

MODULE II

CONTROL OF COMMON UNIT OPERATIONS

Having covered the essential aspects of control theory, in this module we consider control systems as applied to common unit operations in the process industry. We thus treat simple and complex distillation configurations (including heat integrated sequences), reactors, heat exchangers and miscellaneous systems such as furnaces, compressors, refrigeration cycles and boiler houses. Several of the examples shown here can be found in “Plantwide Process Control” by Luyben, Tyreus and Luyben (McGraw Hill, 1998) and “Process Control Systems” by Shinskey (McGraw Hill, 1996). We have attempted to present these examples afresh in the hope that readers readily assimilate the concepts.

Chapter 5. Control of Distillation Systems

5.1. Distillation Basics

Distillation is unarguably the most preferred unit operation used for separating mixtures. In the design of chemical processes, other separation techniques are considered only if distillation is found to be economically unviable. It is thus not surprising that the final product stream from a plant is typically a product stream from a distillation column. This Chapter provides guidelines for designing effective control systems for distillation columns.

5.1.1. *The Simple Distillation Column*

A proper understanding of the basic physics of a distillation column (or any other process for that matter) is a pre-requisite for designing an effective control system. Figure 5.1 shows the schematic of a simple distillation column along with the control valves. It consists of a tray section, a condenser, a reflux drum and a reboiler. The feed mixture is fed on a feed tray. The trays above the feed tray constitute the rectifying / enriching section and those below constitute the stripping section. The overhead distillate and the bottoms are the two product streams from a simple distillation column. Steam is typically used to provide vapour reboil into the stripping section. The liquid reflux into the enriching section is provided by the condenser. Cooling water is commonly used as the coolant in the condenser. The condenser may be a total condenser, where all the vapour is condensed, or a partial condenser where only a part of the vapour is condensed. The overhead distillate is a liquid stream for a total condenser. A partial condenser column may be operated at total reflux where all the liquid is refluxed back into the column and the distillate stream is a vapour stream. Alternatively (and more commonly) both a vapour and a liquid distillate stream are drawn. The reflux drum provides surge capacity to adjust the reflux and distillate rate during transients. The bottom sump provides the surge capacity for adjusting the bottoms and steam rate.

The vapour generated when a volatile liquid feed mixture is boiled is richer in the more volatile component. The remaining liquid is then richer in the heavier components. Chemical engineers refer to this as flashing a mixture. If the flashed vapour is condensed and partially vaporized again, the vapour from the second flash would be further enriched in the volatiles (light boilers). Similarly, if the liquid from the first flash is further vaporized, the heavies composition of the liquid from the second flash would increase. Theoretically speaking, a sufficiently large number of flash operations on the vapour can result in a final vapour stream that is almost 100% pure lightest component. Similarly a series of flash operations on the liquid can result in a final liquid product that is 100% heaviest component. The array of trays in a distillation column accomplishes this series of flash operations. The temperature difference between the liquid and vapour streams entering a tray causes condensation / vaporization so that as one moves up the column, the composition of the lightest component increases monotonically. Alternatively, as one moves down the column, the composition of the heaviest component keeps on increasing. Since heavier components boil at higher temperatures, the tray temperature increases as one moves down the column with the condenser being the coolest and the reboiler being the hottest. The reboiler and the condenser are the source of vaporization and condensation respectively for the series of vaporization / condensation.

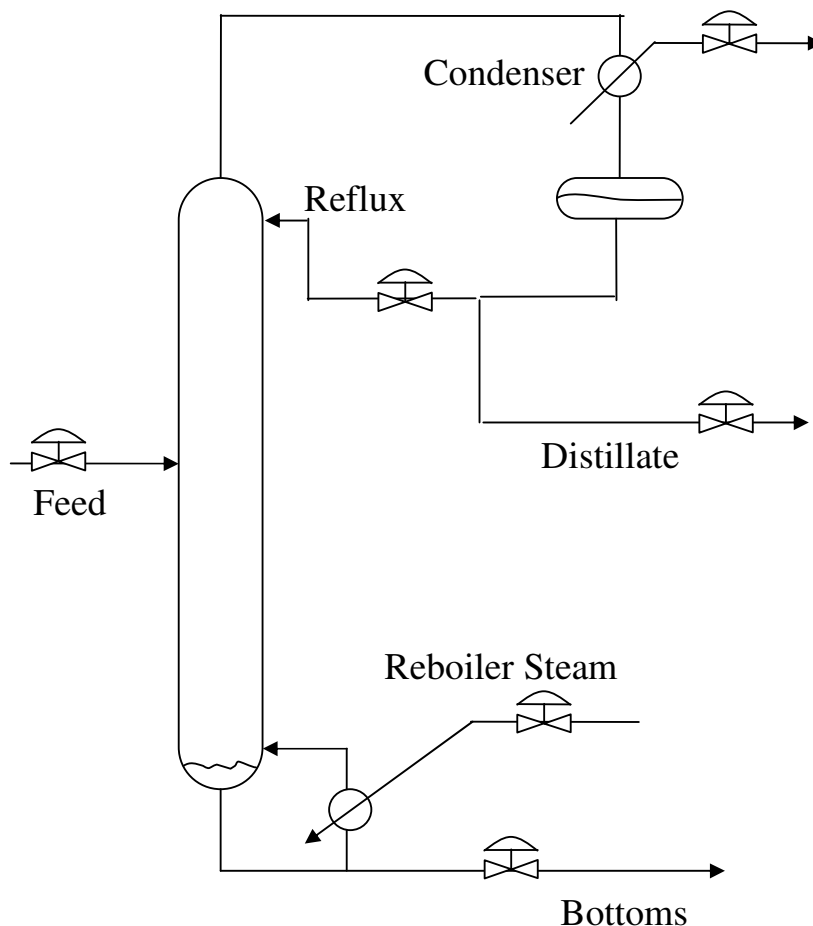


Figure 5.1 Schematic of a simple distillation column along with the control valves.

5.1.2. Splits in a Simple Distillation Column

Consider a five component equimolar ABCDE mixture feed into a simple distillation column. The components are in decreasing order of volatility so that A is the lightest and E is the heaviest. The feed rate is 100 kmol/h. The steady state distillate to bottoms product split is primarily determined by the choice of the distillate (or bottoms) rate. Assuming a sufficiently large number of trays, adequate reboil and reflux, for a distillate rate of 40 kmol / hr, which is equal to the component A and component B flow rate in the feed, essentially all of the A and B would leave up the top so that the distillate would contain traces C, D and E impurities in decreasing order of composition. The bottoms would be a CDE mixture with traces of B and A, in decreasing order of composition. The column thus accomplishes a split between components B and C with the liquid preventing C from escaping up the top and the vapour reboil preventing B from escaping down the bottoms. Components B and C, are referred to as the light key (LK) and heavy key (HK) respectively. The LK is the dominant impurity in the bottoms stream and the HK is the dominant impurity in the distillate stream. The component split is referred to as an AB/CDE split. The component that is the next lighter component than the LK is called the lighter than light key (LLK). The heavier than heavy key (HHK) can

be defined in a complementary manner. Components A and E are respectively the lightest and heaviest and therefore referred to as the lightest key and the heaviest key.

For the ABCDE mixture, there are four possible splits – A/BCDE, AB/CDE, ABC/DE and ABCD/E. The first one, where the light key is also the lightest key is referred to as the direct split. The last one, where the heavy key is also the heaviest key is referred to as an indirect split. The remaining splits where the key components are intermediate boilers are referred to as intermediate splits. It is helpful to categorize the column split into these basic types.

5.2. Basic Control Structures

A simple distillation column with a total condenser has a total of six valves as in Figure 5.1. Of these six valves, the feed valve is usually set by an upstream unit in the process. Also two valves must be used to control the reflux drum level and the reboiler level as liquid levels are non-self regulating. Another valve must be used to regulate the column pressure which represents the vapour inventory in the column. Typically, the cooling duty valve in the condenser is used for pressure control. After implementing the three inventory loops, the position of the remaining two control valves can be set by an operator or a controller to regulate the separation. This gives a operation degree of freedom of two for a simple distillation column. The operation degree of freedom is more for complex column configurations that are considered later.

Four control structure types result for a distillation column corresponding to the choice of valve used for reflux drum and reboiler level control. These are the LQ, DQ, LB and DB structures and are illustrated in Figure 5.2. The nomenclature corresponds to the two control degrees of freedom (valves) that remain to regulate the separation. The LQ control structure corresponds to the distillate (D) controlling the reflux drum level and the bottoms (B) controlling the reboiler level. This leaves the reflux (L) and reboiler duty (Q) as the two valves for regulating the separation achieved, hence the label LQ. In the DQ structure, the condenser level is controlled using the reflux while in the LB structure, the bottoms level is controlled using the reboiler duty. Lastly in the DB control structure, the reboiler duty and reflux are used for controlling the reboiler and condenser levels respectively.

5.2.1. The Energy Balance (LQ) Structure

The LQ control structure is the most natural control structure for a simple distillation column. This is because the separation in a distillation column occurs due to successive condensation and vaporization of the counter-current vapour and liquid streams flowing through the column. Adjusting the cold reflux, the source of condensation, and the reboiler duty, the source of vaporization, is then a natural choice for regulating the separation achieved in the column. The LQ control structure is thus the most commonly applied distillation control structure. It is also sometimes referred to as an energy balance structure as changing L (cold reflux) or Q alters the energy balance across the column to affect the distillate to bottoms product split.

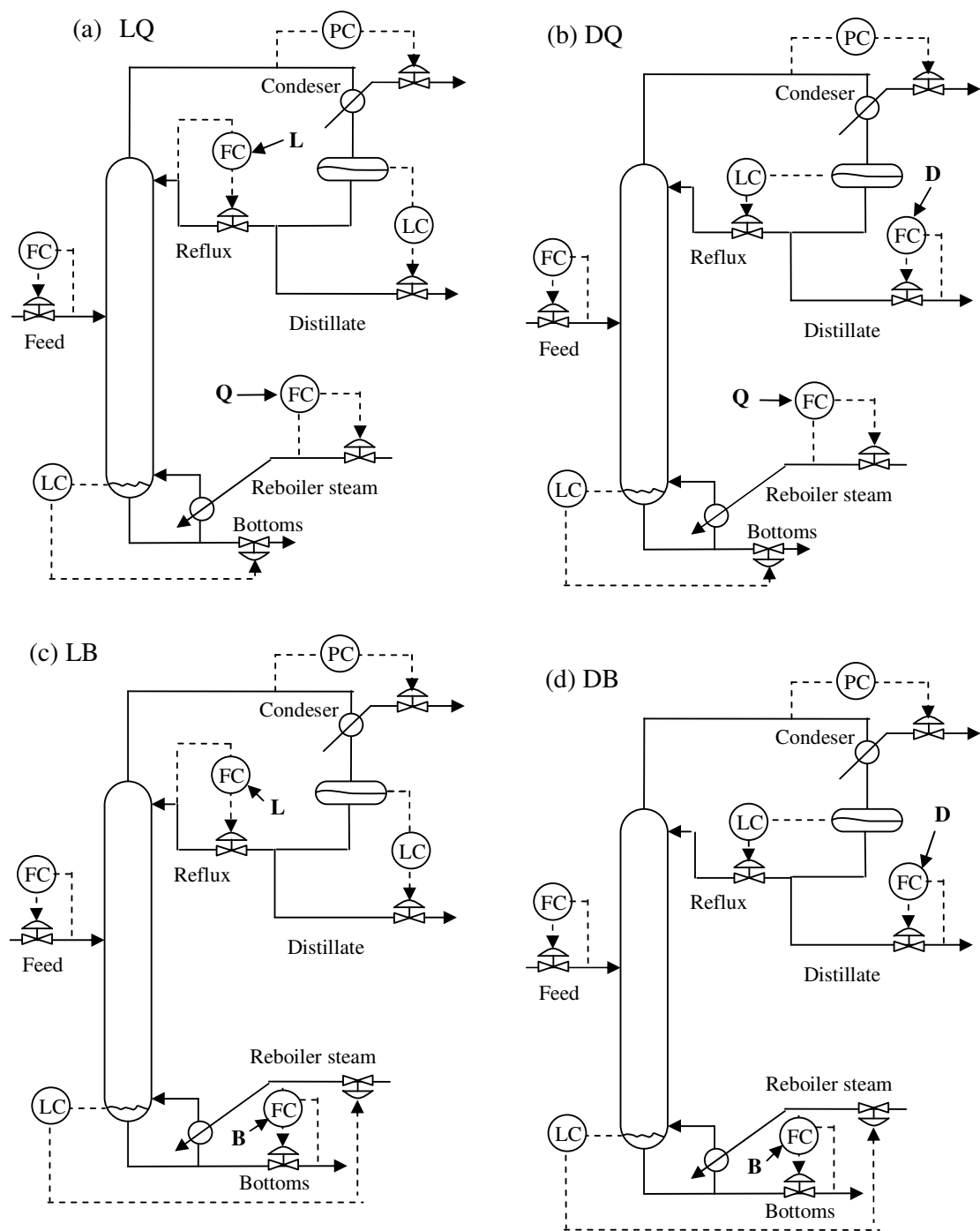


Figure 5.2. Schematics of LQ, DQ, LB and DB control structures

5.2.2. Material Balance Structures

The other control structures are referred to as material balance structures as the product split is directly adjusted by changing the distillate or bottoms stream flow rate. The material balance structures are applied when a level loop for the LQ structure would be ineffective due to a very small product stream (D or B) flow rate. The DQ structure is thus appropriate for columns with very large reflux ratio ($L/D > 4$). The distillate stream flow is then a fraction of the reflux stream so that the reflux drum level cannot be maintained using the distillate. The level must then be controlled using the reflux. The LB structure is appropriate for columns with a small bottoms flow rate compared to the boil-up. The bottoms stream is then not appropriate for level control and the reboiler duty must be used instead. The DB control structure is used very rarely as both D and B cannot be set independently due to the steady state overall material balance constraint. In dynamics however, the control structure may be used when the reflux and reboil are much larger than the distillate and bottoms respectively. This occurs in super fractionators which will be discussed later in this Chapter.

5.2.3. Other Control Structure Variants

Other variants of the basic control structure types include the L/D-Q, L/D-B and D-Q/B. In the first two structures the reflux ratio is adjusted for regulating the separation. In the last structure the reboil ratio is adjusted. These control structures are illustrated in Figure 5.3. Note that when the reflux is adjusted in ratio with the distillate, the distillate stream can be used to control the reflux drum level even as it may be a trickle compared to the reflux rate. Similarly, when the reboil rate is adjusted in ratio with the bottoms, a small bottoms stream can provide effective level control.

Maintaining the reflux ratio is quite common in distillation control as it provides feedforward compensation in the reflux for a change in the distillate rate. Such a feedforward compensation can significantly improve quality control as the column dynamics are slow with respect to a change in the reflux rate due to the slow liquid hydraulics with every tray having a time constant of 15-30 s. Pure feedback adjustment of the reflux can thus result in large purity deviations. Maintaining the reboil ratio is not very popular. This is because all the tray compositions / tray temperatures respond almost immediately to a change in the reboil due to the fast vapour dynamics. Adjustment of the reboiler duty in a feedback arrangement is thus usually sufficient for effective regulation.

5.3. Temperature Based Inferential Control

The distillation column performs a separation between the light key and the heavy key so that heavy key and light key impurity levels respectively in the distillate and bottoms are below design specifications. The primary control objective then is to ensure these impurity levels remain below specifications for load changes. A change in the flow rate and composition of the feed into the column are the two major load disturbances that must be rejected by the control system. Feedback control based on the impurity levels in the product streams is usually not acceptable due to the large delays / lags introduced by composition measurements. Also, control action would only be taken after the product purity deviates in a feedback system. Considering that the column consists of an array of trays, the trays would respond to a load disturbance much before the effect of the disturbance reaches the product streams. It therefore makes sense to control an appropriate tray process variable so that the

disturbance is compensated for before the product purities are affected. This would lead to tighter product purity control.

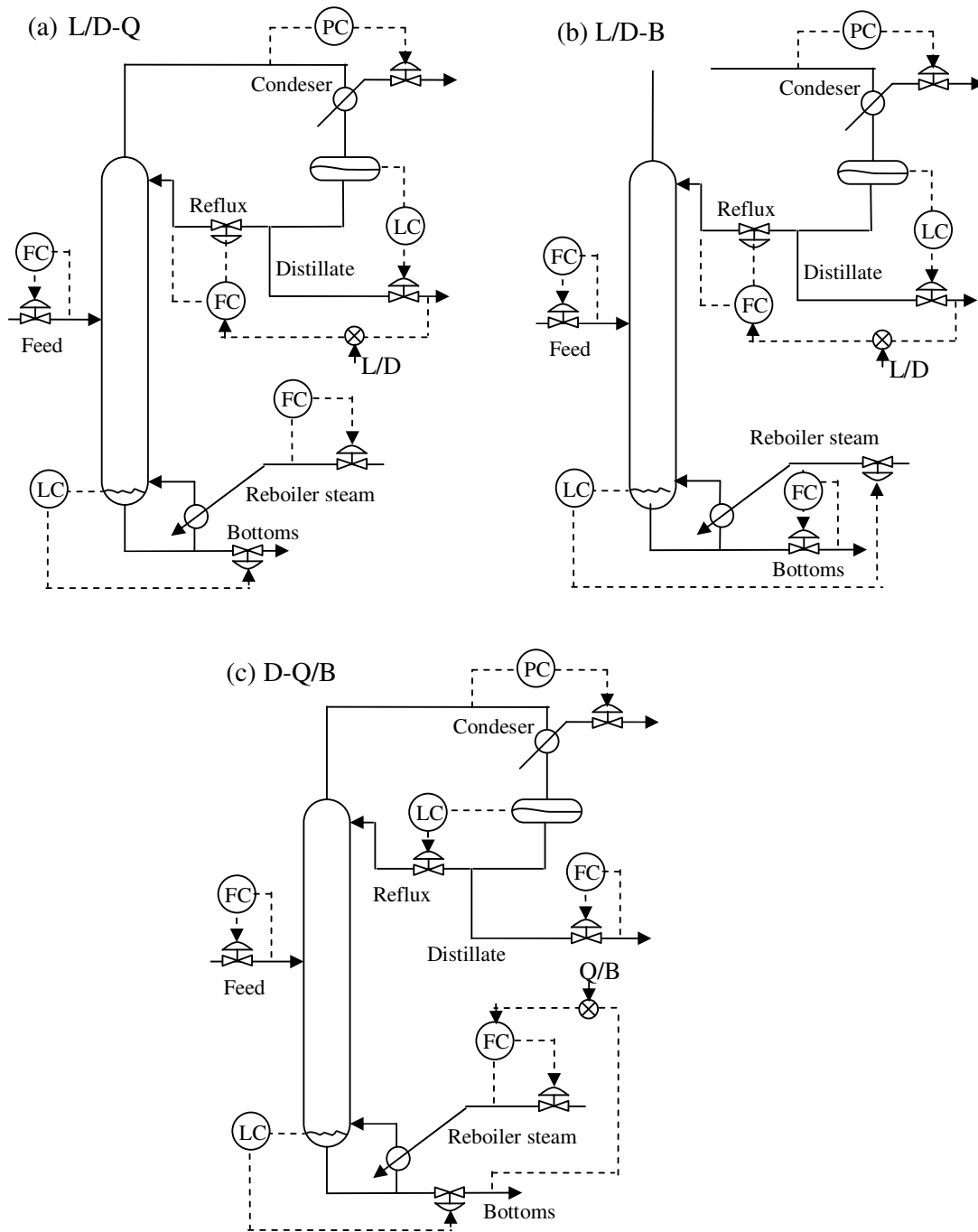


Figure 5.3. Schematics of L/D-Q, L/D-B, and D-Q/B control structures.

The tray temperature is almost always used as an inferential variable for the tray composition. The boiling point of a mixture depends on the pressure and the mixture composition. At a constant pressure, the boiling point increases as the concentration of

heavies increases and vice versa. A change in the tray composition can thus be inferred from a change in the tray temperature. The relationship is exact for a binary mixture and is approximate for a multi-component mixture.

