CHAPTER 3
DISTILLATION COLUMN CONTROL

3.1 Introduction

This chapter introduces the subject of a distillation column control. The interactive nature of the process make it one of the most complicated, interesting, but challenging areas for the process control engineer. Many control configurations have been proposed and used to achieve the objective of producing a quality product from a distillation column. This will show how these may be devised from the controlled and manipulated variable of the system.

3.2 Distillation column objectives

The objective of operating a distillation column is to separate the feed so that the overhead product is enriched in the lighter feed components and the bottom product is enriched in the heavier feed components. Either may be the more valuable product so that operations are adjusted to meet criteria for this product while the other is allowed to vary as the operating conditions require. The valuable product criterion is almost always expressed as a function of composition as a specification which states that the product must contain at least a minimum fraction of certain constituents. This specificaton in most cases becomes a maximum as well, since there may be no reward for exceeding it, but making additional product at specification can lead to additional income. Distillation column control systems are then designed to produce a product stream which has a specified composition and neither exceeds nor falls short of this mark.

In addition to maintaining quality specifications, the control system must assure that column operating conditions remain within the limits imposed by the column operating constrains. The foremost of these is to assure that the overall column material balance is satisfied. Local material balances at the condenser and reboiler also must be observed. Their respective liquid levels are limited within minimum levels in the drum. For effective operation the column may not be permitted to flood, slug, weep excessively, drum, etc. In addition the column conditions may need to be adjusted to protect its components.

3.3 Manipulated variables

To achieve the operating objective, column inputs of the control system must be adjusted. If the feed rate, composition, and condition are determined upstream from the column, only the energy input and the relative product rates are available for control operations. There are four manipulable variables in a conventional column (without intermediate product stream):

(i) the distillate rate

(ii) the reflux rate

(iii) the bottom rate

(iv) the reboiler duty

One of these will be chosen as the primary manipulated variable to achieve the column objective in product quality. Two others are required to maintain operating constraints of the liquid level in the reboiler and in the overhead accumulator. The fourth variable is fixed.

3.4 Controlled variables

Although the operating objective is exactly described in terms of composition, this variable is seldom used in control systems. For reasons of reliability and/or cost it is often replaced by a temperature measurement which, with calibration, is used to indicate deviations from the objective composition, three controlled variables are selected for a conventional column:

(i) a selected composition (x_b or x_d)

(ii) accumulator level

(iii) reboiler level

Most control systems for conventional columns consist of three control loops which manipulate input variables to achieve control objectives stated in terms of the three controlled variables. The effectiveness of the control system, however, is determined from its ability to maintain the quality of the more valuable product at its specified value.

3.5 Conventional control configurations for distillation columns

Conventional control configurations for distillation columns are made up of three single-variable control loops and a single manipulated variable which is free. The operator places the set point on the controller which adjusts this variable and changes this set point only seldomly. The column operation to be considered is that one in which the free manipulated variable is not changed at all.

There are twenty-four possible configurations for the three control loops and the free variable. Many of these may be quickly dismissed. A control loop which includes the entire column between the controlled variable and the manipulated variable may be considered poor control since the process transfer function would contain a large time delay and/or a large time constant. This control will be sluggish and undesirable. An example of this sort of loop would eliminate the use of bottoms rate as a manipulated variable to control the level in the overhead accumulator. When these configurations are eliminated, eight possible combinations remain. These are summarized in Table 3.1. Their actions may be understood more readily if their responses, as described in the following sections, are followed on this table.

TABLE 3.1 Control configurations for conventional columns.

Configuration	Overhead accumulator level	Reboiler level	Composition (x_b or x_d)	Free
1	Distillate rate	Bottom rate	Reflux rate	Reboiler duty
2	Distillate rate	Bottom rate	Reboiler duty	Reflux rate
3	Reflux rate	Bottom rate	Reboiler duty	Distillate rate
4	Distillate rate	Reboiler duty	Reflux rate	Bottom rate
5	Distillate rate	Reboiler duty	Bottom rate	Reflux rate
6	Reflux rate	Bottom rate	Distillate rate	Reboiler duty
7	Reflux rate	Reboiler duty	Distllate rate	Bottom rate
8	Reflux rate	Reboiler duty	Bottom rate	Distillate rate

