

# IMPLEMENTATION OF APC ON REPSOL PUERTOLLANO CDU 1

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# **IMPLEMENTATION OF APC ON REPSOL PUERTOLLANO CDU1**

## **Introduction (slide 2)**

In this paper, the implementation of a new multivariable process controller (MVPC) for Crude Distillation Unit 1 at Repsol YPF refinery in Puertollano is presented. The paper reviews the improvements generated from this work, in two aspects: better product quality control using GCC inferencial model, and enhanced operation and control of the unit achieved with the new MVPC.

## **Repsol YPF and Puertollano refinery site (slides 3 and 4)**

Repsol YPF is an international integrated oil and gas company, operating in over 30 countries. The installed refining capacity of the company is 1.2 MM bpd, 70% of it is installed in Spain and the rest in several South American countries.

Repsol YPF operates five refineries in Spain. The positioning of Puertollano Industrial Complex in the center of the peninsula is ideal for covering the fuel product needs of Madrid and the center area. Puertollano Industrial Complex has a crude oil distillation capacity of 7.5 million tonnes a year. The Puertollano refinery is technically the most complex of all the Spanish refineries, due to both the product variety and the integration of the units.

The Advanced Control group responsible for APC in the Repsol YPF spanish refineries is composed of a central group plus local groups at each refinery site. Nowadays, the efforts of the group are concentrated on the maintenance of existing multivariable controllers, and the implementation of new ones in the most relevant units of the refineries.

## **The rational for Puertollano CDU APC (slide 5)**

Puertollano CDU1 is not a large unit by industry standard, but the unit operational complexity and frequency of crude switches were enough to justify APC. Puertollano CDU1 has gained substantial yield improvement and can also gain throughput increases.

It is strange but appropriate to note that further, not insignificant benefit of APC is that it provides immediacy and focus to get operational problems solved. This was the case at Puertollano. Many of the problems that hindered APC were not control related, but our project provided a catalyst to get the problems investigated and solved. The solutions were a

multi-disciplinary team effort between operations, process, instrumentation, and the advanced control groups working together.

### **The Project team (slide 6)**

As is well known, the key to CDU APC is repeatable inferential models, estimating product qualities at all times and especially during crude switches. Repsol has elected to work with the Petrocontrol GCC (Generalized cutpoint calculation) technology. Petrocontrol implemented GCC together with Repsol technology center engineers as well as Puertollano engineers.

Hall Controls was selected as the consultant in the design and implementation of the application. Repsol provided the strong support of a full time engineer.

The application makes use of MVPC (multivariable predictive control software): Aspen Technology's DMCplus®, for which Repsol has a site wide license.

### **CDU Configuration (slide 7)**

The unit is configured with a prefractionator, separating out light naphtha, and a main atmospheric column. The main column separates a top product of heavy naphtha and four side streams: kero, diesel, heavy diesel and AGO. Commercially the desire is as usual to maximize the sidestreams to cutpoint specifications.

Light and heavy naphtha are combined and taken through a stabilizer and naphtha splitter.

The main fractionator has four cooling circuits:

- Sub-cooled external reflux returned to the top tray
- Top pumparound (TPA) exchanges heat against crude before the desalter and returned to the top tray
- Middle pumparound (MPA) exchanges heat against crude after the desalter
- Bottom pumparound (BPA) which supplies heat for the naphtha stabilizer and crude preheat.

### **GCC concepts (slide 8)**

GCC has been described in several papers (1, 2, 3, 4, 5, 6), and time here permits a summary but not detailed discussion. GCC applies heat balances, mass balances, partial pressure

corrections and boiling curve conversions to estimate the crude TBP (true boiling point) curve. Then from TBP and internal reflux, GCC proceeds to estimate product properties. Similar methodologies have perhaps been applied by other inferential packages, but the advantage of GCC is a quick detection of the TBP curve during crude switches.

### **GCC handling of crude switches (slide 9)**

Once the crude TBP is known, slide 9 illustrates how easy it is to control the products on specification. The yield pattern required to keep product cutpoints at target is known, and hence control execution is trivial. Of course what make it not trivial are the other constraints that must simultaneously be respected.

### **Problems that hampered inferential precision (slide 10)**

Upon calibrating GCC we have encountered problems that were tricky to solve. The first one was an erroneous side stream flow measurement. The TBP curve converts flow into boiling temperature, and in this case it converted flow measurement errors into boiling point inference errors. Repeatable flow errors can be cleaned out by calibration, but if the measurement is erratic - no calibration would work. We have identified this problem from mass balances, noticing that when mass balance is inconsistent, and when it is off the inferential accuracy deteriorates.

Puertollano applies a CL program to update inferential biases from comparison of inference versus lab data. Seeing the biases drift with time we delayed commissioning, thinking that there was a bug to be fixed in GCC.

### **GCC handling of crude switch, (slide 11)**

Following repair of the erroneous side draw measurement and debugging of the inferential bias program, slide 11 shows the actual performance of GCC during a crude switch. This trend covers two days in March 2006. The red signal is the slope of the TBP curve, showing a substantial crude switch starting March 11 early morning. The blue lines are inferences: naphtha, kero, diesel and heavy diesel 95% points. Immediately after the switch some control deviations are apparent, and those are the result of other constraints that became active during the switch. The controller had resolved the constraints quickly and then kept all properties within range.

The diamond shaped dots on slide 11 represent lab tests, three tests for most products and five for heavy diesel. The consistent fit of GCC versus lab tests before, during and after the crude switch is clearly demonstrated.