5.3.1. Single-Ended Temperature Control

Controlling a single tray temperature in a column is usually referred to as single-ended temperature control. Either of the two operation degrees of freedom can be used as the manipulation handle. For example, in the LQ control structure, the reflux rate or the reboiler duty can be manipulated for maintaining a tray temperature. This is shown in Figure 5.4(a). Manipulation of Q is usually preferred due to the fast response of all the tray temperatures to a change in Q . The dynamics with respect to reflux rate are slower due to the associated tray liquid hydraulic lags. The reflux rate can also be used for maintaining a tray temperature if the control tray location is not too far below the reflux (say about 10 trays). The single ended variants for the DQ and LB control structures are shown in Figure 5.4(b) and (c) respectively. In the DQ structure, if the distillate rate is used to control a tray temperature, the temperature controller is nested with the reflux drum level controller. This means that the temperature controller would work only if the reflux drum level controller is working. Similarly, in the LB structure, if the bottoms flow rate controls a tray temperature, the temperature controller is nested with the reboiler level controller. In both these cases, the level controller must be a tightly tuned PI controller, else the temperature control would be extremely sluggish. Note that the reflux and the reboil are the only two causal variables that affect the tray temperature so that any control scheme must directly / indirectly effect a change in these causal variables.

5.3.2. Dual-Ended Temperature Control

Theoretically speaking, since the column degree of freedom is two, two tray temperatures can be controlled in a column. This is referred to as dual-ended temperature control. For example, in the LQ control structure, the reflux rate can be used for controlling a rectifying tray temperature and the reboiler duty can be used to control a stripping tray temperature as in Figure 5.5. Industrial practice is to control a single tray temperature as controlling two tray temperatures usually requires detuning of the temperature controllers due to interaction between the temperature loops. More importantly, the two controlled tray temperatures may not be sufficiently independent so that, in the worst case, the control system may seek infeasible temperature set-points. Dual temperature control is possible for long towers so that two tray temperatures that are far apart are sufficiently independent.

5.4. Temperature Sensor Location Selection

Various criteria have evolved for the selection of the most appropriate tray location(s) for temperature control. Prominent among these are selection of tray with the maximum slope in the temperature profile, sensitivity analysis and SVD analysis.

5.4.1. Maximum Slope Criterion

The maximum slope criterion is the simplest to use and requires only the steady state temperature profile. From the temperature profile, the tray location where the temperature changes the most from one tray to the other is chosen as the control tray. The temperature profile usually also shows a large change / break at the feed tray location. The feed tray should however not be chosen for control as the changes in temperature would be due to

changes in the feed composition / temperature and not due to a change in the separation. A large change in the temperature from one tray to the other reflects large separation between the key components so that disturbances in the separation would affect this separation zone much more than other locations. It therefore makes sense to place the temperature sensor at that location.

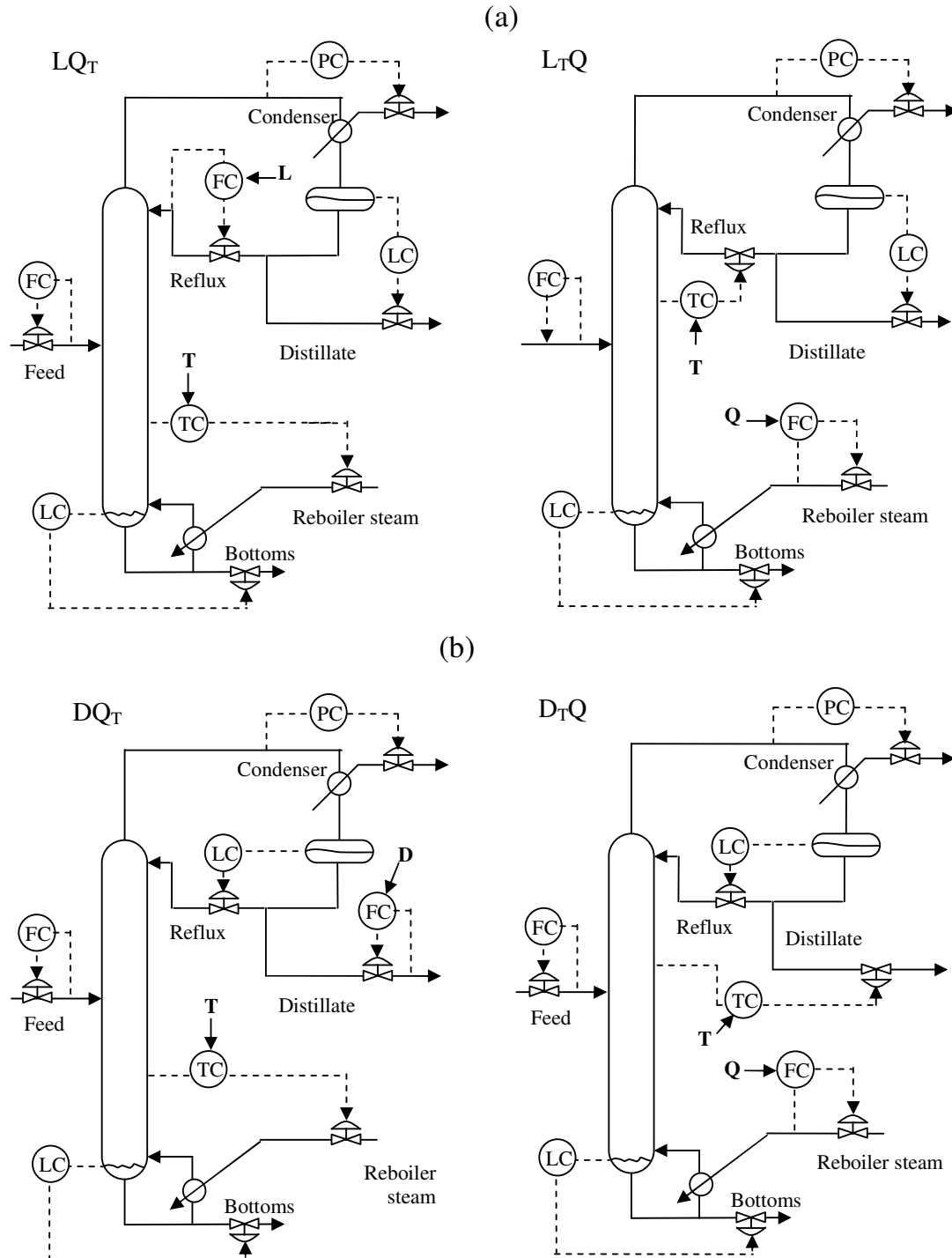
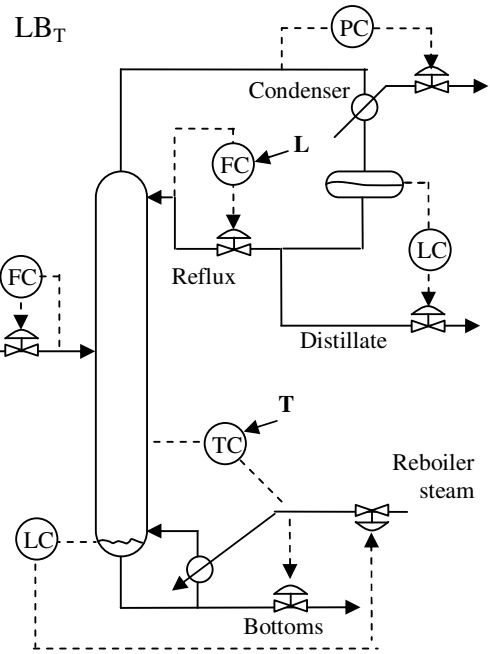
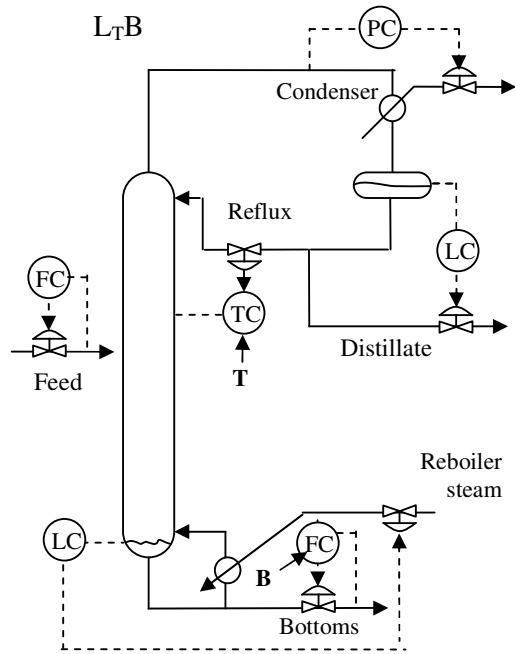


Figure 5.4 Single ended temperature control structures using LQ and DQ scheme

(c)



(d)

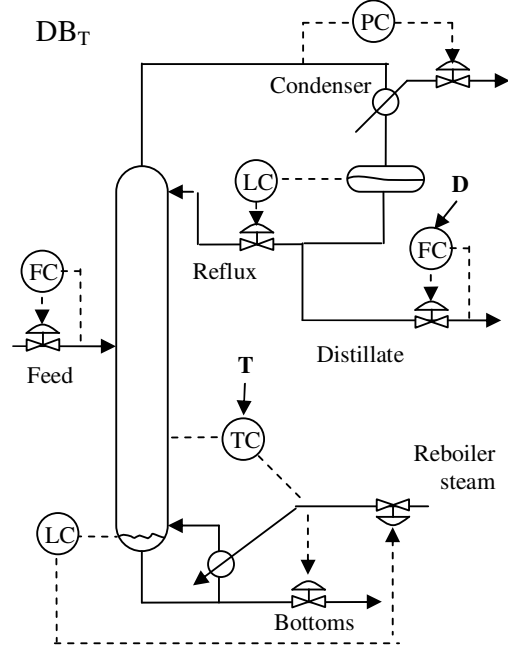
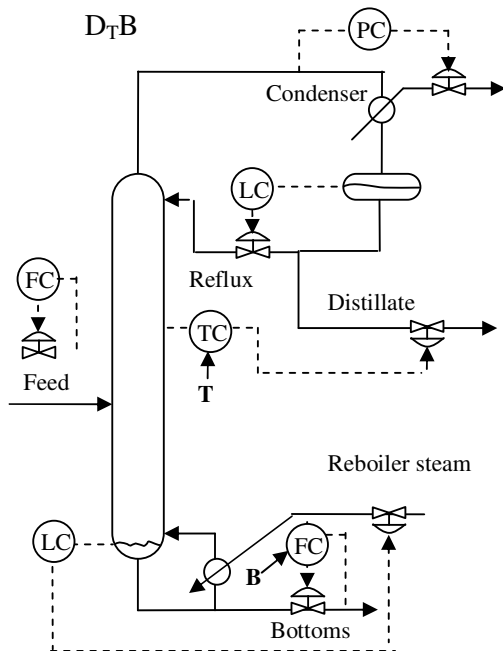


Figure 5.4 Single ended temperature control structures using LB and DB scheme

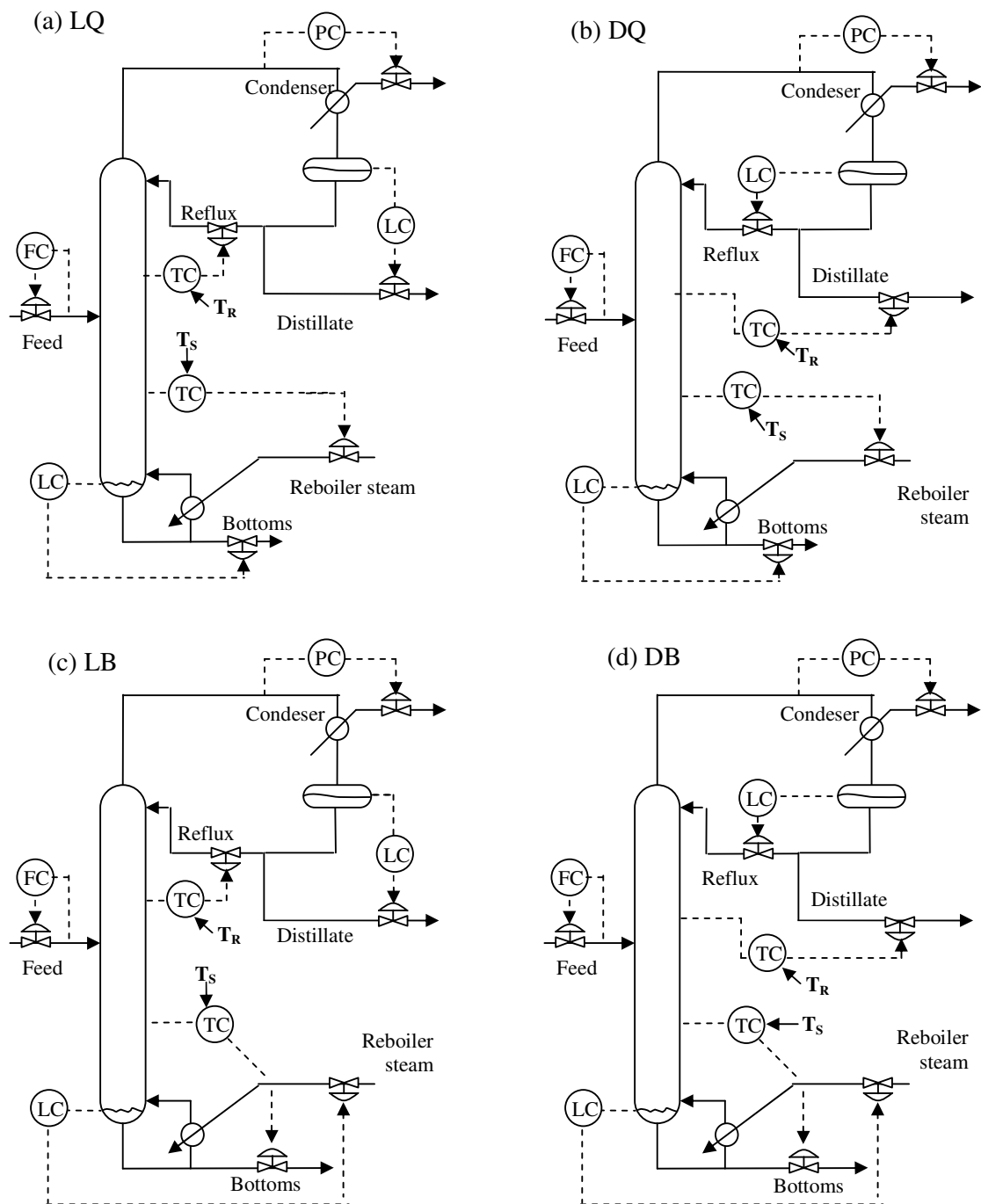


Figure 5.5. Dual ended temperature control structures using LQ, DQ, LB and DB schemes.

5.4.2. Maximum Sensitivity Criterion

Sensitivity analysis recommends controlling the tray with maximum sensitivity to the control input. The causal variables that effect a change in the tray temperature are the reflux rate (or ratio) and the reboiler duty. The sensitivity of the i^{th} tray temperature to the reflux rate (L) and reboiler duty (Q) is defined as

$$S_{iL} = \frac{\partial T_i}{\partial L}$$

and

$$S_{iQ} = \frac{\partial T_i}{\partial Q}$$

respectively. Controlling the most sensitive tray location provides muscle to the controller as a smaller change in the manipulated variable is needed to bring the deviating temperature back to its set-point. The open loop steady state gain is large so that a low controller gain suffices. The low controller gain mitigates sensor noise amplification. Also, a small bias in the temperature sensor can be tolerated. Plotting the sensitivity of all the trays with respect to Q and L would reveal the most sensitive tray location. In case two distinct regions of high sensitivity are observed, dual temperature control should be possible. If not, dual temperature control is likely to result in the two temperature controllers fighting each other.

5.4.3. SVD Criterion

The SVD analysis is another useful technique for selecting the tray temperature locations. The sensitivity matrix

$$\mathbf{S} = [\mathbf{S}_L \mathbf{S}_Q]$$

where \mathbf{S}_L and \mathbf{S}_Q are column vectors of tray sensitivities, is decomposed using the singular value decomposition (SVD) as

$$\mathbf{S} = \mathbf{U}\mathbf{\Sigma}\mathbf{V}^T.$$

In the above \mathbf{U} and \mathbf{V} are orthogonal matrices with the columns constituting the left singular and right singular vectors, respectively. The $\mathbf{\Sigma}$ matrix is a diagonal matrix. A plot of the first two left singular vectors (first two columns of \mathbf{U}) shows the two most independent locations in the column. The index of the element of the first left singular with the maximum magnitude corresponds to the tray location that should be controlled in a single-ended scheme. If dual temperature control is to be implemented, the corresponding index for the second left singular vector gives the tray location for the second temperature sensor. The feasibility of dual ended temperature control is reflected in the ratio of the two diagonal elements, σ_1 and σ_2 , of $\mathbf{\Sigma}$. The diagonal elements in $\mathbf{\Sigma}$ are always in decreasing order of magnitude. If the two singular values are comparable, ie the ratio σ_1/σ_2 is not too large (say < 10), dual temperature control should be possible.

Of the above three criteria, the maximum slope criteria is the simplest to use. Sensitivity analysis and SVD analysis requires the availability of a rating program to calculate the tray temperature sensitivities to the manipulated variables. The SVD technique further requires a module to obtain the SVD decomposition of the sensitivity matrix \mathbf{S} . In most distillation column studies, the three techniques would agree on sensor location. However, for columns with highly non-ideal columns, the use of the SVD technique is recommended for selecting the tray location.

5.5. Considerations in Temperature Inferential Control

5.5.1. Effect of LLK / HHK

The temperature composition relationship is not exact for multi-component mixtures. If the feed LLK composition increases, the LLK must leave up the top of the column. For the same LK/HK split, the tray temperatures in the enriching section should be lower as the LLK composition must increase due to the increase in its feed composition. This dip in the temperature would be more as one moves up the enriching section since the LLK accumulates at the top. If a tray temperature near the top is controlled using the reboiler duty and the tray temperature set-point is not reduced on increase of feed LLK, more of the HK would be pushed up the top of the column by the action of the controller. Controlling a tray temperature near the feed tray would mitigate this effect. Another option is to measure the HK composition in distillate and use it to compensate the control tray temperature set-point. Note that the volatilities dictate that any LLK entering the column must exit up the top. In case LLK in the distillate is not acceptable, action must be taken upstream to ensure LLKs do not enter the column. Troubleshooting the process would typically reveal an upstream column not doing its job.

Similar to LLK, if a stripper tray temperature low down the column is being controlled using boilup, an increase in feed HHK would cause more LK to leak down the bottoms, unless the tray temperature setpoint is appropriately increased.

5.5.2. Flat Temperature Profiles

When the key components in a mixture are close boiling, the column temperature profile is flat with only a small change in adjacent tray temperatures. This is typical of superfractionators that use a large number of trays and a high reflux ratio as the separation is inherently difficult. Controlling a tray temperature is then not desirable as variations in the tray pressure with changes in column internal flow rates would swamp any subtle variations in the tray temperature due to composition changes. Controlling the difference in two tray temperatures that are located close by mitigates the effect of pressure variation as the change in the local pressure for the two trays would be about the same. The differential temperature measurement then reflects the change in the HK (or LK) composition between the trays. Care must be exercised in the use of a differential temperature measurement as the variation in ΔT with the bottoms composition depends on the location of the separation zone inside the column. If the measurement trays are below the separation zone, ΔT increases as the steam rate is decreases. Once the separation zone passes below the ΔT trays, a decrease in the steam would cause the ΔT to decrease. The gain thus changes sign depending on the location of the separation zone inside the column.

5.5.3. Easy Separations

The other extreme to a flat temperature profile is an extremely sharp temperature profile. This occurs when the separation is very easy so that the separation zone shows a large change in temperature over a few trays ie a sharp temperature profile. During transients, this sharp separation zone may move up or down the column leading to temperature transmitter saturation. Once the separation zone moves up and continues to move up, the error signal that the controller sees does not change so that the burden of bringing the profile back falls on the integral action. The problem is compounded by the low controller gain due to the extreme

sensitivity of the tray temperature to a change in the manipulated variable. The problem is solved by controlling the average temperature of the trays over which the profile moves.

5.6. Control of Complex Column Configurations

5.6.1. Side-draw Columns

Side product streams are sometimes withdrawn from a column when the product purity specifications are not very tight and there is small amount of impurity in the feed that must be purged. Two common configurations are a liquid side-draw from a tray above the feed tray or a vapour side draw below the feed tray. Consider an ABC ternary mixture. If the component flow rate of A in the fresh feed is small, the liquid side stream withdrawal above the feed tray allows most of the B to be removed in the side stream. The side-draw must be liquid as A being the LLK would be present in smaller amounts in the liquid phase. The vapour side draw below the feed tray is used when there is a small amount of C (compared to A and B) in the fresh feed. The C HHK would separate into the liquid phase so that a vapour side stream that is mostly B with small amounts of C can be withdrawn below the feed tray. The side stream (liquid or vapour) provides an additional operation degree of freedom and its flow rate may be adjusted to maintain the B purity in the side draw. The control schemes are illustrated in Figure 5.6.

