The impact of process design decisions on operability and control of an LNG process

Finn Are Michelsen[a 1], Ivar Johan Halvorsen[a] and Berit Floor Lund[a]

[a]*SINTEF ICT, Department of Applied Cybernetics, N-7465 Trondheim, Norway*

Abstract

This paper describes a framework for integrated process and control structure design, and applies this framework to a Liquefied Natural Gas (LNG) process design. The overall aim of the work is to contribute to the methodological basis for improved design and operability of gas processing plants. Good operability means essentially that a plant can be operated easily, i.e. that it can cope with unknown disturbances, offsets and other uncertainties with the smallest possible profit loss and without frequent shut-downs. This is obtained both through the design of the process itself and the design of the control system. There is a potential for improved operability of process plants, and thereby reduced profit loss, by considering these two aspects together. The main message is that this is handled by considering control structure design when process design changes are made.

The main steps for developing an improved procedure for integrated process and control design are suggested. Such a procedure includes analyses of how altering key parameters in the process design affects the best possible control structure in the presence of defined disturbance scenarios. An important ingredient is the use of a dynamic, control relevant simulation model. The paper presents an analysis of how altering compressor size affects the choice of control structure for the Tealarc LNG process. This analysis has been carried out using a self-optimizing control methodology.

Keywords: process design, process control, optimization, control structure, natural gas processing, LNG
[2]

1 Introduction

Higher profit requirements force new processing plants to operate the most expensive units close to their constraints. For a Liquefied Natural Gas (LNG) plant, this typically implies minimizing volumes and compressor sizes. However, altering the process topology or design of equipment components also has an impact on the behavior of the plant. This means that in the process design phase of new gas processing plants there is a potential economic benefit of also considering operational issues like controllability, stability and robustness of the controlled plant. Hence, the design procedure should include the possibility to analyze how process design decisions impact on process behavior and control structure design. Careful choice of control structure can improve profitability of the new plant in normal operation, but also reduce the risk of operational problems, losses and redesign when a new process is first started up. This paper addresses the steps that should be included in an integrated process and control design procedure.

---

[1] Corresponding author. Email: Finn.A.Michelsen@sintef.no, Phone:+47 73594397, Telefax:+47 73594399

[2] Abbreviations: CV - controlled variables, DV – disturbance variables, LNG – liquefied natural gas, MV - manipulated variables, NG – natural gas, PID - proportional-integral-derivative, Pr – compressor pressure ratio

Thereby, the design of both the process equipment and the control system is coordinated in a systematic manner. An objective should be to use such an integrated method as the mainstream design procedure for industrial plants (Seferlis and Georgiadis (2004)). A literature overview shows that the research area is somewhat immature, still there are a number of publications addressing the topic. Many of the papers propose detailed procedures for integrated process equipment and control system design. The high level of detail makes reuse of the procedures for different, or even only slightly different processes difficult. The reason for this is that different applications have different requirements to what issues that are important with respect to the specification of process units and operability. This paper therefore outlines a more general and reusable framework for integrated process and control design. When applied to a specific process, the framework should be further detailed and expanded with various design and operability analysis methods which are relevant for the specific case.

The framework has been applied to the TEALARC LNG process. However, as a full analysis of the entire process would be too comprehensive for a paper, the impact of compressor sizing on the choice of control structure has been chosen as an illustrative example. This particular analysis includes self-optimizing control methodology for operability analysis.This control methodology aims at finding the set of controlled variables which gives the smallest steady state profit loss, despite changes in unknown disturbance variables and implementation errors at a nominal operating point.

Operability is defined as the ability (goodness) of a system to be operated as required. Operability analyses require models which are input-output causal and have good convergence qualities. Moreover, such analysis takes advantage of dynamic, control relevant models which are, computationally light, and therefore suitable for comprehensive optimization calculations (Foss and Halvorsen (2009)). Often, design analyses are made based on a comprehensive steady state simulator model implemented in e.g. commercial process simulation tools like HYSYS. When such a model is available, it is advantageous to adapt unit models (e.g. compressor and expansion valve models) in the control relevant model to the design model in a nominal design point.

In the following, section 2 gives a literature overview regarding integrated process and control system design. Section 3 gives a description of the TEALARC LNG process. Section 4 describes the steps that are proposed for integrated process and control design procedures. A demonstration of how these steps can be applied for design of the TEALARC LNG process is discussed in section 4. A discussion is given in section 5 and conclusions are given in section 6.

