

## Scale-Up Analysis of Three-Phase Oxygen Reactor in the Cu-Cl Thermochemical Cycle of Hydrogen Production

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### Abstract

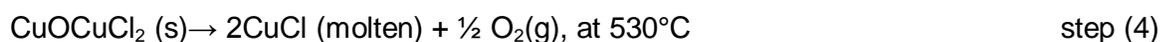
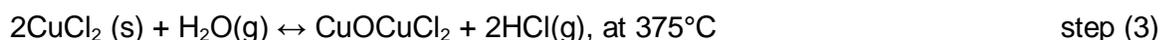
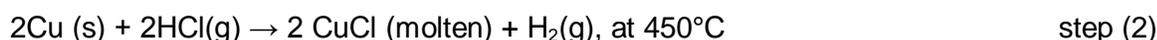
The copper-chlorine cycle is one of the most promising thermochemical cycles to produce hydrogen. In this cycle, water releases oxygen from the thermolysis of an intermediate compound. The thermolysis process is a three-phase system involving molten salt, solid reactant and gaseous oxygen. This paper examines the scale-up feasibility of the chemical reactor from the perspectives of the optimum size and number of oxygen reactors for different oxygen and hydrogen production rates. It specifically analyzes the factors contributing to the oxygen reactor size. It is shown that the reactor size is significantly influenced by residence times, mass and heat transfer.

**Keywords:** hydrogen, clean energy, scale-up feasibility

## 1. Introduction

The usage of hydrogen as a substitute for fossil fuels can mitigate the effects of climate change by reducing greenhouse gas emissions. It is widely believed that hydrogen as an energy carrier will be a significant contributing factor to the sustainable energy supply in the future [1], in spite of challenges such as the scale-up, cost and “green” method of hydrogen production. Thermochemical hydrogen production is one of the promising alternatives that may deliver a low cost and sustainable technology of decomposing water into oxygen and hydrogen thermally through a series of intermediate reactions, by using nuclear or solar thermal energy.

The thermochemical Cu-Cl cycle was identified as one of the most promising lower temperature cycles [2, 3]. This cycle consists of four reactions; three thermal and one electrochemical. A significant attraction of the Cu-Cl thermochemical cycle is its low temperature requirements that may reduce the cost of thermal energy. The four reactions of the Cu-Cl cycle are:



where aq in brackets means aqueous solution, s means solid, and g means gas.

As a summation result of the four reactions, the net input of the cycle is water, while the products coming out of the cycle are only hydrogen and oxygen, as shown in Fig. 1. It can be found that the oxygen generation reaction (step 4) has the highest temperature requirement (530°C) in the Cu-Cl cycle. Therefore, the scale-up feasibility of the oxygen production step would significantly influence the economics and industrialization of the Cu-Cl cycle. In this step, an intermediate compound, solid copper oxychloride (CuOCuCl<sub>2</sub>), is decomposed into oxygen gas and molten cuprous chloride (CuCl). To accommodate this reaction, the reactor must be a high temperature gas-solid-liquid three-phase system. In this paper, the factors that affect the scale-up of the oxygen reactor are discussed briefly. Then the mass flow streams of the oxygen production step are quantified to examine the reactor size for different oxygen and hydrogen production scales.

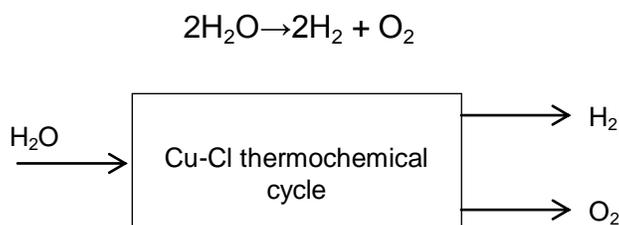


Fig. 1 Schematic of the thermochemical Cu-Cl hydrogen and oxygen production system.