Alternative simpler control schemes are possible when the light A or heavy B impurities occur in very small amounts in the fresh feed. The purge rate is flow controlled with a set-point corresponding to the maximum expected impurity component flow in the feed. When the impurity is below this maximum, small amount of LK or HK would be lost with purge. However, the loss is acceptable due to the very small purge rate. The alternative simpler control schemes for the two common side draw configurations are shown in Figure 5.7.

5.6.2. Side Rectifier / Side Stripper Columns

The side rectifier and side stripper columns are an extension of the side-draw column discussed above. As with side-stream columns, these are used when there is a small amount of light or heavy impurity in the feed that is removed as a small purge stream. However the purity specs on the main products are tight. The vapour or liquid side stream respectively, must then be further rectified or stripped to ensure that the impurity is pushed back into the main column and does not escape with side-product stream to ensure high purity. The side stripper and side rectifier column arrangements are shown in Figure 5.8(a) and (b). An additional operation degree of freedom is introduced in the form of the reflux rate or the reboiler duty. The side draw rate and the reflux rate or reboiler duty can then be adjusted to maintain the two impurities in the side-product. The corresponding control schemes are shown in Figure 5.8(a). Along with the two composition loops in the main column, these schemes represent a highly coupled 4X4 multivariable system. Simpler control schemes with only one temperature (or composition) being controlled in each of the main column and the side-column are much more practical.

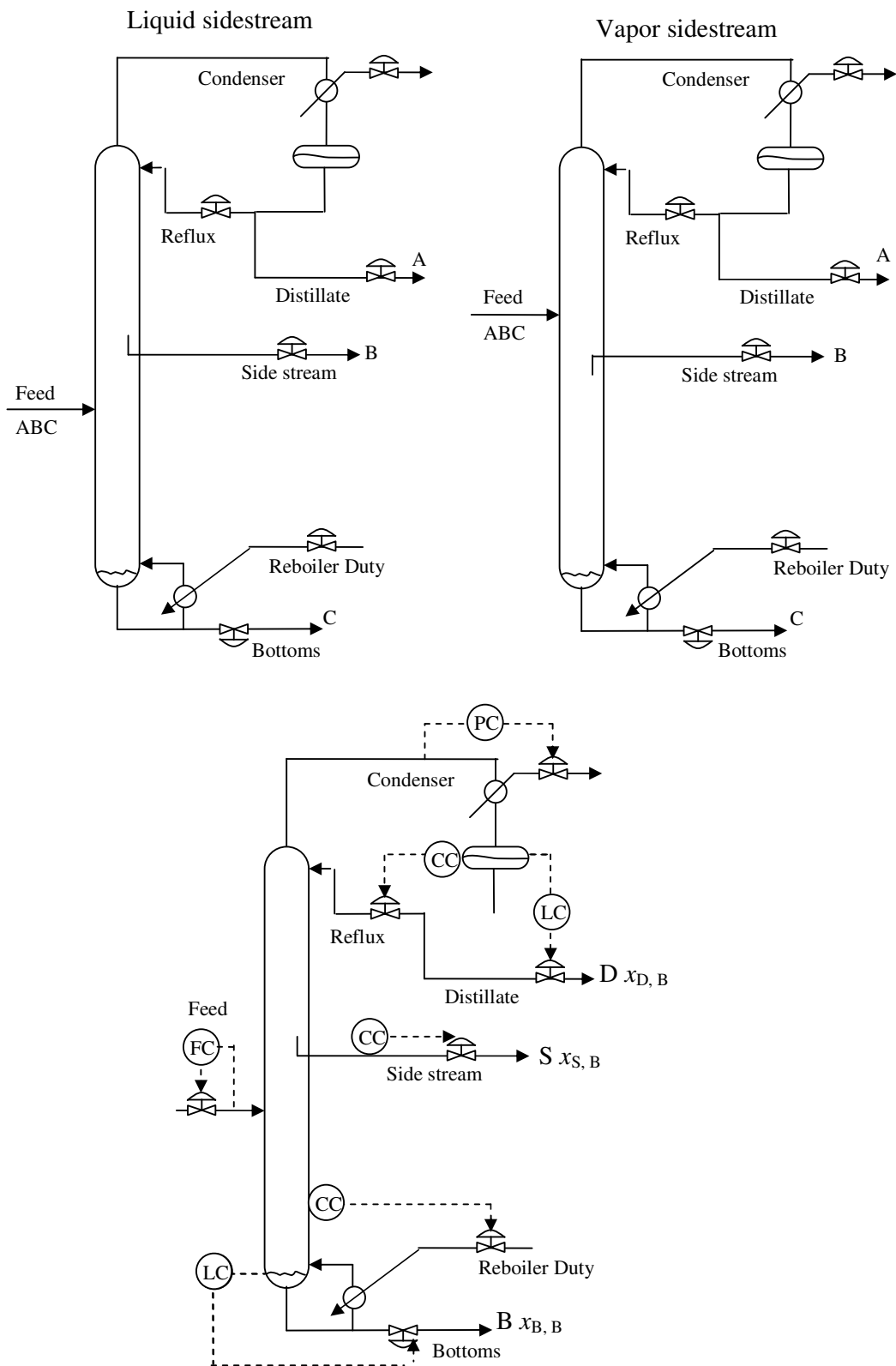


Figure 5.6. Control of side stream column

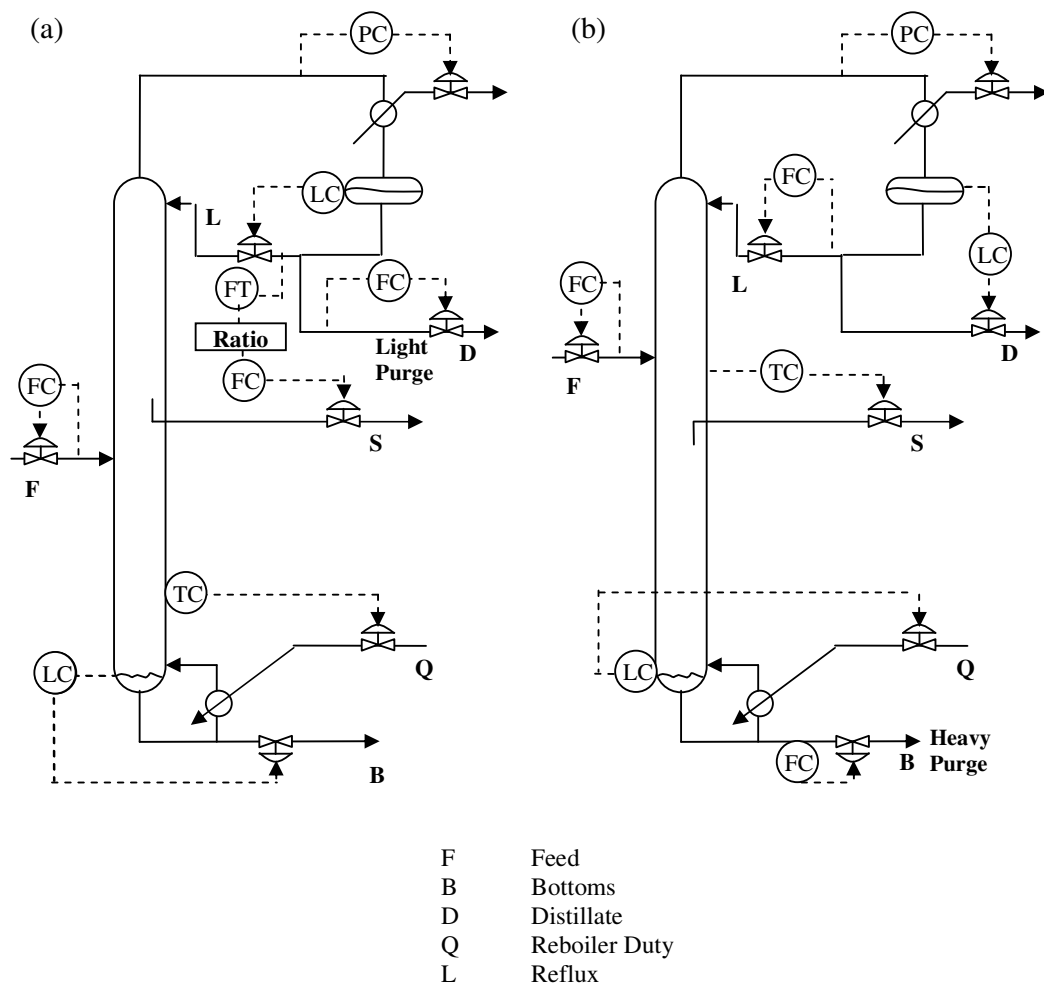


Figure 5.7. Control of purge columns of (a) Liquid side draw and (b) Vapor side draw

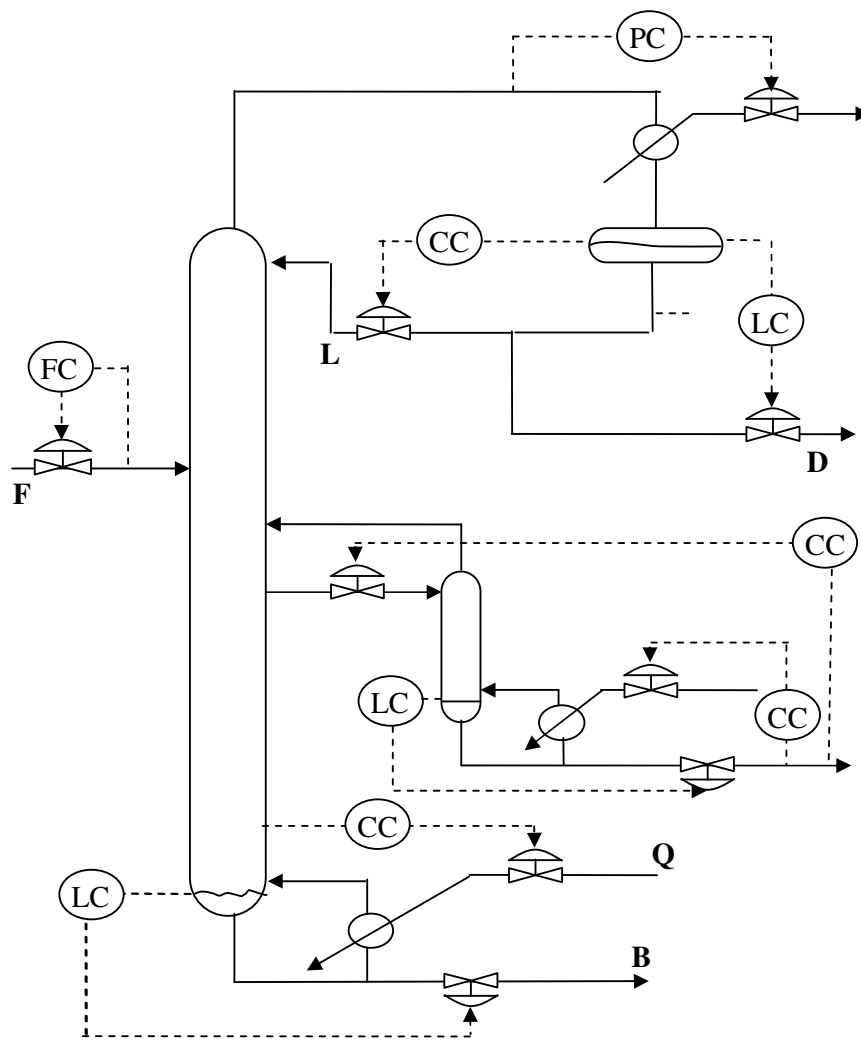


Figure. 5.8(a). Side stream column with stripper

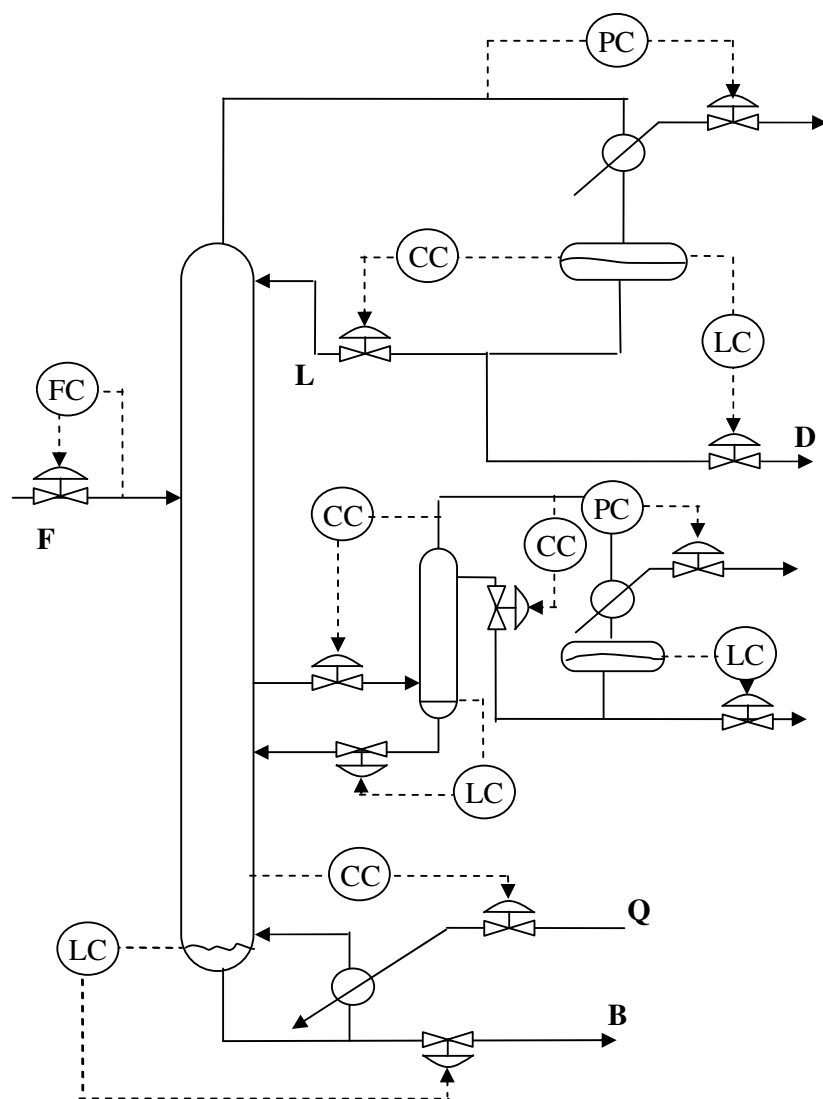


Figure 5.8(b). Side stream column with rectifier

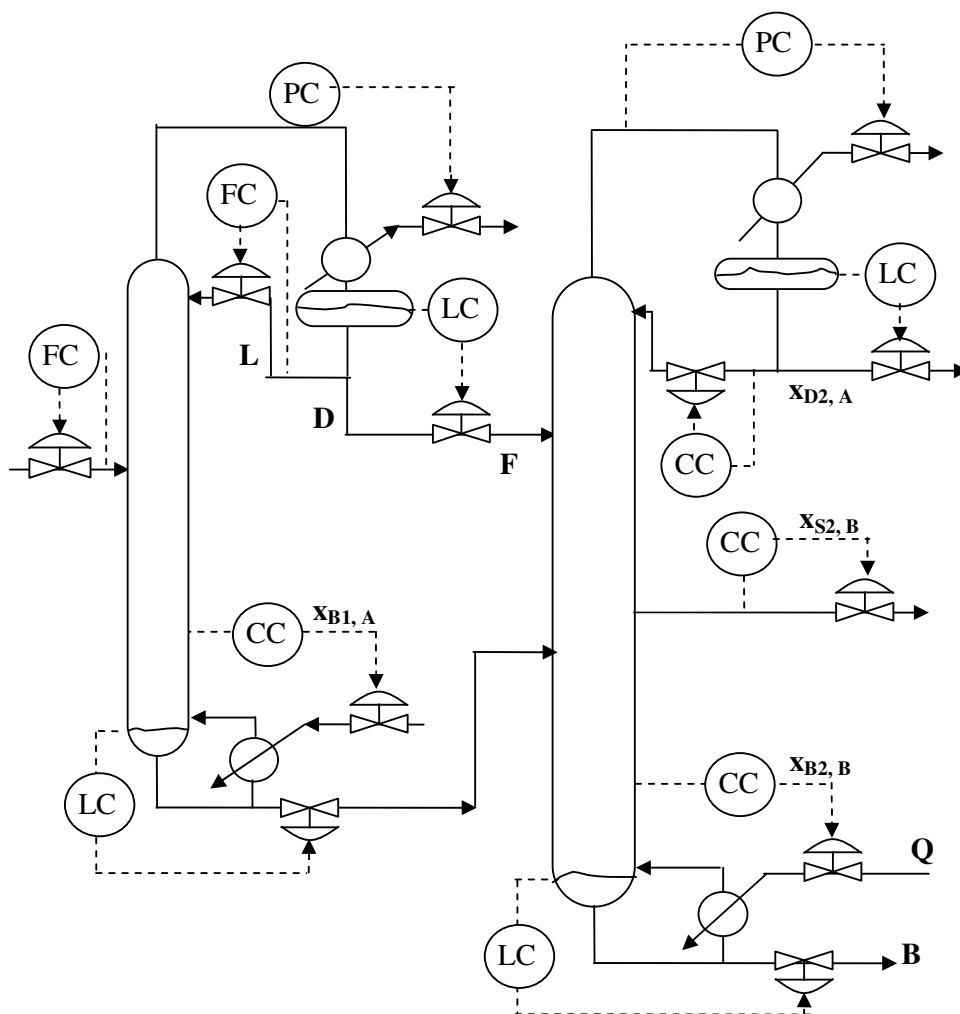
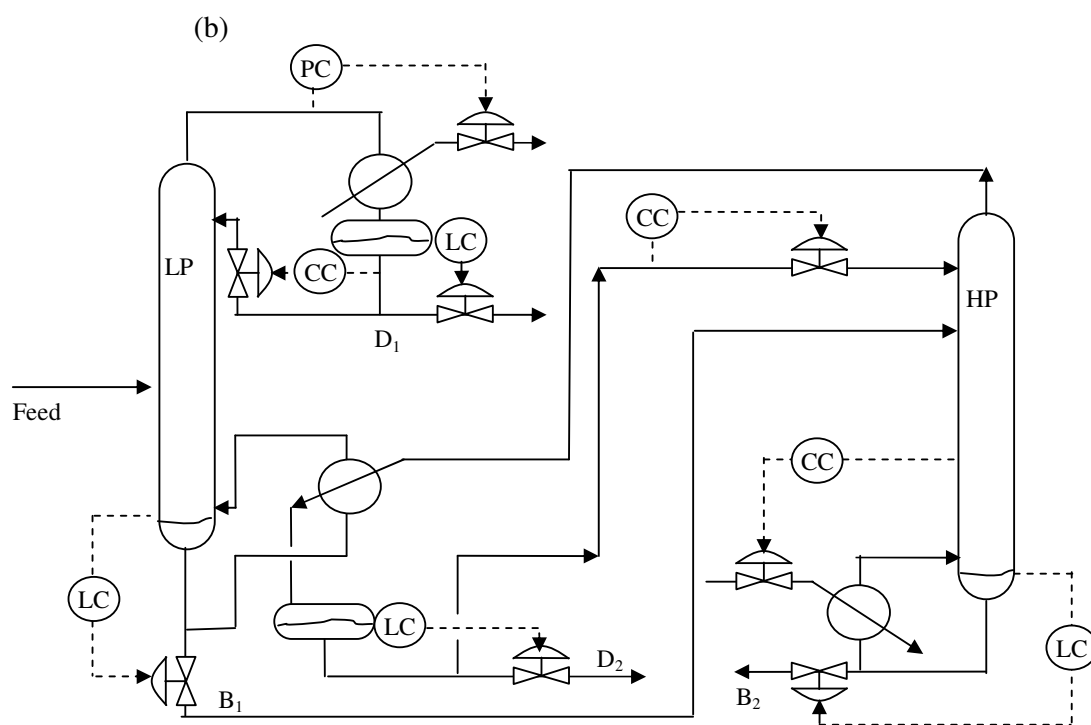
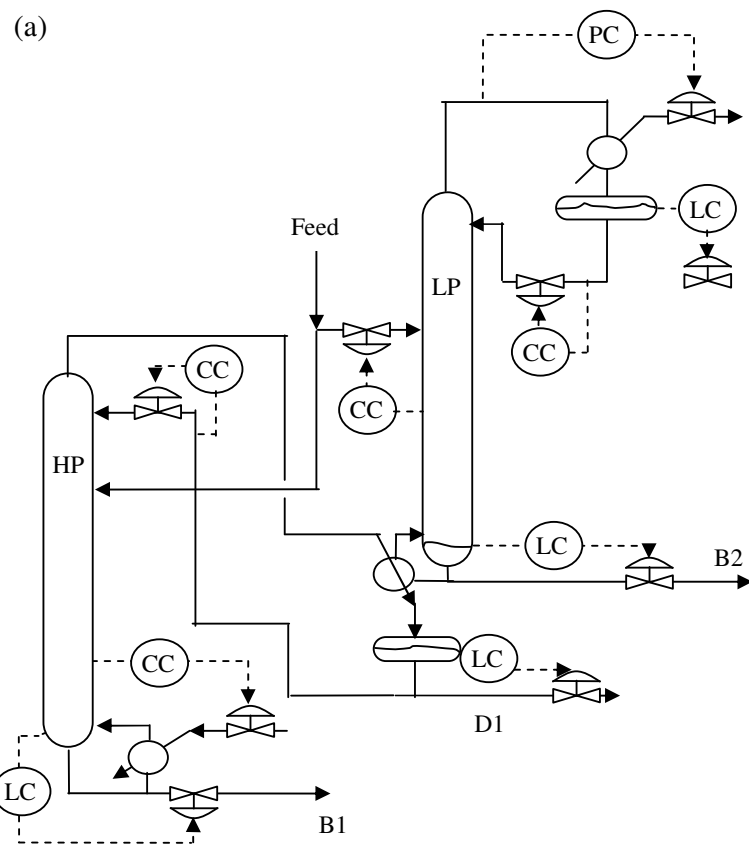


Figure 5.9. Side stream column with prefractionator

5.7. Control of Heat Integrated Columns

Heat integration arrangements in columns consist of the hot vapour from high pressure column providing the energy for reboil in a low pressure. The reboiler for the low pressure column then also acts as the condenser for the high pressure column. Three possible heat integration schemes are shown in Figure 5.10. In the feed split scheme (Figure 5.10(a)), a binary fresh feed is split and fed to two columns. One of the columns is operated at high pressure and the other at low pressure. The pressure difference is chosen so that the hot vapour is 10-15 C hotter than the low pressure column reboiler temperature. The temperature difference provides the driving force for reboiling the liquid in the low pressure column. In the control structure shown, note that the feed to the low pressure column is adjusted so that the bottoms composition is maintained. This is because the reboiler duty in the low pressure column cannot be manipulated. Heat integration thus leads to the loss of a control degree of freedom. Also note that this heat integration scheme is used for a binary separation as the presence of LLK / HHK components can affect the column temperature profiles sufficiently so that the temperature driving force necessary for heat transfer in the low pressure column reboiler disappears.