2 Literature overview of integrated process and control design methods

Process design (steady state considerations) and control design (dynamic considerations) was traditionally handled sequentially by first designing the process, and later designing the control system. One reason for this was that process design engineers and control engineers worked separately. Control design means control structure design, controller design and controller tuning. The attention to achieve the 'best' dynamic performance under closed loop control, i.e. to maintain the specified controlled variable at a setpoint value was then ignored during the design phase of the processing system. The process design engineer worked on the selection of the best process flow-sheet with the minimum capital and operating cost based on steady state considerations. Then control engineers optimized the dynamic performance to implement the best control for a given design (Chawankul et al. (2005)). An improved concept is to integrate these steps in an iterative sequence including evaluation. Hence, this approach is also denoted as a sequential process and control design procedure. A reason to this improvement is that a process which is optimal at steady-state may not be optimal in light of the changes it may face (Schweiger and Floudas (1997)). Ziegler and Nichols (1943), in their frequently quoted work, gave early a good elaboration of this important issue. They pointed out that the process design problem neglects the fact that the dynamic controllability is an inherent property of its design. Hence, design changes can remove operability problems without leading to cost increases. A much referenced example is given by Anderson (1966), where a plant shut down and equipment replacement could have been avoided if the dynamic aspects of the operation had been considered during the design phase. The first ideas for systematically integrate these ideas in process design was made by Nishida and Ichikawa (1975) (Morari and Perkins (1994)). A further development of the integrated design procedures has lead to a simultaneous approach. This solves the process and control design problem in one simultaneous step. Thereby both steady state and dynamic operability are considered in the joint design of process and control. These methods include mixed integer dynamic optimization where explicit consideration of structural process and control design aspects (such as number of trays in distillation columns, pairing of manipulated and controlled variables) is formulated by binary variables (see e.g. Mohideen et al. (1996; Sakizlis et al. (2004)). Georgiadis et al. (2002) argued that this approach leads to significant economic benefits and improved dynamic performance during plant operation as compared to the sequential approach. However, although optimization provides a possible solution to large problems, to find a solution especially for non-convex problems is still a challenge. This may be a larger problem for the simultaneous approach as compared to the sequential approach. Another advantage of the sequential approach is that it improves the possibility to learn about the process during the design process. Extensive research during the past decades has resulted in a huge amount of new methods for integrated process and control design (see e.g. Morari and Perkins (1994), Georgakis et al. (2003) Seferlis and Georgiadis (2004)).

3 A framework for integrated process and control design

Two important factors that determine the steady state optimal performance and the optimal closed loop behavior of process plants are the disturbances that act on the plant and the constraints in the plant equipment. This means that the constraints have to match the impact of the disturbances. Hence, a disturbance sensitivity evaluation has to be a central part of an integrated process and control design procedure. Further, such a procedure should take advantage of a dynamic, control relevant model for optimization calculations. This can be used to screen out design options and analyze the implications for the operability of the plant. Thus, feedback from the control system design phase to the process design phase can be made early in the design project.

Based on these considerations, an integrated process and control design procedure should include the following steps:

1. Make a process design including types of process units, their characteristics and how they are organized in a process structure. At this step, the operational constraints for the plant are identified, i.e. sizing of equipment etc.

2. Define the process design objectives and the operational objectives for the plant. These constitute a set of multi-objective objective functions to be minimized. An example is minimum energy consumption, which may result from maximum energy integration in the plant. Another example is the negative profit of the operation of the plant, i.e. total costs minus total income. The annual costs related to investments, man hours, environmental load, safety and security issues, and taxes etc. are other examples. An operational objective, which is an operability measure, should describe how well the plant is controlled.

3. Define the disturbance scenarios, i.e. what disturbances that affect the plant and the size and direction of these disturbances over time. The best strategy to handle disturbances may be to avoid them. Hence, at this step investigate the sources for the disturbances and evaluate whether it is possible to easily avoid any of them. Disturbances may be known

(i.e. available) either by direct measurements or estimation. In this case, they can partly be handled by feed-forward control when the effect of the disturbance on the manipulated variables (MVs) is known by a model. Otherwise, they are unknown, and have to be rejected by feedback control. MVs are control variables as the output from the controller, which is available for manipulations. Examples are valve command and compressor speed.

4. Build a dynamic, control relevant model of the plant. This model should be input-output causal and computationally light.

5. Based on the steps above, calculate the steady state optimal performance. This includes the minimum value of the process design and operational objective functions, but may also include other steady state performance measures such as the steady state operational range for certain process variables.

