This is a repository copy of *Gas Phase Train in Upstream Oil & Gas Fields: PART-I Model Development* .

White Rose Research Online URL for this paper:
http://eprints.whiterose.ac.uk/109286/

Version: Accepted Version

A Process Model Representing Gas Phase Train in Upstream Oil & Gas Fields: PART-I

Y. H. Al-Naumani *,** **J. A. Rossiter** ** **S. J. Bahlawi** ***

Quality Measuring Instruments and Metering Supervisor - OSGOH432, Petroleum Development Oman LLC, P.O.Box 81, Postal code 100, Muscat, Sultanate of Oman (yahya.yn.naamani@pdo.co.om)
*** Department of Automatic Control and Systems Engineering University of Sheffield. Sheffield S1,3JD UK (yhal-naumani1@sheffield.ac.uk) (j.a.rossiter@sheffield.ac.uk)*
**** Harweel Production Support Coordinator - OSGOH2, Petroleum Development Oman LLC, P.O.Box 81, Postal code 100, Muscat, Sultanate of Oman (Said.jh.bahlawi@pdo.co.om)*

Abstract: The prime contribution of this paper is to provide a large scale system (LSS) model for the gas phase operation in upstream oil and gas plants. The process model consists of the three main gas conditioning processes which exist in most upstream oil and gas gathering stations; these are gas sweetening, gas dehydration, and hydrocarbon dew-pointing. The function of such a model is to provide a realistic process representation in order to test and verify different process control approaches, specifically those control themes which deal with highly interactive control loops.

1. INTRODUCTION

Process models representing upstream oil & gas processes are scarce in the literature. The majority of the process models available in the literature represent single chemical processes. In order to investigate different control structures and proposals it is necessary to have a suitable benchmark model and/or scenario reflecting realistic upstream oil & gas operations. Such a model would not only be of benefit to studies on upstream oil & gas processes, but also to Large Scale Systems (LSS) in general and control research on how to deal with system interaction as well.

The scarcity of process models in the literature could be due to the fact that most of the upstream process control goals can be achieved sufficiently well by implementing Single Input Single Output (SISO) control strategies. This comment is true to some extent for simple process control operations such as level control of storage tanks and furnace temperature control. On the other hand the situation is different with complex oil & gas dynamic processes such as control of fractionation columns, compressor surge control or crude stabiliser column control. These units contain a number of interactive control loops and thus it is often difficult to tune SISO loops to control such processes effectively. Nevertheless in practice these process are often controlled using simple control strategies with one consequence being that their performance and stability are sensitive to disturbances and load changes.

Looking to the past, Shell Oil's heavy oil fractionater model was one of the earliest models presented in the literature to represent a multivariable interactive control process. The distillation column model introduced by Prett and Morari (1987) has three controlled variables and three manipulated variables which are highly interactive with each other. For decades, this model was the base for many studies of different control approaches and strategies for distillation columns control.

A few years later, the first and most famous plant-wide process control problem 'Tennessee Eastman' (TE) was proposed by Downs and Vogel (1993) as a control challenge test problem. It was based on an actual industrial plant and consists of number of linked chemical process units with multivariable control loops which can be subdivided into four or five interacting subsystems. The TE process characteristics were described by sets of flow sheets and steady-state material balances rather than transfer function models or model equations.

The intention of this paper is to provide the first model representing typical gas train processes in upstream oil & gas plants and based on transfer functions; that is, a model which is easy, simple to understand, and fit for purpose. The proposed model is aimed to be the base element for process control designers to overcome the disturbance growth issue in the series connected processes (which will be investigated in the second part of the paper). The model provides good opportunity for control engineers to test different variety of process disturbances, malfunctions, and load changes on the process operation and verify its signif-

icance. The model can be used to develop a specific Model Based Predictive Controller (MBPC) which is expected to significantly reduce the plant shut down occasions due to process disturbances and also to drop down operating cost by properly controlling the disturbance growth in the process, hence reducing energy fluctuations in the process and saves fuel.

The process description is illustrated in section two, while section three presents the plant-wide process model and section four gives some model validation and verification. Section five finishes with conclusions and future work recommendations.

2. PROCESS DESCRIPTION

Natural gas processing trains in upstream gas plants contain processes to purify the raw natural gas extracted from underground oil & gas fields and brought up to the surface by production wells. Raw natural gas typically consists primarily of methane, the lightest hydrocarbon molecule. It also contains significant amounts of ethane and varying amounts of heavier hydrocarbon products like natural gas liquids (NGL) compromising of propane, butane, and pentane and even higher molecular weight hydrocarbons such as crude oil. In addition, the gas contains undesirable impurities, such as liquid or vapour water, carbon dioxide (CO_2), hydrogen sulfide (H_2S), and mercaptans molecules (Baker and Lokhandwala (2008); Kidnay et al. (2011)).

