CONTROLLABILITY ANALYSIS OF INDUSTRIAL FIVE EFFECTS EVAPORATOR SYSTEM

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Abstract

This paper reports on the controllability analysis of an industrial five-effects-evaporator process using the Dynamic Operability Framework. The process is associated with the liquor-burning unit of the Bayer process for alumina production. The control design problem is to reach a target density and to maintain the level on each flash tank as well as product temperature within limits with minimum operational costs. The problem is addressed by examination of various operational modes and control loops in the process superstructure. It involves the solution of a dynamic Mixed Integer Nonlinear Problem, as well as geometric representation of feasible operating space and process dynamic responses within the Dynamic Operability Framework. It is shown that the target density is achieved by using four effects operation and minimum variability is kept at minimum by appropriate control loops selection.

Keywords

process control, controllability, dynamic operability, optimization

Introduction

As one stage of the Bayer process for alumina production, the liquor-burning evaporation unit has been receiving limited attention, especially on control studies (To *et al.*, 1998, Kam and Tade, 2000, Sidrak, 2001). The reasons include that the liquor being evaporated is an intermediate product and hence its economic value is yet to be determined. However, the process is notoriously nonlinear and interacting, with limited online sensors for measurement of mineral, chemical and physical properties, therefore making optimal control design difficult. Improved liquor-burning evaporator performance will lead to increased throughput, with improved organics removal and hence increase refinery liquor yield.

This paper reports the systematic controllability assessment of an industrial five-effects-evaporator associated with a liquor-burning unit in the Bayer process, using Dynamic Operability Framework. It addresses the problem of handling high interaction between liquor levels, product density and temperature with proper selection of operating condition and control pairs in a multi-loop PI-control strategy. In the following sections, a brief review of Dynamic Operability Framework features is presented. This is followed by evaporator superstructure, control and optimization strategies. Finally, an application of the framework to find the optimum operating mode and conditions including control loops selection is demonstrated.

Dynamic Operability Framework

The Dynamic Operability Framework applied in this study is an extension of the original framework (Bahri,

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1996) to incorporate Operability Index (Vinson and Georgakis, 2000) for dynamic regulatory case, specifically the geometric representation of the feasible operating conditions and system responses in process system design (Ekawati and Bahri, 2001).

The framework determines the optimum operating condition within the feasible space, such that any disturbances and uncertainties affecting the system will not cause constraint violations, whilst maintaining the best control and economic objective function possible. For that purpose, the framework applies several high dimensional geometric spaces to represent the process as follows:

1. *Expected Disturbance Space (EDS)* as the set of disturbance values expected to affect the process

$$\theta \in EDS = \{ \theta : \theta^l \le \theta \le \theta^u \} \tag{1}$$

2. *Desired Output Space (DOS)* as the set of desired values of the output and measured variables

$$\begin{aligned} &DOS(z, y, \theta, x, \dot{x}, w, u, p, t) = \\ &g_j(z, y, \theta, x, \dot{x}, w, u, p, t) \le 0 \quad j \in I \end{aligned} \tag{2}$$

3. *Achievable Output Space due to Disturbance (AOS_d)*, as the set of variation on output variables caused by members of EDS

$$\begin{aligned} &AOS_d(z, y, \theta, x, \dot{x}, w, u, p, t) = \\ &h_i(z, y, \theta, x, \dot{x}, w, u, p, t) \quad i \in E \\ &\theta \in EDS \end{aligned} \tag{3}$$

4. *The regulatory Output Controllability Index (r-OCI)* of the process

$$r - OCI = \frac{\mu(AOS_d \cap DOS)}{\mu(AOS_d)} \tag{4}$$

Subsequently, the extension of the original framework (Bahri, 1996) is formulated as follows:

Outer level:

$$\begin{aligned} &\min_{z,y} \Phi(z, y, \theta^k, x, \dot{x}, w, u, p, t) \\ &s.t. \quad h_i(z, y, \theta^k, x, \dot{x}, w, u, p, t) = 0 \quad i \in E \\ &\qquad r-OCI(z, y, \theta^k, x, \dot{x}, w, u, p, t) = 1 \\ &\qquad z \in Z = \{ z : z^l \le z \le z^u \}, \\ &\qquad y \in \{0,1\} \qquad \theta^k \in EDS \\ &\qquad x \in X = \{ x : x^l \le x \le x^u \} \\ &\qquad w \in W = \{ w : w^l \le w \le w^u \} \\ &\qquad u \in U = \{ u : u^l \le u \le u^u \} \\ &\qquad p \in P = \{ p : p^l \le p \le p^u \} \end{aligned} \tag{5}$$

