CHAPTER 8: ADVANCED CONTROL STRUCTURES

8.1 CASCADE CONTROL

Cascade control is one of the most successful configurations to improve the performances of the single-loop feedback control. It can provide more effective control by reducing both the maximum deviations and integral error for the disturbances response.

The concept of cascade control was introduced in chapter 3. The block diagram for cascade control is shown in Fig. 3.6 which consists of two control loops; the inner and the outer loop. The primary (also called outer, master) controller maintains the primary variable, say y_1 at its set point by adjusting the set point y^{sp}_2 of the secondary controller. The secondary (also called inner, slave) controller responds both to the output of the primary controller and to the secondary controlled variable y_2.

The key point in cascade control is the selection of secondary variable. Two guidelines must be observed:

- The secondary variable must indicate the occurrence of an important disturbance
- The secondary variable dynamics must be faster than that of the primary variable.

Let us study the cascade control through an example. In the main menu shown in Fig. 1.13, select “Multi-stage Flash Desalination” . This will bring the following submenu:

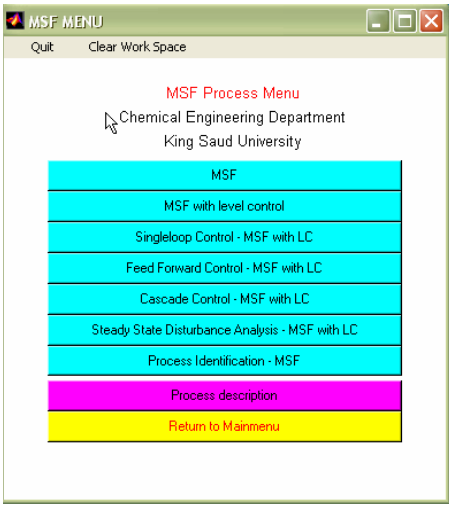


Fig. 8.1 The MSF Process Menu

Selecting "Singleloop control" in the MSF menu shown in Fig. 8.1 will bring up the following Simulink module:

