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MULTIVARIABLE CONTROLLER IMPLEMENTATION FOR A CRUDE UNIT: A CASE STUDY

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1. OVERVIEW

In end-1998, Koa Oil (now a part of Nippon Oil Corporation) embarked on a comprehensive advanced control implementation at its two refineries in Marifu and Osaka. The advanced control implementation involved implementing multivariable controllers and optimizers across all the major process units in both refineries plus a separate oil movements & blending optimization project.
Koa selected Honeywell Hi-Spec Solutions' Profit® Suite for multivariable control. At the completion of the project in mid-2001, a total of 22 Profit® Controllers (Formerly Robust Multivariable Predictive Control Technology - RMPCT controllers) and 2 Profit Optimizers® were implemented at the two refineries. The estimated benefit from all these applications is estimated as about 4 million dollars per year.

This paper specifically deals with the multivariable controller implementation at the #4 crude unit at Marifu refinery and describes the major highlights of this application.

1.1 Crude unit at Marifu refinery

The Marifu Refinery #4 CDU operates at a nominal feed capacity of 127,000 BPSD. The unit is divided into a crude section and a gas plant section (Fig – A). The crude section is made up of two preheat trains on both sides of a desalter, a prefractionator, a crude heater, and a 52 tray crude fractionator, The gas plant section is made up of a 35 tray debutanizer column and a 40 tray depropanizer column. The products from the prefractionator and fractionator are light and heavy naphtha streams (LSR and HSR) that are combined to form a wide-cut naphtha (WSR) product, a wide-cut naphtha stream to CCR, three kerosene streams (light kero, 2nd kero, and 3rd kero), gasoil (GO) and topped crude (TC). The overheads of the prefractionator and fractionator go to the Gas Plant where they are separated into butane and propane.

The crude unit operation is complex in two respects. First there are many operational modes calling for entirely different operating conditions. Second the unit undergoes severe crude switches.

To illustrate the mode complexities, the three kerosene and gasoil sidestreams are combined in various ways. There are different property specifications for the product streams in each of the operating modes.

Following are the most common crude types and operating modes.

Table 1 Major crudes processed and operating modes

Crude Processed	Operating Modes
Middle East (ME)	ME-#1 Kero
	ME – JP
	ME – Jet
Ratawi (RW)	RW - #1 Kero
	RW – JP
	RW – Jet
Sumatran Light (SL)	SL – Molex

The differences between the operating modes are due the way the side-streams are blended together. The following table illustrates sidestream combinations during the different modes.

Table 2 Sidestream product blending

Operating Mode				
Sidestream	**ME/RW #1 Kero**	**ME/RW #1 Jet**	**ME/RW JP**	**SL Molex**
Lt. Kero (LK)	#1 Kero	#1 Kero	#1 Kero	BGO
2nd Kero (2K)	#1 Kero	#1 Kero	#1 Kero	BGO
3rd Kero (3K)	#1 Kero	#1 Kero	JP	Molex
GasOil (GO)	GasOil	GasOil	GasOil	BGO

The crude unit processes a number of crude slates with crude switches occurring every 2 to 5 days. Some of those switches are so severe to the point that without APC it takes a long time to line out the unit and occasionally this leads to incidents such as pump cavitation, off specification products and furnace overheating.

1.2 Multivariable controller objectives

The following broad controller objectives were decided during the functional design stage.

- Meet the unit feed target. (The controller never needs to maximize the throughput).
- Maximize the production of most valuable products by minimizing product quality giveaways.
- Maximize operations efficiency e.g. by maximizing COT, and minimizing Crude feed preheat bypass flows, etc.
- Honor various process constraints.
- Optimize crude switch handling by
    - Handling process constraints during Crude Switches
    - Automating the change of quality targets and optimization objectives
    - Minimizing Crude Switching time

Crude switch optimization was expected to be one of the major benefits of implementing Profit Controller.

2. DESIGN CONSIDERATIONS

There were several major design considerations throughout the course of implementing the project. Some of these are explained below.

2.1 Non-linear crude TBP model

A high fidelity crude TBP model was thought to be essential both for developing accurate inferentials and for handling crude switches. Petrocontrol's GCC (Generalized Cutpoint Control) model (Ref-1) was chosen for this purpose.

The GCC model is based on first principle calculations such as heat balance and TBP/ EFV conversions. It gives fast, accurate results and the models do not require much tuning.
The model first calculates naphtha cutpoint for the fractionator based on prefractionator overhead flow, fractionator top temperature and partial pressure correction. Next, flash zone vapor flow is calculated from heat balance across the column. This is followed by flash zone cutpoint calculation based on flash zone vapor flow, flash zone temperature, pressure and steam flow.

The GCC model initially assumes a linear TBP curve between Naphtha and Flash Zone cutpoints. The non-linearities are then estimated from the column temperature profiles. Finally, the nonlinear TBP curve is used to estimate the sidestream cutpoints. Fig – B shows a plot of the sidestream flows vs. the linear and non-linear TBP cutpoints.

During a crude switch, the column can be out of mass balance for a long time but it cannot be out of heat balance. Since this method is based on heat balance, it gives fairly accurate results for the cutpoint predictions during crude switches.

The GCC model replaced an old crude TBP model that was in use for the unit. The old model had two inaccuracies

(a) It ignored the prefractionator completely and considered the fractionator feed as the crude. This led to errors when the prefractionator column operation was changed.

(b) It did not consider TBP slope non-linearity. The TBP curve for the SL crude was found to be significantly non-linear.

2.2 Inferential CV's

A total of 15 inferentials for product properties were developed using the GCC model. These inferential values were inputted to Profit Controller as CV's. Overall, the inferential models were excellent as demonstrated in figures C through I and discussed below.

Figures C through I show the comparison of the inferentials with the lab measurements for JP End Point, Kero 95%, Kero Freeze Point, JP Flash Point, BGO CFPP, Molex 10% and Kero Flash Point respectively. The predictions even from difficult inferences like FlashPoint, CFPP and Freeze Point were quite good. **<u>This is demonstrated by the fact that in almost all the cases no laboratory or analyzer bias update was needed even though all the inferentials were used for online control.</u>**

All of the inferential properties except for the cold properties (e.g. freeze point) were based on sidestream cutpoints. The cold properties required additional density measurement through Watson K factor.

The inferentials were directly used as Profit Controller control variables (CV's) and were developed for the product sidestreams for each of the crude types/ operating modes. During a crude switch, these CVs were dropped or brought in as per the product specifications as shown in the following table.

Table 3 Inferential CVs used in various operating modes

CV	SL Molex	ME/Rw #1Kero	ME/Rw Jet	ME/Rw JP
#1 Kero Flash		✓	✓	
#1 Kero 95%		✓	✓	
#1 Kero Freeze Pt			✓	
#1 Kero Smoke			✓	
GasOil 95%		✓	✓	✓
GasOil CFPP		✓	✓	✓
JP Flash Pt				✓
JP Freeze Pt				✓
JP 50%				✓
JP EndPoint	✓			✓
#1 Kero Flash (JP)				✓
Molex 10%	✓			
BGO Flash	✓			
BGO CFPP	✓			

2.3 Decision regarding size/ scope of controller

Initially it was thought that the CDU crude section and the gas plant would reside in the same Profit Controller. During detailed design this was changed to two controllers, one for the CDU crude section and the other for the gas plant. This design change was done for a number of reasons

- A smaller controller is easier to operate and maintain when compared to a larger controller. The CDU section controller itself was quite large (24MV's by 58 CV's).

