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# Distributed MPC for Upstream Oil & Gas Fields - a practical view

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**Abstract:**
This work aims to improve corporate functional departments' confidence in adopting modern control approaches in new scenarios and thus presents control structure solutions based on MPC for two control problems facing existing upstream oil & gas production plants; these are the disturbance growth in the series connected process and the control system dependency on operators. The proposed approach integrates distributed MPC (DMPC) as a master controller for the existing classical control of each subsystem, with a focus on those with high interaction phenomena. The proposed DMPC also considers safeguarding, constraints and the enhancement of plant-wide optimal performance. The suggested control solution reduces the role of control room operators which is shown to reduce the growth in the impact of process disturbances. Compared with some alternative control structures (centralised MPC, decentralised MPC, distributed MPC (DMPC), and hierarchical DMPC) this proposal is simple, inexpensive to implement, and critically, builds on the local team operational experience and maintenance skills.

*Keywords:* Distributed model predictive control; Process Control; Upstream Oil & Gas

## 1. INTRODUCTION

Control system application has continued to improve since the discovery of the first commercial oil well in the mid-nineteenth century (Habashi (2000)). The growth in demand for upstream hydrocarbon gathering and production plants stimulated the need for changes from manual control systems to fully automated ones. Today the majority of the upstream production plants still mainly utilise classical Proportional-Integral-Derivative (PID) control laws to regulate process variables; indeed these are largely sufficient for oil & gas production plants and will certainly continue to play an important role in the process industries. PID control is robust and transparent, but its main weakness is in being Single Input Single Output (SISO), thus giving a decentralised process control system. The risk here arises from the lack of coordination between controllers because each controller has to cope alone in meeting its objectives (except in the cases where a cascade approach is applied).

On the other side of the petroleum industry, PID control architectures were clearly an obstacle to optimal operation of refinery processes. The majority of the control loops in refineries and power generation plants are Multi Input Multi Output (MIMO) control loops where each controller output not only regulates a particular process variable but also effects other process variables within the system. Poorly tuned interacting controllers severely limit the best achievable closed loop performance and thus incur extra operational costs (Christofides et al. (2013)). Despite the vast array of tuning tools (Seborg et al. (2010); Romagnoli and Palazoglu (2012)), tuning MIMO PID controllers is still difficult and may not give good solutions (Johansson et al. (1998)). Based on this rationale, Model based Predictive Control (MPC) was developed as a systematic multivariable control scheme for refineries and power plants. MPC relies on an explicit mathematical model of the process to predict the future response of the plant and the definition of a cost function (measure of performance). MPC computes the sequence of optimal future control actions (inputs) over a specified future time horizon in order to optimise the expected performance of the system. The first input is then used by the plant, and the rest of the sequence is discarded repeating the procedure at subsequent control intervals. At each instant the horizon is displaced towards the future (Camacho et al. (2007)). Due to its ability to deal with process constraints, multivariable and/or complex dynamics systems, MPC has become a standard approach and it is popularity in the chemical process industries has increased steadily.

The globally rising demands for fossil fuel leads to over-consumption of the valuable resources alongside a deficiency in the discovery of new reservoirs. Therefore there is a necessity to optimise the production operation of the current assets. Unfortunately the control difficulties

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of the upstream oil & gas fields have not received the same research attention as the downstream processes. The two major control issues affecting the current upstream production plants are:

- The disturbance growth in the series connected process.
- The control system dependency on operators.

The intention of this paper is to target these control issues and to provide an inexpensive feasible control concept based on MPC for the benefit of existing upstream oil & gas plants. As per the authors' knowledge, no specific study is reported so far to tackle these control issues and to provide a friendly inexpensive upgrade based on MPC to the **existing** upstream oil & gas plant and its classical control system.

This paper brings much needed attention to the control challenges facing upstream oil & gas production plants, especially for existing (not new) plant and discusses different solutions to handle the challenges; the prime focus is on improving disturbance rejection and the potential for integrating MPC (cheaply) for optimal plant operation. The control challenges are illustrated in section 2. Section 3 presents a discussion of the current available solutions and the practicality of retrofitting existing plant, while section 4 provides a feasible solution to the control challenges. Section 5 recommends future work and gives some conclusions.

