Exercise 12 Meeting The Frequency Specifications On Certain Flows And Tuning The Level Controllers

I. OBJECTIVE

The objective of exercises 10 to 14 is to demonstrate how one can tune loops once a plantwide control architecture has been selected. The Tennessee Eastman simulation [1] is used and in exercise 14 several candidate architectures are evaluated. Once the 10 inner cascade loops are tuned, then the level loops can be tuned. This exercise deals with tuning the level loops, and meeting the frequency specifications on certain flows. Before loops of comparable speed to the level loops can be tuned the level loops need to be closed. If the level loops are not closed, their integrating nature results in the maximum or minimum level specifications being violated during the bump tests for other loops. When a level specification is violated the Tennessee Eastman simulation shuts down.

II. CONTROL TECHNOLOGY

a) OVERVIEW: The following nonlinear dynamic model is used to simulate the Tennessee Eastman process [1]:

$$\dot{x} = f(x,u) \quad (1)$$

$$y = g(x,u) \quad (2)$$

where x is the state vector, u is the vector of manipulated variables, and y is the vector of process measurements. The vector, y, contains all the available process measurements, including those for the 10 inner cascade loops. Several MATLAB m-files have been written to interface with the FORTRAN simulation of eqns. 1 and 2. These m-files allow one to carry out reaction curve tests on the process, and to simulate it with various loops closed. Once a plantwide control architecture is decided upon, its loops can be tuned by starting with the fastest loops and proceeding to the slowest loops. All the controllers used in exercises 10 to 14 are PI controllers. They are implemented in velocity form as:

$$\Delta mv(t) = K_C(\varepsilon(t) - \varepsilon(t-1) + \varepsilon(t)\Delta t / T_R) \quad (3)$$

where Δmv is the change in manipulated variable, e(t) is the error at time t, K_C is the controller gain, and T_R is the reset time. The integration time step used in the simulation is 1 sec. Note that some of the routines ask for simulation times in seconds, and others in minutes. Results for tuning the 10 inner cascade controllers are given in Table 1.

Table 1. Final Tuning Constants From Closed Loop Tests

Valve	K_C	T_R(min)
D	.026(%/kgh^{-1})	.10
E	.017(%/kgh^{-1})	.10
A	150(%/kscmh)	.075
C	10.(%/kscmh)	.10
Purge	200(%/kscmh)	.06
Sep. Exit	2.0(%/m^3h^{-1})	.12
Product	3.0(%/m^3h^{-1})	.12
Steam	2.5%/kgh^{-1})	1.5
React. CW	-10(%/^oC)	1.0
Cond. CW	-8.5(%/^oC)	2.6

The inner cascades involving flows with frequency constraints need to be considered. There are 4 such inner cascades and they involve the A, D, and C feeds and the product flow. Table 2 gives the frequency constraints placed on these loops in the problem statement [1].

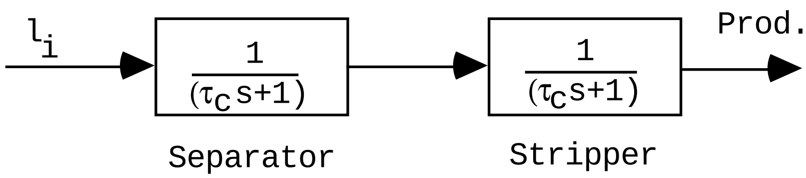
Table 2 Frequency Specifications On Flows

Flow	Frequency Range h^{-1}	Comments
A Feed	8-16	Protect from Variability
D Feed	8-16	Protect from Variability
Product	8-16	Less than ±5% Variability
C Feed	12-80	Minimize Variability

First consider the A and D flow inner cascade loops. To meet the frequency specifications on the A and D loops one can proceed as follows. First, the inner flow loops for these variables are tightly tuned. Then normal bump testing is used to tune the outer loops that manipulate the set points of the A and D flows for control. Next, each of the disturbances and expected set point changes is checked to see if it results in a violation of the frequency specifications, given in Table 2. If a violation occurs then the outer loops need to be detuned. The specifications given in Table 2 are somewhat vague, and the comments "protect from variability" need to be quantified further in order to decide if a violation has occurred, and how much detuning is necessary. Mc Avoy and Ye [2] have discussed how to use Fourier series techniques to assess the frequency content of flows. Since this process of detuning the outer loops utilizing the A and D feeds is somewhat complicated, and the specifications in Table 2 are somewhat vague, and the various upsets and set point changes do not produce very high frequency effects in the A and D flows, this aspect of the Tennessee Eastman test problem is not addressed in this exercise. For actual application, the frequency constraints on the A and D flows definitely need to be evaluated.