6. Based on the steps above, define a control structure and design the controllers. The latter includes the structure of the controller (e.g. linear proportional-integral-derivative (PID), which is the most applied controller structure in the process industries) and tuning of controller parameters.

7. Based on the steps above, calculate the optimal closed loop behavior of the process plant, i.e. the operability measures. These are given by the control performance and stability requirements for the operation of the plant. This step is also a validation of the total design.

8. Re-do from any of the steps above dependent on whether it is desirable and possible to either change the

    - process design and/or operational objective function
    - process constraints (process design)
    - process structure design (e.g. related to energy integration)
    - controller structure and/or controller design

    Another option is again to investigate the sources for the disturbances and evaluate whether it is possible to avoid any of them in order to decide whether this is cost effective. If none of these options are relevant, the resulting design is either not feasible or must be accepted.

The sequence of steps above can be regarded as an extension of the plant wide control design procedure by Skogestad (2004), which is a sequential iterative control design procedure. Thus, the sequence of these steps may be used to develop an improved sequential iterative design procedure. However, the implementation of the steps does not necessarily have to follow a sequential iterative approach in the sequence they are listed above. Hence, there are several ways of refining and structuring this integrated process and control design procedure by including various design and operability analysis methods from the literature. Some of them are mentioned in section 1, constituting either a sequential iterative or a simultaneous design procedure.

4 Application of the procedure for design of the TEALARC LNG process

This section first gives a short introduction to the Tealarc LNG process. The second section shows how the framework described in section 3 may be applied to the TEALARC LNG process. Since a complete design would include a large number of design options, the design example is simplified for illustration, focusing on compressor design for which all the design steps are described. The design procedure is structured as a sequential iterative approach in the sequence they are listed above.

The plant may be affected by large disturbances. Moreover, higher profit requirements mean that the process has to be operated close to its constraints. Hence, the operability analysis in step 7 may reveal findings which may be important from an economic perspective. The last sections give some examples of how to handle these issues by using important information from control analysis to achieve an optimally designed process which is easy to operate.

4.1 The TEALARC LNG process and model

The TEALARC process was patented by TECHNIP (1974), and has been described by Paradowski and Dufresne (1983). Based on that description, the process used in this study is illustrated in Figure 1 and is the same process as studied by Wahl (2007). The expander (Exp) and the storage tank (ST) are not included in the present model. Non-condensed NG is either used as fuel gas or recirculated to the feed inlet for NG. This is not shown in the figure. Two refrigeration cycles, named the liquefaction and pre-refrigeration cycles, are used in the plant. Both cycles contain refrigerants that are mixtures of components. The liquefaction cycle cools the natural gas in three heat exchanger steps (HE1-HE3). A dynamic, control relevant model was developed for the TEALARC LNG process by Michelsen et al. (2009a). The model contains simplified thermodynamics. For example, the vaporization and condensation processes in the heat exchangers are not modeled, and the streams are considered to be single phase. Instead, constant heat capacities are assumed for each stream. These are used as tuning parameters to adapt to a steady state point obtained from a rigorous design model for the same plant as developed by Wahl (2007). For operational system analysis and overall control structure design, however, the model still includes enough complexity for steady-state operability analysis. If a composition and temperature dependent heat capacity model was chosen instead of the fixed average heat capacity model for a certain heat exchanger stream, there would be some deviations from the present model. However, the qualitative response from any perturbation will normally be similar. Hence, the quite simple assumptions of constant heat capacities can be used and still provide a very good picture of how the overall plant will behave locally.

Figure 1 Flow sheet of the TEALARC LNG process used in this study. Control loops are not shown.

As an example of a process unit model, the compressor operating charts are modeled by cubic polynomials according to Jensen (2008), who used the method of Moore and Greitzer (1986) with some adjustments. Figure 2 shows the compressor characteristics. Pr is the pressure ratio defined as the discharge pressure divided by the surge pressure. The dashed arrow line in the upper figure shows the surge line through compressor curves at higher compressor speeds. The green line in the upper figure shows the control line located at the right hand side of the surge line.

Figure 2: Compressor curves as function of compressor speed N according to Michelsen et al. (2009a).

The compressors are allowed to operate individually at different speeds.

4.2 Resizing of compressor for profit increase and the impact on the choice of controlled variables

The choice of controlled variables (CVs) is a very important step in control design in order to obtain optimal operation in practice. The best choice should be robust in presence of unknown disturbances and process model uncertainties, and the final control implementation should be simple. There are many possible selections, and it is not always obvious to pick the best ones. For this analysis, methodology from self optimizing control is used. The methodology was proposed by Skogestad (2004) and further developed by Halvorsen et al. (2003), Alstad and Skogestad (2007) and Alstad et al. (2009).

