The relation between size of plant and risk: traditional processing versus intensive continuous processing.
Proceedings of European Congress of Chemical Engineering (ECCE-6)
Copenhagen, 16-20 September 2007

The relation between size of plant and risk: traditional processing versus intensive continuous processing

Parastoo Khoshabi
Parastoo.khoshabi@postgrad.manchester.ac.uk
Paul. N. Sharratt
paul.sharratt@manchester.ac.uk

[a]*Department of Chemical Engineering, University of Manchester, PO BOX 88 Manchester M16 1QD*

Abstract

There have been always problems regarding scale up of a process to move from laboratory to a commercial scale plant. In general, this is considered in terms of process performance, while safety and environmental issues are considered based on the full scale design. The relation between safety and environmental risk and the size of plant is an important consideration in design but one that is often considered only indirectly. The magnitude of hazards change with scale in ways that depend on their nature, as well as the response time of equipment, inventories in process, changes in the ability to control etc.

The IMPULSE project[1] aims to deploy innovative process equipment such as micro reactors, thin–film devices and other structured components to attain step-change performance enhancement for whole processes, including intensification, thereby contributing to significant improvements in supply–chain sustainability. One theme pursued is the numbering up (rather than scale-up) of processing devices, so that in IMPULSE the size of commercial devices is the same size of the size of equipment in the lab. By this means it is expected to have less problems regarding performance change on scale up. Also, it has been widely claimed that process intensification leads to safer manufacture. However, for intensive plant there may be many small devices, so while the hazardous consequences of failures might be low, the frequency of occurrence of hazardous events might be higher. This paper analyses the issues that could arise. For example, by having smaller plant, several parallel streams might be needed for the same rate of production. In this circumstance, the total leakage (for example) might more for the IMPULSE plant than a conventional one. In addition the (small) equipment will be very closely spaced, raising questions as to the risk profile of such a plant.

1 Introduction

If a system remains non-hazardous when subject to all deviations that might lead to danger, the system is called inherently safe. According to (Kletz 1998) this arises from designing a safe plant and not by adding equipment to the system to make it safe. This can be achieved by preventing problems at their root causes; therefore inherent safety has an important role during process design.

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[1] This research is a part of the EU FP6 IMPULSE (Integrated Multi-scale Process Units with Locally Structured Elements) project. http://www.impulse-project.net/

Nowadays, industries are looking for shorter lead times, higher quality for their products as well as lower environmental discharges and safer plant. It has been suggested that in order for the chemical industry to survive in the developed world, radical and novel approaches are needed. The main aim of the IMPULSE Project (Integrated Process Units with Locally Structured Elements) is to innovate through application of structured process equipment such as micro reactors, compact heat exchangers and thin film device (Sharratt, Matlosz et al. 2006). Substitution of batch by continuous plant in pharmaceutical and fine chemical plants is another aim. The IMPULSE approach is application driven – in other words the novel devices being deployed with the aim of delivering the best possible process outcome; this is distinct from the early approaches of Process Intensification where the target was equipment size, not business success. In this context, business benefit is taken to include safety and environmental performance.

The novel, IMPULSE devices will be smaller compared to conventional devices. There should be a considerable difference between traditional and IMPULSE technologies in terms of fire and explosion risks, harmful emissions and efficiency. In order to find out whether, and in what ways IMPULSE continuous plant is inherently safer than batch, a comprehensive assessment is required.

A main strategy in developing inherently safer chemical process is process intensification. Reduction of inventory of hazardous substances or energy leads to reduction of the consequences of failure to control that hazardous substance (Barton and Rogers 1993). Safety of a plant should be based on reduction of possible damage. Safety devices are not perfect and will probably fail at some point; they are not totally reliable. In a chemical plant with large content of hazardous material or energy, the result of the failure of the add-on safety devices can be large. In a small plant the inherent capability to cause damage is reduced, so small plant can be considered safer (Stankiewicz and Moulijn 2004). Nevertheless, there is also a need to reduce probability of hazards as much as possible – it is possible that the increased complexity of small plant might result in more frequent problems and therefore increased risk.

2 Inherent safety and IMPULSE

It is important to find the best choice of technology in order to apply in a system. Choosing between IMPULSE continuous (IC) and conventional batch manufacturing methods requires deep consideration. IMPULSE is trying to achieve inherent safety through one or more of

- having all equipment as small and safe as possible
- allowing substitution of dangerous materials with less dangerous by being able to process them in ways not possible in traditional plant
- attenuation of the operating conditions and
- Intensifying the process to minimise inventories.

