

## Experience with “Grey Box” Modelling and Optimisation of Refinery Processes

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### Abstract

This paper discusses experience with an alternative approach to process modelling and real time optimisation at the Chevron refinery at Pembroke, UK.

Traditionally, real time optimisation has been almost synonymous with the use of steady-state, first-principles process models within on-line applications. Over the past decade, however, there has been a growing interest in dynamic optimisation using empirical linear models or block-oriented non-linear models in structures that are known as Hammerstein or Wiener systems. The inclusion of non-linear terms in the dynamic process model not only provides improved model accuracy over a wider range of operation, it also enables the optimiser to solve a variety of problems that cannot readily be solved correctly using linear models, such as those involving parallel equipment or blending processes. On the other hand, the inclusion of non-linear terms also further complicates the problem of identifying suitable process models using purely empirical “black box” modelling techniques. Hence, it is often desirable to be able to make use of some fundamental *a priori* knowledge about the process without necessarily having to go all the way back to rigorous first-principles modelling. This leads to a hybrid approach, which is sometimes referred to as “grey box” modelling because it attempts to combine the advantages of traditional steady-state process models and “black box” dynamic models.

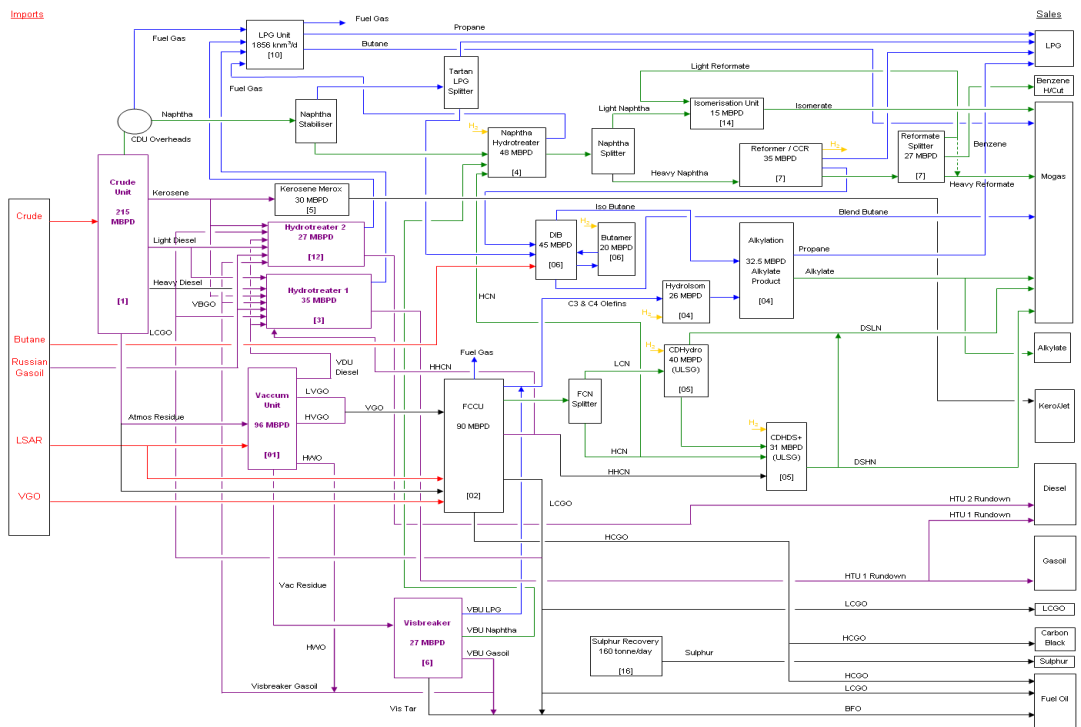
Such a “grey box” modelling approach was applied to the optimisation of a large section of the processing facilities of the Chevron Pembroke refinery, including crude and vacuum distillation, visbreaking and two parallel diesel hydrotreaters. The paper focuses primarily on the optimisation of the Heat Exchange Network of the crude unit as an example of the modelling and real time optimisation of non-linear systems using a “grey box” modelling approach.

