

**THE USE OF FIRST-PRINCIPLES INFERENCE  
MODELS  
FOR CRUDE SWITCHING CONTROL**

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# **THE USE OF FIRST-PRINCIPLES INFERENCE MODELS FOR CRUDE SWITCHING CONTROL**

## **OVERVIEW**

Crude units (CDU) advance process control (APC) is one of the most lucrative refinery control application, and it would have been even more profitable if not hindered by the performance of such controllers during crude switch disturbances. If a CDU APC application cannot work during crude switches – economic penalties ensue:

- Throughput reduction
- Product downgrading
- Occasional pump cavitation and other equipment problems

Why then are many CDU control applications turned off during crude switches? To control the unit during crude switches the APC must have a reliable inferential package that would be robust enough to work through the switch. Most inferential packages rely on regression analysis of steady state data. Such statistical packages have reliability problems even at steady state, and dynamically they are simply not designed to cope with the crude switch disturbance. Appendix A Provides more information about the draw backs of regression analyses.

This paper describes how Petronas has successfully integrated crude switch capability into their CDU APC application by using first principles inferential modeling.

### **Slide 2. Starting point for the project**

About five years ago Petronas had implemented APC on its PPMSB PSR2 crude unit. The application included an RMPCT (Honeywell's multivariable) controller plus inferential models based on Honeywell's fractionator toolkit and other regression models. The application worked, but its MV's were heavily clamped, and during crude switches it was turned off. In addition to the inferential modeling problems, the unit has gone through a revamp, and the control matrix was not taking into account all unit constraints.

Following a reference site visit (to NPRC Marifu, Japan), Petronas has commissioned Petrocontrol to modify the existing application as follows.

- Implement automatic crude switching using the GCC inferential package
- Redesign RMPCT to handle all current unit constraints

### **Slide 3. Project highlights**

Petronas desired a turnkey project by Petrocontrol, although given the experience of Petronas engineers we had a joint team of two Petrocontrol engineers and two local engineers. Not only that was a very economical way to accomplish the project, but also it guarantees good local knowledge of the application and effective maintenance.

#### **Slide 4. Project schedule**

The project went smoothly through quite an ambitious schedule of less than six months, starting with an initial visit in Early August, and commissioning in Early January. After the January commissioning the refinery started operating on a certain crude-oil with high LPG and gas content, causing many operational problems and forcing throughput sacrifices as shown in slide 5. We came back in February to configure the compressor loading as a throughput constraint and that has stabilized the throughput.

#### **Slide 6. PSR2 crude unit**

PSR2 is a crude unit with naphtha overhead and three sidestreams. This slide shows the main manipulated variables, and for simplicity it does not show the constraints. We are manipulating top temperature, product draws, pumparounds, furnace COT (coil outlet temperature), and last but not least – throughput. Given that the current throughput is much higher than design, the unit has a number of constraints in the furnace, fractionator and compressor area.

There is also an unusual desalter constraint worth mentioning. Both the TPA and MPA exchange heat against crude before the desalter. That renders the desalter too hot, to the point that desalter temperature must be controlled very precisely or else the crude may partially evaporate and cavitate the furnace charge pump.

#### **Slides 7, 8. GCC summary**

We will now take a few slides to explain how GCC works.

GCC is a fractionator first principles inferential package designed to handle crude switches. The model involves:

1. Identification of the crude TBP (true boiling point) curve from unit measurements
2. Use of TBP information to model the product cutpoints. Hence the name of the model: GCC (Generalized cutpoint calculator).
3. Product properties prediction
  - ⇒ ASTM distillation is calculated as a function of cutpoints, plus internal reflux to present heavy tail effect.
  - ⇒ Flashpoint is calculated as a function of cutpoints and stripping steam
  - ⇒ Freezepoint is calculated as a function of cutpoints and Watson Kw. Watson Kw is an indication of the aromatic content, which affects cold properties. GCC requires a density reading of one of the sidestreams to calculate the crude Kw factor.

#### **Slides 9, 10, 11. GCC features**

- **Heat balance**

The reliance on heat balance is one of the main reasons for the robustness of the model. During major disturbance the unit can operate off mass balance, but it is always in heat balance, permitting the model to work during crude switches. Further, fractionator cooling load is sensitive to crude TBP and the use of heat balance permits a quick detection of TBP curve changes.

Slide 10 shows another interesting inferential control phenomenon. Models that read product flows directly see their own actions as disturbances. That is a completely unintended feedback loop that tends to drive controllers unstable. People deal with the instability by lagging the model inputs, but that interferes with the ability of such models to handle disturbances.

- **Internal reflux model**

Pumparound load distribution has a profound effect on unit economics. The wrong distribution would cause either poor separation or fuel efficiency penalties, and in the extreme it may cause tray dry-out and pump cavitation.

GCC estimates the internal reflux again by heat balance. A top-down heat balance model would have good accuracy in the TPA and MPA region, whereas at the bottom of the column it loses accuracy and the overflash model takes over.

- **Overflash model**

GCC estimates overflash from temperature profile and heat balance around the fractionator lower section. That is again an inference that is absolutely necessary for smooth control during crude switches. In the absence of such model the lowest sidestream may occasionally be contaminated by bottoms carry-over, a danger which forces the operator to shy away from APC and operate very conservatively during crude switches

About 40 GCC applications have been implemented to-date, and almost invariably they have demonstrated crude switch capabilities [1, 2, 3, 4], inferring product qualities and working with an MVC (multi-variable controller) to manipulate the unit smoothly to a new steady state. Early implementations applied MVC as a part of GCC, whereas more recent applications interface to other commercial MVC products.

### **Slide 12. GCC handling of crude switches**

Once the crude TBP is known, slide 12 illustrates how easy it is to control the products. The yield pattern required keep the product cutpoints at target is known, and control execution is trivial. (Of course what make it not trivial are the other constraints that must be respected).

## **Slides 13, 14, 15, 16, 17. PSR2 application structure**

Our application covers all of the unit constraints and manipulated variables, including furnace balancing and heat exchanger train balancing, in total about 40 CV's and 30 MV's. For simplicity we will address here only the key inferences and constraints.

Slide 13 shows the key GCC inferences

- Naphtha 90% point
- Kero flash, 90% and freeze point
- LGO and HGO 90% point
- Overflash model, replacing an unreliable overflash measurement
- Tray loading in the TPA section

Slide 14 shows the main measured constraints

- Desalter temperature
- HGO tray dry-out: a dry-out indicator is calculated as a function of stripper level, level rate of change and valve position
- Furnace constraints
- Compressor and flaring constraints
- Pumparound load ratios
- Hydraulic constraints

Slides 15 and 16 show some RMPCT dynamic models. It is of interest to note how quick the response is, without any dead time. That is typical of GCC inferences and such a response permits tuning for the application that is much more aggressive than common practice. Initially we were worried about how quick one can tune a multi-variable controller, and we prepared a fall-back option of special disturbance variables that would become active during crude switches and speed up the response. We were pleased that there was no necessity to activate these disturbance variables.

## **Slide 18, 19. Petronas crude mix**

Petronas goes through frequent and major crude switches. Slide 18 gives a table of crude-oils operated during the month of January. Announced switches occur every three to seven days, and there are also unannounced switches, which we would discuss later.

Slide 19 shows some GCC trends: crude TBP slope, °C / % volume in the sidestream region. Clearly there are some significant crude switches and some gradual changes. Those gradual changes are related to tank layering. The second, blue curve on slide 19 is the overflash model, showing that the overflash generally stayed in a safe region. Due to HGO tray damage, overflash below 25 M<sup>3</sup>/Hr can cause a tray dry-out as we will see in an example later.

The high level of noise during the first half of the period stems from the flaring problem and resulting unstable crude throughput. Once that problem was solved the level of noise was substantially reduced.

### **Slides 20 through 26. Inference versus lab trends**

There are two yardsticks to measure the performance of inferential model. First they must trend well against the lab. Daily changing of biases is not satisfactory because it amounts to controlling the inference model instead of the process. GCC has been calibrated against 2003 production data, but upon coming to commission the application we found some errors in the old input data and decided to recalibrate. Slides 21 through 26 show the fit of this new model.

It can be seen that with the exception of HGO, inferences track the lab well. Considering HGO, one must remember that lab repeatability deteriorates as the product becomes heavier, due to sample cracking, and it is plausible that the real fit is better than what slide 24 shows.

We are especially pleased with the performance of kero flashpoint and freeze point. Note that Petronas does not have a density analyzer and the Watson Kw is calculated from operator input of ASTM 50% and density. The correlation then is only guaranteed for as long as the same crude is run.

### **Slides 27 through 44. Dynamic performance of the application**

We come now to the second yardstick for inferential control performance. How responsive the inference is to operational changes, and whether the predicted off specification product can be corrected in time.

Slide 29 shows a PI trend of overflash flow, overflash low and high targets and HGO stripper level. Due to insufficient training the operator specified infeasible APC operation, causing RMPCT to ignore the overflash limit and operated below it. That in turn caused a tray dry-out and a drop in stripper level. Tray dry-out has a high control priority and RMPCT quickly recovered. While it is clear that we have to spend some more time training operators, this event also illustrates the value and accuracy of the overflash model.

Note that the tray dry-out control variable is quicker in its response than the level, as it takes into account level rate of change and level valve position.

Slides 30 through 38 show trends of TBP curve slope, overflash flow, product flows and product ASTM 90% inference during three crude switches. The switches often come in several steps, the exact times of which are unknown to the operators. We can see that the product draws, and in particular LGO, are moved aggressively to control the product properties, and that the typical deviation during switches is around 5°C, a very good performance, considering that there are other constraints on the

furnace, overflash, desalter temperature and valve positions, which must also be respected.

Slides 39, 40, 41 show the same trends, except that crude switch was completely unannounced. Officially the crude tanks did not change but the tank farm operators would occasionally introduce slops, adjust crude ratio of heavy to light crude, or switch tanks of supposedly the same crude. These trends demonstrate that the application does not rely on operator help for its good performance. The operator had no idea that a substantial crude switch was taking place at the time.

Slides 42, 43, 44 show again the same trends, except instead of an abrupt crude switch these there was a gradual change in crude quality. This change has to do with tank layering. It is common that the bottom of a crude tank is fairly heavy, and the crude gets lighter as it is being pumped into the unit. Again our controller gradually adjusted unit conditions while the operator had no idea that the crude quality was continuously changing.

#### **Slide 45. Benefits**

Since commissioning the application stayed on, registering a service factor of 99% since February 1<sup>st</sup> and was used through several dozens announced and unannounced crude switches. That is no small feat because major crude switching at Melaka involves going from about 40% white products to 60% or vice versa. The unit throughput is higher than nominal design capacity and the unit is always constrained. Switching crude while keeping the unit within constraints, staying at a high throughput and at the same time keeping the products on spec without giveaways are the most important control objective for the unit. We have clearly quantified the following benefits

- Crude switch time shortened from several hours down to one hour
- The throughput is kept high against constraints
- Product downgrading is almost completely eliminated
- Overflash control prevents tray dry-outs
- Internal reflux profile probably improved energy utilization. This is difficult to measure.
- Reduced number of alarms (and incidents)
- Reduced lab support requirements
- Useful engineering information: crude TBP and other indicators

## **LITERATURE CITED**

1. Friedman, Y. Z., "Control of Crude Fractionator Product Qualities during Feedstock Changes by Use of a Simplified Heat Balance". American Control Conference, 1985.
2. Friedman, Y. Z., "Crude Unit Advanced Control Experience". Hydrocarbon processing Journal, February 1994.
3. Mark Schuler et al, "Refinery uses column data to infer and control product properties". NPRA Computer Conference, November 2000, later published in O&G Journal, February 19, 2001.
4. Pradeep Singh et al, "Multivariable Controller Implementation for a Crude Unit: A case Study". NPRA Computer Conference, October 2002, later published in O&G Journal, November 4 2002.



## **APPENDIX A**

### **A discussion of reliability of regression versus first principles models**

This section explains the difficulties of applying empirical regression to infer distillation column product purities. There is no need to discriminate between neural net versus simple regression models, because both rely on Gaussian statistics, and the empirical modeling issues are all related to limitations of Gaussian theory. The authors are not generally against statistical models, especially for equipment where first principle knowledge does not exist, but in distillation there are several strikes against empirical models.

#### **A) Regression requires independent inputs**

As shown in figure 2, inference models input flow, temperature and pressure measurements and output estimated product properties. Gaussian theory requires that all inputs be independent, however that is not feasible. Temperatures, pressures and flows are related in several ways: mass balances, heat balances and equilibrium equations. Ignoring these relations makes the modeling process theoretically incorrect. A more correct statistical approach would involve Bayesian theory, which takes into account a-priori knowledge of dependence among input variables. Of course that a-priori knowledge brings us back to the first principles of distillation.

What then happens to models that use dependent variables? They have incorrect coefficients and hence would drift upon changes in process conditions.

#### **B) Empirical models require large volumes of lab data**

Statistical regression requires hundreds of laboratory data and that poses a problem. A fair percentage of daily lab data is biased and reliable process data is obtainable only by test-runs. During a test-run the unit is kept at steady state for several hours, then samples are drawn by careful procedure, in properly sealed sample bottles, taken to the lab without delay and tested immediately in presence of the process engineer.

There is no hope that the quantity of lab data needed for regression would come from high quality test-runs, and empirical model developers would have to rely on imprecise every-day lab data. Imprecise because it permits occasional spike contamination, sampling during process changes, inappropriate sampling procedures and long delays between sampling and testing.

Can this lab data still provide the basis for a regression model? That might be possible, if the error is normally distributed without bias. Model developers often assume that those lab data that do not fit their correlation are wrong. They label such data "outliers", removing them from the original set, thus obtaining excellent fit, only to find out later that regression fitting does not guarantee future prediction. How many

outliers are acceptable? Eliminating just a few points from a large set of random data could entirely alter the regression formulae. Can we possibly delete over 1% of data from a set and still call this data reliable? What about 5%?