## 2. Impact of H<sub>2</sub> production rates on the materials processing scales in the O<sub>2</sub> reactor

The thermochemical cycle decomposes water into oxygen and hydrogen, through intermediate copper and chlorine compounds. To evaluate the scale-up of the oxygen reactor, the dependence of the processing capabilities of the oxygen production reactor on the hydrogen production scales must be known. The dependence could be derived from the reactant mass flow rates calculated from the material balance, which is based on the fundamental material balance equation [4]:

$$\left\{ \begin{array}{l} \text{Accumulation} \\ \text{w within the} \\ \text{system(buildup)} \end{array} \right\} = \left\{ \begin{array}{l} \text{Input through} \\ \text{system} \\ \text{boundary} \end{array} \right\} - \left\{ \begin{array}{l} \text{Output through} \\ \text{system} \\ \text{boundary} \end{array} \right\} + \left\{ \begin{array}{l} \text{Generation} \\ \text{Within the} \\ \text{system} \end{array} \right\} - \left\{ \begin{array}{l} \text{consumption} \\ \text{w within the} \\ \text{system} \end{array} \right\} \quad (1)$$

which may be written in a mathematical form as:

$$\frac{dm}{dt} = \dot{m}_i - \dot{m}_o + \dot{G} - \dot{C} \quad (2)$$

where  $\frac{dm}{dt}$  denotes the rate of change of the material  $m$ , and  $\dot{G}$  and  $\dot{C}$  denote the rate of generation and consumption, respectively. Element balances have no generation or consumption terms, and the mass balance is simplified to continuous and steady-state processes wherein input always equals output. The material balance is based on the stoichiometric relationship, for example, the number of the moles of each compound must be multiplied by the stoichiometric coefficients. In this paper, continuous process operation at the steady-state will be analyzed.

Using the stoichiometric material balance for the overall system shown in Fig. 1, the mass flow rates of H<sub>2</sub>O and O<sub>2</sub> can be calculated as follows:

$$\dot{m}_{H_2O} = \dot{n}_{H_2O} M_{H_2O} \quad (3)$$

Since  $\dot{n}_{H_2O} = \dot{n}_{H_2}$  and  $\dot{n}_{H_2O} = \frac{\dot{m}_{H_2}}{M_{H_2}}$ , then Eq. (3) becomes:

$$\dot{m}_{H_2O} = \dot{m}_{H_2} \frac{M_{H_2O}}{M_{H_2}} \quad (4)$$

For the mass flow rate of O<sub>2</sub>,

$$\dot{m}_{O_2} = \dot{n}_{O_2} M_{O_2} \quad (5)$$

Since  $\dot{n}_{O_2} = \frac{1}{2} \dot{n}_{H_2O}$ ,  $\dot{n}_{H_2O} = \dot{n}_{H_2}$  and  $\dot{n}_{H_2} = \frac{\dot{m}_{H_2}}{M_{H_2}}$ , then Eq. (5) will be:

$$\dot{m}_{O_2} = \frac{1}{2} \dot{m}_{H_2} \frac{M_{O_2}}{M_{H_2}} \quad (6)$$

Fig. 2 shows a linear relationship between the flow rate of each component of the Cu-Cl cycle (in the case of complete conversion of the components) and the hydrogen production rate.

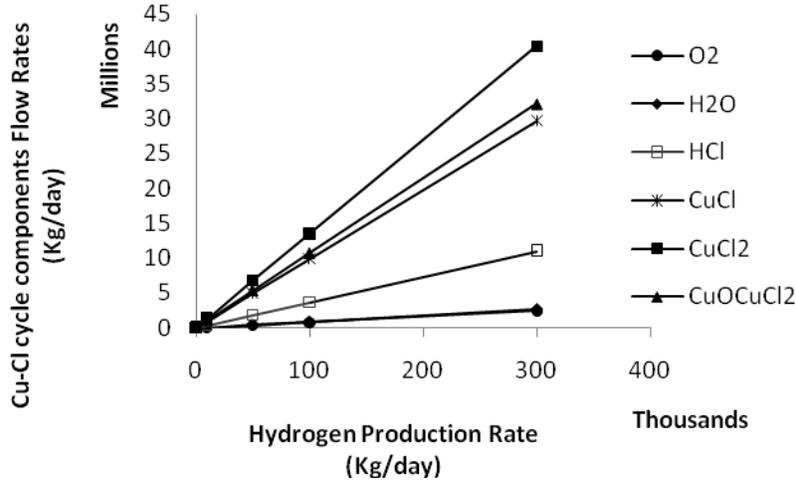


Fig. 2 Flow rates of Cu-Cl cycle mass streams versus various hydrogen production rates.

