406d A Plantwide Control Procedure Applied to the Hda Process

Antonio C. B. Araujo, Norwegian University of Science and Technology (NTNU), Chemical Engineering, Sem Sealands vei 4, Trondheim, N7491, Norway and **Sigurd Skogestad**, Chemical Engineering, Norwegian Univ of Sci & Tech (NTNU), Trondheim, N7491, Norway.

Introduction

The HDA process, due to McKetta (1977), was first presented in an American Institute of Chemical Engineers contest problem. Through the years, it has been exhaustively studied by several authors, e.g Stephanopoulos (1984), Brognaux (1992), Cao and Rossiter (1997), Wolff (1994), Herrmann et al. (2003), Ng and Stephanopoulos (1996), Ponton and Laing (1993), Brekke (1999), Luyben et al. (1998), Luyben (2002), and Konda et al. (2005) and many others. The focus of these studies have included steady state design, controllability and operability of the dynamic model and control structure selection and controller design. Many authors have proposed control structures based heuristic procedures and have not focused on the economics. Thus, most of the proposed configurations are far away from the optimal operation. In particular, economically optimal active constraints are not necessarily controlled. The resulting back-off from the active constraints, which is necessary to guarantee feasibility, results in an economic loss.

Plantwide control procedure

In this paper, we apply the plantwide control procedure of Skogestad (2004). The procedure consists of two main steps.

I. "Top-down" steady-state approach where the main objective is to identify the primary controlled variables

A steady-state analysis is sufficient provided the plant economics depend primarily on the steady state. First one needs to quantify the number of steady-state degrees of freedom. This is an important number because it equals the number of primary controlled variables that we want to identify. The next step is to optimize the steady-state operation with respect to the degrees of freedom using a steady-state plant model. This requires that one identifies a scalar function J to be minimized. Typically, an economic cost function is used:

J = cost of feed + cost of energy - value of products

Other operational objectives are included as constraints. A key point of the optimization is to identify the active constraints, because these must be controlled to achieve optimal operation. For the remaining unconstrained degrees of freedom, the objective is to find sets of "self-optimizing variables", which have the property that near-optimal operation is achieved with these variables are fixed at constant setpoints.

A simple rule for identifying controlled variables is:

* Look for sets controlled variables that maximize the minimum singular value of the scaled gain matrix.

For scalar cases, the minimum singular value is simply the gain |G'|. The optimal variation (plus the implementation error) enters into the scaling factor. For the scalar case: G' = G / span(c) where span(c) = optimal variation in c + implementation error for c

II. Bottom-up design of control structure with the primary aim of identifying secondary controlled variables (denoted y2) and to pair these with available manipulated inputs (denoted u2).

The number of possible control structures is usually extremely large, so in this part of the procedure one aims at obtaining a good but not necessarily the optimal structure. The main aim is "stabilization" and to make the control problem easy as seen from the top layer. Basically, the variables y2 are variables that should be controlled in order to "stabilize" the operation in the sense of avoiding drift. Typical variables are liquid levels, pressures in key units and some temperatures (e.g. in reactors and distillation columns). Some flows may also be controlled at the lowest level. This results in a hierarchical control structure, with the fastest loop (typically the flow loops) at the bottom of the hierarchy. Note that no degrees of freedom are lost as one closes loops, as the setpoints of the secondary variables (y2s) are the "inputs" for controlling variables in the layer above. The "maximum gain rule" proves useful also for identifying the secondary variables y2, but note that the gain should be evaluated at the frequency of the layer above. The secondary outputs y2 need to be "paired" with manipulated inputs u2. Some guidelines:

Eventually, as loops are closed one also needs to consider the controllability of the "final" control problem which has the primary controlled variables y1=c as outputs and the setpoints to the regulatory control layer y2s as inputs. In the end, dynamic simulation may be used to check the proposed control structure, but as it is time consuming and requires a dynamic model it is usually avoided.

Application to HDA process

The HDA plant has 14 steady-state degrees of freedom. Optimization shows that there are 11 active constraints, namely:

1. Pure toluene feed rate
2. By-pass valve around FEHE
3. Reactor inlet hydrogen-aromatics ratio
4. Flash inlet temperature
5. Methane mole fraction in stabilizer bottom
6. Benzene mole fraction in stabilizer distillate
7. Toluene mole fraction in benzene column bottom
8. Benzene mole fraction in benzene column distillate
9. Diphenyl mole fraction in toluene column bottom
10. Toluene mole fraction in toluene column distillate
11. Quencher outlet temperature (equality constraints)

Therefore, 3 unconstrained degrees of freedom are left. To identify promising self-optimizing variables, the minimum singular value of alternative 3x3 steady-state gain matrices was evaluated. From this,a nd controllability considerations the following self-optimizing controlled variables were selecred

1. Benzene mole fraction in quencher outlet
2. Compressor power
3. Separator pressure.

Alternative self-optimizing variables have evaluated, but these give good results and are simple to implement.

Next, the bottom-up part of the design procedure is applied to the HDA process. The resulting structure has been assessed by dynamic simulations using the commercial simulator AspenDynamics from AspenTech. The simulations so far show that there are no serious control limitations for this process and that a decentralized supervisory control layer would perform well.

In summary, the objective of this paper is to synthesize an economically and yet dynamically attractive control structure for the HDA process based on the previous approach and compare its performance with the classical configurations afore mentioned.

References

- McKetta, J. J. (1977). Benzene design problem. Encyclopedia of Chemical Processing and

Design. Dekker, New York, USA.

- Stephanopoulos,G. (1984). Chemical process control. Prentice-Hall International Editions. New Jersey, USA.

- Brognaux, C. (1992). A case study in operability analysis: The HDA plant. Master thesis. University of London, London, England.

- Cao, Y. and D. Rossiter (1997). An input prescreening technique for control structure selection. Computers and Chemical Engineering 21(6), 563–569.

- Wolff, E. A. (1994). Studies on control of integrated plants. PhD thesis. Norwegian University of Science and Technology, Trondheim, Norway.

- Herrmann,G., S. K. Spurgeon and C. Edwards (2003). A model-based sliding mode control methodology applied to the HDA plant. Journal of Process Control 13, 129–138.

- Ng, C. and G. Stephanopoulos (1996). Synthesis of control systems for chemical plants. Computers and Chemical Engineering 20, S999–S1004.

- Ponton, J.W. and D.M. Laing (1993). A hierarchical approach to the design of process control systems. Trans IChemE 71(Part A), 181–188.

- Brekke, K. (1999). Optimization and control of the HDA process. Master thesis. Norwegian University of Science and Technology, Trondheim, Norway.

- Luyben,W. L., B. D. Tyréus and M. L. Luyben (1998). Plantwide process control. McGraw-Hill. USA.

- Luyben, W. L. (2002). Plantwide dynamic simulators in chemical processing and control. Marcel Dekker, Inc.. New York, USA.

- Konda, N. V. S. N. M., G. P. Rangaiah and P. R. Krishnaswamy (2005). Simulation based heuristics methodology for plant-wide control of industrial processes. In: Proceedings of 16th IFAC World Congress. Praha, Czech Republic.

- Skogestad, S. (2004). Control structure design for complete chemical plants. Computers and Chemical Engineering 28, 219-234.