### **Diesel quality improvement (slides 12)**

Slide 12 trends one significant key performance indicators for the CDU: heavy diesel 95% Point. Before implementation of our controller, errors of 30, or even 50°C were not uncommon, whereas after commissioning the quality control precision has shrunk ten fold.

### **Problems and solutions (slide 13)**

The recently revamped CDU had many process and control problems, the most significant ones were:

- Instrumentation Issues
- Desalter Instability
- Side Stripper Problems
- Pumparound Instability
- Naphtha Stabilizer Loading
- Naphtha Splitter
- Plus frequent feed quality changes

Although finding solutions to all the problems were important, time does not permit addressing all of them. We will concentrate on two issues to illustrate the nature of our problems: pumparound stability issue and naphtha stabilizer loading.

### **The pumparound instability problem (slide 14)**

The newly revamped CDU suffered from pumparound instability problems, frequently tripping TPA and MPA pumps. Top temperature control, moving either the external reflux or TPA was erratic, trying to keep up with the trips. With heavy crudes the required pumparound heat duty is low and then process stability was even worse.

Two main issues caused this instability:

1. Oversized MPA pump
2. Insufficient liquid level on the MPA draw tray, due to the collector design.

To satisfy minimum flow requirement for the large MPA pump, the MPA flow had to be kept

at a high value. Excessive heat removal in the fractionator middle section would condense too much liquid, and shortage of liquid at the top section would trip the TPA pump. Further, lack of liquid on the MPA draw tray often resulted in MPA pump trip. Light diesel product is taken from the MPA draw, which further complicates the MPA pump suction problem. This instability in the TPA and MPA rendered overhead temperature control difficult.

Once the problems were understood we set out to solve them.

- The Process design group engineered a hot bypass of the MPA that could be quickly implemented. The bypass permitted high MPA flow while keeping the heat duty low, satisfying minimum flow requirement of the oversized pump. The MPA bypass became a manipulated variable in the MVPC controller.
- We added an MPA suction pressure indication as a controlled variable in the MVPC controller to ensure that the MPA pump's net positive suction head (NPSH) was satisfied.
- The TPA was permanently paired with the top temperature controller. External reflux was reduced to minimize long term corrosion.
- A heat integration strategy for the pumparounds was added to the MVPC controller to manage MPA and BPA while respecting the minimum suction pressure constraint, providing room for the TPA to maintain top temperature control and adequate fractionation in the upper section of the column. Crude preheat and stabilizer reboiler requirements are also a part of the pumparound strategy.

### **Trend of cooling duties before and after control (slide 15)**

Slide 15 demonstrates the difference these changes made in stability of the crude unit. Before the MVPC was in service, frequent trips of the pumparounds are evident. These disturbances are noticeable even though this slide uses hourly average data. Large movements of heat duties occur as the top temperature struggled to maintain heat balance. After commissioning there are fewer trips and much improved operation. Also note that MPA and external reflux cooling duties have been decreased while BPA heat removal has been generally increased. This is due to the larger heat consumption of the naphtha stabilizer as loading was increased to process all available naphtha.

### **Naphtha Stabilizer Loading problem (slide 16)**

CDU operators had difficulty keeping LPG C5 content within specification, and stabilized naphtha RVP often was higher than desired. Column flooding was suspected and to cope

with this problem operators bypassed atmospheric naphtha around the naphtha stabilizer and directly to the naphtha splitter. That bypassing has resulted in a loss of LPG recovery.

We investigated stabilizer operation as part of the unit testing for controller dynamic model identification, with the help of extra laboratory analysis to determine product qualities. It quickly became apparent that meeting the C5 LPG specification using the existing base level control scheme was impossible. The base level controls consisted of three PID controllers for:

- reboiler outlet temperature
- feed temperature
- reflux flow.

Using this configuration, the operators could achieve either:

- essentially no C5's in LPG but a very high naphtha RVP or
- RVP within limits but high C5's in LPG.

It was very difficult to operate near the 0.5% C5 in the LPG target with good naphtha RVP result. Upon seeing a high C5 lab result, the operator would typically bypass more naphtha around the stabilizer assuming that it was flooded.

We then modified the base level control configuration from a reboiler outlet temperature controller to a tray 6 temperature controller. Using the control input from a sensible tray permitted operating near the desired LPG specification. Reflux flow is then manipulated to fractionate harder and improve the RVP of the stabilized naphtha. The flooding issue turned out to be a myth.

The MVPC controller was tuned to push fractionation and minimize naphtha RVP while respecting LPG specification. The stabilizer energy consumption is managed to provide adequate heat but not at the expense of middle distillate yield in the main column.

### **Trend of Stabilizer operation before and after control (slide 17)**

Slide 17 demonstrates the difference in operation before and after the controller was put in service. The fraction of naphtha processed is now much higher, while the LPG C5 content is steady near the specification.

### **Controller statistics (slide 18)**

The controller is divided into 4 sub-controllers:

- Preflash and Furnaces
- Atmospheric Column
- Stabilizer
- Splitter.

The sub-controllers allow the independent use of the heater and fractionator control functions when either is unavailable due to instrumentation or maintenance. The controller has 61 controlled variables and 31 manipulated variables among the sub-controllers. Four inferential controlled variables are the key to maintaining product specifications during the frequent crude quality changes. The controller is designed to MAXIMIZE MIDDLE DISTILLATE YIELD. Available degrees of freedom are consumed by the controller to satisfy additional objectives:

- Process available naphtha while maintaining the C5 LPG specification and minimize the light naphtha RVP.
- Feed maximization capability.
- Heat recovery/ Energy conservation.

