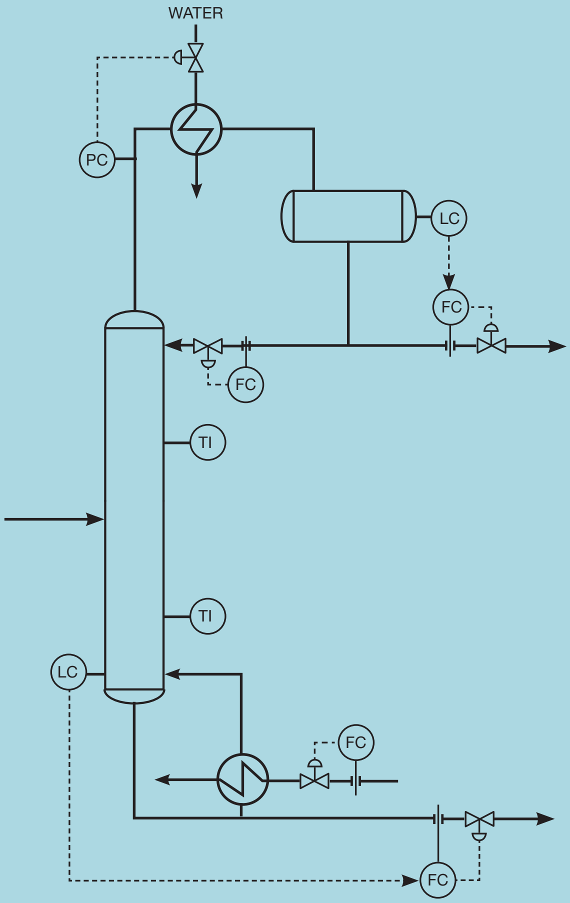
FEATURE SERIES: PRACTICAL PROCESS CONTROL

# 25: Distillation – Part 2

*In the last issue, we highlighted the importance of adjusting both cut and fractionation variables in response to column disturbances or setpoint changes. Here, we examine the problems that arise from their interaction*

**OUR CASE STUDY, as before, comprises a simple binary separation of propane and butane. Propane is deemed on-grade if the C4 content is less than 5%. Similarly, butane must contain less than 5% C3. Figure 1 shows the commonly applied energy balance control strategy *(see TCE 992)* in which the reflux drum level is controlled by manipulating the distillate flow and the column level controlled by the bottoms flow.**

**Figure 1: Energy balance scheme**




## QUICK READ

- **Tray Temperatures Reflect Composition:** Controlling tray temperatures helps maintain product purity in binary separations – assuming stable pressure
- **Controller Interaction Can Destabilise Systems**: Reflux and reboiler controls can conflict, so detuning one controller may stabilise dual composition control
- **RGA and MPC Support Robust Control**: Relative gain analysis (RGA) and condition number help assess controller pairing and predict model predictive control (MPC) performance

## TRAY TEMPERATURE CONTROL

Liquid on the trays in the column is at its bubble point, which is composition dependent. So, by measuring temperature, we can infer composition.

Figure 2 shows the correlation between propane composition and the tray temperature measured a few trays from the top of the column. While there will be some change in composition between this tray and the vapour leaving the column, manipulating reflux to control this temperature at around 60°C