To move beyond the simple argument that “smaller is better” needs a more detailed analysis. It is difficult to do this at an aggregated level but can be achieved by using quantification methods to assess each of the hazard scenarios that exist in the traditional and IC plant. In this paper we develop a method to support such an assessment. Here, we consider the hazards associated with a hydrogenation plant to produce a pharmaceutical intermediate. The approach is to

- Identify the hazard scenarios that could exist for traditional batch and IC plant;
- Pick representative scenarios for assessment;
- Quantify the hazard for each selected scenario for different scales of manufacture in both batch and IC processes.

In this paper the probability is not considered as there are limited failure probability data available for failure of novel components. In this study the scenarios which are seen as credible are considered in lieu of explicit consideration of probability.

3 Method

Table 1 identifies a series of high-level hazardous situations in a hydrogenation unit. From these, fire and explosion is explored in more depth in this paper as the risk of fire and explosion is often the main safety issue in hydrogenation plant. The selected methodologies used to quantify the hazards were applied (these are explained in sections 3-1, 3-2, 3-3). The aim is to show how hazard changes with changing the size of vessels and feed pipes used in a unit, which would in turn depend on the scale and the technology adopted (continuous or batch).

Table 1: Selected hazards in hydrogenation unit

Hazard issues	How it changes for macro-scale	How it changes for meso-scale	How it changes for micro channel-scale
Fire and explosion	Greater, Due to having high inventory, of hydrogen, flammable solvent and catalyst	Lesser due to lower inventory	Further reduced by very low inventory and Micro-channels can act as flame arrestors
Runaway reaction	Contamination of material in tanks by incompatible material or material with wrong temperature.	Lower inventory may reduce hazard	Lower inventory may reduce hazard
High pressure	Higher volumes of compressed gas but pressure not generally very high.	Reduced volumes but higher pressure may be accessed.	Much reduced volumes but higher pressure may be accessed. Feed system volumes may dominate.
Ignition of PD/C catalyst.	Due to contact with a source of ignition or flammable solvent	Lesser – if catalyst quantities are smaller and contained within structure.	Lesser – if catalyst quantities are smaller and contained within structure.

The following hazard phenomena were considered in order to carry out a quantified comparison of IMPULSE continuous plant and conventional plant in the fire and explosion aspect.

- Jet fire
- Explosion

3.1 Jet fire

Jet fire happens when gas is released with high pressure from limited size opening accidentally or intentionally (AIChE 2003). The hazard arising from a jet fire can be estimated based on a simple approach (AIChE 2003). In this approach, by estimating discharge rate, the flame size can be estimated, consequently heat transfer. The heat release rate from a jet fire is controlled by the mass flow rate of the fuel which is released. Nature of release such as pressure, size, and shape of discharge point and fuel properties affect the discharge of fuel.

The assumed scenario here is the failure of a pipe to give a leak that is 30 % of the total cross-sectional area.

3.1.1 Heat release rate

According to CCPS (AIChE 2000), the heat release rate can be calculated by estimating the mass flow rate according to the equation 1. The data which needed in each equation can be estimated from AIChE (2003).

$$m = C_d \rho_{ambient} A \sqrt{\frac{k g m}{R_g\ T}} \left(\frac{2}{k+1}\right)^{(k+1)\ (k-1)} \qquad \mathbf{1}$$

$$Q = m \quad \Delta hc \qquad \mathbf{2}$$

3.1.2 Jet flame size

The heat release which is calculated from equation 2 helps to estimate the length of flame according to the equation 3.

$$L = 0.2 Q^{2/5} \qquad \mathbf{3}$$

3.1.3 Heat transfer

It is assumed in point source model that the fire is a point that is radiating to a target at a distance. In order to estimate the incident heat flux per unit surface area on a target the following equation is used. From equation 4 the heat radiated from fire can be estimated (kW) and then by using equation 5 the incident heat flux (kW/m^2) can be calculated.

$$Q_r = X_r Q \qquad \mathbf{4}$$

$$q'' = \frac{Q_r \cos\theta}{4\pi R^2} \qquad \mathbf{5}$$

In order to identify the possibility of equipment failure, it is possible to estimate the Ts (Surface temperature) , Ts between 500 – 550 C° is a typical failure criterion though some equipment fails at temperatures as low as 50- 250 C°.