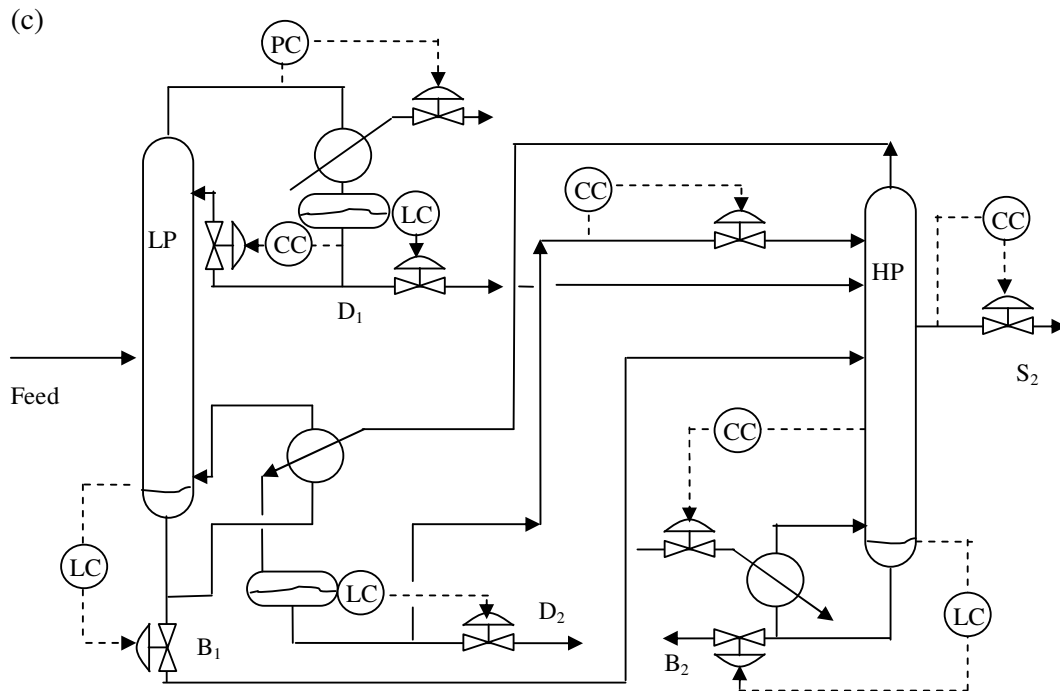


Figure. 5.10. Heat integrated columns (a) Split of feed (binary); (b) Light split reverse (binary); (c) Prefractionator reverse (ternary)

Figure 5.10(b) shows the reverse light split heat integration scheme. Approximately, half the light component is removed as the distillate from the first low pressure column. The bottoms is fed to the high pressure column to remove the light and heavy components as the distillate and bottoms respectively. The hot vapour from the top of this column is condensed to provide vapour reboil into the low pressure column. The direction of heat integration is reverse to that of the process flow. Hence the name, reverse light split. The forward light split configuration is also possible. The control structure for this heat integration scheme is self-explanatory.

Figure 5.10(c) shows another heat integration scheme that can be used for ternary mixtures. The scheme is the same as prefractionator side-draw complex column discussed previously (Figure 5.9) except for reboiler in the pre-fractionator (low pressure column) acting as the condenser for the main column (high pressure) through heat integration. The control structure for this configuration is again self explanatory.

5.8. Control of Homogenous Extractive Distillation System

Homogenous extractive distillation is used to separate a binary AB mixture that cannot be separated due to relative volatility approaching one or the presence of a binary azeotrope. As shown in Figure 5.11, A heavy solvent S is added near the top of the first column, the extractive column, to soak in one of the components (say B). The distillate from the first column is then near pure A. The bottoms, a mixture of S and B, are fed to the solvent recovery column that recovers the heavy S in the bottoms and recycles it back to the extractive distillation column. The distillate from the solvent recovery column is pure B. The control scheme depicted in the Figure manipulates the reboiler duty in the two columns to

keep respectively A and B from falling down the bottoms in the two columns. The solvent into the extractive column is ratioed to ensure fresh feed rate to ensure enough solvent for extracting component B. Note that the bottom sump level is not controlled and the sump must provide enough surge capacity for handling the expected variation in the fresh feed flow rate. Any loss of the solvent over long time is made up by a make-up solvent stream (not shown).

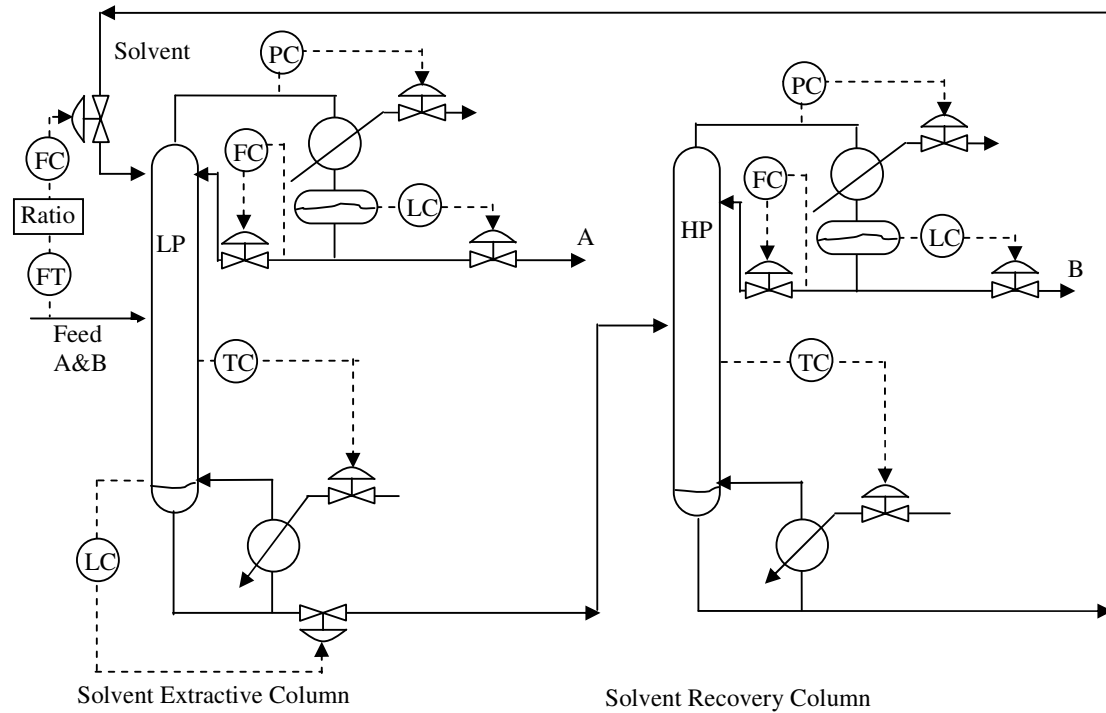


Figure 5.11. Control of Extractive distillation column

5.9. Plant-wide Considerations

In the plant-wide context, the distillate and / or bottoms would feed into a downstream unit such as another distillation column. The variation in the distillate / bottoms flow rate then acts as a disturbance into the downstream unit. The LQ control structure is particularly preferable as the reflux drum and reboiler levels are controlled using the P only controller which results in a smooth flow change into the downstream unit. If however, the DQ (or LB) control structure is used in a high reflux ratio (reboil ratio) column and a tray temperature is controlled using D, the reflux drum level controller manipulates L and must be tightly tuned for a fast dynamics of the of closed loop temperature controller. The D would then show large changes disturbing the downstream unit. Feedforward control action can and should be used to mitigate the propagation of variability to downstream units. For example, the distillate rate may be ratioed to the feed rate with the composition / temperature controller setting the ratio set-point. Alternatively, the distillate may be moved in ratio with reflux with the composition controller setting the reflux ratio set-point. The reflux level controller is then tuned as a P only controller for smooth changes in the reflux and hence the distillate. The

variability in the distillate rate can thus be greatly reduced improving the overall plant-wide control performance.

The vapour distillate from a partial condenser can be manipulated to control the tower pressure. This is however not a good idea if the vapour stream feeds directly into a downstream unit and not a surge tank. In such a scenario, the column pressure should be controlled using the condenser cooling duty, the reflux drum level controlled using the reflux rate and the vent rate moved in ratio with the reflux. This arrangement mitigates the propagation of variability downstream.

The control of energy integrated distillation columns can also be problematic as a disturbance on the hot vapour side necessarily affects the boil-up in the reboiler using the hot vapour as the heat source (instead of steam). To maintain the control tray temperature in the heat integrated column, an auxiliary reboiler (or condenser, as appropriate) is provided. The heat integrated reboiler and the auxiliary reboiler may be arranged in a parallel or a series arrangement. The series arrangement is preferred as the temperature variations in the hot vapour are attenuated due to variation in the temperature driving force in the auxiliary reboiler as shown in Figure 5.12(a). In the parallel arrangement, the auxiliary reboiler must adjust for the variability on the hot vapour side after it has entered the column. One way to prevent this is to use a total heat input controller as shown in Figure 5.13.

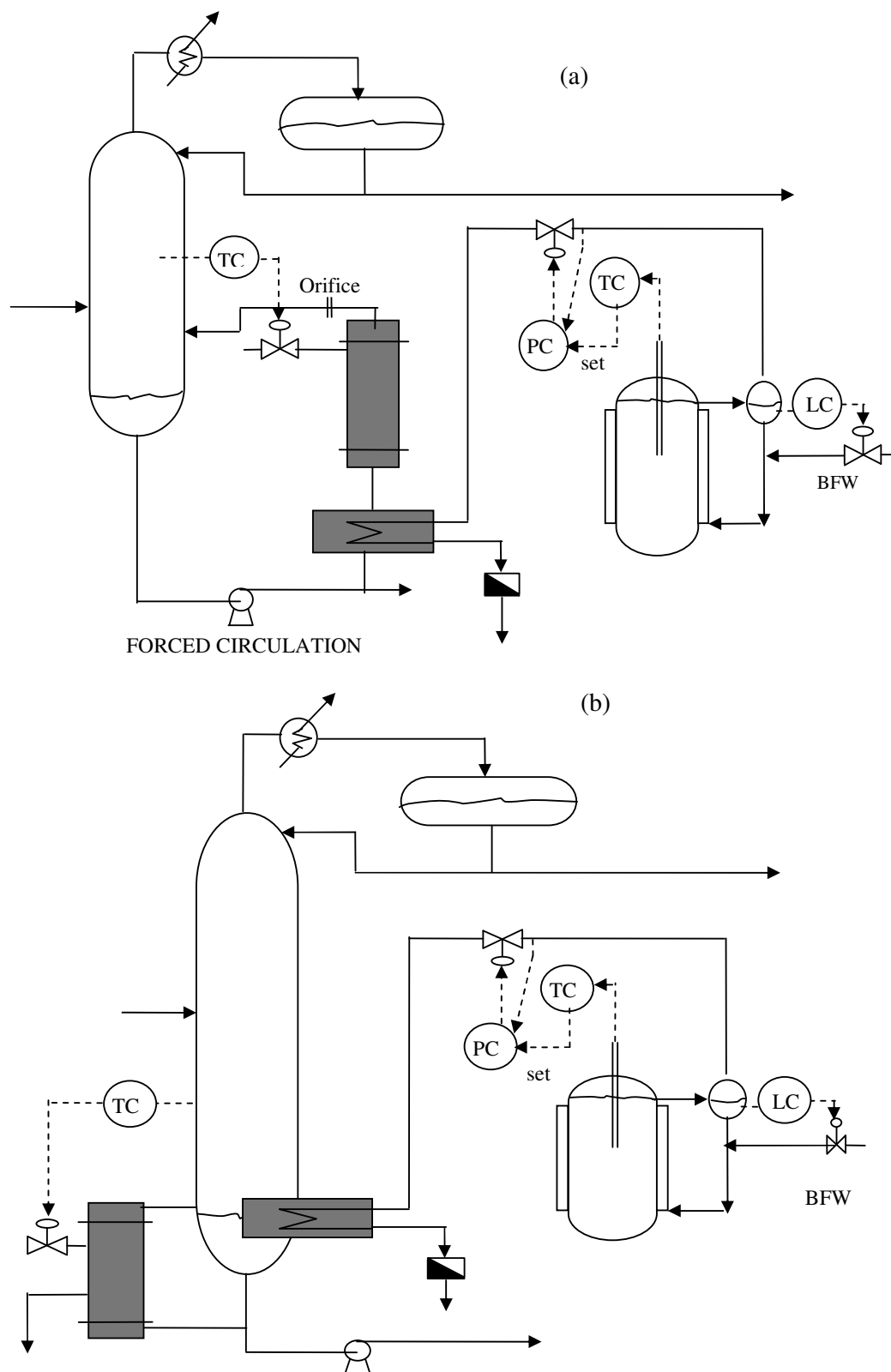


Figure 5.12. Reactor/column heat integration with auxiliary reboiler in (a) series (b) parallel

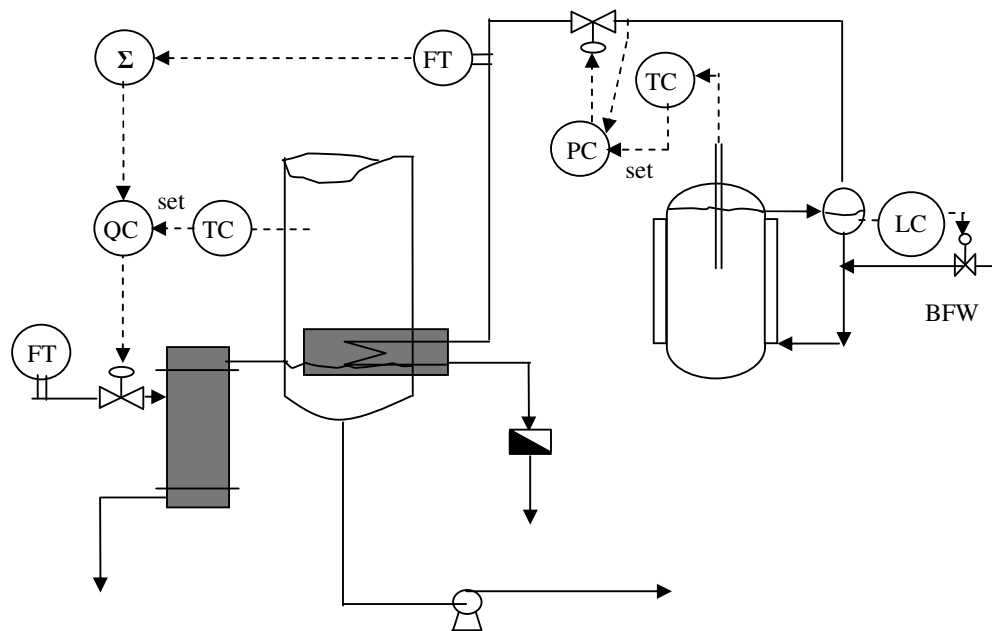


Figure 5.13 Reactor/column heat integration with auxiliary reboiler in parallel and Q controller

Chapter 6. Control of Reactors

A reactor is the heart of a chemical process where the reactants undergo the desired chemical transformation to the products. The transformation is usually incomplete and is also accompanied by undesirable transformation to by-products through side reactions. The reactor operating conditions of temperature, concentration and flow rates determine the production rate of the main product and the side-products. The downstream separation load for separating the unreacted reactants from the products and recycling them back to the reactor is also determined by the reactor. Proper operation and control of the reactor is then crucial to determining the overall process operating profit. The reactor conversion (or yield) and selectivity are the two most commonly used reactor performance metrics that are directly related to the economics of the process. The conversion is defined as the fraction (or %age) of reactant in the feed that reacts to form product(s). The yield is the conversion of a key reactant, usually the limiting reactant. The selectivity is defined as the desired product generation rate relative to the total product generation rate (including all undesired side products). The yield and selectivity represent two key economic objectives of any process. A lower conversion would result in greater energy consumption to separate the unreacted reactants from the products and recycle them back to the reactor. The energy cost per kg product would then go up. A low selectivity represents an economic loss as more of the costly reactant gets converted to the undesired product with lower (or worse, a negative) profit margin. Typically, as the conversion increases, the selectivity decreases so that the usual philosophy is to operate the reactor near the maximum conversion for which the selectivity is acceptable (say >95%).

Even as proper reactor operation is the key to profitability, controlling a reactor offers unique challenges as most reactions are accompanied by the generation or consumption of heat. The reaction heat generation / consumption alters the temperature of the reaction mixture which in turn affects the reaction rate and hence the rate of heat generation / consumption. The coupling of the thermal and reaction effects leads to a highly coupled non-linear system. The reaction rate, r ($\text{kmol.s}^{-1}.\text{m}^{-3}$ or $\text{kmol.s}^{-1}.\text{kg}^{-1}$ catalyst), for an irreversible reaction $A + B \rightarrow C$ would generally vary as

$$r = k.c_A^\alpha.c_B^\beta.$$

In the above expression, c_A and c_B are the concentrations (kmol.m^{-3}) of A and B respectively, k is the reaction rate constant, and α and β are the reaction order (typically > 0) with respect to A and B respectively. The units of the reaction rate constant depend on the reaction order and it follows the Arrhenius temperature dependence as

$$k = k_0.e^{(-E/RT)}$$

where E is the activation energy and k_0 is the Arrhenius frequency factor.