The optimisation system was commissioned in August 2006. The paper includes a discussion of the achieved benefits and the criteria for acceptance among the direct users as well as other stakeholders in the refinery optimisation process.

### Chevron Pembroke ULSD Production Optimisation Project

The Chevron refinery at Pembroke (UK) is a 215 kbpd fuels refinery where vacuum gasoils and atmospheric residue are processed in an FCC unit and vacuum residue is subjected to visbreaking. The crude unit undergoes frequent feed changes causing significant operational disturbances. Nevertheless, the operation of most of the process units can generally be characterised as a single-mode operation, where the downstream units process the intermediate products directly from the crude unit, in some cases supplemented by imported feedstocks. A simplified block diagram of the Pembroke refinery is provided in figure 1.

Figure 1  
Simplified Block Diagram of Pembroke Refinery



In the second half of 2005, a feasibility study was performed for a real time optimisation system covering the following process units of the Pembroke refinery:

- Atmospheric Crude Distillation Unit
- Vacuum Distillation Unit
- Visbreaker
- Middle Distillate Hydrotreaters HTU1 and HTU2

The feasibility study was kicked off with a series of meetings with operations, scheduling, planning and engineering staff, followed by a statistical analysis of process and laboratory data and a review of various existing Multi-Variable Control (MVC) applications. Finally, a simplified model of the process system was built using TCA's Generic Dynamic Optimisation Technology (GDOT) and this model was used to quantify the combined effects of the individual benefits estimated from the statistical analysis.

The outcome of the feasibility study was that fairly significant benefits, amounting to several million USD per year, would be achievable by optimisation of the above process facilities. Some of the primary contributors to these benefits were:

- Increased HTU capacity utilisation
- Reduced ULSD quality giveaway (primarily density)
- Increased upgrade of LCGO to diesel
- Reduced average naphtha cutpoint
- Improved Heavy Diesel cutpoint control and increase average cutpoint

The recommendation was to create a two-layer solution, where the optimum targets established by the on-line optimiser would be implemented via a number of MVC applications. However, the review of the existing MVC applications confirmed that significant effort had to be put into this layer in order to improve its utilisation rate and make it sufficiently robust to support the future optimiser. Hence, the first phase of the actual project was an overhaul of a number of existing applications for the "black oil" units (CDU, VDU and VBU). The overhaul included review and modifications to the conceptual design of the regulatory control layer as well as the MVC layer and rework of various critical

inferred properties. Also, most of the dynamic models used by the MVC applications of these units had to be rebuilt due to the changes to the conceptual design.

In parallel with these activities, the design of new applications for the hydrotreaters was started. Previously, these units had been equipped with just three small MVC applications, which were too limited in scope to support the optimiser. A further MVC activity, which was not foreseen in the original project schedule, was the migration of all MVC applications to a different software platform.

In the second half of 2006, when most of the MVC work had been completed, the design and implementation of the on-line optimiser began. The optimisation system is relatively large and complex with about 200 independent variables to be optimised subject to about 400 constraints. Hence, the optimisation problem was initially broken down into five separate optimisers, one for each process unit. After having tested the individual optimiser modules in closed loop and familiarised the operators with the key functionality of the optimiser, an area-wide optimiser for the entire complex was built and commissioned in August 2006.

