CONTROL OF CRUDE FRACTIONATOR PRODUCT QUALITIES DURING FEEDSTOCK CHANGES BY USE OF A SIMPLIFIED HEAT BALANCE

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Abstract

This paper describes a technique, which applies a heat balance around the crude fractionator to predict changes in the feedstock volatility. Volatility changes are then fed forward to the sidestream yield controllers, significantly improving sidestream quality control during the several hours following feedstock switches. The application was tested at one of Exxon affiliated refineries with good results.

I. Summary and Conclusion

A computer control scheme capable of stabilizing the pipestill sidestream qualities during crude switches was developed. The novelty of this scheme is that it does not use any operator input data and it does not require a prior knowledge of the crude TBP curve, an important advantage because crude characteristics are often not known or only approximately known.

Conventionally, refineries deal with crude changes by applying an engineering analysis to the crude oil to be run next, and determining off line what the new yields are likely to be. Then, when the actual switch takes place, the operator does his best to approach the off line guidance. However, uncertainties as to what the new yields should really be and what the proper timing is, dictate conservative draw off rates, resulting in less naphtha and less middle distillates than need be, and costly disturbances in downstream units.

The crude switch feedforward scheme aims at eliminating much of the yield loss by estimating the volatility of the crude and by automatically changing the sidestream yields whenever changes of volatility are detected. The application performs volatility estimation and feedforward in three steps as follows:

1. A simplified heat balance around the crude tower to calculate the fraction of crude evaporated in the flash zone.
2. Estimation of the True Boiling Point (TBP) curve, taking into account tower temperatures and partial pressures.
3. Feeding forward sidestream flow corrections in proportion to the volatility changes to maintain the end point of each sidestream constant. (We define volatility as the reciprocal of the TBP slope, i.e. % yield per degree of distillation range).

As of today the application has been tested on about one hundred crude switches and has shown consistently capable of keeping the sidestream end points and/or cloud points within 2°C of targets. This is an 80% improvement over deviations as high as twelve degrees experienced before implementation of the application. Also, the duration of giveaway operation has shrunk from about four hours to about two. Figures 7,8, and 9 are example trend plots of the sidestream analyzer reading during a switch from light to heavy crude, showing how well the sidestream specifications are kept with and without the application.

II. <u>Background. The Need for Improving Yield Control During Crude Switches</u>

As is so typical in many refineries, the refinery where this control strategy was applied has six crude tanks and it operates on about 20 crudes regularly. Thus, each tank may contain a mix of crudes with different distillation curves and different specific gravities. Mixing in the tank is poor, and when a tank starts discharging its content to the pipestill there is no way to foretell what the sidestream yields should be. Averaging the crude assays can provide a rough estimate of the yields, but stratification causes the bottom layer, which flows first into the pipestill to often be heavier than expected.

Furthermore, the pipestill takes its feed via two crude headers. During each crude switch the operator changes the header connections from the previous two tanks to new tanks. At that time the operator's main concern is to make sure there is no plugging in the newly connected headers. He cannot pay attention to the problem of timing his actions in such a way as to minimize quality swings. And the swings are large because:

- Once the change over is made, the time at which it would start affecting the pipestill is not known to the operator.
- The new crudes in the headers do not reach the pipestill at the same time. This may (and usually does) cause the feed to temporarily be much lighter or much heavier than either of the steady state values.
- Whenever possible, static headers are filled with light crude to prevent freezing.

III. <u>Advantages of Heat Balance as a Tool for estimating Crude Volatility</u>

Suppose that during a switch from heavy to light crude the following key pipestill operating conditions are kept constant (and bear in mind that this is not an exotic demand, most of the parameters below vary little from one crude to another):

- Tower top temperature
- Tower flash zone temperature
- Tower pressure
- Stripping steam ratios

Then, when light crude reaches the pipestill, the amount of vapor in the flash zone increases, and tower cooling load increases proportionally. I.E. the heat duties of either the reflux condenser or pumparound circuits must increase instantly to condense the additional vapor, otherwise the tower pressure and temperatures cannot be kept constant. This change of cooling load is independent of whether the side streams are increased or not. If the side streams are not increased the extra vapor must still be condensed and returned to the bottom as overflash. Thus, a heat balance based method offers two advantages. First, it is timely! The cooling load goes up at the same time that the lighter crude starts entering the tower. Second, the cooling load is roughly proportional to the amount of flash zone vapor. This being the case, even a simple feedforward scheme that would increase the side streams in direct proportion to the tower cooling load could do a reasonable job.

However, the cooling load in itself should not be used directly as a feed forward signal, because it is affected not only by crude volatility but also by operational variations which are not the result of crude changes, for example: tower temperatures, pressure or stripping steam ratios. A more precise feedforward scheme, based on estimation of crude volatility will be described next.

IV. <u>Crude Volatility Calculation</u>

Figure 3 shows the section of the pipestill column above the flash zone; this is the heat balance envelope. Enthalpy comes into the section with the flash zone vapor, and leaves with the products, and via the three cooling circuits. The basic heat balance equation is:

$$E_f - \sum_{i=0}^{5} E_i = \sum_{k=0}^{2} Q_k \qquad (1)$$

Where
- E_f is the flash zone vapor enthalpy
- E_i is the enthalpy leaving with product i. There are six products as follows: Overhead (i=0), four sidestreams (i = 1, 2, 3, 4), and overflash (i = 5).
- Q_k is the cooling load on cooling circuit k. There are three circuits as follows: Reflux Condenser (k=0), Top Pumparound (k = 1) and Mid Pumparound (k = 2).