- If a single controller was installed, it would require a model between CDU feed rate and debutanizer/ depropanizer constraints. After step testing, it was realized that for such a controller, some of the models would have a setting time of almost 5 hours. For the crude section the shortest responses were around 15 minutes. Since having a combination of long and short settling times can reduce the effectiveness of the control solution, it was decided to split the controllers.

The interactions between the CDU controller and the Gas Plant controller were handled by incorporating crude unit CVs in the gas plant controller and gas plant CVs in the crude unit controller.

2.4 Quality estimation during crude Switches

The GCC model is best known for its ability to continually estimate the slope of the crude TBP curve. During a crude switch, as soon as new crude enters the crude column the heat balance is affected (e.g. for a more volatile crude the cooling load goes up to cope with more vapor) and hence the TBP slope changes.

As the calculated TBP slope changes, the calculated TBP cutpoints and the inferred product quality CVs also change. Since the responses between the sidestream product flow MVs and the inferred product quality CVs are fast, the multivariable controller is able to maintain the CVs at target values during the crude switch.

The timing of the TBP slope update and the resulting multivariable controller manipulations is exactly right. There is no need to lead or lag the calculation as the change in heat balance occurs exactly in phase with the change in vapor flow in the column. As the crude switch progresses and the crude TBP slope transitions to a new value, the multivariable controller continues to simultaneously manipulate the sidestream product flows (on a minute-by-minute basis) to maintain product qualities.

2.5 Crude Switch handling

Having achieved a good estimation of product properties through crude switches, we were faced with the challenge of moving the unit fast enough to control those properties while at the same time avoiding violation of unit constraints. Large multivariable controllers are generally not tuned aggressively enough to move the unit quickly and we needed to implement that ability by adding disturbance variables (DV's) as well as gain scheduling. Additionally we wanted to incorporate features for smoothly moving the unit through mode switches during the crude switches.

The controller was designed for the following crude switches.

Table 4 Typical crude switch scenarios

SN	CRUDE SWITCH FROM	TO
1	ME/Rw #1Kero	SL Molex
2	ME/Rw Jet	SL Molex
3	ME/Rw JP	SL Molex
4	SL Molex	ME/Rw #1Kero
5	SL Molex	ME/Rw Jet
6	SL Molex	ME/Rw JP
7	ME/Rw #1Kero	ME/Rw JP
8	ME/Rw Jet	ME/Rw JP
9	ME/Rw JP	ME/Rw #1Kero
10	ME/Rw JP	ME/Rw Jet

A crude switching program was developed and implemented in DCS. This program works in conjunction with the multivariable controller and automates almost all of the operator actions that would otherwise be required, specifically, the program …..

- Turns certain CVs ON and OFF depending on the crude switch type.
- Changes optimization coefficients
- Changes the high and low limits of certain CV's/ MV's.
- Changes product rundown valves
- Changes analyzer sampling points

The operator only has to select the type of Crude Switch, set the various targets (e.g. final feed rate, etc.) and start the crude-switching program.

2.6 Crude Switch DV

The ME → SL & SL → ME crude switches were identified as being particularly severe. It was decided to use a crude switch DV in such cases. (The ME → RW & RW → ME switches were not severe and the built-in robustness of Profit Controller was found to be sufficient to handle them.)

This crude switch DV was triggered automatically when the crude preheat temperature (6TI31) crossed a certain temperature (45 degC). Since the heavy crudes were hotter, crossing the 45 degC line indicated moving to (or from) a heavy crude.

The models for the DV were obtained by regressing the data over past several crude switches. Some of the DV models were with flow CVs e.g. crude unit feed (CV03) and debutanizer feed (CV58). These flows undergo a significant change during crude switch due to differences in crude characteristics. Other models for the DV were with desalter inlet temperature (CV04) and desalter pressure control output (CV05). These variables were primarily affected due to the differences in the crude feed temperatures.

The crude switch DV could be used successfully because the changes during crude switches were significant and predictable and also because there was a good indicator (preheat temperature) for detecting the crude switch event.

2.7 Gain Scheduling

Gain scheduling had to be introduced in a number of cases to account for the non-linearity of the models as shown in the following table. A DCS program based on the current values of respective CVs and MVs implemented the automatic gain scheduling.

Table 5 Gain Scheduling implemented between CVs and MV's

SN	CV	MV's	REASON
1	CV54 (fractionator GasOil ΔP)	MV04 & MV05 (furnace COT's), MV10 (fractionator ovhd temp), MV11 to MV15 (side-stream flows) and MV18 (fractionator steam)	When close to flooding limit, the gains become very high
2	CV52 (gas oil color)	MV01 (furnace feed), MV10 (fractionator ovhd temp) and MV11 to MV15 (side-stream flows)	Highly non-linear when close to limit.
3:	CV24 (FT receiver pressure)	MV22 (FT receiver pres. controller output)	Non-linear because the pressure control is split-range.
4	CV55 (overflash flow)	MV11 to MV15 (sidestream flows)	To make the control fast when overflash flow is close to low limit.
5	CV02 (debutanizer ovhd C5 analyzer)	MV02 (debutanizer top temperature) and MV03 (debutanizer duty)	The analyzer is highly non-linear.

6	CV04 (desalter temperature)	MV06 (crude heat exchanger bypass flow)	Gain could be changed by the balance of heat and hydraulics (it often becomes inverse)
7	CV08 (crude heat exchanger bypass valve OP)	MV02 (desalter pressure) MV06 (crude heat exchanger bypass flow) MV07 (crude heat exchanger pressure)	Gain could be changed by the balance of heat and hydraulics (it often becomes inverse)

3. DETAILED CONTROLLER DESIGN AND TYPICAL OPERATIONS

3.1 List of MV's, CVs and DVs

The full list of MV's, CVs and DVs for both the controllers is described below. There were a number of changes to this list over the course of the project right upto the final commissioning.

Table 6 CDU Profit Controller list of MV's, CVs and DV's

Type	Tagname	Description
Manipulated Variables		
MV01	6X7FY10.SP	Crude Heater Total Flow Controller
MV02	6PC1.SP	Desalter Pressure
MV03	6TC1.SP	Prefractionator Top Temperature
MV04	6X7TC21W.SP	Crude Heater W cell COT
MV05	6X7TC21E.SP	Crude Heater E cell COT
MV06	6FC40.SP	Crude Preheat Bypass Flow
MV07	6PC32.SP	Crude Feed Pressure
MV08	6FC105.SP	Furnace air flow
MV09	6FC23.SP	FT LSR Return Flow
MV10	6TC5.SP	FT Top Temperature
MV11	6FC12.SP	HSR Product Flow
MV12	6DMC14.SP	Light Kerosene Product Flow
MV13	6FC14.SP	2nd Kerosene Product Flow
MV14	6FC15.SP	3rd Kerosene Product Flow
MV15	6FC16.SP	GO Product Flow
MV16	6FC17.SP	GO Reflux Flow
MV17	6FC13.SP	HSR Reflux Flow
MV18	6FC109.SP	FT Bottoms Steam Flow
MV19	6X4ZY41.SP	Light Kerosene Stripping Steam Ratio
MV20	6X4ZY43.SP	3rd Kerosene Stripping Steam Ratio
MV21	6FC111.SP	Prefractionator Bottoms Steam
MV22	6PC3.OP	FT Column Pressure
MV23	6FC22.SP	FC LSR Return Flow
MV24	6TC205.OP	Air Preheater Bypass Valve

