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Controlled Variables Selection for Liquefied Natural Gas Plant

Arjun Singh*. Morten Hovd**
Vinay Kariwala***

**Department of Engineering Cybernetics, Norwegian University of Science and Technology Trondheim 7491 Norway (Tel: +47-735- 94482; e-mail:arjun.singh@itk.ntnu.no).*
***Department of Engineering Cybernetics, Norwegian University of Science and Technology Trondheim 7491 Norway (e-mail: morten.hovd@itk.ntnu.no)*
**** Division of Chemical and Biomolecular Engineering, Nanyang Technological University Singapore 637459 (e-mail: vinay@ntu.edu.sg)*

Abstract: The primary operational objective of liquefied natural gas (LNG) plants with low production capacity is to maximize LNG production. A robust control structure is required to meet economic objective while ensuring safety of the plant. The selection of appropriate controlled variables (CVs) is critical for design of the control structure. In this paper, general process understanding and the concept of self-optimizing control are used to select CVs for the liquefaction unit of an LNG process. The analysis is carried out using a steady state model developed in gPROMS. It is shown that with appropriate selection of CVs, the effect of disturbances and uncertainty can be minimized on the operational objective of the LNG plant.

1. INTRODUCTION

Natural gas (NG) is a gaseous fossil fuel consisting primarily of methane but including significant quantities of ethane, butane, propane, carbon dioxide, nitrogen, helium and hydrogen sulfide. NG is a major source of electricity generation through the use of gas and steam turbines. Other applications of NG are in residential domestic use and automobile industry. NG also has an advantage that it burns cleaner than other fossil fuels, such as oil and coal, and produces less carbon dioxide per unit energy released.

The major difficulty in the use of NG is transportation and storage because of its low density. NG pipelines are economical, but are impractical across oceans. The discovery of NG fields at remote location also makes pipeline transportation infeasible and economically unacceptable. Liquefaction of NG reduces the specific volume by a factor of 600, thus making transport of liquefied natural gas (LNG) in ships or trucks attractive (Hammer, 2004). Liquefaction of NG, however, requires huge investments and is an energy demanding process (Zalm, 2002). It is quite common to have a heavy upfront investment in large industrial plants for producing LNG since cost per unit of gas volume is relatively low over plant lifetime. In such plants, multi-component refrigerants are commonly used to achieve low temperatures required for liquefaction of NG (approximately -160 ^{0}C at near atmospheric pressure).

There is, however, a growing need for liquefaction of NG at places where it is not possible or economically acceptable to invest heavily. This includes local distribution of NG in small markets, where plant needs to be built close to the gas pipe, while the LNG is transported by trucks and small ships. For such plants, low investment costs are considered to be more important than optimal energy utilization. This happens despite the fact that the relative investment cost for small-scale LNG plants increases almost exponentially with decreased production capacity from about 50,000 tones per annum and below.

The Foundation for Scientific and Industrial Research at the Norwegian Institute of Technology (SINTEF) has developed a low capacity plant, which requires low investment cost and is easy to construct at desired sites. The plant design has been patented by SINTEF (Neeras and Brendeng, 2004). In the subsequent discussion, this patent is referred as the SINTEF patent and the plant as the SINTEF LNG plant.

Singh and Hovd (2007) developed a dynamic model for this plant. The developed model was validated against experimental results obtained from a pilot plant facility located in Trondheim, Norway. The comparison showed that the model predictions closely match the experimental values at steady state. For operation of this plant in face of disturbances and uncertainty, a robust control structure is required. Singh and Hovd (2007) proposed a control structure, but the controlled variables (CVs) were selected arbitrarily.

Finding an appropriate set of CVs is a key step in the design of the control system for the plant. It is important to minimize the economic loss which occurs when plant experience disturbances and optimal operating point shifts. Various methods for selecting CVs have appeared in the process control literature over the past few decades; see *e.g.* Van de Wal and de Jager (2001) for a review. Recently, Skogestad (2000) introduced the concept of self optimizing control for selecting CVs. This concept involves minimization of the economic loss incurred in indirectly optimizing the operation by holding the selected CVs constant, as compared to frequent online optimization.

