HEURISTICS FOR CONTROL STRUCTURE DESIGN

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Abstract: A heuristic approach is proposed to solve material balance control problems for chemical plants. The heuristics are derived from structural and controllability analysis and validated through simulation cases. First, the plant analysis is decomposed into two parts: a reaction section and a separation section. Second, these two sections are combined to evaluate the heuristic procedure implementation to solve a plant-wide control problem. Some control structure designs based on the proposed heuristic procedure are tested in the Tennessee Eastman Case Study.

Keywords: plant-wide control, control structure design, decentralised control, snowball effect, heuristics.

1. INTRODUCTION

For the case where the control structure is well selected, decentralised control is a useful strategy to perform control targets. The advantages to apply decentralised control instead of centralized ones are: easy implementation, low cost, reliability and comprehensive operation.

The problems of decentralised control application are derived from interactions of coupled process. Mass and energy integration usually increase the coupling among the process variables, what makes more difficult to apply decentralised controllers. Therefore it is crucial for the success of a decentralized control strategy to select the right manipulated and controlled variables.

In this paper, four guidelines are proposed to develop good control structures for process control using decentralised controllers. The main idea is to identify which variables and variable ratios should be kept constant to eliminate the effect of the process disturbances in the process quality automatically.

The paper is structured as follows: First, an isothermal reactor is analysed to show how the residence time and inlet flowrates can affect the component material balance. Similar analysis is made for separation process and for a small process with a reactor, separation unit and a recycle stream. These examples are used to developed the guidelines, which are finally applied to propose new control structures for the Tennessee Eastman Case Study (Downs and Vogel, 1993).

2. DEVELOPMENT OF THE GUIDELINES

In this section are shown the model applied to develop the IO-controllability analysis of a reactor and a separator. The heuristics proposed in this paper are derived from these simple cases.

2.1 Reactor

The reactor considered is an isothermal CSTR, with the first order kinetic A $\rightarrow$ B. The corresponding mathematical model is given by:

$$\frac{dV_R}{dt} = u_1 - u_2 \tag{1}$$

$$\frac{d\left(V_R C_A\right)}{dt} = u_1 C_{A0} - u_2 C_A - kV_R C_A \tag{2}$$

The variables used in the model (1) and (2) are summarized in Table 1.

Table 1: Variables description for the reactor model.

Variable	Symbol	Unit
Reactor volume	V_R	m^3
Reactant A inlet composition	C_{A0}	$kgmol/m^3$
Reactant A outlet composition	C_A	$kgmol/m^3$
Flowrates	u_1 and u_2	m^3/h
Kinetic constant	k	h^{-1}

The RGA matrix is an effective tool to evaluate variables coupling. The system is decoupling when the main diagonal of the RGA matrix is close to 1 (Skogestad and Postlethwaite, 1996). If the controlled variables of the reactor are V_R and C_A, we have a 2X2 system and RGA matrix can be expressed by its 1,1-element, λ_{11}. If we choice the control pairs u_1-C_A and u_2-V_R (control structure CS1) or u_1-V_R and u_2-C_A (control structure CS2), we will have distinct values for the λ_{11}. These values can be analytically attained from linearization of (1) and (2), resulting the expressions (3) and (4).

$$\lambda_{CS1} = 1 - \frac{1}{\tau_R}\frac{1}{s} \tag{3}$$

$$\lambda_{CS2} = \frac{1}{\tau_R}\frac{1}{s} \tag{4}$$

In (3) and (4), λ_{CS1} and λ_{CS2}, are λ_{11} calculated for CS1 and CS2, respectively. The τ_R is the reactor residence time and s is the Laplace domain variable. From (3) and (4), we can conclude two remarks:

- **Remark 1:** The λ_{11} depends on the reactor residence time. If τ_R is not constant, it can change λ_{11}, keeping away from 1.

- **Remark 2:** The control structure CS1 is more appropriated than CS2 for high frequency, since $\lambda_{CS1} \rightarrow 1$.

From remarks 1 and 2 we see that it is interesting to propose a control structure, which can keep the reactor residence time constant as possible and to select the inlet flow rates to control the reactor composition.