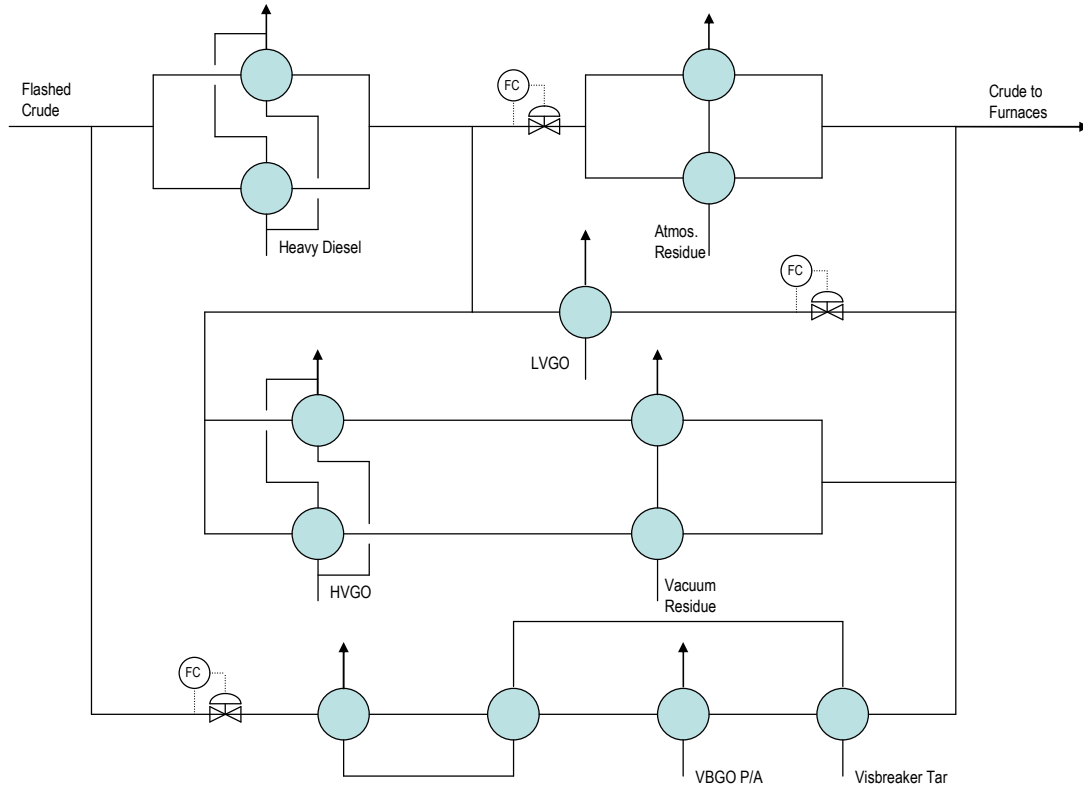
### **Optimiser Model Structure**

The backbone of the GDOT optimiser model is based on TCA's proprietary model of ULSD complexes. This model, which has now been implemented at three refinery sites, is fairly general and can be applied quite easily to a variety of different configurations, both in terms of the number of parallel process units and the design of the individual units. As illustration of the versatility of the model, the Pembroke crude unit has a conventional column with partial draws, whereas in the two other implementations some or all of the side draws are total draws with pumped and metered refluxes. The model has been designed to provide an optimum balance between complexity and accuracy. In particular, it has been designed to ensure global material balance consistency throughout the system.

Chevron is committed to energy efficiency and optimisation of the crude preheat train represents another significant opportunity with each degree Celsius of preheat worth in excess of \$100k per year. Hence, it was decided to create a GDOT model also of the flashed crude Heat Exchange Network (HEN). A simplified diagram of the network is provided in figure 2. The key aspect of the network is that the flashed crude is split into four parallel

branches, one branch recovers heat from the atmospheric residue, two branches recover heat from the vacuum tower products and refluxes and the fourth branch recovers heat from the visbreaker fractionator products and refluxes. The regulatory control system is such that the second vacuum unit branch takes the marginal flow; the other three streams are on flow control.

Figure 2  
Flashed Crude Heat Exchange Network



The model of the HEN is used as example to illustrate some of the modelling techniques used in GDOT.