### **Project Lessons (slide 19)**

Every unit has idiosyncrasies. At Puertollano CDU they came in the form of:

- Real column operational difficulties
- Some bad measurements
- Coding mistakes
- Operational myths.

It is not enough to commission the application and go home. All problems must be addressed before APC starts making money. This must be a team effort among available resources including the Process, Control, Operations departments along with available Consultants. These lessons are not new.

### **Conclusions (slide 20)**

Post project APC audit was carried out by the Puertollano process group by turning the controller off for a week. This exercise has shown an additional refinery profit of 560,000 €/year diesel alone. A comparison of the 8 month period before and 10 month period after



commissioning of the advanced control shows a 2.8% increase in the yield of middle distillate. During this same period the total naphtha processed increased 7.7% to 94.4%.

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# Implementation of APC on Repsol Puertollano CDU1

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**Repsol YPF/ Hall Controls/ Petrocontrol**

# Overview

- **Puertollano Refinery CDU1 Project Summary**
- **Product Quality Measurement Problems and Solutions**
- **Process and Control Problems and Solutions**
- **Project Lessons**
- **Happy Ending**

# Repsol YPF

- Integrated company: upstream, downstream, petrochemicals, gas
- Repsol YPF downstream activities regionally based in Europe and South-America

- **9 refineries operated:**

- 5 refineries in Spain

- **Refining capacity:**

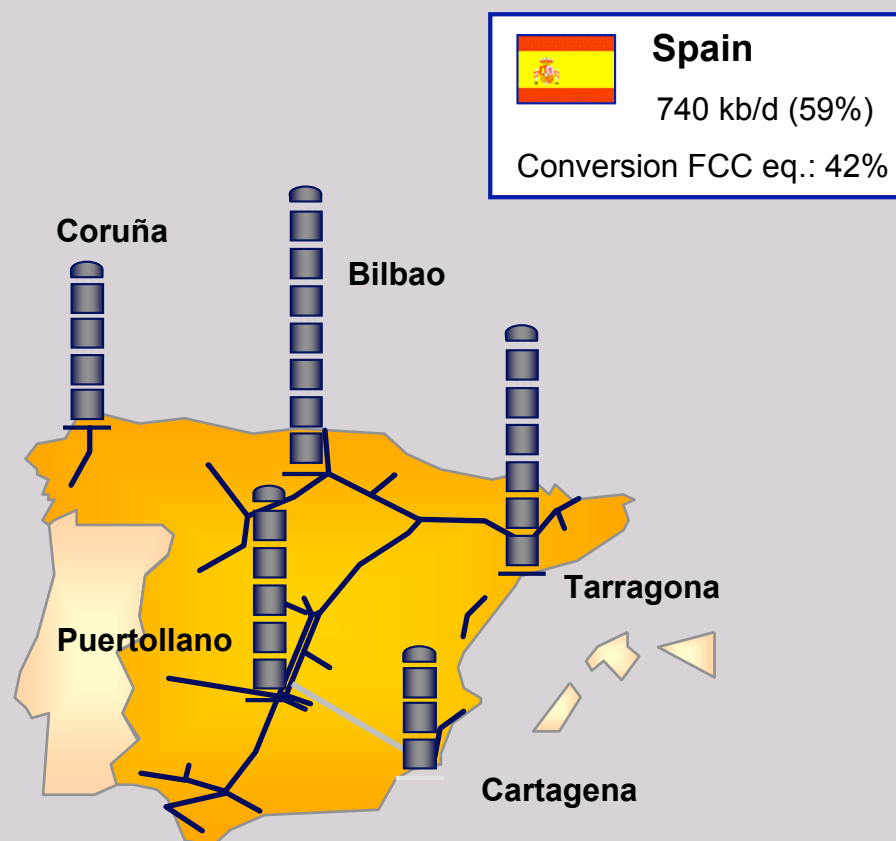
1.2 million bbl/day

- **Puertollano refinery site:**

7.5 M tons/year

Strategic central location

High complexity



# Repsol YPF

- **Advance Control Structure in Spain:**

Technical group in each refinery and a Central Group in Madrid

- **APC situation of Repsol YPF refineries in Spain:**

Implementation of DMCplus® controllers in relevant units

Unit
<b>Crude Distillation Units</b>
<b>Gas Concentration Units</b>
<b>FCC</b>
<b>Vacuum Distillation Units</b>
<b>Coker</b>
<b>Reforming</b>
<b>Hydrocracker</b>
<b>Ethylene plant</b>

