

Predictions for APC in refinery gas plant

Inferential models enable the development of advanced process control for a FCC gas plant

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FCC units convert three-quarters of their feed into light products: naphtha, LPG and gas. These are first separated from heavier products in the main fractionator, and then taken into the gas concentration plant for further separation. FCC LPG is olefinic and therefore of high value. Unsaturated butane is an alkylation reaction ingredient, whereas unsaturated propane is rich in propylene and usually separated further in a propylene splitter. Heavy penalties are paid

for separation mistakes, and that is an opportunity for advanced process control (APC), provided such APC is able to control product separation precisely.

Figure 1 shows a simplified gas concentration process diagram. Liquid feed is taken into a stripper, whose task is to strip out ethane; vapour feed is taken into an absorber. There is a substantial amount of LPG in that vapour and absorber lean oil captures the LPG, returning it to the feed drum. C_2 is

also absorbed in the process, and the stripper strips out the C_2 and returns it to the absorber. Stripper bottoms, by and large free of C_2 , go into the debutaniser, where LPG is separated from naphtha. A depropaniser further splits LPG into propane and butane components. Debutanised C_5+ naphtha is used as absorber lean oil, and hence the absorber overhead contains a small amount of C_5 material. Downstream of the absorber, a sponge absorber absorbs the C_5+ material and