Next consider the product flow. In all plantwide control systems to be studied, the product flow is used to control the stripper level. The product flow is constrained in how fast it can be manipulated. Product flowrate changes greater than ±5% with significant frequency content in the range 8-16 hr^{-1} are to be avoided. To meet this specification one can proceed as follows. To smooth out fluctuations in the product flow, both the capacity of the separator and the stripper reboiler can be used, and a form of averaging level control applied. Assume that the liquid level controllers for both systems are tuned so that each has an identical first order response as illustrated in Figure 1.

Figure 1



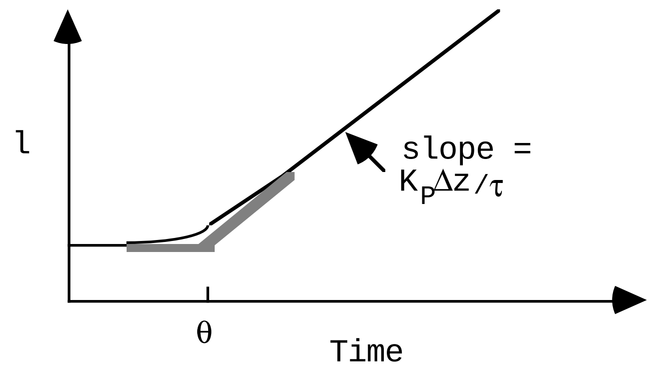
Further assume that a unit increase in the liquid entering the separator, l_i, results in a unit increase in product flow, i.e. the overall system has a gain of 1.0. If l_i is a unit sinusoid then the specifications in Table 2 can be taken to mean that the resulting magnitude of the sinusoidal change in product should be .05. For a second order system with identical time constants, the magnitude of its frequency response is:

$$M = 1/((\tau_c\omega)^2 + 1) \qquad (4)$$

Using the most conservative specification of ω, and M = .05, allows one to calculate τ_C. Once τ_C is known, the stripper and separator level loops can be tuned in a closed loop manner to achieve this value. For all three level loops in the Tennessee Eastman process a large reset time, T_R = 200 min. is used. For the stripper and separator level loops, increasing the value of K_C in these loops decreases τ_C, and vice versa. Both loops can be tuned to give a 63% response time at t = τ_C for a step in their setpoint. One can start with the stripper level/product loop, and then tune the separator level loop.

After the stripper and separator level loops are tuned, the reactor level loop can be tuned, and the tuning should be tight. Tight tuning can be accomplished by using a closed loop approach and step changing the reactor level set point. The resulting response of the reactor level should not have too much overshoot. To generate an initial estimate for the tuning parameters, the m-file *primary* can be used. The set point of the inner cascade loop manipulated for reactor level control can be bumped to develop a reaction curve. The resulting curve will look like that shown in Figure 2.

Figure 2. Reaction Curve
Level Variables



The level does not reach steady state, but rather continues to increase or decrease. The following transfer function model is used to describe the response shown in Figure 2:

$$l(s) = \frac{\Delta z}{s}\left(\frac{K_P}{\tau s} e^{-\theta s}\right) \qquad (5)$$

