



Review

Reaction engineering: Status and future challenges

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ABSTRACT

This manuscript summarizes the plenary lecture delivered at the ISCRE 20 meeting in Kyoto on Tuesday, September 9, 2008. The scope, history and status of our chemical reaction engineering (CRE) discipline are briefly outlined complementing the broad review presented by Professor J. Schouten in a plenary lecture a day earlier (Schouten, 2008). It is argued here that the key challenge for CRE is the development of new more efficient and profitable technologies. This is to be accomplished via an *improved science-based scale-up methodology for transfer of molecular discoveries to sustainable non-polluting processes that can meet the future energy, environmental, food and materials needs of the world*. Available foundations for such scale-up are introduced and the role of science in developing the improved methodology for multiphase systems is described. It is also argued that progress in implementing the desired environmentally benign processes depends, in addition to overcoming technical challenges, on achieving changes in the socio-economic and political arena. Global regulations and peer pressures are needed to provide the economic incentives for cleaner and sustainable processes.

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Contents

1. Introduction	3
2. Chemical reaction engineering methodology (status)	4
3. Reactor scale-up (current status)	6
4. Multi phase reactor selection, modeling and scale-up	8
5. Summary and conclusions	10
Acknowledgments	10
References	10

1. Introduction

It is important to recognize the major role of chemical reaction engineering (CRE) in providing the means for sustaining the life styles in modern technologically developed societies. It is also important to understand the basis for the CRE methodology that is used so successfully in a myriad of industries. The domain of chemical reaction engineering consists of all chemical transformations (and that includes biological) of starting materials, derived from non-renewable and renewable resources, into a variety of products for the market on which we depend to support our life style (see Fig. 1). The key to economically, environmentally friendly and energy efficient processes is in choosing the right

chemical transformation, the right catalyst for it, and the right reactor type, and in being able to scale-up these transformations to commercial production while maintaining high material and energy efficiencies. Those of us who are educators emphasize to students that while science provides us with the understanding needed to engineer the required transformations only profitable processes are commercialized. Past and present practices suggest that when there is a conflict between profitability and environmental friendliness it is improved profitability that always wins. The challenge for CRE is not just to make environmentally benign processes more profitable than the alternatives based on conceptual analysis, but to provide the scientifically based methodology that will significantly reduce the risk of their implementation. It seems thus that we need progress on political as well as technical fronts (Huesemann, 2003). Progressive tax structures that reward environmental friendliness and sustainability, and global regulations or peer pressures that demand certain standards of environmentally benign performance would provide the improper incentives for consideration of more

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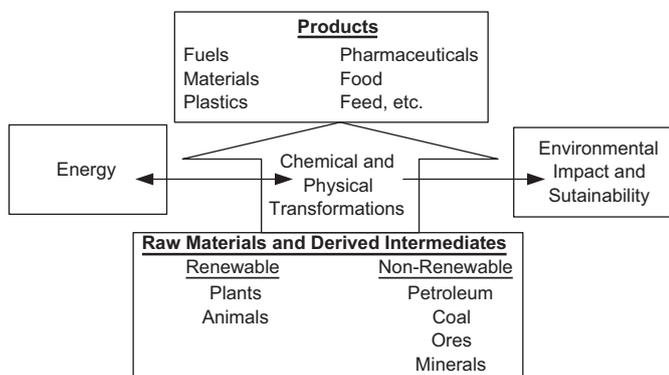


Fig. 1. Schematic of reaction engineering activities and their impact.

efficient and sustainable technologies. Advances in the scientific base of scale-up would reduce the risk of their implementation. We must develop a sound scientific base to meet the world's need for manufactured goods while reducing the environmental impact of our activities. Only with acceptance of environmentally friendly alternatives as profitable and less risky will the Kyoto protocol and other environmental standards be universally accepted.

It is clear from Fig. 1 that our activities have a direct impact on the world energy challenge. We are involved in producing energy carriers, and the choice of starting materials we use for this purpose has a great impact on the environment and sustainability. The challenge of identifying and exploiting future sources for clean energy, however, is a separate very complex topic. Instead we focus here only on the general issues of energy and materials efficiencies of our reaction systems in the process industries and their effect on the environment.

$$\left(\begin{array}{l} \text{Global damage} \\ \text{to the environment} \end{array} \right) = \left(\begin{array}{l} \text{Overall process} \\ \text{inefficiency} \end{array} \right) \times \left(\begin{array}{l} \text{Total} \\ \text{population} \end{array} \right) \times \left(\begin{array}{l} \text{Consumption} \\ \text{per capita} \end{array} \right) \quad (1)$$