In general, the operational objectives of LNG plants include maximizing LNG production, while minimizing energy consumption. For plants with low capacity such as the SINTEF LNG plant, however, energy consumption has insignificant contribution to the overall economics. Thus maximization of LNG production is considered as the only operational objective in this paper. We use general process understanding and the concept of self-optimizing control to select CVs to maximize throughput. For this purpose, a steady state model for the plant is developed in gPROMS, using Multiflash for thermodynamic property calculations. It is shown that with appropriate selection of CVs, the effect of disturbances and uncertainty can be minimized on the operational objective of the SINTEF LNG plant. We also propose a control structure based on the physical proximity of selected CVs and manipulated variables.

2. PROCESS DESCRIPTION

In this section, we briefly describe the SINTEF LNG plant. Interested readers can find the complete description of this process in (Neeras and Brendeng, 2004). Fig 1 represents the simplified flowsheet of the SINTEF LNG plant, where some features have been omitted for clarity.

In Fig.1, sub-components of the plant are numbered and referred as units. Units 13-18 are heat exchangers (HX). The HX numbered 13, 15 and 17 are called the 'Refrigerant HX' and HX numbered 14, 16 and 18 are called 'LNG HX'. Units 3 and 5 are separators and units 4, 6-9 and 12 are valves. Units 10 and 11 are ejectors, while units 1, 2 and 19 represent the condenser, cooling water stream, and the compressor, respectively.

In this plant, the refrigerant is compressed in unit 19. Subsequently, the refrigerant is partially condensed in unit 1 by water cooling, where all the heat absorbed by the refrigerant from NG is removed. The two-phase refrigerant now enters unit 3, where most volatile components are separated at the top and partially condensed in unit 13. The liquid stream from unit 3 goes through pressure reduction in unit 4 and is then mixed with two refrigerant streams, coming from units 15 and 16, in unit 11. Partially condensed refrigerant from unit 13 enters unit 5, where again most volatile components are separated at the top, which are condensed and sub-cooled in units 15 and 17. The sub-cooled stream is flashed to a pressure of about 2-4 bars providing the cold refrigerant for units 17 and 18. The liquid stream from unit 5 undergoes pressure reduction in unit 6 and is mixed with the outlet refrigerant streams from units 17 and 18 in unit 10. After mixing, the refrigerant flow is divided and distributed among units 15 and 16. Evaporating streams in units 13 and 14 are mixed together before they enter compressor. The NG is pre-dried and CO_2 is removed to a level, where no solidification (freezing) occurs in the heat exchangers. Further, NG is cooled in unit 14, condensed in unit 16 and sub cooled in unit 18.

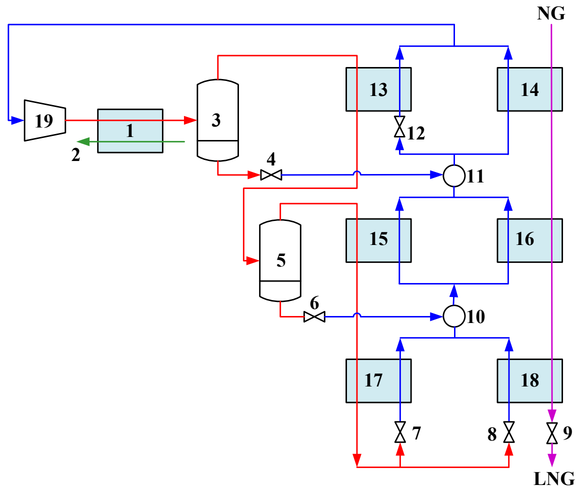

Fig 1: Flow sheet of SINTEF LNG plant

Remark 1. The arrangement of heat exchangers and compressor shown in Fig. 1 is to ensure that oil, which follows refrigerant in the cycle, does not reach the coldest part of the plant. This avoids freezing of the oil and plugging of conduits. In such arrangements, effect of gravity is also minimized since evaporating streams flow upward and condensing streams flow downward.