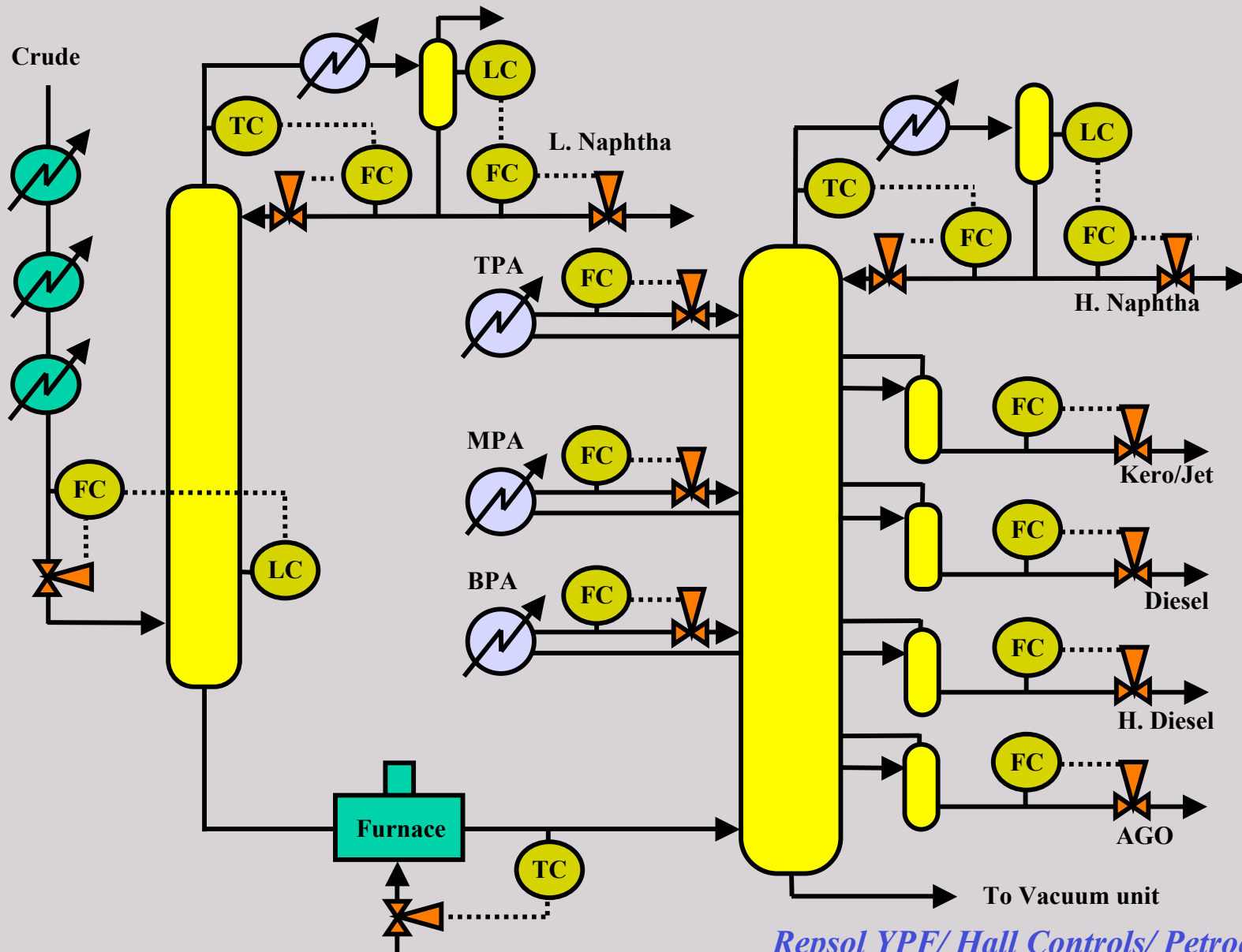
# Project Summary – The Challenge

- **Crude Unit**
  - **360 M3/Hr unit**
  - **10 – 20 crude switches per month (Desire to minimize giveaways/off-spec production during and after crude switches)**
  - **Objective to maximize middle distillate production**
- **Recently revamped unit**
  - **Numerous process problems**
  - **Many control issues**
  - **There was a MVPC installed before**
  - **1 year between the revamping and the new controller**

# Project Summary -The Team

- **Inferentials**
  - **Zak Friedman and engineers from Technology Center in Madrid and Puertollano Refinery**
  - **Petrocontrol selected for inferential model GCC (Generalized cutpoint calculation) known to work during crude switches**
- **Control**
  - **Jim Hall and engineers from Puertollano Refinery and from Central Group**
  - **Hall Controls selected on the basis of prior good experience at Repsol**
- **Multivariable Software**
  - **Aspen Technology's DMCplus®**