There is a wide spread agreement that the global damage to the environment (total pollution) can be represented as a product of consumption per capita, population and overall process inefficiency, as indicated by Eq. (1) above. Accordingly, global damage to the environment (i.e. global pollution) can be reduced effectively by either controlling population, which is currently rising, and/or by reducing consumption per capita. Population growth control is anathema to two major religions (Moslem and Catholic) and is deemed politically incorrect in western democracies. Overall it is unpopular and thus unfeasible. Consumption per capita is proportional to the gross domestic product (GDP); hence, reduction in per capita consumption is anathema to the global capitalist system. The current global economic system is based on growth, and expects the GDP of China, India and many other countries to rise! Dr. Y. Kobayashi, President and CEO of Mitsubishi Chemical Corporation, in his introductory plenary address to ISCRE 20 shared the unwelcome news that Japanese GDP dropped slightly in the past year. No-one applauded that as a Japanese contribution to reducing the global damage to the environment! In view of Eq. (1) rational people conclude that the only hope that we have in reducing the global pollution is in increasing all measures of process efficiency (atom, mass, energy) via increased application of science in process intensification and scale-up. However, since process efficiency increases only asymptotically to unity, and any improvement requires considerable investment of capital and time, we must make a long term commitment to this endeavor. Thus, we must recognize that the

key long term challenge for CRE is the development of new more efficient economic technologies via an *improved science-based scale-up methodology for transfer of molecular discoveries to sustainable processes in meeting the future energy, environmental, food and materials needs of the world.*

To succeed in introducing greener environmentally friendlier processes we must get involved as citizens in all countries of the world to demand a system of incentives and penalties that will encourage innovation and green processing and close the doors for transferring last century technologies to the developing countries. This will not be easy as the lack of universal adherence to the Kyoto principles clearly illustrates. At present, it is more profitable to license proven processes, which often we do not want to handle in the western democracies, to the developing world, than to take the risk and develop superior processes that are environmentally more benign. This should change. As engineers and scientists we must reinvigorate our approach to reactor and process selection, scale-up and design and rely increasingly on scientific fundamentals. This will reduce the risk of implementing new technologies and open the door for greener processing. Hence, let us consider the methodology that we have available and assess what is needed for further improvements.

2. Chemical reaction engineering methodology (status)

Various concepts of chemical kinetics and kinetic-transport interactions have been introduced into branches of physical and industrial chemistry since the end of the 19th century. Chemical reaction engineering (CRE) arose as a discipline in the post WW2 period, spread from Europe (Kramers and Westerterp, 1964) to the USA and then globally (Levenspiel, 1962). The powerful chemical reaction engineering (CRE) methodology, developed over the last 60 years, offers a rational way for quantifying reactor performance based on mass, energy and momentum balances by proper consideration of the prevalent multi-scale transport and kinetic phenomena (see Fig. 2). Understanding these multi-scale transport kinetic interactions is essential for the rational selection of the best reactor type for a given chemistry and catalyst and is imperative for successful scale-up. The relevance of CRE to waste minimization and environmentally benign processing was established early (DeLasa et al., 1992; Dudukovic and Mills, 1999).

The backbone of CRE is the ability to quantify kinetic transport interactions on a variety of scales and utilize them in assessing the effect of reactor performance on the whole process. As suggested in Fig. 2 a reactor model requires mass and energy balances to be performed on an appropriately selected control volume. The connectivity of these volumes depends on the level of understanding of the flow field (and distribution of phase holdups in multiphase systems) in the reactor. The source terms in these conservation equations contain the description of kinetic transport interactions on smaller scales and ultimately the molecular level information. While the knowledge of molecular kinetics and local transport interactions is essential for proper quantification of the reactor performance, so is the flow pattern in the reactor.

The choice of the proper reactor type and operating conditions for a given process chemistry determine volumetric productivity (i.e. space time yield) and selectivity. While the reactor typically represents between 5% and 15% of capital and operating costs of the plant, it is well understood that the reactor choice determines the number and load on pre-reactor and post-reactor separation units and dictates the cost of the whole process. Therefore, the choice of the proper reactor type is essential and it should be based on a rational multi-scale approach (Krishna and Sie, 1994; Lerou and Ng, 1996; Dudukovic et al., 2002). This requires a