3. MODEL DEVELOPMENT

A steady state model for the plant is developed in gPROMS using Multiflash for calculation of physical properties for the NG and the refrigerant. The Soave-Redlich-Kwong (SRK) equation of state is used for both the NG and the refrigerant. We have first developed individual models for the main components in the plant, namely the heat exchangers, valve, compressor, condenser, separator and ejector. The models of the heat exchangers and condenser are based on the same principles. Subsequently, models of the individual unit operations are connected together to form the model of the whole plant. A brief description of the individual models is given below:

3.1 Heat exchangers

For heat exchangers, a one dimensional distributed model with heat exchange between two streams is developed using energy (internal energy) and mass balance. Pressure drop in the heat exchanger is neglected. The composition of each stream is assumed to be constant throughout the heat exchanger. A constant heat transfer coefficient is assumed for each stream. Streams (evaporating and condensing) are assumed to exchange heat through the metal wall, which is assumed to have negligible thermal conduction in the axial direction and infinitely fast thermal conduction in the radial direction (Qu et al., 2006). A separate energy balance is used for the internal energy of the metal wall. Wall ends are assumed to be adiabatic.

3.2 Compressor

This model describes the relation between the gas mass flow rate and the pressure head across the compressor. The hold up and inertia of the refrigerant are considered negligible. Fan laws (affinity laws) are used to model speed dependent variations in the performance so that a single characteristic curve (head vs. flow) can describe the behavior at any speed. The compression process is modeled as polytropic. Constant efficiency is assumed for compressor *i.e.* efficiency is not assumed to vary with flow rate.

3.3 Separators

At steady state, separators (flash drums) are equivalent to flash calculations at constant temperature and pressure for the given composition. Hence these separators are modeled using T-P flash calculations.

3.4 Valves

Valves are assumed to have equal enthalpy at the inlet and the outlet. Valve model is described by an equation relating the pressure drop across the valve to mass flow rate through the valve.

3.5 Ejectors

Modeling ejector is a challenging task since it involves mixing of liquid phase refrigerant with two–phase refrigerant. As procurement of geometrical data for the ejector used in the plants is difficult, ejectors are modeled as pure mixers. The mixer model accounts for the component mass and energy balances. The ejector is modeled as an isobaric unit, where the pressure is calculated from the relationship between the mass flow rate and the pressure drop for the exit stream from the ejector.

4. SELF OPTIMIZING CONTROL

The optimal operating point of a plant changes with disturbances (d). It is optimal to track these variations in the degrees of freedom (u) using an online optimizer. Recently, Skogestad (2000) presented a simpler feedback based strategy for achieving nearly-optimal operation.

To present this methodology, let the economics of the plant be characterized by the scalar objective function J. The central idea is to use feedback controllers to indirectly update u such that CVs (c) are held close to their set points. In this case, in addition to u and d, J is affected by the implementation error (n) arising due to uncertainty and measurement noise. As compared to the online optimizer, the feedback based strategy results in a loss and self optimizing control is said to occur when the loss is acceptable (Skogestad, 2000). Based on this concept, the CVs can be selected by comparing the losses for different alternatives.

The choice of CVs based on the general non-linear formulation of self-optimizing control can be time consuming. Halvorsen et al. (2003) have presented a local method to quickly screen the alternatives. The final selection of CVs is done in a subsequent step, where the losses for the promising set of candidate CVs identified using local analysis are compared using the nonlinear model.

To present local method, let the linearized model, obtained around the nominal point, be given as

$$y = G^y u + G_d^y W_d d + W_n n \tag{1}$$

Here y denotes the process measurements and the diagonal matrices W_d and W_n contain the magnitude of expected disturbances and implementation errors associated with the individual measurements, respectively. Denoting $G = HG^y$ and $G_d = HG_d^y$, the CVs (c) are given as

$$c = Hy = Gu + G_d W_d d + HW_n n \tag{2}$$

It is assumed that the dimension of c is the same as u, and G = HG^y is invertible. Halvorsen et al. (2003) have shown that the local worst-case loss over the allowable set of d and n is given as

$$L_{worst} = 0.5\sigma_{\max}^2\left(\begin{bmatrix} M_d & M_n \end{bmatrix}\right) \tag{3}$$

where $\sigma_{\max}$ is the maximum singular value and

$$M_d = J_{uu}^{1/2}(J_{uu}^{-1}J_{ud} - G^{-1}G_d)W_d \tag{4}$$

$$M_n = J_{uu}^{1/2}G^{-1}HW_n \tag{5}$$

Here, $J_{uu} = \partial^2 J/\partial u^2$ and $J_{ud} = \partial J/\partial u\partial d$ are evaluated at the nominally optimal operating point.