# CDU1 simplified structure

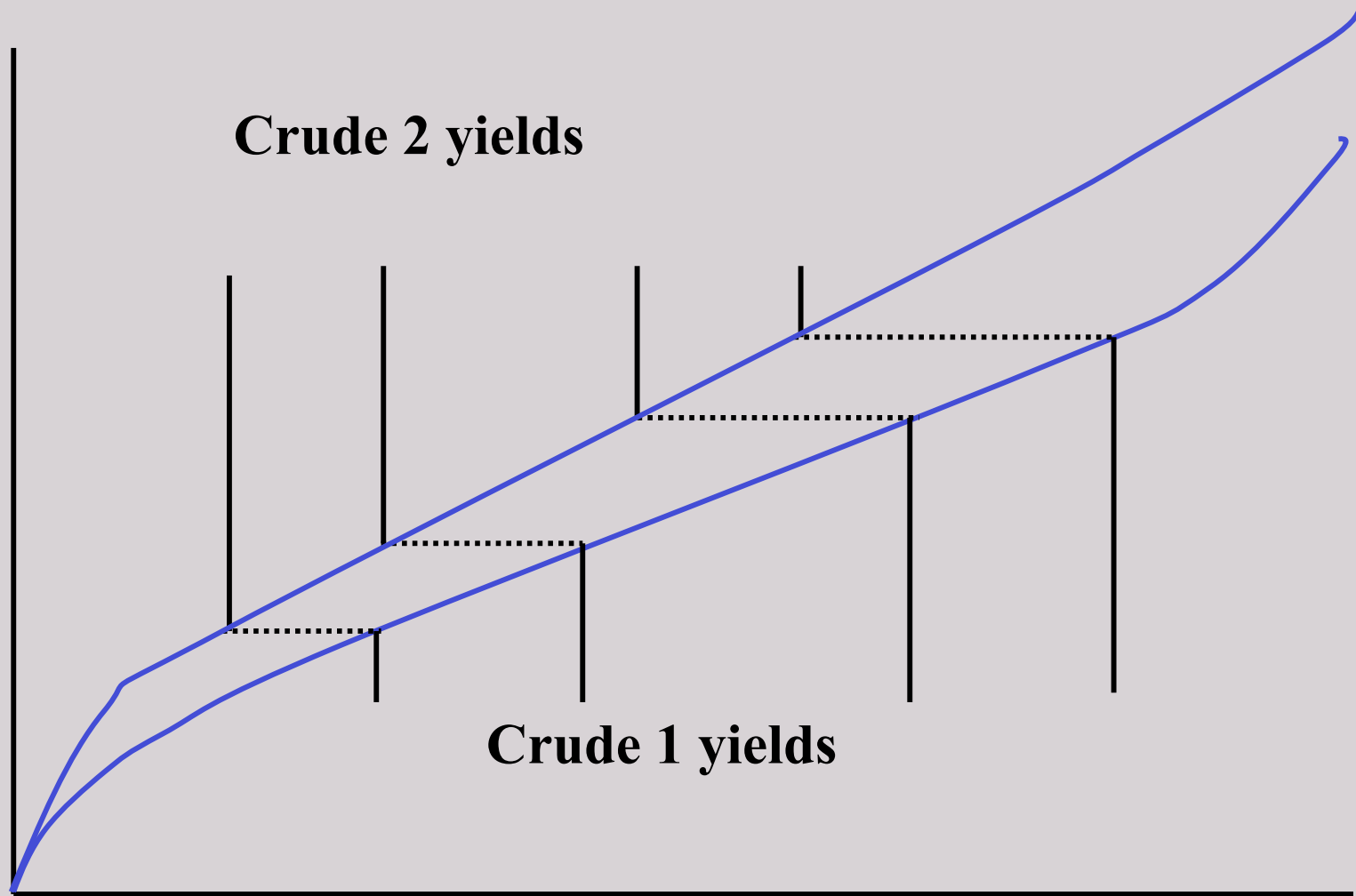




# How GCC works

- Estimate crude TBP curve from unit conditions and heat balance
- Estimates internal reflux profile
- Then estimates product qualities:
  - **ASTMX% = f (cutpoints, internal reflux)**
  - **Flash = f (cutpoints, steam)**
  - **Freeze = f (cutpoints, Watson K factor)**
  - **(We didn't need flash or freeze at Puertollano)**