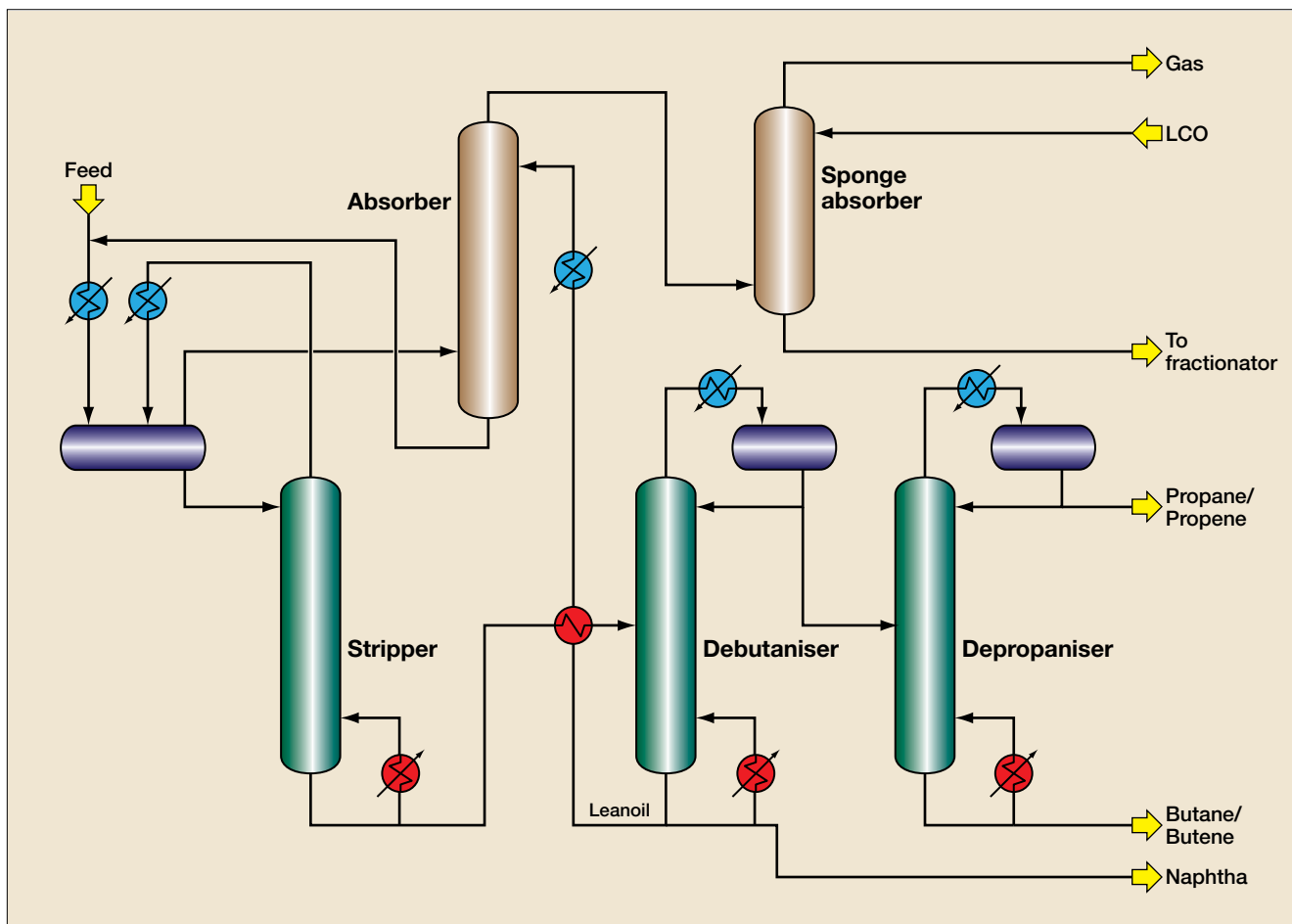


Figure 1 FCC gas plant configuration

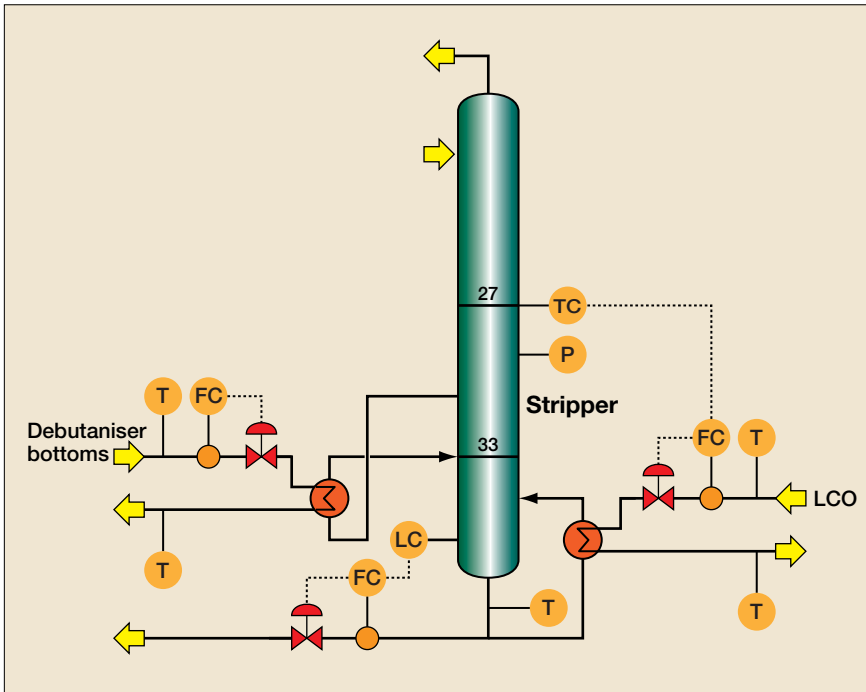


Figure 2 Stripper configuration

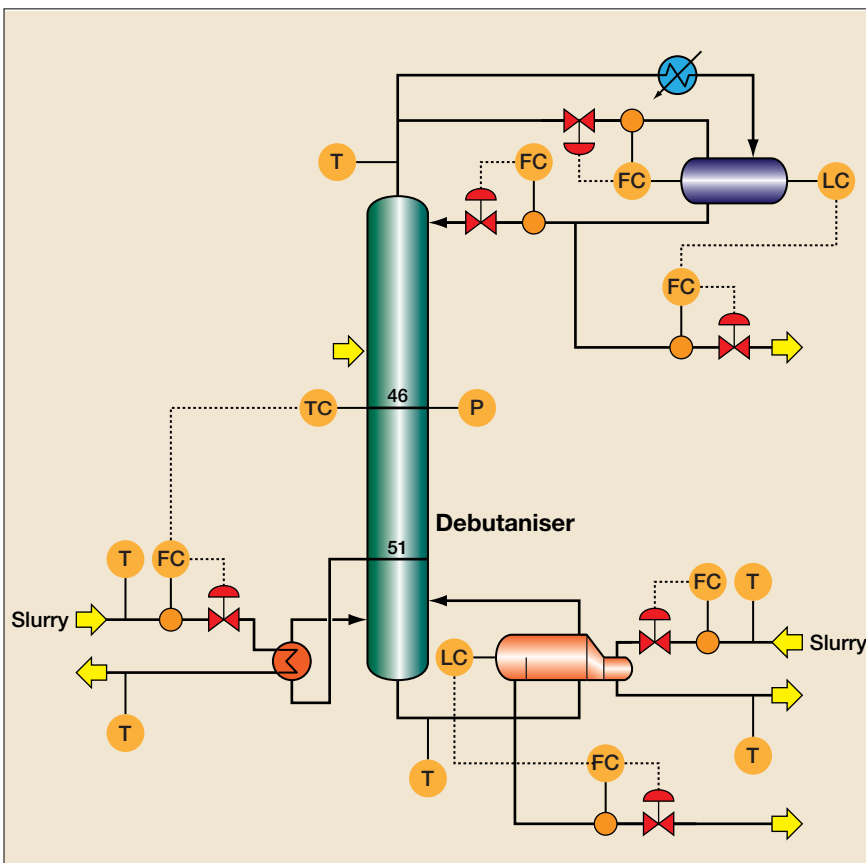


Figure 3 Debutaniser configuration

recycles it back to the main fractionator.

This gas plant belongs to a US refinery. The unit design permits good separation, in the order of 0.5% cross product contamination, except that for the operator it is not easy to determine correct unit

settings to take full advantage of equipment capability. In a steady laboratory test results, the operator should be able to optimise propane recovery and butane separation, but steady state operation is illusory. FCC reactors are weather

sensitive and the same is true of the gas plant. Absorption is temperature dependent and the coolers shown in **Figure 1** in blue are usually air coolers. Test results from a lab sample taken at 06:00 come out at 10:00, after the sun is up and process conditions are different. Lacking information, operators resort to conservative settings, accepting loss of propylene to fuel gas. A loss of 2-5% of the propylene production to fuel gas is not uncommon. The debutaniser aims to remove C_4 , leaving 0.5% C_4 in naphtha, but many sites experience 2-4% C_4 slippage, causing a loss of alkylation unit feedstock. C_4 product can reach 2-3% C_3 contamination, taking capacity and causing operational problems in the alkylation unit.

Is the DCS control structure adequate?

The question applies to **Figures 2** to **4**. Consider first the depropaniser of **Figure 4**. A tray temperature controller (TC) manipulates the reboiler whereas operators manipulate reflux flow. Is that a good structure? Distillation control is a two-by-two problem, the two degrees of freedom being 'cut' and 'fractionation'. 'Cut' is essentially distillate yield, to be manipulated when one product is too pure, the other too contaminated. 'Fractionation' is column loading, which typically depends on the reflux. Fractionation is to be manipulated when both products are too pure or too contaminated. The depropaniser tray TC is a cut control device. When the tray is too cold, the TC will increase reboiling, sending more vapour to the overhead drum and eventually to the distillate, increasing the cut. It helps that tray temperature is also a rudimentary inference of product purity. Such DCS structure is convenient because there are essentially no interactions between cut and fractionation. Changes of cut do not alter the reflux. Changes of reflux and column loading have some dynamic effect but do not alter the steady state tray temperature, and only minimally alter the cut.

The debutaniser of **Figure 3** conceptually has the same structure. The tray TC manipulates the reboiler while operators control fractionation by manipulating reflux. The second reboiler adds a degree of freedom, and the loading balance between those two reboilers is another APC consideration.