The process model illustrated by Fig. 1, describes the three main processes which are commonly found in the upstream fields; upstream fields are classified as those with high gas to oil ratio (GOR). The processes are Gas Sweetening, Gas Dehydration, and Hydrocarbon Dew Pointing. Table 1, provides a description of the abbreviations used in figure 1.

In line with the strict global regulations to reduce levels of sulphur and carbon dioxide content inside gaseous hydrocarbons used as fuel, it is necessary to remove sulphur and carbonic dioxide from the gas to meet the specifications on acid gas imposed by safety procedures, transport requirements, and distribution specifications (Ohshiro and Izumi (1999)).

| Acronym	Description
DPIC	Differential Pressure Indicator Control
FIC	Flow Indicator Control
FCV	Flow Control Valve
GDU	Gas Dehydration Unit
GSU	Gas Sweetening Unit
HCDP	Hydrocarbon Dew Pointing Unit
IGV	Inlet Guide Vanes
LIC	Level Indicator Control
LCV	Level Control Valve
PIC	Pressure Indicator Control
PCV	Pressure Control Valve
QIC	Quality Indicator Control
TIC	Temperature Indicator Control
TCV	Temperature Control Valve

Table 1. Process figure key
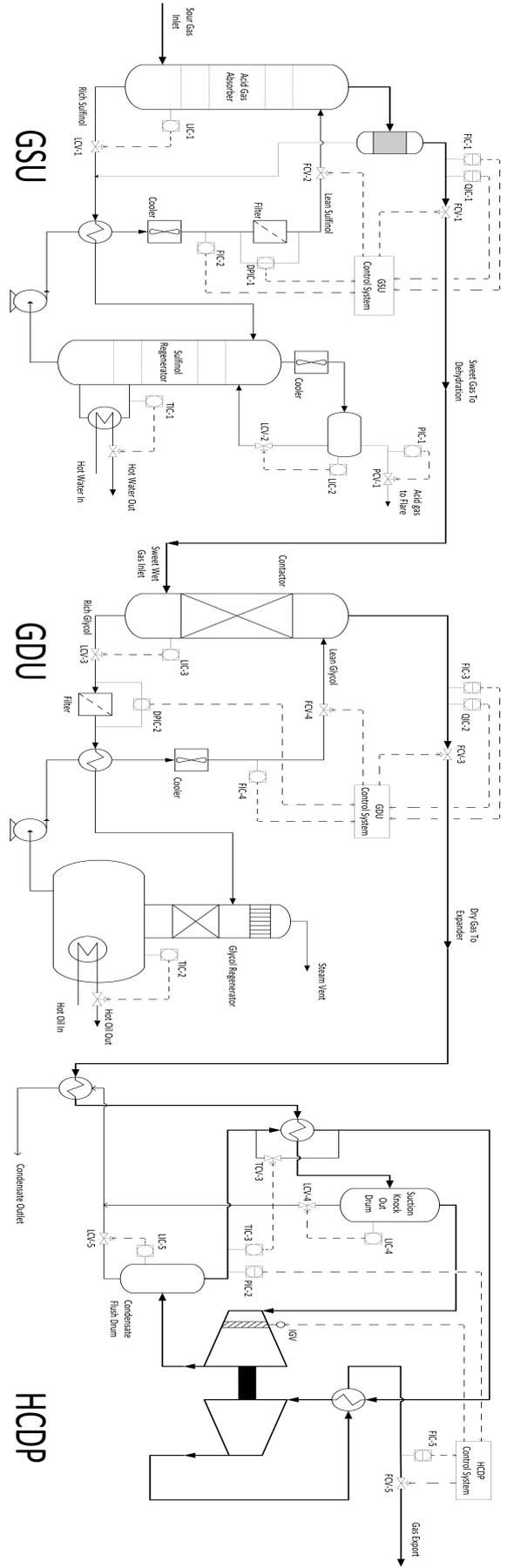
Fig. 1. Gas Processing Train

Sour gas produced from the oil and gas wells is separated from the crude in the production separators and then routed to the Gas Sweetening Unit (GSU) where it is treated to meet the hydrogen sulphide and carbon dioxide export gas specification and in addition the bulk of the mercaptans (RSH) and carbonyl sulphide (COS) contaminants are removed.