To estimate the height of the oxygen production reactor, the liquid level of molten CuCl salt must be known, which is determined by the volume of molten salt, gas bubbles, and unreacted reactants. The minimum volume of the oxygen reactor is the total volume of the molten salt and unreacted reactants if the volume of gas bubble is neglected. The volume of the solid CuOCuCl<sub>2</sub> and molten CuCl can be calculated as:

$$V_{CuOCuCl_2} = \frac{m_{CuOCuCl_2}}{\rho_{CuOCuCl_2}} \quad (7)$$

$$V_{CuCl} = \frac{m_{CuCl}}{\rho_{CuCl}} \quad (8)$$

The data needed to calculate the volume of the solid CuOCuCl<sub>2</sub> and the molten CuCl, in addition to the minimum total volume of the oxygen reactor are given in Table 1. The bulk density of CuOCuCl<sub>2</sub> is used for the calculation, because the voids always exist among the solid CuOCuCl<sub>2</sub> particles even as all the solid particles are compacted together for the conveyance from the hydrolysis unit to the oxygen production reactor. The bulk density depends on the particle size and structure which influences the volume of the reactor, feeder and discharge vessels. The structure and shape of solid CuOCuCl<sub>2</sub> are formed in the upstream CuCl<sub>2</sub> hydrolysis reactor (step 3). The reasons for using the bulk density instead of the actual density for CuOCuCl<sub>2</sub> include [5]: 1) lack of actual density data of CuOCuCl<sub>2</sub>; 2) the bulk volume is close to the practical conveying and packaging state; 3) the bulk density is a more conservative value to avoid space deficiency of equipment.

Property	Value
Bulk density of $\text{CuOCuCl}_2$ [15]	4.853 g/cm <sup>3</sup>
Actual density of molten $\text{CuCl}$ at 430 °C [5]	3.692 g/cm <sup>3</sup>
Actual density of solid $\text{CuCl}$ at 25 °C [5]	4.14 g/cm <sup>3</sup>
Melting point of $\text{CuCl}$ [5]	430 °C

Table 1. Properties for oxygen reactor size analysis

For a hydrogen production rate of 100 tons/day, the processing volumes per day of the solid  $\text{CuOCuCl}_2$  and molten  $\text{CuCl}$  can be calculated from Eqs. (7) and (8) as:

$$V_{\text{CuOCuCl}_2} = \frac{m_{\text{CuOCuCl}_2}}{\rho_{\text{CuOCuCl}_2}} = \frac{10,700,000}{4.853 \times 1000} = 2204.82 \text{ m}^3/\text{day} \quad (9)$$

$$V_{\text{CuCl}} = \frac{m_{\text{CuCl}}}{\rho_{\text{CuCl}}} = \frac{9,900,000}{3.692 \times 1000} = 2681.47 \text{ m}^3/\text{day} \quad (10)$$

The volume ratio of the molten salt product to the solid reactant is:

$$\frac{V_{\text{CuCl}}}{V_{\text{CuOCuCl}_2}} = \frac{2681.47}{2204.82} = 1.216 \quad (11)$$

As indicated by Eqs. (9-11), if the molten salt is not removed from the reactor immediately after it is produced, its volume is larger than that of the solid reactant by a factor of 1.216, which is larger than 1. Thus, in the design of an oxygen production reactor, as shown in Fig. 3, the molten  $\text{CuCl}$  can be prefilled into the oxygen reactor before the reaction is started. Once the reaction takes place and the molten salt of  $\text{CuCl}$  is produced, the produced  $\text{CuCl}$  overflows to the outside of the reactor, and the addition of the solid reactant of  $\text{CuOCuCl}_2$  also causes the overflow of the molten  $\text{CuCl}$ . In this design, the reactor volume can be represented by the prefilled volume of the molten  $\text{CuCl}$ , which is assumed to be proportional to the processing scale of the molten  $\text{CuCl}$ .

