14: Pressure Compensated Temperatures

Myke King shows how to pressure compensate distillation column tray temperature controllers

THE TEMPERATURE, measured on a tray in a distillation column, can be a good indication of product composition. Separation by distillation relies on the components having different boiling points. So, there will be a correlation between composition and boiling point. Tray temperature measures boiling point (at the operating pressure) and so provides a simple *inferential property*. **Figure 1 shows a typical control scheme. Our example is a column separating a mixed propane/butane stream. The product specifications permit 5% propane in the bottom butane product and vice versa.**

Figure 2 show the correlation between bottoms composition and tray temperature. If we were to control the tray temperature at 84.4°C, this should keep the butane on grade. However,

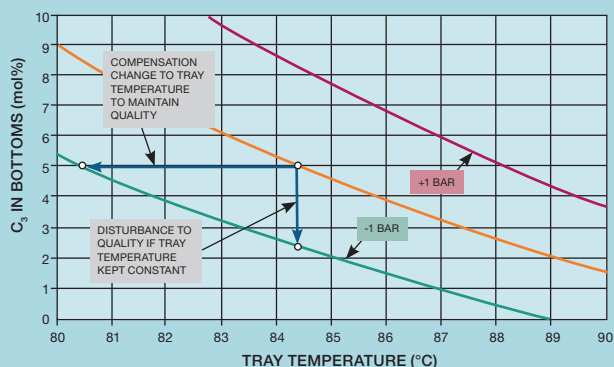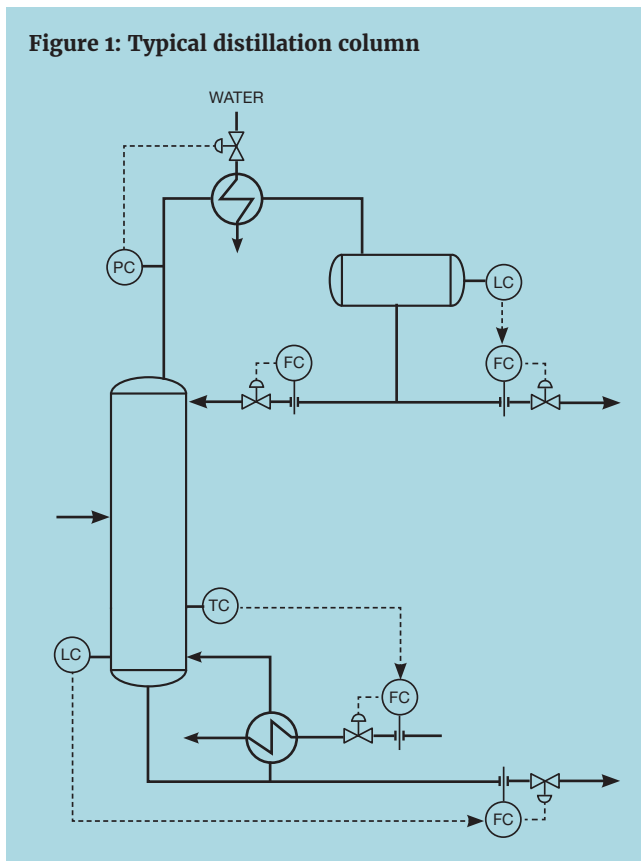Figure 2: Effect of pressure on bottoms composition


Figure 1: Typical distillation column

the relationship between composition and tray temperature is not fixed. Firstly, the tray will be a few trays up from the bottom of the column, so we are effectively controlling the composition on the tray. Any change in the vapour and liquid flows below this tray will change the product composition. However, such changes are small when compared to the effect of column pressure. Secondly, boiling point is a function of pressure. While pressure is likely to be controlled (in this example at 12 barg), it can be attractive to manipulate its setpoint. Reducing pressure saves energy or debottlenecks a limiting reboiler. Increasing it helps relax any condenser constraint, potentially allowing feed rate to be increased. Adjusting pressure can be economically attractive and is often included in a multivariable predictive controller (MPC).

LINEAR COMPENSATION

Figure 2 shows the effect of reducing the column pressure from 12 to 11 barg. Keeping the tray temperature fixed will reduce the propane content of the butane to 2.4%. We can see, that to stay on grade, the tray temperature should be reduced to 80.5°C. This is the key to pressure compensation; we require a change of 3.9°C per bar of pressure.

Figure 3 shows the same effect at the top of the column. If we were to have a temperature controller in this section, then dT/dP would be about 3.1°C/bar.