The GSU consists of an absorber where the acid gas is removed by a counter current contacting with sulfinol solvent and a regeneration loop where the sulfinol is regenerated via desorption of the acid gas components. The treated gas from the absorber is further washed in the Treated Gas Water Wash Vessel to minimize carry over of solvent to the downstream process. The treated gas subsequently flows to the Dehydration Unit (GDU) and Hydrocarbon Dew Pointing Unit (HCDP) for further treatment to remove moisture, and condensate in order to reach the final product quality.

The lean sulfinol flows downward through the GSU absorber contacting the upward flowing natural gas. Sulfinol absorbs acid gas components and other impurities from the natural gas, and leaves the bottom of the absorber as rich sulfinol under level control. Rich sulfinol then flows to the Lean/Rich Heat Exchangers where it is heated by the hot lean sulfinol from the Regenerator column.

The pre-heated rich sulfinol is then introduced to the top of the regenerator column, where the sulfinol solvent is regenerated by contacting with the stripping steam and recycled back to the system as lean sulfinol.

Presence of water vapour in the hydrocarbon gas raises several problems on the production operation and in the transportation. The water vapour may condense and cause the formation of hydrates, solidify or favour corrosion if the gas contains acid components (Kvenvolden and Lorenson (2001); Sloan (2003)). Henceforth the wet sweet gas stream from the gas sweetening units is sent to the Gas Dehydration Unit (GDU) where it contacts with a liquid stream of glycol which has a greater affinity for the water vapour than does the gas. Afterwards the dehydrated gas is then sent to the Hydrocarbon Dew Pointing Unit (HCDP).

After contacting the gas, the water-rich glycol is regenerated in the glycol regeneration package by heating at approximately atmospheric pressure to a temperature high enough to drive off almost all the absorbed water. The regenerated glycol is then cooled and re-circulated back to the contactor.

To meet the required quality for export gas specification, the gas is further processed in the Hydrocarbon Dew Pointing unit. The purpose of Hydrocarbon Dew Pointing Unit is to achieve the export gas specification of Gross Heat Value (GHV), Wobbe Index, and hydrocarbon dew point. This is done by expanding the gas from the GDU through a Turbo-Expander and removing the condensed heavier hydrocarbon as a liquid stream from the Cold Condensate Flash Drum. The gas is then compressed in a re-compressor and flows to Gas Export metering.

3. PROCESS MODEL

The model of the gas phase train is developed using first order transfer functions (where possible) with dead-time. Those simple models are sufficient to represent many chemical processes and moreover favoured in the industry (Forbes et al. (2015)). The benefits of using simple models may not be seen during design and commissioning phases when an expert control engineers are mostly present. However the benefits will be clearly visible during the operation phase when process engineers or plant operators can easily identify a model's gain, delay, and time constant and compare the information with the real process data. Hence use of such simple models builds confidence amongst the operation team and reduces the risk of large model errors (model-plant mismatch) which may arise due to staff difficulty in understanding and identifying higher order models for a large scale system.

The developed process model, as it sketched in figure 1, is based on plant with a natural gas specification of around 100 $barg$ pressure, $45^{\circ}C$ gas temperature and a throughput gas flow rate up to 3.0 $MMSCMD$ (Million Metric Standard Cubic Meter per day).

3.1 Gas Sweetening Unit (GSU) Dynamical Model

Referring to the GSU in figure 1, sour gas enters the bottom of the absorber column where the acid gas components are removed in a counter-current contact with the sulfinol flowing downwards from the top. The GSU system has two variables that have to be controlled: these are the throughput gas flow measured by *FIC-1* and the acid concentration in the gas outlet measured by the process analyser *QIC-1*. The manipulated variables are the absorber gas outlet flow through *FCV-1* and the absorber sulfinol input flow through *FCV-2*. The specification of the acid concentration in the outlet gas is fixed by operational goals and must be kept no more than 0.5% of its setpoint at steady state. *FIC-2* provides lean sulfinol flow measurements to the GSU control system, whereas the differential pressure sensor *DPIC-1* across the sulfinol filter provides measurements of the sulfinol flow disturbances.

The dynamics of the GSU system (inputs *FIC-1*, *QIC-1* and outputs *FCV-1*, *FCV-2* respectively) found to be well defined by the following model:

$$G_{GSU} = \begin{pmatrix} \dfrac{-13.5}{18.6s+1}e^{-2s} & \dfrac{16.7}{23.5s+1}e^{-6s} \\[2ex] \dfrac{7.3}{9.5s+1}e^{-13s} & \dfrac{20}{15.4s+1}e^{-6s} \end{pmatrix} \quad (1)$$