**Figure 2: Propane composition**

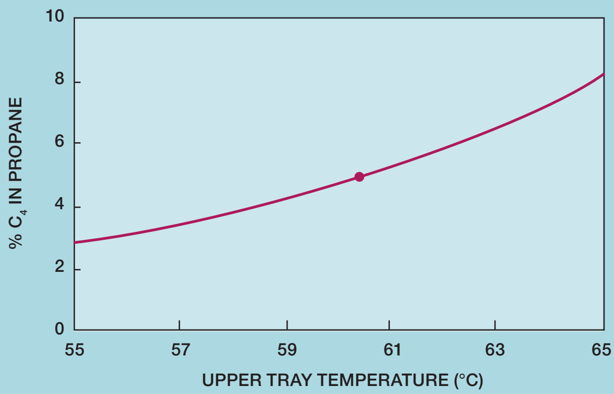
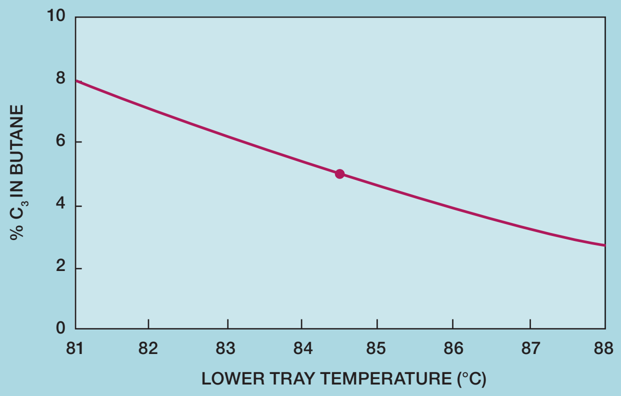


**Figure 3: Butane composition**




would maintain the C4 content of propane close to the target of 5%. Similarly, as Figure 3 shows, manipulating the steam flow to the reboiler to control the bottom tray temperature at around 84°C would maintain the bottoms composition. Of course, bubble point is also a function of pressure. The correlations, as drawn, assume operation at a fixed pressure. To allow pressure to be adjusted, pressure compensated temperature measurements would be used, as described in TCE 996. And the underlying assumption is that temperature is sensitive to changes in composition. This is the case when the components being separated have very different boiling points. Tray temperature control would not be feasible on, for example, a propane/propene splitter.

## INTERACTIONS

While we can usually control the composition of one product relatively easily, controlling both is usually much more challenging. The lower curve in Figure 4 shows the effect that reflux has on the top tray temperature. Setting it at 56.5 m³/hr will maintain the temperature at the required value. However, if this controller takes corrective action (for example to accommodate a change in temperature setpoint), the upper curve shows that bottom tray temperature will also change. Its controller will take corrective action by manipulating the reboil steam but, as Figure 5 shows, this will also change the top temperature. The two controllers will fight each other – often to the point of becoming unstable.

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In this example, the solution may be relatively straightforward. Shown on the figures are the process gains. These are the slopes of the lines at the operating point. For example $(K_p)_{11}$ is the process gain between $PV_1$ (top tray temperature) and $MV_1$ (reflux):

$$(K_p)_{11} = \frac{\Delta PV_1}{\Delta MV_1} = -2.0\ ℃/(\mathrm{m}^3/\mathrm{hr})$$

The impact that reflux has on the lower temperature is significantly less, with a process gain of -1.0. The same is true of reboiler duty; the impact it has on the temperature lower down the column is (marginally) greater than on the temperature in the upper section. This situation may be exploited by tuning one of the controllers conventionally and, with this in automatic mode, performing step-tests to determine the dynamics of the other. These are used to tune the second controller, but

**Figure 4: Interaction with distillate composition controller**

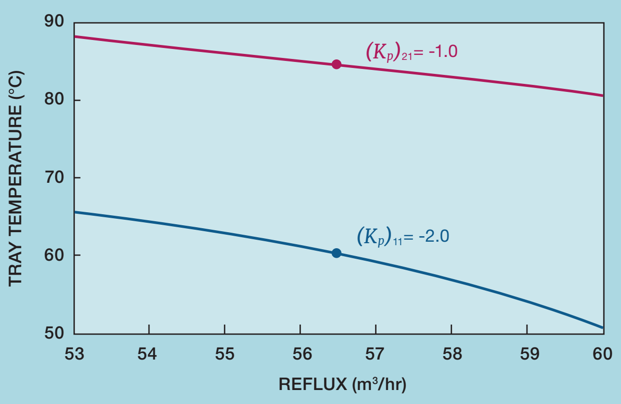
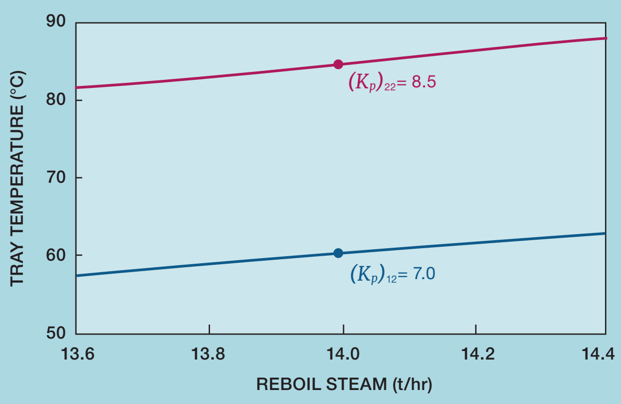