The enthalpy of the flash zone vapor is a sum of the enthalpies of the components in the flash zone. Standard process engineering procedures consider each would-be-product as a “component”, and the enthalpy of each component is a sum of the sensible and latent heat required to bring it to that temperature in a vapor phase. The summing is done by equation 2.

$$E_f = \sum_{i=0}^{5} [H_{vi} + C_{pvi} * (T_f - T_{Ri})] * F_i \qquad (2)$$

Where
- F_i is the flow of product i
- T_f is the flash zone temperature
- T_R is a reference enthalpy temperature at which the enthalpy of liquid is taken as zero. Normally enthalpy tables use T_R = 0°F, T_R = 0°C or T_R= -273°C, but theoretically any reference would do, as long as it is consistent. It is also

possible to use different reference temperatures for different products, hence T_{Ri} is the reference enthalpy temperature for product i.

- H_{vi} is the heat of evaporation of product i at the reference temperature T_{Ri}.
- C_{pvi} is the specific heat of vapor product i at constant pressure.

The enthalpy leaving with the vapor overhead product is also a sum of latent and sensible heat.

$$E_o = [H_{vo} + C_{pvo} * (T_o - T_{Ro})] * F_o \qquad (3)$$

Where:

- T_o is the tower top temperature
- H_{vo}, C_{pvo}, T_{Ro}, F_o are as defined under equation 2 for i = 0.

And for the enthalpy leaving with the liquid products only sensible heat needs to be considered.

$$\sum_{i=1}^{5} E_i = \sum_{i=1}^{5} C_{pli} * (T_i - T_{Ri}) * F_i \qquad (4)$$

Where:

- T_i is the temperature at which product i leaves the tower.
- C_{pli} is the liquid specific heat of product i

A look at equations 2,3 and 4 shows that it is convenient to choose for each product a reference enthalpy temperature equal to the temperature at which that stream is exiting the tower, i.e.

$$T_{Ri} = T_i; \qquad i = 0,1,2,3,4,5 \qquad (5)$$

This can be done without a loss of generality as long as the heat of evaporation H_{vi} is selected at the proper temperature T_i.

Such choice of reference temperature changes equations 2,3 and 4 into 6, 7 and 8 respectively:

$$E_f = \sum_{i=0}^{5} [H_{vi} + C_{pvi} * (T_f - T_i)] * F_i \qquad (6)$$

$$E_o = H_{vo} * F_o \qquad (7)$$

$$\sum_{i=1}^{5} E_i = 0 \text{ (by definition)} \tag{8}$$

And the basic heat balance of equation 1 can be rewritten as:

$$C_{pvo} * (T_f - T_o) * F_o + \sum_{i=1}^{5} [H_{vi} + C_{pvi} * (T_f - T_i)] * F_i = \Sigma Q \tag{9}$$

A question presents itself here: If the flash zone vapor F_i's are assumed to be exactly identical to the product flows, then this also assumes instantaneous mass balance, and obvious error. The tower hold up is large enough to permit some mass imbalance during crude switches. The answer is: the application does not use any measurement of F_i. Measuring F_i in order to consequently feed forward to F_i would be redundant. Instead, the F_i's are calculated from heat balance in a way that will be shown below. The liquid flows F_i's (i = 1,2,3,4,5) should therefore be considered as the potential products and overflash that are in a vapor form in the flash zone and are condensed in the tower. The only product measurement used in the model is F_o, a measurement which is not redundant because there is no feed forward to F_o; the overhead flow is determined by the top temperature controller setpoint.

So far no assumptions were made and the heat balance is general. However, at this point we need to introduce two simplifying assumptions to permit the use equation (9) for calculating the crude volatility. It is assumed first that the combined liquid product enthalpy can be calculated as if the liquid products all leave the tower at an average tower temperature, and second that the crude *True Boiling Point* slope is a straight line. The error introduced by the average temperature assumption is within 3% of the cooling load, and it has almost no impact on the calculation of liquid product flow (97% accuracy is more than adequate for feed forward purpose). The error introduced by the straight line TBP assumption will be discussed later.

Incorporating the average temperature assumption into equation (9) enables calculation of the instantaneous liquid product flow, in equation (10) F_{liq} is the flow of all sidestreams plus overflash.

$$F_{liq} = [\Sigma Q - C_{pvo} * (T_f - T_o) * F_o] / [H_v + C_{pv} * (T_f - T_o) / 2] \tag{10}$$

After calculating the liquid products, the application proceeds to estimate the crude volatility, which we loosely define as the reciprocal of the crude TBP slope between 140 and 400°C. The usefulness of that definition is demonstrated in Figure 2. The figure shows that during crude switches the sidestream changes should be proportional to volatility changes, and hence once the volatility is known, accurate feed forwarding is possible.

At this stage of the calculation two points on the TBP curve are known. One is the overhead flow F_o and its end point T_o, and another is the flash zone flow (F_{liq} + F_o) and the flash zone temperature T_f. Although T_f and T_o do not exactly belong on the TBP curve they can be corrected via a standard TBP/EFV conversion method. Having made the assumption that crude TBP curves are linear within the sidestream boiling range, equation 11 is the resultant crude volatility.

$$V = (100 * F_{liq} / F_c) / (T_{fc} - T_{oc}) \tag{11}$$

Where:
- T_{fc} is the boiling temperature corresponding to flash zone % evaporation, corrected to represent a TBP point.
- T_{oc} is the boiling temperature corresponding to overhead % evaporation, corrected to represent a TBP point.
- F_c is the crude flow.
- 100 * F_{liq} / F_c is the combined % yield of overflash and all sidestreams.