Absorber-stripper cut control is also handled by a tray TC. The absorber-stripper fractionation handle is not shown in **Figure 2**. Fractionation is controlled in this system not by reflux but by lean oil, and again the operator is in charge of lean oil flow and thus fractionation.

Depropaniser and debutaniser pressures are controlled by manipulation of condenser hot vapour bypass, a method commonly used with fully condensed overhead product. When the bypass is open, the condenser partially fills with liquid, condensation is reduced, and the condensate becomes sub-cooled. Closing the hot vapour bypass valve increases condenser pressure difference, draining the condenser, increasing condensation and thus reducing column pressure. While there are possibly better ways to design hot vapour bypass arrangements, the resulting pressure control is stable.

Absorber-stripper pressure floats on the sponge absorber pressure, which is controlled on the off-gas stream. Overall, we conclude that this gas concentration unit DCS structure is quite good.

Can we obtain high fidelity inferential models for this FCC gas plant?

If high fidelity inferential models are possible, then APC becomes simple as follows:

- Manipulate the stripper tray TC setpoint to control C_2 in propylene. Manipulate lean oil to control absorber-stripper loading. That is a two-by-two dynamic problem, non-interacting and easily solved.
- Manipulate the debutaniser tray TC setpoint to control C_4 in naphtha; manipulate reflux for column loading; another easy two-by-two set-up.
- Manipulate the depropaniser

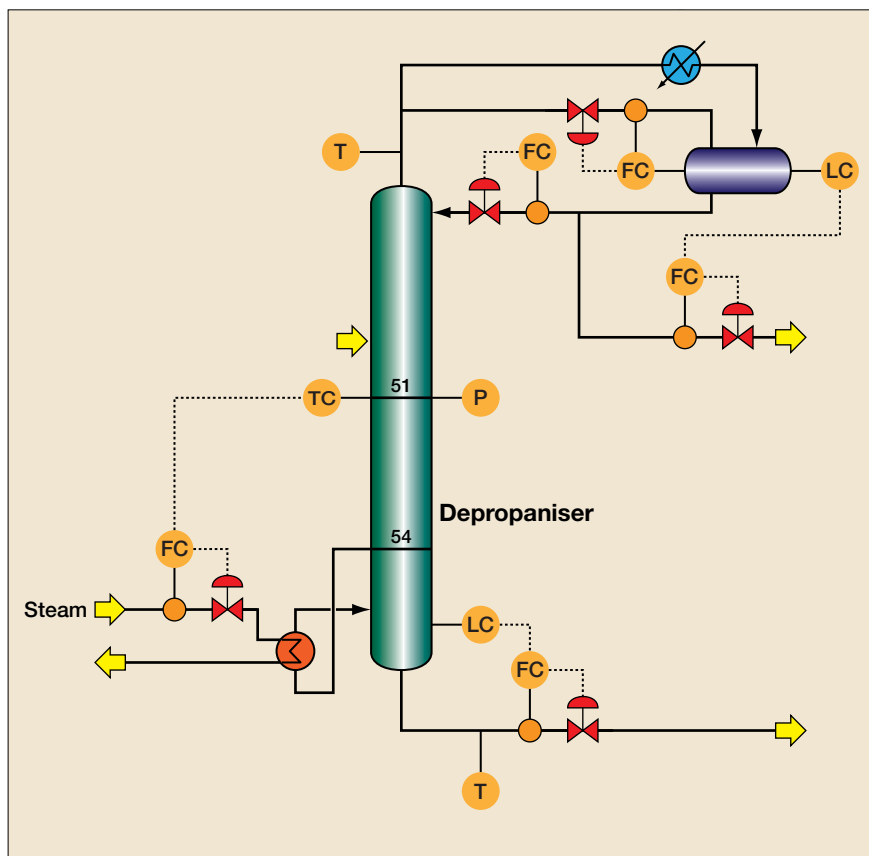


Figure 4 Depropaniser configuration

tray TC setpoint to control C_3 in C_4 . Manipulate reflux for column loading.