Table 6 CDU Profit Controller list of MV's, CVs and DV's

Type	Tagname	Description
Controlled Variables		
CV01	6XF1MAX.PV	Crude feed pumps constraint
CV02	6PI31.PV	Crude Feed Pump Suction Pressure
CV03	6FC1M.PV	Total Crude Feed Compensated Flow
CV04	6TI39.PV	Desalter Crude Inlet Temperature
CV05	6XP1A.PV	Desalter PC Output (transformed)
CV06	6XPDFC.PV	Prefractionator Column ΔP
CV07	6XFCCP.PV	Prefractionator Ovhd Cutpoint
CV08	6PC32.OP	Feed pressure controller output
CV09	6FC2.OP	Prefractionator Ovhd Flow Controller Output
CV10	6FC38.PV	Prefractionator Reflux Flow
CV11	6TI50.PV	Prefractionator Ovhd Seawater Condenser Inlet Temp.
CV12	6TC205B.PV	Air Preheater Dewpoint Temperature
CV13	6FC105.OP	Air Flow Controller Output
CV14	6AO2.PV	Crude Heater Stack O_2
CV15	6XTSPOW.PV	Crude Heater Pass Outlet Temperature Spread – Cell 2 (West) (Max. POT - Min. POT)
CV16	6XTSPOE.PV	Crude Heater Pass Outlet Temperature Spread – Cell 1 (East) (Max. POT - Min. POT)
CV17	6XTMHNGW.PV	Max. Crude Heater (Cell 2 – West) Hanger Temperature
CV18	6XTMHNGE.PV	Max. Crude Heater (Cell 1 – East) Hanger Temperature
CV19	6XTMWALW.PV	Max. Crude Heater (Cell 2 – West) Wall Temperature
CV20	6XTMWALE.PV	Max. Crude Heater (Cell 1 – East) Wall Temperature
CV21	6PC103A.PV	Crude Heater Draft Pressure
CV22	6PDI136.PV	Air Preheater ΔP
CV23	6PC107.OP	FT Blower Suction Pressure Controller Output
CV24	6PC3.PV	FT Pressure
CV25	6PI37.PV	FT Overheads Receiver Pressure
CV26	6FI84.PV	FT MSR Settler Water Flow
CV27	6XTDFTT2.PV	FT Tray 2 Temp-Dewpoint ΔT
CV28	6XTDFTT.PV	FT Dewpoint-Top Temp ΔT
CV29	6X1KYD.PV	#1 Kero Yield (ME/#1 Kero mode, ME/JP mode)
CV30	6FC20.PV	FT Topped Crude Rundown flow
CV31	6FC20.OP	FT Topped Crude Rundown FC Output
CV32	6XFMN3OP.PV	Min. Crude Heater Pass Flow Valve Position
CV33	6TI80.PV	FT Ovhds Seawater Condenser Inlet Temp.
CV34	6FC54.OP	FT Reflux Flow Controller Output
CV35	6FC11.PV	FT Overheads Product Flow
CV36	6FC11.OP	FT Ovhds Flow Controller Output
CV37	6XWSR90.PV	WSR 90% Point

Table 6 CDU Profit Controller list of MV's, CVs and DV's

Type	Tagname	Description
CV38	6X1KFLP.PV	#1 Kero Flash Point (ME/#1 Kero mode)
CV39	6X1K95C.PV	#1 Kero 95% Point (ME/#1 Kero mode)
CV40	6X1KFP.PV	#1 Kero Freeze Point (ME/#1 Kero mode)
CV41	6X1KSM.PV	#1 Kero Smoke Point (ME/#1 Kero mode)
CV42	6XGO95.PV	GasOil 95% Point (ME/#1 Kero mode, ME/JP mode)
CV43	6XGOCFPC.PV	GasOil CFPP (ME/#1 Kero mode, ME/JP mode)
CV44	6XJPFLP.PV	JP Flash Point (ME/JP mode)
CV45	6XJPFP.PV	JP Freeze Point (ME/JP mode)
CV46	6XJP50.PV	JP 50% Point (ME/JP mode)
CV47	6XJPEP.PV	JP End Point (ME/JP mode)
CV48	6XKFLP2.PV	#1 Kero Flash Point (ME/JP mode)
CV49	6XMOL10C.PV	Molex 10% Point (SL/Molex mode)
CV50	6XBGOFLC.PV	BGO Flash Point (SL/Molex mode)
CV51	6XBGOCFC.PV	BGO CFPP (SL/Molex mode)
CV52	6AXCL1V.PV	GasOil color analyzer
CV53	6XPDFT.PV	FT Column ΔP
CV54	6PD48.PV	FT GasOil Section ΔP
CV55	6XFOF.PV	FT Overflash Flow from GCC Model
CV56	6XAMHSRR.PV	HSR Reflux Pump Amperage
CV57	6X7TD21.PV	Delta temperature between crude heater cells
CV58	6X5FX01.PV	Debutanizer column feed rate
Disturbance Variables		
DV01	6XMESLSW.PV	ME-SL Crude switch

Remarks

1. The existing total unit feed, 6FC1M proved to be a very noisy signal. Instead, a new tag based on Prefractionator overhead flow and furnace feed was used.

2. The desalter pressure controller output was transformed using the equal percent type valve equation.

3. Two additional constraints for the controller were the side stripper levels for the 2^{nd} and 3^{rd} kero streams. Due to the fast responses of the levels in the side strippers, this control was provided outside Profit Controller. For this purpose, a custom DCS code was implemented that reduces the sidestream flow if the level goes too low.

Table 7 Gas Plant Profit Controller list of MV's, CVs and DV's

Type	Tagname	Description
Manipulated Variables		
MV01	6PC4.SP	DB Pressure
MV02	6XT126A.SP	DB Top Temperature
MV03	6X5FC11.SP	DB Reboiler Duty
MV04	6PC5.SP	DP Pressure
MV05	6XT9A.SP	DP Top Temperature
MV06	6X6FC02.SP	DP Reboiler Duty
MV07	6FC26.SP	C3 Recirculation Flow
MV08	6FC27.SP	C4 Recirculation Flow
Controlled Variables		
CV01	6XDBBC4V.PV	DB Bottoms %C4
CV02	6XA14C.PV	DB Overheads %C5 (analyzer)
CV03	6XDPBC3V.PV	DP Bottoms %C3
CV04	6XDPOC4V.PV	DP Overheads %C4
CV05	6PC4.OP	DB Pressure Controller Output
CV06	6DB1_LL.PV	DB Tray #1 Calculated Liquid Load
CV07	6XDB35.PV	DB Tray #35 Calculated Liquid Load
CV08	6LC15.OP	DB Bottoms LC Output
CV09	6PC5.OP	DP Pressure Controller Output
CV10	6FI60.PV	LSR Product Flow To Storage
CV11	6FI62.PV	C3 Product Flow To Storage
CV12	6FI64.PV	C4 Product Flow To Storage
CV13	6FC35.PV	DB Overheads Product Flow
CV14	6XFD26.PV	Difference between C3 and C4 recycles flows
Disturbance Variables		
DV01	6XF101.PV	DB feed rate
DV02	6XFFCLPG.PV	GCC Model Predicted LPG Flow

Remarks

1. The old top temperature controllers for debutanizer and depropanizer were replaced by pressure compensated temperature controllers and used as Profit Controller MV's.

3.2 Optimization Strategy

Optimization strategy was decided based on discussions with operations, planning and unit engineers. For the multivariable controller, this optimization strategy was achieved by configuring cost coefficients on the CVs and MV's. Some of these cost coefficients change depending on the type of crude and operating mode.
The numerical value of cost coefficients was adjusted based on trial and error so as to achieve the desired objective.