**Figure 5: Interaction with bottoms composition controller**

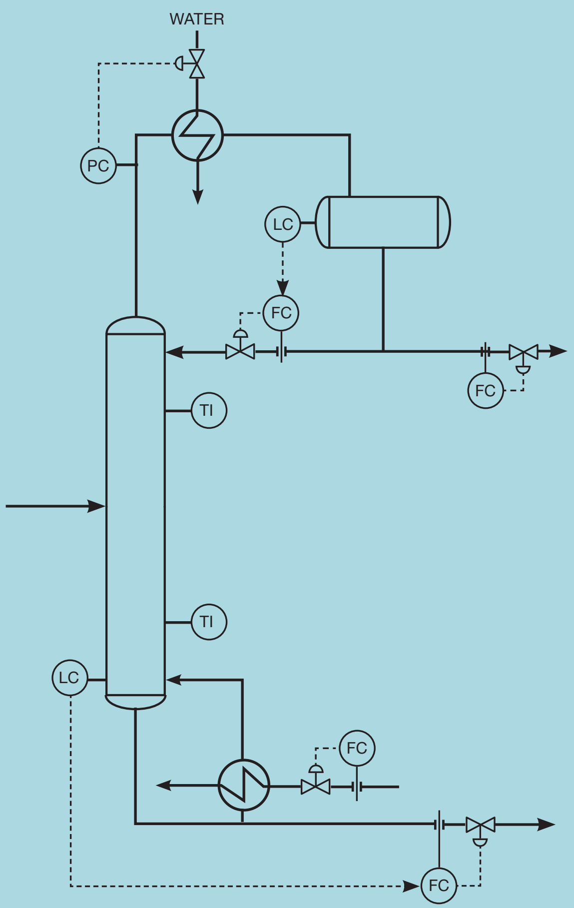


its controller gain is reduced by at least 50%. The first controller will operate as normal. The disturbances it causes to the other will be corrected very slowly – at a speed with which they can be dealt with by the first. While not an especially elegant solution, it may be appropriate when precise composition control of one product is significantly more critical than that of the other.

## RELATIVE GAIN

Relative gain analysis (RGA) is a long-established technique that identifies the best *pairing*, ie which manipulated variable (MV) should be used to control which process variable (PV). For our example, this will only confirm the pairing we have selected intuitively. But it can be very useful for more complex problems such as, for example, columns with a side-draw. It also quantifies the level of interaction between controllers. The technique has been somewhat eclipsed by the advent of model predictive control (MPC) but is making something of a resurgence. The methodology is proving useful identifying where an impractically large MPC can be simplified.

**Figure 6: Alternative drum level controller**




Relative gain ($\lambda$) is defined as the ratio of the process gain determined with all the other controllers in manual mode to the same process gain with all the other controllers in automatic mode. For our simple 2×2 case, it is the process gain of the upper temperature with respect to changes in reflux (with reboil constant) divided by the gain determined with the other temperature kept constant:

$$\lambda_{11} = \frac{\left(\frac{\Delta PV_1}{\Delta MV_1}\right)_{\Delta MV_2=0}}{\left(\frac{\Delta PV_1}{\Delta MV_1}\right)_{\Delta PV_2=0}}$$

While a simple calculation, the problem is that we have yet to install any controllers, so none exist to enable us to determine the denominator. However, knowing all four process gains, we can derive relative gain from:

$$\lambda_{11} = 1 - \frac{1}{1 - \frac{(K_p)_{12}(K_p)_{21}}{(K_p)_{11}(K_p)_{22}}} = 1.7$$