2.2 Separation Unit

A simple model of one stage separation unit can be used to show how it is possible to control material balance without composition controllers.

For an ideal binary mixture this model permits to write the equation (5).

$$(\alpha - 1)(1 - \psi)x^2 + [1 + (\alpha - 1)(\psi - z)]x - z = 0 \tag{5}$$

From (5), α is the relative volatility, ψ is the vaporisation fraction, x is the liquid phase composition of the light component and z is the light component composition into the feed flowrate. Considering that there are no feed composition changes and knowing that the relative volatility is almost constant, x can be inferred by ψ. In the other words, the component material balance can be controlled by a fix ratio V/F (inlet vapour flowrate/feed flowrates).

A dynamic model proposed in the literature for one stage unit separation (Luyben, 1990) was implemented to show the composition response under feed flowrate changes. Considering hypothetical conditions, it is possible to show this behaviour when: a) the liquid phase composition has a feedback controller and b) a feedforward control strategy is implemented to fix the ratio V/F. These results are shown in the Figures 1 and 2. It is easy to see that composition variability is lower when the ratio V/F is direct controlled.

This idea can be extended to separation columns where the ratios inlet vapour flow rate/feed flow rate and reflux flow rate/feed flow rate are fixed by a feedforward strategy. This application and its effect on the plant economics are explored by self-optimising techniques for controlled variables selection procedures (Skogestad, 2000).

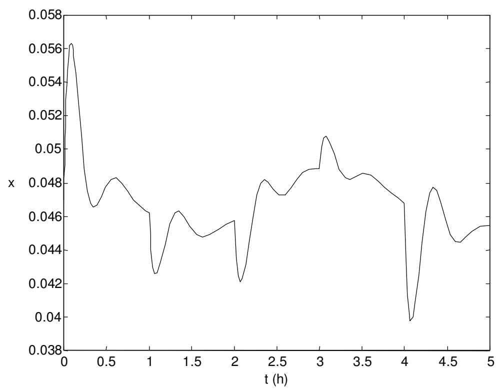


Fig. 1. Typical profile of the liquid phase composition change under flow rate disturbance when a feedback composition controller is used.

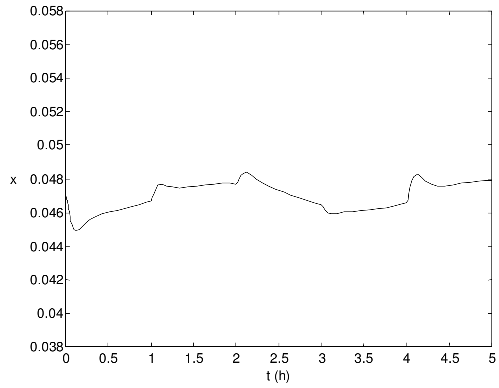


Fig. 2. Typical profile of the liquid phase composition change under flowrate disturbance when a feedforward strategy to fix the ratio *V/F* is applied.

A kind of the problem takes place when this control structure is implemented: under composition variations, it does not work. Solving this problem we can introduce a feedback composition controller through cascade configuration with feedforward strategy.

Base on these results two additional remarks can be written:

- **Remark 3:** We can implement feedforward control structures through constant flow rate ratios to control material balance under inlet flow rate changes.

- **Remark 4:** The feedback composition controllers can be introduced through cascade configuration with basic feedforward control structure.

2.3 Plant-Wide Control

In this section a hypothetical plant will be considered. The process flowsheet of this plant is shown in Figure 3.

The process is composed by liquid phase CSTR reactor, distillation column, and a recycle stream. The reactor is isothermal and the kinetic system is a single first order reaction, A → B. The process variables are described in the Figure 3.