## 2. PROBLEM FORMULATION

There is an obvious importance for upstream oil & gas companies in confronting the control weaknesses in order to cope with an increasing level of process complexity, demanding product specifications, profitability, safety, and environmental sustainability challenges and of course, to ensure they meet production revenue targets. Upstream oil & gas plants constitute a number of processes (the number and complexity depends on crude type) connected in series and physically distributed over a wide area. Fig. 1 illustrates the most common process. In addition to the main processes, plants also contain utility processes like the 'instrument air system' and the 'produced water treatment facilities' etc.

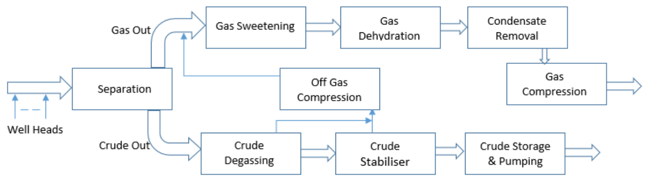


Fig. 1. Upstream Oil and Gas main Process

For safe and stable operation, plants rely on hundreds of PID control loops driven by a Distributed Control System (DCS). Since the control structure is too basic to act protectively in advance, companies usually dedicate a number of staff to work as control room operators. Their main task is to monitor the process deviation and amend the controllers' reference values to achieve safe and profitable optimal plant operation. The plant control optimisation and problem solution are totally dependent on the respective operators' efficiency and significantly, also on their speed of observation at the time a process deviates from one operation scenario to another. Human operators have a habit of operating within their comfort zone and their decisions can be exaggerated by the control room environment and the sudden assigned responsibilities and commitments. Bello and Colombari (1980) provides a detailed discussion about the risks caused by the control room operators of process plants.

Feed disturbance and equipment failure are the two common causes of major process disturbances in upstream production plants. Such disturbances have the potential to cause significant deviation of the process and potentially cause violation of operation constraints. In series connected systems where one process output is the feed to the successor process, the effect of disturbances can be magnified due to system gain. If an extraordinary or more than one antagonistic condition develops at the same time, the operator may not be able to react satisfactorily and the consequences are a larger risk of a major disturbance event.

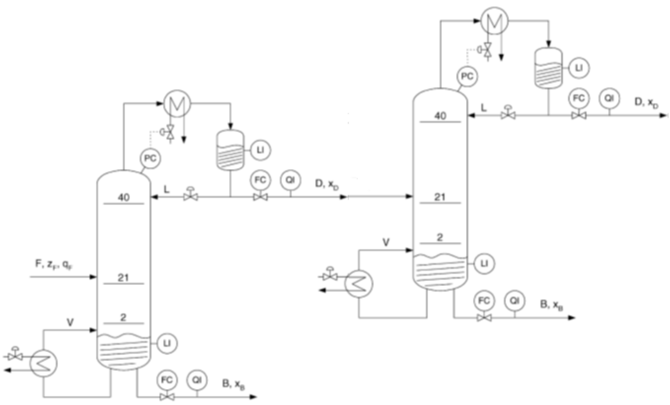


Fig. 2. Two Columns Process

To illustrate the impact of the disturbances on series connected LSS (Large Scale Systems), consider the two column process shown in Fig. 2. Feed enters the first column and the overhead distillate flow is connected as inlet feed to the second column. In this example, the aim is to maintain the overhead composition on both columns at "0.9 Molfrac" (to aid disturbance comparison); continuous measurement by process analysers is available. Distillation column dynamics are presented in Muske and Badgwell (2002). A disturbance of "-5 %" is introduced on the first column feed as illustrated in Fig. 3. The disturbance's effects on the overhead composition of both columns are presented in Fig. 4 and Fig. 5. While the overhead composition of the first column is only effected to a small extent as expected, the impact of the disturbance on the second column product was substantial and indeed caused a violation of the desirable/required operating conditions.

In summary, poor coordination between the controllers for successor processes (here a series connected LSS) means that constraints and safeguarding limits are more likely to be violated. However, this issue has received relatively little attention in the literature.