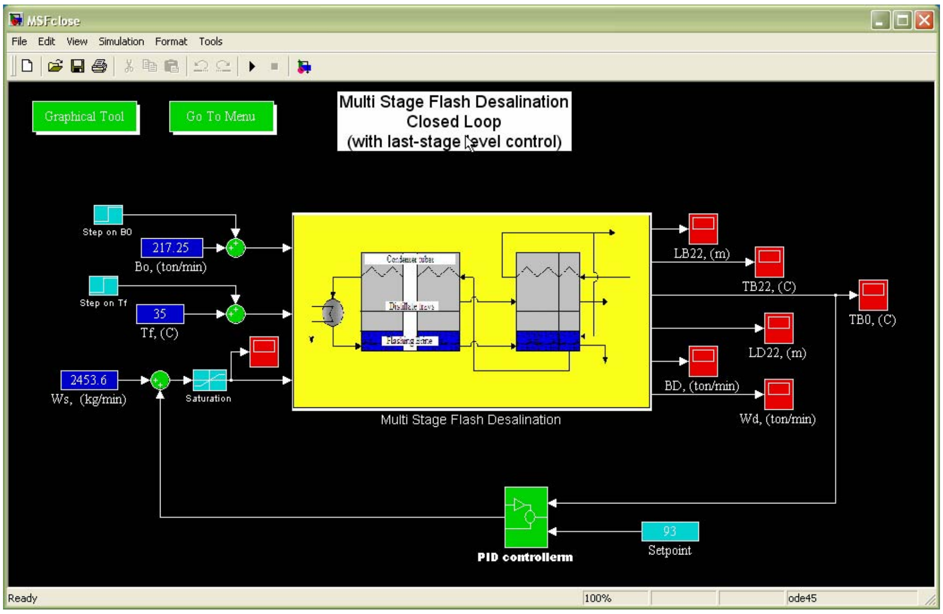


Fig. 8.2 Single loop control for MSF process

The module shown in Fig. 8.2 illustrates a control loop that regulates the top brine temperature, T_{B0} by manipulating the steam mass flow rate, W_s. The top brine temperature plays important role in the MSF operation as it affects the production rate, W_d significantly. Since the distillate product W_d, is the main output of the MSF, one may try to control the distillate product via manipulating the top brine temperature. However, the control objective can not be achieved directly. Therefore, one can work through a cascade loop. Selecting "cascade control" in Fig. 8.1 will activate the module shown in Fig. 8.3. The module shown in Fig. 8.3 contains a cascade control loop. The inner loop in that cascade structure is the T_{B0}-W_s control loop. This is the standard loop shown in Fig. 8.2. However, the set point of this loop is adjusted from an outer loop, which is the one designated for controlling the distillate product. Hence, the outer loop has W_d as its controlled variable and T_{B0}^{sp} as its manipulated variable.

Cascade control can use the standard PID feedback controllers in the two loops. The secondary loop must have the proportional mode but it does not require the reset mode. Integral model may be used in the secondary controller if it is desired to suppress completely the disturbance entering the primary or when the primary controller is not in operation (sensor not functioning or calibrated, etc). Derivative modes are not advised in the secondary loop since the derivative action is designed to overcome some lag in the controller loop and if applied to set point changes may result in excessive valve motion and overshoot. The cascade controller is tuned in a sequential manner. The secondary controller is first tuned satisfactorily before the primary is tuned. Conventional tuning guidelines for PID, such those discussed in chapter 6, apply for both control loops.

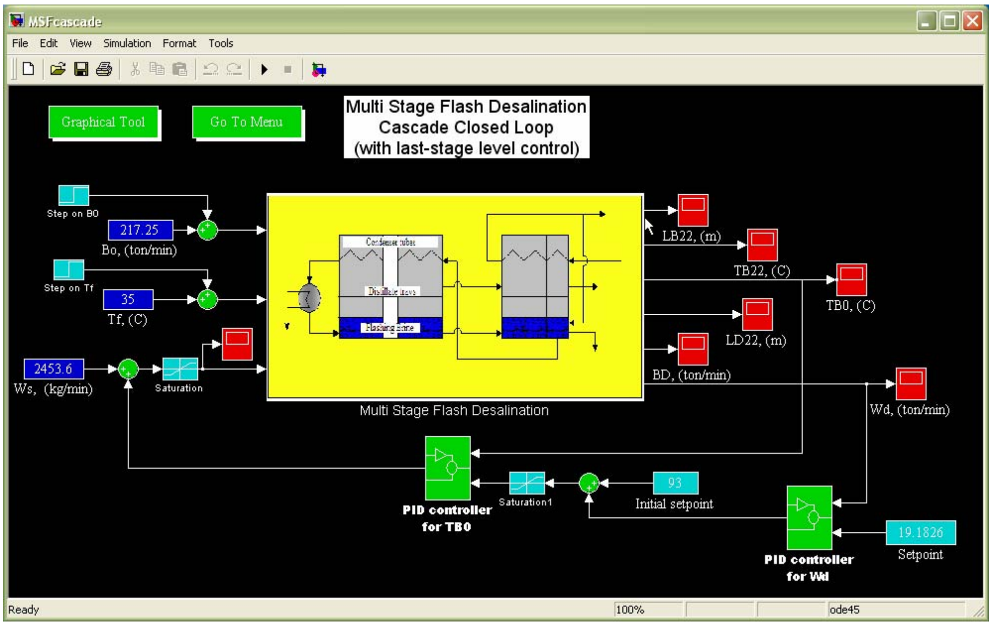


Fig. 8.3 Cascade Control Structure for MSF process

For satisfactory cascade control application, the inner loop must have faster dynamics. To check the speed of response to the inner loop, one can step change the steam mass flow rate with all loops disabled. As mentioned in chapter 6, disabling the control loop is achieved by settings all PID parameters to zero in the designated boxes, i.e. *PID controller for* T_{B0} and *PID controller for* W_d.