Changes of this volatility are fed forward to all the sidestreams. But before proceeding with the feedforward part of the application it is of interest to pause here and confirm that after all assumptions are introduced, the calculation of crude volatility does not violate the simple logic of the previous section. The previous section had argued that the feedforward should be about proportional to the cooling load in the three cooling circuits of the fractionator, and only small deviations from that proportion are needed to account for unrelated variations of operating parameters. A close look at equations 10 and 11 reveals the validity of that statement. In the numerator of equation 10 there are two terms. The first is the combined fractionator cooling load ΣQ and the second is the sensible heat of cooling the overhead vapor. Numerically, the second term is about ten percent of the first one and F_{liq} is nearly proportional to ΣQ. The proportionality is even further enhanced because the overhead flow itself normally also goes up or down with ΣQ. The denominator of equation 10 is usually stable; it might vary five percent or so crude-to-crude. The same is true for the denominator of equation 11. Thus it is reconfirmed that the volatility calculation predominantly depends on the fractionator cooling load.

The discussion will now go into feedforwarding the volatility changes to the sidestream draw off rates. In the initial stages of testing the scheme feedforward was done exactly as shown in figure 2 for all four sidestreams, i.e.

$$\text{Delta } Yi = Yi * \text{Delta } V / V \tag{12}$$

Where Yi is the yield set point of sidestream i

It was then found that the feedforward to the second, third, and fourth sidestreams was nearly 100% accurate in magnitude whereas the feedforward to the first sidestream was always underestimated. This has to do with the nonlinearity of the TBP slope in heavy crudes and the fact that they contain less first sidestream than what is predicted by a straight line TBP curve. The problem was handled by pretending that the first sidestream flow is higher than the actual measurement:

$$\text{Delta } Y_1 = (Y_1 + YE_1) * \text{Delta } V/V \qquad (13)$$

Where YE_1 is the extra first sidestream flow, determined by tuning

To summarize, the volatility calculation has been simplified to rely on two basic equations: (10) and (11), plus some corrections for temperatures. Such a simplification helps reduce the number of critical measurements and increase reliability.

V. Interface with Other Pipestill Control Applications

This section provides a short summary of the Pipestill control schemes that interface with the crude switch application. The interfacing schemes can be divided into two groups: sidestream quality control and tower internal reflux control. The crude switch application reads internal reflux information, performs heat balance and volatility calculation, and feedforwards to the sidestream control schemes. A graphical illustration is shown in Figure 1. A summary of the interfacing control schemes follows.

1. Sidestream Quality Control

This is a package of three applications, one for each of the first three sidestreams. Their objective is to maintain sidestream distillation and/or cloud point at operational targets. The important features of the quality control schemes are:

- Reliance on analyzer measurements for feedback control of sidestream quality
  - ⇒ The first sidestream strategy aims at maintaining the distillation end point at a specified target.
  - ⇒ The second sidestream strategy aims at maintaining the distillation 95% point at a specified target.
  - ⇒ The third sidestream has two analyzers: a cloud point and a 90% distillation point. Analyzer feedback can use either or both analyzers.
  - ⇒ (The fourth sidestream is drawn at a constant yield).

- The sidestream strippers plus analyzer sampling delays cause considerable dead times in the quality loops, and to compensate for these dead times the sidestream applications all use dynamic quality predictors.
- The program incorporates a sidestream to sidestream decoupling logic; when sidestream i-1 changes end point, sidestream i’s draw off rate would be adjusted to keep its endpoint constant.
- Crude volatility change are fedforward to all sidestreams.

2. Internal Reflux Control

- Tower top temperature control.
  Minimizes the tower top temperature subject to a dew point constraint.
- Tower pumparound control.
  Maximizes the top pumparound subject to top section internal reflux constraint.
- Overflash control.
  Minimizes furnace load subject to wash section temperature difference constraint.

VI. Application Performance

This section compares the sidestream quality control during a crude switch with and without crude volatility feedforward. Figures 4 through 9 are reconstruction of trend plots of interest taken during crude switches. Each figure has two plots: a solid line which refers to a crude switch when the application was working in closed loop control, and a broken line which refers to data when the application was not working. At the time of the open loop test the operator had already been trained in usage of the application, and it appears that he relied on the volatility calculation for timing of sidestream adjustments; but even so, it is felt that the comparison demonstrates the precision of the application well. Explanation of specific figures follows:

Figure 4 is a plot of the crude mix temperature. The heavy crude tanks are heated to prevent pumping problems, and thus the change of temperature indicates the time at which new crude started coming into the unit, or it may also indicate changes of crude mix ratios. The figure shows that both switches started around 10:00 am, that it took about one hour for the crude mix to stabilize again, and that both switches went from light to heavy crudes. This figure is given for reference only. The crude temperature is not taken into account in the application.

Figure 5 is a plot of crude volatility as calculated by the simplified heat balance procedure, equations 10 and 11. It shows several interesting points. First, when the crude mix is stable, the calculation is also stable. This is true for both switches before 10:00am and after 12:00pm. Such stability is a must for eliminating extraneous feedforward. Second, the volatility trend is more complex than the crude temperature trend of Figure 4. During the switches there were periods where the crude was lighter than either steady states, then heavier than either steady states and only then the volatility settled to a new value. As explained in section III such variations are the result of switching in one new tank and then another, and of unused headers being filled with light crude. A comparison of the dotted line against the solid one shows that the open loop switch was somewhat more severe in that the change of volatility was larger. On the other hand, the solid line of the closed loop switch shows some irregular multiple patterns of header switching, which make it a more difficult control case. The solid temperature line of Figure 4 also hints of a certain intermediate step between 10:15 and 10:30. Taking that complexity into account it is felt that the two crude switches are of about the same difficulty and a comparison between them is fair. The solid line in

Figure 5, when the application was working, with all its ups and downs, was directly fed forward to the sidestreams as specified in equations 12 and 13.