To evaluate the model response of the GSU model, a disturbance of 5% sulfinol filter chock has been introduced to the system at sample time 200. Filter chock is expected to limit sulfinol flow to the absorber and hence the gas flow through the absorber will kick off due to the reduction on the opposing flow. In response, the acidic gas is predicted to increase sharply driven by the sudden rise in the gas volumetric flow rate and the reduction of the solvent flow rate.
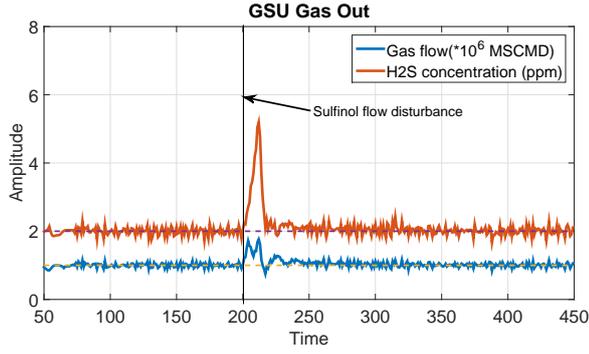
GSU Gas Out

Fig. 2. GSU gas outlet responses for a solvent filter chock under GPC control.

The GSU model trended in figure 2 responds to solvent filter chocks exactly as expected of a real gas sweetening process. The results are obtained using a multivariable Generalised Predictive Controller (GPC) with a prediction horizon of 30, and a control horizon of 5. (Rossiter (2013), Camacho et al. (2007)).

The absorber bottom liquid level is maintained by the level controller *LIC-1* which acts on the level control valve *LCV-1*. Level is one of the most common variables in the process industry. The model transfer function of the absorber level control can be approximated to:

$$LIC1 = \frac{1.2}{54s + 1}e^{-12s} \qquad (2)$$

The rich sulfinol is then routed to a low pressure Flash Vessel (not shown) where most of entrained and absorbed hydrocarbons, some of the sour components like H_2S, CO_2, COS, RSH, and water vapour are flashed off. Rich sulfinol then flows through the lean/rich heat exchanger (where it will be preheated) towards the top of the sulfinol regeneration column. The absorbed acid gases will be stripped off by the counter-current contacting with a stripping vapour produced by the re-boiler beneath the column. The most important controls here are the vapour pressure and temperature. The rich Sulfinol is heated in the re-boiler and the vapour is returned to the column for stripping the absorbed acid gas components from the solvent. The flow rate of heating media, that is hot water, is controlled through *TIC-1*. The vapour outlet of the regeneration column passes through overhead condenser and is then routed to the overhead separator to capture any hydrocarbons or sulfinol carried over by the gas; this is then recycled back to the regeneration column as a reflux.

3.2 Gas Dehydration Unit (GDU) Dynamical Model

The Gas Dehydration Unit is downstream of sweetening train as shown in figure 1. The GDU mainly consists of an export gas glycol contactor and dehydration regeneration package. The wet gas enters into the bottom section of the contactor column and then flows into the inlet scrubber section of the column where any entrained liquid is removed before the gas is introduced into the dehydration section of the contactor. All the liquids recovered in the bottom of the inlet scrubber are drawn down under level control *LIC-3*. A transfer function model of the level control at the bottom of the contactor is quite similar to the *LIC-1* of the GSU absorber.

Lean glycol enters at the top of the column and is equally distributed over the whole section of the column. The dehydration by absorption takes place as the gas flows upwards through the packing, contacting the wetted surface of the packing. The GDU system has two variables that have to be controlled; these are the throughput gas flow measured by *FIC-3* and the water vapour load in the gas outlet measured by the process analyser *QIC-2*. The manipulated variables are the contactor gas outlet flow through *FCV-3* and the contactor lean glycol input flow through *FCV-4*. The specification of the water vapour concentration in the outlet gas is fixed by operational goals and must be kept to within 0.5% of its setpoint at steady state. *FIC-4* provides lean glycol flow measurements to the GDU control system, whereas the differential pressure sensor *DPIC-2* across the rich glycol filter provides measurements of the glycol flow disturbances. The dynamics of the GDU system with these inputs (*FIC-3*, *QIC-2*) and outputs (*FCV-3*, *FCV-4*) are well defined by the following model:

$$G_{GDU} = \begin{pmatrix} \dfrac{-8}{15s+1}e^{-3s} & \dfrac{19}{30.3s+1}e^{-7s} \\[2ex] \dfrac{6.2}{13.5s+1}e^{-13s} & \dfrac{10}{16.7s+1}e^{-7s} \end{pmatrix} \qquad (3)$$

To evaluate the model response of the GDU model, a disturbance of 5% glycol filter chock has been introduced to the system at sample time 200. Filter chock is expected to limit glycol flow to the glycol regeneration package and causes disturbances to the regenerated glycol quality. Lean glycol flow to the contactor column is expected to be effected after a while which causes a small fluctuation in the gas flow rate. GDU control fluctuations are predicted to take a longer time to settle because the disturbance effects both operations in the system; those are the glycol regeneration package and the export gas dehydration.