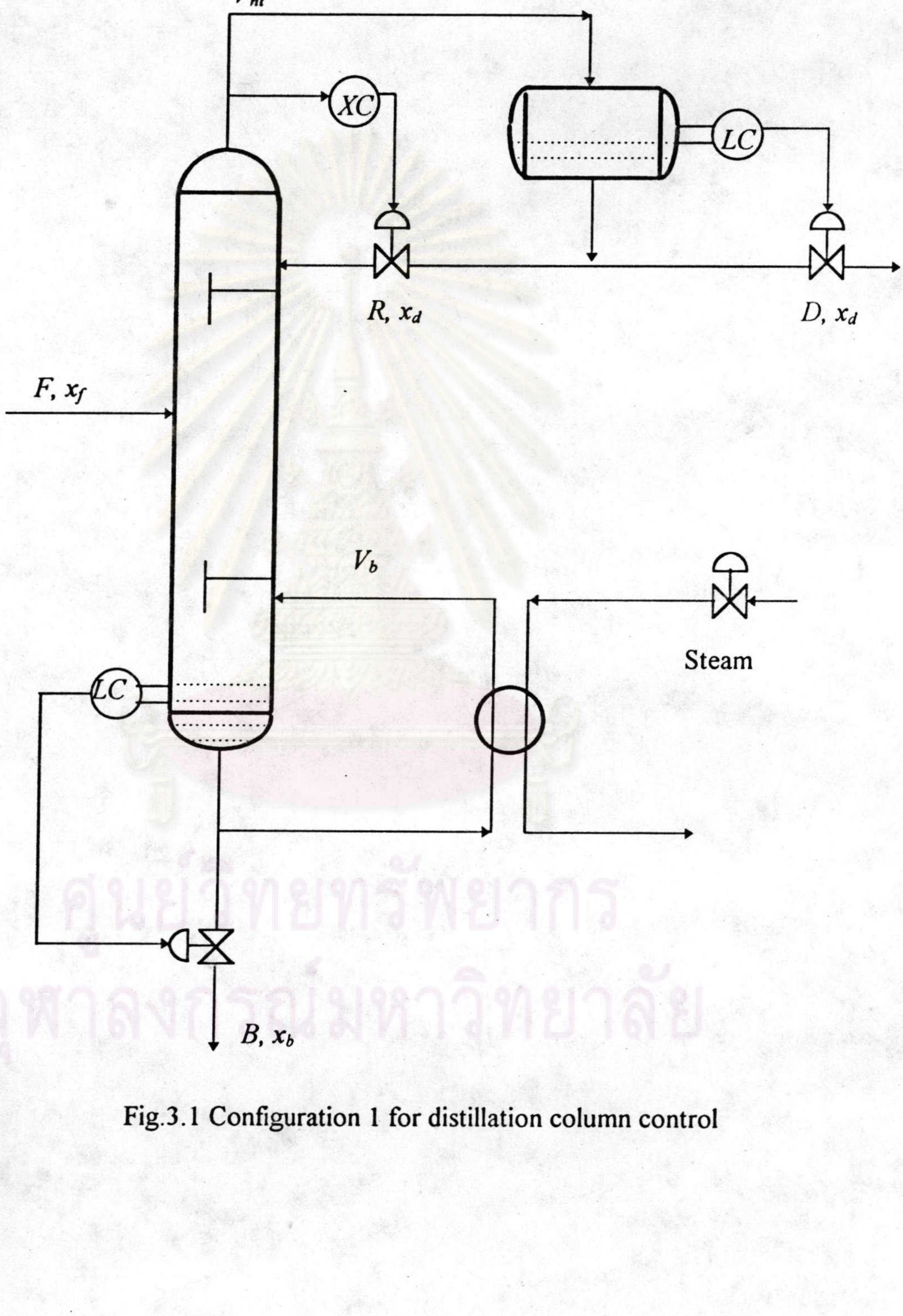

Fig.3.1 Configuration 1 for distillation column control

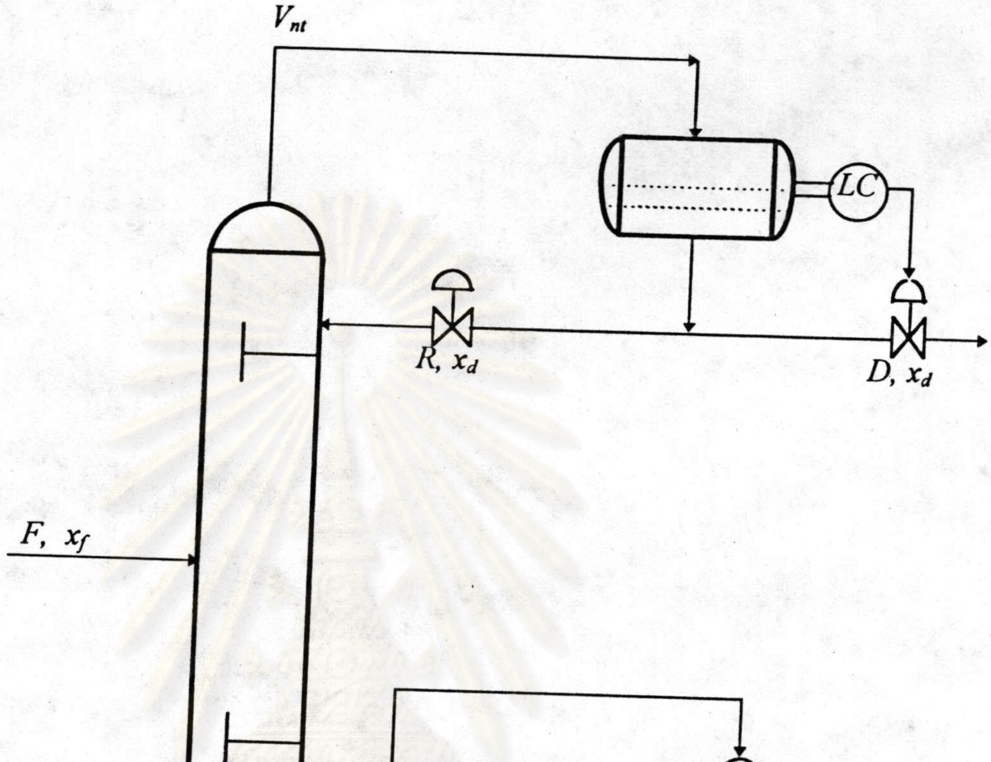

Fig.3.2 Configuration 2 for distillation column control

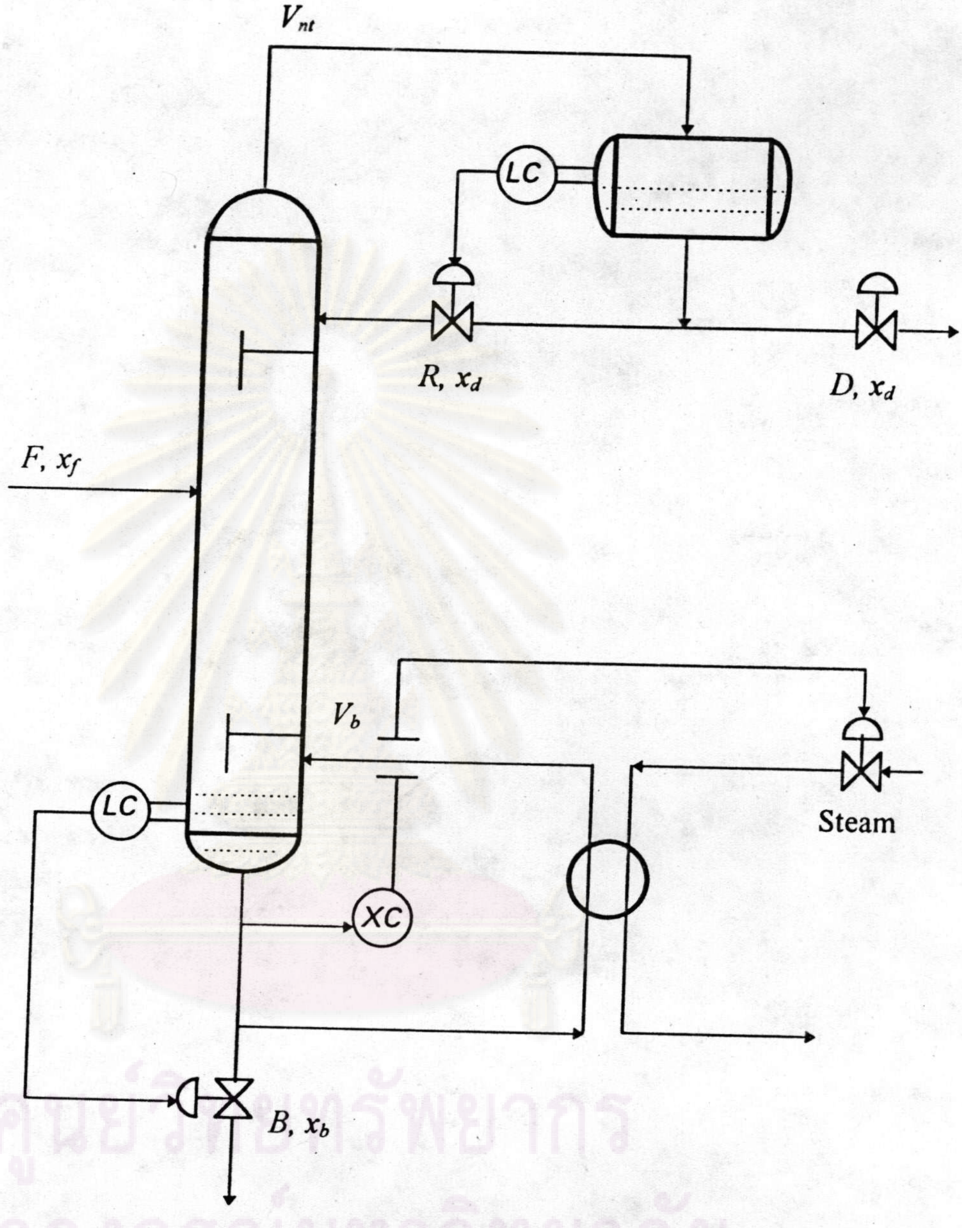

Fig.3.3 Configuration 3 for distillation column control

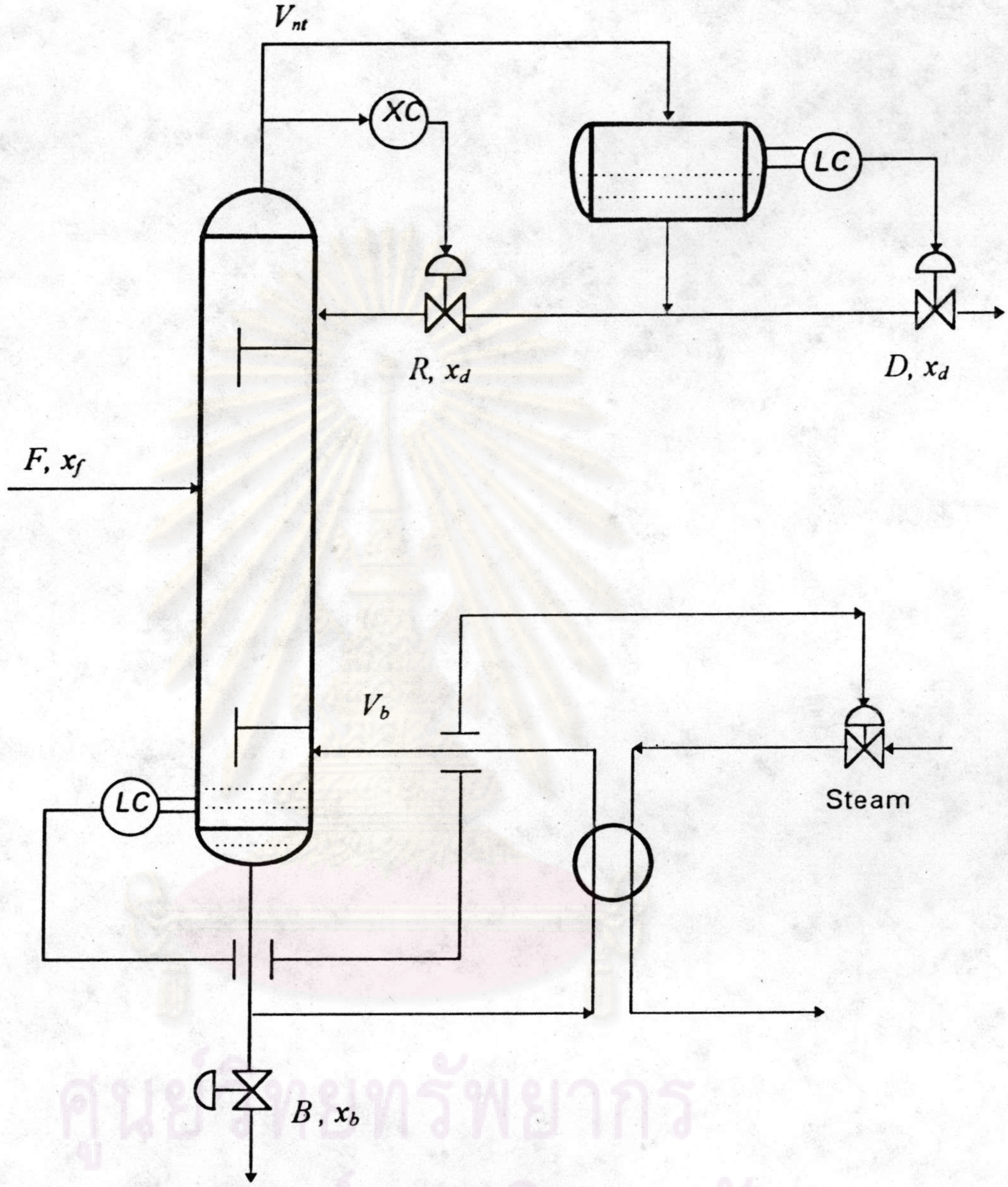

Fig.3.4 Configuration 4 for distillation column control

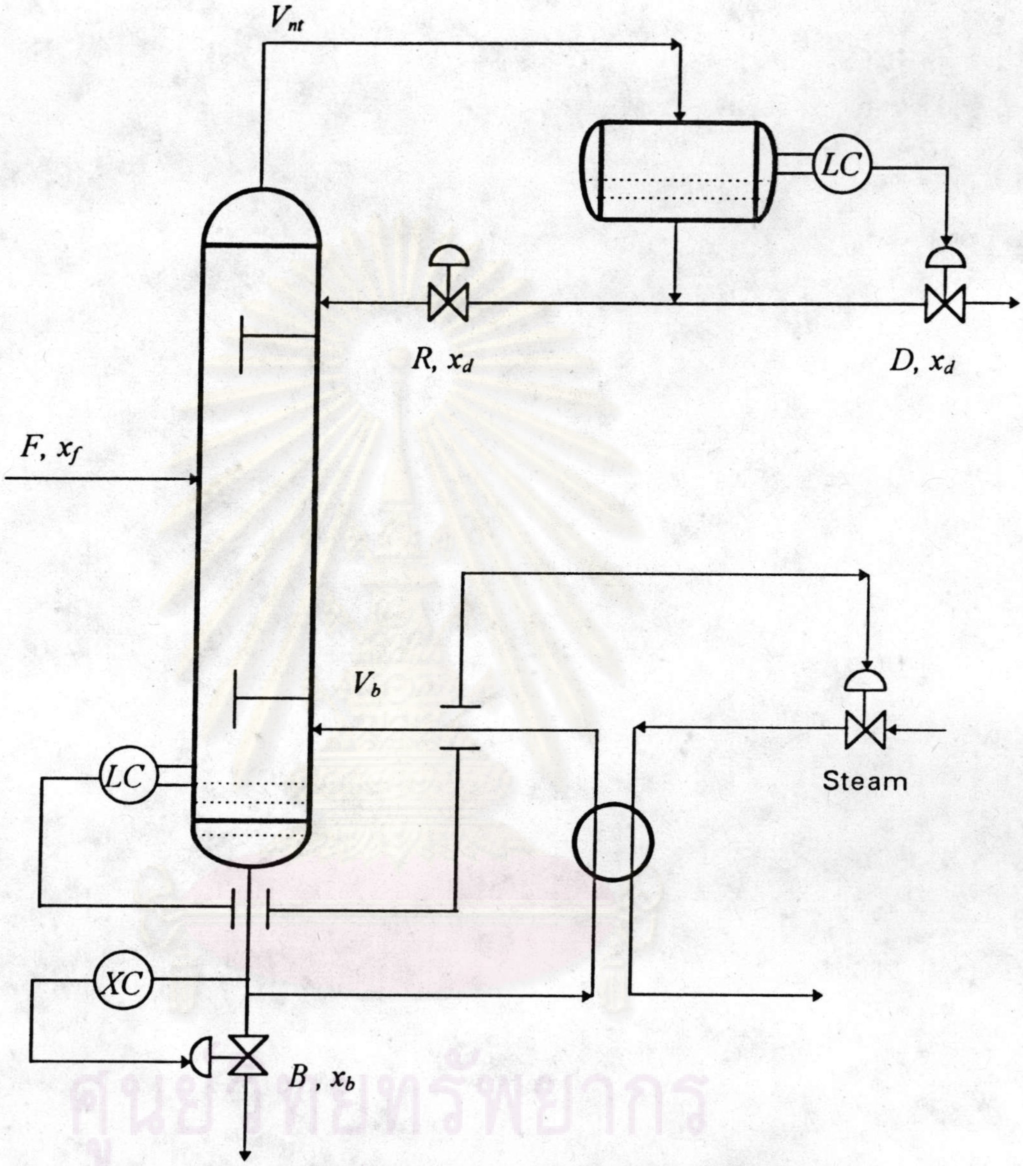

Fig.3.5 Configuration 5 for distillation column control

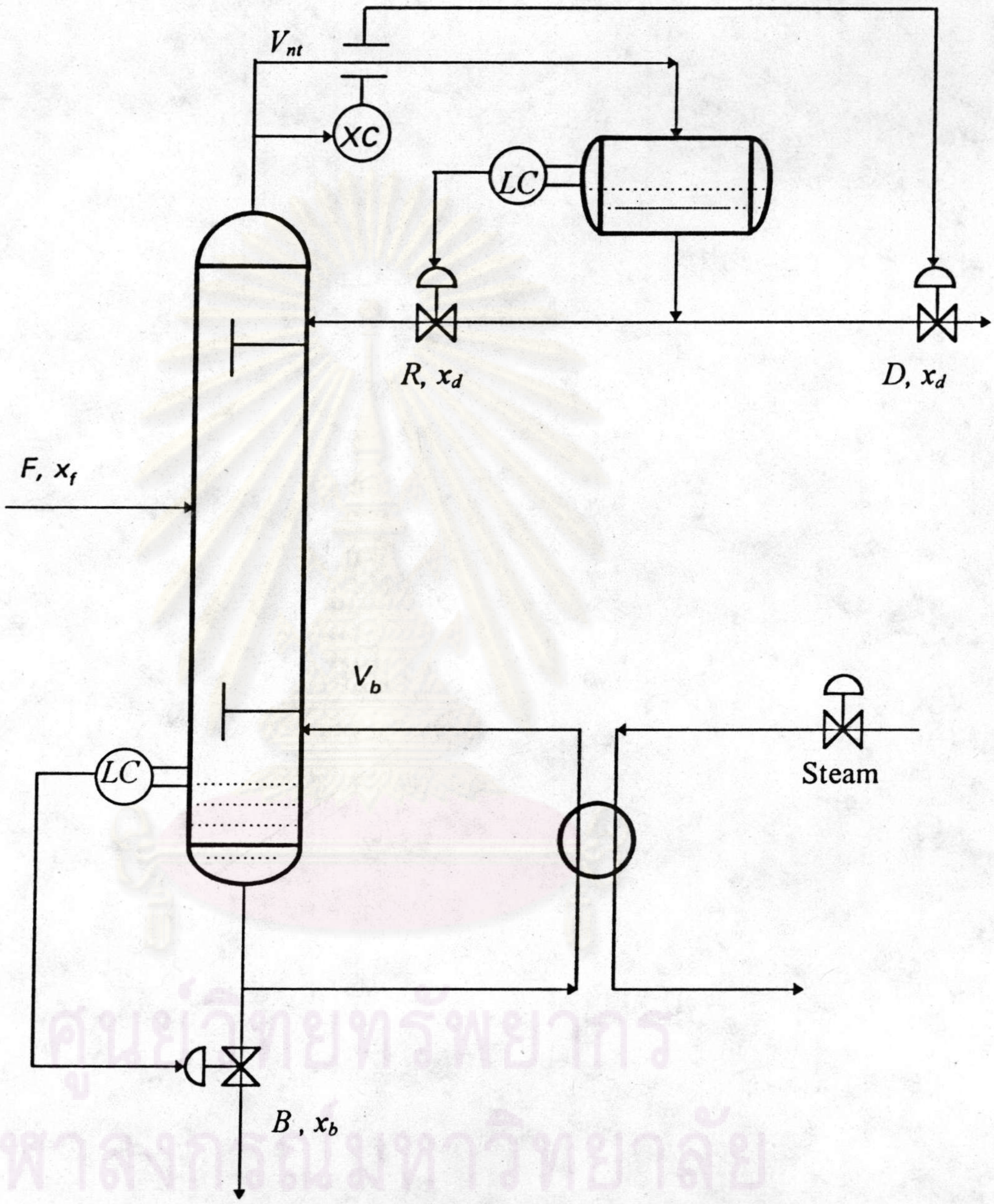

Fig.3.6 Configuration 6 for distillation column control

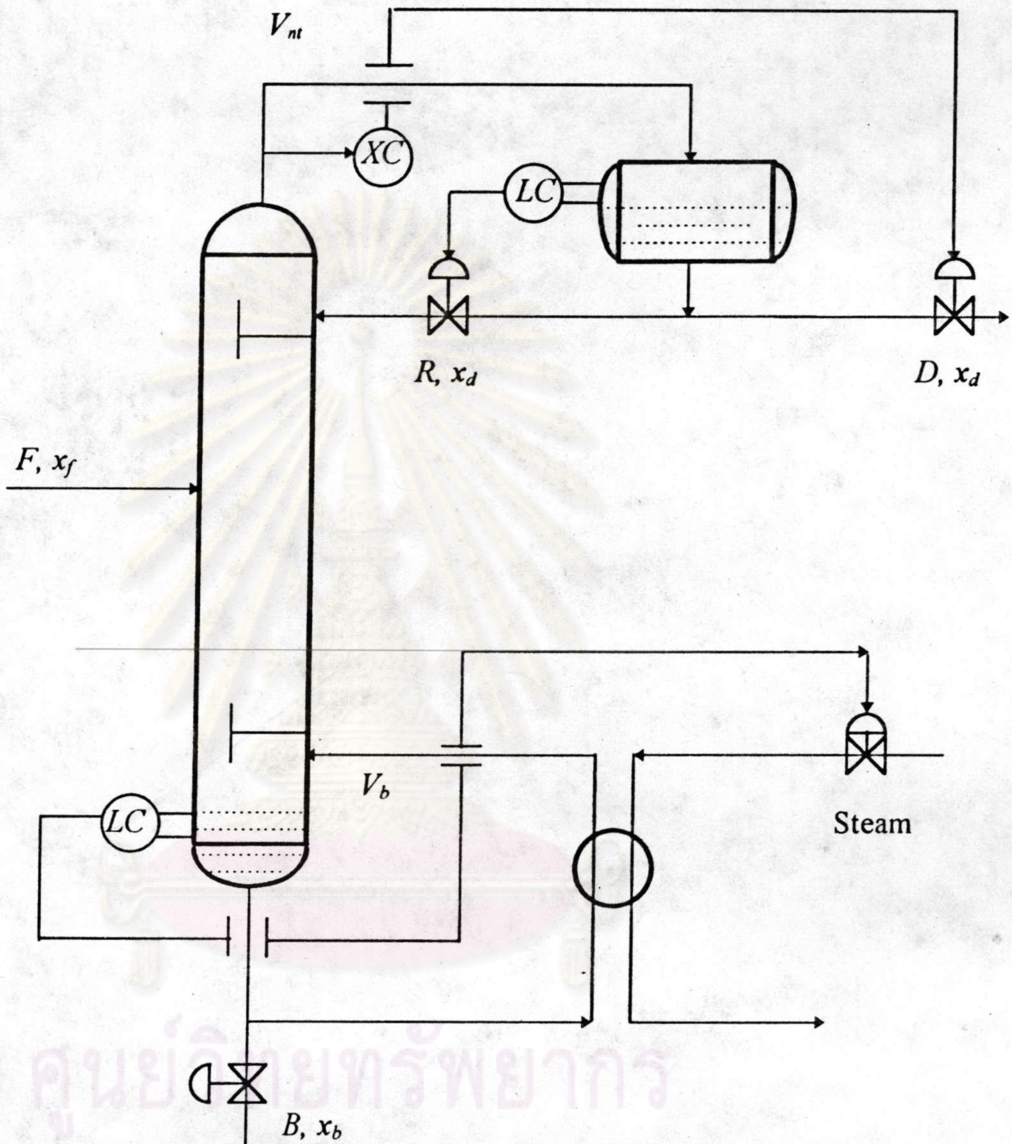

Fig.3.7 Configuration 7 for distillation column control

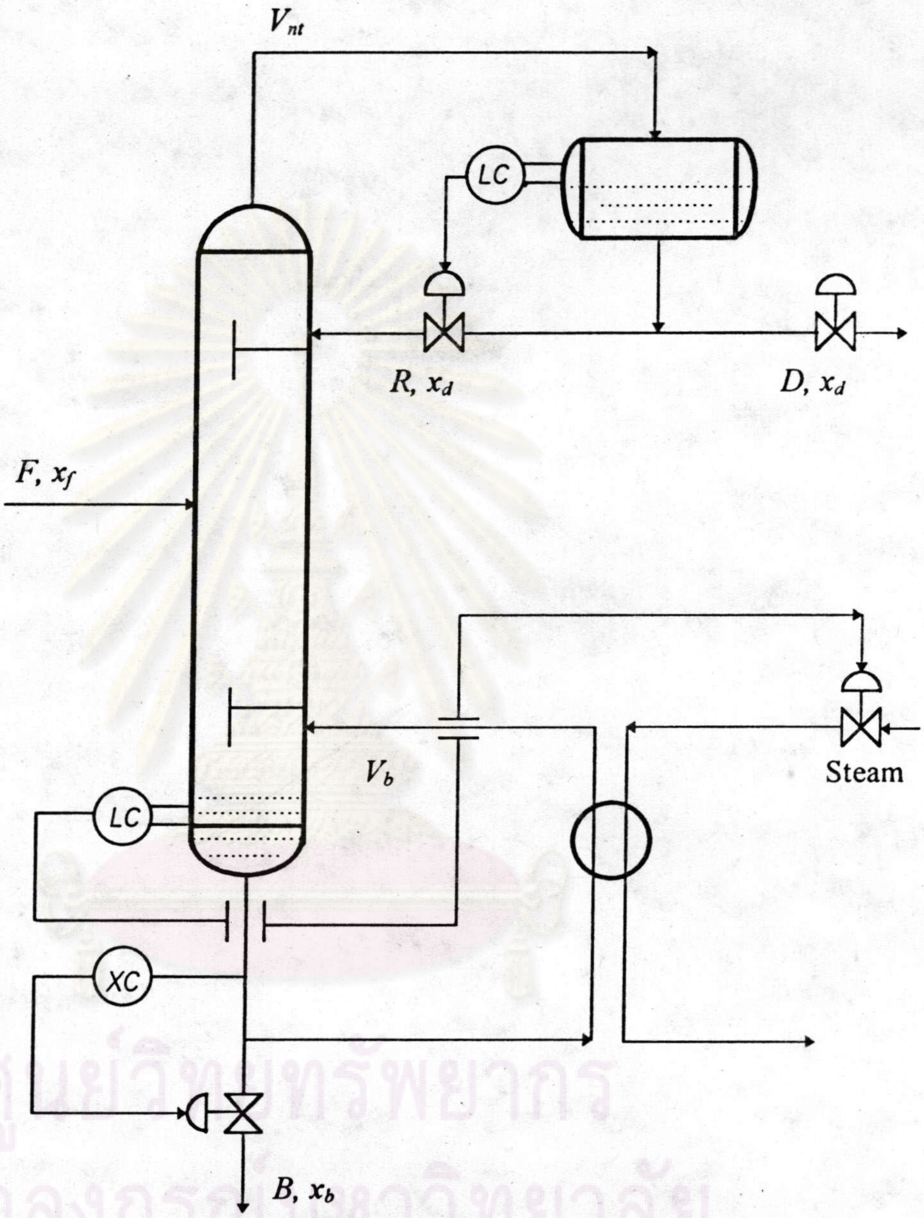

Fig.3.8 Configuration 8 for distillation column control

3.6 Material balance and recycle control