- Keep the sponge absorber on minimal loading. There is no desire to absorb much C_2 material and return it to the main fractionator, because such recycle increases the wet gas compressor load and takes up FCC capacity.

We have studied first principle distillation inferential models for

many years and have developed a methodology called GDS (generalised distillation shortcut). GDS inferential model performance has been reported in the early 2000s.^{1,2,3,4} Over a decade and about 200 applications later, we find it informative to publish again, first because the technology has improved over time, and second there are unusual reboiler configurations that we have been able to address. Shown in **Figure 2**, the stripper has two reboilers, one at the bottom and another one feeding off the lowest tray. The upper reboiler operates at lower temperatures and that permits the use of naphtha as the heating fluid. The debutaniser (see **Figure 3**) also has two reboilers. One siphon reboiler is fed from the lowest tray, and a second kettle reboiler is fed from the bottom product. Such complications make the inferential modelling task more difficult.

GDS theory

GDS uses DCS measurement data around a specific section of a distillation column to come up with an inferential model. The green area of

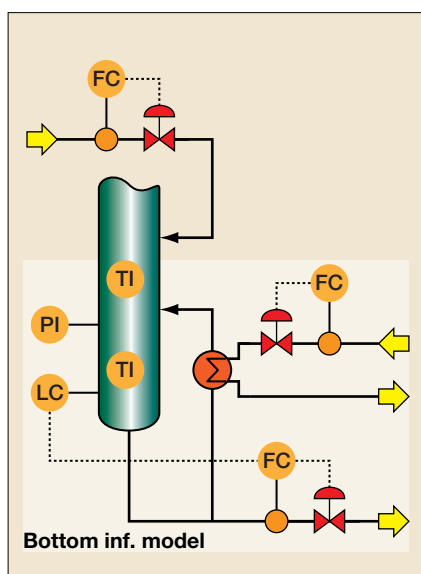


Figure 5 Stripping section GDS input data

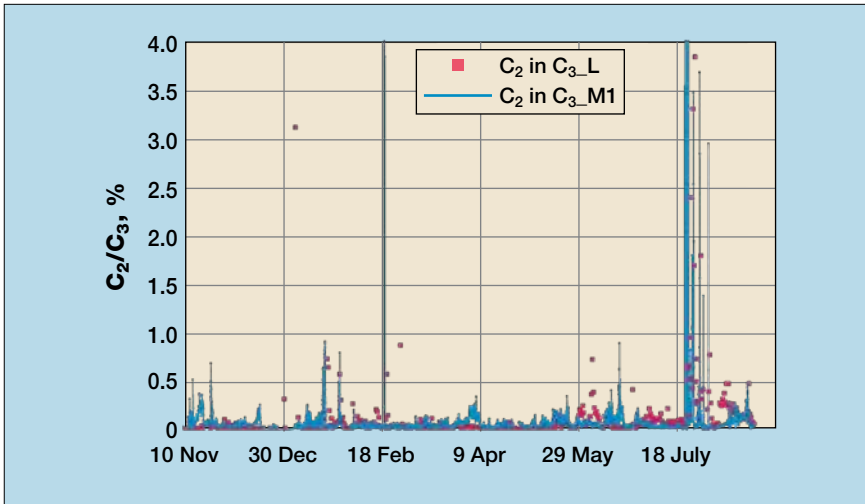


Figure 6 Trend of C_2 in C_3 over nine months

Figure 5 shows a stripping section example. The inputs are pressure, temperatures plus enough measurements to permit vapour and liquid traffic calculation by heat balance. A minimum of two temperature points are needed, one at the bottom and the other on a tray, distant enough from the bottom to have a light key component content of 10% or so.