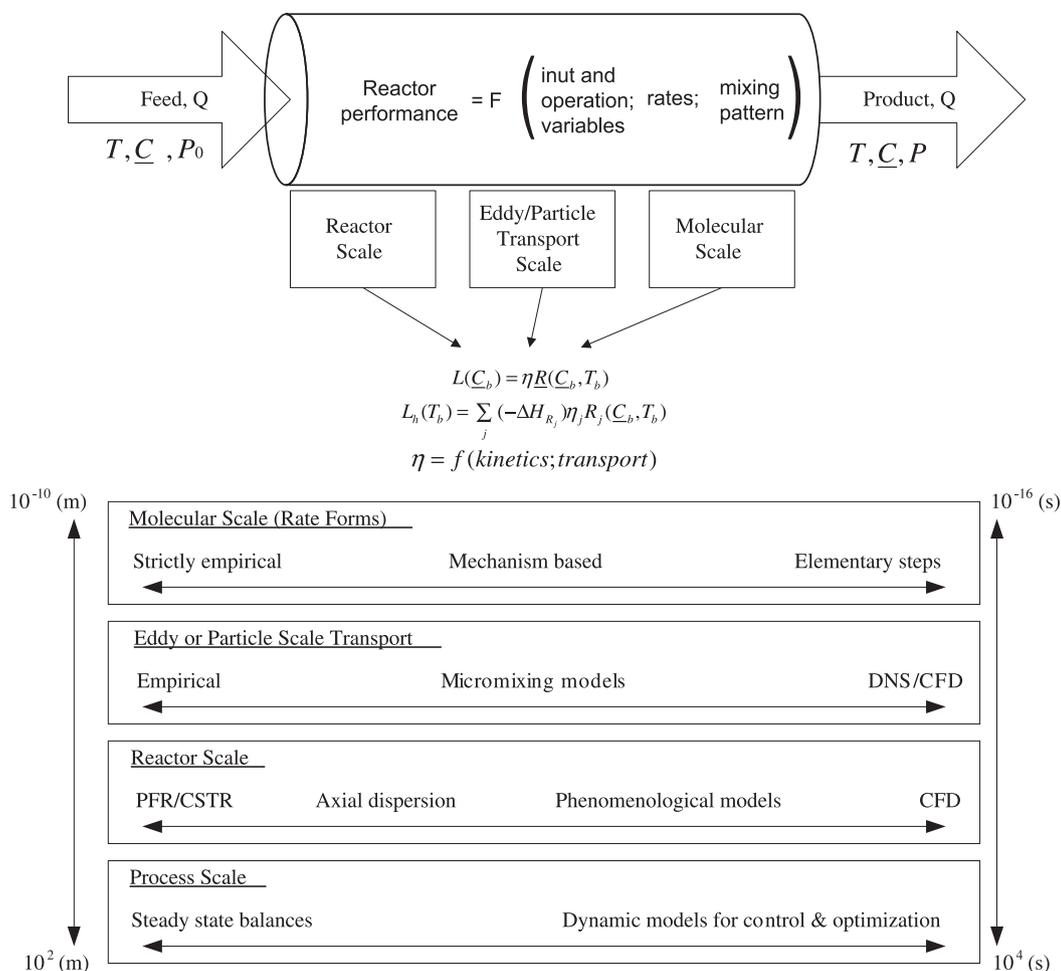


Fig. 2. Schematic of the multi-scale reaction engineering methodology.

reactor model which can capture the events on a multitude of scales at the right level. The complexity arises from the fact that the interactions of events on various scales are dependent on the scale of the equipment. It is increasingly necessary, in order to accommodate and use novel more active catalysts, to understand the change of the flow pattern with change in reactor scale and the interaction of this flow pattern with meso-scale transport and molecular scale events. This again requires appropriate models. Hence, moving our level of understanding of all scales affecting reactor performance (Fig. 2) to higher level of being more quantitative and predictive is needed for reducing the risk of scale-up and for design of the next generation of reactors and processes.

There is plenty of evidence that industries that practice reaction engineering at a higher level, and have better quantitative model based descriptions of their reactors, are doing better than others in reducing environmental damage and waste. The environmental impact factor E, expressed as the mass ratio of waste to wanted product produced, is low in petrochemical (< 0.1) and basic chemical industries (< 3). So-called high tech industries, which are really high value added industries (e.g. electronic industry, pharmaceutical industry, etc.), have unacceptably high E-factors (in excess of 100) and are not high tech at all from the environmental standpoint. However, because the production rates are so high in fuels and commodities we still must improve the processes for these industries in order to have a serious impact on reducing the global damage to the environment. These further improvements in E-factors and environmental foot

prints of large scale industries require serious long term investments in research and technology, followed by large capital investments in new plants and a favorable political and economic climate that will reward such investments. It is not surprising that many global companies, driven by the profit motive and current political climate, opt for implementing old established technologies in developing countries that welcome it. This transfers the local pollution burden from the developed to the developing world without reducing the global pollution burden. The air quality in China and India is as bad as or worse than the air quality was in the US in the 1950s. In the western world the global companies focus on new high value added technologies where the improvement of E-factors is easier to achieve with less investment due to historically high E factor levels and the neglect of this issue in the past. While any improvement is welcome news the impact on the global environment is not as good due to small foot prints compared to large scale industries.

It is important to note that the CRE methodology, sketched out in Fig. 2, and briefly outlined above, has served us well in the past. It has contributed significantly to increase efficiencies in the petroleum, petrochemical and chemical industrial sectors, for which it was originally designed. It has also enabled rapid commercialization of new technologies such as specialty polymers, electronic materials, optical fibers, combustion syntheses of ceramics and metal alloys, manufacture of inorganic and organic composites, and has played an important role in advances in biotechnology and bioengineering. Notably in all these new technologies some of the basic well established principles of

CRE were 'rediscovered' (e.g. surface reaction–transport interactions in CVD, Thiele modulus and effectiveness factor in cell growth and immobilized enzyme promoted reactions, ideal stirred tank and plug flow reactor in biotechnology, etc.). Each technology requires a thorough understanding of the key time scales and a quantitative model to relate them to key process parameters. Chemical reaction engineers are best skilled in accomplishing this.

Adoption of green manufacturing principles (Douglas, 1992; Allen and Shonnard, 2002; Tunca et al., 2006) besides political changes will require technical innovations on all scales. Such innovations are possible and increasingly depend on increased level of science. Much investment has been made in advancing the state of understanding of the molecular scale in chemistry, biology and physics. While this is good, and should continue, the discovery of new molecular scale phenomena and chemistries alone is not sufficient for establishment of environmentally benign and efficient processes. We need a science based approach for quantifying the transport–kinetic interactions on a variety of scales, beyond the descriptions provided in the past, in order to accomplish an efficient and timely transfer of these new discoveries to commercial practice.