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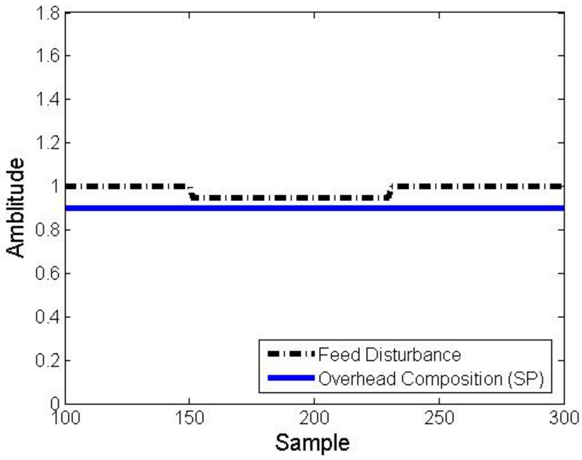


Fig. 3. References and Disturbances

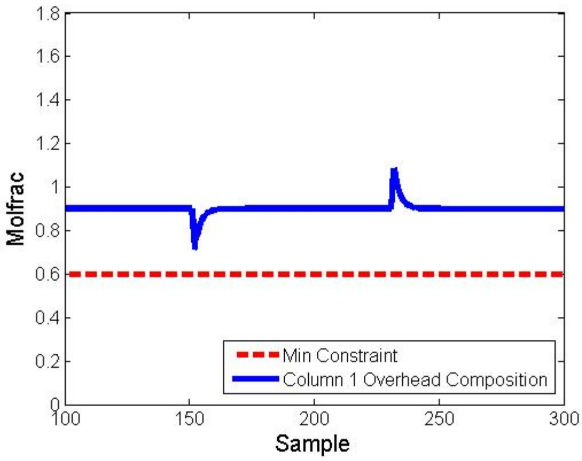


Fig. 4. First Process Output

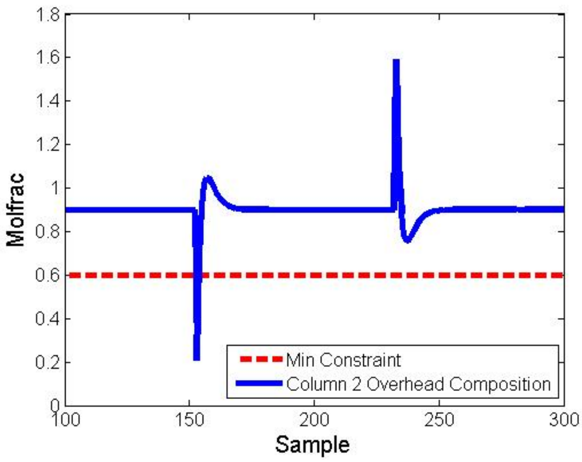


Fig. 5. Second Process Output

## 3. CURRENTLY AVAILABLE SOLUTIONS

There are four main control system structures based on MPC algorithms which are successfully implemented in the industry. These are centralised MPC, decentralised MPC, distributed MPC (DMPC), and hierarchical DMPC. They differ in the implementation structures but all of them apply a receding horizon strategy and employ a model of the process to obtain the control output as the optimum solution of an associated cost function minimisation. An important question for process operators is to determine which structure best suits their plant requirements and moreover, fulfils current and future commitments? The best overall control structure will depend upon typical control objectives, possible process disturbances, all constraints, robustness obstacles (Vogel and Downs (2002)), and of course costs of retrofitting and staff education.

### 3.1 Centralised MPC

A control structure is considered to be centralised when the complete plant-wide process is modelled and all control inputs are computed in one controller. In other words all plant-wide interactions are dealt with in a single optimisation problem as illustrated in Fig. 6.

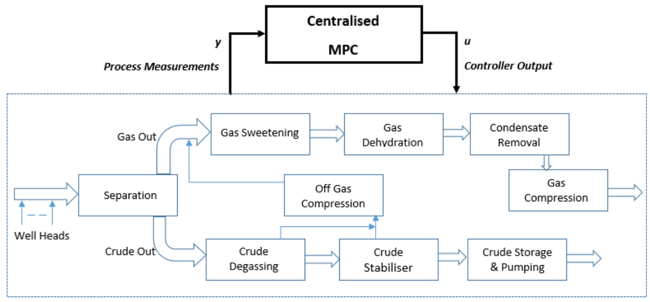


Fig. 6. Centralised MPC

In the last decade, most DCS vendors have upgraded their systems capability to handle predictive control benefiting from the substantial advances in computational power. Evolution in electronics engineering, specifically the memory and processor microchips enhances the development of faster optimisation software, higher speed communications, and extra powerful computers. Consequently and with a precisely designed control algorithms for large scale system (LSS), the adoption of a centralised control structure may seem to be a reasonable choice (Stewart et al. (2010); Pannocchia et al. (2007)).