CDU Profit Controller

Maximize CDU feed if desired by Planner
Minimize CDU Feed Bypasses (in order to maximize the heat exchange between the colder crude unit feed and the hotter flow streams such as fractionator bottoms flow and side-stream flows.)
Maximize Prefractionator Ovhd Flow
Maximize Prefractionator Bottom Steam
Maximize Furnace COT
Minimize Fractionator Pressure
Minimize flaring from FT Blower (PC107)
Maximize Fractionator Bottom Steam
Maximize HSR Reflux
Minimize Light Kero
Maximize GasOil
Minimize Topped Crude
Minimize LK Stripping Steam
Minimize 3rd Kero Stripping Steam
Minimize GO Stripping Steam
Minimize FC LSR Return
Minimize FT LSR Return

For the following MV's, the optimization strategy was different for different operating modes.

Table 8 Optimization strategy for different operating modes

CV	SL Molex	ME/#1Kero	ME/Rw Jet	ME/Rw JP
GasOil Reflux	Min	Min	Max	Min
2nd Kero Flow	Min	Max	Max	Min
3rd Kero Flow	Max	Max	Max	Max

GAS PLANT Profit Controller

Minimize DB, DP Pressure
Min DP Duty
Minimize DB Duty
Minimize C3 Circulation
Minimize C4 Circulation

3.3 Typical operating scenario

Even though several models exist between all the MVs and CV's, most of the times MVs move because of only a small set of CV's. In fact the whole controller can be broken down into several small sets of CVs and MVs which act like sub-controllers for normal operation (e.g. 95% of the time).
During occasional disturbances or unusual operations, the other models (CV-MV relationships) come into play.
This kind of a subdivision of the controller is useful for explaining and understanding the controller behavior.

Table 9 Groups of related CVs and MVs for typical operation

DESCRIPTION	MV	CV
CDU Profit Controller		
Crude feed control	MV01: Feed	CV03: CDU Feed
Crude Furnace control	MV04: COT-E MV05: COT-W MV08: Air	CV14: Furnace O2 CV21: Draft CV32: Furnace Min O/P CV57: COT E-W
Preheat Section Control	MV02: Desalter P MV03: FC Top T MV06: FC40 MV07: PC32 MV23: LSR Ret to FC	CV04: Desalter T CV05: PC1.OP (transformed variable) CV08: PC32.OP CV09: FC2.OP CV11: FC seawater
Fractionator Pressure Control	MV22: PC3.OP	CV23: 6PC107.OP CV24: 6PC3 CV25: 6PI37
Fractionator Top Temperature Control	MV10: TC5	CV28: FT Dew Point CV58: DB Feed CV33: FT Seawater
HSR Reflux Control	MV17: HSR Ref	CV56: HSR Reflux Amps
Side-streams Flow Control	MV11: HSR MV12: Light Kero MV13: 2nd Kero MV14: 3rd Kero MV15: GasOil	CV37: WSR 90% CV38: Kero Flash - ME#1 Kero CV39: Kero 95% CV40: Kero Freeze CV41: Kero Smoke CV42: Gas Oil 95% CV43: Gas Oil CFPP CV44: JP Flash CV45: JP Freeze CV46: JP 50% CV47: JP/ Molex Endpoint CV48: Kero Flash - ME/JP CV49: Molex 10% CV50: BGO Flash CV51: BGO CFPP CV52: GasOil/ BGO Color CV54: Gas Oil delta P CV55: Overflash
Gas plant Profit Controller		
Debutanizer column control	MV02: DB Top Temp MV03: DB Duty	CV01: DB Bottom C4 CV02: DB Ovhd C5 CV06: DB Tray -1 Load CV07: DB Tray -35 Load
Depropanizer column control	MV05: DP Top Temp MV06: DP Duty	CV03: DP Bottom C3 CV04: DP ovhd C4
Debutanizer overhead flow control	MV07: C3 Recycle MV08: C4 Recycle	CV13: DB ovhd Flow CV14: C3 - C4 Recycle flow

4. BENEFITS ACHIEVED

A benefit study was carried out after several months of stable operation following the controller commissioning. These are described below.

4.1 Increase in absorbed duty of CDU Furnace

After Profit Controller, absorbed duty for CDU Furnace increased due to following reasons (Fig - J):

(a) Ability to operate at higher COT by tighter control of O2 (CV) and optimum airflow (MV) manipulation.

(b) Ability to operate at higher COT by pushing against the pass flow minimum O/P constraint.

(c) Lower column pressure

Higher absorbed duty leads to better fractionation, which improves gasoil yield. This benefit was quantified as about 400 kUS$ per year.

4.2 CDU Fractionator pressure minimization

As a result of Profit Controller implementation, CDU fractionator could be operated at a lower pressure (Fig – K). This was due to the following reasons.

(a) Use of a better control handle. Profit Controller manipulated the pressure controller output (6PC3.OP) instead of 6PC3.SP. Earlier, the PID control of pressure was poor because of non-linear behavior due to split-range operation. The control was significantly improved (see Fig – K) as a result of choosing a direct control handle (pressure valve output) and implementing gain-scheduling to take care of the non-linearity.

(b) Better constraint control. Pressure was minimized while satisfying the constraints for following 3 CV's:

- Fractionator pressure 6PC3.PV
- Receiver pressure 6PI37.PV
- Blower suction pressure output 6PC107.OP

Earlier the pressure had to be conservatively operated higher because the operator had to handle the constraints manually.
As a result of lower pressure operation, gasoil recovery improved due to better fractionation. This benefit was quantified as about 175 kUS$.

4.3 Better Overflash Control

Profit Controller could accomplish better overflash flow control than before due to two reasons:

- Overflash flow being a CV could be controlled directly
- Overflash CV was based on a combination of flowmeter and the calculated flow from GCC model. (This was helpful because the overflash flowmeter reading often went out of range).

This helped in maximizing gasoil flow sometimes during JP Operation. This benefit was not quantified.

4.4 Gas plant operation flexibility

For Gas Plant operation with Profit Controller, the operator now has more flexibility than before to regulate debutanizer bottoms %C4 based on requirement. For example, the operator now can either

- Minimize %C4 in Debutanizer Bottoms by operating close to flooding (this was not possible before) OR
- Set a target for %C4 based on requirement.

4.5 Improved crude switch handling - CDU

Before implementing Profit Controller, a number of actions had to be done manually during crude switch such as:

- Operator initially decided/ guessed the correct TBP cutpoint setpoints depending on the crude type and desired product specification. Then he had to make small adjustments to get the right product quality. (Our APC application does this automatically based on 15 inferential CVs as shown in Table-6.)
- Operator manipulated the fractionator sidestream flows to control overflash and gasoil delta P. GasOil delta P is the pressure drop in the gasoil section of the fractionator and is an important indication of the onset of flooding in the fractionator. (Profit Controller does this automatically since overflash and GO deltaP are CV's. There is also a crude switch DV for GO deltaP.)
- Operator manipulated furnace feed rate to achieve the correct CDU feed target. (Profit Controller does this automatically. CDU Feed is a CV and it also has a crude switch DV model.)
- During crude switch, operator used to manually increase COT while observing oxygen and gasoil delta P. (Profit Controller does this automatically because it has these CVs and in addition to DV model.)

None of the above actions are required after Profit Controller.

