CHAPTER II

Literature Review

The concept of plantwide control structure synthesis is not new to the chemical process industry. The first documented plantwide control approach was suggested by Buckley in 1964. In the 1980s the research in this area grew, in part because of the number of heat management systems that were being installed. However, compared to the number of methods available for the synthesis of control structures for individual unit operations, the number of plantwide control synthesis methods has remained modest. This sparsity reflects the complexity inherent in the plantwide control problem, which includes a multitude of objectives, conflicting objectives, interactions, large dimensionality, and non-linearity. The research studies that have appeared in the open literature for plantwide control can be broadly classified into three categories.

A large number of contributions on plantwide control structure synthesis can be classified into the first category, process-based experience and engineering judgment. One of the most notable contributors is Luyben and co-workers.

A recycle behavior in process is studied both open loop and closed loop. The result is recycle stream has an impact on the dynamics of the process and the overall time constant can be much different than the sum of the individual unit. Moreover, the behavior of a recycle system depends strongly on the recycle loop gain and somewhat less strongly on the dynamics of the individual units in the recycle loop. (Luyben, W. L., 1993)

Price, R., and Georgakis, C. (1993) presented a five-tiered approach based on inventory (material balance closure), production, and quality control. The control design development is done in stages that correspond to the goals and tasks of the proposed control system. The five tiers: production rate; inventory maintenance; product specification; equipment and operating constraints; and economic performance. There are differences between this method and that proposed by Luyben et al.

Randel M. Price and co-worker (1994) have proposed guidelines for the development of production rate and inventory controls within the plant. These rule help a designer to ensure that these parts of the control structure are effectively designed before proceeding to the designs of product quality control. The guidelines is obtained by using process internal streams as throughput manipulator and illustrated on a CSTR/column system and the Eastman process. Both examples provide insight into the problem of throughput manipulator selection and control of recycle systems. They agreed in that both performed best when an internal throughput manipulator was used.

Luyben, W. L. (1994) presents a mathematical analysis of the problem for several typical kinetic systems. In the simple binary first-order case of $A \rightarrow B$, an analytical solution can be found for the recycle flow rate as a function of the fresh feed flow rate and fresh feed composition. Two different control structures are explored. It is shown analytically why the control structure proposed by Luyben prevents snowballing and why the conventional structure results in severe snowballing. Two other kinetic systems are studied numerically: consecutive first-order reactions $A \rightarrow B \rightarrow C$ and a second-order reaction $A+B \rightarrow C$. Results confirm that snowball problems can be prevented by using a control structure that fixes the flow rate of one stream somewhere in a liquid recycle loop. In processes with one recycle, the flow rate of the reactor effluent can be set. In processes with two or more recycle streams, the flowrate of each recycle can be fixed.

The steady state disturbance sensitivity analysis method is presented. It provide a preliminary screening of candidate plantwide control structures in order to eliminate some poor structures and it required only steady state information. This method is illustrated on a complex process with one reactor, three distillation columns, and two recycle streams. A number of alternative control structures were screened and eliminated some unworkable control structure. The effectiveness of the remaining structure is demonstrated by dynamic simulation. Some control structures are found to have multiple steady states and produce closed loop instability. (Chang K., and Luyben, W. L., 1995)

Michael L. Luyben and Bjorn D. Tyreus (1996) analyzed the effect of the process design on control structure for a system with a reactor, two distillation columns, and two recycle streams. The reaction A + B→ C occurs in a reactor. They present an analysis that explains the fundamental problem with control structures in which one fresh feed is fixed and no reactor composition is measured. They show that this control structure can work if modifications are made in the design from the steady-state economic optimum. This highlights the potential trade-off between steady-state economics and dynamic controllability and illustrates considerations that ought to be included during the conceptual design procedure. A modified control structure is proposed that provides effective control of the economically optimal process design. It permits throughput to be directly set and does not require a composition measurement. The basic idea is to use the flow rates of the recycles from the separation section to infer reactor compositions. Dynamic simulation studies on both simplified and rigorous models are used to evaluate the performance of the proposed control system over a wide range of reactor sizes.