It is clearly seen that, the GDU model responses as shown in figure 3 respond to glycol filter chocks exactly as expected of a real gas dehydration process. The results are obtained using a multivariable Generalised Predictive Controller (GPC) with a prediction horizon of 30, and a control horizon of 5.
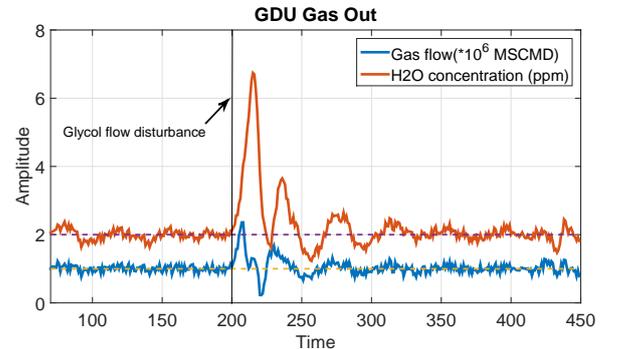

GDU Gas Out

Fig. 3. GDU Gas outlet responses for a glycol filter chock under GPC control.

The rich glycol leaves the contactor towards the Glycol Flash Drum (not shown). The Glycol Flash Drum is a horizontal three phase separator which separates the hydrocarbon liquid phase from the glycol phase and vents

any remaining gases. Thereafter the rich glycol flows through the Glycol Cartridge Filters to remove any solid particles from the rich glycol stream to prevent these solids from fouling the heat transfer surfaces within the glycol regeneration package. Filter chocks are continuously monitored by the differential pressure indicator *DPIC-2* installed across the filter cartridge.

The filtered glycol flows to the Lean/Rich Glycol Exchanger provided for energy conservation and reduces the total heat input required for the regeneration process. The rich glycol enters the Glycol Regeneration Column and flows down through the packing in a counter-flow stream to the upward water vapours from the reboiler beneath. In the reboiler, the rich glycol is heated to $202\,°C$ by a hot oil bundle to remove any volatile material. The temperature in the glycol reboiler is maintained by the temperature controller *TIC-2* which controls the flow rate of the hot oil. Finally the regenerated lean glycol flows to the glycol pump which circulates the glycol back into the system.

Control of temperature, like pressure and level, is one of the most common objectives in the process industry. The model transfer function of the repoiler temperature control *TIC-2* can be approximated to:

$$TIC2 = \frac{0.5}{45s + s}e^{-45s} \tag{4}$$

3.3 Hydrocarbon Dew-pointing Unit (HCDP) Dynamical Model

The export gas then flows through a further gas conditioning process called Hydrocarbon Dew Pointing (HCDP) to remove hydrocarbon liquids from the natural gas in order to achieve a defined export gas specification of Gross Heating Value (GHV), Wobbe Index and hydrocarbon dew point. The process consists in cooling the natural gas under the dew point temperature of the heavy hydrocarbons which also maximises the production of the natural gas liquid obtaining LPG (C3 and C4) from the raw gas.

The feed gas from the GDU, at $45\,°C$ and $95\ barg$ approximately, is cooled in the first heat exchanger by exchanging heat with the cold condensate return from the condensate flush drum. It is further cooled in the second heat exchanger by exchanging heat with separated gas from the condensate flush drum. The feed gas then flows to the suction knock out drum, where the temperature is further reduced to around $2\,°C$ by flashing. Thereafter the gas flows to the Turbo-Expander where it is expanded to $65\ barg$ and due to that the gas is cooled to around -15℃. The exit gas from Turbo-Expander flows to the Cold Condensate Flash Drum, in which the condensed hydrocarbon liquid is removed. The dry gas from the Cold Condensate Flash Drum is then heated up by exchanging heat with incoming feed gas. Afterwards the gas enters the Re-compressor section, where it is pressurised to around $70\ barg$ and then flows to the export pipeline after which it is being cooled down via the third heat exchanger.

The performance of the HCDP unit is mainly driven by the operating pressure and temperature. The two main controllers for this function are *PIC-2* and *TIC-3*, see figure 1.

The temperature control of the export gas Turbo expander is achieved by *TIC-3* located at vapour outlet of the condensate flush drum. *TIC-3* will throttle the control valve *TCV-3* provided in the cold bypass line of the second heat exchanger to maintain the Turbo expander inlet temperature. Achieving this temperature is very important to remove the liquid condensate and attain the export gas specification.