Inner level:

$$\begin{aligned} &\max_{\theta} AOS_d(z^*, y^*, \theta, x, \dot{x}, w, u, p^*, t) \\ &s.t. \quad \theta \in EDS \end{aligned} \tag{6}$$

Where Φ is the objective function, z is the vector of continuous design variables, y is the vector of binary design variables, θ is the vector of external disturbances and/or process uncertainties, x is the vector of state variables, w is the vector of output variables, u is the vector of manipulated variables and p is the vector of process parameters. h_i, $i \in E$ is the set of equalities defining process and controller models and g_j, $j \in I$ is the set of constraints defining the feasible operating region. The vectors z*, y* and p* are optimum decision variables found from the previous outer level. The vector θ^k is the worst disturbances/uncertainties combination found from the previous inner level. Z, X, W, U, P and EDS are the sets of all possible realizations of their respective variables, assuming uniform probabilistic distributions.

EDS, DOS and AOS_d are multi dimensional polyhedrons. DOS represents feasible operating space of the process and AOS_d envelopes dynamic responses due to possible combinations in EDS. The sizes of the polyhedrons are defined in terms of their convex hulls μ.

As shown in Eqs. 5 and 6, the algorithm used in this framework is iterative, consisting of outer and inner levels. The outer level solves dynamic Mixed Integer Non-Linear Programming (MINLP) problem to deliver both optimal operating condition and system structure. As the process responses to θ^k forms an AOS_d, the constraint r-OCI = 1 states 100% process feasibility by placing the AOS_d fully inside DOS.

The optimum values z*, y* and p* found from the outer level are investigated further in the inner level. This time, the process forms a new AOS_d as responses to all members of EDS. The convex hull of this AOS_d is transformed back to EDS, adding new members to θ^k for consideration in the next outer level. The algorithm iterates until all the operating conditions meet the feasibility assessments in spite of θ^k.

The AOS_d volume represents process controllability. It provides a General Integral Absolute Error (GIAE) value of the system. This controllability index involves all output variables, instead of single variable in conventional IAE.

The Case Study – A Five Effects Evaporator

Process Structure

The system in this study is a five-effects-evaporator associated with the liquor burning facility in the Bayer process, as shown in Figure 1. It consists of one falling film unit, three counter current forced circulation units and one super concentrator unit. The superstructure facilitates distribution of live steam to the last three stages and vapor mixing between 3rd – 4th stages. It allows choices of four or five effects structures as well as different control loops.

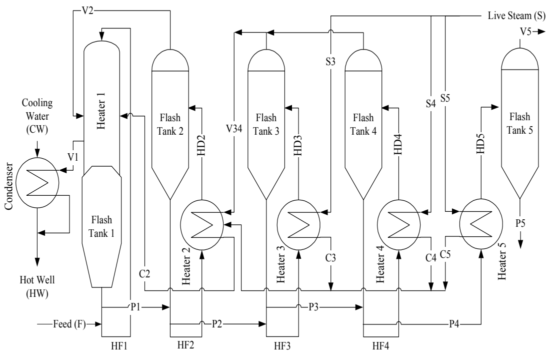

Figure 1 Process Flow sheet

The dynamics considered are associated with liquor level h_{Pi} (i = 1-5), concentrations C_{Pi} and temperatures T_{Pi} in flash tanks, assuming significantly faster changes elsewhere. Disturbances affecting the system include fluctuations on feed flow rate Q_F, composition C_F and temperature T_F. It is assumed that there is no uncertainty in process parameters. Heat transfers on heaters involve live steam $\dot{m}_{Si}$ or vapor from next stage(s) $\dot{m}_{Vi+1}$, condensates $\dot{m}_{Ci+1}$ and recycle streams $\dot{m}_{HFi}$.

The binary design variables of the process are the distribution of live steams, ys_3, ys_4 and ys_5. These variables determine the process structure, equivalently deciding the operation mode of four or five effects on reaching the target density. At least one of stages 3 or 4 shall operate; and if both are operating, vapor from both stages shall mix.

