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Mechanical and Aerospace Engineering

**Design of the Dual Circulating Fluidized Bed
Prototype for Chemical Looping Combustion
System**

by

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Declaration

We hereby declare that this report entitled “Design of the Dual Circulating Fluidized Bed System Prototype for Chemical Looping Combustion System” is the result of our own project work except for quotations and citations that have been duly acknowledged. We also declare that it has not been previously or concurrently submitted for any other degree at Nazarbayev University.



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Abstract

The aim of this project is to design, build, and test a lab-scale cold flow model of a dual circulating fluidized bed (DCFB) for application in chemical looping combustion (CLC) systems. Chemical looping combustion is a potential carbon capture and storage (CCS) technique that can mitigate CO₂ emissions and achieve efficient fuel conversion. However, it is complicated to investigate the hydrodynamic characteristics of solid and gas mixture in large-scale CLC systems because of high temperature and gas-solid dynamics. Therefore, to solve this problem, a cold flow system working under ambient conditions was designed and developed to study the critical hydrodynamic variables like pressure drop and solid circulation rate (SCR) profiles.

Stable circulation was achieved at AR and FR velocities of 1.86 m/s and 1.23 m/s, respectively. The results showed that the pressure drop decreased with increasing geometrical height, from 5.2 to 1.28 kPa in the air reactor (AR) and from 2.22 to 0.39 kPa in the fuel reactor (FR), reflecting the transition from dense to dilute particle regions. In addition, increasing the air flow rate in the AR from 67 to 90 m³/hr significantly affected the SCR, corresponding to a more than 200% rise. Increasing the solid inventory from 55 kg to 65 kg also enhanced circulation stability and increased the SCR in both reactors.

Lastly, Computational Fluid Dynamics (CFD) simulations were developed to investigate the gas-solid flow dynamics and compare pressure drop profiles with experiments. Conclusions regarding the pressure decrease with the height of the reactors were validated with CFD simulations. Differences between CFD and experimental pressure values may be attributed to real operating conditions, simplifications in the numerical model, and limitations in drag coefficient correlations used to describe gas-solid interactions. The obtained results provided valuable information that can be used to optimize and scale up CLC systems in the future.

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1 Introduction and Problem Statement

Rising global population and economic advancements are the primary drivers behind the growing worldwide demand for energy. Currently, the world's energy mix consists of coal, gas, oil, hydropower, biofuels, solar, wind and nuclear. Despite the abundance of options, over 80% of the worlds' energy supply comes from fossil fuels. As can be seen in Figure 1, the overwhelming majority of the power produced in developed countries is provided by fossil fuels, followed by renewables and nuclear energy [1, 14].

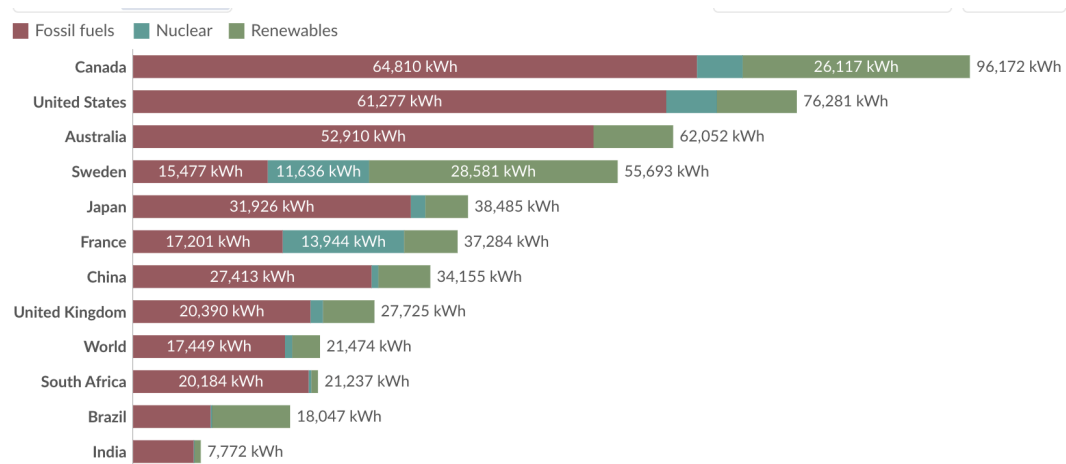


Figure 1. Energy from fossil fuels, nuclear, and renewables per capita in 2024 [14].