The Turbo Expander has two variables to be controlled to maintain the quality of the product; these are unit pressure measured by *PIC-2* which is located at the vapour outlet of the condensate flush drum and the load demand on the unit measured by *FIC-5*. The manipulated variables are the re-compressor outlet flow measured through *FCV-5* and the expander inlet flow through *IGV* (Inlet Guide Vans). The dynamics of the Turbo Expander can be described by the following model (inputs *PIC-2*, *FIC-5* and outputs *FCV-5*, *IGV* respectively):

$$G_{HCDP} = \begin{pmatrix} \dfrac{0.2}{2s^2 + 4s + 1}e^{-s} & \dfrac{1}{2s + 1} \\ \dfrac{0.3}{0.4s^2 + s + 1}e^{-0.5s} & \dfrac{-0.3}{0.1s^2 + 3s + 1}e^{-1.3s} \end{pmatrix} \tag{5}$$

To evaluate the model response of the HCDP, the flow set point of *FIC-5* stepped down from 1.2 to 1.7 *MMSCMD* at sample time 240. It is expected that at a time when the *IGV* decreased the angle opening in order to decrease the load demand through the turbo expander, there will be a slight reduction in pressure and then a small overshoot as expected due to the load reduction. The delay of pressure stream fluctuation is due to the fact that the pressure sensor *PIC-2* is physically located in the downstream of the condensate flush drum while the *IGV* is located in the inlet of the expander.

The HCDP model responses as shown in figure 9 react to step disturbances on gas flow rate exactly as expected of a real hydrocarbon dew-pointing process. The results obtained by applying a multivariable Generalised Predictive Controller (GPC) with a prediction horizon of 23, and a control horizon of 3.
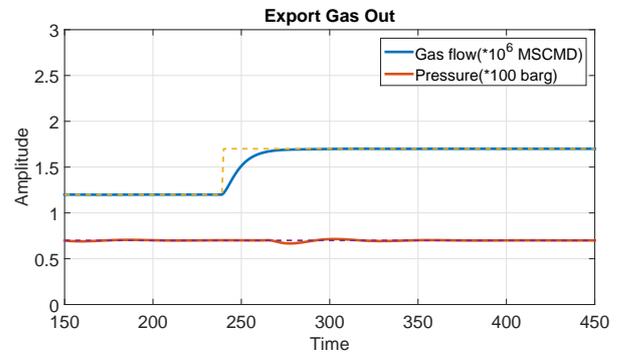

Fig. 4. HCDP Gas outlet responses with a step disturbance in gas flow under GPC control

4. MODEL VALIDATION

Model validation and verification is the most important step in the model building sequence. The ultimate goals of creating a model representing the gas phase train in upstream oil & gas fields are to aid decision making and to provide engineering solutions to operational problems. The obtained models need to accurately reflect the real process scenarios which they will be compared to. Nevertheless, the developed models of $G_{GSU}, G_{GDU}, G_{HCDP}$ are destined to represent general processes and dynamics and not specific units; therefore the models can be validated by graphical comparisons and descriptions of model outputs with data from commonly available industrial processes (Jain et al. (2011)).

4.1 Gas Sweetening Unit (GSU) Model Validation

Figure 5, below captures the simulation results of the GSU model stimulated by a step increment of almost 33% in the throughput gas flow. The results obtained by applying a multivariable Generalised Predictive Controller (GPC) with a prediction horizon of 30, and a control horizon of 5.

The model outcome is then compared with a real GSU response for almost the same size step increment taken from PDO Harweel site in Oman shown in figure 6. It is noticeable from the trends that the model responds -of the two concerned signals, the gas flow rate and the H_2S concentration- almost in a similar manner to a real gas sweetening unit.
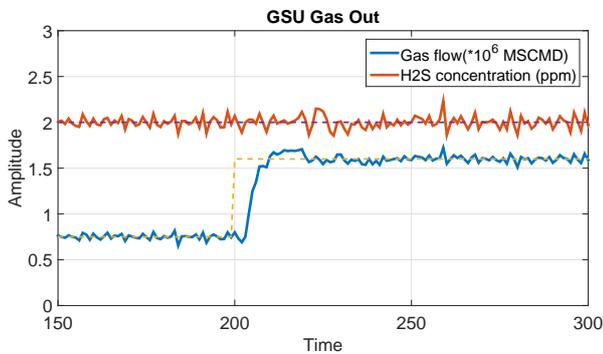

Fig. 5. Model: GSU Gas outlet responses with a set point increment of 33% in gas flow rate

4.2 Gas Dehydration Unit (GDU) Model Validation

Figure 7 trends the simulation results of the GDU model stimulated by a step increment of almost 30% in the throughput gas flow. The results obtained by applying a multivariable Generalised Predictive Controller (GPC) with a prediction horizon of 30 and a control horizon of 5.