Figure 6 is a plot of the pipestill bottom yield. It is shown here because the bottom yield variations summarize all control actions, feedforward or otherwise, taken on the sidestreams. The solid line plot of Figure 6 is nearly synchronized with that of Figure 5. This is of course due to the combined effect of volatility feedforward to the sidestreams. When the sidestreams are decreased the bottom yield increases. When the application was not in use the operator had also tried to time his sidestream changes together with volatility changes but he made two minor errors. First, his feedforward was too fast. This is hard to see on Figure 6 but it will be clearly seen in the sidestream quality trends. Second, the mismatch of the dotted lines of Figures 5 and 6 between 11:00 am and 12:00 noon indicates that he failed to see the period of heavier than normal feed. The effects of these relatively small mistakes will be observed in Figures 7,8, and 9.

Figures 7,8 and 9 are trend plots of the sidestream quality analyzers: heavy naphtha end point, kerosene 95% distillation point and light gasoil cloud point respectively. The naphtha end point and kerosene 95% point targets remained constant during both switches, but the gasoil cloud target was changed in both crude switches from -11°C before to -4°C after the switch.

As discussed above, in the open loop crude switch the operator committed two minor errors: first, the sidestream yields were reduced a little too early, and second, he failed to notice a period of heavier than expected crude. The first mistake had driven all of the dotted lines in Figures 7,8 and 9 down at 10:40 or so, showing lighter than necessary sidestreams. Then, at about 11:10 the second mistake started driving all sidestreams to be heavier and off specification. In the case of Figure 9, the operator changed cloud set point around 10:30, but the loop has an hour of dead time before any results can be observed, and the sharp rise of cloud is therefore not the result of target change. So much for dynamic precision of operator action; on the other hand, the steady state accuracy of the open loop test was unusually high, save a 2°C (about 0.5% on crude) mistake on the first sidestream, which in turn caused also a kerosene deviation. The operator used planner's prediction of expected yield based on averaging the crude assays of the mixture. This prediction was good on the day of the test, but it may on occasions be less than adequate because as explained in section III, good mixing of crudes inside the tanks cannot be relied on.

To summarize the open loop crude switch, the planning information and operator actions were as precise as can be expected, and the dotted lines of Figures 7,8 and 9 present a smooth crude switch, better than can usually be achieved in open loop.

The solid lines of Figures 7,8 and 9 demonstrate that automatic volatility feedforward can achieve a much better crude switch control than even the best open loop yield setting. The trends speak for themselves. During that particular crude switch, distillation points of naphtha and kerosene were within 2°C of targets, and cloud point of gasoil climbed up smoothly to its new target without overshooting. In fact, these variations of two degrees around target are only slightly larger than normal deviations of sidestream qualities on days of steady crude. It is also worth noting that other tower parameters, not plotted here had also maintained excellent stability. Tower temperatures hardly moved and downstream disturbances were minimized.

VI. Nomenclature

C_{pli}	Liquid specific heat of product i
C_{pv}	Average vapor specific heat of the liquid products
C_{pvi}	Specific heat of vapor product i at constant pressure
E_f	Enthalpy of flash zone vapors
E_i	Enthalpy of product i as it leaves the tower
F_c	Crude oil flow
F_i	Flow of product i
F_{liq}	Combined flow of the first, second, third and fourth sidestreams plus the overflash flow
H_v	Heat of evaporation of an average liquid product (sidestreams and overflash) at an average product draw off temperature
H_{vi}	Heat of evaporation of product i at a reference temperature T_{Ri}
i	Index denoting pipestill products as follows: + Overhead (i=0) + Sidestream 1 (i=1) + Sidestream 2 (i=2) + Sidestream 3 (i=3) + Sidestream 4 (i=4) + Overflash (i=5)
k	Index denoting pipestill cooling circuits as follows: + Reflux condenser (k=0) + Top Pumparound (k=1) + Mid Pumparound (k=2)
Qk	Heat duty of cooling circuit k
ΣQ	Combined heat duty of all three cooling circuits
T	Temperature
T_f	Flash zone temperature
T_{fc}	Flash zone temperature corrected from EFV to TBP
T_i	Temperature of product i as it leaves the pipestill tower
T_R	A reference enthalpy temperature at which the liquid enthalpy is taken at zero
T_{Ri}	A reference enthalpy temperature for product i
T_{oc}	True boiling endpoint of the overhead product
V	Crude volatility (reciprocal of the slope of the crude TBP curve)
Y_i	Yield of product i (%)

YE_1 Additional imaginary first sidestream yield needed to linearize the crude TBP curve. If the TBP curve of the crude were linear, the first sidestream yield would have been $Y_1 + YE_1$.

