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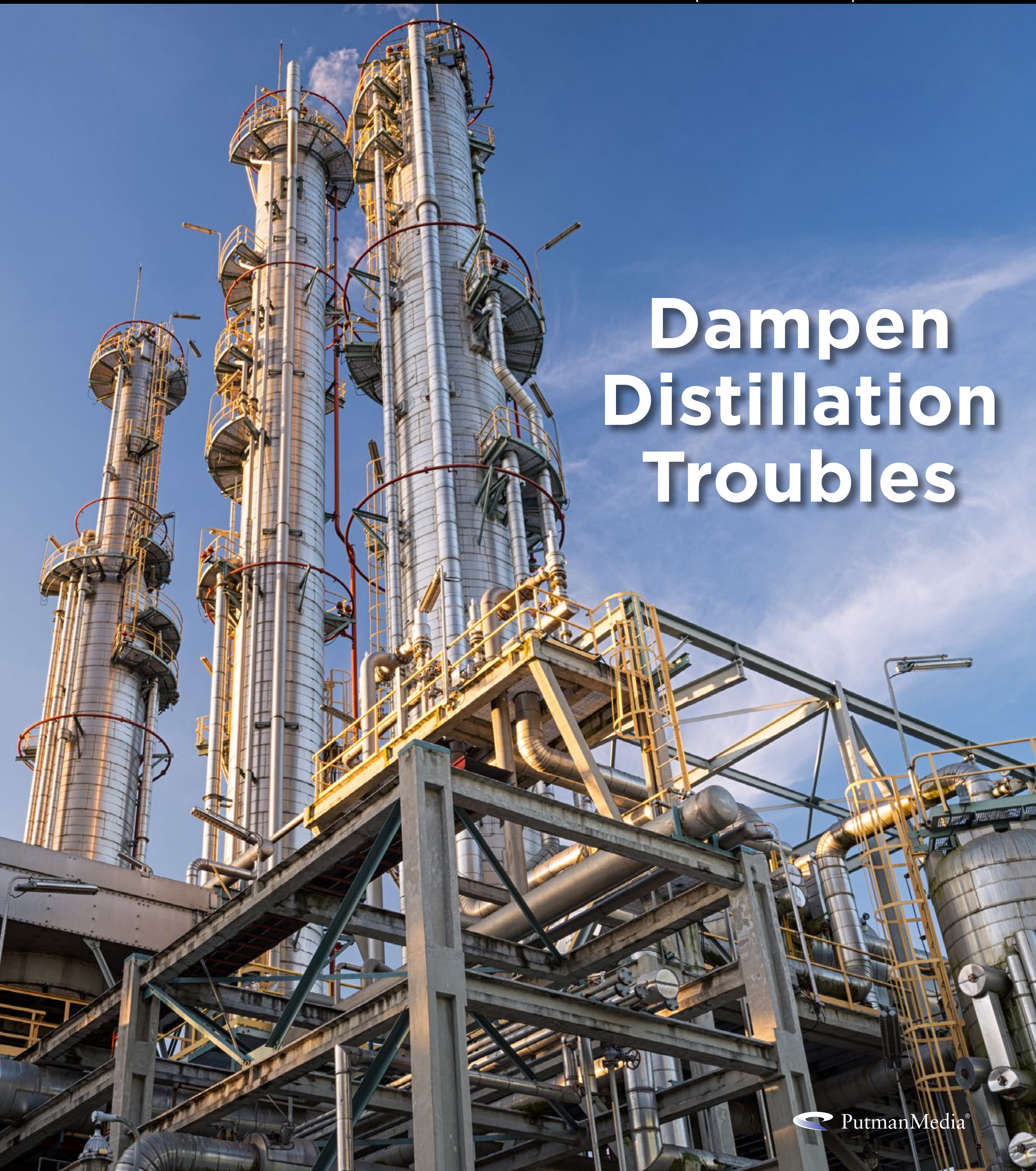

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Improve Distillation Control

Three design changes can reduce energy, improve capacity and prevent off-spec product

By Charles Herzog, Guest Contributor

Designing a distillation system to consistently achieve a given capacity at specified product purities involves more than sizing equipment and pipe. Feed changes, equipment constraints, uncontrolled disturbances, and periodic analyzer downtime can result in significant variations in product compositions (i.e., large standard deviation, σ). For this reason, it's common to set target product purities significantly higher than specifications.