The model outcome is then compared with a real GDU response for almost the same size step increment taken from PDO Harweel site in Oman shown in figure 8. It is visible from the figures that the model responds similarly as a real gas dehydration unit when comparing the two concerned signals, the gas flow rate and the water vapour load in the gas.
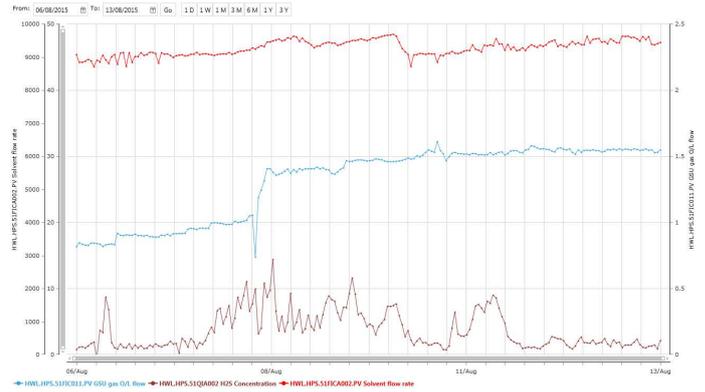

Fig. 6. PDO Harweel GSU Gas outlet responses with a set point increment of 33% in gas flow rate. Blue: Gas flow rate (*Range 0 - 2.5 MMSCMD*), Brown: H_2S concentration (*Range 0 - 50 ppm*), Red: Solvent flow rate (*Range 0 - 10000 m^3/d*)
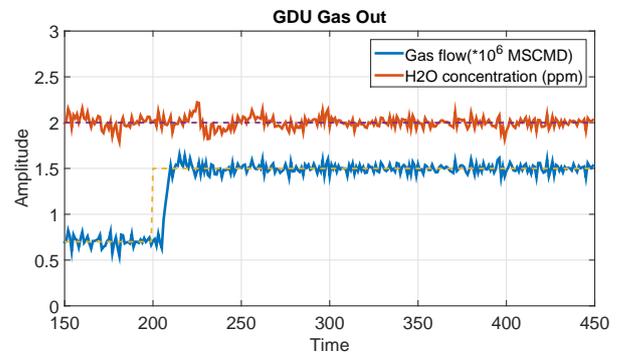

Fig. 7. Model: GDU Gas outlet responses with a set point increment of 30% in gas flow rate
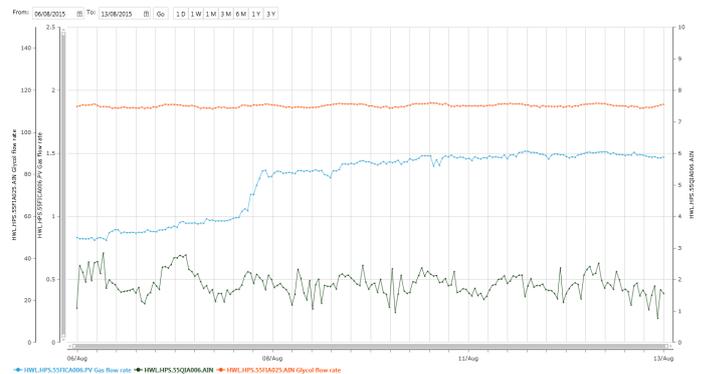

Fig. 8. PDO Harweel GDU Gas outlet responses with a set point increment of 30% in gas flow rate. Blue: Gas flow rate (*Range 0 - 2.5 MMSCMD*), Green: H_2O concentration (*Range 0 - 10 ppm*), Red: Glycol flow rate (*Range 0 - 140 m^3/d*)

4.3 Hydrocarbon Dew-pointing Unit (HCDP) Model Validation

The HCDP model responded as shown in figure 9 to a step disturbance on gas flow rate exactly as expected of a real hydrocarbon dew-pointing process when compared with a real HCDP process response for almost the same size step increment taken from PDO Harweel site in Oman shown in figure 10.