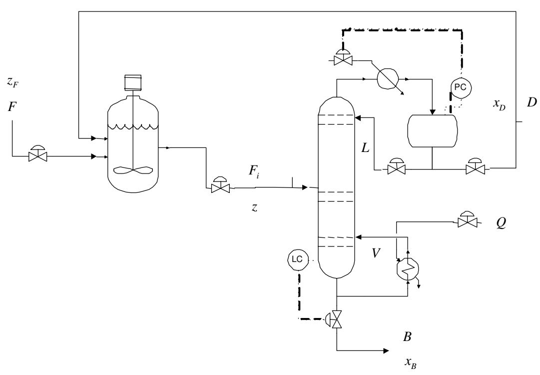


Fig. 3. Process flowsheet of the hypothetical plant.

In steady state we can write the material balance of this plant by equation (6).

$$F_i = \frac{F(x_D - x_B)kV_R}{kV_R x_D - F(z_F - x_B)} \tag{6}$$

The V_R in (6) is the reactor hold-up, expressed in molar base.

If the feed is the pure reactant A and considering an ideal separation, we have $z_F = 1$, $x_B = 0$ and $x_D = 1$. Thus, we can simplify (6), resulting (7).

$$F_i = \frac{FkV_R}{kV_R - F} = \left(\frac{k\tau_R}{k\tau_R - 1}\right)F \tag{7}$$

The expression (7) has been used to describe the snowball effect (Russel *et al.*, 2002). The snowball effect is a usual behaviour of systems with recycle streams with constant reactor inventory (Luyben *et al.*, 1998). Figure 4 shows two typical curves from expression (7) when the process inlet flowrate is increased. This trends shows the behaviour of the recycle stream, F_i, when V_R or τ_R are fixed.

The high increasing of the recycle stream is verified when the reactor hold-up is controlled at its set-point. This means that high changes of the recycle streams are expected to keep controlled compositions under inlet flowrate variations.

The control of the reactor residence time is a way to minimize the snowball effect. Figure 4 shows a linear behaviour of the recycle stream when the feed flowrate is increased, when the residence time is constant, whereas increases exponentially when the inventory is maintained constant. Therefore, there are two advantages to maintain the residence time constant. First the outlet flow increase linearly and second the outlet composition suffer small variation. Base on equation (7) it can be easily concluded that to maintain residence time constant, we need just to keep the F/F_i ratio constant. In other words, if it is controlled the F/F_i ratio, it is possible to operate the plant without snowball effect and with automatic composition control. These conclusions agree with the remarks 1 and 3. Now we can apply the guidelines summarized in the remarks 1 to 4, to build a control structure for the hypothetical plant.

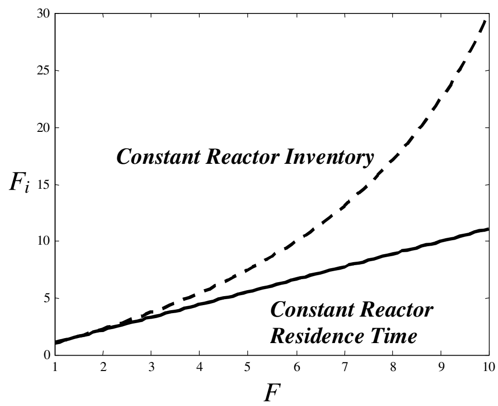


Fig. 4. Recycle stream behaviour.

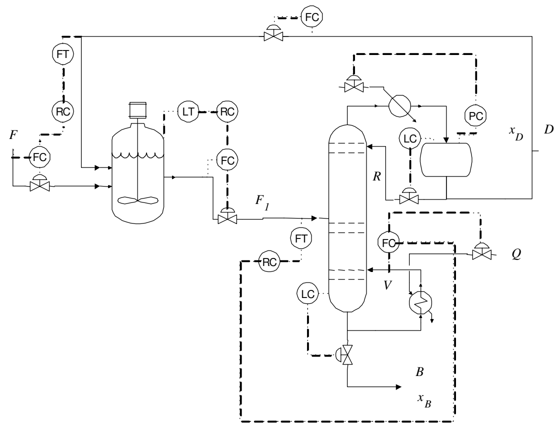


Fig. 5. CS1 flowsheet.