At present, there is a wide spread belief among company managers that once a new chemistry (or biology) is identified, the body of knowledge needed to provide commercialization is readily available. This belief is not supported by reality, as most new chemistries are abandoned in attempts to design a reactor and process for them. Thus improved understanding of the reactor scale is often needed to make new chemistries successful. Let us consider for a moment how one goes about selecting a reactor for either old process chemistry with new catalyst or for new green chemistry. Two choices are available. The first one is based on the old paradigm of 'do one thing at a time' that guided our discipline for decades. Thus, one examines the figures of merit and attributes of the well established reactor types such as packed bed, catalytic wall reactor, bubble column, stirred tank, riser, fluidized bed, moving bed, etc. Based on multi-scale analysis one selects the reactor type that has, on the relative basis, the smallest transport resistances and maximizes selectivity and volumetric productivity on the reactor scale. Such a reactor, operated optimally, reduces the number of needed separation units in the process flow sheet. The second approach, popularized during the last couple of decades, is to implement proper means of process intensification into the reactor operation. Process intensification often relies on combining reaction and separation in the same unit (e.g. reactive distillation, catalytic distillation, membrane reactors, etc.) leading to so called multifunctional reactors (Westerterp, 1992). Process intensification also results from dynamic rather than steady state operation (e.g. regenerative adsorption, reverse process for exothermic reactions, coupling of exothermic and endothermic reactions, etc.). Another popular method is reactor miniaturization which results in vastly reduced transport resistances. A partial list of process intensification concepts introduced in the past few decades is listed in Table 1 but will not be discussed in detail here. Among intriguing novel ideas for process intensification are those of Coppens (2005) and Van Ommen et al. (2007) who advocate mimicking nature at the micro, meso and reactor scale in achieving efficient distribution of reactants to active sites.

One should point out that while innovative concepts on the meso and reactor scale have been introduced (see Table 1) the quantitative description of transport kinetic interactions in 'old' established reactor types, as well as in 'new' ones, has not been advanced much in the last several decades. The reactor scale flow description is at primitive levels. In multiphase reactors, which dominate the scene by appearing in over 95% of all processes, the understanding of the effect of reactor scale on the extent of flow

Table 1

Various means of process intensification.

Combination of reaction and separation
Reactive/catalytic distillation
Membrane reactors
In-situ adsorption
Distributed feed
Miniaturization and transport enhancement
Micro-reactors
Structured packing
Monoliths
Rotating packed bed
Dynamic forced periodic reactor operation
Gas–solid exothermic reactions
Gas–solid exothermic and endothermic reaction coupling
Gas–liquid–solid systems (trickle beds)
Use of solvents and supercritical media
Phase transfer catalysis
Enhancement of selectivity and ease of catalyst separation

and mixing pattern departures from the ideal reactor flow patterns has not advanced much. Volumetric mass transfer coefficients are still described based on decades old models. They are estimated based on outdated correlations that represent the whole reactor with a single coefficient which cannot capture the effect of reactor scale on the key transport phenomena. Hence, for more active catalysts, when reactor scale may become the limiting step, scale-up is uncertain and difficult. Scale-up is the dominant issue in industrial practice and it has been largely ignored in academic research and teaching (Bisio and Kabel, 1985; Euzen et al., 1993; Trambouze and Euzen, 2004).

3. Reactor scale-up (current status)

The pioneering work of Amundson and Aris generated great interest and follow up in the area of reactor modeling. This mathematical approach to quantification of phenomena in chemical reactors is best illustrated in the special volume containing the selected papers of Neal Amundson (Aris and Varma, 1980). Such models have been extremely useful in guiding our thinking. Advances have been made in ensuring that the models capture multi-scale information at the desired level of accuracy. Nevertheless, a priori reactor design, based solely on models, remains illusive and experimentation of some kind is always required. Actually, the typical process development effort can be best summarized as follows. Once the product for the market has been identified, different starting materials and reaction pathways are examined. At this stage life cycle analysis and the 12 green chemistry principles lead us to identify the environmentally best and most sustainable pathways. Process flow sheets are assembled and environmental impact analysis is done together with the usual economic assessment. These assessments rely on standard computerized tools (e.g. Aspen, WAR, etc.). The elements of these programs often are based on idealized assumptions and contain some information obtained on the bench scale. Hence the question arises whether we will be able to reproduce on commercial scale the data obtained on the bench scale. We consider below the basic options we have for accomplishing this and what level of science each requires. But first let us examine how we obtained the bench scale data. Ideally, we should always seek the best chemistry that maximizes atom, mass and energy efficiency. Based on the understanding of the reaction pathways involved, we should then seek the reactor with the best flow pattern and phase contacting pattern for implementation of that chemistry. At the same time we should examine

opportunities for effective coupling of reaction and separation or for other means of process intensification. Only then bench scale experiments should be performed and scaled-up. However, most of the process development effort in the world currently is not conducted in this manner. The chemistry, and/or catalyst that will do the job, is found by trial and error (e.g. combinatorial analysis, etc.). Usually the reactor type that the company is well familiar with is tried next, and the best operating conditions are sought by statistical approaches with limited understanding of the underlying phenomena. Experimentation on different larger scales is avoided as too costly and plants are built based on computerized procedures that essentially are based on half a century old correlations dressed up in Excel, Power Point and Visual Basic. Often expected design figures of merit are not reached by the full scale plant and experimentation continues with full scale plants, a really expensive activity. Thus, avoiding the use of pilot plants due to high cost leads to non-performing plants that end up being far costlier. No wonder most managers prefer not to take the risk and stick with the available technology.