Weaknesses of a centralised control structure are mainly related to system complexity, speed of control, and organisational issues. Development of plant-wide interaction model either by mathematical modelling or by utilising system identification methods is a complex task. Major modelling difficulties are due to the addition of unmeasured disturbances and system uncertainties in each subsystem. The developed model should be as representative as possible to the plant, otherwise the MPC controller may fail to stabilise the plant or even to give sensible control strategies. In addition to the complexity issue, the new control loops should execute at a higher sampling rate or at least equivalent to the current classical control. Current DCS in upstream fields executes sampling at sup-second to one second (Darby and Nikolaou (2012)). Notwithstanding the evolution in the computers computational power and microchips processers, a typical DCS is not utilised for superior control performance only but also to do other operational tasks like alarm management, history records,

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high resolution graphical interface, etc . Accordingly, the computational time needed to solve the centralised control problem may be significantly prolonged which in turn hinders the MPC ability to perform real time calculations (Christofides et al. (2013)). Furthermore, Stewart et al. (2010) noticed the organisational objections to the implementation of centralised MPC for LSS plants. Maintenance and troubleshooting of a mega dimension and complex central controller is a tricky practice and will consume a lot of valuable efforts and time. In simple terms, the potential improvements in coordinated behaviour and performance are unlikely to be realised in practice.

### *3.2 Decentralised MPC*

A decentralised control structure is the most common control framework implemented in industry for LSS (Scattolini (2009)). In a decentralised architecture (presented in Fig. 7) each subsystem control is locally centralised by means of one or more non-cooperative controllers depending on the subsystem complexity. Each controller focuses on its own local optimisation problem only and doesn't exchange information with other controllers.

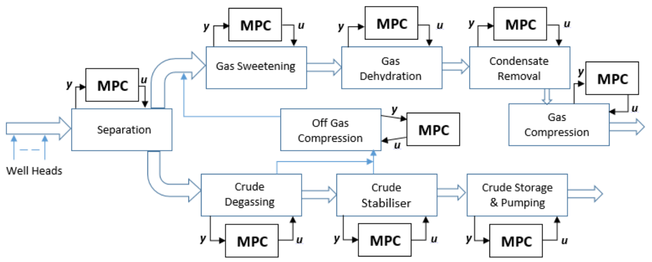


Fig. 7. Decentralised MPC

Unlike the centralised control structure, a decentralised structure is far easier to design and maintain as well as the real time implementation is not an issue. Nevertheless, since there is no information exchange between subsystem controllers, the decentralised structure can't optimise the plant-wide control problem and thus could result in poorer performance. Decentralised control systems are successful for LSS which have weak interaction between subsystems, for example where these interactions can be considered as disturbances which can be compensated through feedback (Christofides et al. (2013)). A decentralised structure is not recommended for LSS with strong interconnections between the subsystems due to stability concerns and optimum performance achievements.

A key message of this paper is to note that many existing oil & gas production plants utilise decentralised control system structure underpinned with PID controllers. Hence the potential for implementing a decentralised control based on MPC is straightforward in principle and companies may achieve a better optimised subsystem operational control, However, the expenditures on training and building up the operational and technical expertise and demonstrating the potential benefits are key obstacles.

### *3.3 Distributed MPC*

A distributed MPC (DMPC) control structure (shown in Fig. 8) is relatively similar to the decentralised structure except that the local controllers (agents) exchange information and communicate cooperatively among themselves to solve the overall plant-wide control problem (Negenborn and Maestre (2014)). A distributed MPC control structure reduces the overall achievable performance limitations associated with a decentralised structure. In a distributed structure each controller espouses the interaction between the subsystems with the local control objectives and constraints to optimise the local control problem. Sometimes the controllers are forced to sacrifice their own control objectives in order to achieve the required plant-wide performance. The controllers' communication load and decisions on with whom to communicate, are dependent on the level of interaction between the subsystems and the status of the communication network. Controllers can be constructed to communicate information like their next control move with the neighbouring agents or specific agents or even with all agents in the system.