The model results obtained by applying a multivariable Generalised Predictive Controller (GPC) with a prediction horizon of 23, and a control horizon of 3.
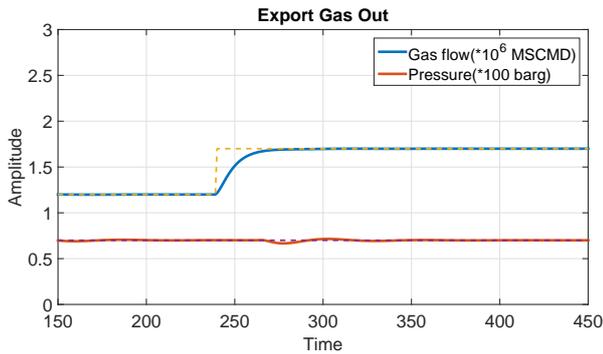

Fig. 9. Model: HCDP Gas outlet responses with a step disturbance in gas flow rate
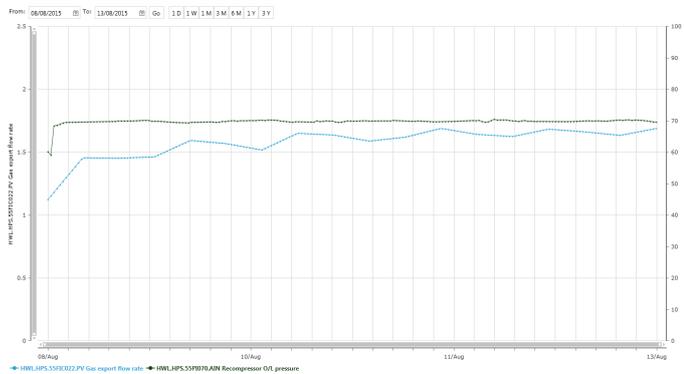

Fig. 10. PDO Harweel HCDP Gas outlet responses with a step disturbance in gas flow rate. Blue: Gas flow rate (*Range 0 - 2.5 MMSCMD*), Green: Gas pressure (*Range 0 - 100 barg*)

5. CONCLUSIONS AND FUTURE CHALLENGES

Large scale series processes are rather common in the upstream oil & gas industry. Consequently, representative models are a key demand for control and automation engineers to test and verify different control approaches and strategies. The intention of this 'PART I' paper is to deliver simple and easy to understand process models based on transfer functions for a complex gas processing operations. Processes like gas sweetening and gas dehydration are deemed as difficult control tasks for both process and control engineers. Henceforth the presented model is aimed to ease these control challenges by providing an authentic framework for engineers to design, analyse and evaluate different control solutions. The model provides good opportunity for control engineers to test different variety of process disturbances, malfunctions, and load changes on the process operation and verify its significance in order to design a precise control system. Control system thats able to solve a major control challenge problem the disturbance growth effecting the series connected processes in LSS.

'PART II' paper will investigate the use of this models as a base to test the control concept of "Integration of Distributed Model Predictive Control (DMPC) and the plant safe guarding system with the classical control system" presented by AlNaumani and Rossiter (2015). These models will be used to develop a specific Model Based Predictive Controller (MBPC) which is expected to significantly reduce the plant shut down occasions due to process disturbances and also to drop down operating cost by properly controlling the disturbance growth in the process, hence reducing energy fluctuations in the process and saves fuel. Accordingly the process element fatigues will be reduced which increases the process safety measure. As a result, the process operators work load will be decreased.

REFERENCES

AlNaumani, Y. and Rossiter, J. (2015). Distributed mpc for upstream oil & gas fields - a practical view. In *9th International Symposium on Advanced Control of Chemical Processes*, 325–330. IFAC.

Baker, R.W. and Lokhandwala, K. (2008). Natural gas processing with membranes: an overview. *Industrial & Engineering Chemistry Research*, 47(7), 2109–2121.

Camacho, E.F. et al. (2007). Model predictive controllers. In *Model Predictive control*. Springer.

Downs, J.J. and Vogel, E.F. (1993). A plant-wide industrial process control problem. *Computers & chemical engineering*, 17(3), 245–255.

Forbes, M.G., Patwardhan, R., Hamadah, H., and Gopaluni, B. (2015). Model predictive control in industry: Challenges and opportunities. In *9th International Symposium on Advanced Control of Chemical Processes*, 532–539. IFAC.

Jain, S., Creasey, R., Himmelspach, J., White, K., and Fu, M. (2011). Verification and validation of simulation models.

Kidnay, A.J., Parrish, W.R., and McCartney, D.G. (2011). *Fundamentals of natural gas processing*, volume 218. CRC Press.

Kvenvolden, K.A. and Lorenson, T.D. (2001). *The global occurrence of natural gas hydrate*. Wiley Online Library.

Ohshiro, T. and Izumi, Y. (1999). Microbial desulfurization of organic sulfur compounds in petroleum. *Bioscience, biotechnology, and biochemistry*, 63(1), 1–9.

Prett, D. and Morari, M. (1987). Shell process control workshopbutterworths. *Stoneham, Mass.*

Rossiter, J.A. (2013). *Model-based predictive control: a practical approach*. CRC press.

Sloan, E.D. (2003). Fundamental principles and applications of natural gas hydrates. *Nature*, 426(6964), 353–363.