The objective function in this case is to minimize the operational and utility cost as follows:

$$\begin{aligned}\Phi &= +0.5\times 10^{-6}\,\dot{m}_{CW}^2 + y_{S3} + y_{S4} + y_{S5} + \dots \\ &\quad + 0.01(\dot{m}_{S3}^2 + \dot{m}_{S4}^2 + \dot{m}_{S5}^2)\end{aligned} \tag{7}$$

Subject to operational constraints as follows:

$$\begin{aligned}&g_{1-5}: h_{Pi,L} \le h_{Pi*} \le h_{Pi,U} \qquad i = 1,2,3,4,5\\ &g_6: 125.08\% \times \rho_F \le \rho_{P5}\\ &g_7: 1 \le y_{s3} + y_{s4}\end{aligned} \tag{8}$$

The subscripts L, U and * indicate lower, upper and optimum values, respectively. Meanwhile, disturbances Q_F, C_F and T_F are assumed as step functions with amplitudes varying within 10% of their nominal values.

Control strategies

Control design problems for this process include maintaining product density $\rho_{P5}(T_{P5},C_{P5})$ above the target density, e.g. 125.08% $\rho_F(T_F,C_F)$; and maintaining the liquor levels and final product temperature T_{P5} within operational limits. The possible manipulations are product flow rates Q_{Pi}, live steam rates $\dot{m}_{Si}$ and the vaporization rate in final stage $\dot{m}_{V5}$.

The levels are open loop unstable, while product's density and temperature are self-regulated; however, all variables are strongly interactive. Flow rates Q_{Pi} are favorable candidates to manipulate levels, because of their direct effects. However, flow rate manipulations also directly affect densities. This conflict can be worse if integral actions are applied on h_{Pi}-Q_{Pi} pairs, due to high fluctuation on Q_{Pi}. To assess the problem, several possible control loops for multi-loops PI-control are supplied to the process superstructure, which are listed in Table 1. Binary variables y_L are attached to reset times τ_i of level controls on stages 1-4, to both proportional gain and reset time on loop h_{P5}–Q_{P5}, and to all density and temperature control loops. More constraints are also added in the selection of best loops as shown in Eqs. 9.

Table 1 Control loops and parameters

Integer var.	Output variable	Manipulated variable	Kc	τ_i (hr)
y_{L1}	h_{P1}	Q_{P1}	-25	0.75
y_{L2}	h_{P2}	Q_{P2}	-25	0.75
y_{L3}	h_{P3}	Q_{P3}	-25	0.75
y_{L4}	h_{P4}	Q_{P4}	-25	0.75
y_{L5}	h_{P5}	Q_{P5}	-30	2
y_{L6}	ρ_{P5}	$\dot{m}_{V5}$	300	0.5
y_{L7}	T_{P5}	$\dot{m}_{S5}$	0.2	0.01
y_{L8}	ρ_{P5}	Q_{P5}	-5	0.01
y_{L9}	h_{P5}	$\dot{m}_{V5}$	-1.5	0.75

$$\begin{aligned}&g_{8-12}: 90\% \le h_{Pi*} \le 100\% \qquad i = 1,2,3,4,5\\ &g_{13}: 98\% \le T_{P5*} \le 102\%\\ &g_{14}: 99\% \le \rho_{P5*} \le 101\%\\ &g_{15-19}: 75\% \le Q_{Pi*} \le 125\%\\ &g_{20-22}: 50\% \le \dot{m}_{Si*} \le 150\%\\ &g_{23}: y_{L1} = y_{L2} = y_{L3} = y_{L4}\\ &g_{24}: y_{L5} = y_{L6}\\ &g_{25}: y_{L8} = y_{L9}\\ &g_{26}: y_{L6} + y_{L8} \le 1\end{aligned} \tag{9}$$

The framework was developed in MATLAB™. The geometric computation is performed by calling a C code *Qhull* (Barber *et al.*, 1996), the dynamics equations are solved using ODE/DAE solver in MATLAB and MINLP problem is solved using branch and bound method.

Results and Discussion

The framework converges in two iterations. The first iteration uses nominal disturbance values, and delivers optimal steady state open loop values. In the inner level, dynamic response to all possible disturbance combinations on closed loop process for 10 hours operating time is assessed. The responses form an 8-dimensional AOS_d, containing time, h_{P1-5}, ρ_{P5} and T_{P5} respectively. The critical disturbance combinations as the result of mapping AOS_d convex hull to EDS are the lowest and highest on their respective ranges. The second outer level selects the best loops and optimizes the closed loop process with consideration of critical disturbance effects. The feasibility of optimum structure and operating condition are verified in the second inner level. The results are shown in Table 2.