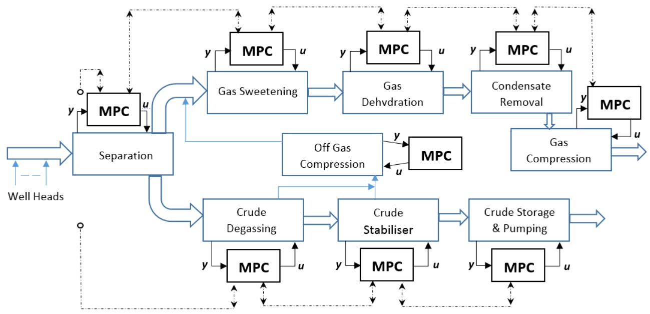


Fig. 8. Distributed MPC

Although co-operation between agents to solve a global optimisation problem is clearly a sensible proposal, nevertheless co-operation in some cases may lead to a poor local control behaviour and consequently deterioration in the plant-wide control performance. Negenborn and Maestre (2014) surveyed a number of different DMPC approaches and theories which designed to foster co-operation based on process, theoretical, and control architecture commonalities. One of the findings was that out of thirty five DMPC schemes, only one was designed for transfer function models. It is worth mentioning here that the majority of oil & gas processes are described with transfer function models. Also, the survey suggested the need for researchers to develop flexible DMPC architectures able to modify the control network topology and the communication burden depending on the circumstances.

Practically speaking a DMPC structure is recommended for any new oil & gas plants (greenfield) but it is rather costly to retro-fit on existing plants (brownfield). Moreover, the required operation and maintenance skills might take long time to build among the team which may affect the company's confidence in the efficacy of introducing a new control architecture.

### *3.4 Hierarchical Distributed MPC*

A hierarchical DMPC system is structured from two or more control layers which coordinate among themselves to control the process. As presented in Fig. 9, the higher layer receives system wide information to perform the real

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time optimisation and manage the global objective of the process and provide reference signals for the agents in the lower control layer which cooperatively control and regulate the plant control elements. Dividing the overall control system structure into layers helps to ease the control problem and to speed up the control cycles in a Large Scale Systems. The fast system dynamics are being controlled by the faster lower control loops referencing to the latest set-points provided by the higher control layer and without waiting for the real optimisation problem solution.

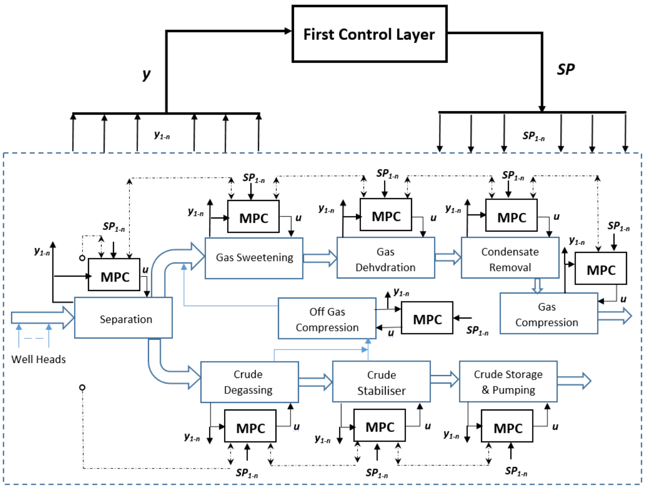


Fig. 9. Hierarchical DMPC

Due to the complexity of LSS, there exists number of hierarchical control structures in the process industries. Each of these structures are tailored for controlling particular classes of processes. For example Scattolini (2009) reviewed four main hierarchical control architectures. These are, the hierarchical control for coordination where an algorithm at the higher control level coordinates the actions of local regulators placed at the lower control level. The hierarchical control of multi-time scale systems to control systems with slow and fast dynamics. The hierarchical of cascade control structure and the hierarchical control for plant-wide optimisation.

Even though a plant-wide control strategy will be enhanced by a hierarchical DMPC control structure, the costs of implementing it in brownfield processes to replace existing classical control will be too expensive. Also the new structure may be unwelcome by the operation team due to the same reasons as for the Distributed MPC control structure discussed earlier.

## 4. A FEASIBLE SOLUTION FOR EXISTING OIL AND GAS FIELDS

The previous section had demonstrated that while there are many proposals in the literature, and indeed already being used in practice, these are far more likely to be feasible for a greenfield project but not necessarily for brownfield. The feasibility of retro-fitting a new control structure is influenced by factors like project cost, system simplicity, process safety, running cost, and anticipated gains compared with the existing control system. Critically, from an operational standpoint, the feasible control solution to enhance the current classical control system in the existing oil & gas plants must also inexpensively integrate the **team experience and operational knowledge within it**. Consequently, this section proposes what is considered to be a more pragmatic alternative.