The key scale-up issue can be summarized as follows: Once the reaction system was successfully run in the laboratory to produce the desired conversion, yield and selectivity, how to reproduce the results at a commercial scale. Two approaches are available: horizontal scale-up (scale-up in parallel, or scale-up by multiplication, or scale-out) offers one alternative, while vertical scale-up offers another. Only the latter must account for the effect of equipment scale on the interplay of transport and kinetics. The former keeps the geometry, flow, contacting pattern and flow regime the same but has to deal with the logistics of system integration and flow distribution. Key scale-up considerations require us to match the mean residence times or mean contact times in multiphase systems. This requires the knowledge of phase holdups and their flows. To ensure the same performance, we must also match the variance of residence times in the reaction environment and/or the covariance of sojourn times in different phases (Dudukovic, 1986). In systems that use scale-up in parallel this requires that we have the same flow and the same amount of catalyst in each parallel branch. Hence our scale-up challenge is reduced to the task of implementing perfect flow distribution and control.

In recent years micro-reactors (Jensen, 2001; Lowe and Ehrfeld, 1999; Losey et al., 2001; De Mas et al., 2003; Khan et al., 2004; Hessel et al., 2005) have received a lot of attention because they combine the possibility of scale-up by numbering up with process intensification achieved by reduction of transport resistances due to miniaturization. Micro-reactors offer a whole series of advantages such as: (1) high surface to volume ratios and, due to small dimensions enhanced mass and heat transfer coefficients by one to two orders of magnitude, (2) laminar flow conditions and low pressure drop but ability to make RTD narrow by introduction of another phase (e.g. segmented flow), (3) controllable RTD and back mixing, (4) high volumetric productivity, (5) low manufacturing and operating costs, (6) increased safety due to small amount of material, (7) scale-up in parallel (scale out). Jensen and colleagues have illustrated the potential use of micro-reactors in a variety of reactions (Losey et al., 2001; De Mas et al., 2003) and in material synthesis (Khan et al., 2004). In their comprehensive review papers Hessel et al. (2005) established the foundation for micro-reactor technology (Lowe and Ehrfeld, 1999) and summarized the contacting principles in gas–liquid and gas–liquid–solid micro-reactors Hessel et al. (2005). They reviewed the characteristics of a variety of contacting patterns attempted and reported a vastly improved mass and heat transfer coefficients, much larger interfacial areas, controllable RTDs, increased volumetric productivity, ease of scale out. They offer demonstrations of successful bench scale use of multiphase micro-reactors in

direct fluorinations, oxidations with fluorine, chlorinations, sulphonations and hydrogenations.

The following question then arises. With all their perceived advantages, and the technologies available to manufacture micro-reactors in silicon, in glass and in steel, or other metals, why aren't they more widely used? Part of the answer lies in the fact that generally they require very fast reactions and active stable catalysts (usually these two do not go together). Most importantly micro-reactors are, due to small dimensions, more prone to fouling and clogging, leaks between channels, and their reliability and life on stream are unknown. All of these disadvantages present potentially solvable problems on a case by case basis. However, the perceived risk factor is too large at the moment for micro-reactors to replace existing installations especially for commodity scale production. Most likely acceptance of micro-devices will occur in consumer products, distributed power systems, highly energetic fast reactions, in-situ production of hazardous chemicals. Other applications will come later due to the attractiveness of scale-up by numbering out.

Scale-up in parallel is suggested for monoliths also, were the argument that monolith performance can be predicted based on a single channel performance is often made. Nevertheless, during the last four decades monoliths made inroads only in few applications: automobile exhaust (multiple distributed units), and power plant gas cleanup (SCR of NO_x). Applications in other areas, especially for gas–liquid–solid catalyzed reactions, are slow to come. The customer resistance factor is too large in spite of the fact that Liu and Roy (2004) and Roy et al. (2004) showed that monoliths can outperform trickle beds. The studies of Kreutzer et al. (2005a, b) established, based on firm theoretical principles, the conditions when good flow distribution can be obtained in monoliths in gas–liquid flows and provided guidance for scale-up. Their work will be helpful in finding a wider acceptance of monoliths in the future. New technology such as gamma ray computed tomography and NMR are able to provide confirmation of theories for prediction of uniform flow distribution and identify conditions when maldistribution is expected. Studies in our laboratory by gamma ray tomography (Roy, 2006) reveal that in gas–liquid flows uniform distribution in monolithic channels cannot be taken for granted as there is only a narrow window of flow conditions that allows it. Outside that window of operation the flow distribution can be highly non-uniform and considering the monolith to be a bundle of identical channels would not model reality right. Hence, more science is needed for understanding monolith bundles as the scale-up in parallel approach at some conditions is not as simple as originally envisioned.