4.6 Improved crude switch handling – Gas Plant

Profit Controller has several features to handle crude switches, e.g.:

- Debutanizer feed DV. The debutanizer feed changes significantly in the event of a crude switch and this DV gives advance information to the controller before the disturbance hits the column.
- LPG in debutanizer feed DV. The LPG in debutanizer feed is calculated by the GCC model. Similar to the debutanizer feed DV, this gives advance information to the controller.
- Debutanizer overhead flow CV. During some of the crude switches the debutanizer ovhd flow drops too low. This CV enables Profit Controller to keep this flow in control by manipulating the recycle flow MV's.

With Profit Controller, crude switching requires almost no action by the operator. Before Profit Controller, the operator had to perform several actions during crude switches such as:

- Ramp setpoints of debutanizer bottoms RVP and overhead C5 control
- Adjust debutanizer and depropanizer duties
- Sometimes adjust the fractionator ovhd. temperature in order to prevent overloading the debutanizer. (Profit Controller now takes care of this constraint automatically since CDU controller has a max debutanizer feed CV limit.)

4.7 Light Kero flow minimized

After Profit Controller, the light kero flow could be minimized (Fig – L). This results in more heat recovery in the crude preheat section because light kero doesn't take part in heat exchange while the 2nd kero and 3rd kero do. This also leads to improvement in fractionation.
This benefit was also not quantified.

4.8 Better constraints handling

With multivariable controller, a lot more constraints could be handled easily as compared to the regulatory schemes earlier. As a result :

- Better optimization could be achieved.
  (Since Profit Controller is controlling all the CVs every minute, it does a better job at pushing them against the limits as compared to the operator).

- Operator load was reduced.
  (Almost all of the possible constraints were built into Profit Controller as CV's, including the variables which only constrained occasionally.)

The following 'new' constraints (CV's) were handled by the controllers.

Table 10 New constraints handled by Profit Controller controllers

CDU Profit Controller		
Desalter Pressure OP Crude Line Pressure OP (6PC32.OP) Prefractionator Ovhd Flow OP Prefractionator Sea-water Temp FT Ovhd Blower suction OP FT Ovhd Receiver pressure (6PI37) FT DewPoint - Top Temperature FT Ovhd Sea Water Temp	Minimum Furnace Pass Flow OP HSR Reflux Flow Amps Overflash Flow Gas Oil Delta Pressure Debutanizer Feed WSR90% Kero 95% Kero Freeze	Kero Smoke Gas Oil 95% JP Freeze JP 50% Molex / JP End Point Molex 10% Gas Oil 95% Gas Oil Color
Gas Plant Profit Controller		
Depropanizer Overhead C4 Depropanizer Bottom C3	Debutanizer Tray-1 Load Debutanizer Tray-35 Load	Debutanizer Overhead Flow (6FC35)

5. CONCLUSION

A number if tangible and intangible benefits were achieved by multivariable control implementation. An annual benefit of about 0.6 million dollars per year could be readily quantified.

The operator load for normal plant operation and crude switch handling for CDU and gas plant was reduced. Almost all the possible constraints for the crude unit were incorporated in Profit Controller. A few additional benefits existed which could not be readily quantified e.g. those due to high quality inferential CVs for CDU products.

For the future there is potentially a huge benefit if the economics change so as to maximize the CDU feed.

6. REFERENCES

1. “Control of crude fractionator product qualities during feedstock changes by use of a simplified heat balance”.
Y. Zak Friedman, 1985 American Control Conference.