Figure 3: Effect of pressure on overheads composition
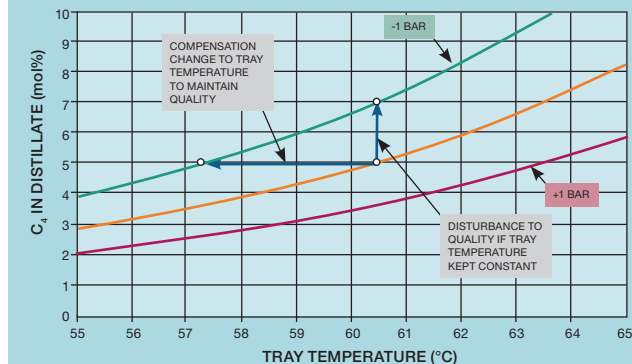
The simplest way to apply this technique is to use signal conditioning. The measurement (PV) of the temperature controller becomes the pressure compensated temperature (PCT):

$$PV = PCT = T_{measured} - \frac{dT}{dP}\left(P_{measured} - P_{reference}\right)$$

The reference pressure is any sensible constant value; it might be design pressure or the average operating pressure. The use of the minus sign might appear incorrect; we want the tray temperature to increase with pressure. But remember that this is the PV of a controller. By "tricking" the controller into believing the temperature has dropped, it will take corrective action to increase it.

It would be unlikely that the pressure measurement is on the same tray as the temperature. Indeed, as is shown in Figure 1, it is common for the pressure to be measured (and controlled) at the top of the column. However, although far from the tray, this measurement is normally suitable. The pressure drop across the column will be small compared to the operating pressure and, in any case, we are concerned with only changes in pressure. The exception to this can be vacuum distillation columns. Distillation under vacuum is used to reduce

> **On most columns, the pressure measurement we use for compensation will also be the measurement used by the pressure controller. This offers the option of using the pressure controller setpoint, rather than its PV, for compensation**

the boiling point of heavy components, for example, to avoid thermal cracking. As we have seen, the heavier the hydrocarbon, the higher the value of dT/dP. Under these circumstances, this can be as high as 800°C/bar – increasing the impact of using the wrong pressure. Secondly, the pressure drop across a vacuum column will be comparable to the operating pressure. Any change in pressure drop would cause a significant error. Under these circumstances the pressure measurement should be close to the tray used for temperature control.

USE THE PRESSURE CONTROLLER SETPOINT

On most columns, the pressure measurement we use for compensation will also be the measurement used by the pressure controller. This offers the option of using the pressure controller setpoint, rather than its PV, for compensation. The most compelling reason to do this is when the pressure controller output is *saturated*. Caused by reaching the condenser duty limit, the controller output has reached 100% but the PV is still higher than the setpoint. If we were to use the PV for compensation then the temperature controller would increase the tray temperature by increasing the reboiler duty. This would overload the condenser further and cause a further increase in pressure. Further, using the setpoint also means that the pressure compensation will not unnecessarily respond to PV disturbances that will quickly be corrected by the pressure controller. Another advantage is that the setpoint will be noise-free. If an operating issue requires the pressure controller to be switched to manual, PV tracking means the setpoint will follow the measured pressure and so can still be used in the PCT.

However, as with our example of vacuum distillation, it may only be feasible to use a measurement different from the one used for control. Then, it may be necessary to filter noise and include logic to ensure that the temperature controller will not aggravate any violation of the condenser limit.

ANTOINE

There are several ways in which dT/dP might be determined – one of which is based on the Antoine equation:

$$\ln(P) = A - \frac{B}{T+C} \qquad \therefore \frac{dT}{dP} = \frac{B}{P(A - \ln(P))^2}$$

P is the absolute pressure and T the temperature. A, B, and C are coefficients that are readily available for most components. Table 1 gives them for propane and butane (for pressure measured in bara and temperature in °C). The calculated values of dT/dP assume an operating pressure of 12 barg (~13 bara).

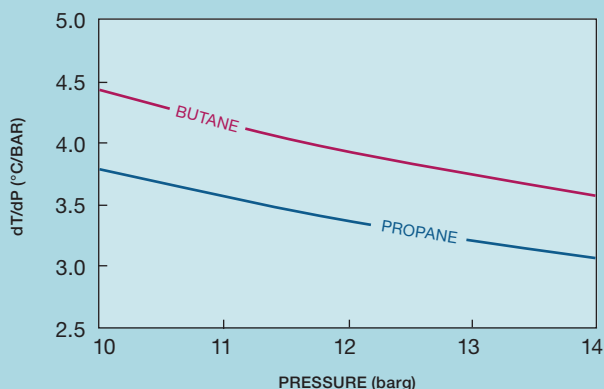There are two limitations to this approach. First, Antoine coefficients are available only for pure components and, even if we could deduce them for a mixture, we do not know the composition on the tray. However, distillation is usually used to separate components with similar properties – as in this case, and so the estimates dT/dP for the two key components will be similar. Little error would be introduced by interpolating between the two values, perhaps based on the position of the tray in the column.

