to τ. This gives

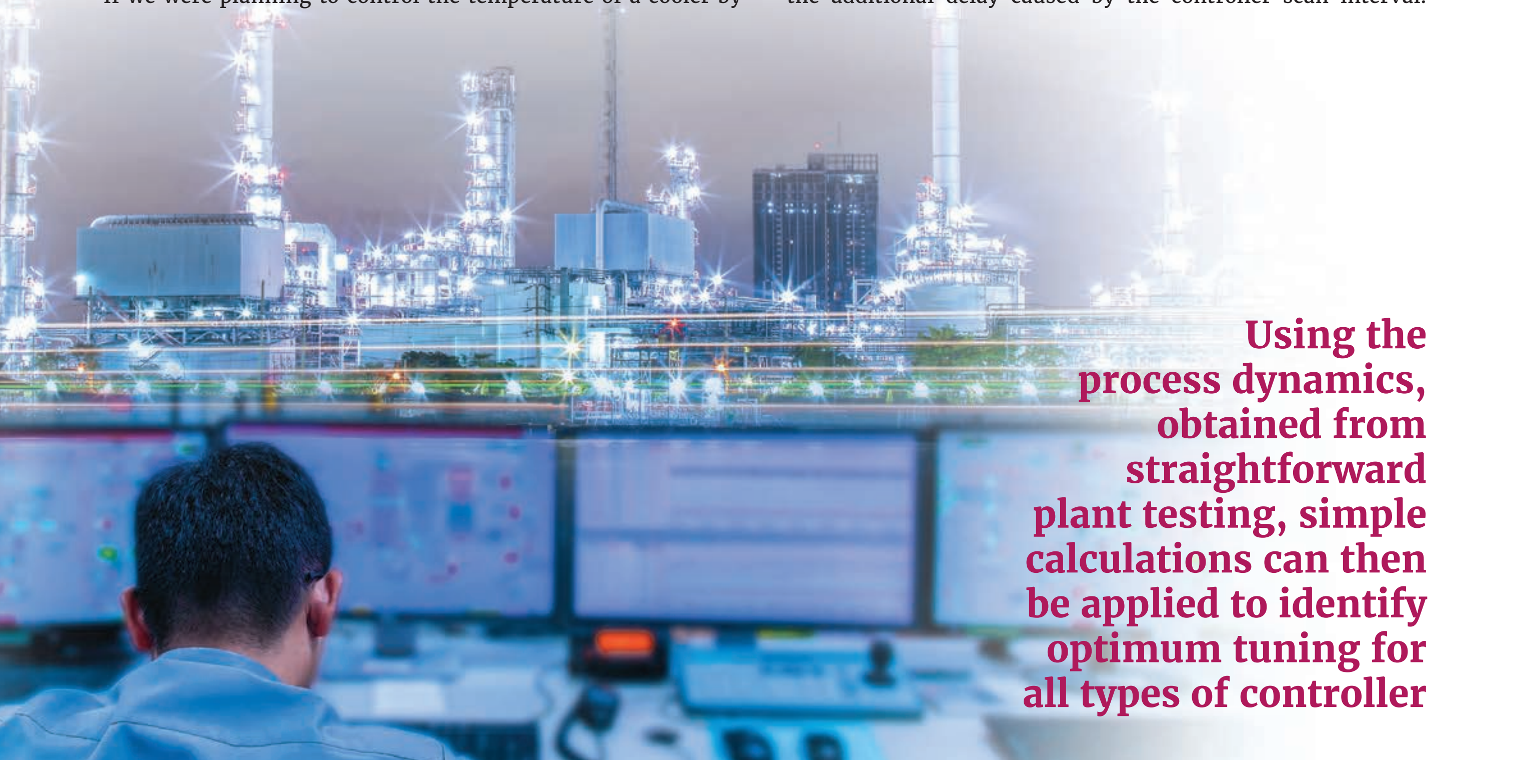
$$\Delta PV = 0.63 K_p \Delta MV$$

We can use this formula to determine τ; it is the time taken (after the deadtime has elapsed) for the *PV* to reach 63% of the steady state change. Knowing only $\theta + \tau$, we of course need to also determine θ. One approach is to identify two points on the response curve. One of the more reliable methods is based on identifying t_{25} and t_{75} – the times taken to reach 25% and 75% of the steady state change. Fitting our exponential curve through these two points gives