No, is the logical answer. Data should be deleted from a set only upon evidence that the point in question is erroneous. In that respect, first principle models provide a mechanism for identifying erroneous data. The development of first principle models is done without regard to the lab data and often before lab data is collected. Trending the un-calibrated model against lab data provides a powerful tool. While the un-calibrated inference model may not be in agreement against the lab, the two should surely trend together.

### **C) Empirical models must identify a large number of coefficients**

Scientific models incorporate model gains inherently, and the calibration procedure amount to adjusting one or two parameters. For example tray efficiency and weight in a weighted average formula. The effect of signal to noise ratio on the calibration procedure is minimal.

Empirical models on the other hand must identify at least one gain coefficient associated with every measurement. That is a problem because normal day-to-day operation may not provide enough movement in the data to give meaningful information. Large enough process moves endanger the product specification and cannot often be permitted. When data movement is too small the regression would simply model noise patterns.

### **D) First principle models provide the means for checking instrument errors.**

The sister problem of erroneous lab data is erroneous instrument data. Instrument errors occur due to poor calibration, partial plugging of orifice meters, improper installation, incorrect meter range and finally also computer interface errors. Inference models that were developed from erroneous instrument data would necessarily be weak. Eventually instrument problems are identified and corrected and what would happen to a regression model then? It would have to be re-developed of course, but from what data? The old data sets with erroneous readings cannot be used. Assuming lab tests are carried out once a day, creating a new set of hundreds of lab data would take many months, during which time the APC would be off.

We conclude that before starting inferential model development it would be prudent to survey all input measurements and identify all problems. The best way to accomplish that is via employing first-principles knowledge; testing the readings against mass balance, heat balance and equilibrium equations. If those readings cannot be reconciled against basic thermodynamic laws they cannot be correct. There is an obvious conflict here because people choose the empirical approach to avoid the

more time-consuming first principle models, only to find out that the use of first principles is unavoidable, if one desires to identify erroneous data sets.

**E) There is no replacement for process engineering**

And what if the measurements set is inadequate? A key measurement is missing, or is in the wrong location? To obtain a good model the set of measurements ought to “have the inferential information in them”. A first-principles modeler would identify an insufficient set of inputs at the outset by a simple sensitivity study. He would then halt the modeling effort until the missing measurement is installed. The empirical modeler would go through model development and the problem would only be found out at the time of model validation. At that time he would have to employ first principles to analyze the problem, wait for the measurement to be installed, then months until a quantity of lab data is available.

**F) Ability to survive process modifications**

During unit turnarounds distillation columns are often modified by the way of replacing trays, cleaning condensers or reboilers, etc. Any inferential model would need to be re-calibrated upon modifications of column equipment. First-principles models might require changes of equation coefficients, but empirical models would be turned off for a period of several months until a meaningful set of lab data is accumulated and the model re-developed from scratch.

# THE USE OF FIRST-PRINCIPLES INFERENCE MODELS FOR CRUDE SWITCHING CONTROL

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# Petronas CDU 2 starting point

- **RMPCT implemented 5 years ago**
  - **No crude switch capability**
  - **Some constraints not accounted for**
  - **MV's heavily clamped**
- **This project goals**
  - **Implement automatic crude switching using the GCC inferential package**
  - **Redesign RMPCT to handle all current unit constraints**

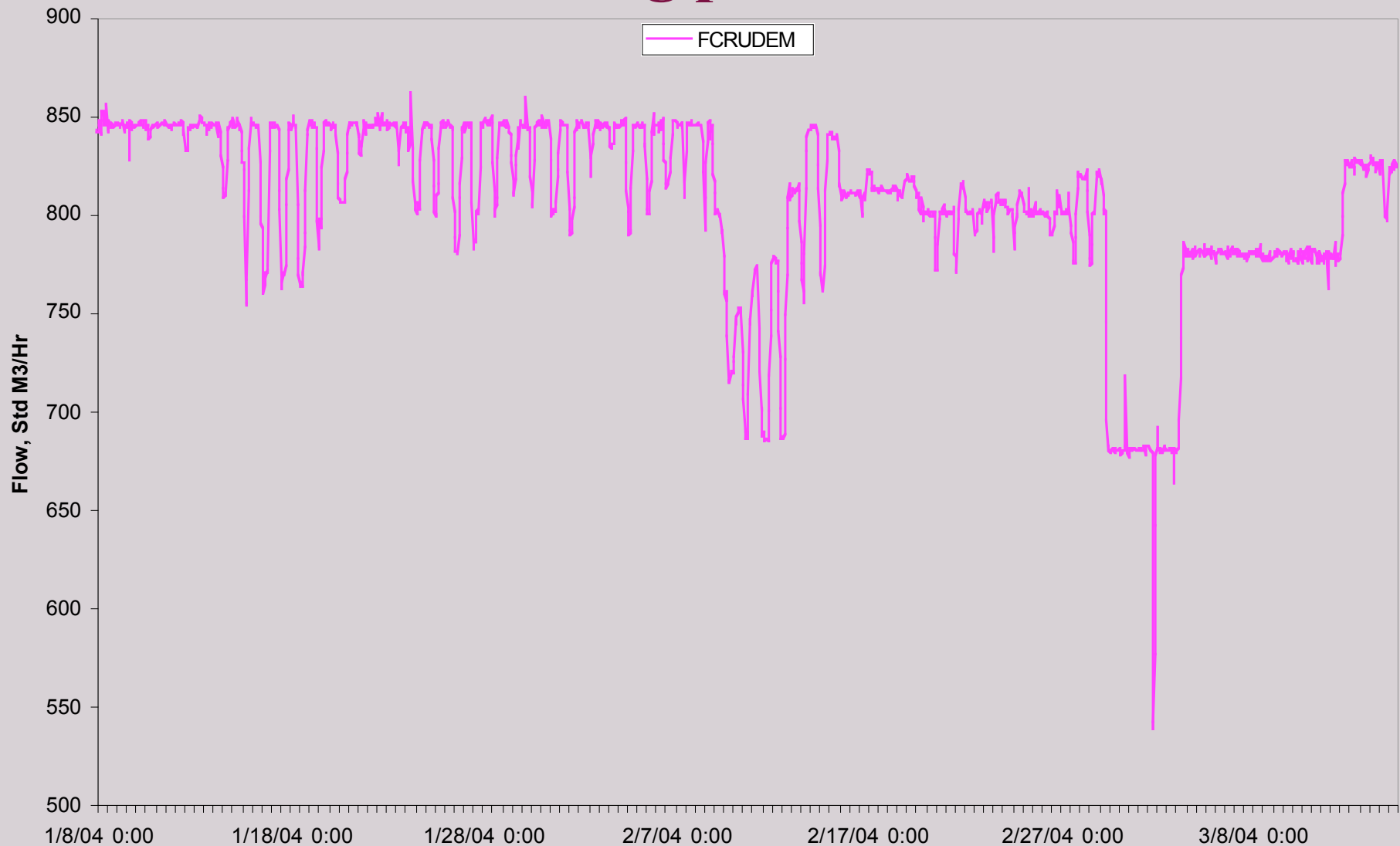
# Project highlights

- **Turnkey project by Petrocontrol**
  - **Functional design**
  - **GCC model**
  - **Step testing support**
  - **RMPCT commissioning**
  - **Overall project responsibility**
- **Heavy participation by Petronas**
  - **Step testing**
  - **Operator training**
  - **DCS work**
  - **Commissioning support**
  - **Performance testing**
- **Service factor Feb – April: 99%**

# Tough Schedule

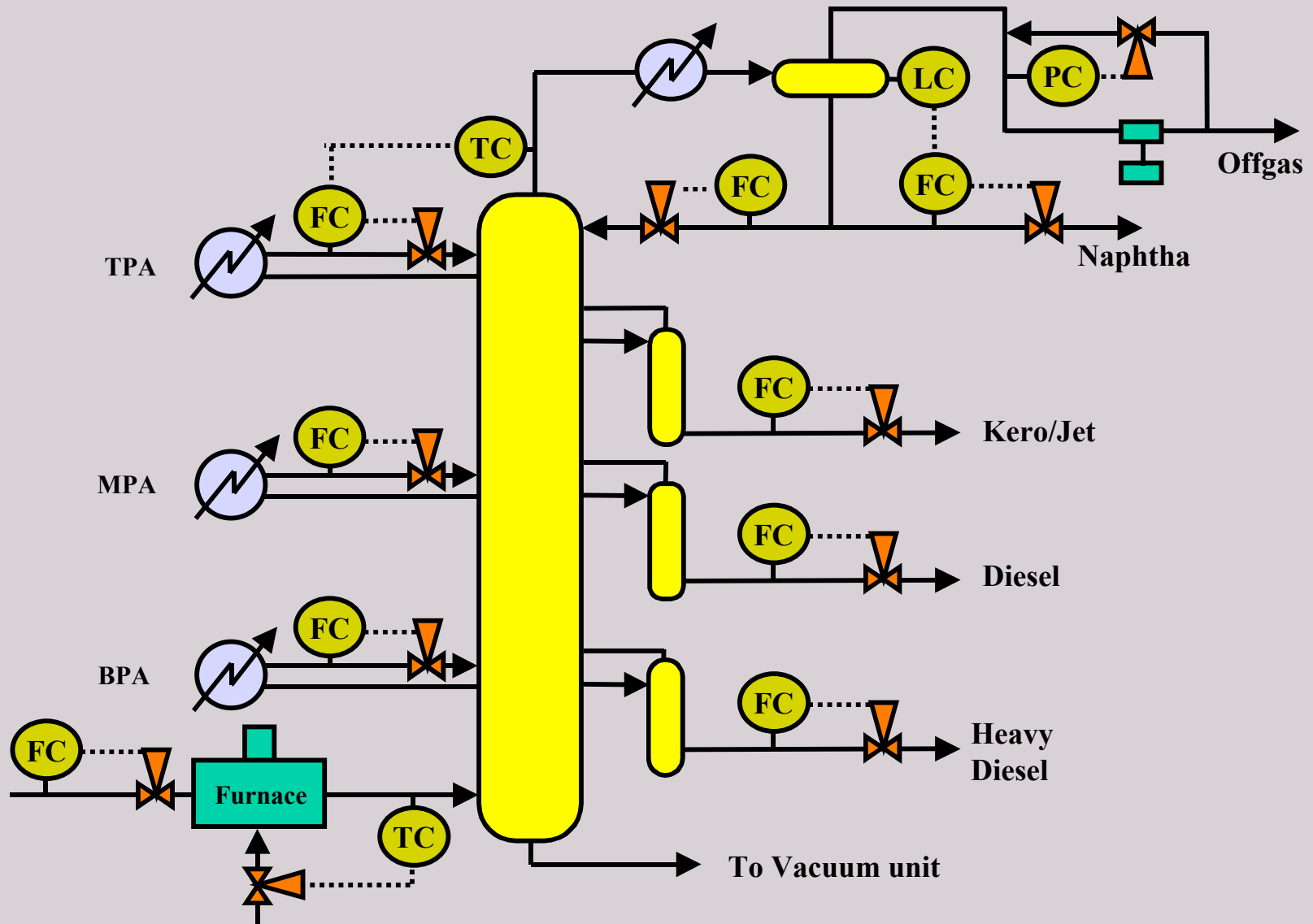
<b>Kickoff meeting</b>	<b>August 03</b>
<b>Functional Design</b>	<b>September 03</b>
<b>Step testing</b>	<b>October 03</b>
<b>Commissioning</b>	<b>January 04</b>

# We came back February to deal with a flaring problem





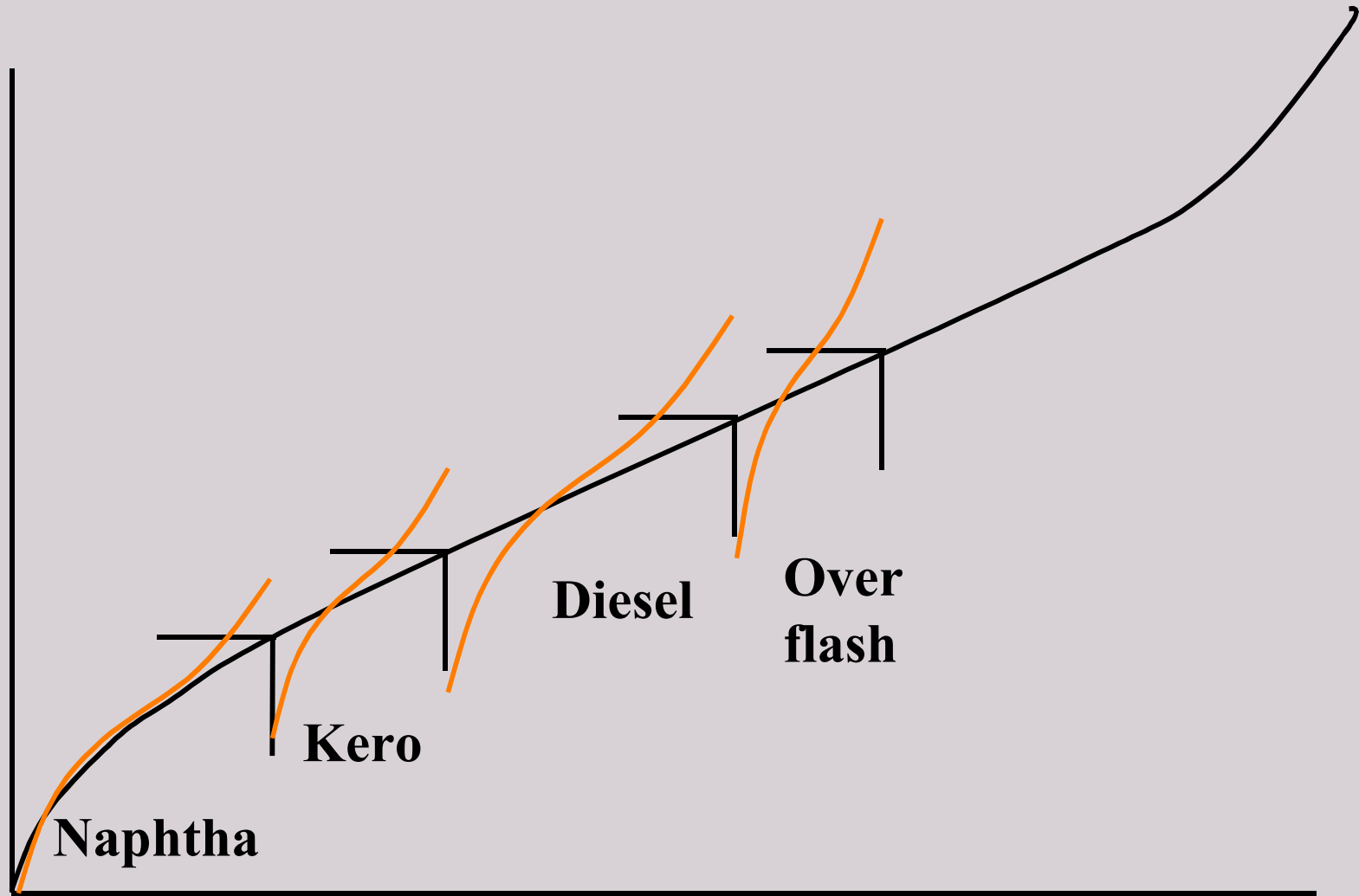
# Petronas CDU 2



# GCC (Generalized cutpoint calculation)

- **Estimates crude TBP curve from unit conditions**
- **Then estimate product qualities:**
  - **ASTMX% = f (cutpoints, internal reflux)**
  - **Flash = f (cutpoints, steam)**
  - **Freeze = f (cutpoints, Kw)**
- **About 40 models implemented to-date**