Self-optimizing control analysis focuses on steady state performance of the process and involves constrained steady state optimization, which normally is made as a part of process plant design. The analysis aims at finding the set of CVs which gives the smallest steady state profit loss according to the operation objective function, despite changes in unknown disturbance variables (DVs) and implementation errors at a nominal operating point. The profit loss is defined as the difference between the profit using constant controller setpoints and the optimal profit.

At the nominal optimum, some process variables will normally lie at their constraints. Then a corresponding number of manipulated variables (MVs) are removed from the optimization problem, and the objective function is reformulated into an unconstrained optimization problem. The unconstrained, reduced objective function is then the basis for finding the controlled variables as a linear combination of measurements. Hence, one set of CVs are calculated for each unconstrained MV involved. This choice of CVs reduces the need for frequent real time re-optimization in operation. DVs may include exogenous changes affecting the system, process changes, changes in the specifications (constraints) and changes in the parameters (prices) that enter the objective function. They may also include parameter variations in the process. Implementation errors include control input deviation and measurement errors.

In the self-optimizing control design methodology the use of a model is only necessary in the design phase and not in the operation phase of the process. Further, the optimal choice of CVs has impact on recommendations for type and location of sensors for such plants. Michelsen et al. (2009b) applied the self-optimizing control design methodology to the TEALARC LNG process. The simulation results below include those results as well as another case of control structure design.

There are several options in how to select a compressor. In the following two design iterations, the size of one of the compressors is changed. The outset is a large measurement vector consisting of temperatures, flow rates and pressures in every stream in the plant and the mole fraction of methane in the gas and liquid outlet streams from the separator. This gives a basis of 66 possible measurements for finding the important variables to measure for control. Thus, measurements with minor significance can be left out in the final implementation.

The design procedure proceeds as follows (see Michelsen et al. (2009b) for more details):

1. First, a steady-state rigorous model (in Aspen HYSYS®) was developed by Wahl (2007) to perform a basic process design and to carry out energy analysis. Important constraints in the process variables, $(g(x,u,d))$, include:

   - 100% LNG fraction out of the sub-cooler HE3
   - Dew point margin on suction side of compressors, e.g. 10°C superheating (Jensen (2008)). These are safety constraints to avoid damage of the compressors.
   - Separator level [0-100%]

- Valve opening [0-100%]
- Compressor capacity (maximum drive power and rotational speed, surge limitation)
- Capacity in single components (e.g. maximum heat exchanger duty)
- Maximum feed (import limitation)
- Maximum production (export limitation)
- Storage capacity
- Product specifications (composition and state)
- Content limit of components which may freeze out in heat exchangers.
- Capacity limit of sea water cooling circuits

In the first iteration, all units operate at open loop, i.e. with constant inputs at their constraints.

2. The profit given by the difference between income from sale of LNG and the energy costs that are required to produce the LNG is chosen as the operational objective function *J*:

$$J = q_{\ln g} v_{\ln g} - q_{energy} v_{energy} \quad (1)$$

where

q_{lng} is the Production rate of LNG [kg/h]
v_{lng} is the unit price of LNG
q_{energy} is the flow of consumed energy
v_{energy} is the unit price of energy

The worst case loss is applied as operational objective, i.e. operability measure. This is expressed by:

$$L_{wc} = \frac{1}{2}\left(\bar{\sigma}\left[M\right]\right)^2 \quad (2)$$

where $\bar{\sigma}$ is the largest singular value of the matrix *M*, which is given by:

$$M = -J_{uu}^{1/2}(HG_y)^{-1}HF \quad (3)$$

using the nullspace method from the self-optimizing control methodology. The nullspace method finds the set of CVs which gives zero steady state profit loss, despite changes in the DVs as described in the introduction to this section above. The matrix G_y defines the sensitivity at the optimum operating point for the measurement vector *y* with respect to the MVs *u*:

$$\Delta y = G_y \Delta u \quad (4)$$

where $\Delta y=y-y^*$, $\Delta u=u-u^*$ and the asterisk defines the nominal optimum operating point. *F* is the optimal sensitivity matrix of *y* with respect to the disturbances *d*, defined by:

$$F = \frac{\partial y_i^{opt}(d_j^*)}{\partial d_j} = \begin{bmatrix} \frac{dy_1^{opt}}{dd_1} & \cdots & \frac{dy_1^{opt}}{dd_{n_d}} \\ \frac{dy_2^{opt}}{dd_1} & \cdots & \frac{dy_2^{opt}}{dd_{n_d}} \\ \vdots & \ddots & \vdots \\ \frac{dy_{n_y}^{opt}}{dd_1} & \cdots & \frac{dy_{n_y}^{opt}}{dd_{n_d}} \end{bmatrix} \text{ at } d=d^* \tag{5}$$

where n_d is the number of disturbances and superscript *opt* denotes the re-optimized measurements after perturbation dd_j of the disturbance d_j. J_{uu} is the Hessian matrix of J with respect to u. The matrix H gives the optimal linear combinations of the measurements y:

$$c = Hy \tag{6}$$

see Michelsen et al. (2009b) for more details.