The remaining configurations illustrate three basic procedures for the control of distillation columns:

(i) direct material balance control

(ii) indirect material balance control

(iii) internal recycle control

Configurations 3 and 5 implement direct material balance control. In both cases composition is controlled by manipulating the flow rate of the product stream and thereby changing the distribution of product between the overhead product, the distillate, and the bottom product. The overall material balance for the column is changed to effect a composition change in a product stream. It is intuitively seen that if the distillate rate is reduced, the greater part of the reduction will be obtained from heavier components. The distillate will be come richer in light product, purer, but the components transfered from the distillate appear in the bottoms where they are relatively light and cause the bottoms to be less pure. Furthermore this action has caused a reduction in the quantity of distillate product available for sale.

Configuration 1 is most often the one selected for distillation control and is designated as the "standard" configration. Other configurations are often chosen of course, but must be specifically justified as more suitable than configuration 1 for the particular operation under consideration. Configuration 1 is an example of indirect material balance control, as is configuration 2. In both cases, the initial action taken to control composition manipulates the intenal recycle flow by changing the the reflux rate or the reboiler duty, which changes the boil-up rate. Subsequent actions by the other control loops then affect the overall material balance of the column. For configuration 1, a composition controller may call for an increase in reboiler duty, which is reflected in an increase in the boil-up rate. The reboiler liquid level drops as more material is vaporized and its controller reduces the bottoms rate. The increased vapor rate reaches the condenser/accumulator and its level increases. Since the reflux rate is set, the accumulator level controller opens the valve on the distillate product line. The overall effect is a change in the material balance to increase the distillate rate, decrease the bottoms rate, and improve the bottoms product purity in heavy components. Indirect material balance control may be sensitive to upsets in column with a high reflux (to distillate) ratio.

Configurations 4 and 6 are illustrative of still another type of distillation column control. In both, one product stream is set as the "free" variable. The other product stream is placed on level control in order to maintain the column material balance. Deviations from the specified product quality are corrected by manipulating the internal bulk flow rates within the column. As these rates are increased, a better separation is obtained. This type of control may be described as internal recycle control.

3.7 Feedback controller

The control law necessary to control product compositions and levels of a distillation column is studied. Column inventories and product compositions are controlled by proportional (P) and proportional-integral (PI) controllers. Proportional controllers are used for simple level loops that do not required tight setpoint. Proportional-integral (PI) controllers are used for product composition loops.

3.7.1 Proportional controller (P)

The output of a proportional controller changes only if the error signal changes. Since a load change requires a new control-valve position, the controller must end up with a new error signal. This means that a proportional controller usually gives a steady-state error or offset. This is an inherent limitation of P-controllers and why integral action is usually added. The magnitude of the offset depends on the size of the load disturbance and on the controller gain. The bigger the gain, the smaller the offset. As the gain is made bigger, however, the process becomes underdamped and eventually, at still higher gains, the loop will go unstable, acting like an on-off controller.