Philip R. Lyman and William L. Luyben (1996) studies of a ternary process with the reaction A + B → C and two recycle streams have revealed some interesting dynamic problems in transitioning from one production rate to another, even though the new production rate is attainable from a steady-state point of view. If the process has been designed with a small reactor volume, which means large recycle flow rates, the process sometimes shuts itself down when the production rate is increased. The control structure used has fixed reflux flow rates on both columns. The basic problem is a buildup of impurity (component C) in one or both of the recycle streams. This can be prevented by modifying the control structure to use reflux-to-feed ratios for processes with intermediate reactor holdups or to use dual composition control on both columns for processes with low reactor holdups.

The process that consists of a reactor, two distillation columns and two recycle streams are considered with a generic methodology call the capacity-based economic approach. The method combines both steady state economics and process controllability and may be used to screen alternative plant designs on their ability to minimize costs and process variability. The result can infer that the benefit from designing a more controllable process are much more significant than designing a better plantwide control structure and closed loop process dynamics should be studied prior to build the plant.(Elliott, T. R., and Luyben, W. L.,1996)

An easy approach to determine the design degrees of freedom has demonstrated. For a certain class of processes, the number of control valves in the process just be count. The design and the control degrees of freedom are same in number, but different parameters are used to satisfy them. For a broader class of processes, the procedure must be modified to account for nonreactive liquid levels, column sections, and gas phase reactors. Several case studies of increasing complexity have been presented to show the effects of various types of flowsheets and physical properties. The complexity of the phase equilibrium and the physical properties does not affect the degrees of freedom. The structure of the flowsheet is what determines the design degrees of freedom. The practical significance of this approach is that a model is not need and can be avoid the tedious and error prone procedure of accounting for all variables and equations. (Luyben, W. L., 1996)

The presence of recycle streams and energy integration in chemical process creates unique features for plantwide control because of the potential for disturbance propagation and the alteration of the system's dynamic behavior. Therefore, a general heuristic design procedure is presented that generate an effective plantwide control structure for an entire complex process and not simply individual units, The nine steps of the proposed procedure: energy management; production rate; product quality, operational, environment and safety constraints; liquid level and gas pressure inventories; makeup of reactants; component balances; and economic or process optimization. Application of the procedure is illustrated with three industrial process example: the Vinyl acetate monomer process, Eastman process, and HDA process. (Luyben, W. L., Tyreus, B. D., and Luyben, M. L.,1997)

Design details of an industrial process for the manufacture of Vinyl acetate monomer is presented. The process contains many standard unit operations that are typical of chemical plants and it has both gas and liquid recycle streams with real components. The details are about the flowsheet information that required constructing rigorous steady state and dynamic mathematical model of this process, and the process control requirements and objectives. Including the rigorous nonlinear dynamic simulation is constructed for this process by using TMODS program. (Luyben, W. L., and Tyreus, B. D. 1997)

There are two basic control structures for chemical plants: fixed feed and fixed products. Luyben point out that the fixed product, which called on-demand structure, has several inherent dynamics disadvantages compared to the more conventional approach of setting the feed streams to a process. They are illustrated on two processes of increasing complexity. The first is a binary system with the reaction $A \rightarrow B$ and a plant topology of one rector, one stripping column, and one recycle stream. The second is a ternary system with the reaction, $A + B \rightarrow C$ and a flowsheet containing one reactor, two distillation columns, and two recycle streams. Dynamic simulations demonstrate that the on-demand structure introduces larger disturbances into the system, which results in more variability in product quality. (Luyben, W. L. 1999)

Tyreus, B. D. (1999) studied thermodynamically motivated method for the identification of dominant variables used in partial control. The method is applied to the Tennessee Eastman challenge process and shows that the reactor temperature together with the vapor-phase composition of component A are dominant variables. It is also shown that the temperature in the separator is dominant for this unit and that the overhead vapor composition is a dominant variable for the stripper. When a complete partial control structure is formed by feedback control of all the dominant variables, it turns out that an insufficient number of manipulated variables remain to satisfy inventory and component balances. A reduced partial control structure is thus suggested. It is demonstrated that this design is superior to other control solutions to the Eastman process in that it can easily attain and hold the plant at its maximum production rate.