The choice of operating target typically is driven by factors such as:

- Performance record of similar units with a given operating target;
- The magnitude and frequency of anticipated disturbances (e.g., during furnace decoking on ethylene plants);
- Whether off-spec product is flared;
- The priority placed on quality and reputation as a supplier; and,
- Whether customer is willing to buy off-spec product.

If off-spec product must be flared, or an off-spec event requires a plant shut down, a large operating cushion (4σ or more) may

Selling off-spec product at reduced prices can be a bad idea.

be necessary. Often, supplying on-spec product outweighs all other requirements.

A track record of uninterrupted on-spec production enhances a supplier's reputation, leading to a competitive advantage, including higher prices to customers. Conversely, selling off-spec product at reduced prices can be a bad idea. This tarnishes the supplier's reputation and makes it difficult to sell on-spec product at full price. Customers may wait for an off-spec incident to buy at cut-rate prices.

Reduced variation (i.e., smaller σ) helps shift the composition operating target toward the product specification. This decreases heat loads and reflux requirements per unit feed, which improves energy efficiency. It also increases the capacity of the distillation system. Three important design improvements can reduce the variability of primary product composition:

Insulate overhead equipment and piping. This greatly decreases the impact of distillation's most unforgiving disturbance, the summer rainstorm. Insulation isn't a safety requirement on many distillation overhead systems, nor is it necessary for energy conservation in steady-state conditions. However, at the onset of a summer rainstorm, cool raindrops fall onto hot metal equipment and piping, causing a sudden increase in condensing duty and a rapid drop in pressure. Equilibrium liquids on trays vaporize, often leading to off-spec product and several hours of unsteady operation. Operators and the automation systems — even advanced process control systems — often are powerless to keep the system on-spec during and immediately following rainstorms in the absence of overhead insulation.

Increase holdup time for primary product. This is a guaranteed way to reduce product composition variation. A larger overhead accumulator or tower bottom serves as a product protector, dampening temporary upsets, regardless of cause, giving the operator time to make adjustments. Of course, a larger accumulator costs more than a smaller one, but the incremental cost is typically small compared to the cost of a single off-spec incident. Furthermore, a larger accumulator may eliminate the need to buy an off-spec tank.

The oversized accumulator (or bottoms) may be crucial if the distillate is the primary product and the feed is vapor phase. In this case, the tower overhead composition almost immediately reflects any change in feed composition. Similarly, if the bottom is the primary product and the feed is liquid, any change in feed composition goes directly to the bottom product.

Increase size of upstream vessel to dampen feed rate changes. Feed rate disturbances to a distillation system must be addressed. Unsteady feed leads to unsteady product. Increasing the liquid residence time of an upstream drum allows more time for the level controller to dampen feed rate changes. If a smaller drum is exposed to large feed rate changes, more-aggressive controller tuning is required to prevent overfill; this can lead to oscillations.

More opportunities for improving column control and saving energy, including enhanced regulatory process control strategies and the use of inferential process models, are discussed in C. Herzog, "Address Distillation Process Control During Design Phase to Save Energy and Increase Capacity," AIChE Southwest Process Technology Conference, Sugar Land, Texas, October 1–2, 2019. ●

CHARLES HERZOG, guest contributor, can be contacted at chasherz@aol.com

Editor's note: This article was written by a colleague of our regular columnist, Alan Rossiter. Charles Herzog is a process consultant with 40 years of experience in oil refining and petrochemicals, including all phases of plant design, startup and operations.