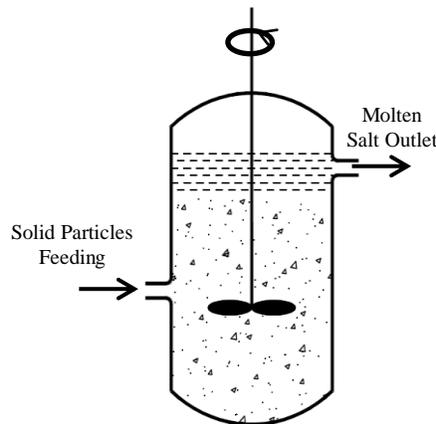


Fig. 3 Schematic diagram of CFSTR.

### 3. Impact of residence time and aspect ratio on the O<sub>2</sub> reactor size and number

In addition to the processing scale discussed above, the reactor volume is also significantly influenced by the residence time, which is the time the reactant stays in the reactor. In this paper, it is the time required by the solid CuOCuCl<sub>2</sub> reactant to decompose completely into the products of molten CuCl and oxygen gas. The reactor volume becomes smaller when the reaction time decreases. It is necessary to know the residence time in order to calculate the volume of the oxygen reactor. Michele et al. [6] have studied the oxygen generation reaction by real-time monitoring of the composition of the effluent gas using a mass spectrometer. A sample of laboratory-produced CuOCuCl<sub>2</sub> was heated to 530 °C and the mass spectrum that showed the evolution of oxygen was obtained. By integrating the area under the peak of oxygen evolution with time, they calculated that the amount of oxygen recovered was equal to 100% ± 10% (experimental uncertainty) of theoretical data, based on the initial amount of CuOCuCl<sub>2</sub>. It was shown from the mass spectrum that the total time required for 100% of the oxygen release was 25 minutes. This time is considered as the minimum residence time in this paper.

To be more conservative in the design of the oxygen reactor, the residence time that is considered in this paper also includes 1 hour and 2 hours. For a residence time of 2 hours, the total volume of the reactor is:

$$V_{total} = \frac{2681.47 \times 2}{24} = 223.45 \text{ m}^3 \quad (12)$$

This result for an industrial scale oxygen reactor is reasonable in engineering, because examples of chemical reactors that have comparable values of volumes for the industrial microbial fermenters are in the range of 100-300 m<sup>3</sup> [7]. Table 2 shows the reactor volumes for different hydrogen production rates and different residence times. As shown in Table 2, the volume of the reactor increases because the volumes of the solid CuOCuCl<sub>2</sub> and the molten CuCl increase with a larger hydrogen production rate.

Residence Time (hr)	H <sub>2</sub> (Kg/Day)	V <sub>CuOCuCl<sub>2</sub></sub> (m <sup>3</sup> )	V <sub>CuCl</sub> (V <sub>total</sub> ) (m <sup>3</sup> )	$\frac{V_{CuCl}}{V_{CuOCuCl_2}}$
2	100,000	183.70	223.46	1.216
	50,000	91.87	111.73	1.216
	25,000	45.93	55.86	1.216
1	100,000	91.87	111.73	1.216
	50,000	45.93	55.86	1.216
	25,000	22.97	27.93	1.216
0.5	100,000	45.93	55.86	1.216
	50,000	22.97	27.93	1.216
	25,000	11.48	13.97	1.216

Table 2. Oxygen reactor volumes for different hydrogen production rates and different residence times

The normal configuration for a process vessel is a vertical cylindrical section closed by dished ends [8]. On the basis of the reactor volume, the reactor height and diameter can be determined if the height-to-diameter ratio (aspect ratio, AR) is known. This ratio is also essential to specifying the reactor's agitator specifications, if an agitator is needed. Reactors with low AR can have restrictive recirculation patterns and those with high AR can increase the path length for recirculation. A high AR presents several disadvantages in the mixing of the oxygen reactor which can affect the heat transfer rate inside the reactor. Also, the design of a longer shaft can be more difficult.