Table 2. Controllability assessment results

	Steady State Optimum	Closed-loop optimum
Φ	3.349	3.351
GIAE		2.523×10^{-13}
y_{S3}	1	1
y_{S4}	0	0
y_{S5}	1	1
y_{L1-4}		0
y_{L5-6}		0
y_{L7}		1
y_{L8-9}		1

The assessment shows that the target density is achieved by running stage 1-3 and stage 5, or four effects operation ($[y_{S3}\ y_{S4}\ y_{S5}] = [1\ 0\ 1]$). While reset times of level control in Table 1 do not affect product density and temperature profiles, smaller values do cause severe oscillations. On the other hand, P-only level controls ($y_{L1-4} = 0$) still keep the offsets well within limits whilst maintaining less fluctuation in flow rates. This condition facilitates easier control of product density and temperature. Product density is tightly manipulated by product flow rate. It leaves level control to vaporization, resulting in slow oscillation both on final level and steam consumption. Nonetheless, its gives the smallest GIAE compared to other alternatives. Overall, the recommended control strategy is P only for h_{P1-3}–Q_{P1-3} and PI for ρ_{P5}–Q_{P5} as well as T_{P5}–$\dot{m}_{S5}$ ($y_{L7-9} = 1$). This result also emphasizes the importance of the super concentrator, since its co-current configuration breaks interaction between product quality and those from previous stages. Any correction is easier done on super concentrator, as it will not be recycled to previous stages, therefore preventing unnecessary oscillations.

Conclusion

A controllability analysis for an industrial five effects evaporator system has been presented. It is demonstrated that the framework works well with highly nonlinear and interacting industrial evaporation process. The target density is achieved by four effects operation and the best control pairs for multi-loops PI controller are determined. Work is underway to include assessment of multivariable controllers in the process, inclusion of General Integral Absolute Error (GIAE) in objective function and the application of a generic function of disturbances.

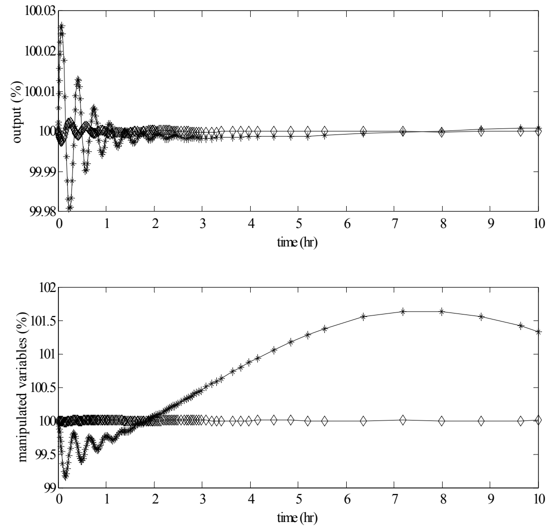

Figure 2 Optimum density and temperature responses to critical disturbance combination
◇- = ρ_{P5}-Q_{P5}, -- = T_{P5}-m_{S5}*

References

Bahri, P. A. (1996) *A New Integrated Approach for Operability Analysis of Chemical Plants*, Dept. of Chem. Eng. University of Sydney, Sydney, Australia.

Barber, C. B., Dobkin, D. P. and Huhdanpaa, H. T. (1996). The Quickhull Algorithm for Convex-Hulls. *ACM Transaction on Mathematical Software,* 22**,** 469 - 483.

Ekawati, E. and Bahri, P. A. (2001). Adaptation of Output Controllability Index within Dynamic Operability Framework. *In Proceedings of 6th IFAC Symposium on Dynamics and Control of Process Systems (DYCOPS-6).* Chejudo Island, Korea,

Kam, K. M. and Tade, M. O. (2000). Simulated Nonlinear Control Studies of Five Effect Evaporator Models. *Computers and Chemical Engineering,* 23**,** 1795 - 1810.

Sidrak, Y. L. (2001). Dynamic Simulation and Control of the Bayer Process. A Review. *Industrial and Engineering Chemistry Research,* 40**,** 1146-1156.

To, L. C., Tade, M. O. and Le Page, G. P. (1998). Implementation of a Differential Geometric Nonlinear Controller on an Industrial Evaporator System. *Control Engineering Practice,* 6**,** 1309-1319.

Vinson, D. R. and Georgakis, C. (2000). A New Measure of Process Output Controllability. *Journal of Process Control,* 10**,** 185 - 194.