3. Changes in heat transfer in the heat exchangers are regarded as the main disturbances. Such changes may typically be caused by composition variations in the natural gas or in the refrigerants. Hence, these are the actual disturbance sources. The relation between such composition variations and the heat transfer in the heat exchangers is, however, modelled in a simplified way as discussed in section 4.1. Hence, variations in the heat capacities of the natural gas and in the refrigerant of the liquefaction cycle are considered as unknown disturbances.

4. Michelsen et al. (2009a) developed a dynamic, control relevant model of the plant as described in section 4.1. A nominal design operating point for the steady-state rigorous model was used as a basis for steady state adaptation of the dynamic model. The initial choice of compressor chart can be regarded as a first rough design step. Next, a full chart was "painted" around the nominal point for each compressor in the dynamic model.

5. The steady state optimal performance is given by the maximum value of the operational objective function J with respect to the MVs for control u:

$$\max_u J \quad [NOK/h] \tag{7}$$

subject to

$$\begin{aligned} f(x,u,d) &= 0 \\ g(x,u,d) &= 0 \end{aligned} \tag{8}$$

where

$x \in \mathbb{R}^{n_x}$ is the state vector
$u \in \mathbb{R}^{n_u}$ are the MVs
$d \in \mathbb{R}^{n_d}$ are the unknown DVs
$g(x,u,d)$ are other constraints in the process variables (step 1)
$f(x,u,d)$ is the process model (step 4)

The NG flow rate is one of the process variables that has a strong impact on the production of LNG and is in some installations available for manipulation (see e.g. Singh et al. (2008) who applied the self-optimizing control methodology to a patented LNG plant designed by SINTEF). Figure 3 shows the profit as function of the NG flow rate q_{NG} as MV.

Figure 3: The profit as function of NG flow rate

The profit increases gradually from low NG flow rates until it drops steeply at a given rate due to the limitations in the refrigeration capacity. Above this point, the refrigeration cycles are not able to reach down to the temperature of total liquefaction of the natural gas. In this operating region, non-condensed NG is either used as fuel gas or recirculated to the feed inlet for NG (c.f. section 4.1). The optimum is located at 8000 kmol/h NG, giving a maximum profit of about 328 kNOK/h.

6. The nullspace method from the self-optimizing control structure design gives the following optimal CV based on q_{NG} as MV:

$$c = 0.76 y_{13} - 0.51 y_{41} - 0.41 y_{50} \tag{9}$$

The calculation of *c* involves steady state optimization calculations based on the model from step 4. Table 1 describes these measurements and Figure 6 shows their location in the plant. A PID controller is chosen as controller structure.

7. The worst case loss L_{wc} (see step 2 above) is zero based on *c* from step 6 above. This is a result of the nullspace method as used for the calculation of *c*.

Operability can also be verified by dynamic simulations of a control structure based on the resulting set of optimal CVs from step 6 and a dynamic model of the plant from step 4. Figure 4 shows the closed loop response of *c* from a positive disturbance change in the heat capacity of the natural gas. The response is acceptable with a small overshoot and fast settling time.

Figure 4: Closed loop response of the CV

Now, the design has to be changed in order to increase the production of LNG. Hence, some of the process constraints have to be changed, i.e. return to step 1. First, the compressor C2 is replaced by a larger compressor, and the NG flow rate is fixed at the optimal value above. All units operate at open loop, except for the compressor C2 which has the speed N_{C2} available as MV for control. This is another process variable that has a strong impact on the production of LNG and is also available for manipulation in some installations (see e.g. Singh et al. (2008)).

8. Figure 5 shows the profit as function of this MV (step 5). The profit increases instantly from low compressor speeds in the range below liquefaction until a maximum point (threshold speed) where it decreases gradually for higher speeds due to increased energy costs by the cooling work. Hence, the optimum is not at maximum compressor speed. The reduction in profit as function of higher compressor speeds is low due to a low energy price. This reduction is smaller at lower energy costs, and it is zero at zero energy costs. In that case, the profit is given by the LNG income only and optimum is achieved at any compressor speed above the threshold speed. The optimum is located at about 68% speed, giving a maximum profit of about 328 kNOK/h as before the design change.