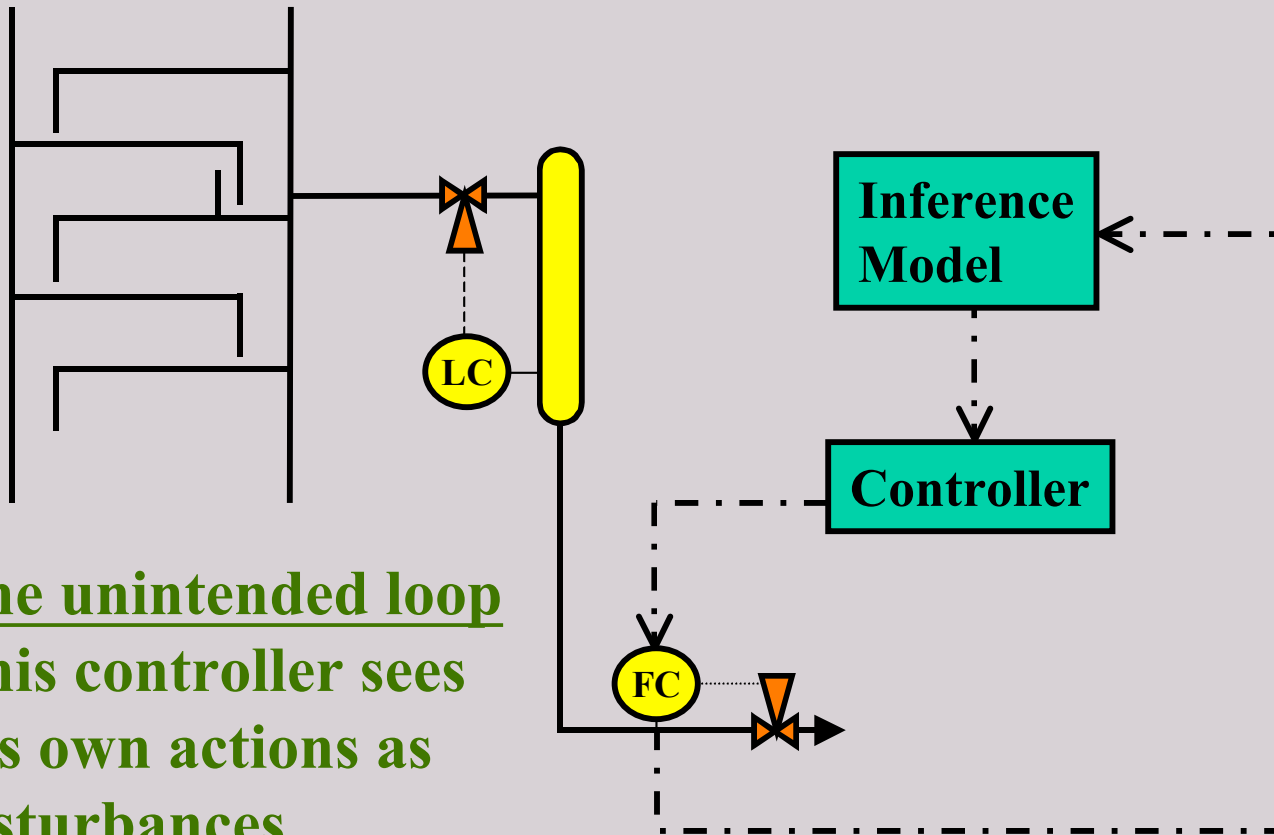
# Crude TBP curve



# GCC main features - 1

- **Heat balance**
  - **During major disturbance the unit can operate off mass balance**
  - **But it is always in heat balance, permitting the model to work during crude switches**
  - **Fractionator cooling load changes with crude**
  - **Quickly detects crude TBP curve changes**

# If you do not use heat balance



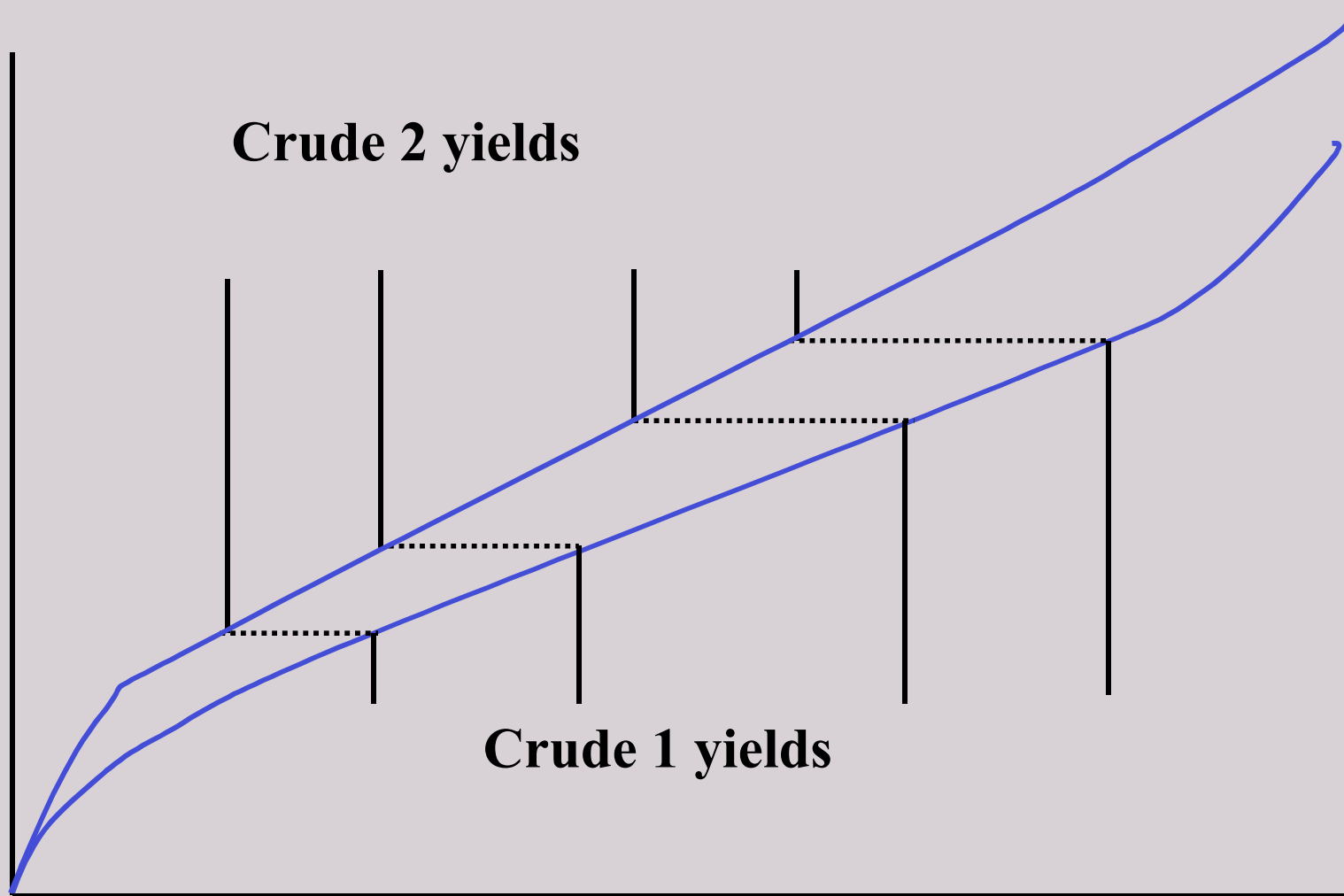
The unintended loop  
This controller sees  
it's own actions as  
disturbances

Heat balance avoids this loop  
by not directly inputting flow

## **GCC main features - 2**

- **Internal reflux model**
  - **Permits precise pumparound control**
  - **Improvement of fuel efficiency or product distillation gap**
  - **Eliminates tray dry-out events**
- **Overflash model**
  - **Eliminates carry-over contamination**
- **These features are key to a smooth crude switch handling**

# How GCC handles crude switches



# Key inferences


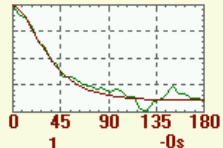







- **Naphtha 90% point**
- **Kero flash, 90% and freeze point**
- **LGO and HGO 90% point**
- **Overflash model, replacing an unreliable overflash measurement**
- **Tray loading in the TPA section**



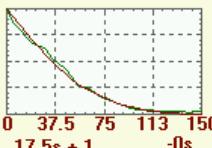

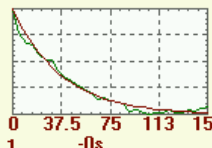



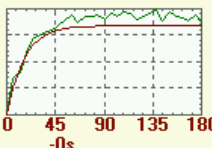
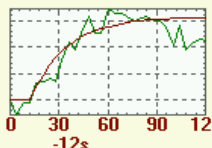
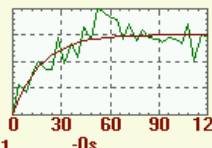
# Other main control variables

- **Desalter temperature**
- **HGO tray dry-out**
- **Furnace constraints**
- **Compressor (**and flaring**) constraints**
- **Pumparound load ratios**
- **Hydraulic constraints**

# Main RMPCT models

	MPA Heat Duty	BPA Heat Duty	TPA Heat Duty
<b>Naphtha 90%</b>	<p>Lap Order 1 Settle T = 120 TfSettle = 124 FIR form = Vel Trial 1</p>  <p><math>G(s) = -3.55 \frac{1}{30.4s + 1} e^{-0s}</math></p>	<p>Lap Order 2 Settle T = 180 TfSettle = 114 FIR form = Vel Trial 3</p>  <p><math>G(s) = -5.53 \frac{1}{349s^2 + 37.4s + 1} e^{-0s}</math></p>	<p>Lap Order 2 Settle T = 180 TfSettle = 108 FIR form = Vel Trial 3</p>  <p><math>G(s) = -4.58 \frac{9.62s + 1}{140s^2 + 32.7s + 1} e^{-0s}</math></p>
<b>Kerosene Freeze Point</b>	<p>Lap Order 2 Settle T = 150 TfSettle = 185 FIR form = Vel Trial 2</p>  <p><math>G(s) = -1.48 \frac{12.3s + 1}{213s^2 + 52s + 1} e^{-0s}</math></p>	<p>Lap Order 2 Settle T = 180 TfSettle = 126 FIR form = Vel Trial 3</p>  <p><math>G(s) = -2.31 \frac{1}{431s^2 + 41.5s + 1} e^{-0s}</math></p>	<p>Lap Order 2 Settle T = 180 TfSettle = 174 FIR form = Vel Trial 3</p>  <p><math>G(s) = -1.91 \frac{9.79s + 1}{63.6s^2 + 46s + 1} e^{-0s}</math></p>
<b>Kerosene 90%</b>	<p>Lap Order 2 Settle T = 150 TfSettle = 190 FIR form = Vel Trial 2</p>  <p><math>G(s) = -4.77 \frac{110s + 1}{4563s^2 + 144s + 1} e^{-0s}</math></p>	<p>Lap Order 2 Settle T = 150 TfSettle = 175 FIR form = Vel Trial 2</p>  <p><math>G(s) = -7.44 \frac{1}{531s^2 + 52.7s + 1} e^{-0s}</math></p>	<p>Lap Order 2 Settle T = 180 TfSettle = 192 FIR form = Vel Trial 3</p>  <p><math>G(s) = -6.15 \frac{13.7s + 1}{143s^2 + 53s + 1} e^{-0s}</math></p>

# Main RMPCT models (cont.)

	MPA Heat Duty	BPA Heat Duty	TPA Heat Duty
<b>LGO 90%</b>	<p>Lap Order 2 Settle T = 150 TfSettle = 150 FIR form = Vel Trial 2</p>  $G(s) = -9.75 \frac{17.5s + 1}{1169s^2 + 56.7s + 1} e^{-0s}$	<p>Lap Order 2 Settle T = 180 TfSettle = 168 FIR form = Vel Trial 3</p>  $G(s) = -15.2 \frac{1}{759s^2 + 55.1s + 1} e^{-0s}$	<p>Lap Order 1 Settle T = 150 TfSettle = 150 FIR form = Vel Trial 2</p>  $G(s) = -12.6 \frac{1}{37.1s + 1} e^{-0s}$
<b>HGO 90%</b>	<p>Lap Order 2 Settle T = 150 TfSettle = 150 FIR form = Vel Trial 2</p>  $G(s) = -10.9 \frac{14.5s + 1}{1084s^2 + 54.2s + 1} e^{-0s}$	<p>Lap Order 2 Settle T = 180 TfSettle = 174 FIR form = Vel Trial 3</p>  $G(s) = -17 \frac{1}{803s^2 + 56.7s + 1} e^{-0s}$	<p>Lap Order 1 Settle T = 150 TfSettle = 155 FIR form = Vel Trial 2</p>  $G(s) = -14 \frac{1}{38.7s + 1} e^{-0s}$
<b>Overflash</b>	<p>Lap Order 1 Settle T = 180 TfSettle = 66.0 FIR form = Vel Trial 3</p>  $G(s) = 7.93 \frac{1}{16s + 1} e^{-0s}$	<p>Lap Order 1 Settle T = 120 TfSettle = 104 FIR form = Vel Trial 1</p>  $G(s) = 12.4 \frac{1}{22s + 1} e^{-12s}$	<p>Lap Order 1 Settle T = 120 TfSettle = 72.0 FIR form = Vel Trial 1</p>  $G(s) = 10.2 \frac{1}{17.4s + 1} e^{-0s}$