The map in Figure 2 further confirms the statement above, demonstrating a high concentration of coal, oil, and gas in the world's energy mix.

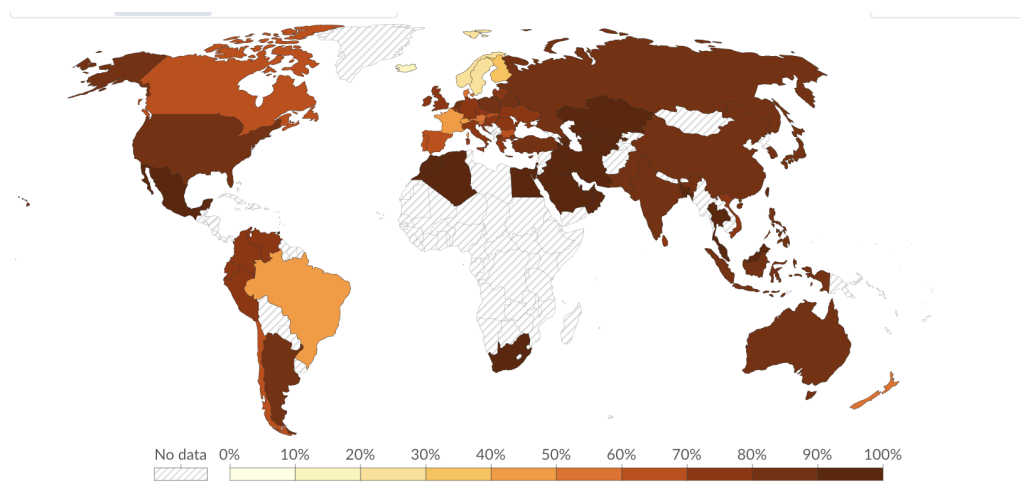


Figure 2. Share of primary energy consumption from fossil fuels in 2024 [14].

It is a well-established fact that fossil fuels emit greenhouse gas (GHG) emissions during their combustion. The burning of oil, coal, and gas for energy purposes accounts for approximately $\frac{3}{4}$ of global GHG emissions. By the year 2100, a 5°C increase in Earth's surface temperature is expected relative to 1990 due to the GHG effects and their release into the atmosphere [14, 21]. Despite their impact on the environment, fossil fuels are still projected to remain the leading source of energy in the near future. Thus, it becomes crucial to develop carbon capture and storage (CCS) technologies that will minimize the carbon footprint of power generation. Earlier, oxy-fuel combustion, post-combustion, and pre-combustion were found as three main methods that have been developed for CO₂ capture in industrial applications and power generation. While post-combustion and pre-combustion technologies are feasible, they suffer from significant energy penalties and high capital costs due to the necessity of complex gas separation units [1]. On the other hand, chemical looping combustion (CLC), a new type of oxy-fuel technology, provides the cheapest and most efficient alternative for these technologies. It is a promising clean energy technology that enables the conversion of hydrocarbon fuels with significantly reduced CO₂ emissions. In some configurations, CLC was shown to have almost ideal efficiency of 100% [2]. In this system, fuel is combusted without direct contact with air. It is achieved through oxygen carriers (OCs), typically a metal oxide, which are used to transfer oxygen between the air and fuel reactors to initiate combustion [5].