The minimization of worst case loss can be conservative as it may not occur frequently in practice. It is more appropriate to minimize the average loss. Kariwala *et al.* (2007) have shown that the average loss is given as

$$L_{average} = \frac{1}{6(n_u + n_y)}\left\|\begin{bmatrix} M_d & M_n \end{bmatrix}\right\|_F^2 \tag{6}$$

where $||.||_F$ denotes the Frobenius norm. Losses in (3) and (6) depend on H and the CVs are selected by minimizing the losses with respect to H. When individual measurements are selected as CVs, the matrix H is a selection matrix with its elements being restricted to be binary.

5. SELECTION OF CONTROLLED VARIABLES FOR SINTEF LNG PLANT

In this section, the self-optimizing control method is applied for selection of CVs for the SINTEF LNG plant.

5.1 Operational objective and constraints

The primary operational objective for the SINTEF LNG plant is to maximize LNG production. In addition, the following constraints must be satisfied during the operation:

- *Superheating:* The refrigerant entering the compressor must be at least 10^0C superheated (Jensen and Skogestad, 2006).
- *LNG Temperature:* The temperature of NG coming out of unit 18 must be below a certain temperature. This is to make sure that NG is converted into LNG and that LNG is stored as a 'boiling cryogen' at near atmospheric pressure. In this state, temperature and pressure of LNG don't increase even if there is heat addition to the stored LNG due to imperfect insulation since LNG vapor boil off due to this heat addition is allowed to leave the storage tank.
- *Compressor power:* The compressor power must be less than an upper bound, which depends on the equipment specifications.

To optimize the operation, this plant has the following 8 degrees of freedom or manipulated variables:

- Compressor speed
- Mass flow rate of cooling water in condenser
- Opening of valve 4
- Opening of valve 6
- Opening of valve 7
- Opening of valve 8
- Opening of valve 9
- Opening of valve 12

Changes in inlet temperature and pressure of NG stream are considered the main disturbances. Composition of refrigerant is assumed constant and active charge in heat exchangers and condenser is neglected.

5.2 Degrees of Freedom Analysis

The SINTEF LNG plant has two separators. Since there may be a significant amount of refrigerant in the heat exchangers and pipes connecting different components, it is possible that either any one or both of the drums empty in the presence of disturbance. Thus it may seem necessary to control liquid levels in two separators. The dynamic non-linear model for the plant, however, shows that the refrigerant hold up outside these two separators is negligible. In this case, the liquid level in only one separator need to be controlled, which is chosen based on the quantity of liquid refrigerant it contains. In other words, we control the level of the separator which has less liquid refrigerant since chances of it getting empty are higher in case of disturbances.

In addition, the compressor speed and cooling water mass flow rate in condenser should be kept at their maximal allowable values, as energy consumption is not considered an issue. It is also necessary to control the degree of superheat at 10^0C from safety point of view. We note that the heat energy absorbed by the refrigerant leaves the cycle only in the condenser. As the flow rate of the cooling water in the condenser is constrained at its maximum value, the heat energy that can be removed from the NG stream is also maximal. Thus cooling of LNG beyond the dew point temperature (approximately -163 ^{0}C at near atmospheric pressure) can only be done at the expense of reducing the throughput. With these arguments, we control the LNG temperature at its dew point temperature.

After implementing the active constraints and controlling necessary variables from safety and operational point of view, we are left with only three degree of freedom. For steady state analysis, pairing of inputs and outputs is insignificant and without loss of generality, we choose remaining degrees of freedom as the openings of valves 7, 8 and 12.

5.3 Local self-optimizing control analysis

To select CVs for the remaining 3 degrees of freedom, we use the concept of self-optimizing control. For local analysis, a linear model of the process is required, where the point of linearization is usually chosen as the optimal operating point for nominal values of disturbances. For the SINTEF LNG plant, finding the nominally optimal operating point is difficult due to numerical reasons. Alstad (2005) has shown that local analysis can still be applied by linearizing by model at a non-optimal operating point, as the differences between losses for different candidate CVs remain the same.