The primary optimisation objective here is of course to maximise the temperature of the combined crude furnace inlet stream

$$T_{AVG} = \frac{F_{CDU} * T_{CDU} + F_{VDU1} * T_{VDU1} + F_{VDU2} * T_{VDU2} + F_{VBU} * T_{VBU}}{F_{CDU} + F_{VDU1} + F_{VDU2} + F_{VBU}}$$

(eqn. 1)

Each heat exchanger in the network is modelled by the following basic equations:

$$Q = cp_{HS} * sg_{HS} * F_{HS} * (T_{inHS} - T_{outHS}) = cp_{CS} * sg_{CS} * F_{CS} * (T_{outCS} - T_{inCS}) = U * A * MTD$$

(eqn. 2)

where

Q is the heat duty of the exchanger, kJ/h

cp is the heat capacity, kJ/kg/°C

sg is the density, kg/m<sup>3</sup>

F is the volumetric flow, m<sup>3</sup>/h

T<sub>in</sub> is the inlet temperature, °C

T<sub>out</sub> is the outlet temperature, °C

U is the heat transfer coefficient, kJ/m<sup>2</sup>/°C

A is the surface area, m<sup>2</sup>

MTD is the corrected logarithmic mean temperature difference, °C

and where subscript HS refers to the hot side and subscript CS refers to the cold side. The corrected logarithmic mean temperature difference is defined as

$$MTD = f * LMTD$$

$$LMTD = \frac{(T_{inHS} - T_{outCS}) - (T_{outHS} - T_{inCS})}{LN(T_{inHS} - T_{outCS}) - LN(T_{outHS} - T_{inCS})}$$

(eqn. 3)

where f is a correction factor that depends on the design of the exchanger shells.

Now, if cp\*sg and U\*A\*f can be assumed to be fairly constant, then it is relatively straightforward to rewrite the above equations in a dynamic open-equations format that can be solved by GDOT. The resulting equation system has the flows and temperatures as

independent variables and the above equations as constraints. In reality, the heat transfer coefficients will of course vary due to exchanger fouling and variations in the product flow rates and product properties. Hence, the heat transfer coefficients are estimated on-line from process measurements. In the general case, this presents a typical data reconciliation problem, where redundant information exists for some exchangers, whereas other exchangers require inputs that are estimated from the heat balance of other exchangers. In the particular case of the Pembroke flashed crude HEN, however, the amount of redundant information was quite limited, and it was fairly easy to decide, based on past reliability, which instruments to use for the on-line estimation of the transfer coefficients.

The heat exchange is optimised subject to various process constraints, e.g. valve positions, pressures and product rundown temperatures. The optimum solution from GDOT is output to the DCS via an existing MVC application. It is worth noting that no additional step testing was required for the GDOT application. Valve characteristics were based on historical data and process dynamics were partly lifted from the existing MVC application and partly estimated from experience and “common sense”.

It is also worth noting that the optimum solution from GDOT differs significantly from the optimum solution that the MVC application would generate in “stand-alone” mode. As control engineers, we sometimes take a too simplistic approach to this type of optimisation problems. In this case, the MVC application has the stand-alone objective to balance the outlet temperatures of the individual branches by driving the difference between the outlet temperature of each flow-controlled branch and the outlet temperature of the VDU2 branch, which takes the marginal flow, to a target of zero. However, by rearranging (eqn. 1), differentiating with respect to one of the control flows, e.g. the CDU branch, and remembering that  $\partial F_{VDU2}/\partial F_{CDU} = -1$  because the VDU2 branch closes the material balance, one obtains

$$\partial T_{AVG}/\partial F_{CDU} * (F_{CDU} + F_{VDU1} + F_{VDU2} + F_{VBU}) = \partial T_{CDU}/\partial F_{CDU} * F_{CDU} + T_{CDU} - \partial T_{VDU2}/\partial F_{CDU} * F_{VDU2} - T_{VDU2}$$

(eqn. 4)

Thus,  $T_{CDU} = T_{VDU2}$  is only a stationary point, and hence probably a maximum for  $T_{AVG}$ , if  $\partial T_{CDU}/\partial F_{CDU} * F_{CDU} = \partial T_{VDU2}/\partial F_{CDU} * F_{VDU2}$ . In other words: The operating point where  $T_{CDU} = T_{VDU2}$  is not necessarily a stationary point for  $T_{AVG}$ , except in the special case where this point

coincides with the point where the effect of the change in the CDU outlet temperature on the duty recovered by the CDU branch is equal to the effect of the change in the VDU2 outlet temperature on the duty recovered by the VDU2 branch. This condition is normally not fulfilled in the Pembroke flashed crude Heat Exchange Network. In fact, the different hot side flow rates in the different branches causes different pinch points, which means that the overall optimum may occur at significantly different branch outlet temperatures.