First, we introduce a feedforward ratio control to maintain reactor residence time constant. Second, flowrate ratios are implemented in the plant. The vapour flowrate to column/column feed flowrate and plant feed flowrate/recycle stream ratios are fixed through feedforward control. The recycle stream is chosen to set production rate. This control structure is called CS1 and it is shown in the Figure 5.

In Figures 7 and 8 the impurity composition of the column product shows a dynamic behavior when two kind of disturbances are applied: 10% production rate increasing and 10% feed composition increasing. We can see that CS1 has a good performance when the production rate is increased, but it does not work under composition disturbances. This problem is solved by introducing feedback composition controllers through a cascade configuration. This complete control structure, CS2, is shown in the Figure 6.

Based on all discussions made until here, we can write three relevant heuristics for plant wide control structure design:

- Use a control configuration (feedforward or feedback) to fix the reactor residence time. For this, the reactor hold-up (inventory) can change.
- Use a feedforward control configuration to fix flowrate ratios.
- The two heuristics above compose the basic design to control the inventory structure. The composition control is made through a cascade configuration

3. TENNESSEE EASTMAN PROBLEM

The Tennessee Eastman case study is described in (Downs and Vogel, 1993) and the process flowsheet is shown in Figures 9 and 11.

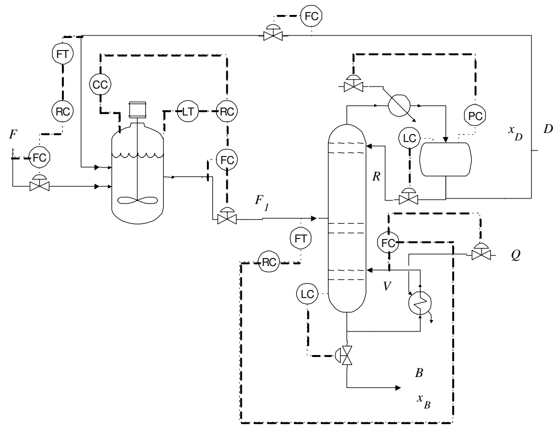


Fig. 6. CS2 flowsheet.

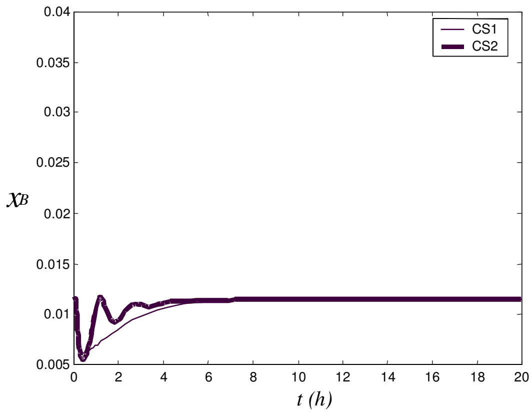


Fig. 7. Dynamic performances of control structures CS1 and CS2 when 10% of the production rate is increased.

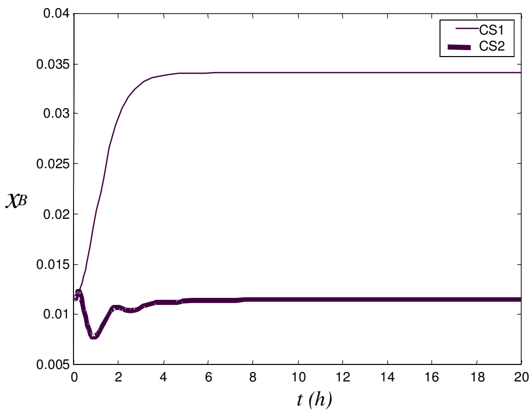


Fig. 8. Dynamic performances of control structures CS1 and CS2 when 10% of the feed composition is increased.

The kinetic system of the process is comprised by exothermic reactions given by the chemical reactions (8) to (11).

$$A(g) + C(g) + D(g) \longrightarrow G(liq) \tag{8}$$

$$A(g) + C(g) + E(g) \longrightarrow H(liq) \tag{9}$$

$$A(g) + E(g) \longrightarrow F(liq) \tag{10}$$

$$3D(g) \longrightarrow 2F(liq) \tag{11}$$

The main reactions are shown by (8) and (9). These reactions yield the main product constituted by G and H components, therefore the G/H ratio defines product quality. The reaction is gas-phase on the catalytic surface and the liquid-phase products are vaporized and partially condensed in a separator vessel. The gas-phase is compressed and recycled to reactor and the liquid-phase feeds the top of the separation column. A portion of the gas-recycle is purged to control inert inventory. This purge performs a meaningful loss of the plant.