Introducing a step change in the steam mass flow rate result in the reaction curve depicted in Fig. 8.4. The curve indicates that the T_{B0}-W_s loop has a time constant of 20 minutes. It is hard to estimate the response speed of the W_d-T_{B0} loop because the top brine temperature can not be stepped independently. Nevertheless, the effect of Ws on T_{B0} is direct and its time constant controlled by the heat capacitance of the brine re-boiler. On the other hand, the effect of T_{B0} on W_d is transmitted through the 19 flash stages. This means that the time constant would be the accumulation of mass and heat capacitance of all stages. According to this reasoning, it is evident that the speed of response for T_{B0}-W_s is faster than that for W_d-T_{B0}.

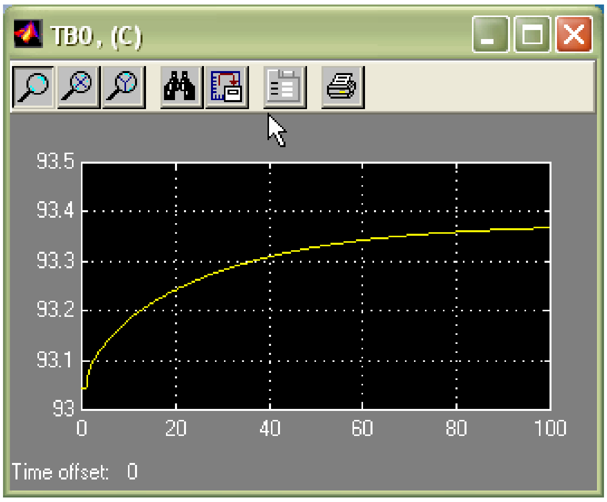


Fig. 8.4 The open loop response for T_{B0} to step change in W_s.

Fig. 8.5 demonstrates how the process responds to a disturbance of -5 ^{o}C in the feed temperature of the sea water without control.

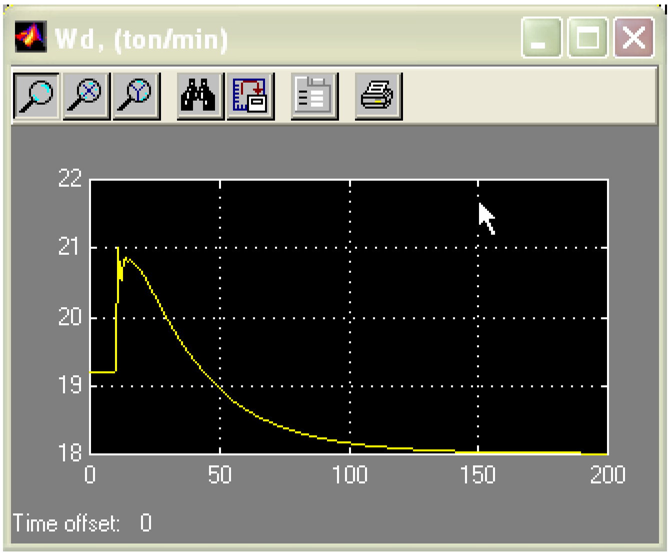


Fig 8.5 Product response to disturbance in feed temperature.

The result indicates a drop of more than one ton of distillate water. The situation demands a good control system. Fig. 8.6 illustrates how the process behaves when a single control loop like that shown in Fig. 8.2 is involved. The PI settings for that loop is k_c = 50, k_I = 10, k_d = 0. The feedback reaction in Fig. 8.6 demonstrates an excellent control of the top brine temperature but the product is left without control. Although the product is increased, the feedback performance is considered poor from control point view.