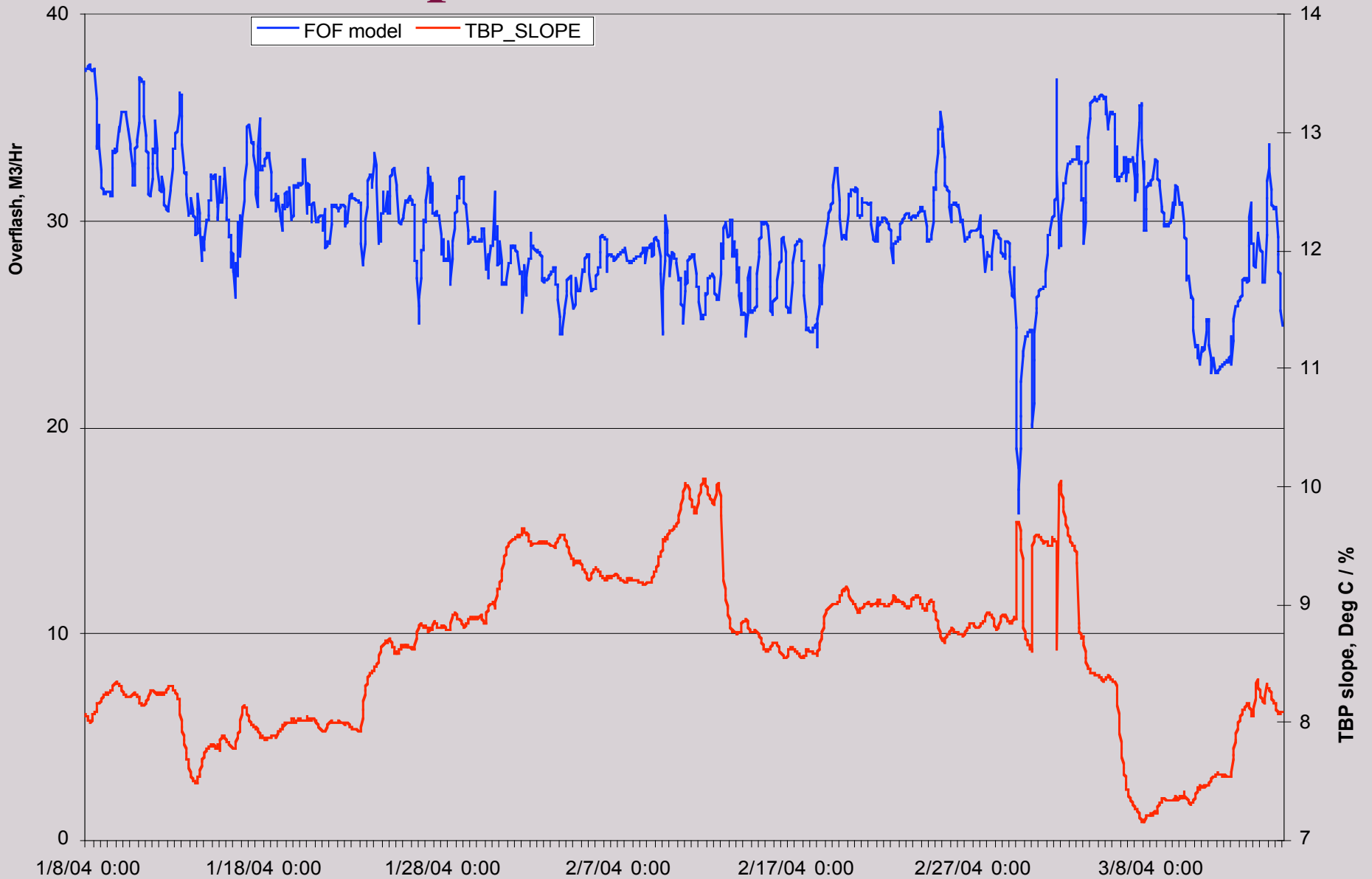
# RMPCT considerations

- **Aggressive Tuning maintained during both normal operation and crude switch**
  - Set RMPCT control horizon as fast as 20% of open loop response time for key controlled variables
  - No need of special crude switch tuning sets
- **Optimization Configuration for**
  - Furnace Inlet Pressure Maximization

# Typical crude mixes

Dates	Crude Rate By Type (MB/D) ME = Middle East M'SIA = Malaysia	CDU/VDU Yields From Crude & LSWR (MB/D)				
		Ovhd	Kero	Diesels	VGO	Resid
8 – 12 Jan	55 ME1 (0.88) / ME2 (0.86) 70 ME3 (0.88) / ME2 (0.86)	25	18	31.2	32.6	25.8
13 Jan	55 ME4 (0.87) / M'SIA (0.86) 70 ME3 (0.88) / ME2 (0.86)	24.7	18	35.2	31.2	23.8
14 – 19 Jan	55 ME4 (0.87) / M'SIA (0.86) 70 ME4 (0.87)	25.5	18	35.7	30	22.9
20 – 22 Jan	125 ME4 (0.87) / ME2 (0.86) / ME3 (0.88)	24	18	32.7	30.5	28
23 – 26 Jan	125 ME4 (0.87)	24.9	18	33.9	29.6	25.7
27 – 29 Jan	125 ME4 (0.87) / ME2 (0.86)	25.5	18	33.5	30.5	24.5
30 Jan – 05 Feb	65 ME5 (0.89) / ME6 (0.86) 60 ME1 (0.88) / ME6 (0.86)	24	18	31.3	30.8	29.5

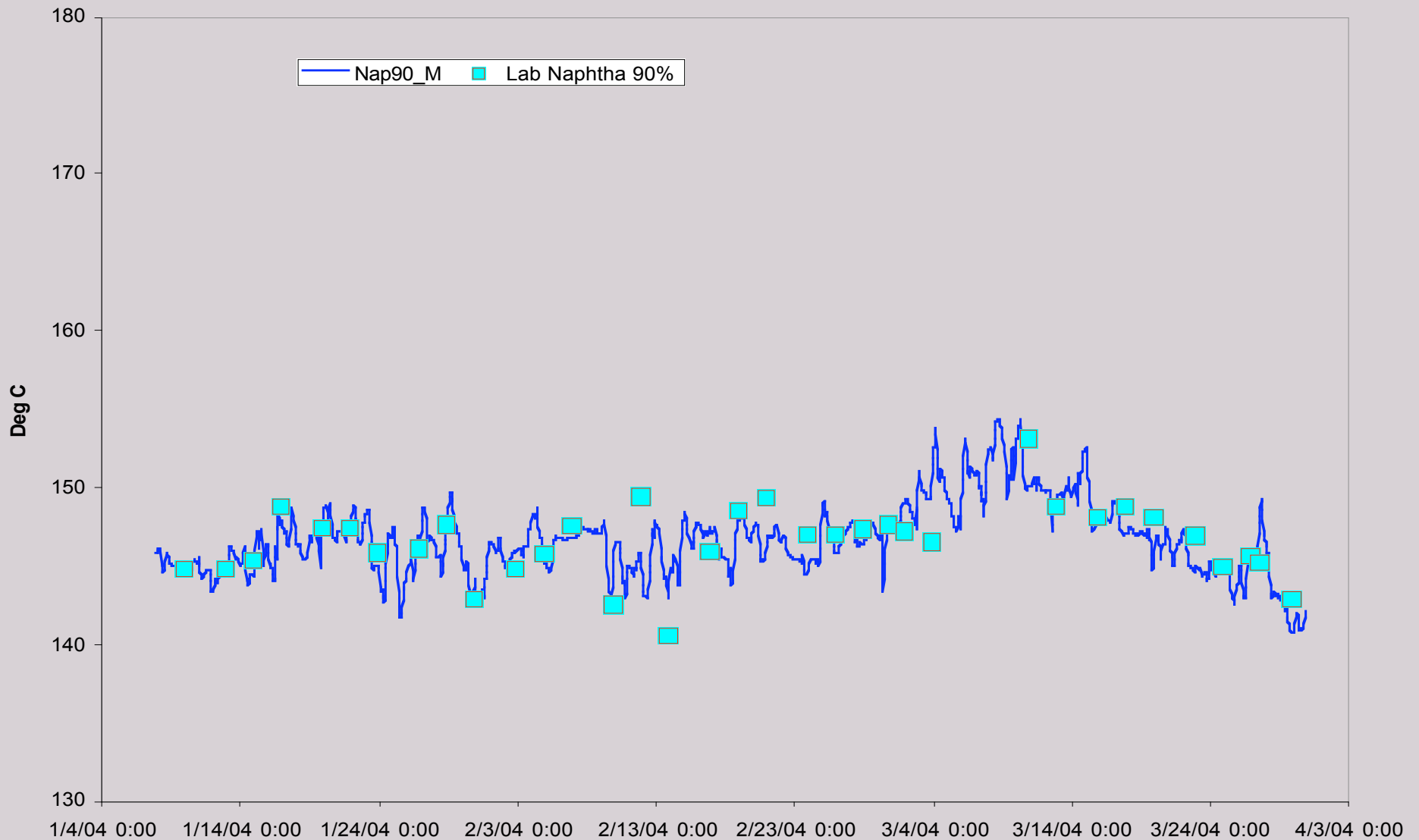
# TBP slope and overflash trend



# Inference versus lab trends

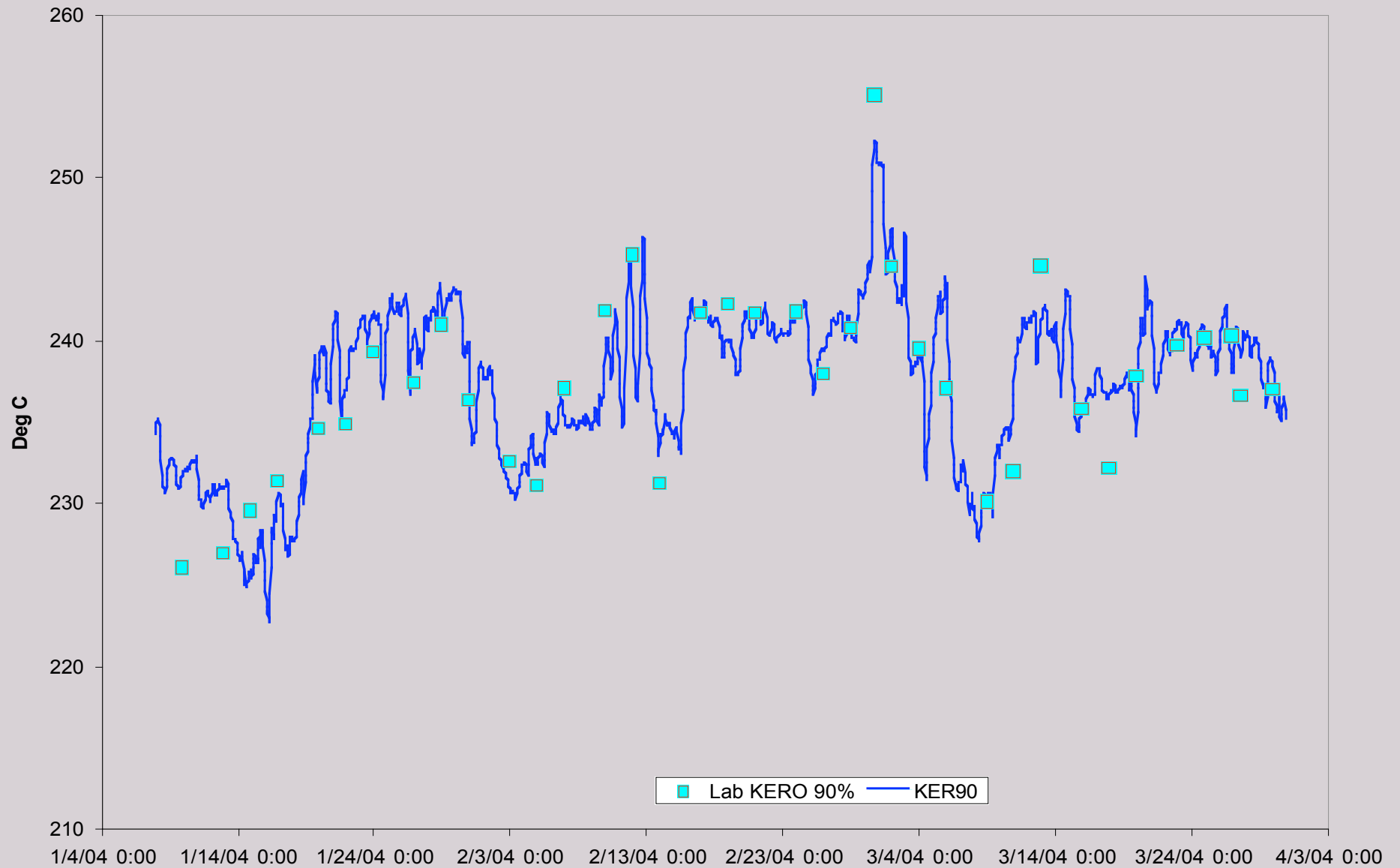
- **Slide 21: Naphtha 90% model versus lab**
- **Slide 22: Kero 90% model versus lab**
- **Slide 23: LGO 90% model versus lab**
- **Slide 24: HGO 90% model versus lab**
- **Slide 25: Kero flash model versus lab**
- **Slide 26: Kero freeze model versus lab**