7. ACKNOWLEDGEMENTS

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8. DIAGRAMS

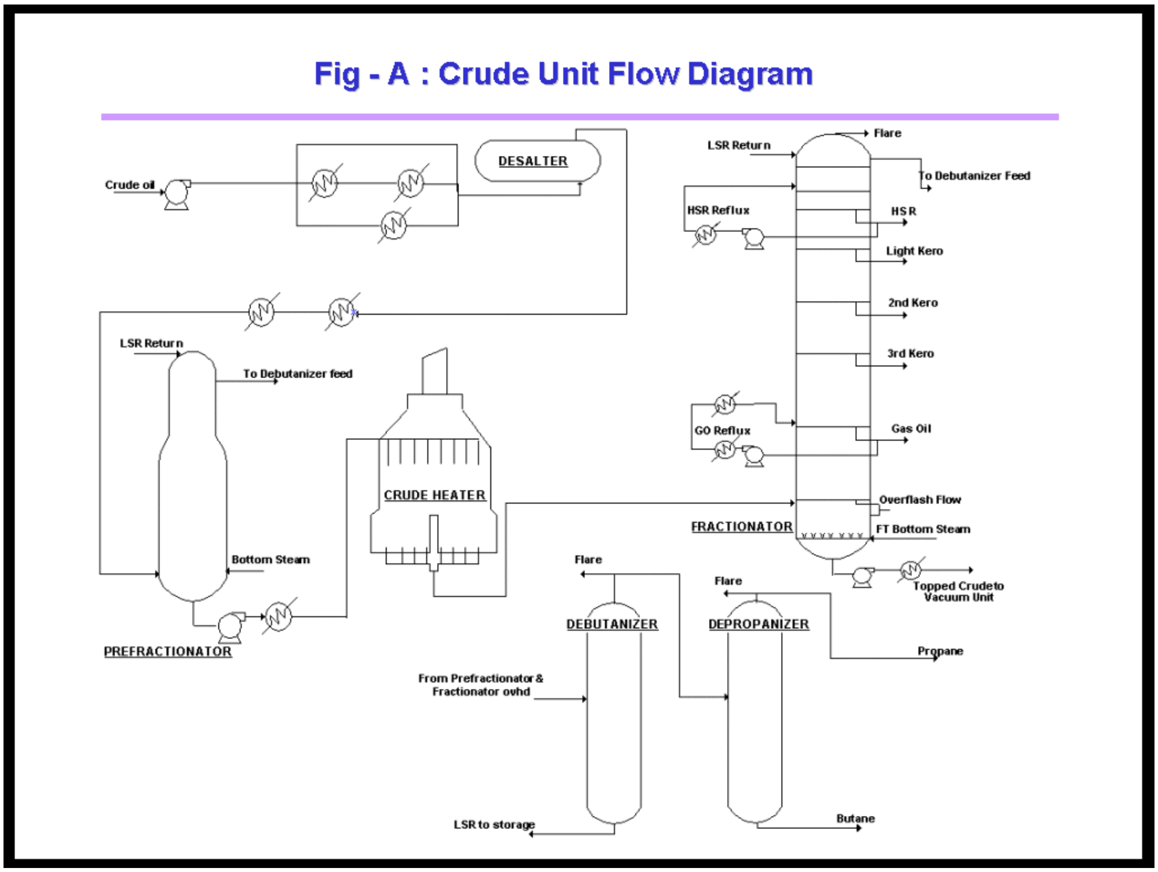


Fig - A : Crude Unit Flow Diagram

Fig - B : Non-Linear TBP Curve

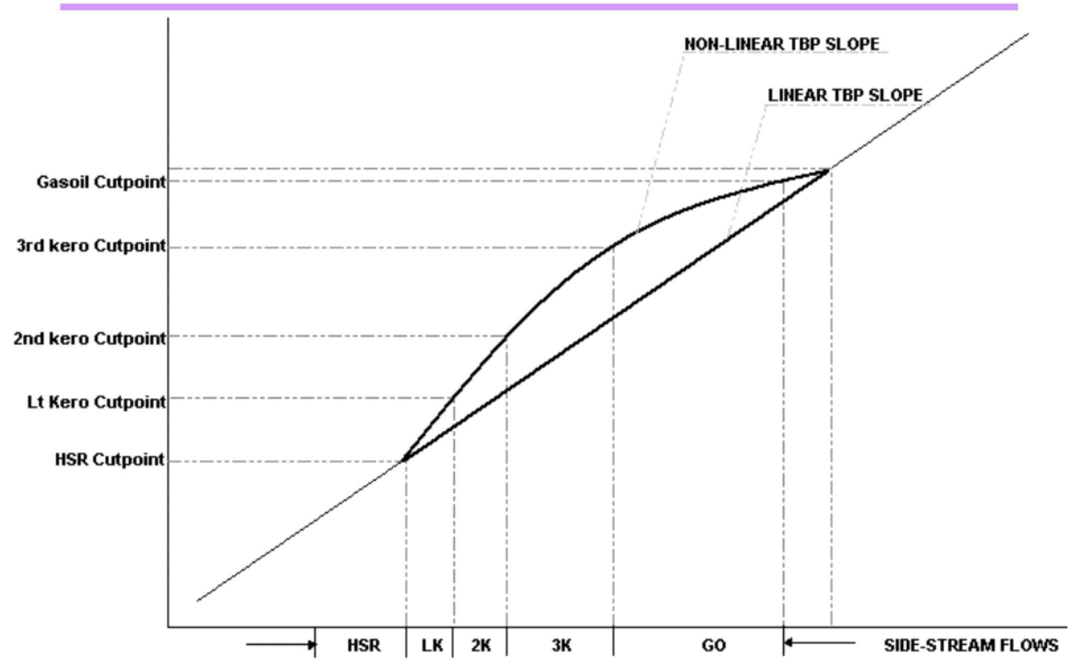


Fig - C : Inferentials - JP EndPoint

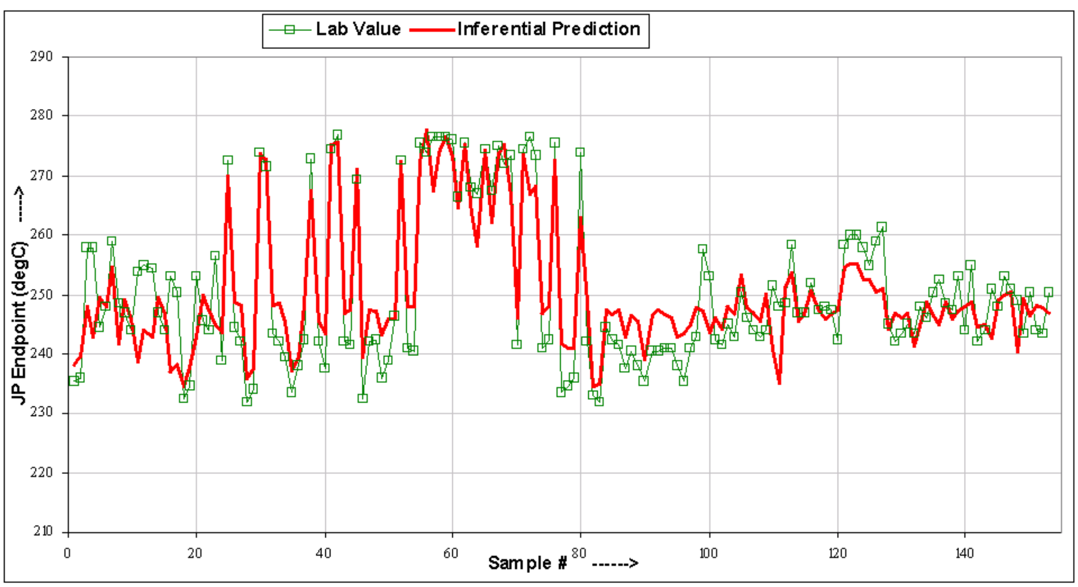


Fig - D : Inferentials - Kero 95%

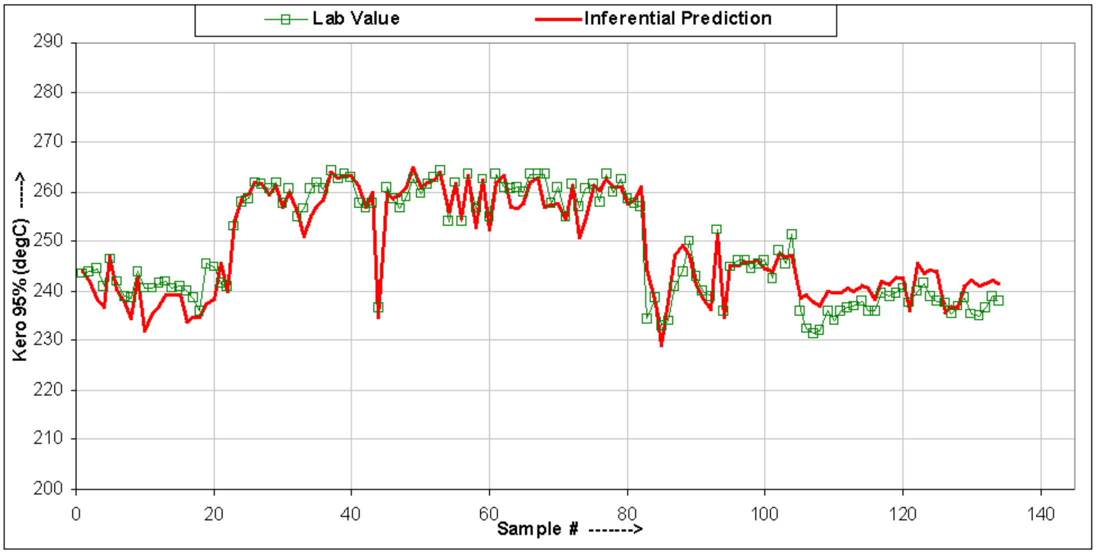