The form of the kinetic expressions above shows that the reactant concentration and the reaction temperature are the two basic manipulation handles for adjusting rate of product generation in a reactor. The reaction rate doubling for every 10 deg C increase in the temperature is an oft quoted rule of thumb. The reactor temperature is thus a dominant variable that significantly affects the reaction rate. When the reactant concentration is adjusted for changing the production rate, altering a key reactant concentration affects the reaction rate more than changing the concentration of other reactants. This is because most reactors must be operated such that one of the reactants is limiting, i.e. in excess of other reactants. The non-stoichiometric environment is necessary to suppress side reactions. For example, consider the main irreversible reaction $A + B \rightarrow C$. The product C can further react irreversibly with B to form an undesired product D as $C + B \rightarrow D$. For this reaction scheme, the reactor must be operated in an excess A environment so that the limited availability of B

for further participation in the side-reaction suppresses by-product generation. Clearly, changing the limiting reactant B concentration would affect the reaction rate more than changing the excess reactant A concentration. In many industrial reactors, the reactor temperature and the limiting reactant concentration are the two dominant variables that are directly / indirectly adjusted for changing the product generation rate.

In exothermic reactions, the use of a selective catalyst lowers the activation energy for the main reaction. The activation energy for the side reactions is thus more than for the main reaction. In case the temperature is increased for increasing the production rate, the Arrhenius temperature dependence of the rate constant causes a larger relative change in the side-reaction rate. Thus if the main reaction rate increases by say 5%, the side reaction rate would increase by more than 5% (say ~10%). The reaction selectivity thus goes down. The adjustment of the reactor temperature for increasing the production rate thus must consider the detrimental effect on selectivity. In many industrial reactors, the reactor temperature is usually adjusted to compensate for catalyst poisoning / deactivation so that the overall reaction rates do not decrease over time.

6.1. Basic Reactor Types

The continuous stirred tank reactor, the plug flow reactor and the packed reactor are the most common reactor types used in the continuous process industry. These basic reactor types are shown in Figure 6.1. The PFR and PBR are similar except that the latter holds a catalyst bed to facilitate the reaction. The CSTR and the PFR (or PBR) differ fundamentally in terms of back mixing. In the PFR (and PBR), the fluid travels along a pipe as a plug so that every atom entering the reactor spends the same amount of time inside the reactor before exiting. This time is also referred to as the reactor residence time. Plug flow, by definition, implies no back mixing. The exact opposite of plug flow is perfect back mixing as in the CSTR. The back mixing is accomplished using agitators, spargers and fluidization.

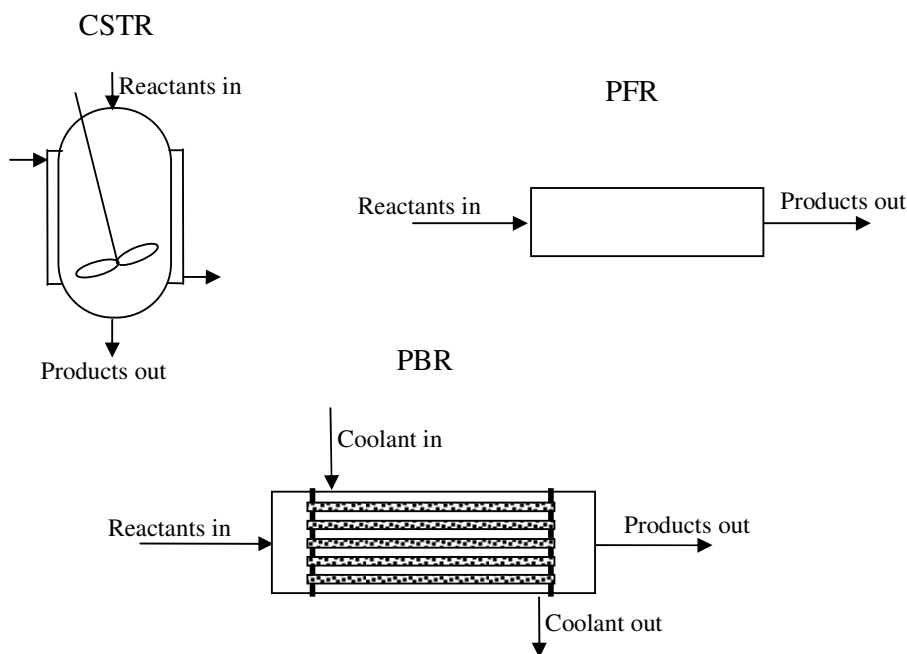


Figure 6.1. Basic reactor types

6.2. Plug Flow Reactor

6.2.1. PFR Basics

To understand the behaviour of a PFR (or PBR), imagine a plug of fluid flowing through the reactor. As it moves through, the concentration of the reactants goes down as they undergo reaction. Assuming adiabatic operation and an exothermic reaction, the heat released due to reaction would heat up the plug. The increase in the plug temperature causes the reaction rate to increase further so that the temperature in the initial part of the reactor increases exponentially. At a sufficient length down the reactor, the reaction rate begins to decrease due to the limited availability of reactants. For a large enough reactor length, the reaction rate would go to zero as the limiting reactant gets exhausted. The temperature profile for an adiabatic PFR thus resembles a sigmoid as shown in Figure 6.2. The difference in the inlet and outlet temperature is referred to as the adiabatic temperature rise. If the reaction is highly exothermic, the adiabatic temperature rise is large, which is usually unacceptable due to reasons such as promotion of side reactions at the higher temperatures, possibility of catalyst sintering in a PBR, increase in the material of construction cost etc. The cooled PFR / PBR is then used

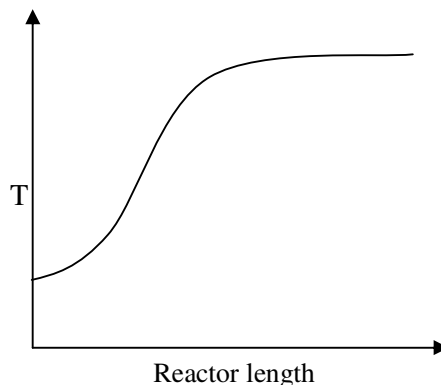


Figure 6.2. Temperature profile for an adiabatic PFR

The most common cooled PBR arrangement is shown in Figure 6.3. Catalyst is loaded in the tubes of a shell and tube heat exchanger. Pressurized water is re-circulated on the shell side. The water carries the reaction heat to form steam in the steam drum. The high recirculation rate ensures a near constant temperature on the shell side. Unlike adiabatic operation, the temperature profile initially increases as the rate of heat generation is more than the cooling rate and later decreases with the reaction rate decreasing due to reactant depletion and the cooling rate increasing due to the higher temperature driving force. The temperature profile thus exhibits a maximum, also referred to as the hot spot. For highly exothermic systems, the reactor temperature profile can be extremely sensitive to the operating conditions, in particular the reactor inlet temperature and the shell side temperature.

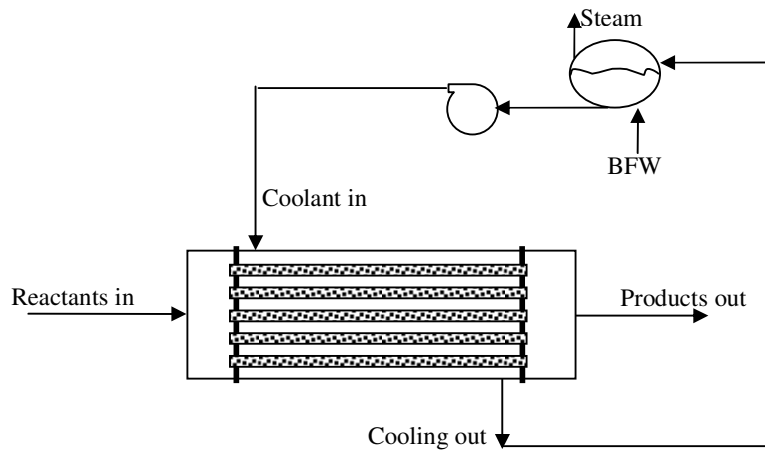


Figure 6.3. Cooled Packed bed reactor

If the reactant inlet temperature (or coolant temperature) is too low, reaction may not kick in leading to the quenched state. If the reactant inlet temperature is too high, the reaction can proceed so fast that only a small fraction of the heat released gets removed resulting in a temperature run-away. The quenched, hot-spot and run-away reactor temperature profiles are illustrated in Figure 6.4.

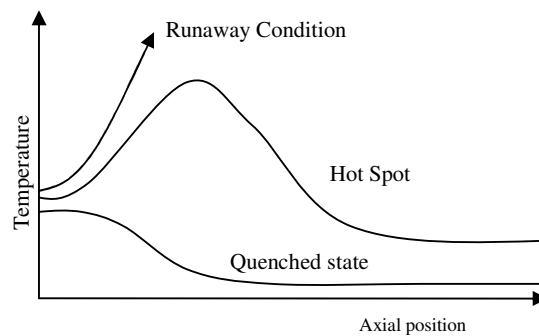


Figure 6.4 Cooled plug flow reactor temperature profiles

The adiabatic or cooled tubular reactor is commonly used in the process industry for gas phase reactions. The reactor is usually operated at the maximum equipment pressure so that the reactant partial pressures are as high as possible for maximum reaction rate. This also reduces the design volume of the reactor for a given conversion. The control of PFRs is considered next.

6.2.2. Control of PFRs

6.2.2.1. Adiabatic PFR

The adiabatic PFR is the simplest reactor configuration and is used when the adiabatic temperature rise is acceptable. The reactants (fresh + recycled) are heated to the reaction temperature using a furnace. The furnace heat duty holds the reactor inlet temperature constant. This is shown in Figure 6.5(a). The adiabatic temperature rise sets the reactor outlet temperature. Sometimes the outlet temperature is also controlled to maintain the reactor conversion and selectivity. The limiting reactant fresh feed rate may be used as the manipulation handle as illustrated in Figure 6.5(b).

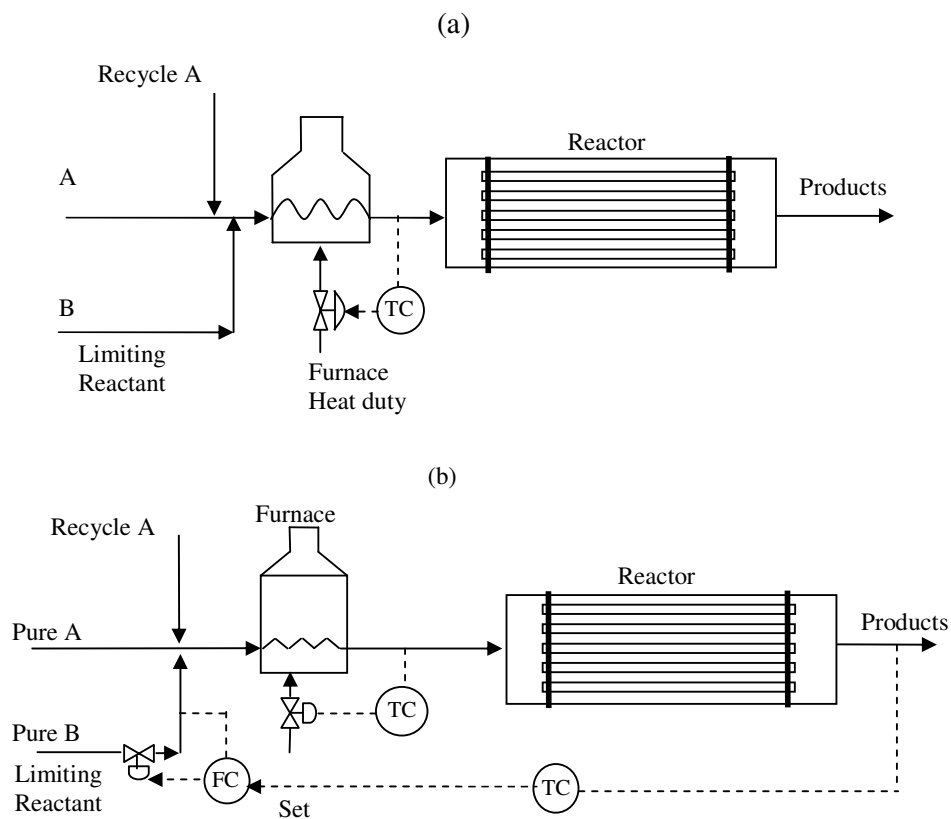


Figure 6.5. Simple adiabatic PFR structures

6.2.2.2. Cooled Tubular Reactors

Cooled tubular reactors are more challenging from the control perspective. The hot spot temperature must be tightly controlled to prevent a runaway. This is accomplished using auctioneering temperature control as illustrated in Figure 6.6. The measurements from an array of thermocouples (or RTDs) placed along the length of the reactor are input to a high selector that passes the maximum temperature to the hot spot temperature controller. This controller typically manipulates the reactor cooling duty. The control structure for the most common cooled PBR arrangement is shown in Figure 6.6. The temperature controller sets the steam drum pressure set-point. A change in the drum pressure alters pressurized water boiling point which in turn changes the temperature driving force for heat removal from reactor.

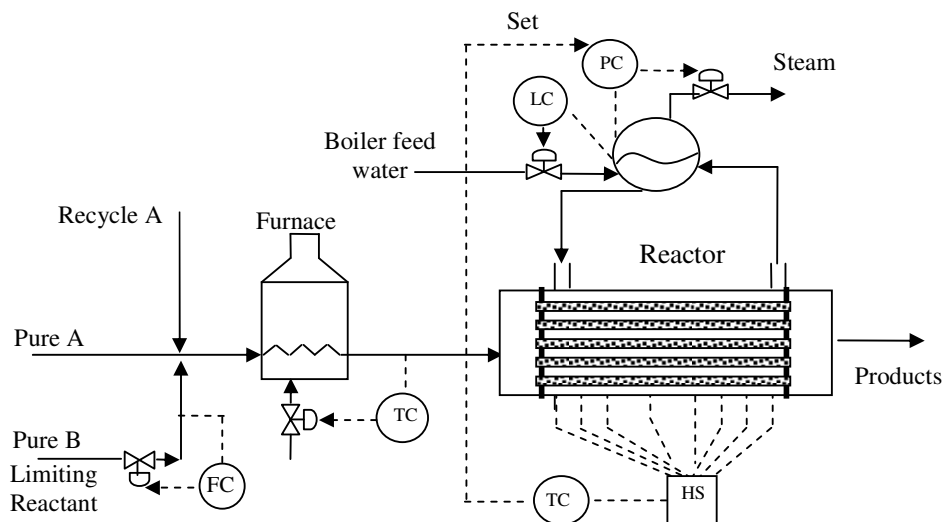


Figure 6.6. Auctioneering temperature control structure

In addition to the reactor cooling duty, there are two other possible manipulation handles for reactor heat management, namely the reactor inlet temperature set-point and the limiting reactant flow rate into the reactor. The schemes are shown in Figure 6.7. Both the schemes work by changing the heat generation due to reaction. The non-linearity between the controlled and manipulated variable is severe in all the three schemes. This may be understood using the analogy of a fire. It is very easy to make a fire whereas extinguishing one requires much effort. Similarly, it requires much more control effort to adjust for an increase in the hot-spot temperature than for the same decrease. The controller thus must be aggressive for deviations in one direction and not too aggressive in the other. Gain scheduling is sometimes used with the magnitude of the controller gain depending on the magnitude and sign of the error signal (or a more sophisticated schedule). The possibility of temperature runaway also requires that large overshoots above the set-point be avoided even as the controller is aggressive. The derivative action is often employed to suppress closed loop oscillations.

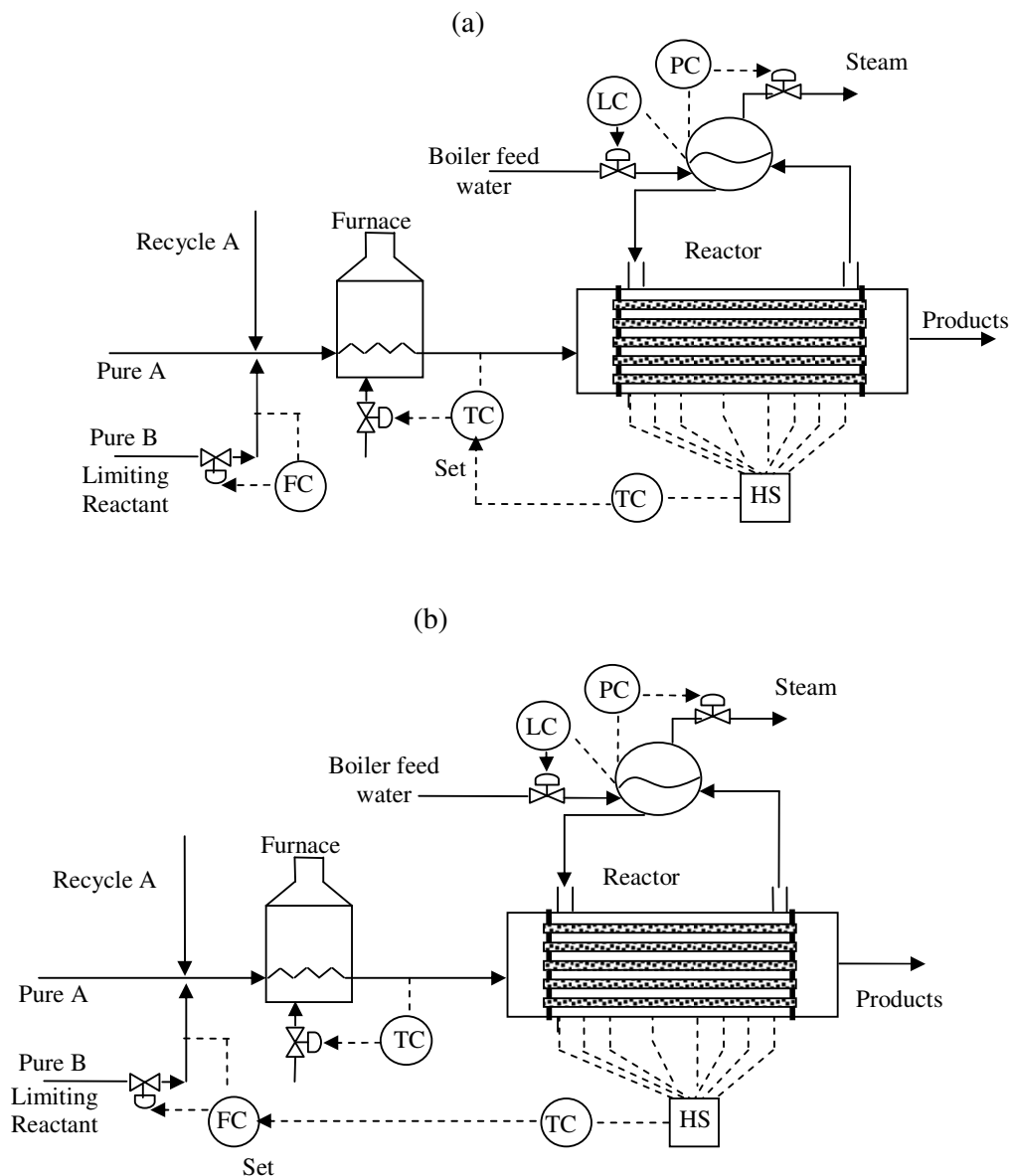


Figure 6.7. Two other possible manipulation handles for reactor heat management

It is noted that controlling the reactor outlet temperature may sometimes provide adequate regulation of the hot spot temperature. The applicability of this much simpler scheme depends on how close the outlet is to the reactor hot spot at the base design condition. When the hot spot is close to the reactor outlet, a change in the outlet temperature correlates well with the change in the hot spot temperature so that controlling the outlet temperature provides adequate regulation of the hot spot temperature.

6.2.3. Intermediate Cooling and Cold-Shot Cooled Reactors

Two other commonly used heat removal configurations, namely, intermediate cooling and cold shot cooling, are shown in Figure 6.8 along with the control structure. Explicit intermediate coolers are provided in equilibrium limited exothermic reactions to increase the overall equilibrium conversion. Cold shot cooling is frequently employed in polymerization reactors where it is extremely important to hold the temperature profile in the reactor to maintain the molecular weight and polydispersity of the product polymer.