According to Czakiert et al. (2022), OCs are the major components in the CLC process that play a principal role in transporting oxygen between the air and fuel reactors, acting like red blood cells. As displayed in Figure 3, the system consists of two interconnected reactors; an oxygen carrier gets oxidized in the air reactor (AR) by fluidizing medium (air), and transfers to the fuel reactor (FR), where it gets reduced during combustion. After being reduced to their metal or less oxidized state, metal oxides return to the AR, where they get reoxidized. This cyclic process repeats continuously, forming a circulation loop within the CLC system. As a result, a concentrated stream of CO₂ free of nitrogen is produced for further utilization, transportation, and storage. In comparison with alternative carbon capture methods, CLC possess an inherent CO₂ capturing capability with no energy penalty, thereby maintaining the effectiveness of fuel combustion [5].

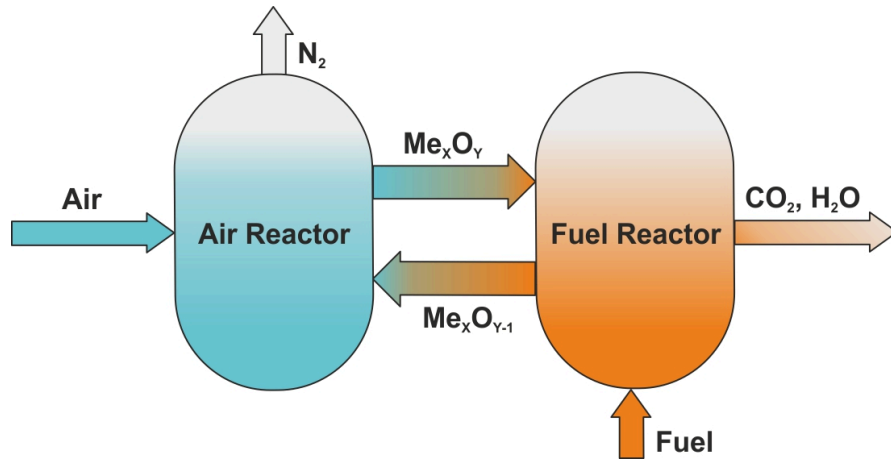


Figure 3. Chemical looping combustion cycle [5].

Despite the promising future and high potential of CLC in contributing to reduced industrial carbon emissions, the widespread deployment of this technology is hindered by several issues [3, 13]. The primary challenge depicted in this capstone project is the complexity of designing and predicting the performance of Dual Circulating Fluidized Beds (DCFBs) since reactors operate at high temperatures of 1000 °C. The second issue is the difficulty of reaching a steady state of DCFBs at once: AR is exothermic, FR is endothermic. These conditions make it challenging to measure and study hydrodynamics of solid particles in an industrial scale hot plant. The study of hydrodynamics has gained a great attention from researchers since it is an essential part of DCFBs technology. It is important to understand solid residence time, gas-solid contact efficiency, solid flux, reactor geometry, and heat and mass transfer to design an optimized and efficient model. Therefore, cold flow models (CFM) are implemented to examine the hydrodynamic behavior of the particles under ambient temperature and pressure. Scaled-down laboratory versions are assembled to resemble the hydrodynamics of a larger hot rig and can be applied to hot models by using scaling relationships (Glicksman scaling law). Easy maintenance, possibility of visual observation, minimal requirements for equipment and control instrumentation make CFMs a convenient method of gathering fluid dynamic information [17]. Thus, this capstone project aims to assemble a cold flow model to study the hydrodynamics of solid particles in a chemical looping combustion system. The main objectives of the capstone project are to:

- Design and assemble a laboratory-scale cold rig model to conduct experiments and obtain relationships between input and output variables;

- Achieve maximum steady state operation of solid circulation in the DCFBs system;
- Develop a CFD model to simulate gas-solid flow behavior and validate experimental results.

Safety and research ethics are crucial in this project to guarantee system safety and experimental results. The project deals with high-pressure air systems, particle characterisation, and laboratory-scale fluidized beds, so all experiments should be conducted in a laboratory environment. Sufficient personal protective equipment, ventilation, and monitoring of air flow and pressure during operation was implemented. Particular care was also taken with the stability of the system because high gas flow rates and pressure inside the reactors may lead to a higher possibility of accidents. Moreover, the project should adhere to the occupational health and safety principles outlined in International Organization for Standardization ISO 45001, which is concerned with risk prevention, safe and healthy working conditions, and safety management in engineering projects [29]. Equipment maintenance and safe disposal of waste should also be practised to ensure a safe environment.