The Tennessee Eastman Plant is an unstable system showing a RHP pole related to reactor temperature. The reactor is jacketed and cooling water cools the reactor content. Thus the low level of the reactor can be harmful to the system stability. When reactor level decreases below 40%, it takes place a high heat increasing (Farina et al., 2000) and below 10% the system exhibits a temperature runaway (Wu and Yu, 1997).

Next, we show two control structures applying the heuristics concepts above. Control performance is evaluated by variability of the product and reactor compositions ($\pm$5% is specified), constraints control and purge flow rate. The simulation examples apply 15% production rate increasing and model implementation is derived from Ricker's Tennessee Eastman Challenge Archive (http://depts.washington.edu/control/LARRY/TE/download.html).

3.1 Control Structure TE-CS1.

The control of the reactor residence time of the Tennessee Eastman plant is a not obvious task. Thus, it was implemented an indirect control applying an inventory-recycle ratio. The reactions take place in gas phase on catalytic surface, then the inventory considered to control must be the reactor vapor holdup. The ratio applied to perform this feedforward control is shown by Figure 9.

The flowrate ratios were applied as feed flow rates to recycle stream ratio. A similar structure with ratio control structure was proposed by Ricker (Ricker, 1996) where it was implemented for all flowrates a ratio control. The difference between the Ricker's control structure and the proposed here is the variable reactor inventory, which is used to keep the residence time constant.

The production rate is changed through recycle flowrate and other inventories are controlled by outlet flows. These alternatives have good io-controllability as shown by Farina et al., (2000).

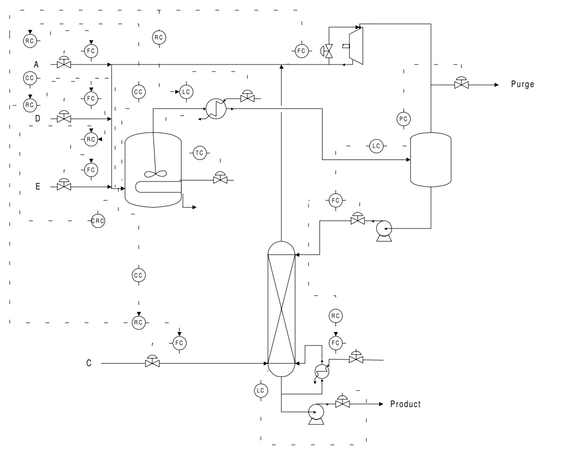
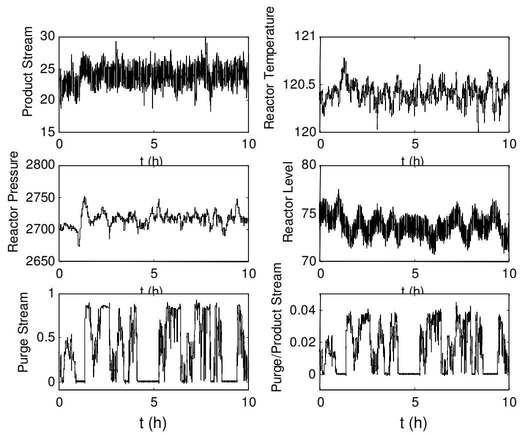
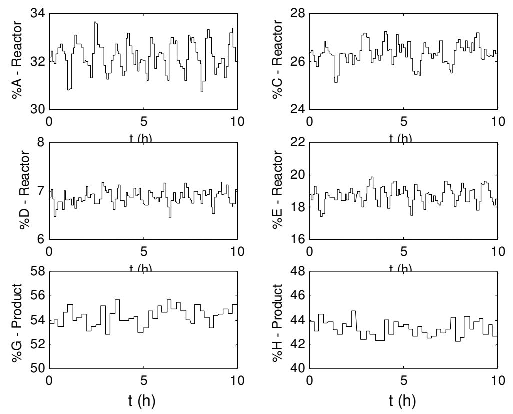


Fig. 9. Flowsheet of the control structure TE-CS1.