The model makes use of Colburn's method,⁵ which estimates the ratio of vapour composition on tray N (the upper temperature measurement tray) to bottom liquid composition. It is a convenient closed form calculation, which correctly takes into account the non-linear effects of column loading and number of trays. For a simple stripping section, the Colburn ratio takes the following form:

$$R_i = 1 + (Z_i - 1) * (K_i - 1) / (U_i - 1)$$

$$K_i = Y_{tray_i} / X_{tray_i} \text{ (Component } i \text{ volatility)}$$

$$R_i = Y_{tray_i} / X_{bot_i} \text{ (Colburn ratio for component } i)$$

(V/L) = vapour to liquid molar flow ratio, calculated by heat balance

$$U_i = K_i * (V/L) \text{ (the effect of column loading)}$$

N = number of theoretical trays in section

$$Z_i = U_i \wedge N$$

Following calculation of volatilities, column loading and Colburn ratios, GDS simplifies the composition into four components and solves a set of four equations with four unknowns. The following

is a GDS equation set for a debutaniser stripping section. The problem set-up has four unknowns to describe column bottom composition:

- NC_4 is the light key component, to be kept at a level of 0.5-1%
- C_5 is the heavy key component, volatile on all stripping trays
- C_6 is a so-called heavy-heavy key component, volatile in lower trays only
- C_{7+} is extra-heavy key, assumed non-volatile even inside the reboiler.

To estimate the bottom composition, GDS comes up with four equations. In their simplest form, the four equations are:

- Bottom mass balance

$$\sum (X_{bot_i}) = 1 \quad [1]$$
 (Sum of bottom molar fractions = 1)

- Reboiler equilibrium

$$\sum (K_{bot_i} * X_{bot_i}) = 1 \quad [2]$$
 (Sum of reboiler vapour molar fraction = 1, that indicates equilibrium)

- Section separation

$$\sum (R_i * X_{bot_i}) = 1 \quad [3]$$
 (Sum of tray N vapour molar fraction = 1)

R_i is the Colburn ratio for component i , the ratio between vapour composition on tray N to bottom liquid composition.

- Tray equilibrium

$$\sum [(R_i / K_{tray_i}) * X_{bot_i}] = 1 \quad [4]$$
 (Sum of tray N liquid molar fraction = 1)
 ($1 / K_{tray_i}$) is the ratio of vapour to

liquid composition on tray N. Hence equation 4 is a tray equilibrium equation.

While the four-by-four matrix coefficient calculations are non-linear, the four resulting equations above are linear and a solution is guaranteed. The calibration procedure for this model involves adjusting tray efficiency for the total section and for tray N (affecting Equation 4) to obtain a good fit between model and lab results.

C_2 in C_3 inference

If the depropaniser overhead drum of Figure 4 were in vapour-liquid equilibrium, then inference of C_2 in C_3 would have been easier. The more C_2 in the drum, the lower the drum temperature, and that phenomenon would have facilitated a reasonable C_2 in C_3 model. It is feasible to design overhead drums to be in vapour-liquid equilibrium, but that is not what we find in this unit. Condensed liquid is sub-cooled and hot vapour bypass is superheated. A similar set-up exists at the debutaniser overhead drum, precluding a model of C_2 in LPG. That brings us to the stripper as the only equipment where an inferential C_2 slippage model is feasible.

GDS has not originally been configured for the staged dual reboiler arrangement of Figure 2. That is not a theoretical limitation, but given that those reboilers are only one tray apart, we decided to lump the two together using a standard GDS configuration for the stripper between the bottom and tray 27. There is imprecision in this assumption in that real vapour/liquid in the lower reboiler is low, hence the lower reboiler, which would typically act as an added separation stage, cannot be fully effective. That imprecision is partly addressed by calibration because in any case average tray efficiency is adjusted to match model readings versus laboratory testing.

The four stripper bottom GDS components are:

- C_2 is light key, to be kept at a level of about 0.05% in the stripper bottom, as the target for C_2 in propylene is 0.5%

- C_3 is heavy key
- C_4 is heavy-heavy key
- C_5+ is extra-heavy key, non-volatile in the stripper conditions.

Considering that all of the C_2 material in the stripper bottom would be concentrated by a factor of about 10:1 into the propane product, our model goes further to estimate concentration using flow measurements, to arrive at a C_2 in C_3 inference.