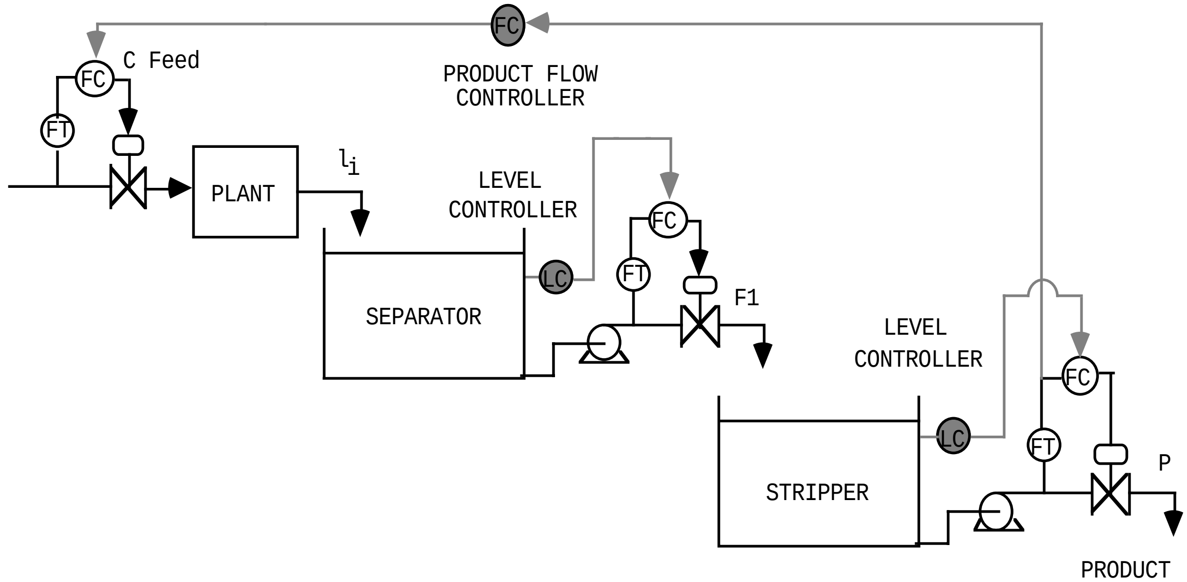
where Δz is the size of the step set point change in the flow used for level control, and the term in parentheses is the process transfer function. For the transfer function given by eqn. 5, a controller with only a small amount of integral action can be used, and tight control can be achieved by using a large proportional controller gain. As illustrated in the exercises, the process itself contains an integrator (1/s), and a proportional only controller for level loops gives no offset for set point changes. For disturbances, the small integral action and large controller gain eventually brings the level back close to setpoint as well. Small integral action can be achieved by setting T_R = 200 min. For approximately proportional only control of a process described by an integrator with dead time, initial tuning values for Kc can be calculated from the expression:

$$K_c = .10 / ((K_p / \tau) * \theta) \qquad (6)$$

From the reaction curve K_P/τ and θ can be determined. The value of K_C obtained from eqn. 6 should be considered as a rough starting value which needs to be fine tuned under closed loop conditions. The closed loop level response can be checked using *primary* , and if necessary the value of K_C can be adjusted. For controlling the reactor level, two manipulated variables are considered, the E-Feed set point, and the Condenser CW Temperature set point.

In all schemes to be tested the C feed is used to control the product flow. Tuning the C feed loop is complicated by the fact that one of the disturbances specified in the Tennessee Eastman problem involves a loss of pressure in the C feed header, IDV(7). Since it is important for the process to recover quickly from such an upset, and since this upset can be corrected quickly by the C feed inner cascade, this loop's tuning should be tight. The use of the C feed for product flow control is shown schematically in Figure 3. Changes in the C feed result in changes in the amount of products produced in the reactor, which in turn results in changes in the liquid feed, l_i, to the separator. In order for the C feed to affect the product flow, both the separator and the stripper level loops must be closed, as illustrated in Figure 3. The C feed acts through both of these loops which need to be detuned to meet the frequency specifications on the product flow. If for example the stripper level loop is open, then changes in the C feed have no effect on the product flow. This is an example of pairing on a zero RGA. The fact that the two level loops have been detuned means that the product flow frequency specification will also be met when the C feed is used for production rate control. At this point the C feed loop has been tightly tuned as one of the inner cascade loops. As discussed above the C feed loop needs to be tightly tuned in order to reject disturbances in the C header pressure (IDV(7)) as fast as possible. However, since a 1% change in product flow requires a 1% change in the C feed, step changing the setpoint of the product flow will result in an approximately equal percent step in the C feed, which in turn will violate the frequency specification for the C flow given in Table 2. One solution to this problem is to allow only ramp set point changes in the production set point. Such an approach allows the C feed loop to function effectively in rejecting header pressure upsets, and at the same time meet the frequency specification placed on the C feed. How to determine the allowable slope of the ramp for the production rate is left as an exercise. Here the tight C feed product flow loop tuning given in Table 1 is used.