The set of measurements, among which CVs are chosen, is shown in Table 1. While most of the chosen measurements are temperature variables, some temperature differences are also considered. For all these measurements, an implementation error of ±1°C is assumed. The allowable ranges of variations for the inlet temperature (d_1) and pressure of NG in HX 14 (d_2) is taken as ±5°C and ±2 bars around the nominal operating values, respectively.

The Hessian matrices required for local analysis are shown in (7) and (8), which are obtained using small perturbations.

$$J_{uu} = \begin{bmatrix} 1 & 0 & 0 \\ 0 & 2.53 & 0 \\ 0 & 0 & 7 \end{bmatrix} \tag{7}$$

$$J_{ud} = \begin{bmatrix} -1.7e-3 & 6e-3 & -1.2e-2 \\ -2.4e-7 & 4e-7 & -6e-7 \end{bmatrix} \tag{8}$$

The most promising CVs identified using local analysis are shown in Table 2. It is noted that the ranking of these CVs is same for worst-case loss and average loss minimization.

5.4 Verification based on non-linear model

It is essential to verify the results obtained from local analysis using the non-linear model. Similar to local analysis, however, a key difficulty in loss calculation for different CV alternatives is the determination of optimal value of objective function J for every disturbance and implementation error scenario. Denoting $J^{opt}(d,n)$ and $J_c(d,n)$ as the objective function values, when the degrees of freedom are updated using online optimizer and by controlling CVs, respectively, we note that

$$L_{average} = \frac{1}{|D|+|N|} \int_{d\in D, n\in N} \left(J^{opt}(d,n) - J_c(d,n)\right)$$
$$= \frac{1}{|D|+|N|} \int_{d\in D, n\in N} J^{opt}(d,n) - \frac{1}{|D|+|N|} \int_{d\in D, n\in N} J_c(d,n) \quad (9)$$

where D and N are the allowable sets for disturbances and implementation errors, respectively, and $|.|$ denotes the size of a set. Now, the first term in (9) remains the same for different CV alternatives. In this paper, we use the value of the objective function for nominal disturbances and implementation errors as an estimate of the first term in (9) without imposing any limitations on the CV selection. Note that this simplification cannot be used for computation of the worst-case loss.

Table 1: Candidates for Selection of Controlled Variables

Variable	Remark
T1	Compressor Discharge Temperature
T2	Inlet Temperature of evaporating refrigerant stream in HX 13
T3	Inlet Temperature of condensing refrigerant stream in HX 13
T4	Exit Temperature of evaporating refrigerant stream in HX 13
T5	Exit Temperature of condensing refrigerant stream in HX 13
T6	Inlet Temperature of evaporating refrigerant stream in HX 15
T7	Exit Temperature of evaporating refrigerant stream in HX 15
T8	Exit Temperature of condensing refrigerant stream in HX15
T9	Inlet Temperature of evaporating refrigerant stream in HX 17
T10	Exit Temperature of evaporating refrigerant stream in HX 17
T11	Exit Temperature of condensing refrigerant stream in HX 17
T12	Exit Temperature of evaporating refrigerant stream in HX 15
T13	Exit Temperature of natural gas stream in HX 15
T14	Exit Temperature of evaporating refrigerant stream in HX16
T15	Exit Temperature of natural gas stream in HX16
T16	Exit Temperature of evaporating refrigerant stream in HX 18
T17	Difference in the temperature of evaporating refrigerant streams coming out of HX15 and HX16
T18	Difference in the temperature of evaporating refrigerant streams coming out of HX13 and HX14
T19	Temperature at the exit of valve 6
T20	Difference in temperature of evaporating refrigerant streams coming out of HX17 and HX18

Using the non-linear model and keeping CVs (chosen based on the linear analysis) at their set points, losses were obtained by changing one disturbance or implementation error at a time and the results are shown in Table 3. Based on the average change in LNG throughput for different alternatives, we observe that the ranking of CVs differs between the linear and non-linear analysis. This can attributed to the process non-linearity and the limited range of disturbance and implementation error changes considered for loss evaluation using non-linear model.