$$Ts = \left[\frac{q''}{\sigma} + \quad (T\infty)^4\right]^{1/4} \qquad \mathbf{6}$$

3.1.4 Results

Table 2 shows the potential leak rates and resultant hydrogen fire sizes for rupture of different size of pipe. It can be seen that a leak from a pipe with 30 mm diameter gives a much higher heat release rate (HRR = 30 MW) compared to a leak from a micro channel (HRR= 0.3 MW). By having the leak rate and heat release rate (HRR), the impact of the jet fire can be estimated. The jet fire evaluation can be seen in Table 3, and according to Figure 2, the heat radiated from a fire of a macro pipe is extremely high compared to a micro channel pipe. In addition it is estimated that the length of flame created by a fire from a macro pipe is higher than the flame length of a micro channel (Figure 3). According to Figure 4 the heat flux on the nearby equipment from a fire of macro and meso pipe is nearly 16 and 8 times more than the channel and micro channel pipe, respectively.

Table 2: Leak and heat release rate (of combusting hydrogen) from different size of pipes

Type of pipe	**Pipe diameter(m)**	**Leak of H_2 at 3 bar (kg/s)**	**Size of hole (m)**	**HRR (kW)**
Macro	30×10^{-3}	0.253	9×10^{-3}	30,360
Meso	5×10^{-3}	7.02×10^{-3}	1.5×10^{-3}	842
Channel	0.5×10^{-3}	7.01×10^{-5}	0.15×10^{-3}	8.7
Macro Channel	0.1×10^{-3}	2.81×10^{-6}	3×10^{-3}	0.34