Figure 6 trends C_2 in C_3 inference (C_2 in C_3 _M1) versus lab results (C_2 in C_3 _L) over a period of nine months. The stripper operation most of the time seems conservative, with hardly any C_2 penetration. That is actually not good data for calibrating an inference, and we have to look at times when there was significant penetration to make model calibration judgments. Beyond that inconvenience, zero C_2 penetration is not called for and it can only be achieved at the cost of significant loss of propylene to fuel gas. Economically, we should see penetration of C_2 in C_3 , in the order of 0.4-0.5%. Eventually, operators would trust APC to control C_2 penetration better, and at that time we intend to recalibrate the inference.

Once C_2 slippage is under control, how would we go about minimising propylene loss at the absorber without an inference of propylene loss? The presence of some C_2 in LPG actually guarantees that, with proper loading of the absorber-stripper system, propylene loss would also be minimised. Typically, a good lean oil flow, about equal to the LPG flow, would reduce the C_3 in off-gas to 1% or so. Bear in mind that, in addition to flow, the colder the lean oil the better it works, and APC designs should actually vary lean oil ratio with absorber temperature, up to a loading constraint such as flooding.

C_4 in naphtha and naphtha RVP inference

For the debutaniser stripping section, our approach is similar to that of the de-ethaniser, combining the two reboilers into one in a GDS model. We expect a similar degree of imprecision due to this simplify-

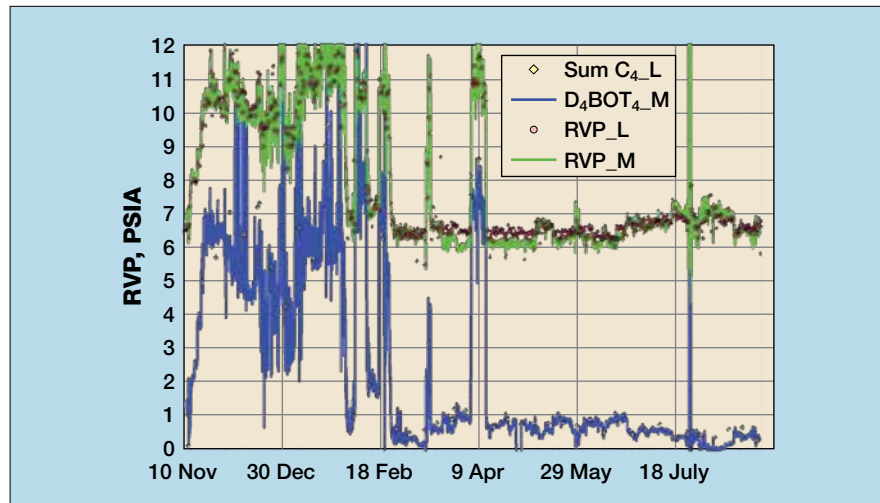


Figure 7 Trend of C_4 in naphtha and naphtha RVP over nine months

ing assumption, and partly the imprecision would be addressed by calibration. As is commonly done, GDS again simplifies debutaniser bottom composition to four components –

- C_4 is light key, actually nC_4 because iC_4 does not normally penetrate into naphtha
- C_5 is heavy key. All C_5 isomers are lumped together into one component
- C_6 is heavy-heavy key. Again, all

Tray measurements on distillation columns permit the construction of first principles, high fidelity inferential models

C_6 isomers are lumped together

- C_7+ is extra-heavy key, non-volatile in debutaniser conditions.

Figure 7 trends C_4 in naphtha inference (D_4BOT_4 _M) versus lab results ($SumC_4$ _L) over nine months. Those are the blue line and magenta diamonds, and we can see that inference and lab are in agreement. Calibration is somewhat influenced by our desire that the model would be most reliable around 0.5-1%. Looking at operational targets, we see that for much of the time operations attempts to control C_4 in naphtha to less than

1%, though, to our surprise on occasions, mostly during the winter months, the content of C_4 in naphtha is actually maximised to RVP constraints, and at those times the content of C_4 in naphtha climbs to 7%. In our opinion, sending olefinic C_4 into the gasoline pool instead of into the alkylation unit is at odds with refinery economics, and gasoline RVP correction should be accomplished by injection of saturated nC_4 . Our argument is based on generic refinery economics and perhaps there are local variations.