Resist the Temptation

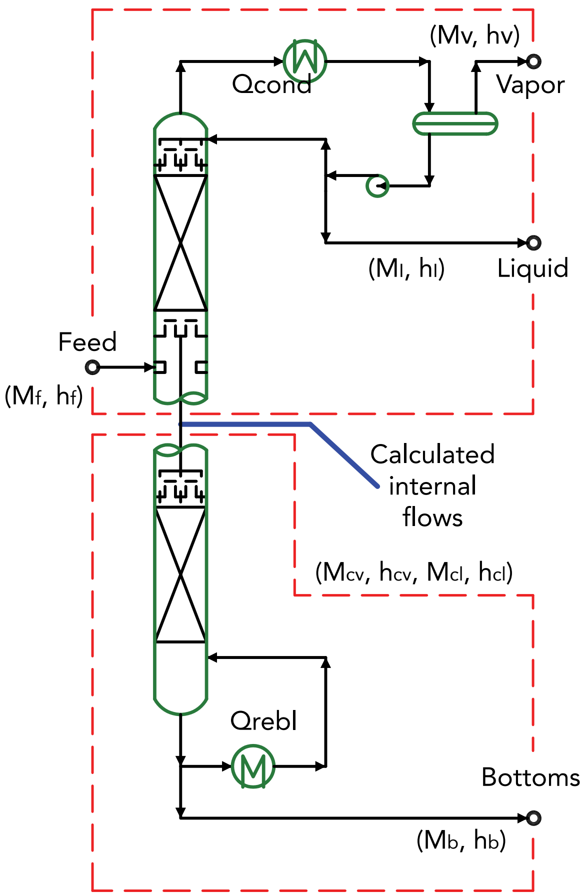
One option to infer internal flow rate in a column rarely makes sense

By Andrew Sloley, Contributing Editor

Some processes reward close control of liquid and vapor rates inside distillation columns. Conventional control approaches use a variation of flow metering based on either imposing a pressure drop inside the tower or drawing the stream out of the tower, metering it, and returning the stream to the tower (see: "Get Some Inside Information," http://goo.gl/g3DNja and "Ease Measurement of Column Internal Flow," http://goo.gl/Aqq2kJ).

Analysis of system fundamentals might suggest a different approach. Figure 1 illustrates a conventional distillation tower. Any heat and material balance (HMB) envelope must have energy in equal to energy out and mass in equal to mass out. Figure 1 shows the HMB envelopes drawn through the tower. An internal vapor stream and an internal liquid stream both cross the HMB boundaries.

With sufficient data, you can solve the heat and material balances to calculate the internal streams, which, in this case, are the mass of the calculated vapor, M_v,



DISTILLATION TOWER
Figure 1. An internal vapor stream and an internal liquid stream both cross the heat and material balance boundaries.

and the mass of the calculated liquid, M_l. You can calculate either stream from analysis of the upper HMB or the lower HMB. The decision to choose between the upper and lower balances depends upon availability of plant data and data accuracy.

The material balance for the upper section is:

$$M_f + M_{cv} = M_{cond} + M_v + M_l + M_{cl} \qquad (1)$$

where M is mass flow, f is feed, cv is the calculated vapor stream, $cond$ is condensate, v is vapor product, l is liquid product and cl is the calculated liquid stream.

The energy balance for the upper section is:

$$Q_f + Q_{cv} = Q_{cond} + Q_v + Q_l + Q_{cl} \qquad (2)$$

where Q is energy flow, which equals the stream enthalpy, h, times its mass flow rate and allows for substituting M_ih_i for the stream energy flows.

With some algebra, we get an equation for calculating the internal liquid rate, M_{cl}, from the top HMB:

$$M_{cl} = [Q_{cond} + M_{cv}(h_v - h_{cv}) + M_l(h_l - h_{cv}) + M_f(h_{cv} - h_f)]/(h_{cv} - h_l) \qquad (3)$$

and an equation for calculating the internal liquid rate from the bottom HMB:

$$M_{cl} = [M_b(h_b - h_{cv}) - Q_{rebl}]/(h_{cl} - h_{cv}) \qquad (4)$$

where b is bottoms and $rebl$ is reboiler.

None of these HMB calculations violate any engineering basics. However, while accurate, are they useful?

Most occasions requiring tight control or measurement of internal liquid rates arise from large incentives from either:

Arbitrary offsets indicate a fundamental flaw in the logic of using the HMB approach.