Luyben, W. L. (2000) studied diabatic tubular reactors. The conventional control scheme for these reactors is to maintain the inlet temperature to the reactor. However, the optimum operation of the entire process often requires that the reactor be run at the highest possible temperature, which occurs under steady-state conditions at the reactor exit if the reactions are exothermic. The process studied in this paper has the exothermic, irreversible, gas-phase reaction $A+B \rightarrow C$ occurring in an adiabatic tubular reactor. A gas recycle returns unconverted reactants from the separation section. Four alternative plantwide control structures for achieving reactor exit temperature control are explored and compared. Manipulation of reactor inlet temperature appears to be the least attractive scheme. Manipulation of recycle flowrate gives the best control but may be undesirable in some system because of compressor limitations. The on-demand structure provides effective control in the face of feed composition disturbance.

Groenendijk et al. (2000) proposed a systematic approach that involves the combination of steady state and dynamic simulations. Several controllability measures (relative gain array, singular value decomposition, closed loop disturbance gain, etc.) are employed to develop the final control structure and to assess its performance. The systems approach is illustrated with a Vinyl chloride monomer (VCM) plant.

Shinnar et al. (2000) introduced the concept of partial control, the identification of a dominant subset of variables to be controlled such that, by controlling only these variables, a stabilizing effect on the entire system results. The methodology to find the dominant partial control set relies predominantly on process experience. The approach was demonstrated on a fluid catalytic cracker unit, not on an entire plant.

Sheng-Feng Chiang et al. (2000) studied the acidic cation-exchange resin catalyzed amyl acetate process. Two design alternatives, coupled reactor/distillation column and reactive distillation column, were evaluated. The phase equilibrium of the quaternary system revealed that a significant two-liquid region exists. Therefore, a decanter was placed on top of the column, where high-purity water was withdrawn from the aqueous phase, and the organic phase was totally refluxed. Systematic design procedures were proposed for these two alternatives. Results, in terms of the total annual cost, indicated that reactive distillation is 4 times more efficient than the coupled reactor/separator for the amyl acetate process. The dynamics of reactive distillation with a decanter is much more complicated than expected. Limit cycles can occur if the control structure is not properly designed. The imbalance in feed must be taken into account in any realistic process operation. This often leads to at least two composition loops (or two inferential variables). Despite the complex dynamics simulation results show that reasonable control can be achieved with the proposed control system.

Francisco Reyes (2001) studied adiabatic tubular reactor systems with liquid recycles and distillation columns used in the separation section. Irreversible and reversible reaction cases have been explored. Both steady-state economics and dynamic controllability have been considered in the designs. For the numerical case studied, which is typical of many real chemical systems, the liquid recycle system is more expensive because of the high cost of the distillation column and the need to vaporize the recycle. The liquid recycle process is also more difficult to control because the large holdup in the recycle loop produces slow composition changes. For irreversible reactions, the activation energy is shown to slightly affect the steady-state design but to drastically impact the dynamic controllability. Steady state economic designs are shown to be very difficult to control because of the severe temperature sensitivity with high activation energies. Changes in the design conditions and changes in the control structure can be used to produce a more easily controlled process. For reversible reactions, the steady-state design is more difficult because of the additional degrees of freedom, but the dynamic controllability is much better because of the inherent self-regulation of exothermic reversible reactions as they encounter chemical equilibrium constraints.

Costin S. Bildea (2002) analyzed the non-linear behavior of several recycle systems involving first-and second- order reactions. The results, presented in term of dimensionless numbers, explain some control difficulties. It is shown that conventional control structures, fixing the flow rate of fresh reactants and relying on self-regulation, can lead to parametric sensitivity, unfeasibility, state multiplicity, or instability, particularly at low conversions. These problems can be solved by fixing the flowrate in the recycle loop, as stated by Luyben's rule. They was demonstrated that a particular location for fixing the recycle flow rate is advantageous, i.e. the reactor inlet. This decouples the reactor from the rest of the plant and avoids undesired phenomena due to mass recycles. For example, the unstable closed-loop behavior observed with non-isothermal PFRs disappears. The HAD plant case study illustrates the proposed strategy.