Just as our four process gains form a matrix, so do relative gains ($\Lambda$). It can be shown that the sum of the relative gains in a column (or row) is 1:

$$\lambda_{12} = \lambda_{21} = 1 - \lambda_{11} \qquad \text{and} \qquad \lambda_{22} = \lambda_{11}$$

$$\boldsymbol{\Lambda} = \begin{pmatrix} 1.7 & -0.7 \\ -0.7 & 1.7 \end{pmatrix}$$

If the controllers do not interact then the process gain will be the same, no matter whether the other controller is in auto or manual, and so the relative gain will be 1. If we have properly paired the controllers, and there are no interactions, then $\Lambda$ would be the identity matrix. The worst possible case would be all four relative gains having a value of 0.5. Our example is well away from this and would probably permit the detuning approach, described above, to be effective.

However, in this example, the story changes if we modify the level control strategy. Figure 6 shows the *material balance* scheme, in which in the drum level controller now manipulates

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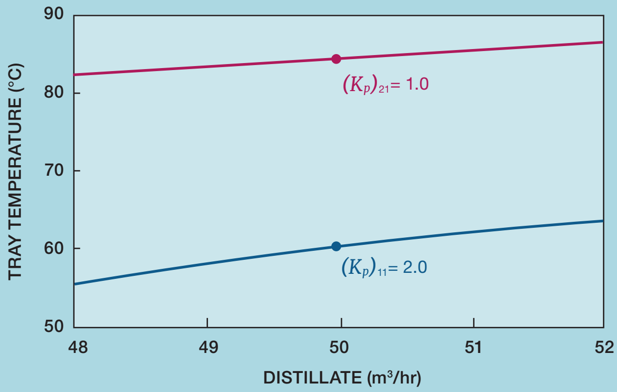



**Figure 7: Effect of distillate flow**

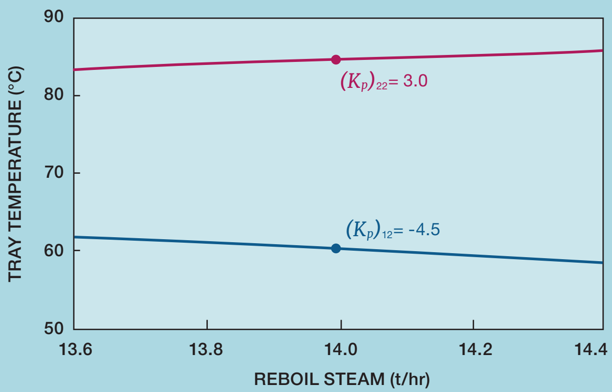



**Figure 8: Effect of reflux flow**

the reflux. The distillate flow is now adjusted to control the top tray temperature. Figures 7 and 8 show the level of interaction. As before, the top tray is more sensitive to changes in distillate flow than the temperature lower down the tower. But now it is also more sensitive to changes in reboiler duty. RGA gives:

$$\Lambda = \begin{pmatrix} 0.6 & 0.4 \\ 0.4 & 0.6 \end{pmatrix}$$

While the pairing remains correct, the interaction is now much more severe – maybe to the point where dual composition control becomes infeasible. Indeed, other issues being equal, this might be an argument for choosing the original level control strategy.

## DECOUPLING

The principle behind decoupling is that when one controller takes corrective action, a compensating change is made to the other controller's MV to keep its PV undisturbed. So, the output of the top temperature controller is multiplied by the ratio $(K_p)_{21}/(K_p)_{22}$ and added to the output of the bottom controller. Similarly, the output of the bottom controller is multiplied by $(K_p)_{12}/(K_p)_{11}$ and added to that of the top controller. However, this provides only steady-state decoupling. It is likely that the moves should be dynamically compensated. This requires the inclusion of two deadtime/lead-lag algorithms (that we introduced in *TCE 999*). While it is possible to build such a scheme in the DCS, experience shows such an approach fails. Firstly, design and implementation are extremely complex, and long-term support can be problematic. Secondly, it is prone to failure. For example, decoupling will break down if an operating constraint is reached. And, because the process is inherently non-linear, changes in process gains can invalidate the compensation it makes. MPC goes some way to resolving these issues but will also fail as the relative gains approach 0.5.