The project will aim to promote cleaner energy technologies that can help mitigate greenhouse emissions and enhance carbon capture in terms of environmental and research ethics. When conducting experiments, one should avoid excessive use of materials, reduce energy consumption whenever possible, and prevent particles or other substances from contaminating the environment. It is also important to ensure data integrity and measurement accuracy, and to document the methodology used in order to obtain reliable results and ensure the integrity and quality of the research.

2 Literature Review

The literature review chapter was divided into two parts: CFM and oxygen carriers. Experimental articles regarding existing CFMs were analyzed to identify major system components, layout, and the dimensioning used by researchers. Moreover, a comprehensive review of published literature was conducted to gather and analyze information about oxygen carriers, including their properties and key characteristics.

2.1 Cold Flow Model

Pröll et al. (2009) constructed a scaled model of the 120 kW hot rig, applying a linear scaling factor of 1:3. A design shown in Figure 4 was implemented, where two cyclones are connected to their respective reactors. The diameters of the AR and the FR are given to be equal to 0.05 m and 0.054 m, respectively. The AR is about 1.6 m tall, while FR is nearly 1.2 m in height. Lower and upper loop seals are used to provide connections between two fluidized beds and beds with cyclones [13].

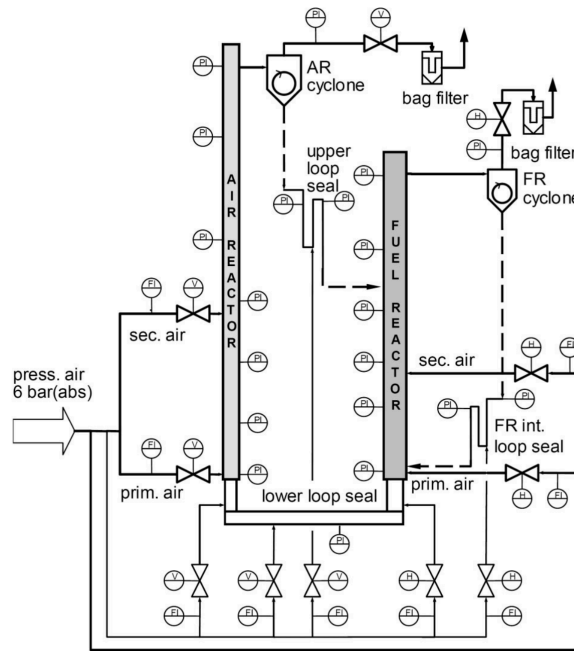


Figure 4. CFM executed by Pröll et al. [13].

The air flow was regulated by flowmeters placed in various locations, followed by manual control valves. U-shaped manometers were used to measure the pressure, some of which had a liquid flow resistance to handle abrupt pressure changes. Bronze was implemented as a solid bed material with an average particle diameter of 50 μm , and air was used as a

fluidizing medium. The researchers conducted experiments with three different bed masses of 4, 4.5, and 5 kg. The air flow to the AR was altered from 50% to 125% of the initial 30.3 m³/hr, whereas the flow rate in FR was varied between three values (5, 10, 15 m³/hr). By changing the air velocity in the AR and FR, the authors aimed to analyze its effect on the solid circulation and pressure profiles. Solid flux is a critical indicator of system performance in circulating fluidized beds, as it determines the efficiency of particle transport between reactors. Moreover, they supplied air to the lower, upper, and internal loop seals, consequently, the influence of fluidization of loop seals on the system was investigated [13].

Figure 5a) illustrates the graph for solid flux in AR plotted against air flow in the same reactor. Across all bed mass conditions, solid flux increased with increasing loop seal velocity and air flow rate to the AR, indicating a strong dependence of particle circulation on gas flow conditions. Moreover, the same relationship can be observed with solid inventory mass; for the same air supply, 5 kg of bronze, which was the maximum weight loaded, showed better SCR than 4 and 4.5 kg [13].