The proposed control system is sketched in Fig. 10; this integrates distributed MPC (DMPC) as a master controller in the existing classical control of each subsystem. The DMPC receives system measurements from the process sensors to compute the subsystem optimal control actions and provide local control goals as set-points (SP) for the critical PID controllers only (high interaction control loops). The DMPC also receives system units status from the process safeguarding system to dynamically update the system constraints. However, a key point is that the DMPC shares information like the current performance factor and the next control move with its neighbour controllers to enhance the plant-wide optimal performance; this communication can help with disturbance rejection.

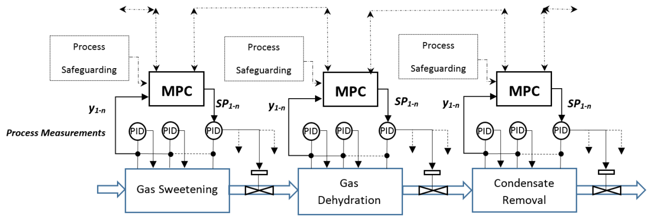


Fig. 10. Integration of DMPC with classical control

The proposed control system is designed on a cascade strategy and thus provides a flexible system control almost like a decentralised structure in dealing with disturbances and unit failures, and at the same time improves the closed loop performance and the plant-wide optimal operation. The DMPC is designed to regulate the critical loops only while the rest of the uncritical PID loops will continue to function in a decentralised fashion; this minimises any design and set up costs, reduces demand on the communication network and simplifies the any associated real time optimisations. The improved local control will reduce the need for control room operator interactions with their associated weaknesses. The one way communication from the process safeguarding enables prompt response to disturbances caused by unit failures while the bidirectional communications with adjacent MPC's in effect enables feed-forward to reduce the impact of process disturbances and enhance optimality.

Fig 10. presents control schematic of three main systems of a gas processing train connected in series. Each of these systems constitutes of number of units like pumps, vessels, contactor columns, and automated isolation and control valves. Depending on traditional control approaches only, the system functionality deteriorates notably when one of these units fails leading to system instability and, in the worst case, process shut down. However, the scenario is totally different with the integration of the DMPC into the control system. When a unit fails to perform to specification, the relevant DMPC will immediately know about it from the safeguarding system before the consequences take effect. Consequently the DMPC updates the system constraints and informs the predecessor and successor system controllers about the new limitations to modify the

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throughput product harmonically. The scenario is more or less similar with feed disturbances. Therefore the proposed control system is expected to reduce process shutdown occasions and to extend the fixing time provided for the maintenance crew.

Compared with the solutions discussed in section three, the proposed control solution is much cheaper and simpler to implement. The DMPC system model is quite easy to develop as well as the control algorithms. Nevertheless it almost delivers the same benefits and does not omit the team operational experience and maintenance skills. In addition its performance can be straightforwardly validated in the DCS by altering the cascade mode between auto and manual.

## 5. CONCLUSIONS AND FUTURE CHALLENGES

A prime contribution of this paper is to identify pragmatic approaches for control system improvements, that is approaches which will be attractive to companies and process operators. The approach should require as little retrofitting as possible, that is to build on existing infrastructure and expertise as much as possible as this reduces cost, training requirements and simplifies validation. Moreover, by utilising the existing structures, system testing and performance compared to the classical control can be easily validated by switching the DMPC cascade mode in the DCS to manual and comparing the DMPC output trends against operators manual set-points. In addition the implementation of the new control strategy will take place in the instrument auxiliary room and will not disturb the field arrangements by any means. Consequently, the proposal of this is expected to be straight forward to implement and test.

Nevertheless, two key obstacles were identified as research challenges in order to progress this theme, to produce stronger evidence, and thus for improving the control of upstream oil & gas plant. These are:

- *Upstream oil & gas process model.* 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 also be of benefit to Large Scale System (LSS) and system interactions control research fields.
- *DMPC control signal segregation.* There are number of suggestions provided by the literature for how to segregate DMPC control signals based on a system's dynamic behaviour (Jogwar et al. (2009)) and physical structure (Motee and Sayyar-Rodsari (2003); Al-Gherwi et al. (2010)) but DMPC control signal partitioning and loop cascading are still a grey area even though much research attention has focused on the DMPC strategies during the last decade. It is difficult to provide a general answer for questions like; how many loops each MPC can manipulate optimally and how to segregate control signals for each loop in multi cascaded systems?

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