Figure 5: The profit as function of compressor speed

This design change results in a change in the model structure as applied in the self-optimizing control analysis, As a result, a different set of self-optimizing CVs are obtained (step 6):

$$c = 0.62 y_1 - 0.74 y_{41} - 0.26 y_{51} \tag{10}$$

Table 1 describes these measurements, which differ from those as found in the previous iteration, see Figure 6.

Table 1: Description of measurements in the optimal measurement combinations

Figure 6: Liquefaction cycle with the measurements in the optimal sets of CVs, c.f. Figure 1

The worst case loss L_{wc} (as operability measure, see step 2 above) is still zero based on c from Eq. (10). The closed loop response of c from a positive disturbance change in the heat capacities of the natural gas (step 7) shows a similar response as Figure 4 (not shown here).

At this stage, it is possible to increase the NG flow rate in order to achieve increased optimal profit. However, this optimum is located at a higher compressor speed, which reduces the control margin to the maximum speed for disturbance rejection. Instead, by making the same sequence of analysis for even larger compressors, increased optimal profit is obtained at higher NG flow rates by keeping necessary control margin. This is illustrated by the set of profit curves in Figure 7 when considering the optimal point above as 80% capacity, c.f. Figure 3.

Figure 7: Profit as function of optimal NG flow rate at increasing compressor size

4.3 Feedback from operability analysis to process design

In the previous section, resizing of a compressor was used as an example. Another example of compressor design is the shape of the compressor curves. The slope of the curves affects the gain from the compressor speed to the pressure difference across the compressor. Steeper compressor curves mean higher gain and thereby lower sensibility for the compressor speed to reach one of its constraints at given disturbances. This is illustrated in Figure 8. At the operating point in the intersection between the two compressor curves, the gain G_1 defined as the derivative dPr/dq is smaller for a large compressor (dashed line) compared to the gain G_2 for a smaller compressor (solid line).

Figure 8: Illustration of a compressor curve for a small compressor (solid line) and a large compressor (dashed line)

The size of an NG cooler and the choice of number of serial coolers are other examples of important process design options for LNG plants. The latter is an option for process structure design. These options affect the temperature differences and gradients in the coolers, which again affect the steady state gain and the dominating time constant for control loops, and thereby how easy the whole plant is to control.

Morari (1983) defined a system with larger steady state gain and smaller time constant as being more resilient because it can handle larger disturbances. This means that the system has better controllability and flexibility characteristics. Iterations of changes in the process design examples above can be used to obtain a plant that is more resilient and thereby easier to operate.

4.4 Improved operability by control design

The plant is often designed such that the most expensive units are operated close to their respective constraints in order to reduce investment costs and increase the profit. Compressors are examples of such units. Hence, good control performance in terms of large rejection of disturbances is used to achieve tight control close to the constraint. This is illustrated in Figure 9. The process capability is the capability of the process to achieve various values *c* of the CV. This can be any process performance measure such as production time, which is closely related to the production profit. Thus, higher profit is achieved by improved control, which is obtained by carefully choosing the control structure, the controller structure and by tuning the controller parameters. As a result, the variance of the CV is reduced. This leads to less dynamic profit loss. Further, this allows for moving the mean value of a CV from c_1 to c_2, which constitutes the best feasible steady state. This is closer to its constraint C, which defines the optimum steady state. Alternatively, a more robust process can be achieved by keeping a large distance to the constraints, which leads to less number of process trips and shutdowns.

Figure 9: The economics of process control

5 Discussion

The suggested set of steps for a design procedure is generic in the sense that a procedure can be developed to improve the design of any new process plant. The topic is still of particular relevance to off-shore conversion/liquefaction plants as the requirements to such plants are extreme, both in terms of profit margins, degree of plant integration and compact size, required plant up-time, small (no) manning etc. Another reason for improving the methodological basis for integrated control structure and process design is that offshore gas processing plants represent new designs, and the engineers can not base their control structure design entirely on operational experience from similar plants.

A key ingredient in the design procedure is the use of a dynamic, control relevant simulation model. As an integral part of the analysis, the model can be used to evaluate various design options for the process units such as the sensitivity from design parameters to controllability and identification of critical parameters such as heat exchanger sizing and compressor sizing.