Delta A mathematical operator denoting difference

Σ A mathematical operator denoting summation

Figure 1 CRUDE SWITCH FEED FORWARD IN GSK KAWASAKI - OVERVIEW

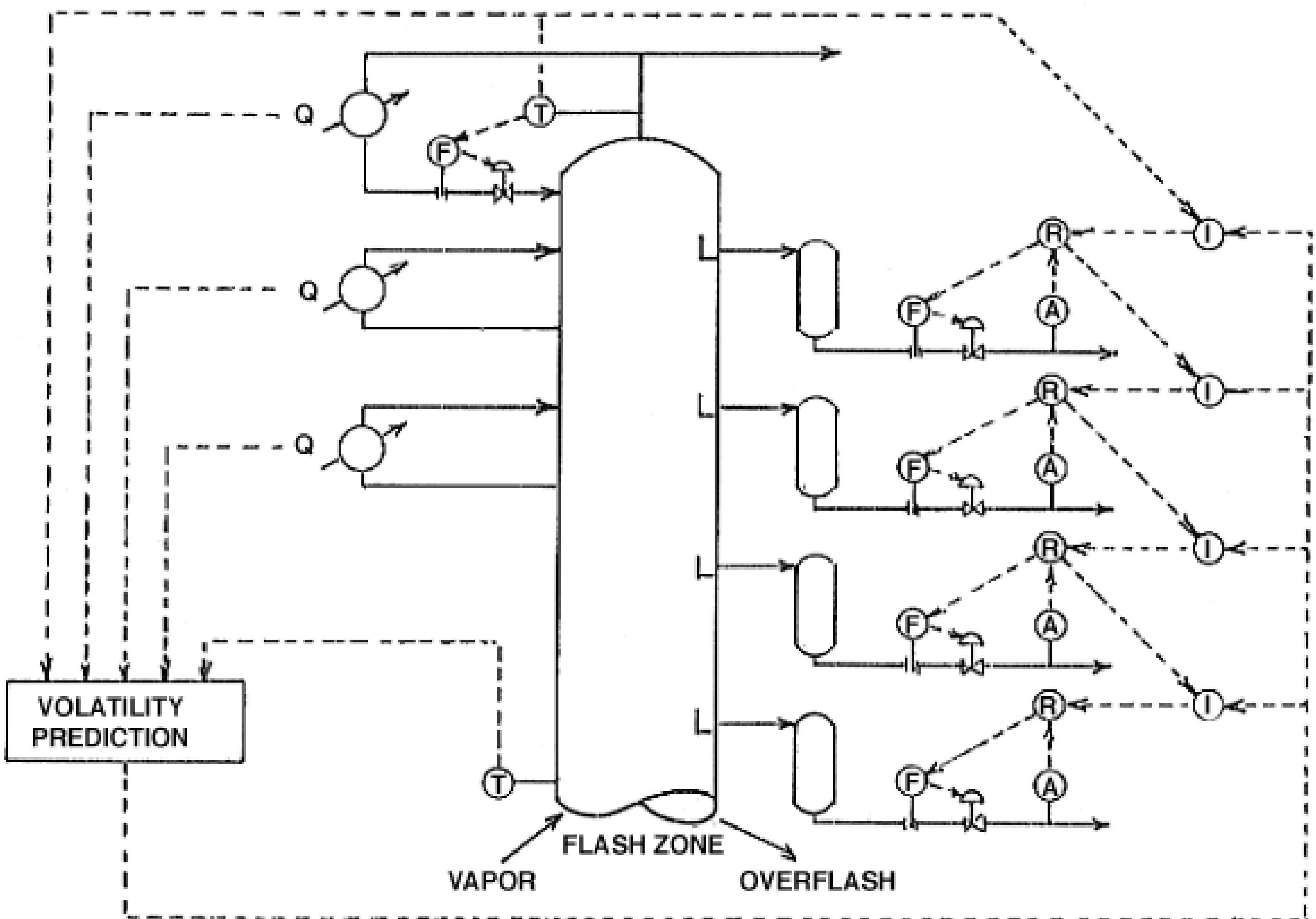