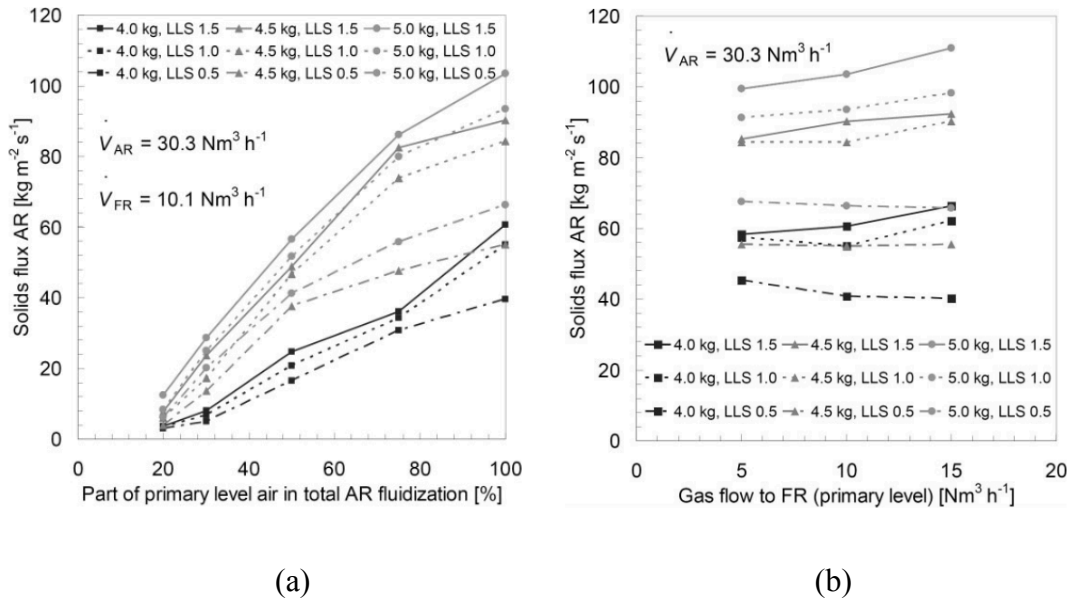


Figure 5. a) Solid flux in AR vs. gas flow in AR with dependence on lower-loop seal's fluidization and mass of solid materials; b) Solid flux in AR vs. gas flow in the FR with dependence on lower-loop seal's fluidization and mass of solid materials [13].

Additionally, the authors investigated the effect of air flow rate change in one reactor on the solid flux of the other one. As seen in Figure 5b), increasing the air supply in the FR had negligible effect on the solid flux in the AR [13].

Pressure profiles of reactors were investigated to study solids holdup and flow behavior. For instance, pressure in both reactors decreased with increasing geometric height, reflecting a transition from a dense bottom region to a more dilute upper region (Figure 6a)).