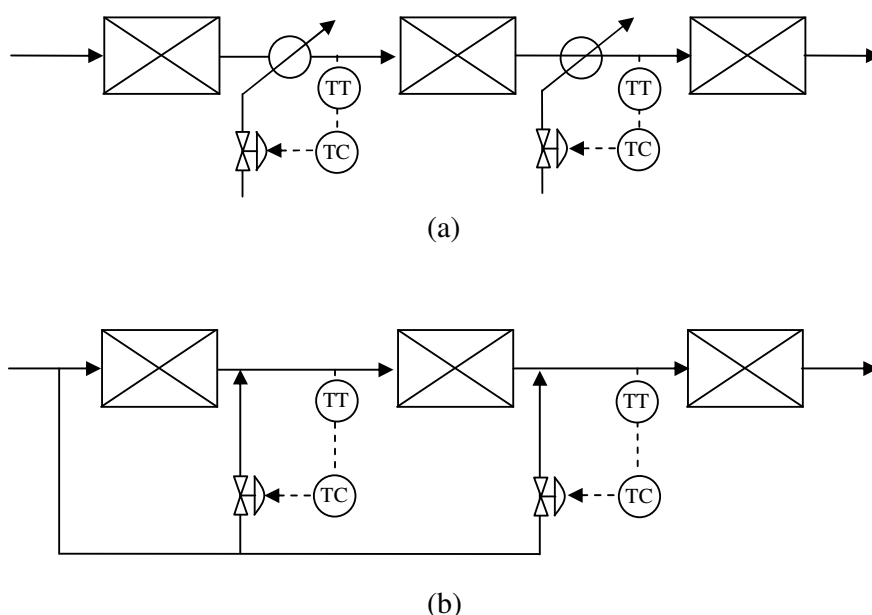


Figure 6.8. (a) Intermediate cooling in sequence reactors
(b) Cold shot cooling

Modern chemical plants frequently recover the reaction heat using heat integration. A common heat integration scheme employs the feed effluent heat exchanger to preheat the cold reactor feed using the hot reactor effluent gases as in Figure 6.9. The heat integration results in energy recycle loop which can lead to instability. Consider the extreme case of a FEHE that heats the reactants to the reaction temperature so that there is no furnace. If the temperature of the cold reactants rises, the reactor inlet temperature will increase. This would cause more reaction accompanied by heat release resulting in an increase in the hot reactor effluent gas. The hotter effluent would cause a further increase in the reactor inlet temperature resulting in a temperature runaway or instability. This instability can be prevented if the reactor inlet temperature (or outlet temperature) is controlled. The furnace performs this function by breaking the positive thermal feedback loop. An alternative to the FEHE is to recover the reaction heat as steam in a waste heat boiler and feed it to the steam utility network. This removes the thermal feedback while ensuring almost 100% heat recovery.

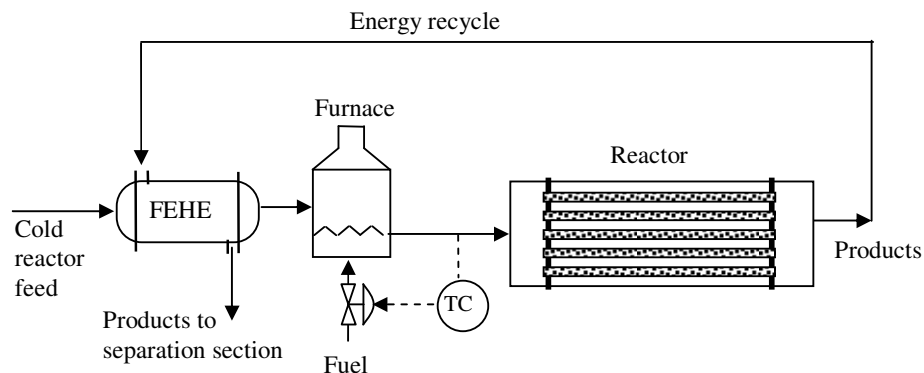


Figure 6.9. Heat integration scheme

Catalytic packed bed reactors differ from simple plug flow reactors in that the catalyst bed constitutes a significant thermal capacitance. The heat capacity of the packed bed can sometimes lead to the outlet temperature exhibiting an inverse response with respect to the inlet temperature. This effect is due to the difference in the propagation rate of the gas and the bed thermal effect through the reactor. If the reactor inlet temperature decreases, the bed temperature does not decrease immediately. The cooler gas plug thus comes in contact with the hot packing and heats up. The reaction now kicks off leading to the outlet temperature increasing. Once the catalyst bed cools down, the outlet temperature of course decreases. This inverse response or “wrong” way behaviour destabilizes the control loop so that a PID controller must be detuned. In cases where the closed loop performance is not satisfactory, the application of advanced control techniques such as the Smith predictor is recommended.

6.3. Continuous Stirred Tank Reactor

Perfect back-mixing occurs in ideal continuous stirred tank reactors. Due to the mixing, the composition at the reactor is the same as the composition inside the reactor. The reaction thus occurs at the reactor outlet concentration. The reactant, upon entering the reactor, thus gets diluted instantaneously to the lower reactor composition. Since the reaction occurs at the exit conditions, the steady state conversion now depends on both the inlet and outlet conditions. This creates higher non-linearity so that the existence of multiple steady states is a distinct possibility. This is in direct contrast to PFRs where the outlet conditions are uniquely determined by the inlet conditions. Back-mixing thus causes material feedback with the possibility of multiple solutions.

Consider a jacketed CSTR as in Figure 6.10. Assume that the coolant flow rate is high so that the jacket temperature is nearly constant. The heat removal rate varies linearly with the jacket-reactor temperature difference. The heat generation due to reaction is an S-shaped sigmoid with respect to the reactor temperature. At steady state the heat removal rate and the heat generation rate must balance each other. The reactor can exhibit a unique steady state or multiple steady states. These scenarios are depicted in Figure 6.11. For multiple steady states, there are three steady states corresponding to a high, intermediate and low temperature. The high and low temperature steady states are stable while the intermediate temperature steady state is unstable. This is because around the intermediate steady state, if the temperature increases slightly, the rate of heat generation increases more rapidly than the rate of heat removal so that the temperature would continue to rise and not return back to the intermediate steady state ie an open loop unstable system. In contrast, at the high / low temperature steady state, the slope of the heat removal curve is more than the heat generation curve in the vicinity of the steady state implying stable open loop behaviour. Typically, reactor operation at the intermediate steady state is desired as the high temperature steady state may lead to catalyst sintering and undesirable side reactions while the reaction rate is small at the low temperature steady state. For such open loop unstable reactors, a temperature controlled must be used to stabilize the reactor. The closed loop system becomes stable for a controller gain above a critical value. At lower gains, the feedback action is not enough to stabilize the unstable system. For extremely large controller gains, the closed loop system becomes unstable due to too much feedback, similar to open loop stable processes. The closed loop system is thus stable only for a range of controller gain. Conventional tuning rules such as the ZN / TLC procedure should therefore not be applied. Multiple steady states are avoided when a large heat transfer area is provided. Unique solution CSTRs are much easier to control and the heat transfer system for CSTRs should be properly designed to avoid multiplicity.

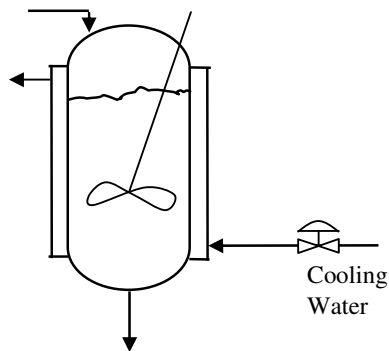


Figure 6.10. High flow rate cooling water temperature control

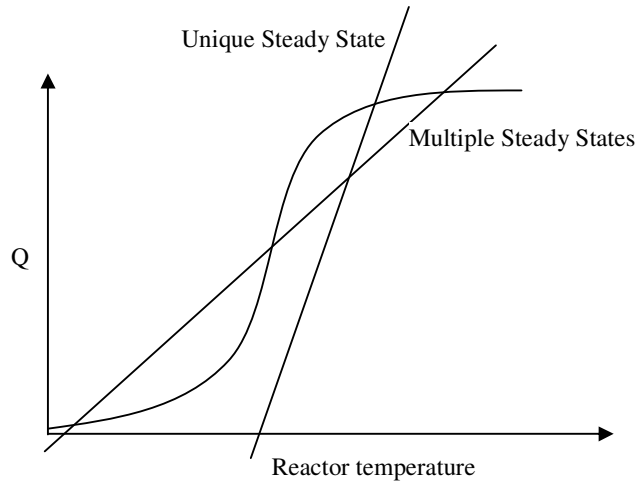


Figure 6.11. Steady states in CSTR

6.3.1. Jacket Cooled CSTR

Evidently, proper regulation of the reaction heat removal is one of the main control tasks. The simplest heat transfer arrangement is that of a jacketed CSTR with cooling water flowing along the jacket. Its flow rate is adjusted to maintain the reactor temperature as shown in Figure 6.12. This approach suffers from drawbacks such as a non-linear process gain due to variation in the heat transfer coefficient with cooling water flow and local reactor hot spots due to the jacket temperature profile. These problems can be mitigated by using a recirculation loop as shown in Figure 6.13(a). The recirculation allows for a constant coolant flow rate so that the jacket temperature is constant. The reactor temperature is maintained by adjusting the fresh coolant flow rate into the recirculation. The recirculation loop introduces and additional thermal lag into the heat transfer loop resulting in a slow closed loop response. The use of a cascade control scheme as in Figure 6.13(b), where the jacket temperature is tightly controlled by adjusting the coolant flow (slave loop) and the reactor temperature loop adjusts this set-point, significantly improves the closed loop reactor temperature control.

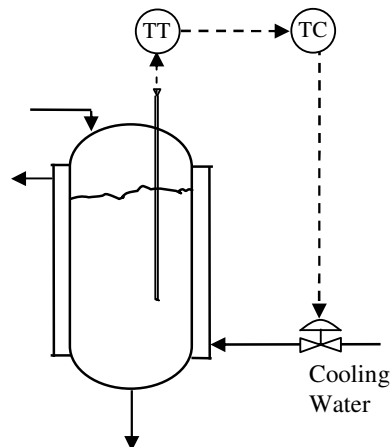
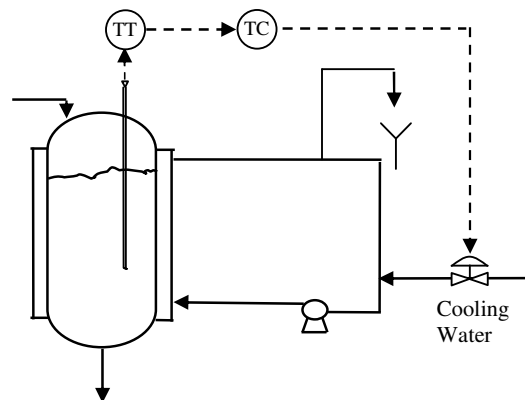
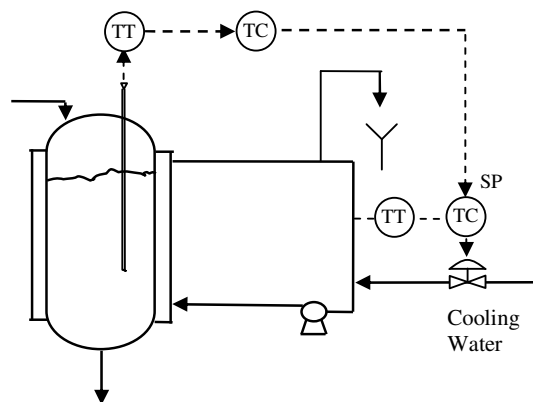


Figure 6.12. Simplest CSTR Control Structure



(a)



(b)

Figure 6.13 (a) Circulating cooling water temperature control
(b) Cascade control

6.3.2. Reaction Heat Removal as Steam

A common arrangement for recovering the reaction heat is to generate steam from the hot pressurized water recirculating in the jacket recirculation loop. The heat removal scheme and the control structure are shown in Figure 6.14. The level in the steam drum would exhibit the inverse response so that the boiler feed water flow into the drum is ratioed to the steam flow with the level controlled adjusting its set-point. This arrangement allows for the feed water to move in the correct direction for load changes.

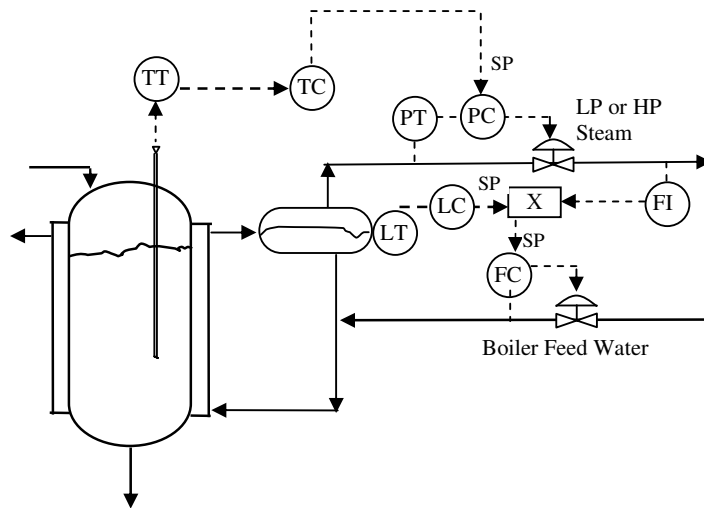


Figure 6.14. Temperature control through steam generation

6.3.3. External Heat Exchanger

In high temperature applications, the generation of steam is not possible so that expensive proprietary coolant oils must be used. An external heat exchanger is then provided as in Figure 6.15 for a closed circuit coolant loop. Cooling water is used as the coolant in the external heat exchanger to remove the reaction heat carried by the hot oil. The control structure shown in the Figure is self explanatory.

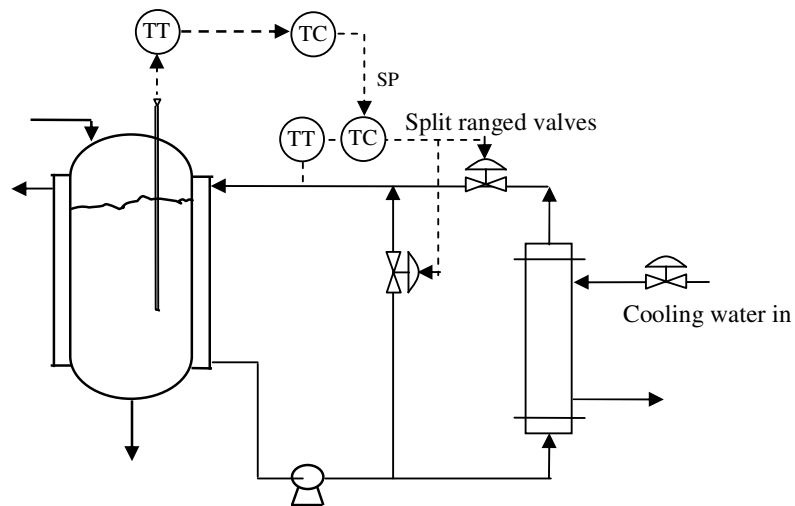


Figure 6.15. Use of external Heat exchanger for temperature control

6.3.4. Cooling Coils

In the jacketed CSTR, the heat transfer area is determined by the reactor volume and may not be sufficient. Cooling coils, as in Figure 6.16, are used for higher heat transfer area per unit volume. The control scheme adjusts the coolant flow rate for maintaining the jacket temperature.

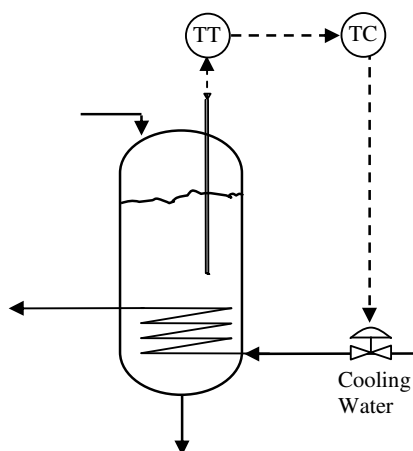


Figure 6.16. Extended heat transfer area through use of cooling coils

6.3.5. External Cooling by Content Recirculation

Another alternative for removing the reaction heat when the jacket heat transfer area is insufficient is to circulate the reaction mixture through an external heat exchanger and feed it back into the reactor. This scheme is shown in Figure 6.17. The reactor temperature is maintained by manipulating the temperature of the cooled reaction mixture from the external heat exchanger. The cooling water flow rate into the heat exchanger is adjusted to maintain the cooled reaction mixture temperature. In this control scheme, the external heat exchanger introduces a significant thermal lag into the slave loop. The dynamics of the slave loop can be significantly improved by a slight design modification providing for bypassing a small fraction of the reaction mixture stream around the external heat exchanger. This is illustrated in Figure 6.18. The thermal lag is thus replaced by the negligible mixing lag as the dominant time constant of the slave loop resulting in significantly improved reactor temperature control.

6.3.6. Boiling CSTR with External Condenser

When the reaction mixture boils, excellent reactor temperature control can be achieved using an external condenser that condenses the vapour and refluxes the cold condensate back into the reactor. The arrangement is shown in Figure 6.19. Note that the condensate flows back into the reactor by gravity so that the condenser should be at a sufficient elevation above the reactor. The U-leg seal is provided to force the vapour to enter the condenser from the correct entry port. Note that the reactor temperature in this case is self-regulatory..

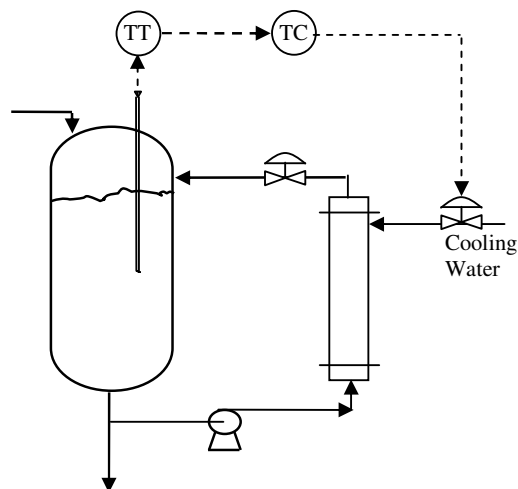


Figure 6.17. Circulation of reactor contents through external heat exchanger

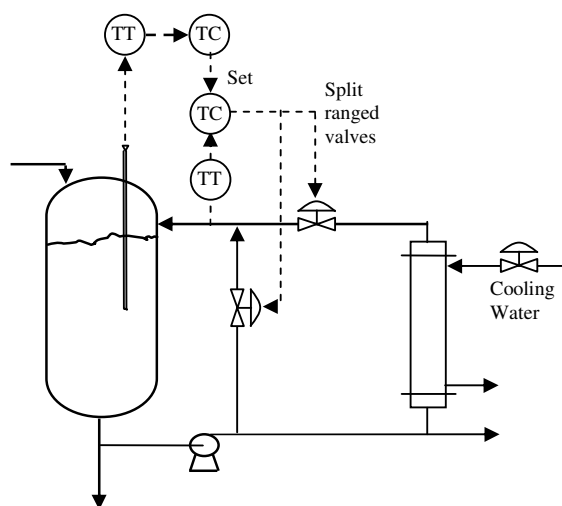


Figure 6.18. Bypass of circulating reactor content around external Heat exchanger

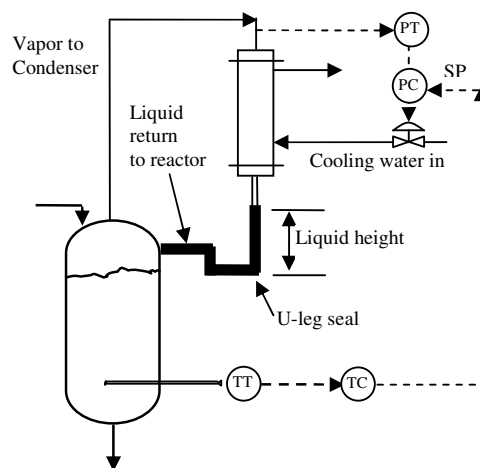


Figure 6.19. Cooling through vaporization of reactor content

6.3.7. Reactor Heat Removal Capacity Constraint

A possibility for controlling the reactor temperature when the heat transfer area is limiting is to adjust the reactant feed rate so that the heat generation due to reaction changes appropriately. The cooling water valve is fully open. The scheme is shown in Figure 6.20. While, appealing in its own right, this control strategy should not be implemented in practice (or used with due caution) as the open loop dynamics of the temperature loop is slow due to the composition lag introduced by the reaction mixture volume. As the feed rate changes, the composition of the reaction mixture changes slowly due to the large reactor hold up. The reaction heat generation thus changes slowly. In case the reactor temperature goes down, the temperature controller would add more feed. The unreacted reactant amount in the reactor thus goes up. Once the reactor temperature begins to increase, reaction would “kick-in” due to the large amount of unreacted reactant inside the reactor. The possibility of a reactor run-away is thus always lurking in the back-drop, especially for highly exothermic reactions. The scheme may be workable for mildly exothermic reactions.