Figure 2

TYPICAL YIELD CHANGES DURING A CRUDE SWITCH

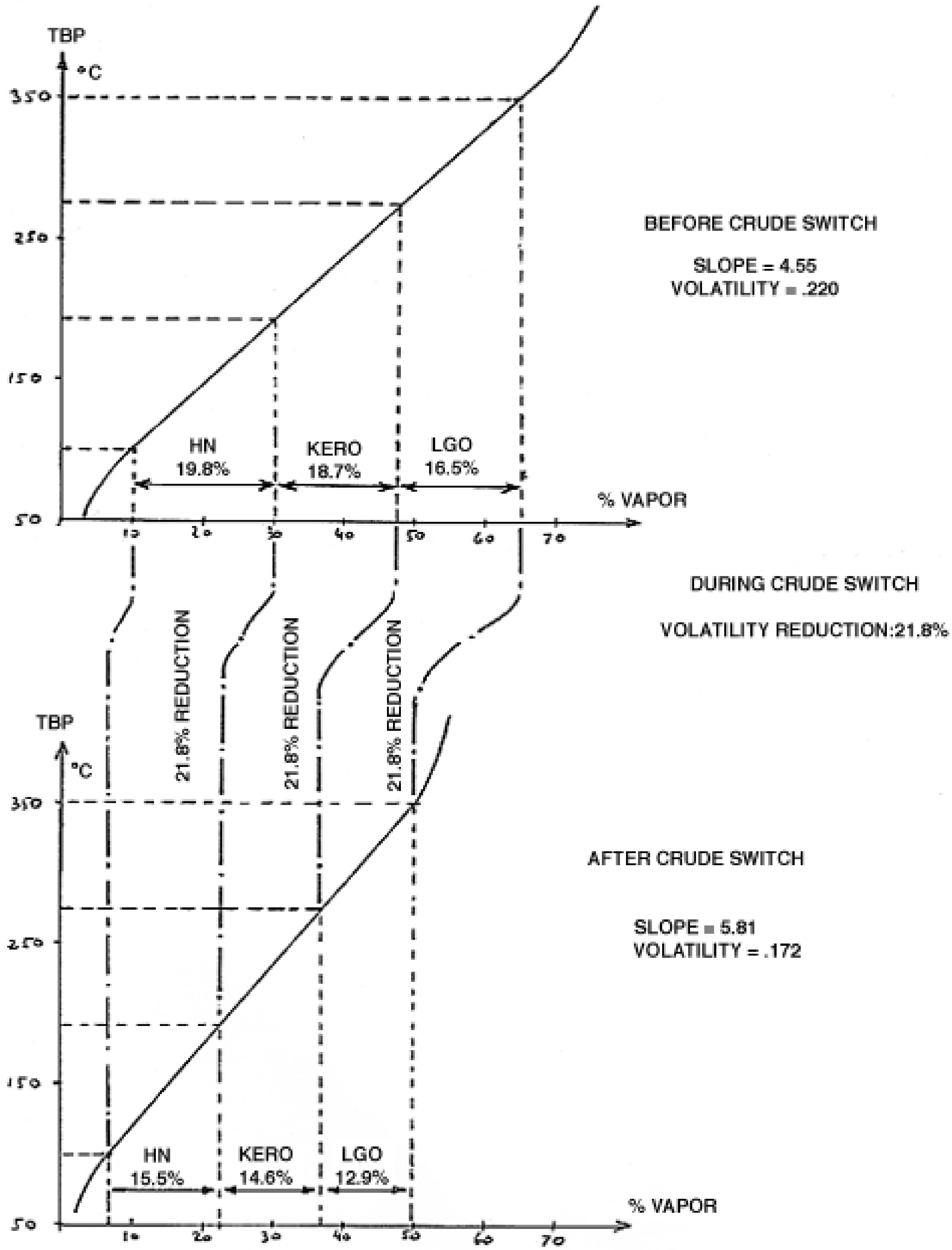


Figure 3. HEAT BALANCE ENVELOPE

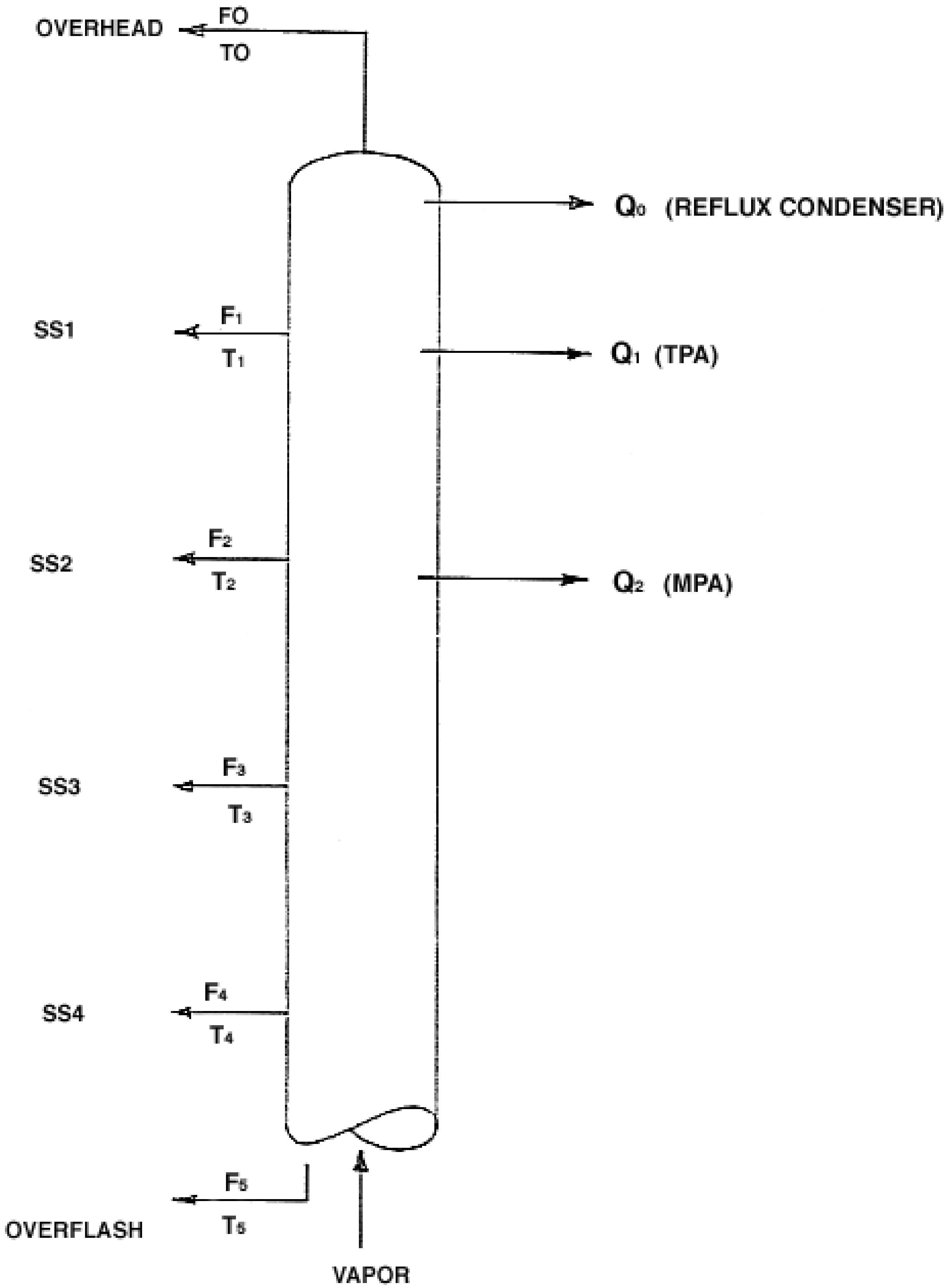


Figure 4. CRUDE MIX TEMPERATURE (°C)