It is clear that in the foreseeable future micro-reactors and monoliths will not replace most of the multi-phase reactors and hence vertical scale-up, i.e. the change in size of the reactor will be needed in transferring bench scale results to commercial practice. Such scale-up effort is much aided by the availability of an appropriate reactor model that can reflect the effect of the change in scale on flow, transport and mixing in the reactor. Significant advances have been made in reactor scale models especially by incorporation of computational fluid dynamics (e.g. Kuipers and Van Swaaij, 1998 and Ranade, 2002) and by better description of reactive turbulent flows (e.g. Fox, 2003). For proper quantitative description of reactor performance it is important to properly describe how are species brought into contact on all scales by flow and mixing. For example, when we have a reactor system with two moving phases it is important to be able to describe the flow pattern of each and the exchange between them. In the past we relied almost exclusively on ideal reactor assumptions and treated each phase as being either in plug flow or perfectly mixed. Available reactor programs embedded in Aspen and other process simulators are based on ideal flow pattern models. When reality

does not conform to these assumptions the axial dispersion model (ADM) is often used to match experimental observations and model predictions. It has been recognized, however, that ADM is not a predictive model and that one needs more accurate flow and mixing models based on the physical phenomena that occur in the system. For reactor types used in practice there are no reliable theories that can predict the change in the flow pattern with the change in reactor size and in operating parameters and feed conditions. Thus in multiphase systems it is hard to predict the variation in phase holdup, its distribution and the change in flow pattern to be able to firmly predict the mean contact times and the change in the variance in contact times with the change in scale of the unit. Recall that maintaining these at constant level or predicting the effect of scale on their change is essential for successful scale-up. Most importantly data on sizeable units on flow patterns, phase distributions and contacting and mixing are sparse and in general not available. We must strive to improve this situation by generating more experimental information to assess the ability of advanced CFD flow models to predict what we observe. Without experimental validation CFD of multiphase flows is an uncertain art. Unfortunately, multiphase systems are opaque and our optical techniques do not work in them. Hence the first order of business is to develop means to measure phase holdup and velocity distribution in opaque systems (Chaouki et al., 1997a, b). We have always advocated the following modern approach to multiphase reactor modeling and scale-up (Dudukovic, 2000): (a) capture the physics of flow by experimental means, (b) use CFD models and validate the results experimentally at different conditions and in systems of different sizes, (c) complete a physically based engineering model for flow and mixing based on validated CFD models and (d) coupling these physically based models with kinetics and catalyst deactivation models when possible. In some situations it may be necessary to replace steps (c) and (d) with a novel type model based on data. To accomplish step (a) we have designed and built two units (Kumar and Dudukovic, 1997; Larachi et al., 1997). One is a gamma ray computer tomography unit (CT) for assessment of phase holdup distributions in various cross sections of the reactor type of interest (Kumar and Dudukovic, 1997). The other is a computer assisted radioactive particle tracking (CARPT) unit which provides us with the complete Lagrangian description of the phase traced by a single radioactive particle (Larachi et al., 1997). The virtues of these have been enumerated in numerous student theses and published papers. We have illustrated how the experimental results obtained by CARPT and CT can be effectively used for validation of CFD codes (Pan et al., 1999; Pan and Dudukovic, 2000; Sanyal et al., 1999; Rafique et al., 2004; Chen et al., 2004) and how such data and the validated codes can be used to build improved fundamentally based models for flow and mixing in bubble columns (Degaleesan et al., 1997; Gupta et al., 2001a, b), risers (Roy et al., 1997, 2005; Roy and Dudukovic, 2001; Roy, 2000; Bhusarapu, 2005; Bhusarapu et al., 2004, 2006) stirred tanks (Rammohan et al., 2001, 2003; Rammohan, 2002; Guha, 2007; Guha et al., 2006, 2007, 2008; Khopkar et al., 2005) and ebullated beds (Limtrakul et al., 2005). These results will not be repeated here as they are readily available in the open literature.

4. Multi phase reactor selection, modeling and scale-up

In this manuscript we argued that an increased level of science is required to significantly reduce the risk of vertical scale-up of multiphase systems. It was also stated that available ideal reactor or axial dispersion model are inadequate for such a task and that modeling and scale-up should be based on models that capture the physics of flow well. We also showed that noninvasive

techniques are now available to provide an actual insight in flow, mixing and phase distribution in the opaque multiphase systems. Here a few examples are given related to green processing as to how such improved models can aid in avoiding pitfalls in scale-up or in providing useful predictions. Two examples are considered: solid acid alkylation and partial oxidation of butane.

The example of *solid acid alkylation* is chosen because the currently licensed processes use mineral acids like HF and sulfuric acid as catalysts and in spite of many improvements present an environmental hazard and generate undesirable waste. A reaction scheme basically involves a second order reaction of olefin and paraffin (first order with respect to each) leading to the desired alkylate and the undesired second order reaction of olefins to unwanted byproducts. It took the industry decades to move to the best flow pattern which at high paraffin to olefin ratio in the feed distributes all the paraffin to flow sequentially through a sequence of reactors while adding olefin (to keep its concentration low) in each reactor. It is not surprising that as long as the licensing business to the developing world is good the industry does not have much incentive to explore environmentally beneficial alternatives involving solid acid catalysts.

Solid acid catalysts offer a potential advantage in selectivity and eliminate liquid waste and environmental hazards but they do deactivate rather rapidly on stream. Slowing down catalyst deactivation and enabling on stream regeneration at mild conditions are areas of current research involving the catalyst particle scale. For any of such catalysts the question arises as to what reactor type should be used. The choice of liquid–solid riser as reactor is only justified if liquid and solids are in plug flow. At large Reynolds number in highly turbulent flow in vessels of very large height to diameter ratios the prevailing opinion in company manuals and texts suggests plug flow for each phase with the slip velocity being tied to solids holdup. One should note that if solids experience back mixing in the riser then their mean activity and selectivity in the reactor will be different than if they were in plug flow. Moreover the extent of back mixing may depend on reactor diameter. A study to examine and quantify solids distribution and back mixing was commissioned in our CREL and the findings have been reported in the thesis by Roy (2000) and in a series of papers (Roy and Dudukovic, 2001; Roy et al., 2005). The pertinent results are summarized here and for details and actual values the reader is referred to the literature. The pertinent illustrations are available in my power point ISCRE 20 presentation on our web site.