A new flowsheet development methodology for synthesizing plantwide control structures has been presented by Vasbinder and Hoo (2003). The method is based on a modified version of the decision-making methodology of the analytic hierarchical process (AHP). The decomposition utilizes a series of steps to select among a set of competing modules. The control structure for each of the individual modules was developed using Luyben's nine steps approach. The decomposition serves to make the plantwide control problem tractable by reducing the size of the problem, while the mAHP guarantees consistency. The modular decomposition approach was applied to the dimethyl ether (DME) process, and the results were compared to a traditional plantwide design approach. Both methods produced the same control structure that was shown to be adequate for the process. Satisfactory disturbance rejection was demonstrated on the integrated flowsheet.

Kapilakarn, K. and Luyben, W. L. present the plant wide control of continuous process multiple products. There are two reversible reactions producing two products. The control structure must be able to achieve different production rates of the two products. Several conventional control structures are studied in which the flow rates of the fresh feed streams are fixed or manipulated by level or composition controllers and the production rates of the two products are not directly set. They present several "on demand" control structures which both product streams are flow controlled. The control system must adjust the conditions in the plant and the fresh feed streams to achieve the desired product flow rates. The most effective on-demand control structure require no reactor composition analyzer and no recycle of product streams.

The second category is to use the hierarchical design approach to develop the control structure. A large contribution in this area was provided by Douglas and co-workers. The plantwide control structure is developed from a simple to a detailed description. Design changes are recommended or another alternative is developed whenever the control and manipulated variables are not balanced. A drawback to this approach is that the entire analysis is based on steady state objectives and models. As Luyben has indicated, assessing performance on the basis of a dynamic model of the plant is equal to the validation of any proposed control structure. A measure of controllability called the "controllability index" was also developed. The basis of this measure is a determination of the minimum amount of addition storage necessary for the plant to optimal control performance.

Alex Zheng et al. (1999) presented a hierarchical procedure for systematically synthesizing a plantwide control system and its basic ideas have been illustrated on the simple reactor-separator-recycle system. The procedure decomposes the problem into a hierarchy of decisions six steps. As one proceeds down the hierarchy, modeling details are added to the flowsheet and alternative are generated. The idea is not to explore every alternative, but to quickly eliminate poor alternatives from further consideration. For an existing plant with an existing plantwide control system, one can move up the hierarchy and compare the economic incentive for redesigning the plantwide control system.

The integration of the process design with plantwide control system design is a concern in the process chemical industries. The advantage of the hierarchical approach is that it takes the idea of concurrent design of the control and process into consideration. Unfortunately, some limitations arise from the fact that control structure synthesis, by necessity, is a dynamic process and the design process is executed only for the steady state conditions. Therefore, it is hard to evaluate the quality of any integrated system until the final flowsheet design is completed.

The third category relies on a rigorous mathematical framework of dynamic theory, constrained optimization, and system analysis. The inspiration is to address the lack of process experience on the part of the learner engineer and to consider more than the regulation of the process.

Arkun and Downs (1990) presented an approach to calculating steady state gain matrices which include integrating variables. This approach one determines the rate of change of the integrating variables. In addition to using the resulting gain matrix for operability analysis, one can also use the matrix for plantwide control system design. An advantage of using a gain matrix that includes integrating variables is that only one matrix is required for a process. This matrix contains all of the information required to assess the steady state operability of a process. However, in this approach a dynamic state space model is assumed to be available and in many cases one does not have such a dynamic model.

Down and Vogel (1993) studied a model of and industrial process for developing, studying, and evaluating process technology. This process consists of a reactor/separator/recycle arrangement involving two simultaneous gas-liquid exothermic reactions. It has 12 valves available for manipulation and 41 measurements available for monitoring or control. This process model is studied both plantwide control and multivariable control problem and has been coded into a set of FORTAN subroutines.