Figure 9: Effect of reflex flow

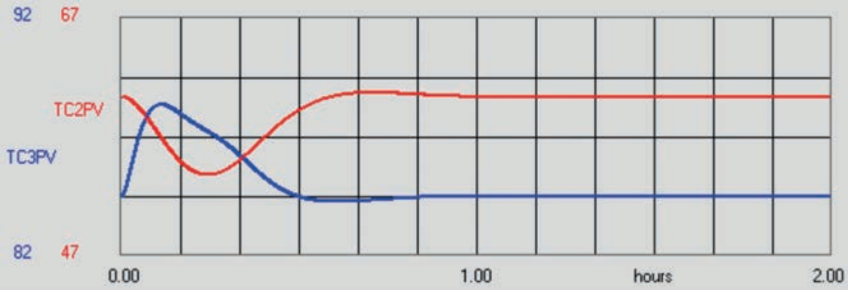
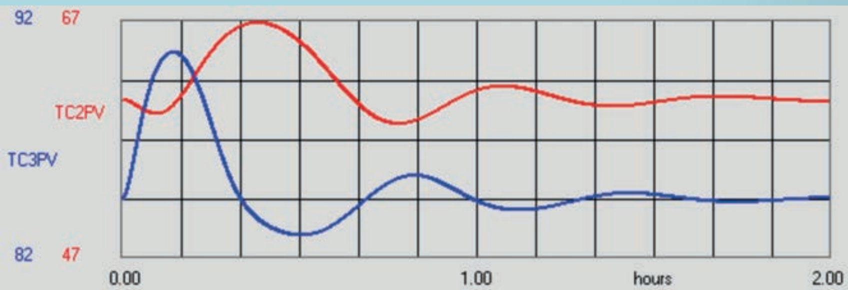



Figure 10: Material balance scheme




## CONDITION NUMBER

Condition number is a methodology often applied to the solution of simultaneous equations. It is a measure of how robust the solution might be. In concept it quantifies how sensitive the solution is to very minor changes in any variable. The same technique can be applied to relative gain matrices. It first requires the absolute value of elements in each column to be summed. We then select the largest total. For example, in our first example, this would be 2.4. We then do the same for the inverted matrix:

$$\boldsymbol{\Lambda}^{-1} = \begin{pmatrix} 0.71 & 0.29 \\ 0.29 & 0.71 \end{pmatrix}$$

Here, the largest total is 1.0. The condition number is calculated by multiplying the two maximum values, resulting in a value of 2.4. As a guide, MPC performs well if the condition number is less than 5 and so would be a realistic solution in this case. Repeating the calculation for the material balance scheme gives a result of 7.0. As a guide, the controller is likely to be unstable if the condition number exceeds 15. So, in this latter case, MPC is likely to work but not work well.

Figures 9 and 10 illustrate the different responses. Both are from a simulation of the column that was subjected to a reduction in feed rate. With the energy balance scheme in place, both tray temperatures return to their setpoints within about 40 minutes without oscillation. As expected, the alternative is very oscillatory and takes almost three times as long to recover. But remember, this is just an example; one should not conclude that the energy balance scheme is the better choice for all columns. Indeed, as we saw in *TCE 992*, there are also other options that can outperform both schemes.

## FINAL WORD OF WARNING

The analysis above has only considered steady-state behaviour. Difficult process dynamics can make control infeasible even if the relative gain and condition number are close to 1. The analysis should be used as a technique which excludes a design that will not work, rather than prove that it will. ■

## NEXT ISSUE

In the next issue we'll make a start on compressor control. We'll be covering load control schemes and surge protection. In particular, we'll show how packaged programmable logic controller (PLC)-based schemes can instead be replicated in the DCS.

*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*