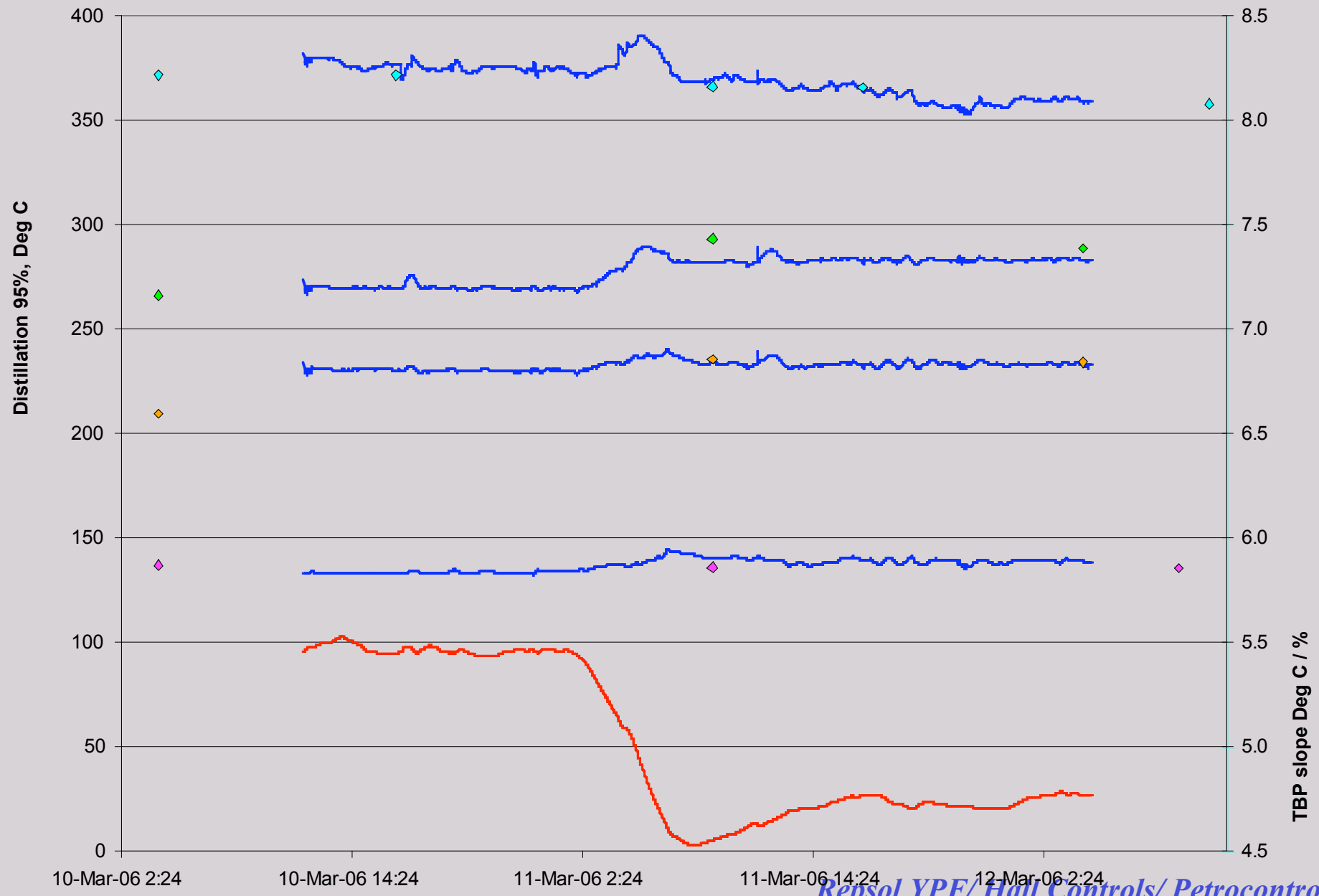
# How GCC handles crude switches



# Inferential Modeling Problems and Solutions

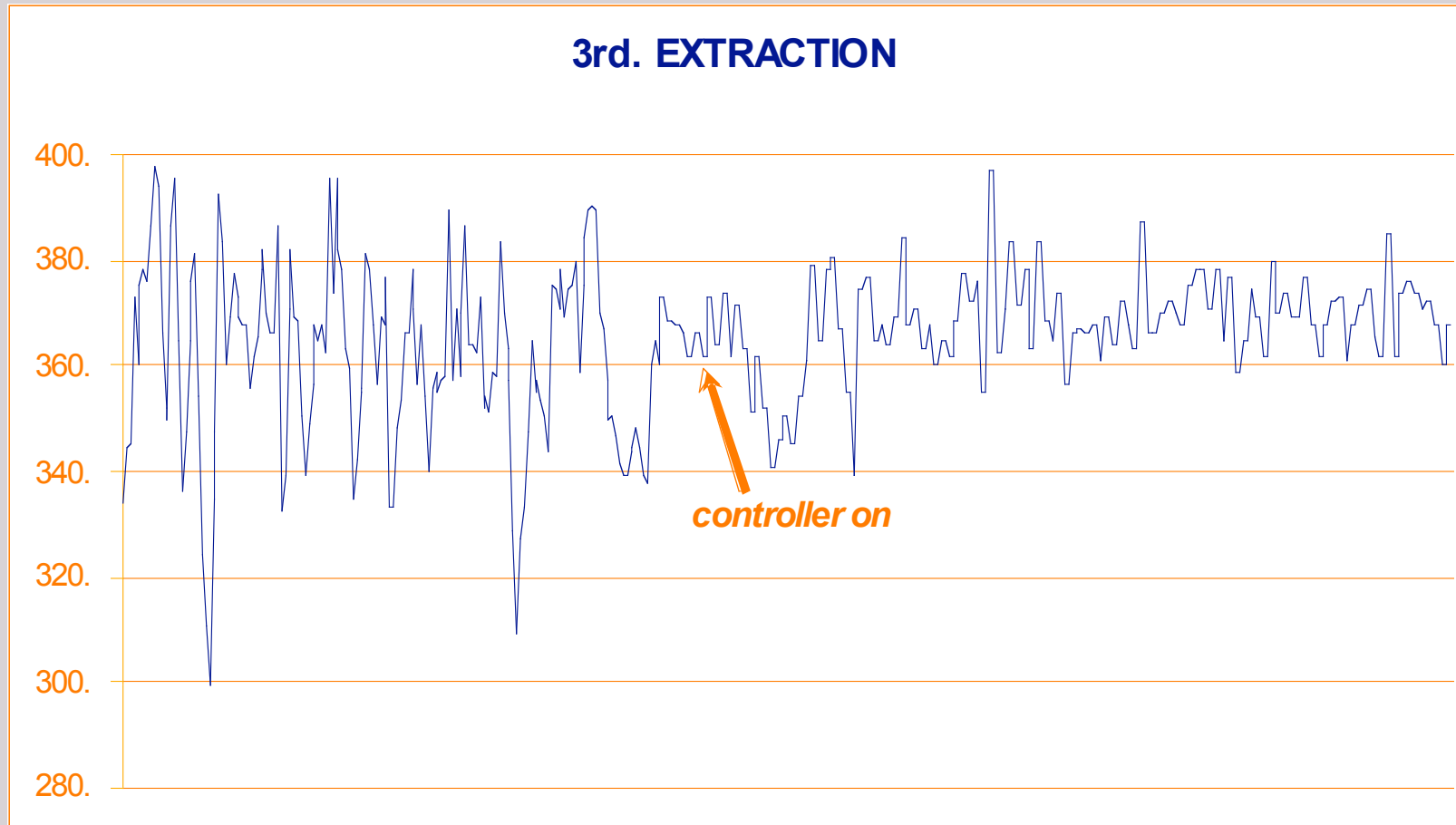
- **Unreliable Instruments**
  - **One of the sidestream flows was erroneous, affecting the GCC model repeatability**
- **Inferential Model**
  - **A bug in the code that determines lab versus inference bias caused slow oscillations**
- **Following slide shows crude switch performance after repairing the problem**

# Predictions before and after switches



*Repsol YPF/ Hall Controls/ Petrocontrol*

# Diesel 95% point before and after control



# Process and Problems and Solutions

- **Many Problems**
  - **Instrumentation Issues**
  - **Desalter Instability**
  - **Side Stripper Problems**
  - **Pumparound Instability**
  - **Naphtha Stabilizer Loading**
  - **Naphtha Splitter**
  - **Plus frequent feed quality changes**

# Pumparound Instability

## • Problem:

- Large MPA pump with insufficient liquid level on draw tray resulted
  - Frequent trips of MPA pump
  - Too much heat removal in MPA caused instability with TPA.