Steady-state error is not always undesirable. In many level control loops the absolute level is unimportant as long as the tank does not run down or overflow. Thus a proportional controller is often the best type for level control. A proportional controller changes its output signal, CO, in direct proportion to the error signal, Err, which is the difference between the setpoint, SP, and the process measurement signal, PM, coming from the transmitter. The bias signal, BO, is a constant and is the value of the controller output when there is no error.

$$CO = BO + KC\,(SP - PM) \tag{3.1}$$

3.7.2 Proportional-Integral controller (PI)

Most control loops use PI controllers. The integral action eliminates steady-state error. The smaller the integral time (τ_I), the faster the error is reduced. But the system becomes more underdamped as integral time is reduced. If it is made too small, the loop becomes unstable.

$$CO = BO + KC\,[Err + (\frac{1}{\tau_I})\int (Err)dt + \tau_D(\frac{d(Err)}{dt})] \tag{3.2}$$

where CO = controller output

BO = value bias

SP = setpoint

PM = process variable

KC = controller gain

Err = error between SP - PM

τ_I = controller reset time

3.7.3 Proportional-Integral-Derivative controller (PID)

PID controllers are used in loops where signals are not noisy and where tight dynamic response is important. The derivative action helps to compensate for lags in the loop. Temperature controllers in reactors are usually PID.

$$CO = BO + KC[Err + (\frac{1}{\tau_I})\int (Err)dt + \tau_D(\frac{d(Err)}{dt})] \tag{3.3}$$

where τ_D = derivative time

The common types of control loops are level, flow, temperature, and pressure. The type of controller and the settings used for any one type are sometimes pretty much the same from one application to another. For example, most flow control loops use PI controllers with wide proportional band and fast integral action. Some heuristics are given below. They are not to be taken as gospel. They merely indicate common practice and they work in most applications.

(a) Flow loops

Proportional-integral controllers are used in most flow loops. A wide proportional band setting or low gain is used to reduce the effect of the flow disturbance. A low value of integral or reset time is used to get fast, snappy setpoint tracking. The dynamics of the process are usually very fast.

(b) Level loops

Most liquid levels represent material inventory used as surge capacity. In these cases it is relatively unimportant where the level is, as long as it is between some maximum and minimum levels. Therefore proportional controllers are often used on level loops to give smooth changes in flow rates and to filter out fluctuations in flow rates to downstream units.

One of the most common errors in laying out a control configuration for a plant with multiple units in series is the use of PI level controllers. If P controllers are used, the process flows rise or slowly down the train of units with no overshoot of flow rates. Liquid levels rise if flows increas and fall if flows decrease. Levels are not maintained at setpoint.

If PI level controllers are used, the integral action forces the level back to its setpoint. In fact if level controller is doing a "perfect" job, the level is held right at its setpoint. This means that any change in the flow rate into the surge tank will intermediately change the flow rate out of the tank. It might as well not even use a tank: just run the inlet pipe right into the outlet pipe! Thus, this is an example of where tight control is not desirable. It want the flow rate out of the tank to increase gradually when the inflow increases so as to not upset the downstream units.

Suppose the flow rate F_0 increases to the first tank. The level h in the first tank will stert to increase. The level controller will start to increase F_1, When F_1 has increased to the point that it is equal to F_0, the level will stop changing since the tank is just an integrator. Now if P controller is used, nothing else will happen. The level will remain at the higher level and the entering and exiting flows will be equal.

If, however, PI level controller is used, the controller will continue to increase the outflow beyond the value of the inflow in order to drive the level back down to its setpoint. So an inherent problem with PI level controllers is that they amplify flow rate changes of this type. The change in the flow rate out of the tank is actually larger (for a period of time) than the change in the flow rate into the tank. This amplification gets worse as it works its way down through the series of units.

Example 1 P-controller

Assuming that the column base level (H_1) is controlled by manipulating bottom rate (B). The model of the column base level is:

$$A\frac{dH_1}{dt} = L_1 - V_b - B \tag{3.4}$$

Integrating by using the Euler's method to estimate error (H_1 - H_0)

$$dH_1 = \frac{(L_1 - V_b - B)\Delta t}{A} = (H_1 - H_0) \tag{3.5}$$

Apply P controller by adjusting bottom rate:

$$B = BO + KC\,(H_1 - H_0) \tag{3.6}$$

Example 2 PI-controller

Assuming that the product composition (x_1) at the top column is controlled by manipulating reflux rate (R). If the dynamic model is:

$$M_{nt}\frac{dx_{nt}}{dt} = x_d R + y_{nt\text{-}1}V_{nt\text{-}1} - x_{nt}L_{nt} - y_{nt}V_{nt} \tag{3.7}$$

Integrating by using the Euler's method to estimate error (H_1 - H_0)

$$dx_{nt} = \frac{(x_d R + y_{nt-1}V_{nt-1} - x_{nt}L_{nt}' - y_{nt}V_{nt})\Delta t}{M_{nt}} \tag{3.8}$$

Apply PI controller by adjusting distillate rate:

$$R = RO + KC\,[(x_1 - x_0) + (\frac{1}{\tau_I})\int (x_1 - x_0)\,dt\,] \tag{3.9}$$

where x_1 $= x_{nt}$ at any time (t)

x_0 $= x_{nt}$ at the old time

RO = initial guess of reflux rate