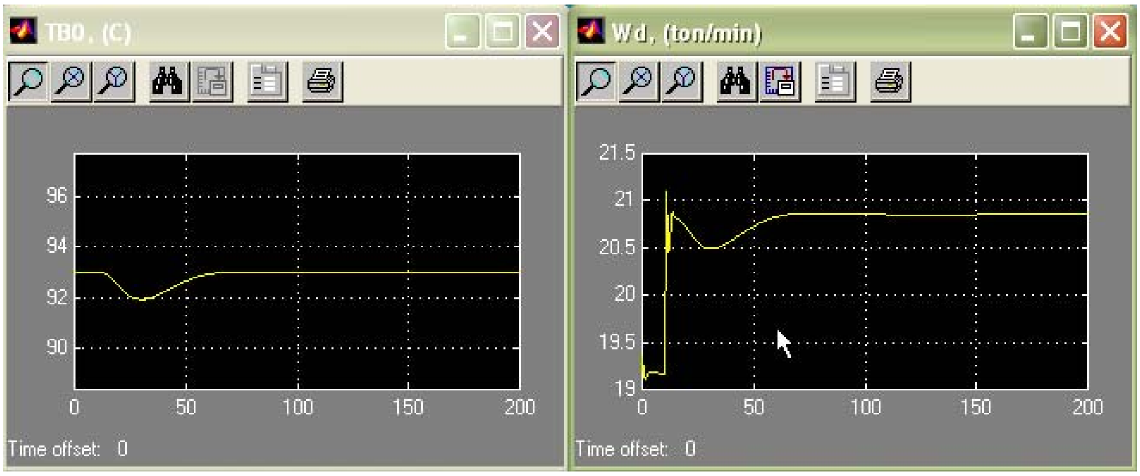


Fig. 8.6 Product response to disturbance when single loop is involved

Now examine Fig. 8.7 which shows the process feedback performance when cascade control is installed. Here, the product flow rate, W_d is well maintained at its reference value. This is achieved by manipulating the top brine temperature to a new value of 88 ^oC. Note that maintaining the top brine temperature at its initial value is not necessary because it is not considered as a desired process output.

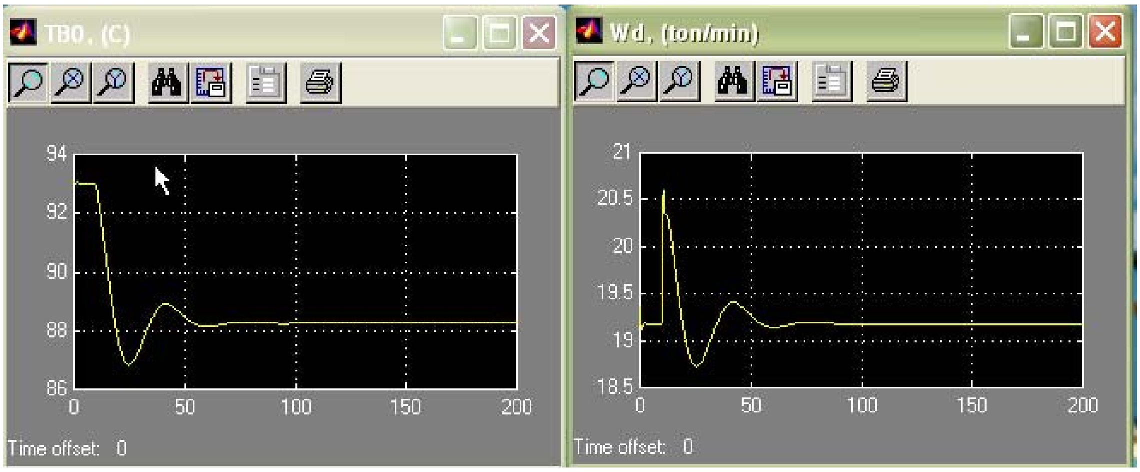


Fig. 8.7 Product response to disturbance when cascade control is involved.

The user can carry several control analysis on this cascade loop. For example, one can test the effect of tuning the inner and/or outer loop on the overall performance. This teaches how these two loops are interacting and hence care should be taken when tuning such loops. One can also test how the overall control loop performs when disturbances enter the process. This teaches him how cascade loop in comparison to conventional loop can improve the feedback performance. Innovative users can create similar cascade control structure for the other case studies and examine their performance.

8.2 FEED FORWARD CONTROLLER (FFC)

Feed forward control attempts to enhance the performance of the single loop feedback control by making use of an additional measurement of process input as shown in Fig. 3.5. The implementation block diagram for FFC is shown in Fig. 3.6 and the design equation is given by (3.12) and (3.13). Feed-Forward should be used when feedback control does not provide satisfactory performance and a measured feed-forward variable is available. Comparison between the feedback and feed forward controllers is listed below.