- operating at close to minimum reflux; or
- operating at close to minimum or maximum equipment capacity.

Drivers for operation at close to minimum reflux include high energy prices and extremely different values between the overhead and bottoms. When the product value differences are high, the purpose of the distillation column is to remove a trace contaminant but with a minimum slip of the high-value product into the low-value one. This is a process consideration driven by system behavior (relative volatility, compositions, stages available) and economics.

Drivers for operation at close to equipment limits normally are capacity and equipment performance. Flooding sets an upper capacity limit on equipment. For trays, either vapor handling or liquid flow regime on the trays may determine the lower capacity limit. For packed beds, liquid distribution quality at low liquid rates usually establishes the lower limit. This is a hardware consideration driven by the installed equipment (diameter, device type as well as fabrication and installation tolerances).

Let's now consider whether the flow rate equations actually are useful for controlling the stream rates very close to an "optimum" value. Equation 5 summarizes the situation:

Flow rate = (large number - large number)/ (medium number - medium number) (5)

Both subtractions include potential errors in measuring composition, flow rate and temperature.

You must calculate stream enthalpies based on stream compositions, flow rates and temperatures. Calculation of condenser duty and reboiler duty must be based on utility flow rates and temperatures. My experience is that as long as the column products remain on-specification the effect of composition errors is small on the calculation.

In contrast, the effect of flow-rate and temperature measurement errors often is large. Routinely, the possible error exceeds 100% of the calculated value. Often, an arbitrary offset is added to prevent negative flow rates from being calculated. These arbitrary offsets indicate a fundamental flaw in the logic of using the HMB approach.

All control systems that use the difference between two large measured values as a target require high-precision measurement. For example, if the internal liquid rate is 10% of the feed rate and the control objective is to restrict the variation in the internal liquid rate to 1% of the feed rate, what's the likelihood of success? It's a rare plant flow meter that has a precision higher than 1% of the flow rate. Most standard thermocouples offer roughly 0.75% accuracy. The combined impact of flow-rate and temperature measurement errors makes precise control difficult.

I routinely see controllers with liquid rates targeted at 3.0±0.3% to 5.0±0.5% of the feed rate. The calculations routinely generate negative flow rates. They can't have a relative accuracy of 10%!

The HMB calculation for internal liquid rates obeys the laws of physics. However, the situation that makes knowing the internal rates most important usually is when the rates are very low. This demands extremely precise flow and temperature measurements. Plant instrumentation rarely can meet these requirements. Unless every other choice simply is unacceptable, don't opt for HMB calculation methods to infer internal liquid rates. If you must use them, conduct statistical checks to confirm their suitability and expected value.

ANDREW SLOLEY is a contributing editor for *Chemical Processing*. Email him at ASloley@putman.net.

Preclude Packed-Column Problems

Carefully dealing with liquids can head off hassles

By Andrew Sloley, Contributing Editor

A fractionating column requires proper mixing of the entering liquid and the liquid already inside. For columns with trays, this rarely poses a problem. Trays are excellent mixing devices. In contrast, columns with packing don't always provide adequate liquid mixing and, as a result, can suffer performance and reliability problems.

Fractionation depends upon heat transfer and mass transfer between the liquid and vapor phases. Heat transfer requires a temperature difference between the phases, while mass transfer requires a composition difference. The basic assumptions for determining packing performance assume uniform flow distribution, temperature and composition across the tower cross-section.

The packing height to provide the equilibrium stages needed for a separation are determined using a height equivalent per theoretical plate or stage (HETP). Operating results from laboratory, pilot plant,

Incomplete mixing or improper liquid distribution may cause many problems.

and industrial equipment can generate data for determining HETPs. These data should account for the non-ideal performance of real packing and distributors, installation practices, and other factors.

Trouble arises when assumptions on equipment performance differ from reality. Fractionation calculations assume perfect mixing of the entering liquid and the liquid already in the column. Incomplete mixing or improper liquid distribution may cause many problems.