There are several ways of refining and structuring the process and control design procedure by including various analysis methods from the literature. In the selection of such methods, considerations should be made regarding the properties of individual process units as well as the total process design that affect the dynamic behavior and controllability of the process. Instability and cyclic behavior are examples of process behavior that may be problematic from an operation perspective. The structuring of the design steps might either lead to a sequential iterative or a simultaneous design procedure. Hence, the detailing of this procedure is case dependent and the methodology should therefore be further explored and developed through further applications.

There are also several options in how to select a certain compressor; the surge margin at the nominal point, the steepness of the chart, the speed range, power ranges and pressure ranges that are needed. Based on such options a large set of compressor designs can be screened out to fulfill the requirements for both the economics and the control performance, and simultaneously make a process that is able to handle the inevitable disturbances.

The example illustrates that the applied self-optimizing control design method is based on the assumption that a new set of controlled variables as controlled by the unconstrained degrees of freedom must be found for each possible set of active constraints. This means that the control structure has to be redesigned when a process unit meets its constraint for new expected disturbances or implementation errors, or when the process constraints are changed as in the example. An option is of course to over-design the process units at the first design iteration so that the process constraints never become active for the complete range of disturbances, although this might be an expensive solution. Also, ignoring optimization of the least important operating variables is an option (Fisher et al. (1988a; Fisher et al. (1988b; Fisher et al. (1988c), Morari and Perkins (1994)).

The self-optimizing control design methodology is based on steady state, local, linear analysis. Hence, there is no guarantee that the controlled variables are globally optimal. However, they give an indication of how sensitive they are to disturbances and measurement error. The profit loss should be verified by simulations of the non-linear model and/or tested in the real (non-linear) plant.

The profit of increasing equipment size, like a compressor, is normally given by the increased value of higher nominal production which must surpass the extra investment. Such process design changes should be made such that operability is improved, maintained or at least aggravated as little as possible. However, inevitable unknown disturbances and model uncertainties contribute to a profit loss in operation, compared to ideal optimal operation as if we knew the disturbances and our models were correct. Thus, resizing equipment may also have impact on the expected economic loss. This might be the case even if the equipment is not a bottleneck for the production rate. The selected control structure also plays an important role here, and this is the focus in this paper. Resizing equipments may lead to a requirement for a change in the control structure as well. The example shows that it is possible to find a new control structure for the resized compressor that also has zero worst case profit loss as operability measure. Then, the steady state operability is maintained after the process design change in this example. In this case the profit increase is only related to the increased nominal production. However, without changing the control structure, some extra loss has to be expected. On the other hand, other operability measures and other profit functions, including dynamic transients of economically important variables such as power consumption, might give increased profit when comparing the operation before the design change with the operation after the design change in other cases. Such type of analysis is a natural follow-up of this work.

6 Conclusions

Integrated methods for process and control design are systematic methods that can improve both design and operation of process plants. This means that there is a potential economic benefit for such approaches. The main message in this paper is that this is handled by considering control structure design when process design changes are made. This area is not yet mature and this paper is a step towards a further development of this methodology. The main steps in such a procedure are suggested.

The design procedure is exemplified by using the TEALACR LNG process. The analysis shows how a critical design issue like compressor sizing might affect the control structure design. A capacity increase, moving the plant bottleneck from a compressor (using the NG flow rate as manipulated variable for control) to the NG flow rate (using the compressor speed as manipulated variable for control), gives a different optimal combination of measurements as controlled variable. This means that a change in process constraints might influence the control design by changing what to control, i.e. the optimal combination of controlled variables. Hence, to decide whether it is a good idea to invest in a bigger compressor, one needs to consider control structure design. When this is made by the described method, the operability as defined by the steady state worst case profit loss is maintained.

Acknowledgements

This publication forms a part of the Remote Gas project, performed under the strategic Norwegian Research program Petromaks. The authors acknowledge the partners; Statoil, UOP, Bayerngas Norge, Aker Solutions, DNV, and the Research Council of Norway (168223/S30) for support.

References

Alstad, V. and S. Skogestad (2007). "Null Space Method for Selecting Optimal Measurement Combinations as Controlled Variables." Ind. Eng. Chem. Res. 46: 846-853.

Alstad, V., S. Skogestad, et al. (2009). "Optimal Measurement Combinations as Controlled Variables." Journal of Process Control 19: 138–148.

Anderson, J. S. (1966). Chem. Engr CE97.

Chawankul, N., H. Budman, et al. (2005). "The Integration of Design and Control: Imc Control and Robustness." Computers and Chemical Engineering 29(2): 261-271.