$$\tau = 0.91(t_{75} - t_{25})$$

$$\theta = 1.26 t_{25} - 0.26 t_{75}$$

Numerically, θ can be negligibly small or many hours. However, with digital control, it will never be truly zero because of the additional delay caused by the controller scan interval.



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Similarly τ can vary greatly from process to process. The unit of measure chosen should be either minutes or seconds. The choice will depend on the units used by the control system for the controller tuning parameters.

ORDER

However, a process response will usually be more complex than that described by a *first order* exponential function. There will be multiple sources of lag. For example, a small lag (τ_1) will be introduced by the flow controller; although we change its set-point as a step, the valve is actually moved by the controller. The firebox will introduce a large lag (τ_2) by absorbing some of the energy provided by the additional fuel. The coil will behave similarly (τ_3). The thermocouple is located in a thermowell, which again has (albeit small) heat capacity that lags the temperature measurement (τ_4). While not precisely correct, we can think of the dynamic behaviour as governed by

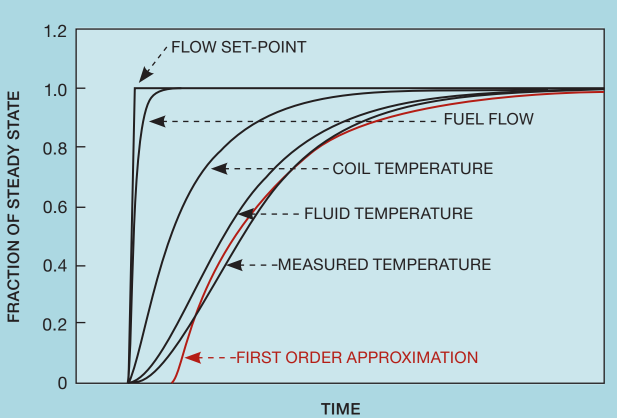
$$\Delta PV = K_p MV_{t-\theta}\left(1-e^{-t/\tau_1}\right)\left(1-e^{-t/\tau_2}\right)\left(1-e^{-t/\tau_3}\right)\left(1-e^{-t/\tau_4}\right)$$

Figure 2 shows the cumulative effect of what is now a fourth order process. It is, however, unrealistic that we can quantify individual lags. Instead we take the *lumped parameter* approach, making the assumption that the process is first order. Included in Figure 2 is the result of applying the two-point method to *model identification*. This approximation is usually reliable enough to design the controller. There are, however, a few exceptions that we will cover in future articles.

LINEARITY

The underlying assumption, so far, is that the process is *linear*, i.e. there is a linear relationship between *PV* and *MV*– meaning that K_p (the slope of this relationship) is constant. To confirm this we need to collect data at more than the two steady state conditions. So, if we first increased the fuel set-point, we should return to the starting condition and then test with an equivalent decrease – using this to give a second estimate of K_p. No process is truly linear. The design technique we will cover in the article on controller tuning gives a *robust* controller that will tolerate variation in K_p of ±20%. If, from our plant testing, the higher estimate of K_p is less than 1.5 times the lower, then both estimates will be within 20% of the average and we can treat the process as linear. If not, then we will need to apply one of the methods we cover in the forthcoming articles on signal conditioning.