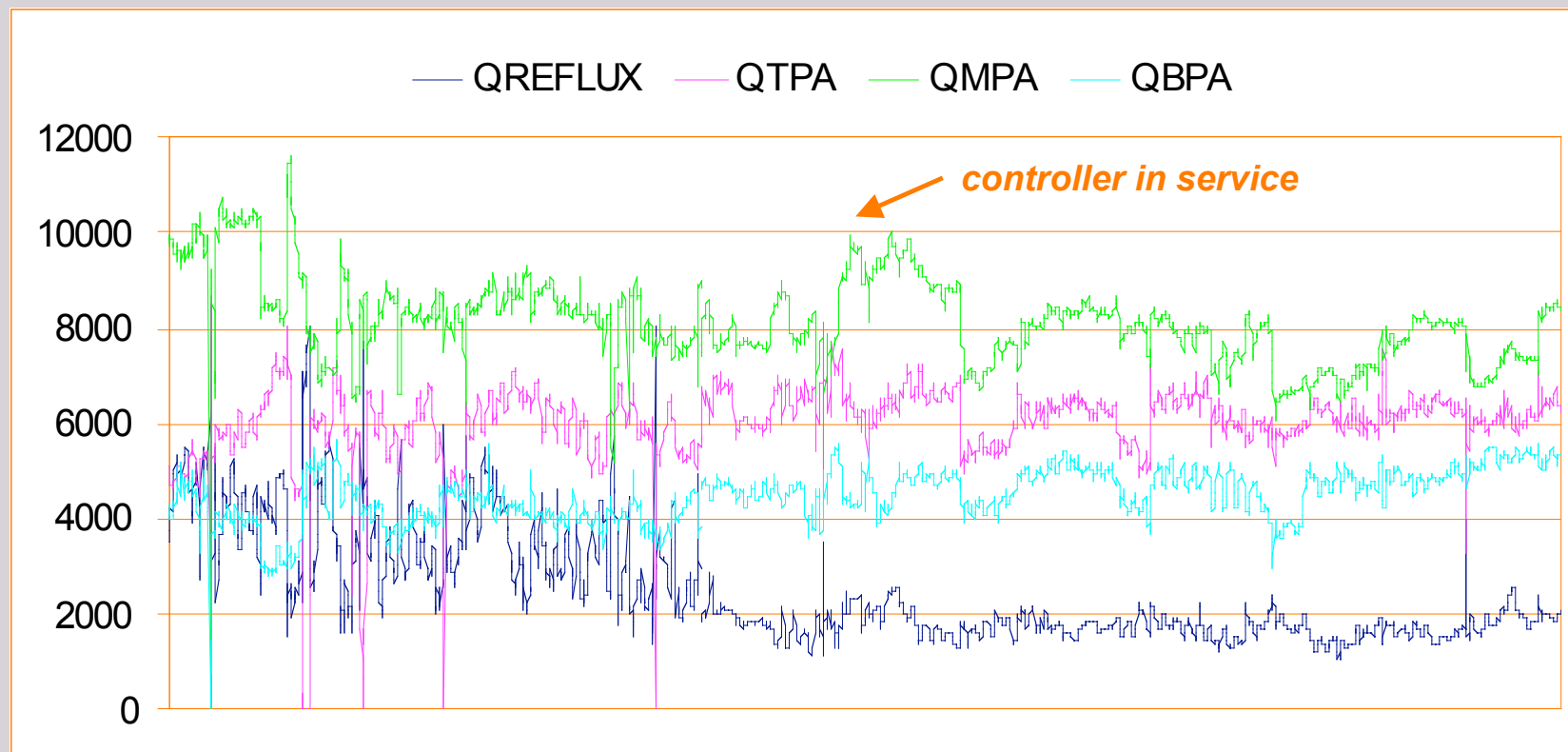
## • Solution:

- Fixes:
  - Implemented bypass of MPA, use as manipulated variable in DMCplus®
  - Implemented MPA pump suction pressure as controlled variable in DMCplus® controller
  - Controller manages MPA and BPA to trade- off crude preheat and maintain fractionation