## **Project Results**

The project has resulted in a significant increase in MVC utilisation rate in this part of the Pembroke refinery. Also, the utilisation rate of the GDOT optimiser is generally very good. Overall, most of the project objectives have been met.

The direct comparison with the baseline established during the feasibility study is somewhat hampered by a significant product specification change that took place during the initial phases of the project. Essentially, the UK specifications for ULSD were aligned with those that now apply throughout most of the EU. This meant a tighter specification on sulphur (now 10 ppm) and a less restrictive specification on density (now 845 kg/m<sup>3</sup>). With the old specifications the pressure was clearly on density and there was essentially no such thing as avoidable density giveaway. With the current specifications, however, it is sometimes not possible to reach the maximum density because the sulphur constraint limits the processing of heavy feed components. This is particularly a problem if the high limit on sulphur has to be lowered to correct a contaminated tank, if the sulphur content of the crude is high or at EOR conditions on the hydrotreaters. On the other hand, the optimisation system is frequently used for crude rate maximisation, something that was not part of the original benefit estimate. Also, the crude preheat optimisation is additional to the expected benefits.

Some of the major challenges when it comes to achieving and sustaining the benefits of this type of optimisation project are the accuracy and reliability of instrumentation, in particular on-line analysers, and the effective usage of the system as a collaboration tool between individual console operators and between operations, coordination (production planning and scheduling) and the process support team.

Significant effort in this project has been devoted to the design and calibration of robust inferred property calculations trying to strike the best balance between accuracy and complexity, e.g. number of required process inputs. Nevertheless, some properties do require biasing by an on-line analyser, e.g. because the content of refractive sulphur in the hydrotreater feed streams cannot be accurately inferred or because some property estimates may be relying on flow measurements that are often less than 10% of the meter range. Unfortunately, despite the efforts of the refinery instrument department, the accuracy of the sulphur analysers in the low ppm range occasionally limits the processing of heavy feed components in the hydrotreaters.

The ULSD production optimiser at the Pembroke refinery is shared between four console operators located in two separate control room buildings, and crosses three LCN's. An important lesson from this project, and from other optimisation projects of comparable size and complexity, is that, with today's technology, any difficulties related to networking infrastructure and computing resources are much easier to overcome than those related to user training and user interface. The optimiser will find a solution for the area-wide optimisation problem subject to the limits that have been entered by the individual console operators. Specifying sensible, non-conflicting limits is the key to achieving benefits with any multivariable control or optimisation tool. Experience shows that this can be difficult enough even with a unit-wide control and optimisation system. In order to effectively use an area-wide optimiser, where imposing irrelevant constraints may adversely affect also units controlled from other consoles, it is absolutely essential that the operators receive adequate training.

As an extension of the control system, the optimiser is owned by the console operators, and, hence, it can be viewed as a productivity tool for the operations department. Nevertheless, it is crucial that the team that supports the optimiser understands the role of the optimiser in the collaboration between operations, coordination and the unit engineers. In particular, the console operators must be able to readily implement the instructions received from the coordination group and the unit engineers in the optimisation system. If this is not the case, the system or the instructions must be changed. Also, the optimiser model and the associated objective function must be monitored and maintained to ensure that the optimiser drives the plant in the direction expected by the coordination group. This all requires significant effort and persistence. On the other hand, an optimisation project that truly supports and enhances

the collaboration between the various stakeholders in the refinery optimisation process undoubtedly produces substantial spin-off benefits that make it all more than worthwhile.