Liquid was indeed in plug flow in the riser as expected and as determined by conventional tracer impulse response studies. More than 20 tanks in series are needed to describe the liquid residence time distribution (RTD). A single radioactive particle, representative in size and density of the solids in the system, exhibits a highly three dimensional trajectory during its rise through the riser including some downturns and loops. However, ensemble averaging of instantaneous particle velocities, obtained from a couple of thousand of such trajectories, produces an axisymmetric picture of the averaged solids flow with upward motion in the center of the column and downward motion of solids close to the walls. CFD computations, based on the Euler–Euler model, reveal a highly complex three dimensional instantaneous solids flow structure (which explains the shape of the single particle trajectory) but 75 s of averaging produces the symmetric flow pattern of rising solids in the middle and falling by the wall. The agreement between simulations and data is excellent for solids velocity, solids holdup distribution and granular temperature, i.e. solids kinetic energy. In addition, the components of the solids eddy diffusivity tensor are obtained directly from the Lagrangian trajectories. Computed tomography (CT) scans shows relatively uniform solids holdup in the column cross section with

somewhat more solids by the walls. This holdup profile is axial position independent testifying to fully developed flow. All the needed information for development of an appropriate level reactor model is now available for coupling with kinetics and deactivation. Most importantly CFD has been validated and, since this technology does not require large increases in column diameter with scale-up, CFD can predict all the model parameters and can be used for scale-up assessment. As a bonus, by monitoring the time of entry and exit of the tracer particle from the riser section by CARPT one obtains the residence time distribution of the solids in the riser. This confirms that solids flow deviates sometimes significantly from plug flow as two to six tanks in series describe the obtained RTD. This makes the riser unsuitable for the current generation of the solid acid catalysts and directs the research to search for slower deactivating catalysts and different reactor type or operation mode.

The second example deals with *partial oxidation of butane to maleic anhydride*. The conventional process uses benzene and wall cooled packed tubular reactors. Replacement of benzene by butane was hailed as a green process, because the theoretical carbon efficiency is increased to from 67% to 100% and the mass efficiency from 44% to 58%. These improvements are noted and recorded in the literature on green processing. A special VPO catalyst was developed for this reaction. Experimentation revealed that the VPO catalyst loses rapidly its activity and selectivity when used in wall cooled packed tubular reactors. It became clear that the optimal activity and selectivity of the catalyst can only be realized in cyclic transient operation in a circulating fluidized bed (CFB) arrangement. The oxidized form of the catalyst is then used to do the partial oxidation of butane (hydrocarbon) in the riser. Upon removal of the product the catalyst is introduced to the fluidized bed where it is reoxidized by air before being returned to the riser. An excellent research effort developed a highly porous strong attrition resistant coating for the catalyst that did not interfere with its functionality or with transport in and out of the catalyst. Thus a CFB was a perfect match for the reactions to be executed and for the developed catalyst. Long term pilot plant studies on a riser several inches in diameter established the desired butane conversion and selectivity to maleic and confirmed the longevity of the catalyst on stream. A large plant was built and never operated to specification so that after many 'plant scale experiments' it had to be decommissioned. Hence, a textbook green process did not make it due to dramatic deterioration in performance with scale-up. This was most likely caused by the belief that after the pilot plant operated successfully construction of the plant reactor is a routine task. That is not so. The most recent CFB 9 International Congress in Hamburg in May 2008 reveals that while know—how exists for operating balanced CFB systems (riser and fluidized bed) in gasification and combustion very little fundamental knowledge is available regarding solids flow pattern and distribution. Knowledge of solids holdup and solids circulation rates is based on old unproven correlations. If solids holdup is not known with certainty and neither is the solids circulation rate, then how can one predict solids mean residence time or its variance with the change in reactor scale? Yet, the constancy of both quantities is required for successful scale-up! The answer is that at present without an improved scientific base one cannot do it with sufficient certainty.

The studies in our CREL with CARPT on cold models of gas–solids risers (Bhusarapu, 2005; Bhusarapu et al., 2004, 2006) reveal that only this technique can provide the true residence time distribution of the solids in the riser since the time of entry of the single tracer particle into the riser and the time of its exit at the upper boundary is precisely known. Studies with injection of multitude of particles do not provide the correct answer as long

ago correctly pointed out by Nauman and Buffham (1983) for these types of systems with 'open boundaries'. The tendency to interpret these impulse responses via the axial dispersion model can lead to errors and masks the true nature of back mixing and its magnitude. Our studies reveal that at conditions when the macro-mixing index of Villermaux (1996) (defined as the mean trajectory length experienced by the solids in the riser divided by the riser height), which can be directly calculated from CARPT experiments, increases indicating more solids back mixing, the axial dispersion model predicts a decrease in the axial dispersion coefficient (increase in Peclet number for axial dispersion) indicating erroneously less solids back mixing. Clearly, improved physically based models and proper CFD computations are needed for gas solid risers to enable development of a rational model for scale-up. CFB technology is a sound one is needed for many partial oxidations, redox systems and chemical looping in carbon sequestration. We must improve the scientific basis for its scale-up. We have shown that using CARPT one can obtain the true descriptions of solids residence time distributions in the riser and map completely the solids flow field. It is the task of CFD modelers to develop codes that can predict the experimental observations of CARPT. They are still working on it.