**Following is a trend of cooling circuit heat duties before and after control implementation.**

**QREFLUX, QTPA, QMPA, QBPA**

# Trend of cooling duties before and after control





# Stabilizer Loading

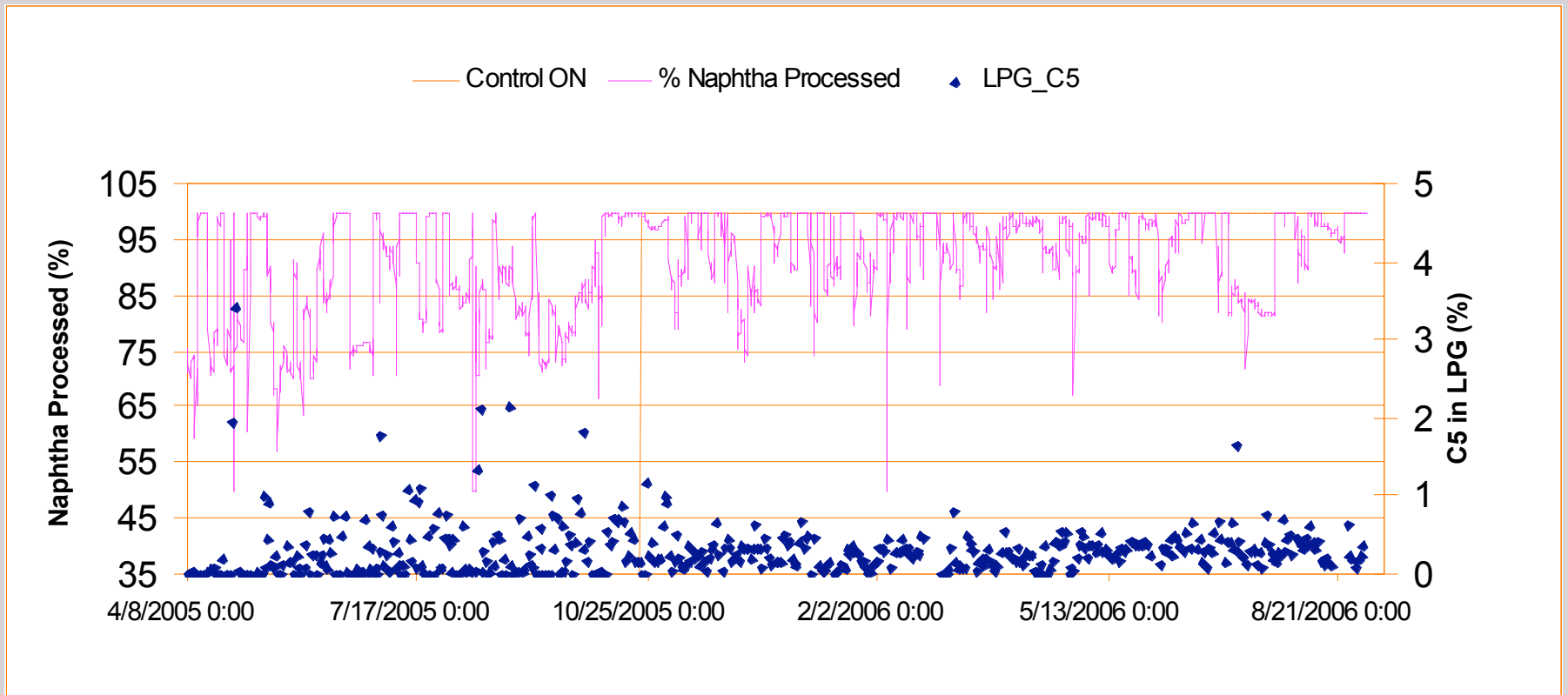
## • Problem:

- Suspected flooding of stabilizer required operators to bypass naphtha around stabilizer
- Control of LPG C5+ content very difficult for operators
- RVP of stabilized naphtha often high

## • Solution:

- During unit step testing found
- Base level control change from reboiler outlet temperature control to Tray 6 temperature control greatly improves operation
  - Simplifies C5+ control for LPG, use of Tray PCT in Controller
  - Inferential prediction for C5+ not required
  - Reflux can then be used as a fractionation handle
- Established proper policy for stabilizer feed temperature
- Found that stabilizer can process almost all of the available nafta. No indication of stabilizer flooding observed.

# Stabilizer before and after control



# DMCplus® Controller Detail

- **Sub-Controllers**
  - Preflash and Furnaces
  - Atmospheric Column
  - Stabilizer
  - Splitter
- **31 Manipulated Variables**
- **61 Control Variables of which**
- **4 Inferential CVs**
  - Naphtha 95%
  - Kero 95%
  - Light diesel 95%
  - Heavy diesel 95%
- **CDU Objective**
  - **MAXIMIZE MIDDLE DISTILLATE YIELD**
  - **Additional objectives include:**
    - **Process available nafta, maintain the C5 LPG specification, and minimize LN RVP.**
    - **Feed maximization capability.**
    - **Heat recovery/ Energy conservation**

# Project Lessons

- **Every unit has idiosyncrasies. At Puertollano they came in the form of**
  - **Real column operational difficulties**
  - **Some bad measurements**
  - **Coding mistakes**
  - **Operational myths**
- **It is not enough to commission the application and go home. All problems must be addressed before APC starts making money.**
- **This must be a team effort**
  - **Process group, Control group, Operations, and Consultants**

# Conclusions – Happy Ending

- **Method 1**

- **Audit done process group: turned the controller off for a week**
- **Additional 560,000 €/year diesel alone**
- **Non-quantified benefits:**
  - **Unit stability during crude changes**
  - **Preventing process upsets (RCM pump discharges, desalter, other).**
  - **Naphtha processing**
  - **Heat recovery**

- **Method 2**

- **Review of 8 months before, 10 months after**

	<b>BEFORE</b>	<b>AFTER</b>
<b>Middle Distillate Yield</b>	<b>39.4%</b>	<b>42.2%</b>
<b>Naphtha Processed</b>	<b>86.7%</b>	<b>94.4%</b>