Fig - E : Inferentials - Kero Freeze Point

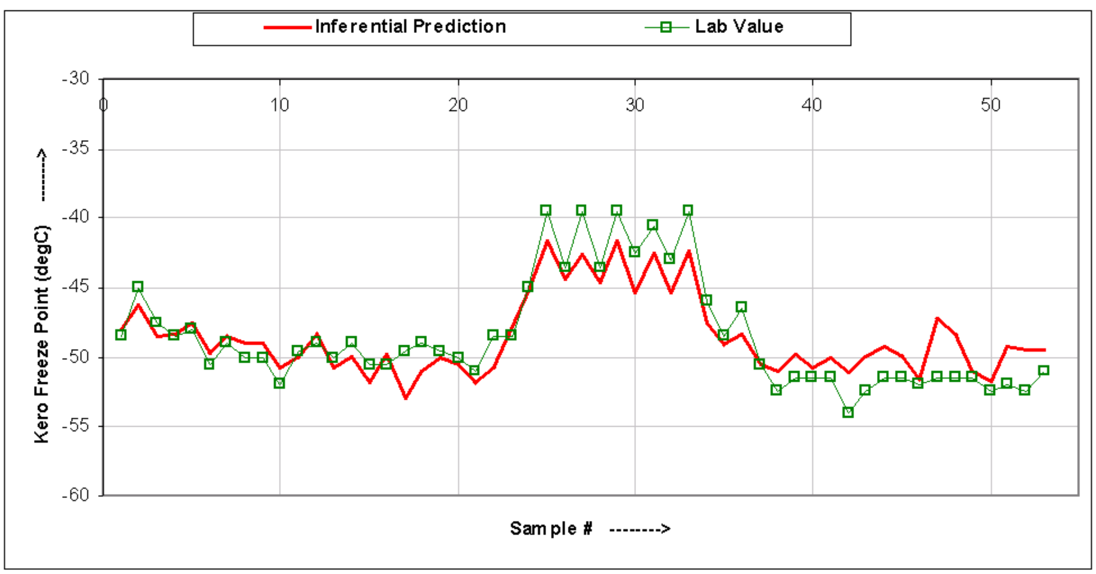


Fig - F : Inferentials - JP Flash Point

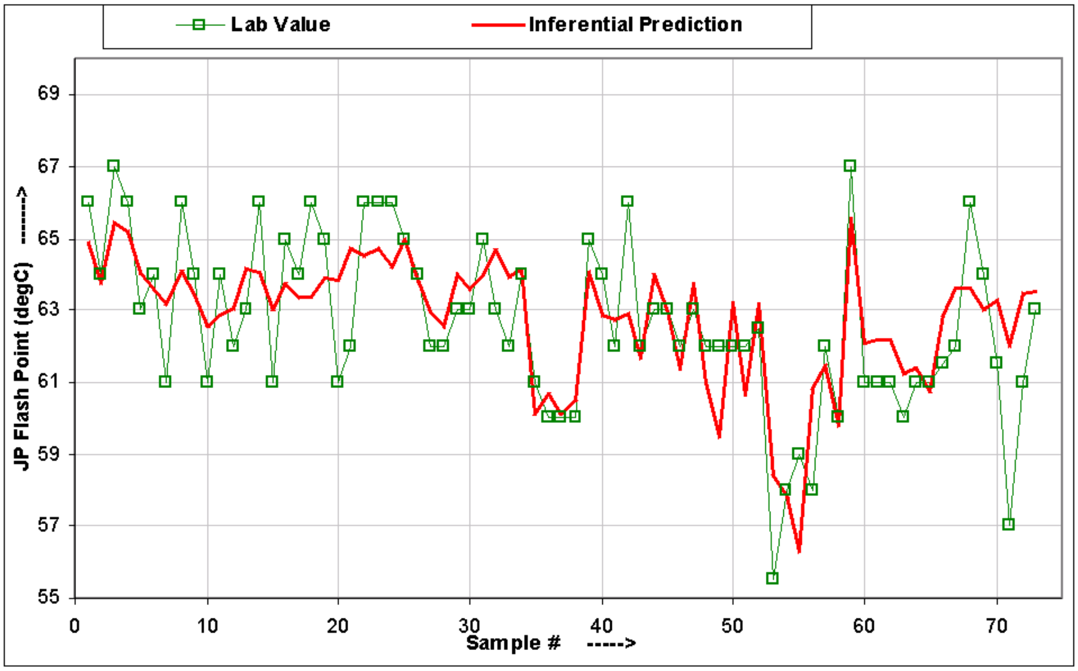


Fig - G : Inferentials - BGO CFPP

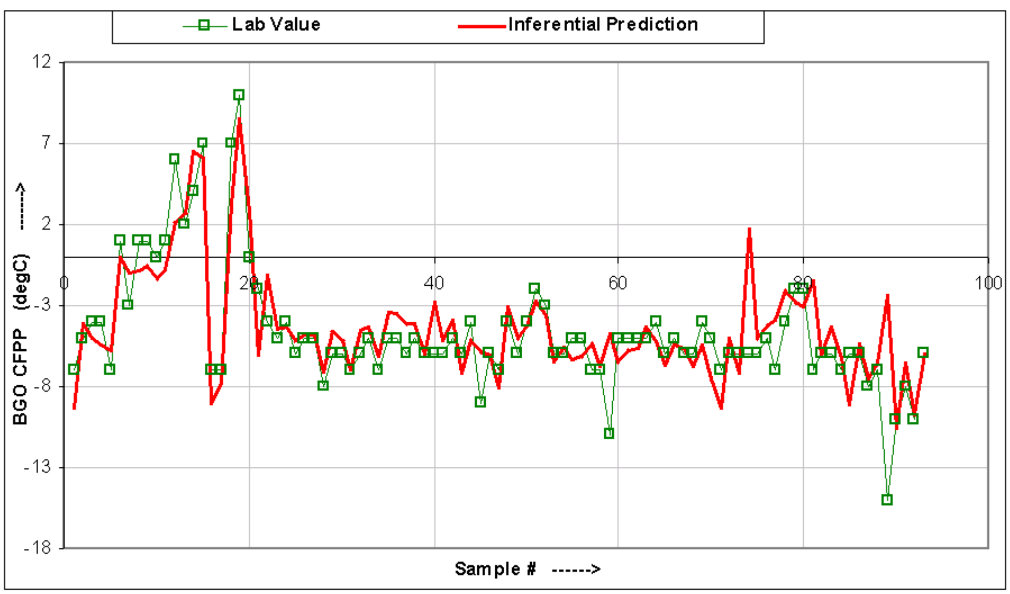


Fig - H : Inferentials - Molex 10%

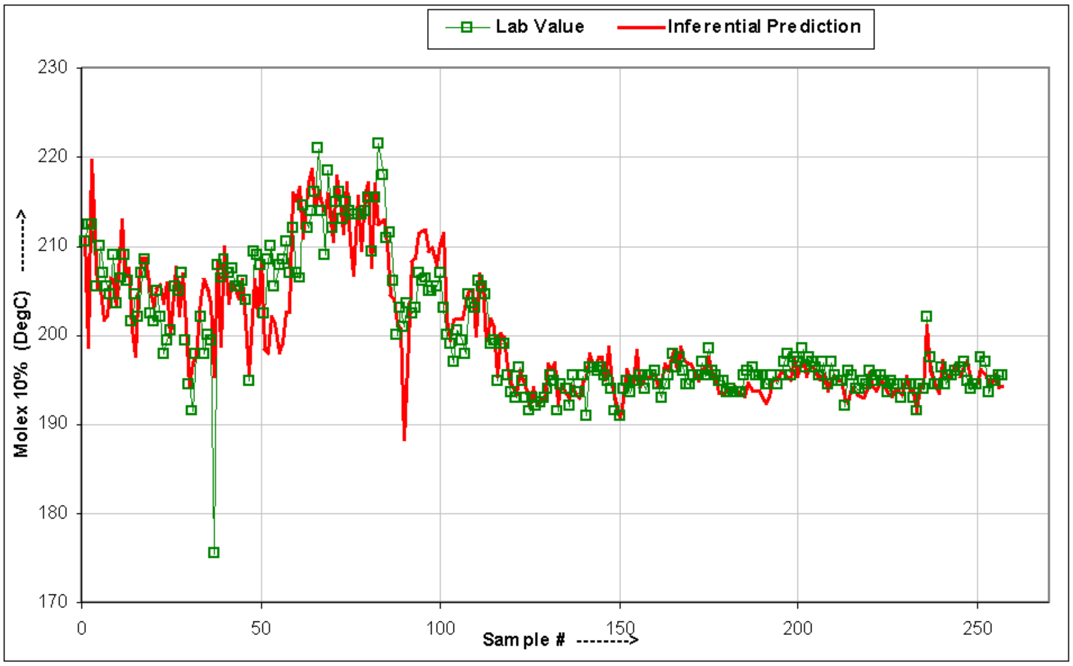


Fig - I : Inferentials - Kero Flash Point

