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Where has engineering judgment gone?

One of my clients has a dozen or so distillation multi-variable predictive control (MVPC) applications, implemented by one of the major advanced process control (APC) vendors. These applications have not been successful. A substantial investment is at risk, and I was asked to investigate what could be done to repair these controllers. The result of this investigation was not what my client wanted to hear. I have seen and written about many APC failures, but it is unusual to come across APC designs that, even if successful in stabilizing the unit, would run worse than the distributed control system (DCS) strategies they are replacing. Let's look at one of the designs—deisopentanizer (DIP) APC controller—in the hope of exposing certain distillation control misconceptions.

The DIP has 60 trays, a healthy reflux ratio and a standard heat balance control configuration (i.e., overhead drum level control is on the top product). There is a large value difference between IC_5 and NC_5. IC_5 is a high-octane MOGAS blending component, whereas NC_5 is sold as naphtha cracker feed at a lower price. The best way to run the column is:

- Maximize the top product to a purity target of about 10% NC_5. Beyond 10% NC_5, the octane penalty makes MOGAS blending difficult.
- Maximize the reboiler heat duty up to a reflux ratio target of 8:1, this ratio being the economic trade-off between higher reboiler duty versus a diminishing yield gain benefit.

It is possible, of course, to state the control problem differently: as a requirement to control both top and bottom impurities. Although, in my opinion, the latter problem statement is not useful. First, in economic terms, we are trying to minimize loss of IC_5, not to control it to a target. Second, dual composition control is a much more difficult problem and should be avoided unless absolutely necessary.

One could argue that the targets of 10% top contamination and 8:1 reflux ratio are only approximate and could vary with the economics of the day. That is, of course, correct. The APC design should permit operator access and ability to change those targets.

The column has a temperature measurement on tray 12, and, before APC implementation, a DCS tray temperature controller manipulated the reflux. You would think that such a temperature control strategy is pretty clever since tray 12 temperature is a not a bad inference of top NC_5. I would have added an inference model to take into account pressure and column *L/V*. But even without those improvements, with constant pressure and reflux ratio, keeping the tray temperature constant would control top NC_5 near its target.

However, the APC design had eliminated that DCS loop and created a top pressure-compensated temperature (*PCT*) control variable (*CV*) instead. We have nothing against pressure compensation; had the control engineer chosen to compensate tray 12 temperature, the control scheme would not have made things worse. But, in this case, he chose a poor inference instead of a reasonable one. The top *PCT* is a poor inference of top NC_5 because it is quite sensitive to the top C_4 content, a 1%–4% light contaminant that comes with the feed from the upstream debutanizer. At tray 12, the C_4 concentration is much lower and the influence less noticeable.

That was the way the application was initially commissioned, with the main *CV* being top *PCT* pressure and another *CV* was set up as the pressure control valve position constraint. The column had never experienced hydraulic limits. But in the application, reboiler duty was set to maximize—and I suppose the valve limit was established to prevent the reboiler from creating a large amount of vapor that the condenser could not handle. On the second commissioning day, APC managed to flood the column. That was a cool day, and the condenser was able to handle all of the reboiled vapor. Reboiler duty went higher and higher until the column flooded. The remedy implemented to avoid flooding was a differential pressure measurement across the entire 60 trays: *bottom pressure – top pressure,* set up as the third *CV*. Normally, a differential pressure measurement across the whole height of the column is not a good flooding detection tool. But, in this case, there is no real need to detect flooding, only to set a reboiler limit, and the differential pressure is a rudimentary measurement of vapor flow or reboiler heat duty. Why not simply clamp the steam flow? I don't know what that engineer was thinking several years ago.

In this form, the application brought the column to a stable operation, but was unable to precisely control the top NC_5 content. Operators still look at tray 12 temperature and adjust the top *PCT* targets to control top NC_5 content. IC_5 recovery has gone somewhat down, while reboiler steam consumption went up.

What can I say to my client now? They have called in a reputable APC vendor to design and implement many distillation tower applications. The vendor engineer has dreamed up control schemes that not only are inconsistent with unit economics but that also ignore known distillation control principles. The clients were disappointed, though they understood the design mistakes. As luck has it, these clients own some other, much more successful APC applications. That is the reason they still believe in APC and want to commit manpower and funds to repair and recommission the mediocre distillation control applications.

What can we say to the vendors who implement unscientific control schemes? Please, why are you destroying your own livelihood? Can you not nominate knowledgeable people to supervise your projects? **HP**

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