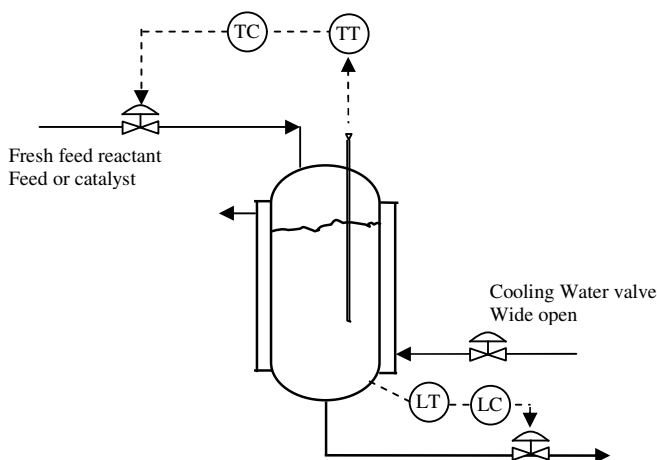


Figure 6.20. Temperature control by manipulating the fresh feed rate

The problem of reactor runaway can be circumvented by the use of a valve positioning scheme as illustrated in Figure 6.21. The reactor temperature is controlled by adjusting the reactor cooling duty. The valve positioning controller measures the cooling duty valve position and slowly adjusts the feed rate so that eventually the cooling duty valve is near fully open and the reactor operates at maximum through-put. In this scheme, the temperature control loop effectively rejects short term disturbances as its dynamics are much faster compared to controlling the reactor temperature directly using the fresh feed rate. Over the long term, the VPC ensures the reactor is operating at near maximum cooling duty, ie maximum through-put. Note that the feed rate can be directly manipulated in a PFR since the material flows through as a plug and there is no back-mixing implying little / no build-up of unreacted reactants inside the reactor.

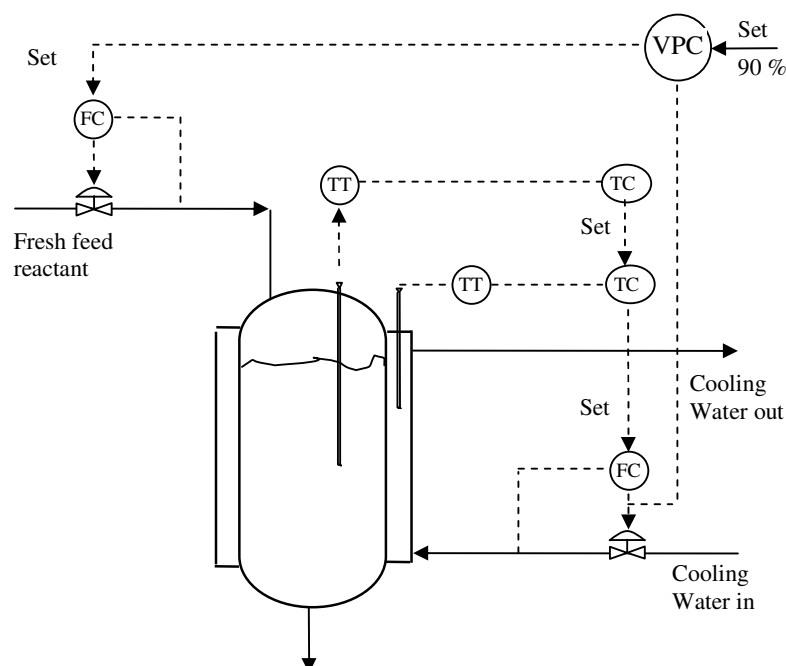


Figure 6.21. Valve positioning control for throughput maximization

Chapter 7. Heat Exchanger Control

Heat exchangers are widely used for heating / cooling process streams to the desired temperature or to change the phase of a stream. The heat exchanger is thus for removing / adding sensible or latent heat. Figure 7.1 shows the schematic of a counter-current shell and tube heat exchanger. The hot stream flows through the tubes and loses its heat to the cold stream flowing through the shell. The heat exchange is driven by the temperature difference between the shell side and the tube side. For a given inlet temperature of the hot and cold streams, the temperature driving force is more for the counter-current flow arrangement. Most exchangers are thus operated with counter-current flow.

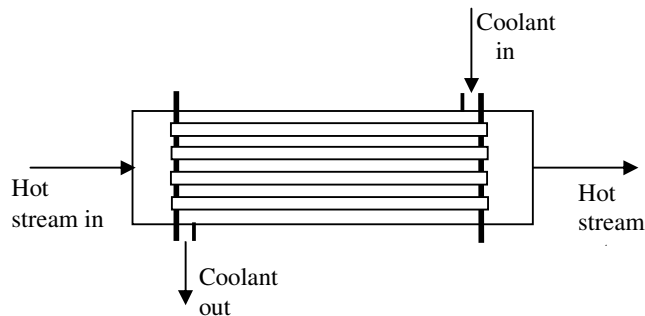


Figure 7.1. Counter current Shell and Tube heat exchanger

The heat exchangers in a process can be usefully classified into utility heat exchangers or process-to-process heat exchangers. Utility heat exchangers typically use steam or cooling water to respectively add or remove heat from a process stream. In process-to-process heat exchangers, both the hot and cold streams are process streams.

7.1. Control of Utility Heat Exchangers

The purpose of a utility heat exchanger is to provide (remove) as much heat as is necessary to maintain the process stream temperature. A simple temperature controller that adjusts the utility flow rate to maintain the process stream temperature accomplishes this function. Figure 7.2(a) shows a cooler that uses cooling water for heat removal. Figure 7.2(b) shows a heater using steam as the utility fluid. The control loops are self-explanatory.

Sometimes the heat transfer is controlled without adjusting the utility flow. For example in a partial condenser that vents the non-condensables as a vent stream, the cooling water valve is fully open and the vent rate is adjusted to control the condenser pressure. The pressure sets the dew point temperature of the condensables in the vapour stream fixing the temperature driving force across the tubes. The control scheme is illustrated in Figure 7.3.

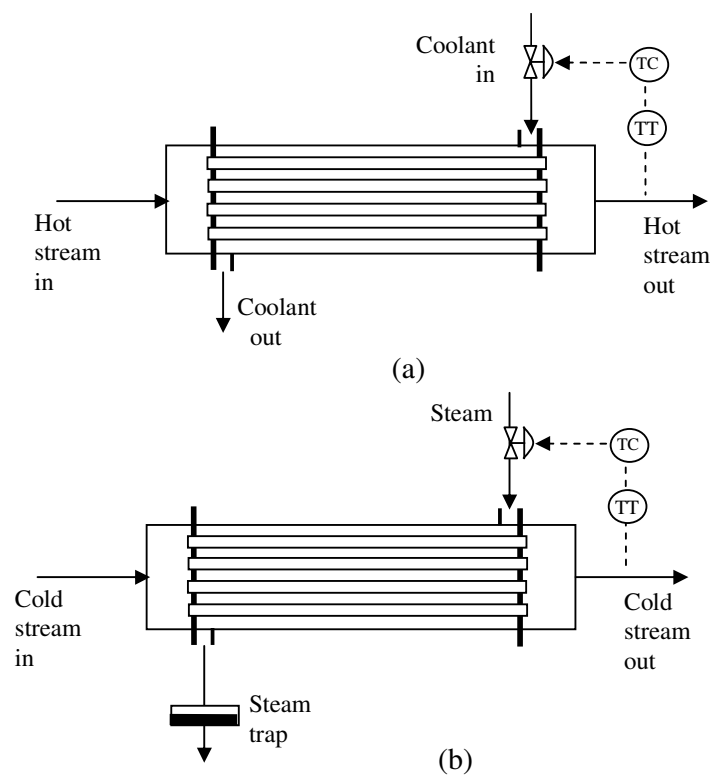


Figure 7.2. Control of utility exchangers

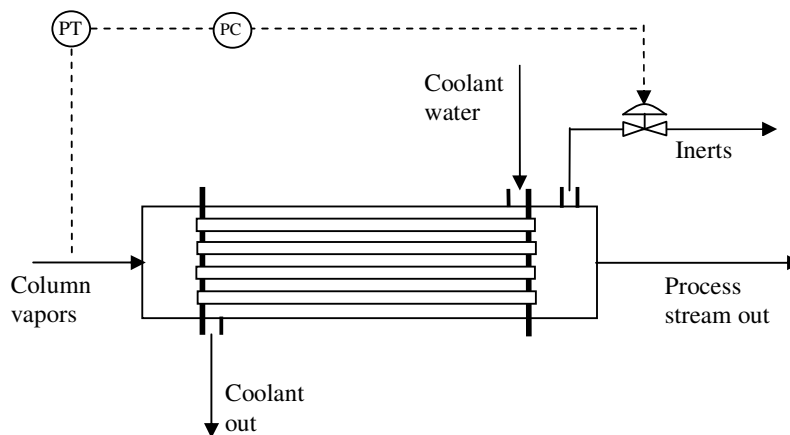


Figure 7.3. Indirect control of Heat exchanger using partial pressure of non-condensable

The flooded condenser is another common arrangement where the level of the condensate determines the number of tubes that are submerged. Heat is thus transferred only across the tubes exposed to the vapour. The cooling rate thus gets adjusted to maintain the pressure by manipulating the condensate draw which affects the level. The liquid hold up inside the condenser represents a significant lag (~2-5 minutes) so that the pressure cannot be controlled very tightly. Flooded condenser arrangement is shown in figure 7.4.

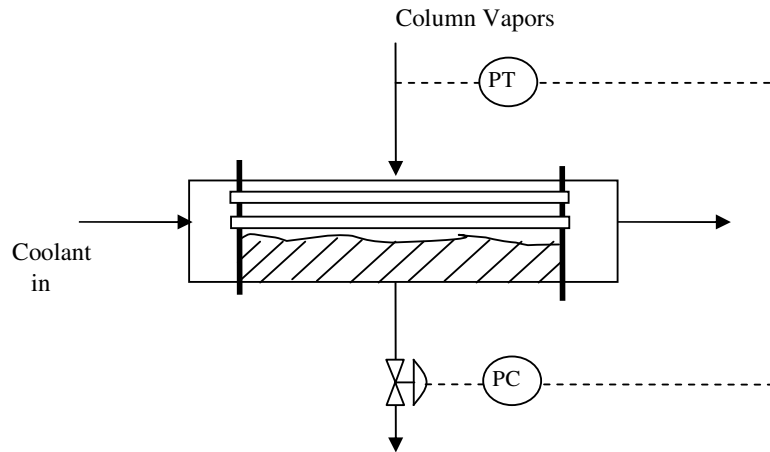
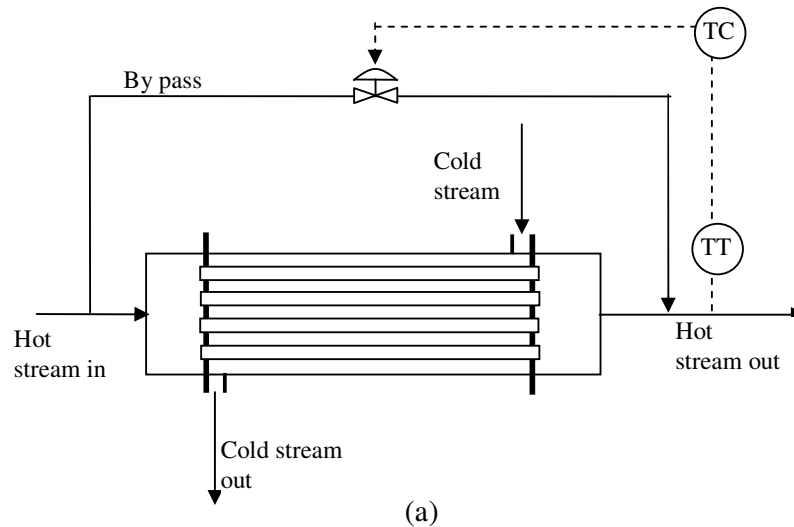


Figure 7.4. Heat transfer control by variable heat transfer area in flooded condenser

7.2. Control of Process-to-Process Heat Exchangers

Process to process heat exchangers transfer heat between two process streams. The flow of these process streams is usually set elsewhere in the plant so that adjusting the flow rate of one of the process streams to regulate the amount of heat transferred is not possible. To provide a control degree-of-freedom for regulating the heat transferred, a small by-pass (~5-10%) of one of the process streams around the heat exchanger is provided. The outlet temperature of this process stream or the other process stream can be controlled by manipulating the by-pass rate. These two schemes are illustrated in Figure 7.5. In the former, tight temperature control is possible as the amount of heat transferred is governed by the bypass. In the latter, a thermal lag of the order of 0.5 to 2 minutes exists between the manipulated and controlled variable.



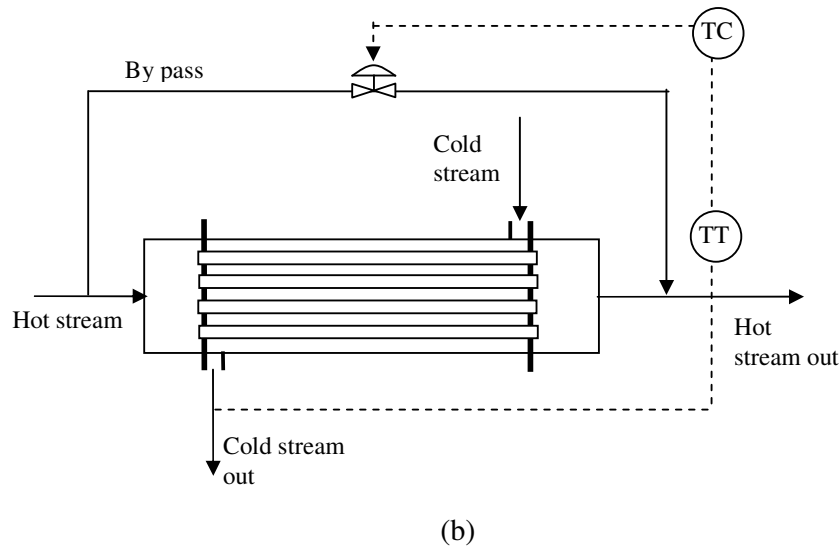


Figure 7.5. By-pass control of process to process heat exchangers
 (a) Controlling and bypassing hot stream
 (b) Controlling cold stream and bypassing hot stream

Process-to-process heat exchangers are increasingly used for heat integration to minimize the energy consumed per kg product. Given that the flow of the two process streams into the heat exchangers is set elsewhere, an auxiliary utility heat exchanger is often provided to control the temperature of the more important process stream. The size of the auxiliary utility heat exchanger should be large enough for effective disturbance rejection.

In the plant-wide context, heat integration using process-to-process heat exchangers causes interaction between the interconnected units. In particular, it is necessary to ensure that in the quest for maximum energy recovery, energy recycle circuits do not lead to instability. Sufficient control degrees-of-freedom should be provided in the form of auxiliary utility exchangers so that the variability is transferred to the plant utility system.

Chapter 8. Control of Miscellaneous Systems

In this chapter, the control of other common units in the industry such as furnaces, compressors, refrigeration systems and plant utility systems is briefly described.

8.1. Furnace Controls

A furnace heats a process stream to high temperature using combustion of a fuel as the heat source. It consists of a fire box or combustion zone with tube bundles carrying the process stream to be heated. Fuel is burnt with air in the combustion zone to heat the tubes to very high temperatures. Typically a convective heat transfer zone is also provided in furnaces to recover heat from the hot flue gases. The furnace is essentially a reactor combusting fuel with air. The control objective is to satisfy the 'on-demand' heat load. The control system shown in Figure 8.1 is typically used. The fuel-to-air ratio must be nearly stoichiometric for complete combustion of the fuel. Excess air is not fed in as that would increase the flue gas discharge rate. Less than stoichiometric air would lead to partial combustion or worse, unburnt fuel remaining in furnace. The flue gas oxygen concentration is a good indicator of the quality of combustion and adjusts the fuel-to-air ratio set-point. The air is fed in ratio to the fuel. The forced draft fan speed is varied to change the air feed rate. An induced draft fan is provided at the outlet to suck the flue gases out of the furnace. Its speed is controlled to maintain the pressure inside the combustion chamber.

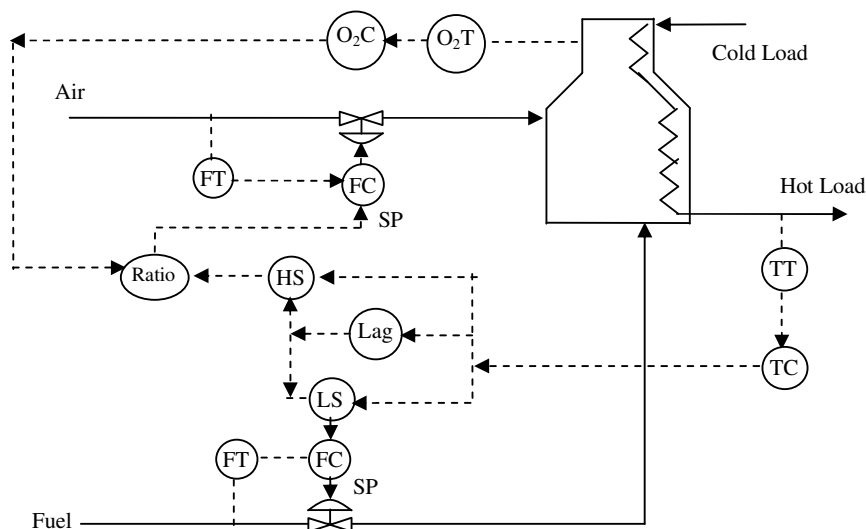


Figure 8.1. Furnace firing controls

A critical safety requirement is to operate the furnace such that the air is fed in excess during transients (load changes). This is necessary to ensure that all the fuel fed into the furnace is burnt and no unreacted fuel remains inside, lest it combust later to damage the furnace. Thus if the heat load increases, the air rate must be increased before the fuel valve is opened. On the other hand, if the heat load decreases, the fuel valve must be closed before the

air flow is reduced. This control action is accomplished by lagging the heat load signal as shown in Figure 8.1. The lagged and the unlagged signals are then input to a high selector and a low selector. The output of the high selector sets the air flow controller set-point. The output of the low selector sets the fuel flow controller set-point. If the heat load increases, the high selector sends the unlagged signal to the air flow controller causing an instantaneous increase in the air flow. The low selector sends the lagged signal to the fuel flow controller. The fuel flow thus lags behind the air flow for an increase in the heat demand. For a decrease in the heat load, the high selector sends the lagged heat load to the air flow controller while the low selector sends the unlagged heat load to the fuel flow controller. The air thus lags behind the fuel for a heat load decrease. Furnace operation in excess air is thus ensured during transients.

8.2. Compressor Controls

Compressors are used to increase the pressure of gas stream. A cooler with a knock-out pot is typically provided at the compressor outlet to cool the hot pressurized gas and remove any condensables that liquefy due to the higher pressure and cooling. There are three important types of compressors used in plants, namely, centrifugal, axial and reciprocating. In reciprocating compressors, the through-put is adjusted by manipulating the strokes per minute or the length of a stroke. A recycle is always provided around the outlet of the compressor for the safety of the compressor.

Centrifugal compressors are similar to centrifugal pumps in that a rotating motor is used to impart energy to the fluid. To control the through-put, three configurations are typically used, namely, exit recycle, suction throttling and motor speed manipulation. These three schemes are illustrated in Figure 8.2. In the exit recycle scheme, a recycle around the compressor back to the inlet is provided which is adjusted to manipulate the through-put. Note that the total (recycle + fresh) flow rate through the compressor remains the same so that compressor operates at a single point on its characteristic curve. This is the most energy inefficient method of compressor operation. Also, note from the figure that the recycle is provided after the cooler so that energy recycle is prevented. In suction throttling, a valve is provided at the compressor suction and the through-put is manipulated by adjusting the suction pressure. At lower through-puts lesser energy is consumed as the amount of material flowing through the compressor is less. The most energy efficient method of throughput manipulation is to vary the rpm of a variable speed drive. High pressure steam, as in a turbine, is used many a times to provide the motive force for rotation. A cascade speed controller that adjusts the steam flow rate set-point maintains the drive speed. The drive speed set-point is input remotely by the through-put controller.