# Naphtha 90% inference versus Lab

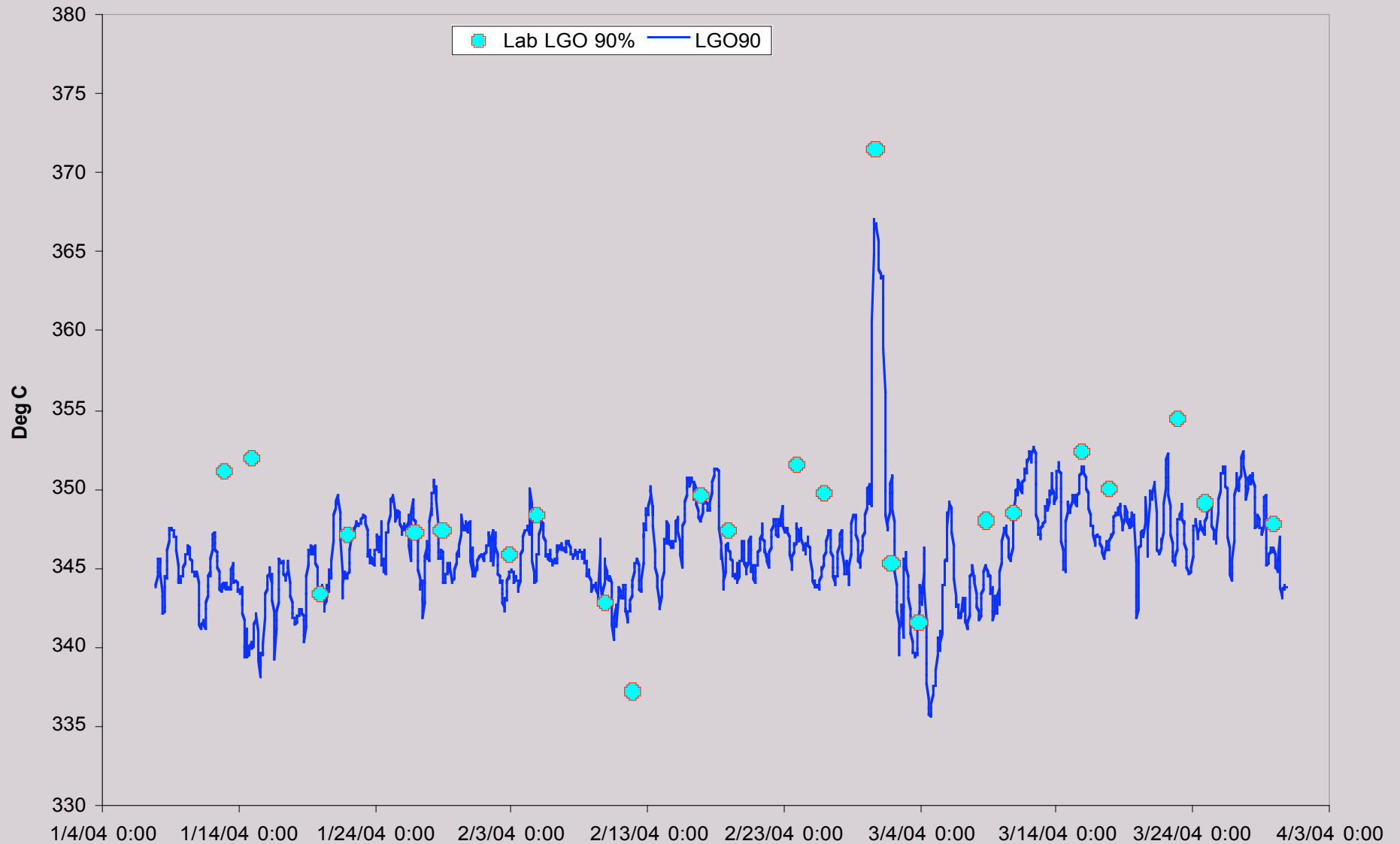




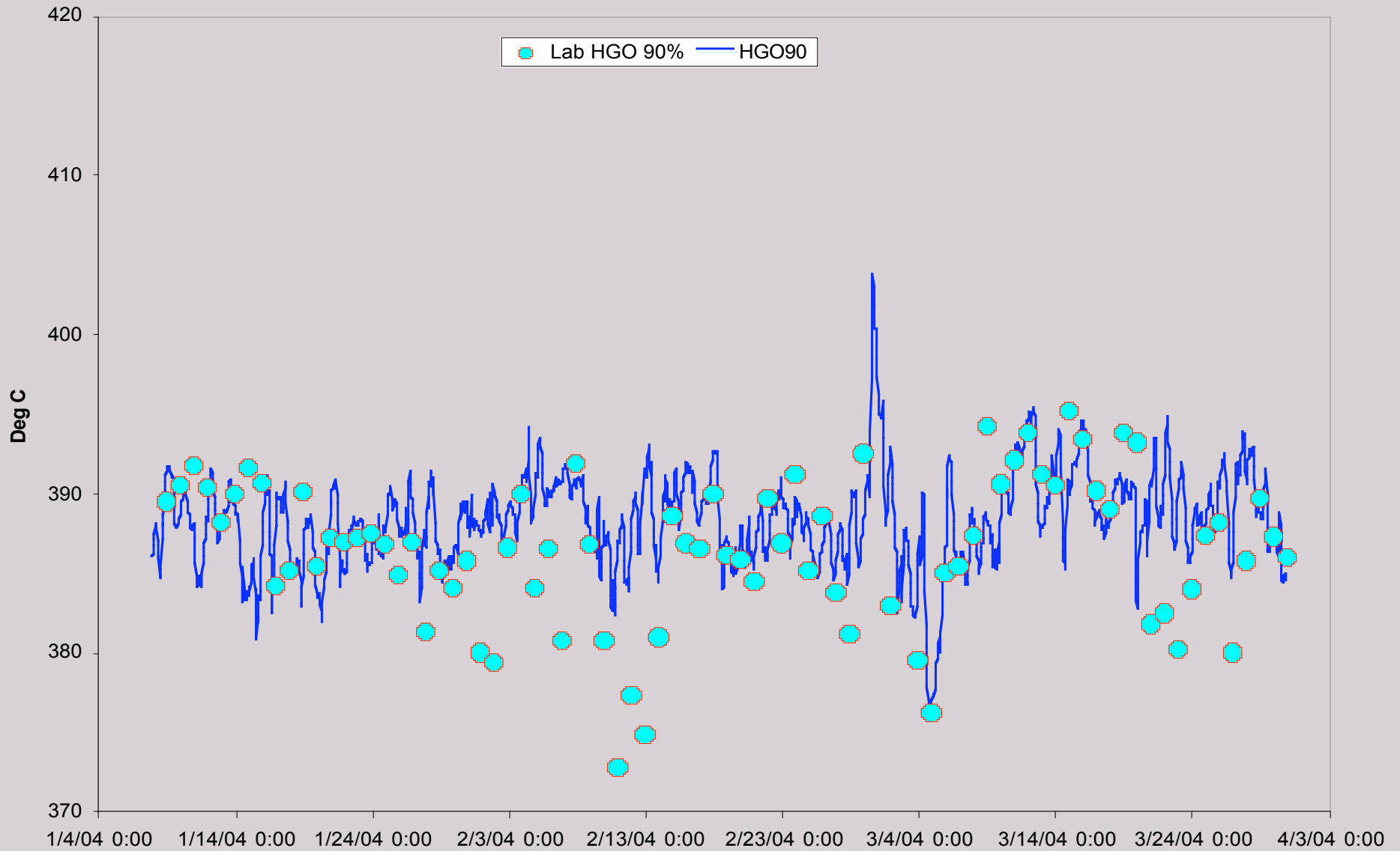
# Kero 90% inference versus Lab



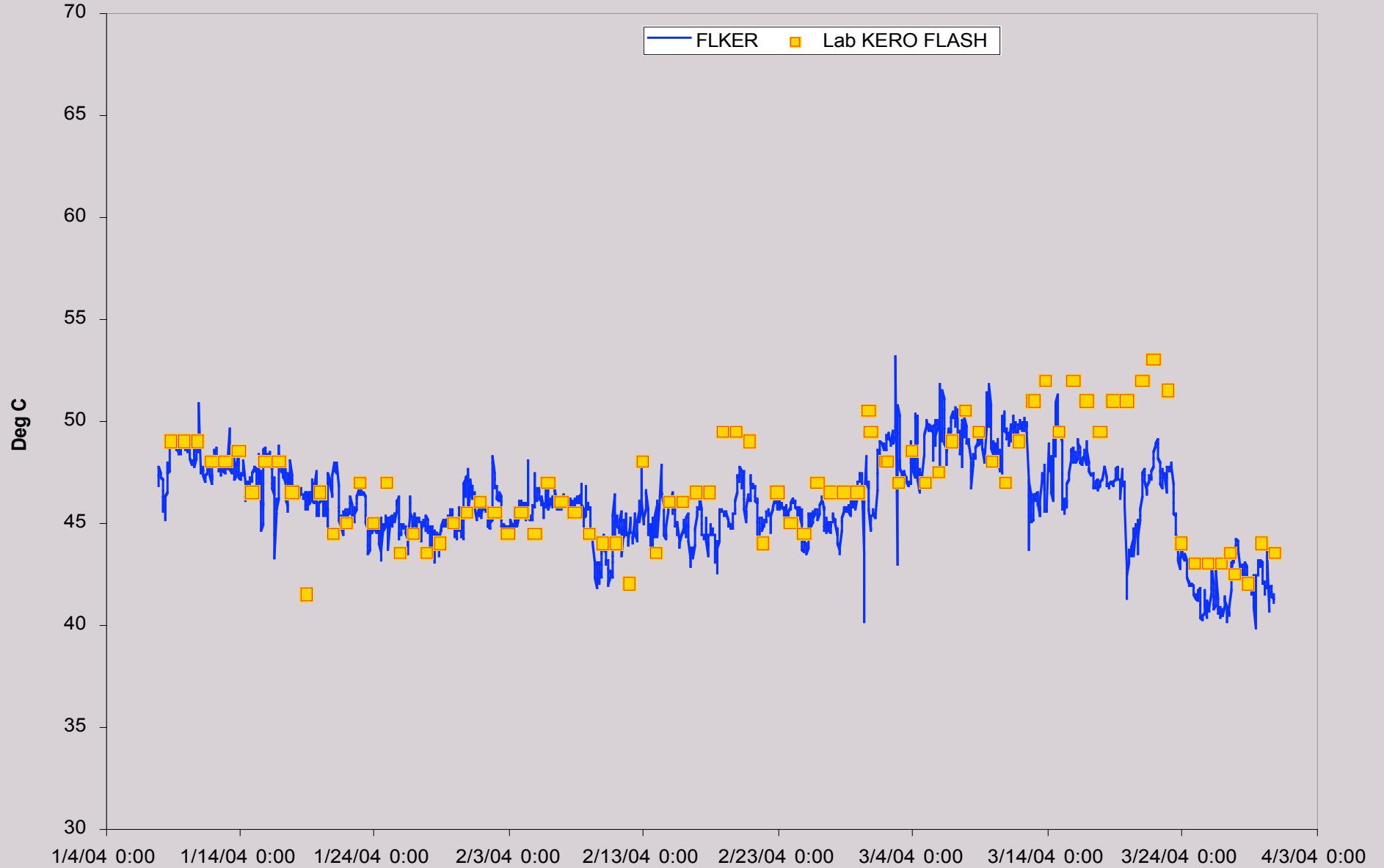
# LGO 90% inference versus Lab



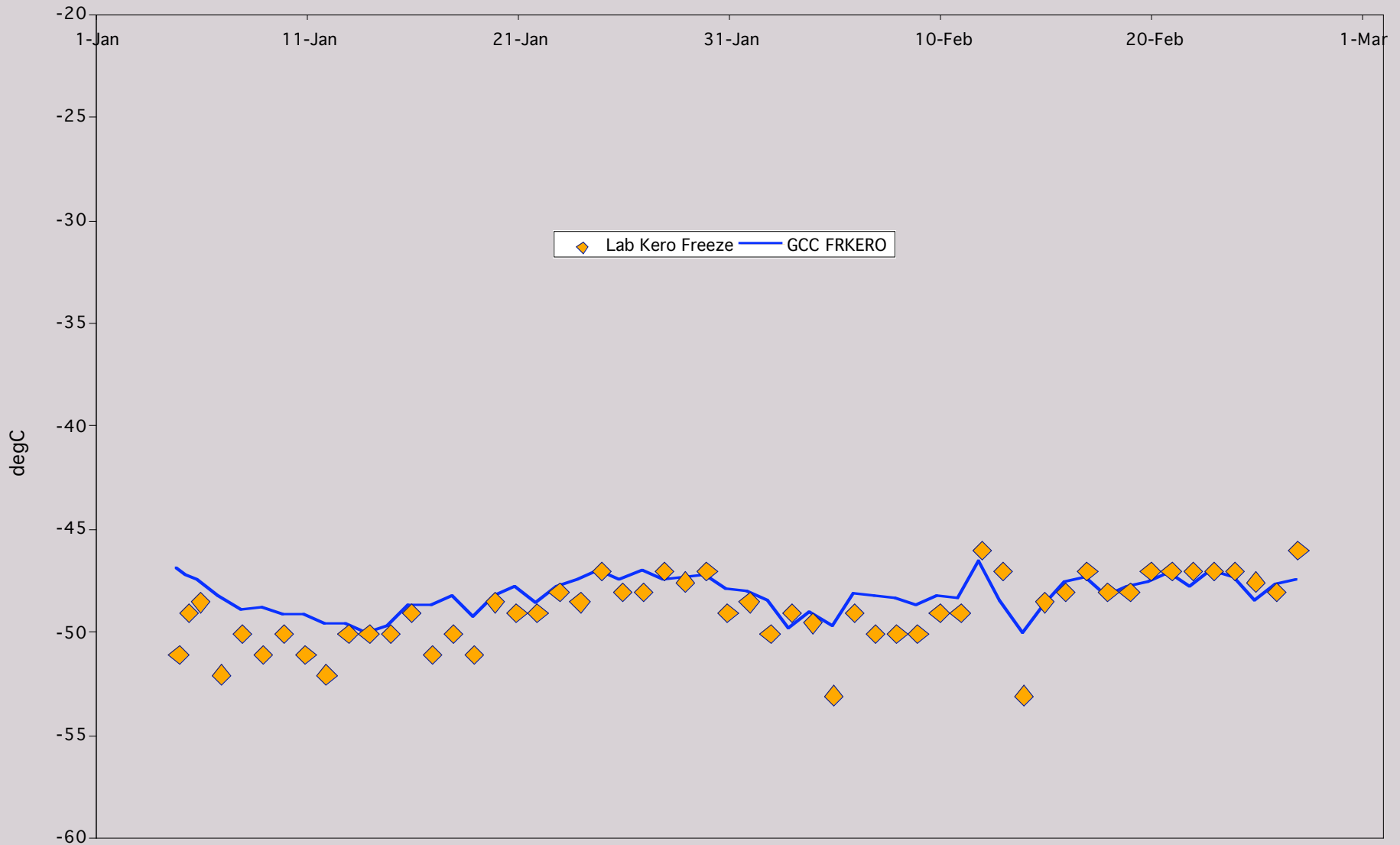
# HGO 90% inference versus Lab



# Kero Flash inference versus Lab



# Kero freeze inference versus Lab



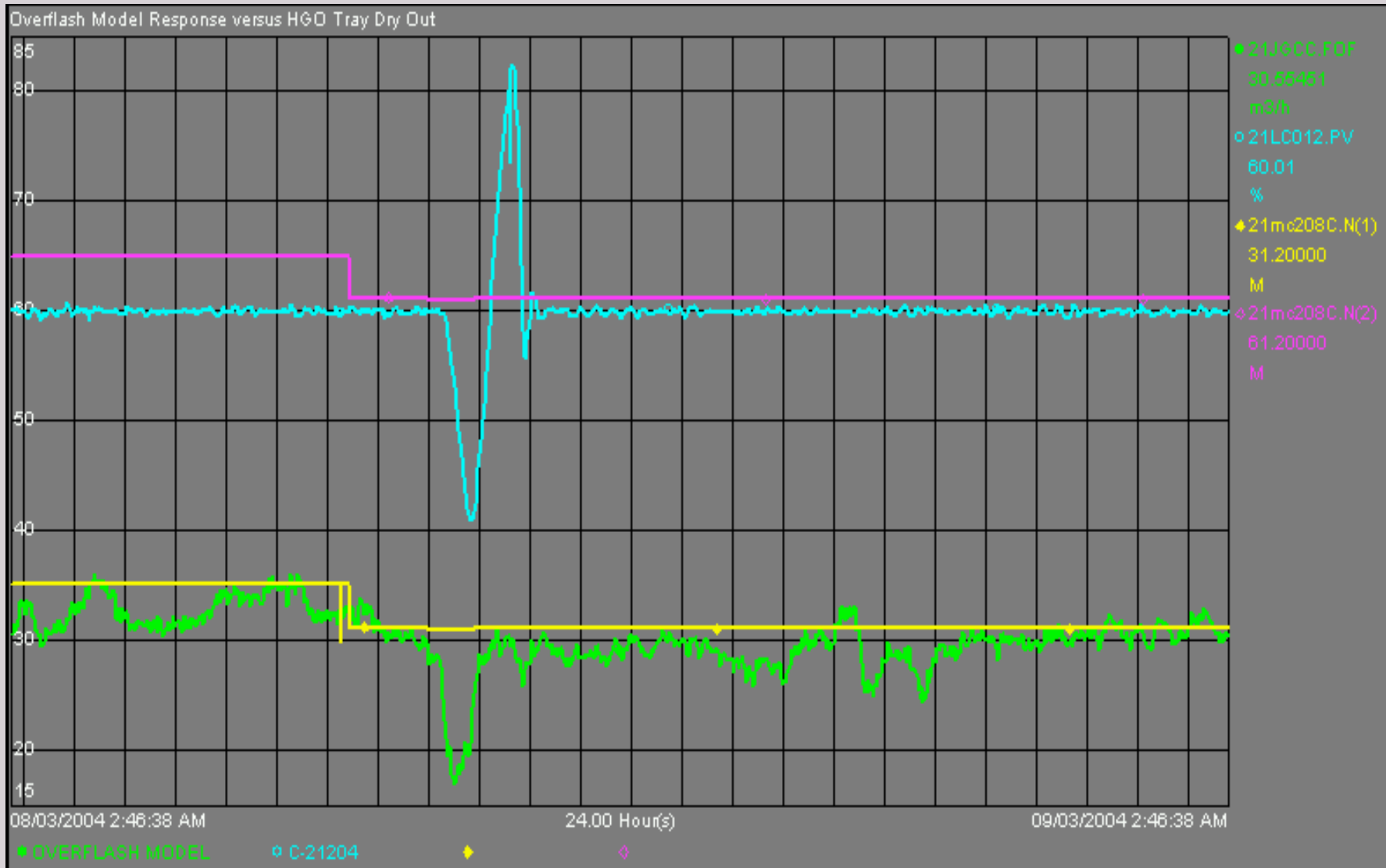
# Dynamic trends of interest

- **Slides 28, 29: overflash and tray dry-out**
- **Slides 30-32: Jan 6<sup>th</sup>, 2004 crude switch**
- **Slides 33-35: Jan 13<sup>th</sup>, 2004 crude switch**
- **Slides 36-38: Feb 12<sup>th</sup>, 2004 crude switch**
- **Slides 39-41: Mar 13<sup>th</sup>, 2004 crude switch (unannounced)**