Table 1: Antoine coefficients

	C_3H_8	C_4H_{10}
A	9.05654	9.05800
B	1850.841	2154.897
C	246.99	238.73
dT/dP	3.38	3.93

The second issue is one of non-linearity between temperature and pressure. As Figure 4 shows, dT/dP varies with pressure. If pressure is likely to change significantly then we can apply the Antoine equation directly. Rearranging for both the current temperature (T) at the current pressure (P); and for the PCT at the reference pressure:

$$T = \frac{B}{A - \ln(P_{measured})} - C \qquad PCT = \frac{B}{A - \ln(P_{reference})} - C$$

Subtracting gives:

$$PCT = T - B\left[\frac{1}{A - \ln(P_{measured})} - \frac{1}{A - \ln(P_{reference})}\right]$$

CLAUSIUS-CLAPEYRON

An alternative approach is to use the Clausius–Clapeyron equation:

$$\frac{dT}{dP} = \frac{RT^2}{\lambda . MW . P}$$

R is the universal gas constant (8.3145 kJ/kg-mole/K). λ is the heat of vaporisation (kJ/kg) and MW the molecular weight (kg/kg-mole). P is again absolute pressure and, in this case, T is the absolute temperature (K). Applying the same manipulation as for the Antoine method, and converting to °C gives:

$$PCT = \frac{1}{\frac{R}{\lambda . MW} \ln\left[\frac{P_{measured}}{P_{reference}}\right] + \frac{1}{T + 273.15}} - 273.15$$

Of course, this equation suffers the same limitation as Antoine, in that we do not accurately know λ and MW for the mixture of components.

REGRESSION

Another, relatively minor, limitation of the techniques we've covered so far is that they compensate only for the change in boiling point. As we reduce pressure in a distillation column, we also improve the relative volatility of the components. To achieve the same product compositions, we should reduce the reboiler duty further. And remember, we're controlling the composition on a tray, not that of the bottoms product. Any

Figure 4: Nonlinearity of pressure compensation

change in vapour rate below the tray will affect the bottom composition. The PCT as it stands, although a considerable improvement, will not maintain exactly constant composition. While we might develop improved compensation from a computer simulation, a more practical approach would be to apply regression analysis to historical data. We will cover this in greater detail in the forthcoming articles on inferential properties but, in principle, we identify the correlation between the % propane in butane (Q) and the pressure and temperature:

$$Q = a_0 + a_1 T + a_2 P$$

differentiating

$$dQ = a_1 dT + a_2 dP$$

When the pressure changes, we want the composition not to change, so dQ is zero and:

$$\frac{dT}{dP} = -\frac{a_2}{a_1}$$

Increasing tray temperature has the effect on composition opposite to that when increasing pressure. So a_2 will be opposite in sign to a_1 and dT/dP will be positive. We still need to include the negative sign in the calculation of *PCT*.

This technique can be extended to produce a non-linear PCT. We know from the theoretical techniques above, that PCT benefits from the inclusion $\ln(P^*)$ – where P^* is the pressure on an absolute basis. So, we include this in the regression analysis:

$$Q = a_0 + a_1 T + a_2 P + a_3 \ln(P^*)$$

$$dQ = a_1 dT + a_2 dP + \frac{a_3}{P^*} dP$$

$$PCT = T + \frac{a_2}{a_1}\left(P_{measured} - P_{reference}\right) + \frac{a_3}{a_1}\ln\left(\frac{P^*_{measured}}{P^*_{reference}}\right)$$

Increasing tray temperature has the effect on composition opposite to that when increasing pressure

Most distillation columns, like our example, have only a single temperature controller. So, while our scheme will respond to a change in pressure by adjusting the reboiler duty (to maintain the tray temperature), the reflux flow remains constant. The overheads composition will therefore change. Without a temperature controller in place, we can use regression to develop a pressure compensation technique for the reflux flow. We correlate the % butane in propane with the reflux flow (R), for example:

$$Q = b_0 + b_1 P + b_2 R$$

And, so obtain:

$$\frac{dR}{dP} = -\frac{b_1}{b_2}$$

Some columns, such as a propane/propylene splitter, usually have no tray temperature controllers. When the boiling points of the components are very similar, temperature is insensitive to composition and so unreliable – particularly on high purity columns. The approach taken for column reflux flow can be extended to reboiler duty.

Such correlations are also developed from step-testing for multivariable predictive control (MPC). For example, those variables stepped would include column pressure (P) and reflux flow (R). Considering only the steady state behaviour, testing would quantify the process gains $\Delta Q/\Delta P$ and $\Delta Q/\Delta R$. In principle, dR/dP can then be determined from the ratio of these two values. However, if both sub-models are included in the MPC, it will adjust reflux to maintain constant composition as pressure is changed. No additional compensation would be required. ■

Myke King CEng FIChemE is director of Whitehouse Consulting, an independent advisor covering all aspects of process control. The topics featured in this series are covered in greater detail in his book Process Control – A Practical Approach, published by Wiley in 2016