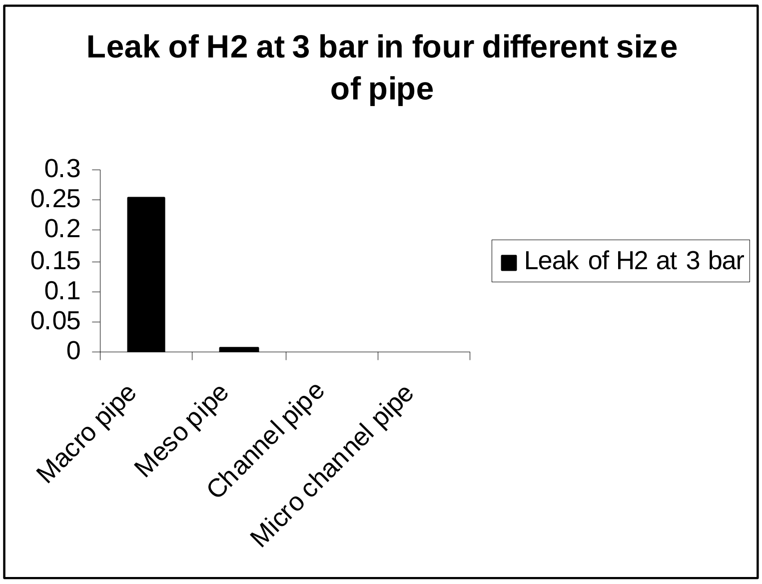

Figure 1: Leak of hydrogen in four different scale

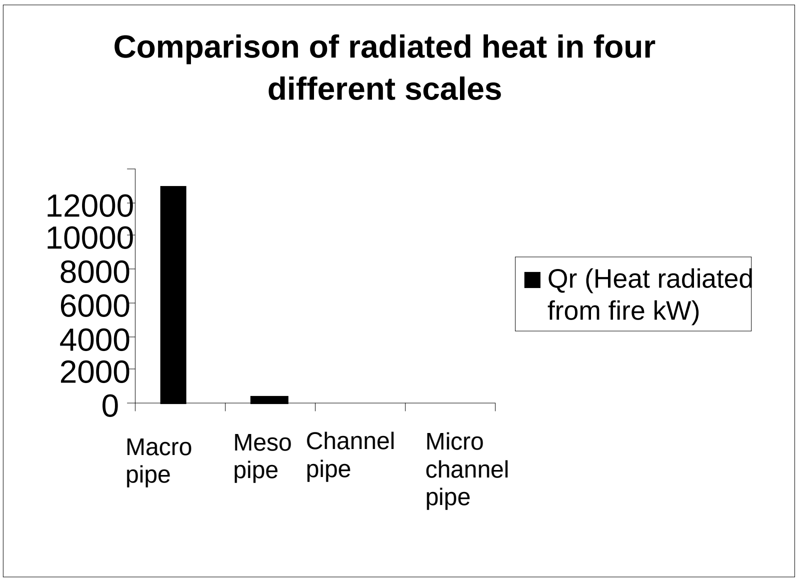

Figure 2: Radiation intensity from jet fires

Pipe	Q(Heat release rate kW)	L (flame length m)	Xr(radiative fraction)	Qr (Heat radiated from fire kW)	H pipe (m)	H target (m)	X (m)	R^2 (m^2)
Macro	30,364	12.4	0.4	12,145	1	2	4	43.1
meso	843	2.96	0.4	337	0.7	1	1	2.39
channel	8.7	0.47	0.4	3.37	0.5	0.5	0.1	0.065
Micro channel	0.337	0.13	0.4	0.13	0.5	0.5	0.05	0.006

Table 3: Data needed for estimating Heat flux and Surface temperature of near by equipment.

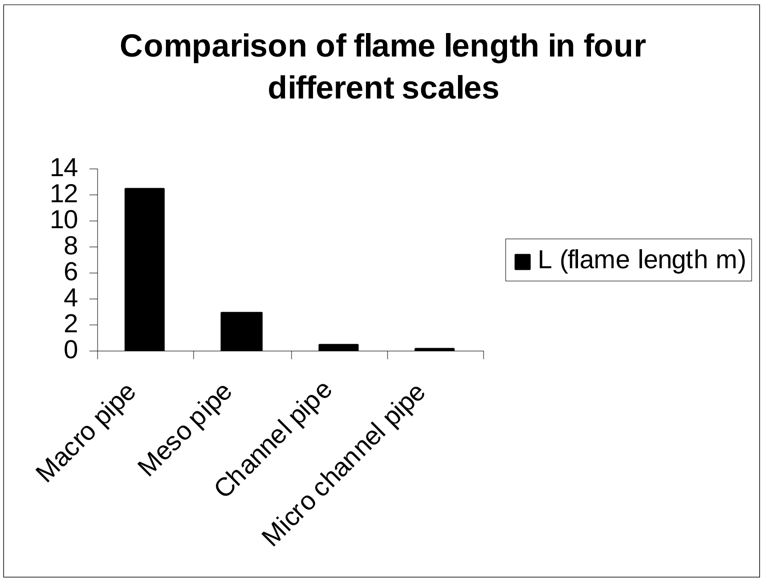

Figure 3: Flame lengths from leaks at different scales

Cos ϴ	q" (incident heat flux on nearby equipment kW/m^2)	σ (stefan constant Kw/M^{2}K^4)	Ts (C°)
0.768	17.2	$5.67*10^{-11}$	474.
0.847	9.5	$5.67*10^{-11}$	374.
0.426	1.7	$5.67*10^{-11}$	171.
0.772	1.2	$5.67*10^{-11}$	142.

Table 4: Evaluating heat flux and surface temperature on nearby equipment

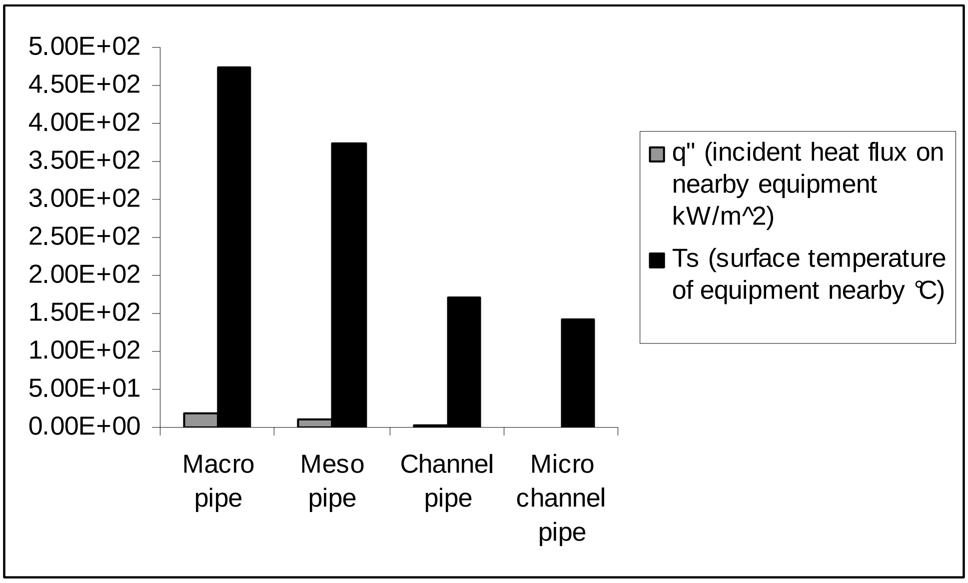

Figure 4: Heat flux and surface temperature of nearby equipment in four different scales