- **Slides 42-44: Feb 14<sup>th</sup>, 2004 typical crude segregation phenomenon**

# Overflash model

- **Slide 29 shows a PI trend of overflash, overflash limits and HGO stripper level**
- **At the time, due to insufficient training the operator specified infeasible limits, causing RMPCT to ignore the overflash limit and operate below it**
- **A tray dry-out occurred**
- **Tray dry-out has a high priority and RMPCT quickly recovered**

# Overflash model accuracy

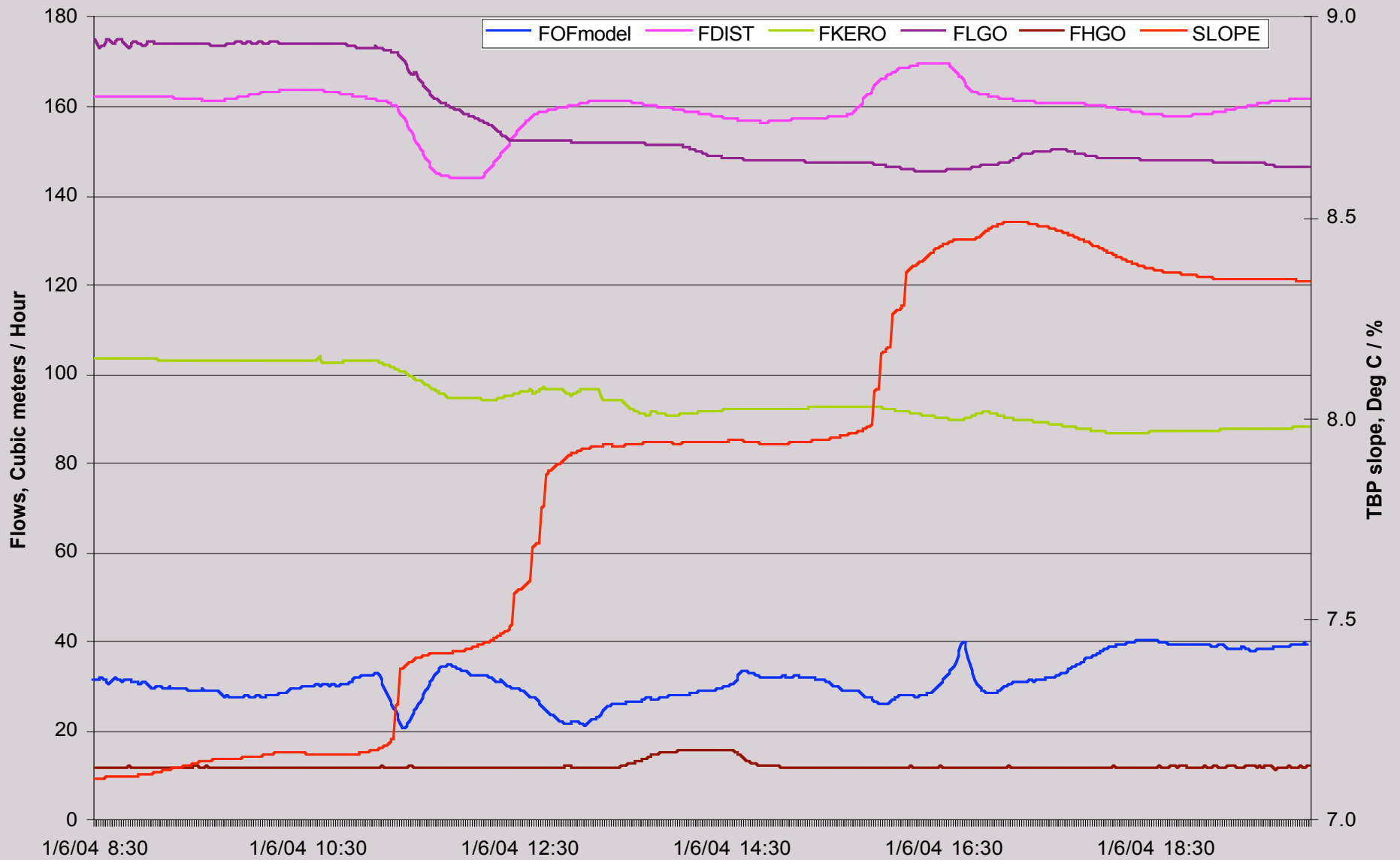




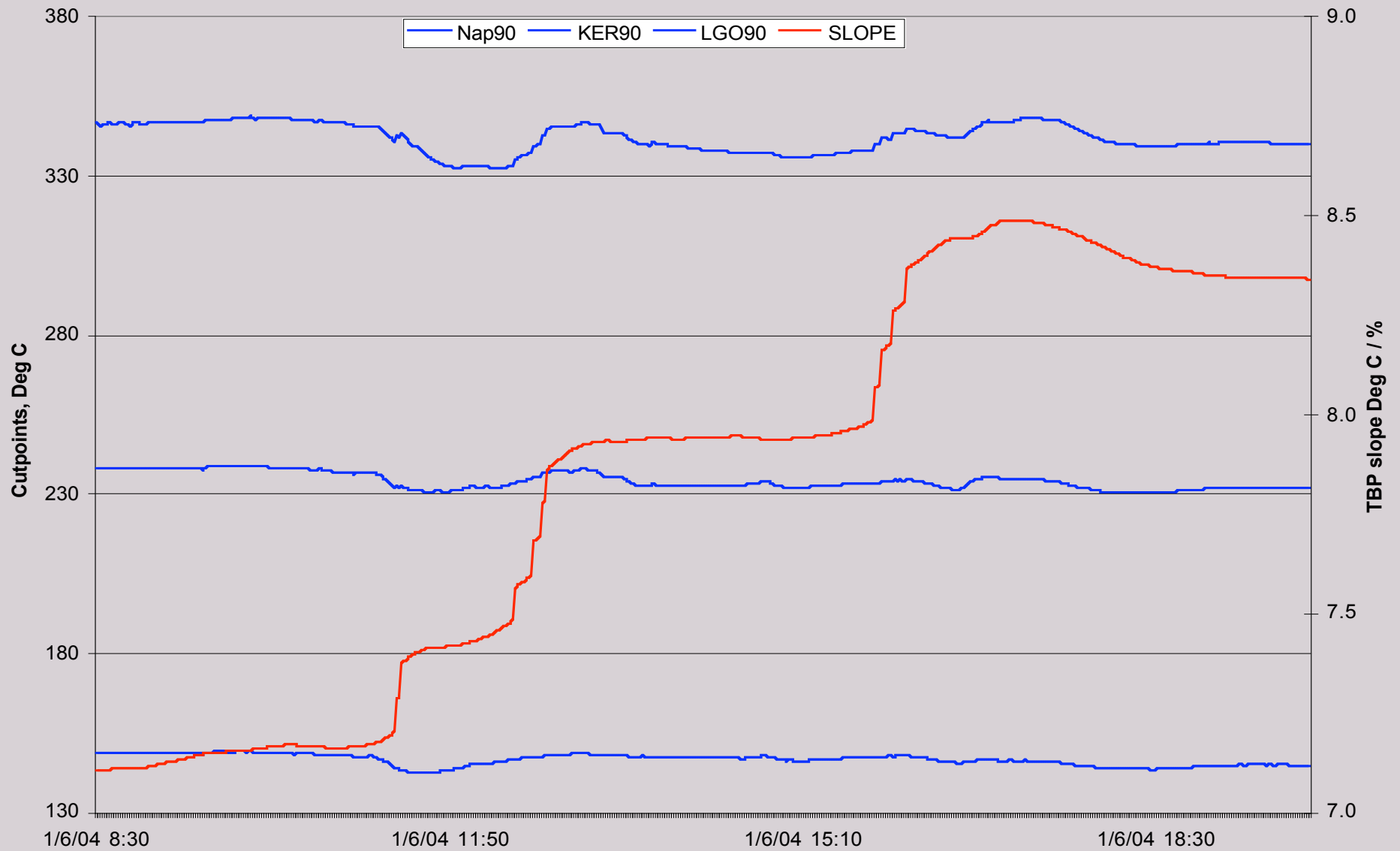
# First crude switch Jan 6<sup>th</sup>

- **Three crude switches on the same day**
- **Under *partial* MVPC control**
- **Many MV's clamped**
- **(Commissioning started Jan 5<sup>th</sup>)**
- **LGO cut deviations ~ 10°C, slide 32**