Figure 3 Schematic Of Tennessee Eastman Plant



III. COMPUTER EXERCISE

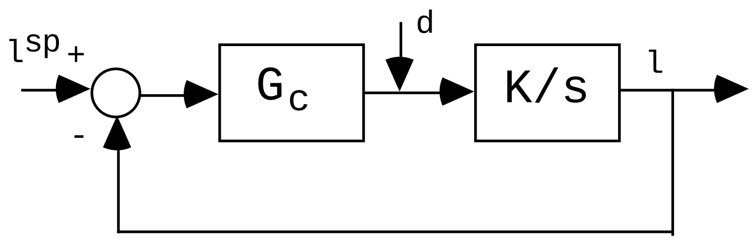
For loops involving flows with frequency constraints, tuning requires the use of the most conservative frequency. For the product flow calculate the value of τ_C that is most conservative from eqn. 4. The m-file *primary* can be used to tune the stripper and separator levels to achieve this value of τ_C in closed loop operation. A 1% step change in level can be used. The reset time, T_R, for each loop should be set to 200 min. The controller gain can be varied for each loop to achieve the desired first order response with a time constant equal to τ_C. Report the final values of K_C for each loop. Next *primary* can be used to carry out a bump test on the reactor level. Two manipulated variables should be studied for this bump test, the E-Feed setpoint and the Condenser CW Temperature set point. Tune the reactor level in closed loop operation to have tight performance for a step change of 1% in its set point. A reset time , T_R = 200 min, can be used. This level loop should not exhibit much overshoot. Report the final values of K_C for each manipulated variable.

IV RESULTS ANALYSIS

What is the value of τ_C that is required for the stripper and separator level loops? What values of K_C are required for stripper and separator level loops to achieve this τ_C. For manipulating the E-Feed to control the reactor level what value of K_C should be used? How fast can the reactor level be changed if the E-Feed is used? For manipulating the Condenser CW Temperature to control the reactor level what value of K_c should be used? How fast can the reactor level be changed if the Condenser CW Temperature is used? Is there a speed advantage in using either the E-Feed or the Condenser CW Temperature for reactor level control? Explain.

V. EXERCISES

1. Consider the block diagram shown below for a level controller:



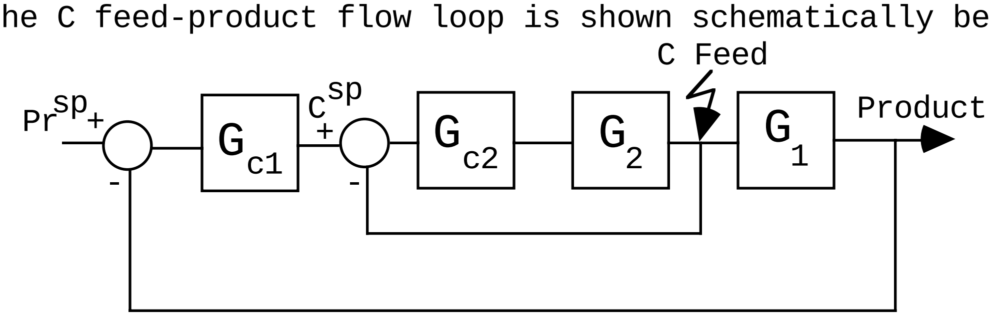
a) Prove that if $G_c = K_c$ that no offset in level, l, occurs if l^{sp} is step changed.

b) Calculate how much offset results in l if a step change in $d = \Delta$ occurs and $G_c = K_c$.

2. Discuss what assumptions are necessary for the system shown in Figure 1 to have a unity gain. Based on the information given in the problem statement [1], is a unity gain reasonable? Explain.

3. For the 3 outer loops illustrated in Figure 3, the 2 level controllers and the product flow controller, write down their open loop gain matrix, showing non-zero elements simply as x's, and zero elements as 0. For the level variables use their rate of change as measurements. Based on this open loop gain matrix what is the RGA for the system?

4. The C feed-product flow loop is shown schematically below:



If the product set point, Pr^{sp} undergoes a step change, then C^{sp} will also be step changed through the proportional term in G_{c1}. Since the C feed loop is tightly tune to reject pressure upsets in the C header, the C feed will respond much too fast to steps in Pr^{sp}. To meet the frequency specifications on the C feed given in Table 2, ramp changes in Pr^{sp} can be used. If "minimize variability" in Table 2 is defined as no more than ±5% content at a given frequency, and if a steady state product change of -15% is desired, estimate the maximum slope of the ramp in Pr^{sp} that should be used. For purposes of calculation it can be assumed that the product measurement is constant, i.e. the error signal to G_{C1} comes only from the step setpoint change, and that the C feed loop responds instantaneously. For G_{C1} use $K_C = .597$ and $T_R = 30$ min.

VI. REFERENCES

[1] Downs, J. and Vogel, E.,"A Plant-Wide Industrial Process Control Problem", Computers and Chemical Engineering, **17**, 245-255 (1993).

[2] Mc Avoy, T., and Ye, N., "Base Control For The Tennessee Eastman Problem", Computers and Chemical Engineering, **18**, 383-413 (1994).