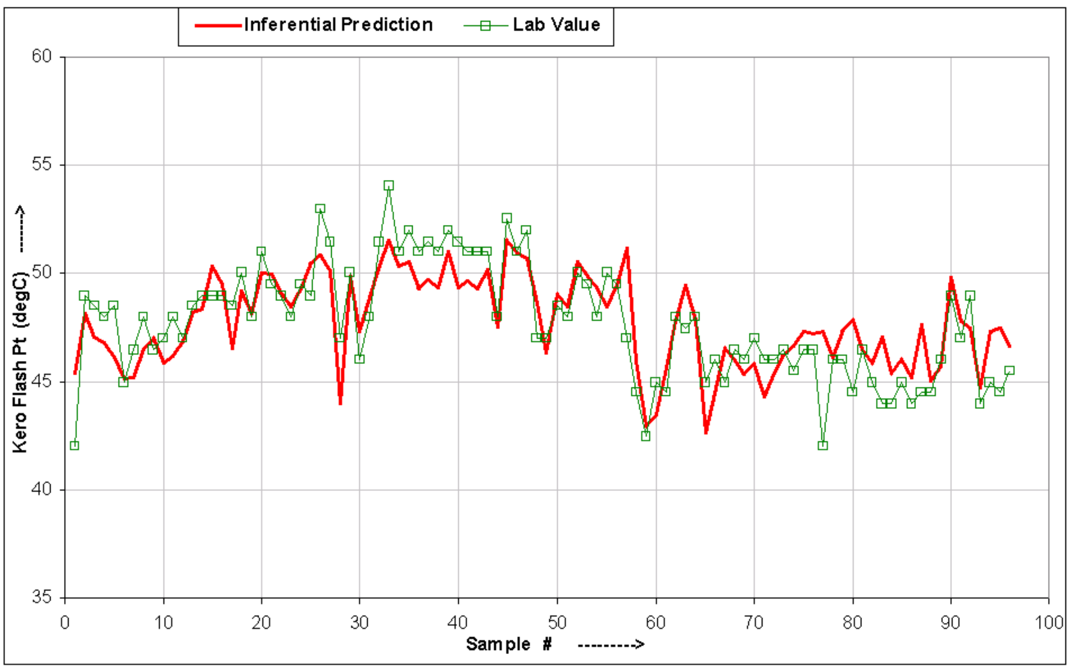


Fig - J : Increase in Absorbed Duty of CDU Furnace

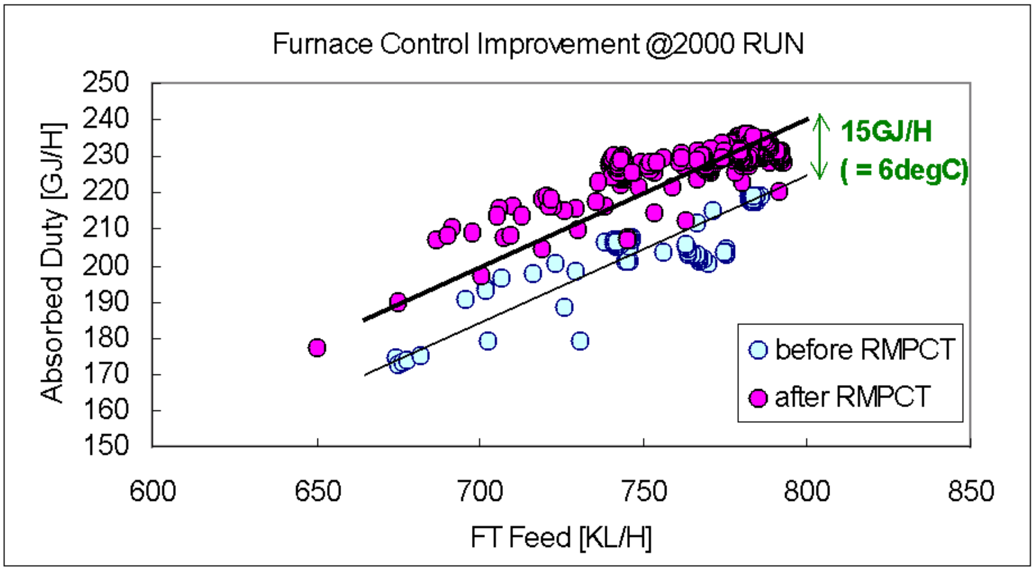


Fig - K : CDU Fractionator Pressure Reduction

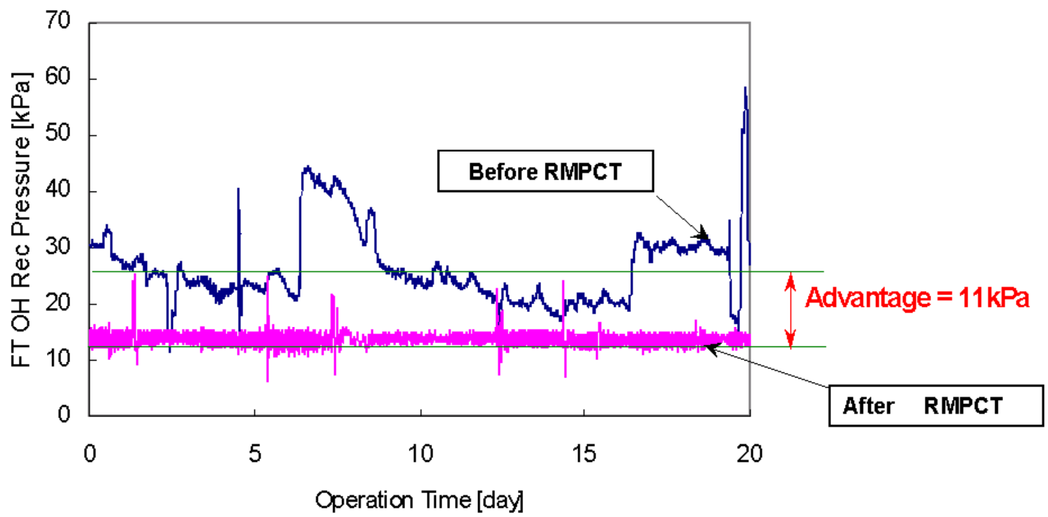
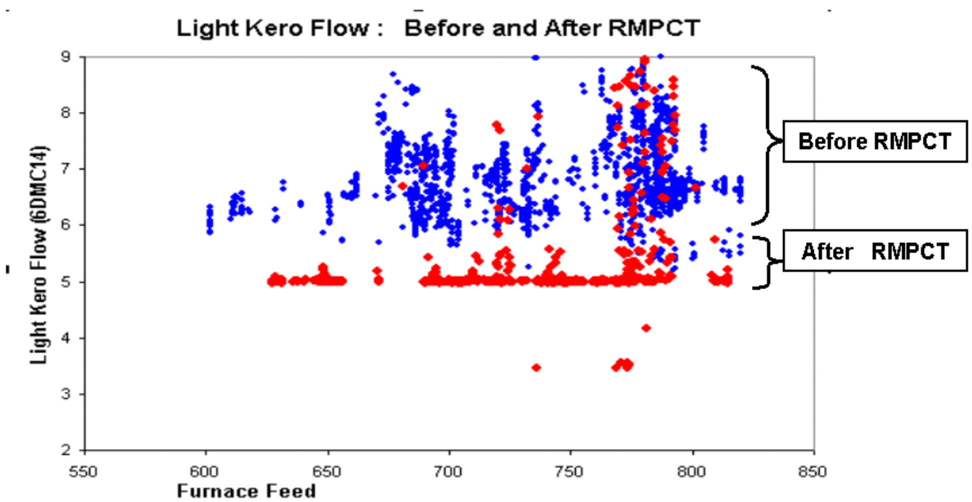


Fig - L : Other Benefits: Light Kero Flow Minimization



ADVANTAGES : **(a) More heat recovery.**

(b) Better Fractionation.