Multiple reactors may be needed if the processing scale is very large and an extremely large reactor is not practical. When limiting the size of oxygen reactor to a specified volume, the number of reactors can be calculated as:

$$N = \frac{V_{total}}{\frac{\pi}{4} D^2 H} \quad (13)$$

Table 3 lists the required reactor size and number for different residence times and different hydrogen and oxygen production scales. For an industrial hydrogen production scale of 100 tons/day and a residence time of 2 hr, the reactor diameter varies in the range of 1 – 4 m, the height varies from 2 to 9 m, and the number of the reactors spans a wide range of 2 – 142. The reactor diameter influences significantly the number of reactors. If the reactor diameter is in the range of 3-4 m, then the number of reactors will be in the range of 2-5 for an industrial hydrogen production scale of 100 tons/day and a residence time of 2 hr.

As reported by Oldshue, (1983), for each type of chemical process, there is not only one optimal or unique design. Several designs may satisfy the process specifications [9]. In this paper, the diameter of 3-4 m is recommended for the upper limit of the scale-up of the oxygen reactor.

Another important parameter to be optimized in the scale-up of the reactor is the aspect ratio. Aspect ratio is a design parameter and there are several issues involved in selecting an aspect ratio. The aspect ratio may influence the capital cost of the reactor, the mixing quality, and the heat transfer. Regarding the capital cost, generally it significantly depends on the amount of required equipment materials, e.g., high quality metals. The weight of metal required to build a reactor of a volume is minimized by using an aspect ratio of about 1 [9]. However, there are also other factors that can make a larger aspect ratio more economical. The issues include manufacturing considerations of mechanical stress, the thermal shock performance, and the pressure inside the reactor.

As for the mixing quality, it often becomes more challenging to achieve a good mixing quality as the aspect ratio increases to above 2 [10]. It has been found that for solid suspension and blending, the optimum ( $L_{liquid}/D$ ) ratio for minimum power consumption should be equal to 0.6 to 0.7 [9]. Walas et al. (1990) indicated that when a single impeller is to be used, a liquid level equal to the diameter is optimum, with the impeller located at the center for an all-liquid system. Economic and manufacturing considerations, however, often prefer higher ratios of liquid depth to diameter [11]. Coker et al. (2001) reported that the optimum proportions for stirred tank reactors occur when the liquid level is equal to the tank diameter, but at high pressures slimmer proportions are more economical [12].

Residence Time (hrs)	H <sub>2</sub> (Kg/Day)	O <sub>2</sub> (Kg/Day)	$D(m) \times H(m) \times N$
2	100,000	800,000	1x2x142 1.99x3.98x18 3.05x6.11x5 4.14x8.29x2
	50,000	400,000	1x2x71 1.99x3.98x9 2.87x5.75x3 4.14x8.29x1
	25,000	200,000	1x1.99x36 2.07x4.14x4 3.29x6.58x1
1	100,000	800,000	1x2x71 1.99x3.98x9 2.87x5.75x3 4.14x8.29x1
	50,000	400,000	1x1.99x36 2.07x4.14x4 3.29x6.58x1
	25,000	200,000	1x1.99x18 2.07x4.14x2 2.61x5.22x1
0.5	100,000	800,000	1x1.99x36 2.07x4.14x4 3.29x6.58x1
	50,000	400,000	1x1.99x18 2.07x4.14x2 2.61x5.22x1
	25,000	200,000	1x1.99x9 2.07x4.14x1

Table 3. Dimensions and number of oxygen reactors (neglecting bubble volume) for different hydrogen production rates and different residence times.

The aspect ratio also influences the heat transfer. It was reported that if the aspect ratio is larger, then the heat transfer area is larger for a given volume. This would improve temperature control [10]. For example, the 200 m<sup>3</sup> reactor would have a diameter of 3.99 m and a heat transfer area of 200 m<sup>2</sup> if an aspect ratio of 4 was used. By comparison, using an aspect ratio of 10, gives a diameter of 2.94 m and a heat transfer area of 271.5 m<sup>2</sup>. If the aspect ratio is 2, then the heat transfer area is 159 m<sup>2</sup>.

A reasonable engineering compromise of these competing effects is recommended and an aspect ratio of 2 is frequently used for chemical reactors [10]. In this paper, therefore,  $H = 2D$  is recommended for the scale-up design of the oxygen reactor.