It is illustrative to assess as to where we are in our ability to predict and scale-up the *performance of stirred vessels*, which are ubiquitous in industrial practice, if we follow the approach advocated earlier of quantifying the flow pattern first and then using it to develop an engineering physically based model for coupling with kinetics. Rammohan et al. (2001, 2003) and Rammohan (2002) have shown that in a standard baffled tank stirred with a six blade Rushton turbine CARPT can, in single phase flow, produce results which are in agreement with LDA and PIV measurements. Mean flows are correctly predicted as well as the distribution of kinetic energies. CARPT misses only the 10% of the energies at the high end of the frequency spectrum since this motion cannot be sampled by the tracer particle. The dead zones at the bottom of the tank are also precisely captured by CARPT. Since in single phase flow available CFD codes (e.g. Euler–Euler method in Fluent) do predict data observed by CARPT, a compartmental model based on CFD is useful for coupling with kinetic information. Such a compartmental model for a stirred tank is outlined in the thesis of Guha (2007). Flows, convective as well as by turbulence, are obtained from CFD computations. The rules used in the discretization scheme ensure that the Damkohler number for each cell is unity or less. The model is then tested to evaluate the effect of the position of the injection port (top feed vs. bottom feed) for the injection of reactant B over a short time period (15 s) into the batch of equal molar amount of reactant A. The reaction scheme is $A+B=R$ plus $R+B=S$ and the experimental results were reported by Paul and Treybal (1971) for the yield of R after the reaction reached completion. The experiment and compartmental model simulation results are in reasonable agreement (Guha, 2007; Guha et al., 2006). It is shown that there are no advantages to use the full blown CFD model that couples directly momentum and species balances (Guha, 2007; Guha et al., 2006). The compartmental model provides the results much faster than the full blown CFD with coupled species balances and this difference increases with the number of reactions.

Extension to two phase stirred tank systems is in principle possible. CARPT and CT measurements have been performed by Rammohan (2002) in a gas–liquid system and by Guha (2007) in liquid–solid systems. Unfortunately, CFD models (Euler–Euler or LES) are not able to produce results in satisfactory agreement with the CARPT–CT measurements in either gas–liquid (Khopkar et al., 2005) or liquid solid systems (Guha et al., 2007, 2008). In absence of a reliable CFD prediction compartmental model cannot be built. These points are to the constant need for validating multiphase CFD codes.

At the end, it is useful to reiterate that mastery of all scales will be needed in providing for the best science based reactor technology of

maximum efficiency. Let us consider here the work of Luo (2003) who in our CREL, under Dr. Al Dahhan's guidance, embarked on maximizing the reactor volumetric productivity of algal growth in biomass production. Many microbiologists study cyanobacteria and algae due to their ability to use photosynthesis in producing either a specific product or create biomass. Luo showed that complicated mechanism for cell growth and photosynthesis can be lumped into a simpler model consisting of three coupled differential equations. The model still had the ability to correctly account for the effect of irradiation on cell kinetics. It was able to capture both photo-inhibition due to too much light and a switch in pathways due to photo-limitation caused by too little light. CARPT experiments were then done in various types of bubble columns, with and without different draft tubes, to evaluate cell illumination time distributions. Coupling this information with the cell kinetics model and the usual mass conservation laws for the reactor, Luo and Al Dahhan (2004) reported the predicted effect of column design on biomass production rate. A split column far out-performed (by a factor of 5) the simple bubble column and column with an internal symmetric coaxial draft tube. Clearly, genetic manipulations can improve cell yield of desired product or the biomass specific growth rate but only the full understanding of the meso-scale and large scale can provide us with the tools to choose the best reactor type and operate it in the optimal manner. This type of approach is suggested always to CREL industrial sponsors.

In closing one should mention that the multi-scale approach to the development of environmentally beneficial processes has been embraced and successfully demonstrated by the Center for Environmentally Beneficial Catalysis (CEBC). This multi-university center is headquartered at the University of Kansas and encourages industrial participation in or quest for greener, safer, more sustainable and economic processes. Case studies from the work of the center will be highlighted at a different occasion.

5. Summary and conclusions

Reaction engineering methodology is the enabling discipline that offers multi-scale avenues for innovations in developing "greener" more sustainable processes. It has been very successfully extended in advancing new materials and biotechnologies. However, the field is far from mature and needs further application of scientific principles in providing better, fundamentally based process development, reactor selection and scale-up procedures.

In addition to the adoption of green chemistry principles, in order to develop greener processes that will be attractive economically and have low risk of implementation, it is vital to master the scale-up methods based on fundamentals and use the multi-scale approach.

Reactor miniaturization provides opportunities for scale-up in parallel. Quantification of multiphase flow fields reduces the risk of (vertical) scale-up. Both are needed for rational process intensification, and research in both areas should intensify!

Incentives should be provided globally to companies for taking the risk to implement new more efficient technologies that utilize process intensification concepts on a multitude of scales.

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