Figure 2: First order approximation



CURVE FITTING

The limitation of most published methods is that they are applicable to a single step change, with the process starting and finishing at steady state. In practice, effective step-testing comprises a series of step tests of varying size and duration – covering the normal operating range. This provides a more reliable result. Further, a series of small steps is likely to be more acceptable, to the process operator, than a single large one. It is advisable to wait for steady state for at least some of the steps. But we might, particularly on processes with large time constants, reduce testing time by taking the next step without waiting. But the main issue is that we don't want to be constrained to making step changes. We'll see later that we will require the dynamics of other variables that might disturb the process. It may be impossible to change these as steps, most notably if we want to include ambient temperature. But the most compelling argument is that most controller tuning is done on existing controllers. In our case the heater outlet temperature controller may have been in service for some time, but we see the potential to improve its performance. It's preferable to perform step-testing with the controller on *auto* or in *closed loop* mode. We maintain control of the process during testing and we can stay within pre-defined bounds for the *PV*. One might think that the temperature controller tuning will somehow affect the model identification. By changing its set-point we cause it to change the fuel flow set-point. Certainly, the way it adjusts the flow set-point will depend on the temperature controller tuning and the changes will certainly not be step changes. But this does not affect the relationship between temperature and fuel flow. It does, however, preclude the use of any identification technique which assumes a step change. Instead we must apply some form of curve fitting.

At this stage we move from analog to digital. Controllers are now largely digital and, although the process is analog, data collected are stored digitally and at a fixed interval. The digital equivalent of a first order process predicts behaviour as

$$PV_n^* = exp(-ts/\tau)PV_{n-1}^* + [1 - exp(-ts/\tau)]\left(K_p MV_{n-\theta/ts} + bias\right)$$

The current PV_n^* is predicted from the previous value (PV_{n-1}^*) and the delayed $MV_{n-\theta/ts}$, where *ts* is the data collection interval.

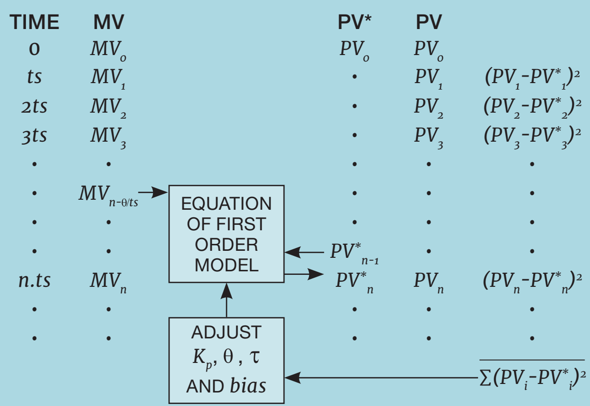
The *bias* is required because it is unlikely that the *PV* will be zero when the *MV* is zero. Because this is a sampled system, θ/ts is restricted to integer values. To overcome this limitation we define the delayed *MV* as an interpolation between adjacent values.

$$MV_{n-\theta/ts} = MV_{n-int(\theta/ts)} - \left(\frac{\theta}{ts} - int\left(\frac{\theta}{ts}\right)\right)\left(MV_{n-int(\theta/ts)} - MV_{n-int(\theta/ts)-1}\right)$$

Figure 3 shows the calculation methodology. From the process data historian, MV and PV data are collected along with their time stamp (in this case in seconds). PV^*_0 is initialized to the actual measurement (PV_0). The remaining predictions are determined using the formulae above. The sum of the squares of the prediction error is minimised by adjusting K_p, θ, τ and *bias*. Plotting the predicted *PV* against the measured *PV* will highlight any nonlinearity. Commercial model identification software is widely available but the calculations above can readily be implemented in Excel, using its Solver function. An example spreadsheet may be downloaded at www.whitehouse-consulting.com/icheme/model.xlsx.

One of the advantages of this approach is that it can be applied to set-point changes made routinely in the past. As a routine, the data historian should be configured to include the set-point, measurement and output of all controllers. Data compression, used by many historians, should be disabled since this will distort the process dynamics. ■

Figure 3: Calculation procedure



NEXT ISSUE

The next article will be one of several on the subject of the PID control algorithm. We'll describe the many modifications made to the classic version and how they are configured in the leading control systems.

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