Table 8.1 Comparison between Feedback and feed forward controllers

Advantages	disadvantages
Feed-Forward	
1. acts before the effect of a disturbance has been felt by the system 2. Is good for slow systems 3. It does not introduce instability in the closed-loop response	1. Requires identification of all possible disturbances and their measurement. 2. Can not cope with unmeasured disturbances. 3. Sensitive to process parameter variation. 4. Require good knowledge of the process model
Feedback	
1. Does not require identification of all possible disturbances and their measurement. 2. It is insensitive to modeling error. 3. It is insensitive to parameter changes.	1. It waits until the effect of disturbance is felt by the system. 2. It is unsatisfactory for slow systems. 3. It may create instability in the closed-loop response.

All case studies of the PCLAB contain feed forward design tutorial. For demonstration purposes we will examine the feed forward controller design for the evaporator process. Refer to the evaporator menu shown in Fig. 1.16 and select the feed forward controller. The following Simulink module shall appear:

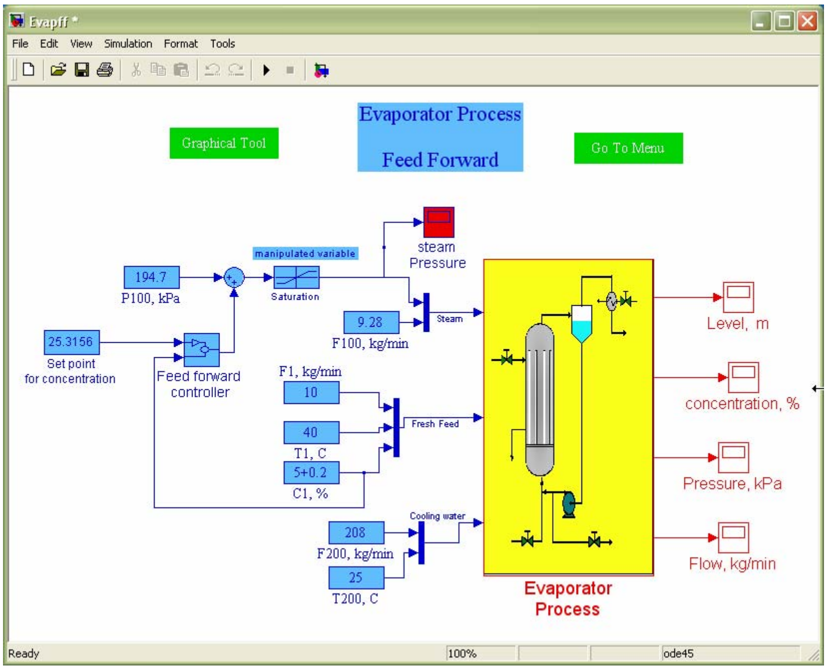


Fig. 8.8 feed forward Controller Structure for the Evaporator Process

Fig. 8.8 depicts a single FFC loop linking the disturbance C_1 (feed concentration) with the steam pressure. It should be noted this structure is not unique. The user can create, or replace it with, another FFC loop. According to the design equation discussed in chapter 3, the feed forward control block require the determination of the static gain and time constant for the manipulated variable and disturbance variable. These values can be inserted directly into the feed forward block as shown in Fig. 8.9. The dialog box for the feed forward controller shown in Fig. 8.9 can be activated by double clicking the "feedforward controller" icon.