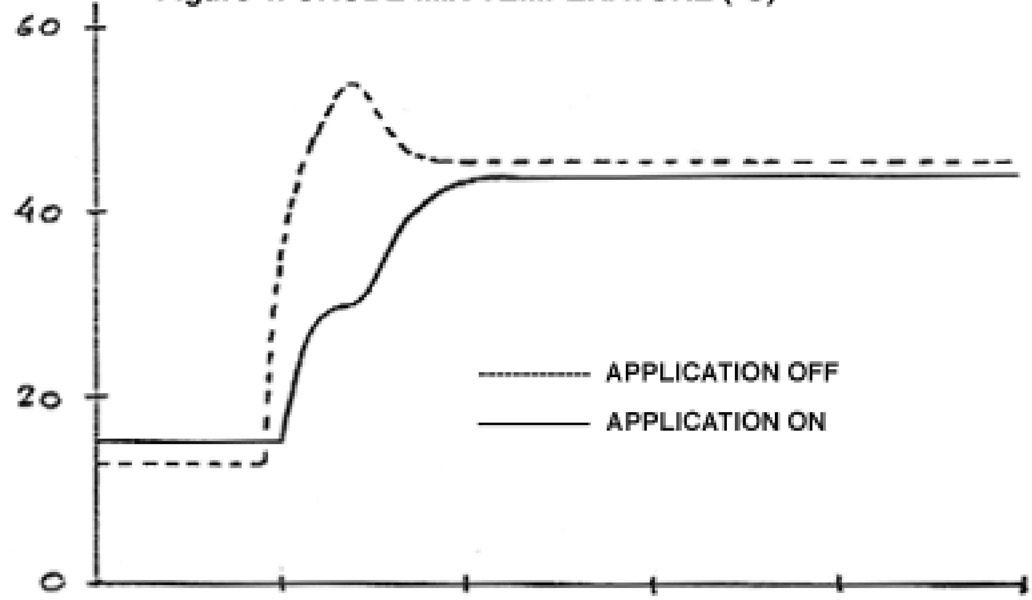


Figure 5. CRUDE VOLATILITY (%YIELD / °C)

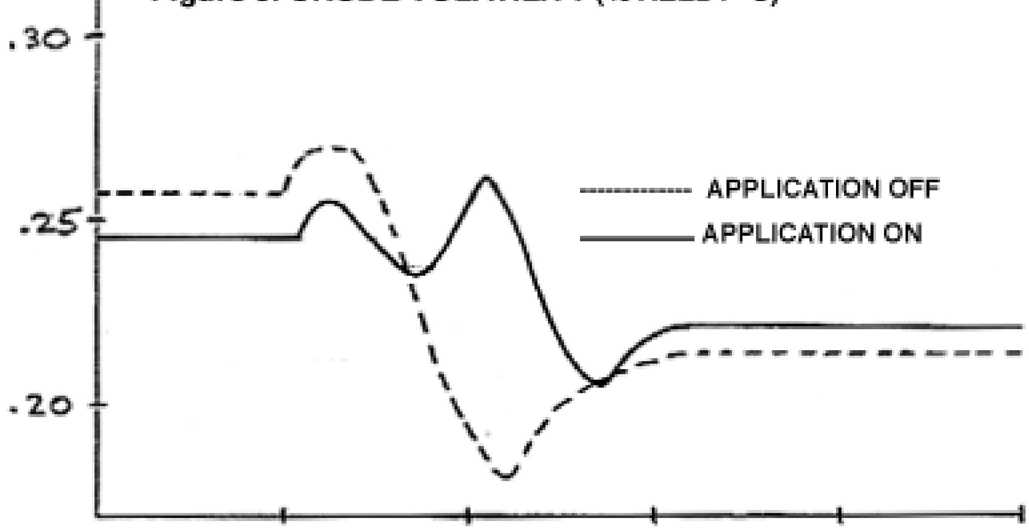
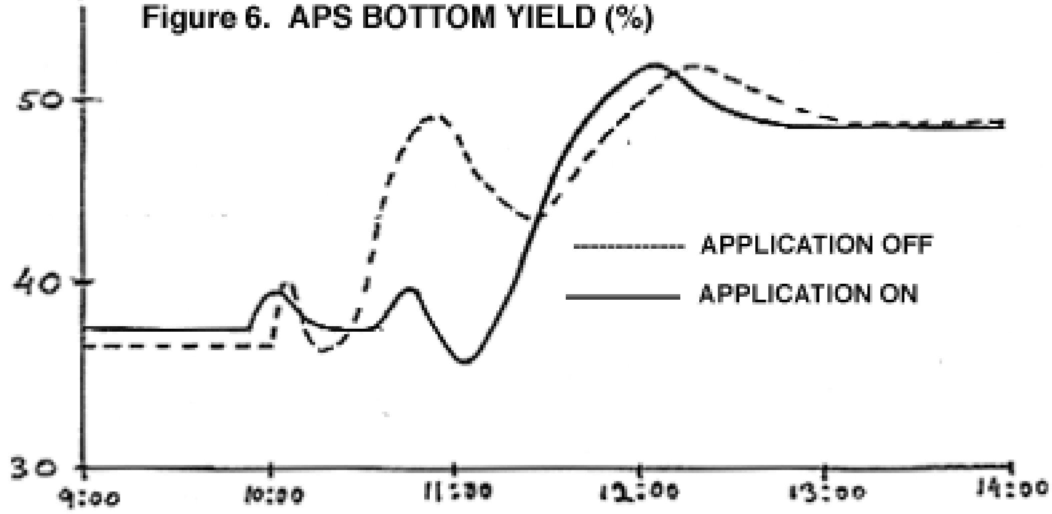


Figure 6. APS BOTTOM YIELD (%)



HOUR OF THE DAY

Figure 7. NAPHTHA END POINT (°C)