### 4. Other factors influencing the scale-up of the oxygen reactor

**(1) Reactant particle size and structure.** The reactor height depends significantly on the descent speed of particles in the molten salt. Despite the fact that the molten CuCl density is smaller than that of the solid CuOCuCl<sub>2</sub> particles [13], there are other factors that can affect the descent speed of particles in a stagnant liquid, such as the size and shape of particles and the viscosity of the molten CuCl. In the oxygen reactor, the descent speed of particles is reduced by the formation of bubbles on the surface of the solid particles. A small descent speed of particles is recommended so as to provide enough residence time for the complete decomposition of particles before reaching the bottom of the reactor. In this way, the height of the reactor can be reduced.

**(2) Feeding temperature.** In the Cu–Cl cycle, the reactant CuOCuCl<sub>2</sub> is a product from the upstream hydrolysis reactor (step 3) that operates at a temperature of 350–450 °C. The oxygen reactor operation can be more efficient if the lowest temperature of the solid particles is equal or higher than the melting point of CuCl, i.e., 430 °C. If the temperature of the entering solid particles is lower than 430 °C, the molten CuCl may be solidified on the surface of the solid particles. This may lead to the confinement of oxygen bubbles and aggregation of the solid particles, molten salt and gas bubbles. If the aggregation of the particles is developed, this may further lead to the floating of aggregations on the surface of the molten salt, and also lead to the reduction of the contact area between the solid particles and molten salt. As a consequence, the decomposition of the reactant particles may be deterred and potentially cause a choking of the reactor, which is a major safety concern. This challenge can be overcome in different ways such as hot feeding of the solid particles, using agitation or protective mixing with hot oxygen (one of the decomposition products) [14].

When using hot feeding, there will be no aggregation of solidified CuCl and entrained bubbles if the temperature of feeding particles is slightly higher than the melting point of CuCl. In this way, the temperature of the liquid surface will be lower than the bulk liquid and this will cause a difference in density within the molten salt bath. The density difference will improve the convection, mixing and flow of liquid [6]. However, this process needs an extra heat exchanger unit before the oxygen reactor unit, to heat the feeding reactant solid.

**(3) Oxygen bubbles.** Oxygen bubbles are generated on the surface of the reactant solid particles of CuOCuCl<sub>2</sub>. Because of the large density difference between oxygen and molten CuCl, these bubbles will leave the particle surface after they are generated. The bubbles will move to the liquid (molten salt) surface, and once they reach the liquid surface they will be removed from the reactor immediately. During the ascent of the bubbles to the liquid level, the size of the bubbles will grow and hence create a required portion of the reactor volume. According to past data [6], the maximum volume fraction of the bubbles is about 15% of the total volume at a life time of 12 minutes.

**(4) Reactor heating rate.** In the oxygen reactor, the decomposition of CuOCuCl<sub>2</sub> to oxygen and molten CuCl is an endothermic reaction that requires a reaction enthalpy of 129.2 KJ/mol and a temperature of 530 °C [5]. Thus, heat must be added to raise and maintain the temperature of the bulk phase inside the reactor. The total amount of the required heat is the sum of the reaction enthalpy and the heat needed to raise the reactant temperature from 375 °C to 530 °C. To provide this amount of heat, a sufficiently high heat transfer rate must be provided inside the reactor.

**(5) Agitation.** Agitation may increase the residence time of  $\text{CuO}(\text{CuCl}_2)$  particles in the molten salt. In this operation, there is no need to decrease the descent speed of particles or increase the height of the reactor. Also, agitation mixes the multiple components well and helps break the aggregation of  $\text{CuO}(\text{CuCl}_2)$  particles and solidified  $\text{CuCl}$ , thus increase the contact area between the particles and molten salt. In addition, agitation will increase the local heat transfer rate by introducing forced convection and increasing the contact area between the  $\text{CuO}(\text{CuCl}_2)$  particles and molten  $\text{CuCl}$ . However, it is unclear if the agitator would increase or decrease the size of the reactor, because the size of the agitator and its auxiliary components may offset its effects of decreasing the reactor volume.