Table 2: Most Promising controlled variables

CVs	Average Loss $\times 10^4$	Worst Case loss $\times 10^5$
T9, T12,T15	3.0414	4.5621
T9, T15,T18	3.0437	4.5656
T6, T9,T12	3.0515	4.5773
T6, T9,T18	3.0538	4.5806

The most promising CVs are the inlet temperature of evaporating refrigerant stream in HX 17, exit temperature of natural gas in HX 16, and the difference in the temperature of evaporating refrigerant streams coming out of HX 13 and HX 14. Selection of T18 as controlled variables is consistent with SINTEF patent, which suggest that it is desirable to have equal temperature for these streams at the exit of HX 13 and HX 14.

To summarize, the available MVs are openings of valve 4, 6, 7, 8, 9, and 12. The CVs are liquid level in unit 5, T9, T15, T18, degree of superheat of refrigerant at compressor suction, and LNG temperature.

A possible approach for implementation of the control structure with selected CVs is shown in Fig. 2, where the input-output pairings are selected based on relative gain array (RGA) analysis (Bristol, 1966). We have chosen these pairing based on sign of steady state relative gains and magnitude of relative gains at around bandwidth frequency. Final proposed pairings (MV ↔ CV) are given below:

- Opening of valve 4 ↔ Degree of superheat
- Opening of valve 6 ↔ Liquid level in unit 5
- Opening of valve 7 ↔ T9
- Opening of valve 8 ↔ LNG Temperature
- Opening of valve 9 ↔ T15
- Opening of valve 12 ↔ T18

6. CONCLUSIONS AND FUTURE DIRECTIONS

This work has identified non-obvious controlled variables for the SINTEF LNG plant using the concept of self-optimizing control. A control structure based on the RGA analysis is also proposed. This feasibility of the proposed control structure is currently being evaluated using the non-linear dynamic model in gPROMS. Upon completion, the control structure will be recommended to SINTEF for testing on the pilot plant.

Table 3: Percentage Loss for different CV alternatives with disturbances and implementation errors changed one at a time

CVs	d_1^+	d_1^-	d_2^+	d_2^-	n_1^+	n_1^-	n_2^+	n_2^-	n_3^+	n_3^-	Average change	Ranking
T9, T12,T15	1.7699	1.8604	0.229	0.043	0.165	0.158	0.150	0.151	0.679	0.612	0.58214	2
T9, T15,T18	1.7601	1.8482	0.235	0.056	0.171	0.163	0.663	0.591	0.150	0.149	0.57906	1
T6, T9,T12	1.7260	1.8136	0.346	0.291	0.749	0.655	0.163	0.155	0.146	0.146	0.61937	4
T6, T9,T18	1.7174	1.8028	0.349	0.295	0.732	0.633	0.168	0.160	0.145	0.144	0.61519	3

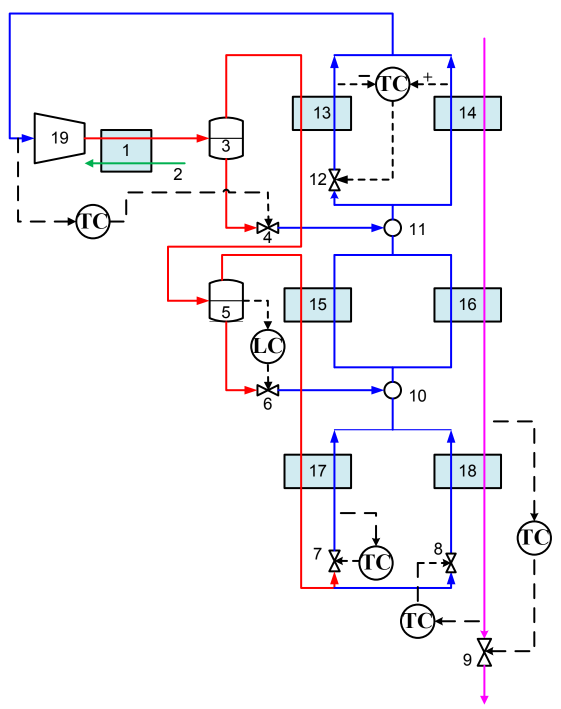

Fig 2. Proposed control structure for SINTEF LNG plant.

7. ACKNOWLEDGEMENTS

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