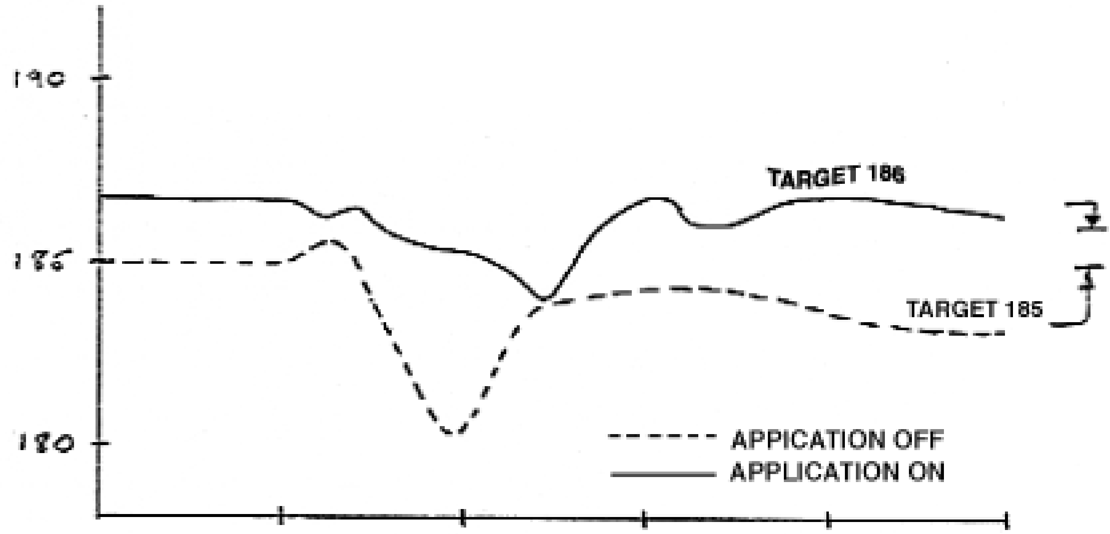
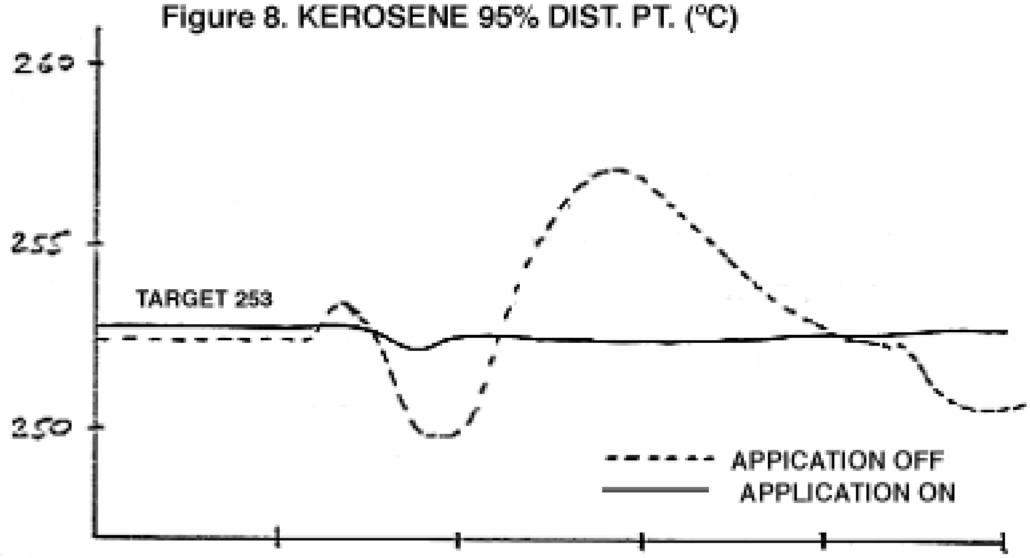


Figure 8. KEROSENE 95% DIST. PT. (°C)



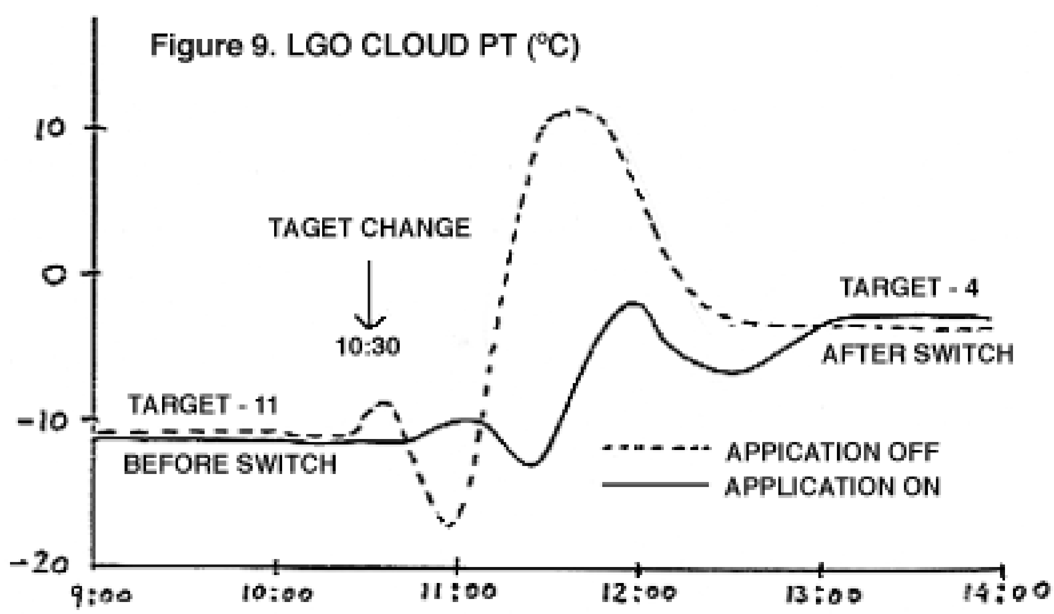


Figure 9. LGO CLOUD PT (°C)