Another important consideration in compressor control is the prevention of surge at low flow rates. The compressor characteristic curve shows a maximum and the compression ratio dips at low flow rates due to compressibility. So much so, that if the flow rate goes low enough, the flow through the compressor can reverse direction. This causes the suction pressure to build and the flow almost immediately reverses direction again (i.e. flows out the compressor). This flow reversal cycle repeats in less than a second. To prevent the compressor surge phenomenon, the compressor discharge is recycled to the compressor suction. An anti-surge controller, as in Figure 8.3 adjusts the recycle rate to prevent the flow through the compressor from dropping below a minimum. Note that this minimum must be sufficiently above the surge flow rate for the particular compressor rpm (or maximum rpm for variable speed drives).

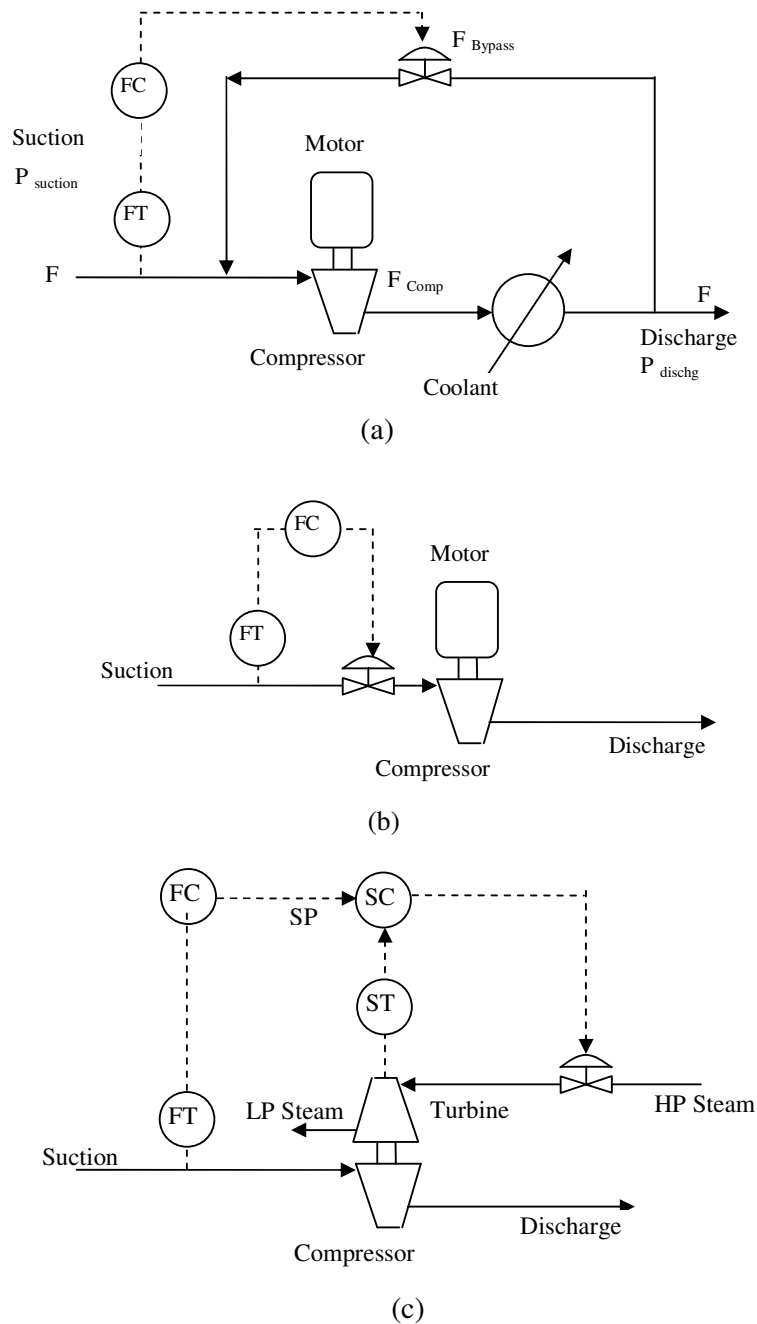


Figure 8.2. Compressor controls
 (a) Exit recycle
 (b) Suction throttling
 (c) Motor speed manipulation

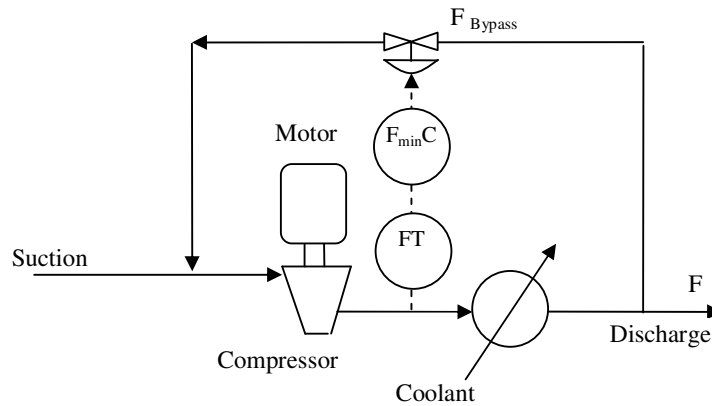


Figure 8.3. Compressor antisurge controller

8.3. Decanter Control

Decanters separate a heterogeneous liquid-liquid mixture into its constituent liquid phases by utilizing the density difference between the liquid phases. The heterogeneous mixture phase separates into a heavy and light liquid phase. Typically, the heavy liquid phase is aqueous while the light liquid phase is organic. Appropriately located withdrawal ports are provided in a decanter for removing aqueous and organic streams. To prevent the aqueous liquid from entering the organic liquid withdrawal port, the level of the liquid-liquid interface must be controlled. Also the organic phase level must be controlled. The simplest scheme, shown in Figure 8.4(a) manipulates the organic and aqueous stream flow rates to adjust the respective levels. The organic level controller must however interact with the aqueous level controller. A simple and effective strategy for removing the interaction is to adjust the total flow out from the decanter to control the organic phase level. Figure 8.4(b) shows the corresponding control scheme. The organic stream flow is manipulated to maintain total flow (organic + aqueous) out of the decanter. The organic level controller sets the total flow set-point.

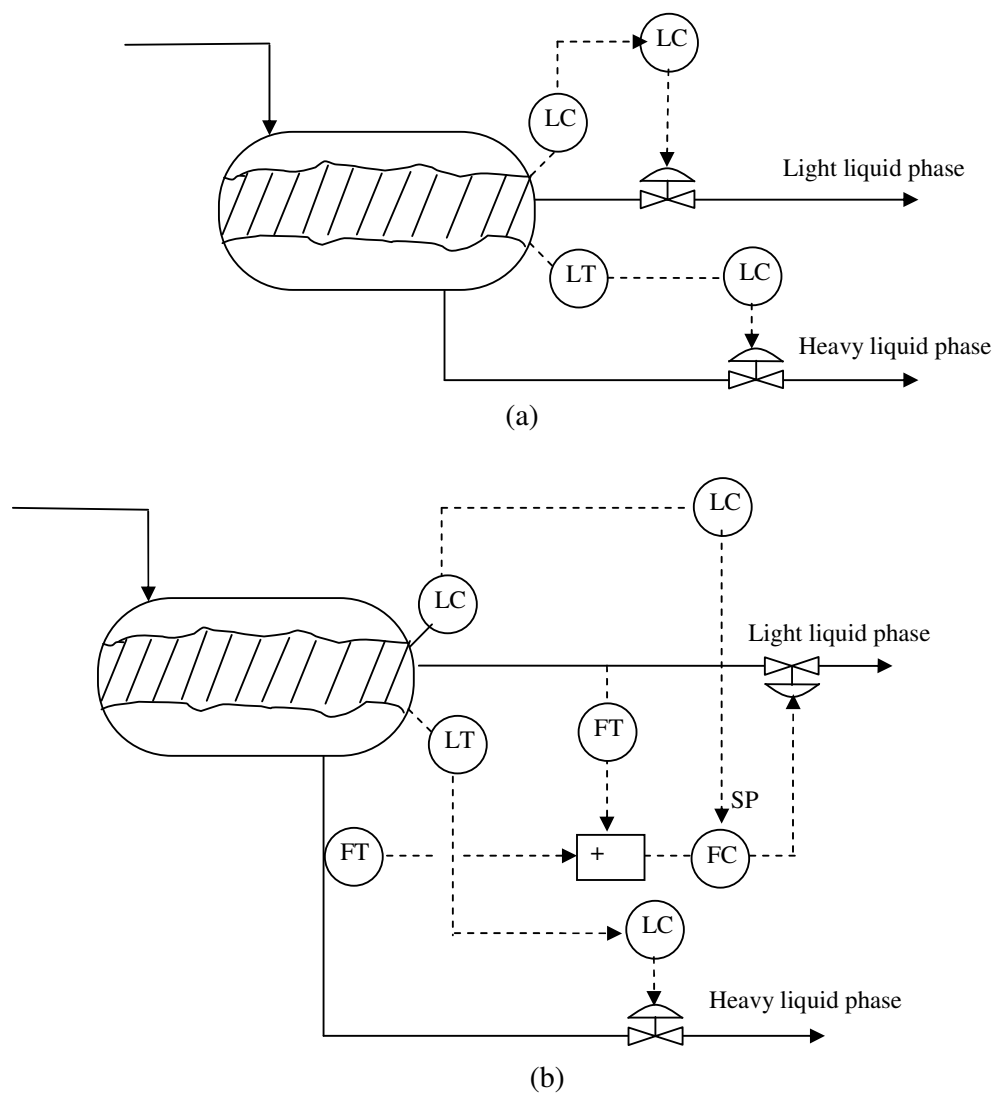


Figure 8.4. Decanter Controls
 (a) Conventional level control
 (b) Buckley control structure to eliminate interaction

8.4. Control of Refrigeration Systems

We study control schemes for the commonly used vapor compression and vapour absorption refrigeration cycle.

8.4.1. Vapor Compression Cycle

The refrigeration cycle typically employs compression. The cold refrigerant absorbs heat from the process stream and vaporizes in the evaporator. The vapour is compressed so that at the higher pressure, cooling water can be used to condense the refrigerant. The condensed refrigerant is collected in a surge drum and fed to evaporator. Figure 8.5 shows control schemes for the compression refrigeration cycle. The chilled process stream temperature controller sets the evaporator pressure set-point. The evaporator pressure is controlled by adjusting the compressor suction valve. The level in the evaporator is controlled by adjusting the liquid refrigerant inlet valve. In case a variable speed drive compressor is used, the pressure controller is done away with and the temperature controller directly sets the drive speed set-point. The pressure controller is necessitated in the compressor suction throttling scheme to compensate for the throttling valve non-linearity. In the variable drive speed compressor, the variation in the suction pressure (evaporator pressure) with respect to the drive speed is relatively linear so the drive speed can be directly adjusted by the temperature controller. The level in the refrigerant surge drum is not controlled as the refrigerant forms a closed circuit. Notice that the heat transfer rate changes as the temperature driving force across the condenser changes due to the dependence of refrigerant boiling temperature on the evaporator pressure.

8.4.2. Vapor Absorption Cycle

In addition to compression systems, refrigerant absorption systems are also applied industrially. The absorption based refrigeration cycle and its control scheme is shown in Figure 8.6. Ammonia (refrigerant) rich strong liquor is distilled at high pressure to recover liquid ammonia as the distillate and ammonia lean weak liquor as the bottoms. The liquid ammonia is fed to the evaporator where it absorbs heat from the process stream to be chilled and evaporates. Vapor ammonia is absorbed by the 'weak liquor' water stream. The 'strong liquor' so formed is fed to the distillation column to complete the closed circuit refrigerant loop. The temperature of the chilled process stream is controlled by adjusting the level set-point of the evaporator. The heat transfer rate is thus varied by changing the area across which heat transfer occurs. The evaporator level controller adjusts the distillate liquid ammonia flow. An increase in the level of the evaporator implies an increase in the ammonia evaporation rate so that the weak liquor rate is increased in ratio to absorb the ammonia vapours. The strong liquor is cooled and collected in a surge drum. The level of the surge drum is not controlled. Liquid from the surge drum is pumped back to the distillation column through a process-to-process heater that recovers heat from the hot 'weak liquor' bottoms from the distillation column. The flow rate of the strong liquor to the column is adjusted to maintain the column bottoms level. Also, the steam to the reboiler is manipulated to maintain a tray temperature.

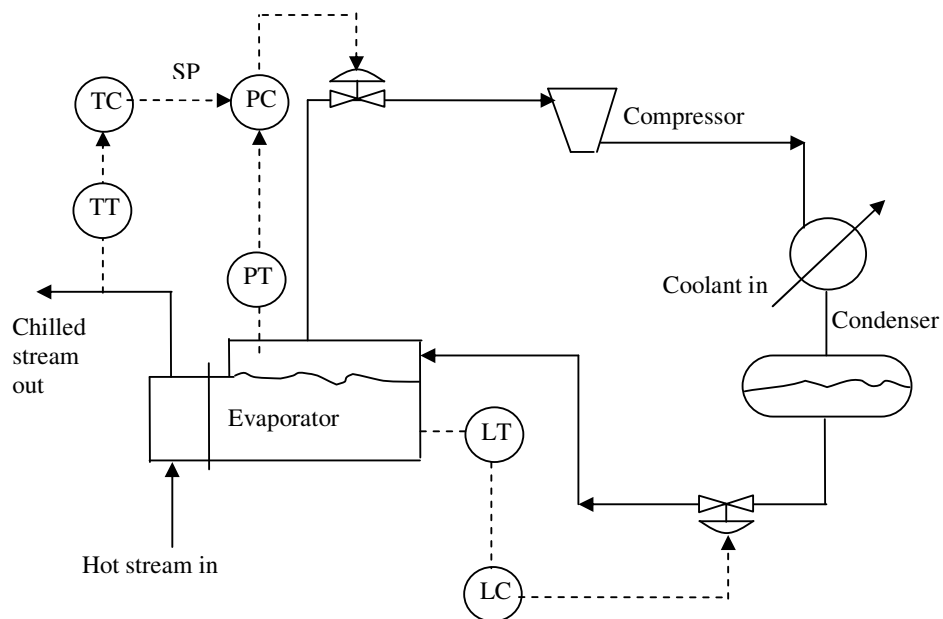


Figure 8.5. Compression refrigeration controls

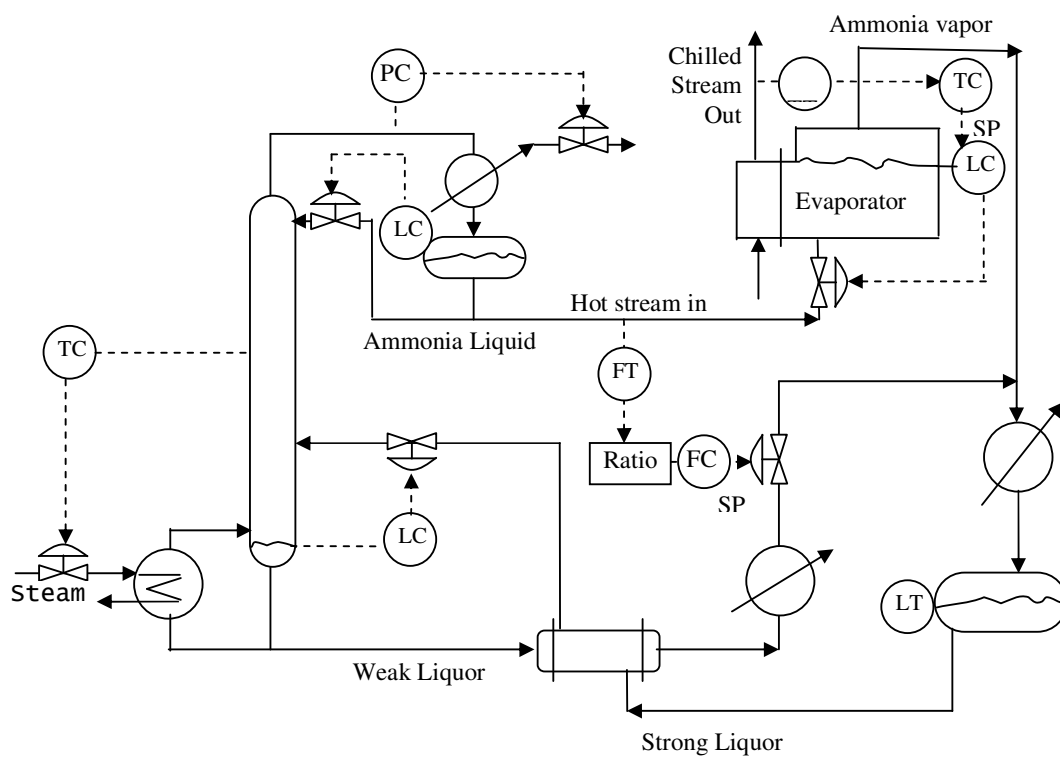


Figure 8.6. Absorption refrigeration controls

8.5. Control of Steam Utility System

Figure 8.7 schematically shows a plant power and utility system. Boiler feed water is heated in a furnace to produce saturated steam. The saturated steam is passed through the furnace to produce superheated steam at 1000 psia pressure. The superheated LP steam drives a turbine to produce electricity. Steam at different pressures is extracted from the turbine for process steam utility requirements. Typically, steam at 300 psia (high pressure steam), 150 psia (medium pressure steam) and 50 psia (low pressure steam) is made available as a heat source at different temperature levels for process use. The pressure of the superheated steam from the furnace is maintained by adjusting the furnace duty. The steam drum level is controlled by adjusting the boiler feed water rate. The pressure of the 300 psia header is maintained in a split range arrangement as shown in the Figure. Steam from the higher pressure header is let in for a decrease in the header pressure while steam is dumped to the lower pressure header for an increase in the header pressure.

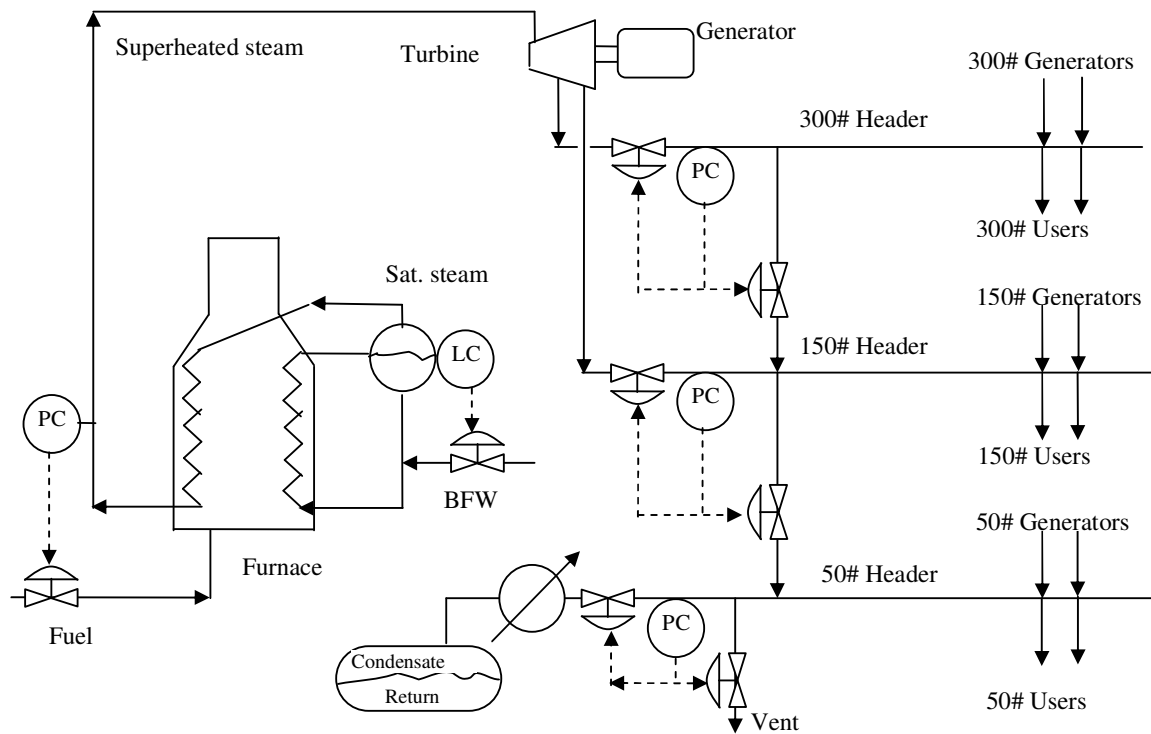


Figure 8.7. Power plant utility system controls