# Jan 6<sup>th</sup>, 2004 crude switch - flows



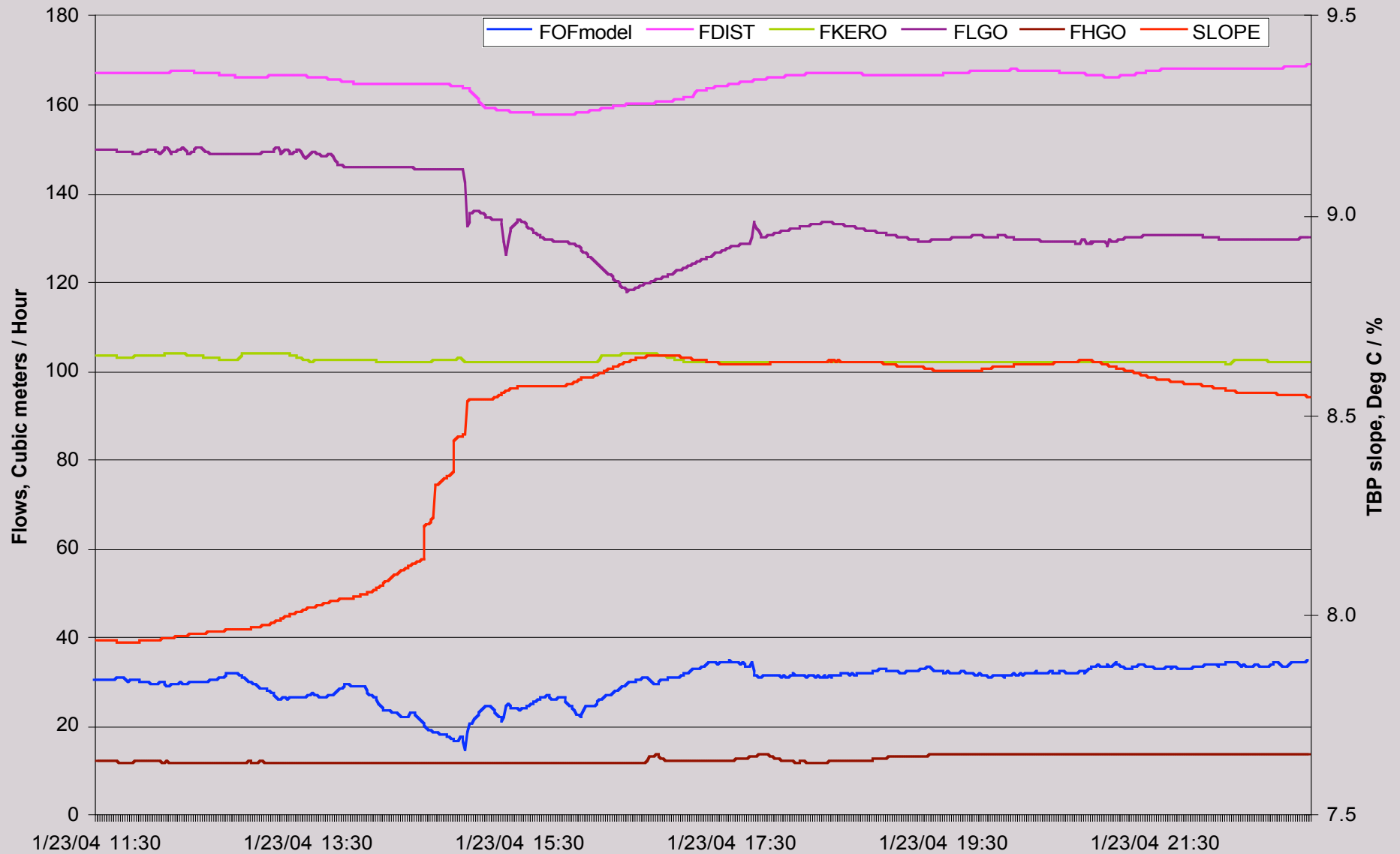
# Jan 6<sup>th</sup>, 2004 crude switch - cuts



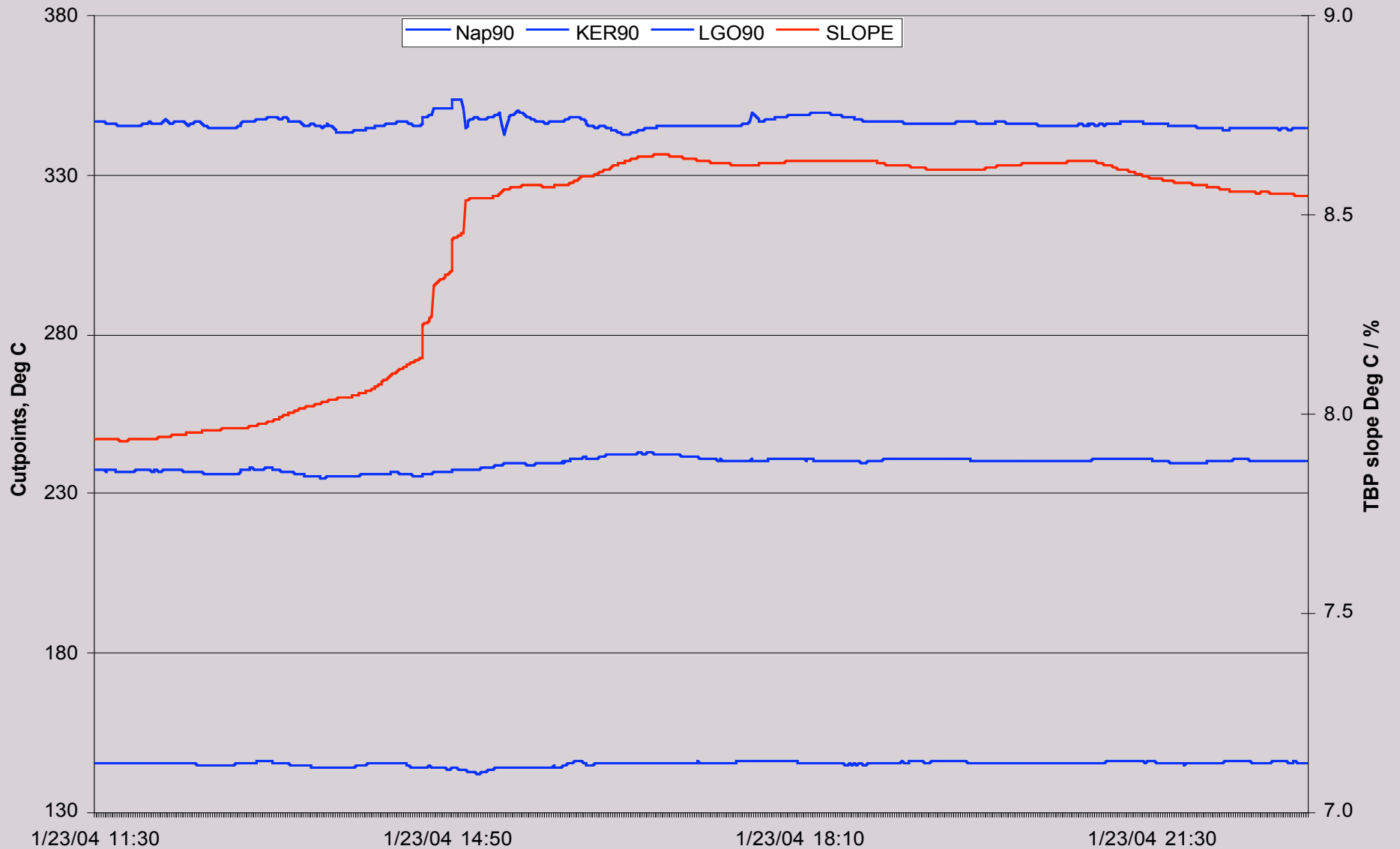
# Second crude switch Jan 13<sup>th</sup>

- **Nearly complete MVPC control**
- **Some MV's clamped**
- **LGO cut deviations ~ 5°C, slide 35**

# Jan 13<sup>th</sup>, 2004 crude switch - flows



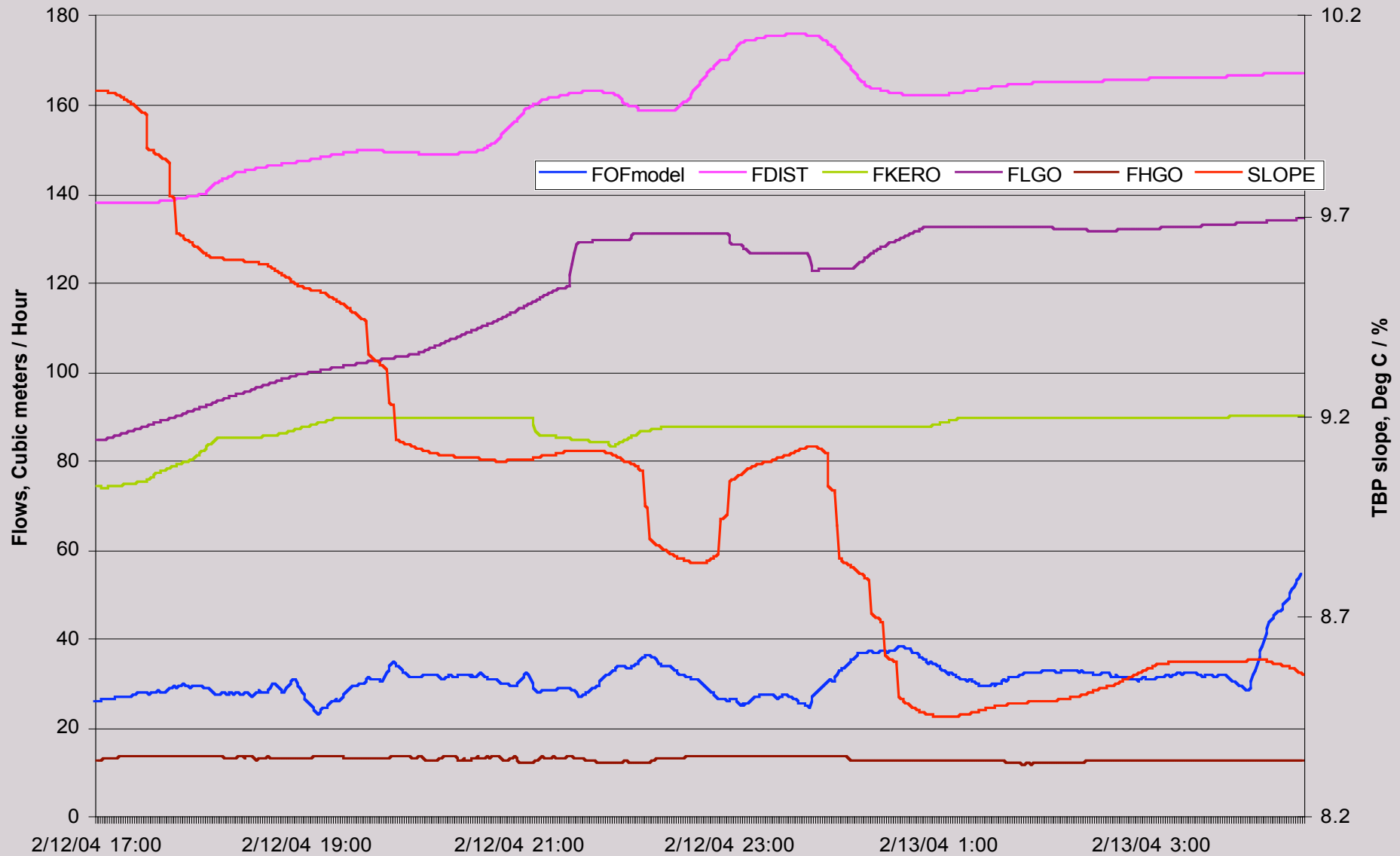
# Jan 13<sup>th</sup>, 2004 crude switch - cuts



# Feb 12<sup>th</sup> crude switch

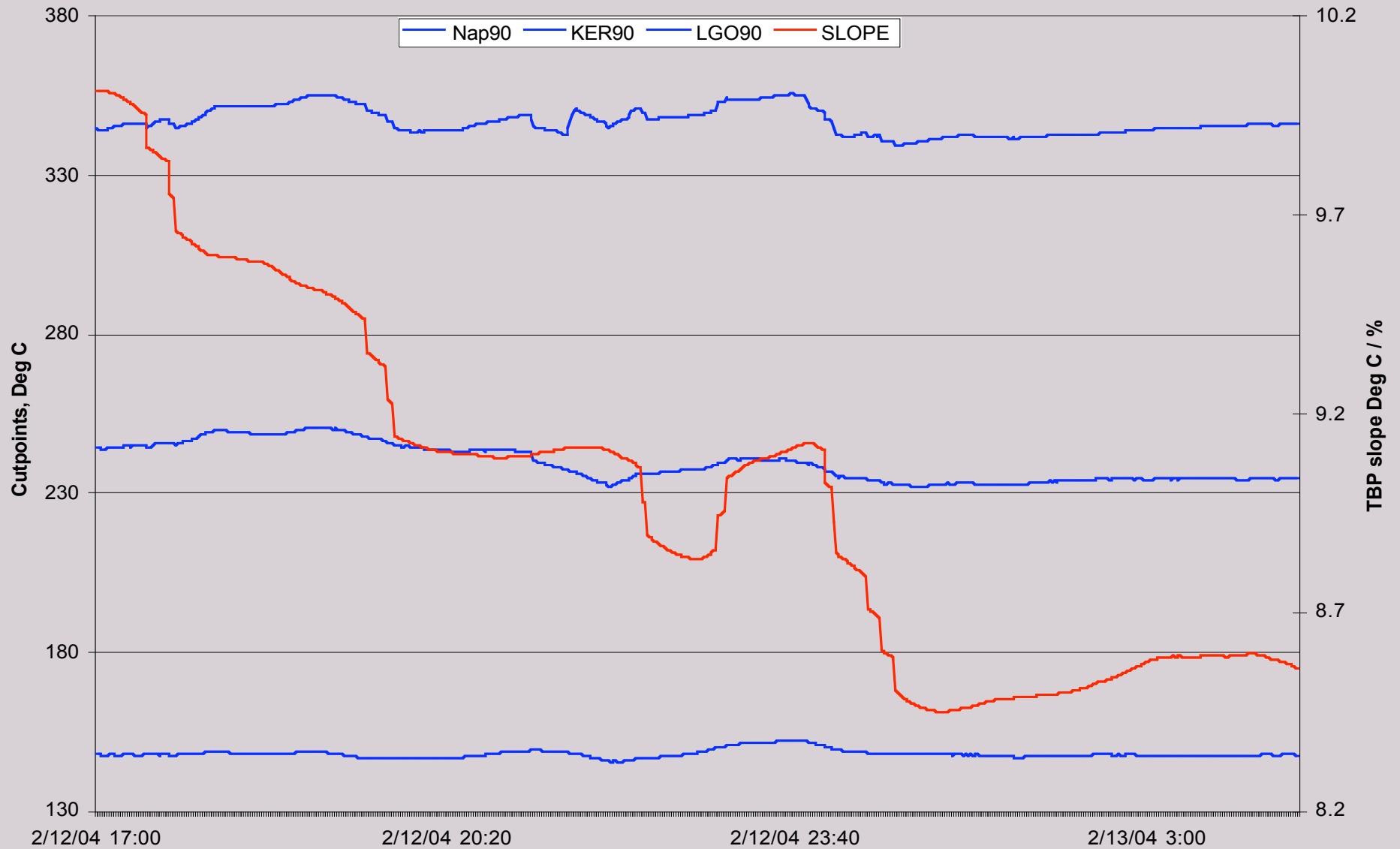
- **Complete MVPC control**
- **No superficial clamps**
- **LGO cut deviations ~ 5°C, slide 38**