Accepting that RVP is sometimes a constraint, we have added an RVP inference (RVP _M). A trend of RVP inference versus lab (RVP _L) is also shown in **Figure 7**. Those are the green line and orange circles. The inferential fit is quite good.

As for the contamination of C_5 in C_4 , normally a good target is 0.5 to 1%. In that range, the impact of C_5 on the alkylation unit would be minimal. We often are able to infer C_5 in LPG, but this debutaniser does not have tray temperature readings in the rectifying section to permit such an inference. Still, as in any distillation column, controlling bottom purity while loading up the column would guarantee reasonable top purity. APC should aim for a reflux to product LPG ratio of 1:1.

C_3 in C_4 model

Tray temperature measurement in the depropaniser stripping section permits a standard GDS model structure with only three meaningful components:

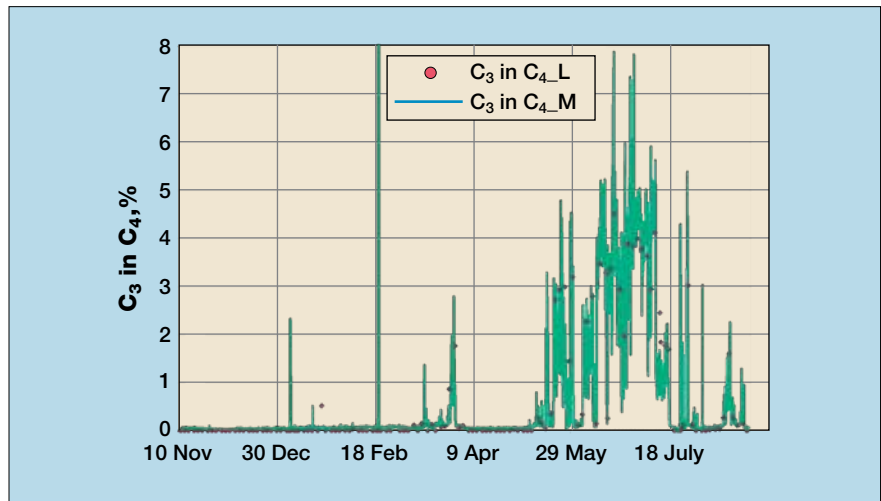


Figure 8 Trend of C_3 in C_4 over nine months

- C_3 is light key, to be kept at a level of about 1%; higher levels may cause difficulties in the alkylation unit
- C_4 is heavy key. All C_4 isomers are lumped together into one component
- C_5 is heavy-heavy key. That is normally iC_5 penetration from the debutaniser.

Figure 8 trends C_3 in C_4 inference (C_3 in C_4 _M) versus lab (C_3 in C_4 _L) over nine months and it illustrates the best inferential agreement. That is not surprising as the depropaniser GDS model incorporates only three clean components and a single siphon reboiler. Looking at how this column is operated, normally C_3 in C_4 is kept very low but occasionally contamination may reach 4%. The refinery in question processes olefinic C_3 in a polymerisation unit and while our instinct would call for a 1% target, the alkylation unit can digest some C_3 and a higher C_3 slippage target could therefore be justified. In any case, with the help of this inference alkylation feed composition can be controlled at the desired value.

Considering propylene purity, this polymerisation unit feed has no special specifications, except that economically we do want to send olefinic C_4 to the alkylation unit. APC should aim at a reflux to propylene product ratio of about 1:1 again, to control the amount of C_4 in C_3 to 1% or less.

Conclusions

We have demonstrated that tray

measurements on distillation columns permit the construction of first principles, high fidelity inferential models. The tray measurements shown in this article are actually not at optimal locations, being too close to the bottom, but still the inferences trend well against the lab. Using GDS models, we can apply APC to FCC gas plants or indeed other gas plants and a wide variety of distillation equipment.

References

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