3.2 Explosion size

Estimation of the explosion pressure with distance can be calculated from an equivalent of TNT. In order to estimate the equivalent mass of TNT, the characteristics of hydrogen are used, such as heat of combustion, total mass of released hydrogen, explosion efficiency and the energy of explosion of TNT. 4688 kJ/kg is a typical value for TNT explosion. In this study the mass of TNT is calculated for four the different scales, again based on the assumption of release from a ruptured pipe.

$$m_{TNT} = 4.57 * 10^{-3} * d^2 * \alpha * P_0 \qquad 7$$

Where, d is pipe diameter (m), P_0 stagnation pressure at operating condition, α ratio of hole size to the pipe area. According to Table , the mass equivalent of TNT is estimated. It can be seen in Figure 5 that for macro size the TNT is quite high compared to other size of pipe.

Table 5: Estimation of mTNT for four scales

Pipe diameter(m)	α	p (pas)	M(TNT)(kg)
$3.00x10^{-2}$	0.3	$3.00x10^{5}$	$3.70x10^{-01}$
$5.00x10^{-3}$	0.3	$3.00x10^{5}$	$1.03x10^{-02}$
$5.00x10^{-4}$	0.3	$3.00x10^{5}$	$1.03x10^{-04}$
$1.00x10^{-4}$	0.3	$3.00x10^{5}$	$4.11x10^{-06}$

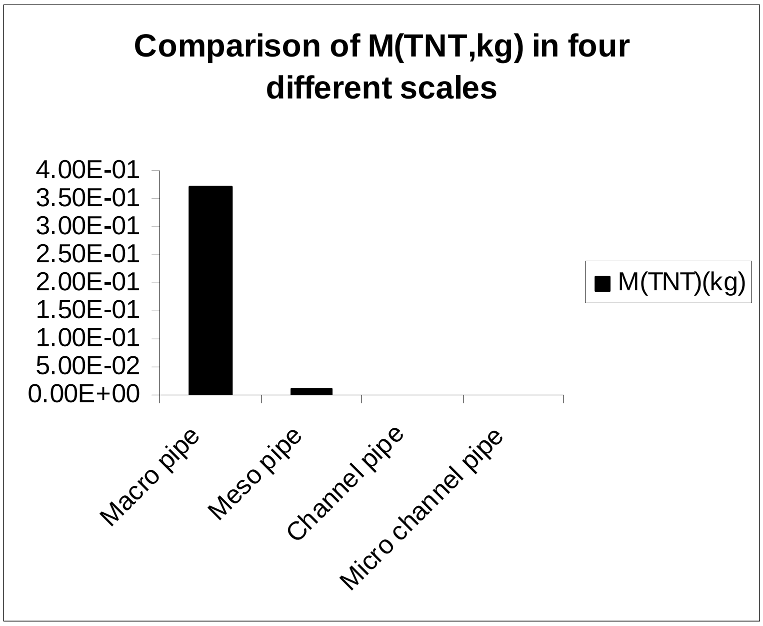

Figure 5: Mass of TNT in four different scales

4 Conclusion

It can be concluded that by reducing the scale of pipe the heat release due to a fire will decrease which stands for the physical size of fire. In addition the heat transferred to the adjacent equipment decreases as well. For example, a leak from a pipe with 30 mm diameter gives a much higher heat release rate (HRR = 30 MW). Compared to a leak from a micro channel used (HRR= 0.3 MW). In addition it is estimated that the length of flame created by a fire from a macro pipe is higher than the flame length of a micro channel. The heat flux on the nearby equipment from a fire of macro and meso pipe is 16 an 25 times more than the channel and micro channel pipe, respectively. This study will continue to apply this methods in order to estimate a fire from a conventional and IC plant and compare them to see whether by reducing the scale the hazardous situation will decrease as well.

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