# Feb 12<sup>th</sup>, 2004 crude switch - flows





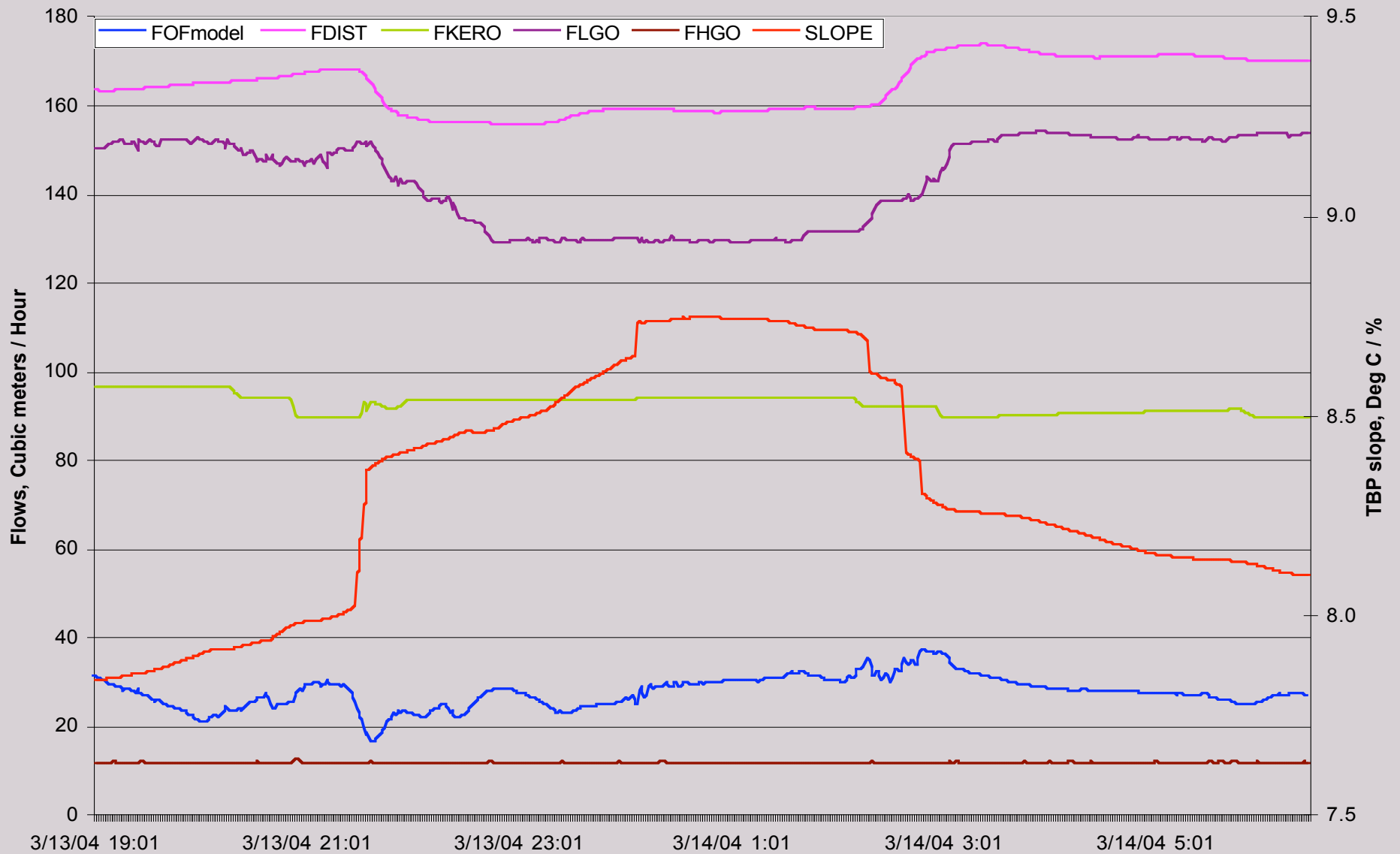
# Feb 12<sup>th</sup>, 2004 crude switch - cuts



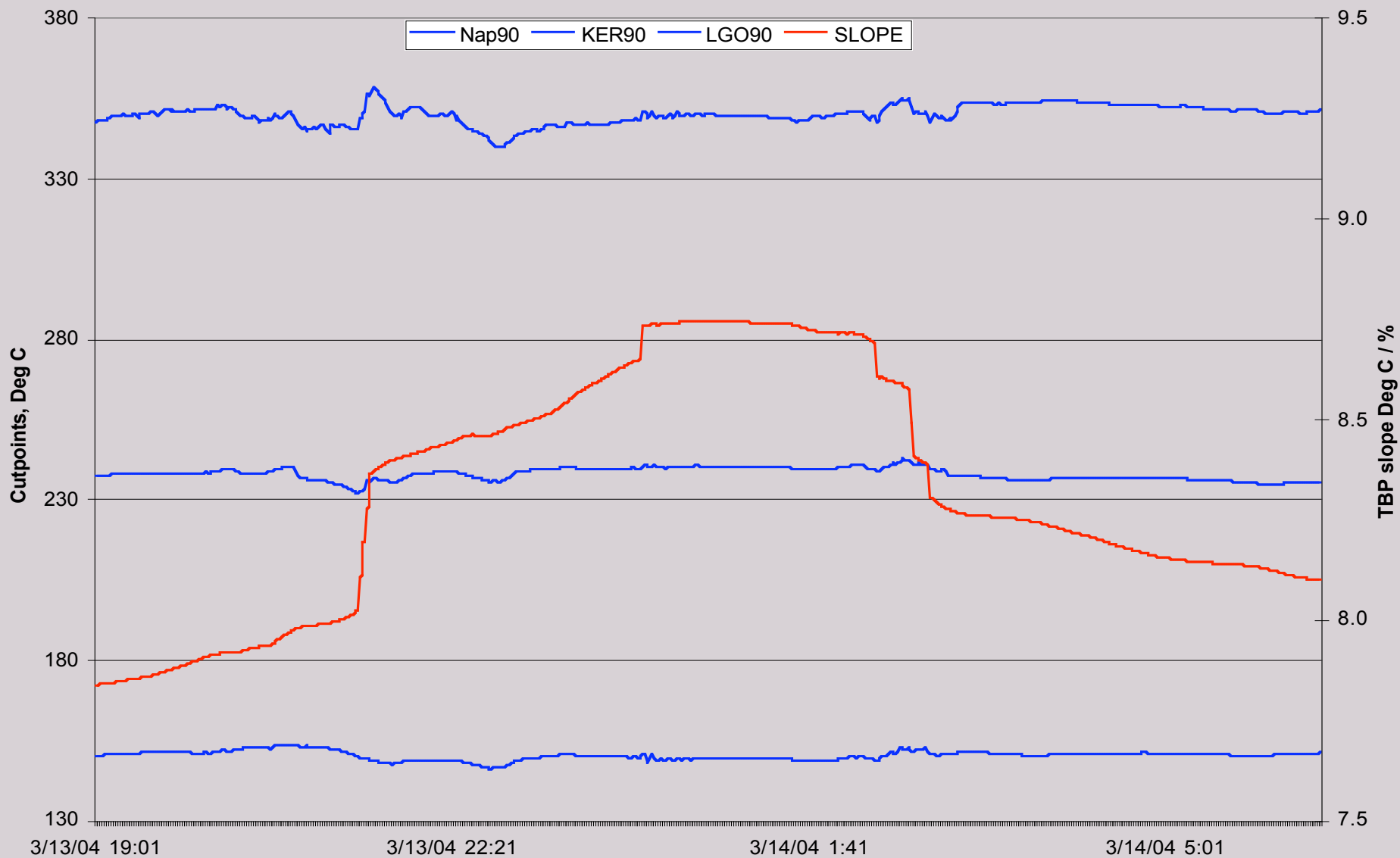
# Unannounced crude switches

- **Crude ratio adjustment**
- **Added slops**
- **Other reasons**
- **March 13<sup>th</sup>, 2004 example, slides 40, 41**

# Unannounced switch, Mar 13<sup>th</sup> - flows



# Unannounced switch, Mar 13<sup>th</sup> - cuts



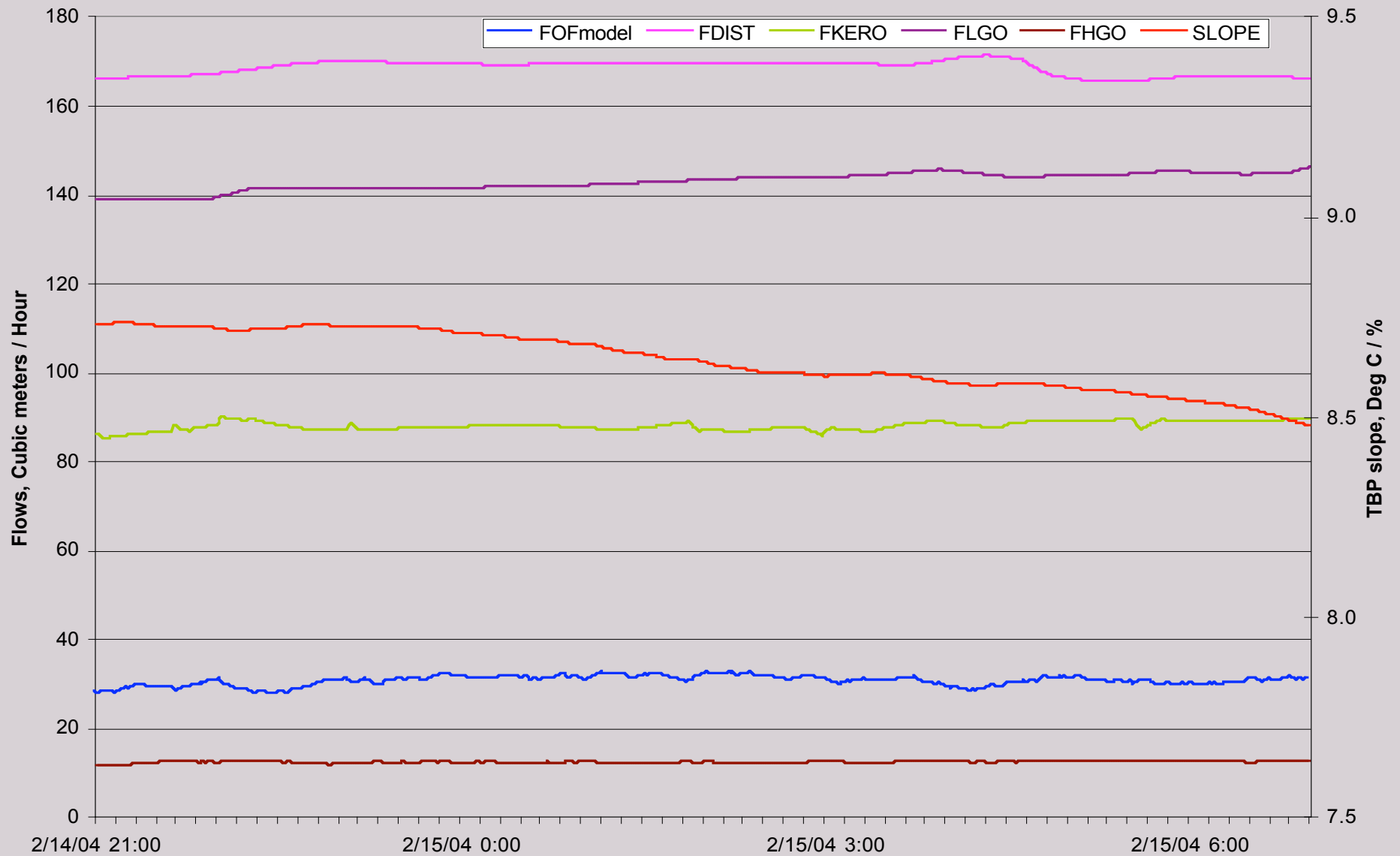
# Tank segregation example

- Layering
- Different heel

$\alpha$ Crude becomes lighter as the tank is depleted

- Feb 14<sup>th</sup>, 2004 example, slides 43, 44

# Tank segregation, Feb 14<sup>th</sup> - flows



# Tank segregation, Feb 14<sup>th</sup> - cuts



# Benefits of GCC

- **Crude switch time**  
**from several hours  $\alpha$  1 hour**
- **Keep throughput high**
- **Eliminate product downgrading**
- **Overflash control**
- **Internal reflux profile**
- **Reduce number of alarms (and incidents)**
- **Reduce lab support requirements**
- **Information: crude TBP, other indicators**