The feedback composition controllers are implemented through cascade configuration. The simulation results are shown in Figure 10.



(a)



(b)

Fig. 10. Simulation results of TE-CS1 for: (a) regulatory control and (b) composition control.

3.1 Control Structure TE-CS2.

In this CS, the reactor inventory is not directly controlled, since no reactor level control loop is used. The production rate is changed by separation drum effluent stream. Thus, the reactor inventory decreases when the plant production rate is increased. The control structure is sketched in Figure 11 and the simulation results are shown in the Figure 12. From these results, there is a higher reactor level variation when we compare it with the first alternative (i.e., *TE-CS1*), but we can see a fast production rate change showing smooth variations.

From these control structures we can see that it is possible to attain the targets of the Tennessee Eastman Plant through energy stabilisation and material balance control, with an almost constant reactor residence time. These control structures also are reliable and comprehensive for the operation staff.

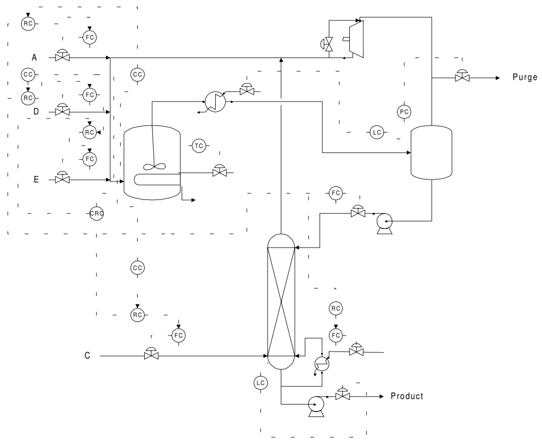


Fig. 11. Flowsheet of the control structure TE-CS2.

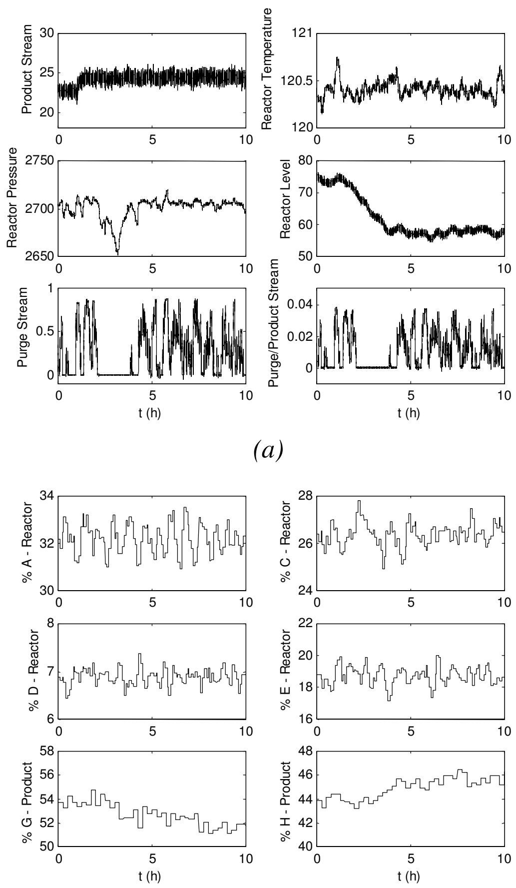


Fig. 12. Simulation results of TE-CS2 for: (a) regulatory control and (b) composition control.

4. CONCLUSION

This paper showed how it is possible to derive guidelines for control structure design based upon a model analysis of simple and general hypothetical plants. These heuristics, when they are applied in a real plant, produce very good control performance and are easy to apply. Finally, the proposed heuristic approach is used to develop two new control structures for the Tennessee Eastman Challenge Control Problem.

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