Different services may have many requirements. However, four common ones are:

- a subcooled liquid feed with heat transfer below the feed as the major objective;
- a subcooled liquid feed with fractionation below the feed as the major objective;
- a bubble-point liquid feed; and
- a mostly liquid feed containing a small amount (by mass) of vapor.

This list shows the services in order of increasing severity. A guideline offered for a more difficult service will work for the easier ones as well.

Heat transfer is a more forgiving service than mass transfer. By themselves, spray headers are effective heat transfer devices. If adequate pressure is available for sprays, allowing the liquid to enter the tower through a spray header may suffice. If the feed liquid rate is much higher (twice or more, for example) than the internal liquid rate, this system will meet the requirements for heat transfer. A reasonable amount of mixing will occur as internal liquid falls from the packed bed above into the feed liquid spray.

Fractionation requires more care in mixing the liquids. Never use sprays for mass transfer services. Instead, opt for a collector for the liquid inside the tower, a feed pipe for the liquid entering the tower, and an internal redistributor for the total liquid to the packing below. Fractionation works best with a smooth flow rate and even composition across the entire tower cross-section.

The piping arrangement must avoid having the feed liquid or internal liquid in only one spot across the tower cross-section. The external and internal liquids rarely have the same composition. Localized composition gradients will make fractionation less effective. In extreme cases, portions of the packed bed may dry out completely due to composition gradients.

If the feed is reliably subcooled, mixing it with the internal liquid will not form a vapor. This makes the equipment simpler as only liquid must be handled.

With bubble-point feeds, mixing the internal liquid with the feed liquid may cause some vaporization. Vapor inside gravity distributors may dramatically decrease their capacity. Because gravity liquid distributors have relatively narrow operating ranges, reduced capacity can seriously affect tower performance.

The most reliable way to de-gas a mixture is to mix the feed and internal liquid on a collector tray. The collector has sufficient volume to allow for de-gassing. An internal overflow then sends liquid to the distributor below the feed entry.

Increasing the amount of gas in the feed may mandate more complex approaches.

If the feed contains gas before entering the tower, the best approach is to de-gas the feed, then mix it with the internal liquid. This may require a two-stage feed gallery.

Keep two key points in mind:

1. Pay attention to the internal liquid distribution device below the feed entry.
    - Spray headers may be able to tolerate some reduced mixing.
    - Gravity distributors require proper mixing of the internal and external streams.

2. Gas formation will mandate more-elaborate feed systems.
    - Gas may form upon mixing the entering liquid with the internal liquid.
    - The feed itself may contain gas.

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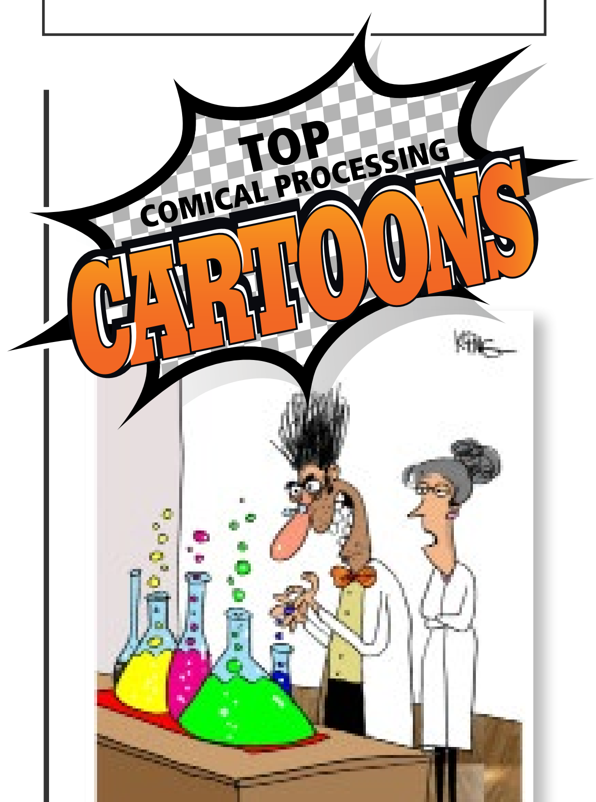

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