Fisher, W. R., M. F. Doherty, et al. (1988a). "Interface between Design and Control. 1. Process Controllability." Industrial & Engineering Chemistry Research 27(4): 597-605.

Fisher, W. R., M. F. Doherty, et al. (1988b). "Interface between Design and Control. 2. Process Operability." Industrial & Engineering Chemistry Research 27(4): 606-611.

Fisher, W. R., M. F. Doherty, et al. (1988c). "Interface between Design and Control. 3. Selecting a Set of Controlled Variables." Industrial & Engineering Chemistry Research 27(4): 611-615.

Foss, B. A. and I. J. Halvorsen (2009). Dynamic Optimization of the Lng Value Chain. Proceedings of the 1st Annual Gas Processing Symposium. E. A. Hassan, G. V. R. Reklaitis and M. E.-H. Mahmoud. Amsterdam, Elsevier**:** 10-18.

Georgakis, C., D. Uzturk, et al. (2003). "On the Operability of Continuous Processes." Control Engineering Practice 11(8): 859-869.

Georgiadis, M. C., M. Schenk, et al. (2002). "The Interactions of Design, Control and Operability in Reactive Distillation Systems." Computers and Chemical Engineering 26(4-5): 735-746.

Halvorsen, I. J., S. Skogestad, et al. (2003). "Optimal Selection of Controlled Variables." Ind. Eng. Chem. Res. 24(14): 3273-3284.

Jensen, J. B. (2008). "Phd Thesis ”Optimal Operation of Refrigeration Cycles”." Faculty of Natural Sciences and Technology, Department of chemical engineering, NTNU, Trondheim, ISBN 978-82-471-7801-0 (printed version).

Michelsen, F. A., I. J. Halvorsen, et al. (2009a). "Modeling and Simulation for Control of the Tealarc Lng Process." Paper submitted to Industrial & Engineering Chemistry Research

Michelsen, F. A., B. F. Lund, et al. (2009b). "Selection of Optimal, Controlled Variables for the Tealarc Lng Process." Paper submitted to Industrial & Engineering Chemistry Research

Mohideen, M. J., J. D. Perkins, et al. (1996). "Optimal Synthesis and Design of Dynamic Systems under Uncertainty." Computers & Chemical Engineering 20(Supplement 2): S895-S900.

Moore, F. K. and E. M. Greitzer (1986). "A Theory of Post-Stall Transients in a Axial Compressor Systems: Part I - Development of Equations." Journal of Engineering for Gas Turbines and Power 108: 68–76.

Morari, M. (1983). "Design of Resilient Processing Plants--Iii : A General Framework for the Assessment of Dynamic Resilience." Chemical Engineering Science 38(11): 1881-1891.

Morari, M. and J. D. Perkins (1994). Design for Operations. Proc. FOCAPD '94 (Foundations of Computer Aided Process Design), Snowmass Village, Colorado, USA.

Nishida, N. and A. Ichikawa (1975). Ind. Eng. Chem. Proc. Des. Dev. 14: 236-242.

Paradowski, H. and J.-P. Dufresne (1983). "Process Analysis Shows How to Save Energy." Hydrocarbon Processing, July: 103-108.

Sakizlis, V., J. D. Perkins, et al. (2004). "Recent Advances in Optimization-Based Simultaneous Process and Control Design." Computers & Chemical Engineering 28(10): 2069-2086.

Schweiger, C. A. and C. A. Floudas (1997). Interaction of Design and Control: Optimization with Dynamic Models. Optimal Control: Theory, Algorithms, and Applications. W. W. P. Hager, P. M., Kluwer Academic Publishers**:** 388--435.

Seferlis, P. and M. C. Georgiadis (2004). "The Integration of Process Design and Control." ELSEVIER.

Singh, A., M. Hovd, et al. (2008). "Control Variable Selection for Liquefied Natural Gas Plant." IFAC world Congress, S. Korea.

Skogestad, S. (2004). "Control Structure Design for Complete Chemical Plants." Computers & Chemical Engineering 28(1-2): 219-234.

TECHNIP (1974). "Wipo Registration Number 411865 Tealarc." TECHNIP, Société anonyme, 170 place Henri Régnault, F-92400 COURBEVOIE (FR).

Wahl, P. E. (2007). Enabling Production of Remote Gas: Pinocchio Lng - Energy Analysis, Internal Report, Sintef Energy Research.

Ziegler, J. G. and N. B. Nichols (1943). "Process Lags in Automatic Control Circuit." Transactions of ASME 65, 433.