A continuous flow stirred tank reactor (CFSTR) could be a suitable type for oxygen reactor. Generally, a CFSTR as shown in Fig. 3 often assumes an isothermal operation (i.e., constant temperature), steady state condition, and constant liquid density [12]. Then in the reactor, the fluid can be considered as perfectly mixed, so that the properties (e.g., concentration and temperature) of the reaction mixture are uniform in all parts of the system. Therefore, the conditions throughout the reactor are the same and the temperature throughout the reactor is equal to that at the reactor outlet.

## 5. Conclusion

In this paper, the factors influencing the scale-up of the oxygen reactor in the Cu-Cl thermochemical hydrogen production cycle were examined. The size of oxygen reactor was estimated on the basis of the hydrogen production scale, mass balance, residence time, and aspect ratio, among others. The diameter range of 3-4 m and the aspect ratio of 2 are recommended for an industrial hydrogen production scale of over 100 tons/day.

## 6. Nomenclature

$D$	Reactor diameter (m)
$H$	Reactor height (m)
$m$	Mass (Kg)
$M$	Molecular weight (g/mol)
$\dot{m}$	Mass flow rate (Kg/sec)
$N$	Number of reactors
$\dot{n}$	Molar flow rate (mol/sec)
$V$	Volume ( $\text{m}^3$ )
$\rho$	Density ( $\text{Kg}/\text{m}^3$ )

## 7. References

- [1] Forsberg, C.W., "Future hydrogen markets for large-scale hydrogen production systems", *International Journal of Hydrogen Energy*, Vol. 32, 2007, pp. 431-439.
- [2] Lewis, M A, Serban, M and Basco, J K., "Generating hydrogen using a low temperature thermochemical cycle", *Proceedings of the ANS/ENS 2003 Global International conference on Nuclear Technology*, New Orleans, 2003.

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- [3] Serban, M, Lewis, M A and Basco, J K., "Kinetic study of the hydrogen and oxygen production reactions in the copper-chloride thermochemical cycle", *AIChE Spring National Meeting*, 2004, pp. 2690-2698.
- [4] Ghasem, Nayef and Henda, Redhouane, "Principles of chemical engineering processes", *CRC Press*, 2009.
- [5] Wang, Z, Naterer, G F and Gabriel, K., "Multiphase reactor scale-up for cu-cl thermochemical hydrogen production", *International Journal of hydrogen Energy*, Vol. 33, 2008, pp. 6934-6946.
- [6] Michele, A Lewis, Joseph, G Masin and Richard, B Vilim, "Development of the low temperature cu-cl thermochemical cycle", *International Congress on Advances in Nuclear Power Plants*, Seoul, Korea, 2005.
- [7] Moo-Young, M. and Chisti, Y., "Biochemical engineering in biotechnology", *International Union of Pure and Applied Chemistry, Great Britain (Technical Report)*, *Pure & Appl. chem.*, Vol. 66, 1994, pp. 117-136.
- [8] Nauman, Bruce, "Chemical reactor design, optimisation and scale-up", *McGraw-Hill*, New York, 2002. pp. 153-159.
- [9] Oldshue, J. Y., "Fluid mixing technology", *McGraw-Hill*, 1983.
- [10] Luyben, William L., "Chemical reactor design and control", *John Wiley & sons, Inc., Publication*, New Jersey, 2007.
- [11] Walas, Stanley M., "Chemical process equipment - selection and design", 1990.
- [12] Coker, A. Kayode, "Modeling of chemical kinetics and reactor design", *Gulf Professional Publishing*, 2001.
- [13] Roberts, WL, et al., "Encyclopedia of minerals", *2nd. Van Nostrand Reinhold*, 1990.
- [14] Naterer, G F, et al., "Thermochemical hydrogen production with a copper-chlorine cycle. I: oxygen release from copper oxychloride decomposition", *International Journal of Hydrogen Energy*, Vol. 33, 2008, pp. 5439-5450.
- [15] Michele, Lewis A, Manuela, Serban and Basco, John K., "Hydrogen production at <550°C using a low temperature thermochemical cycle", *Nuclear Production of Hydrogen*, Argonne, 2003, pp. 145-156.