In order to complete the FFC design, one needs to determine the value of the FFC parameters, i.e. k_p, k_d, τ_p and τ_d. these parameters can be computed using the reaction curve method discussed in chapter 3 and implemented in chapter 5. By step testing the product concentration to pre-defined changes in the steam pressure and feed concentration, the following were estimated:

Table 8.2 FFC tuning values

C-P100	$k_p = -0.018$; $\tau_p = 64$ s
C-C1	$k_d = 4.75$; $\tau_d = 22$ s

Next we insert these values in the feed forward parameter block and test the process for a step change of 0.2% in the feed concentration C_1. Fig. 8.10 illustrates the process response to the load change. Part (a) of the figure shows how the product concentration moves away from its nominal value when no controller of any type is involved. Part (b) demonstrates that the product concentration performance can be improved when a typical feedback controller is used. A typical PI controller is used with the controller gain and integral time magnitudes are set to the values found by Ziegler-Nichols method as listed in table 6.3. The last part of the figure depicts the process performance when FFC is implemented using the tuning values listed in Table 8.2. One can observe that the FFC response has less overshoot and faster settling time than those of the feedback controller. Nevertheless, FFC is expected to provide much superior performance than that of the feedback. The unsatisfactory performance of the FFC may contributed to the accuracy of the estimated parameters in Table 8.2.

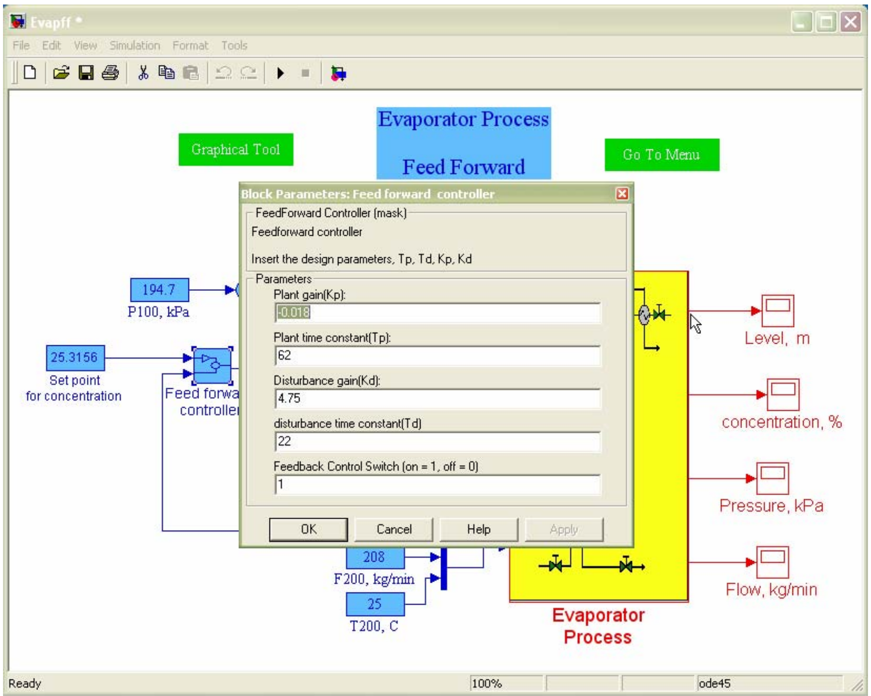
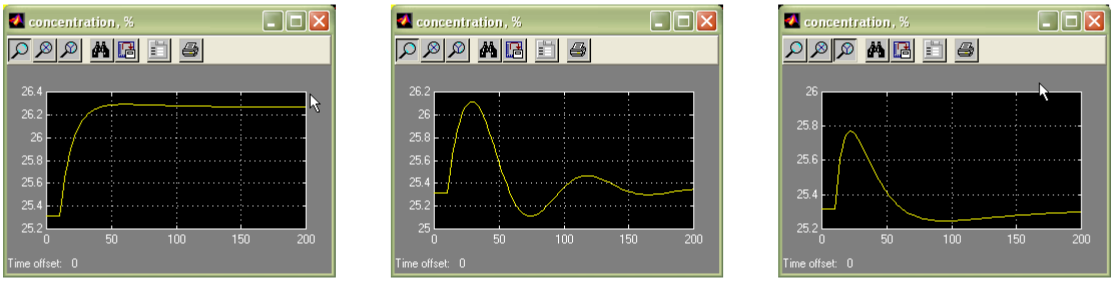


Fig. 8.9 FFC structure for Evaporator process showing the parameter dialog box.



(a) (b) (c)

Fig. 8.10 Process response to step change in the feed concentration