A rigorous design procedure was developed for the complex reactor/stripper with multicolumns and recycle. This study provides a good example for the integration of steady-state design and control design. Effects of various optimization parameters on both design and control were analyzed at the design stage. The flow rate of the recycle stream was found to be one of the major optimization parameters that strongly affected the steady-state economics. The effects of impurities in the recycle stream on economics were minor, however, the effect of impurities on controllability was not negligible. A large number of alternative control structures were explored. One control structure showed multiple closed loop steady-state solutions. An effective control structure was developed and tested for a variety of disturbances. (Chang K., and Luyben, W. L., 1995)

McAvoy, Ye and Gang (1996) presented the approach that using nonlinear inferential parallel cascade control(NIPCC) to improve control system performance. NIPCC is a feedback approach that detects unmeasured disturbance through inferential measurements. It then compensates for these disturbances in a manner that is similar to feedforward control. The effectiveness of NIPCC has been demonstrated on the Tennessee Eastman test process, under two different base control systems. It has been shown that for random feed composition fluctuations the variance in product flow and composition can be reduced significantly by using NIPCC, for all of the process operation modes. The NIPCC control approach is a promising method for improving the performance of existing plantwide control system control system, particularly when a steady state model is available.

McAvoy and Miller (1999) developed the gain matrix as in the Arkun and Downs approach. This approach involves introducing artificial variables that affect the integration variables and that leave the process in the same way as product streams. The artificial variables naturally occurring product streams are used for level control, and this allows the gains for other variables to be calculated. Dynamic models for the integrating variables are also required. Once a steady state gain matrix is calculated, it can be used to assess the operability of plant wide control schemes. The methodology is developed using a distillation tower system, and then it is applied to analyze the "snowball effect" in a three reactor / three distillation tower plant. The methodology is also applied to the Tennessee Eastman process where level dynamics are more difficult to model and accuracy issues arise.

Wang and McAvoy (2001) discussed an optimization-based approach to synthesizing plantwide control architectures. This approach represents an interesting attempt to combine the heuristic and optimization-based methods. The synthesis is split into three stages involving fast and slow safety variables to be controlled, followed by product variables. In each stage a mixed integer linear program is solved to generate candidate architectures. After the safety control system is designed, PI controller for the incorporated into the dynamic model for use in synthesizing the product. The objective function involves a trade off between manipulated variable moves and transient response area. The integer in the formulation determines the control architecture, and integer cuts are employed to generate candidate solutions. The Tennessee Eastman process is used to illustrate the synthesis procedure.

Rong Chen et al. (2003) present the first principles nonlinear dynamic model and simulation of Vinyl Acetate process. The process model is large and in part highly nonlinear and it contains 246 states, 26 manipulated variables, and 43 measurements. The model of Vinyl Acetate process is developed in MATLAB and both the steady state and dynamic behavior of the MATLAB model are designed to be close to the behavior of TMODS model that is an earlier model of the same process was published. Followed shortly, they present a new approach to the design of plantwide control systems of Vinyl acetate process. The approach is based on output optimal control and it assumes that a linear dynamic process model is available. The measurements are determined according to the process gain matrix, and eigen value analysis, and engineering judgment for eliminating and evaluating candidate architectures. The design of plantwide architecture is split into four stages. Four decentralized plantwide designs are generated, and these designs are very closed to one earlier design by Luyben and co-workers.

Larsson et al. (2003) present control structure selection for a simple plant with a liquid-phase reactor, a distillation column, and recycle of unreacted reactants. The starting point is a clear definition of the operational objectives, constraints, and degrees of freedom. Active constraints should be controlled to optimize the economic performance. This implies for this case study that the reactor level should be kept at its maximum, that being economically attractive. Maximizing the reactor holdup also minimizes the "snowball effect". The main focus is on the selection of a suitable controlled variable for the remaining unconstrained degree of freedom, that use the concept of self-optimizing control to search for a constant setpoint strategy with an acceptable economic loss. Both for the case with a given feed rate where the energy costs should be minimized and for the case where the production rate should be maximized, they find that a good controlled variable is the reflux ratio *L/F*. This applies to single-loop control as well as multivariable model predictive control.