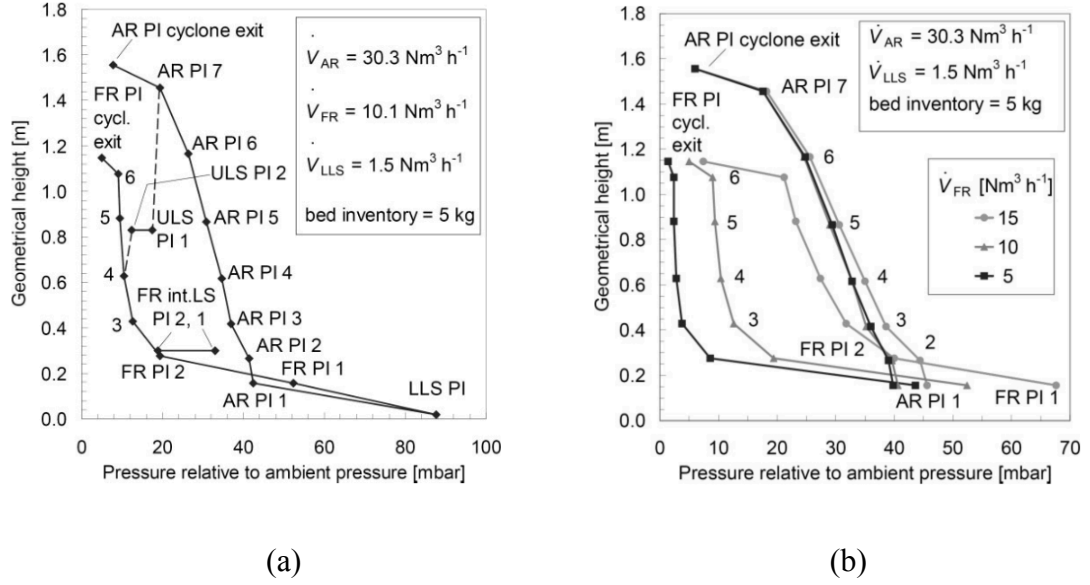


Figure 6. a) Pressure profiles in both reactors at standard operating conditions; b) Pressure profiles in both reactors with increasing flow rate in FR [13].

Furthermore, an increase in the flow rate in the FR led to a considerable increase in pressure within the FR, with only a minor effect observed in the AR (Figure 6b)). The latter can be attributed to the solids being increasingly pushed through the lower loop seal with higher pressure in FR. Also, with growing bed material mass, the pressure required to move metal oxides between reactors increases, resulting in high particle concentration and elutriation in the AR [13].

On the other hand, Shuai et al. (2011) simulated a 2D CLC process consisting of two interconnected reactors on CLC-CFD software, with a high-velocity AR and a low-velocity FR. The Eulerian continuum two-fluid model used in the study treats the solid and gas phases as interpenetrating continua. Kinetic and frictional stresses, drag, and boundary conditions were considered while making simulations. The CFM was designed to scale a 30 MW hot unit (using Glicksman scaling rules) running at 1000 °C and 9 bar. The model had only one cyclone, and the FR was considerably smaller than the AR (see Figure 7). The AR and the FR have the same diameter of 0.19 m and heights of 1.9 m and 0.5 m, respectively. Additionally, the authors utilized a pot-seal as a link and to prevent gas mixing between the reactors. The pot-seal measures 0.15 m in height and 0.14 m in diameter. For the

experiments, 9 kg and 15 kg of silica sand were tested, which featured an average particle size of 150 μm . Gas velocity in the AR, pot-seal, and FR were 1.1, 0.088, and 0.19 m/s, respectively; the AR had the highest velocity, which drove the system's particle circulation. Because the components are directly connected, gravity causes particles to travel from the cyclone to the FR [16].

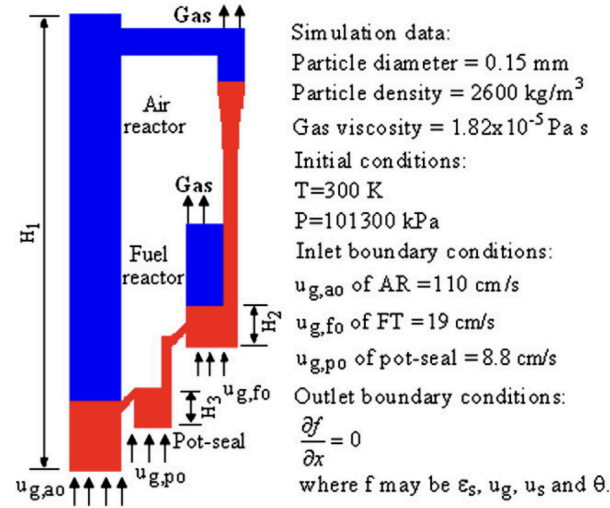


Figure 7. CFM done by Shuai et al. [16].

The following instantaneous images were obtained of the system, where the solid concentration decreased with rising geometrical height (Figure 8).

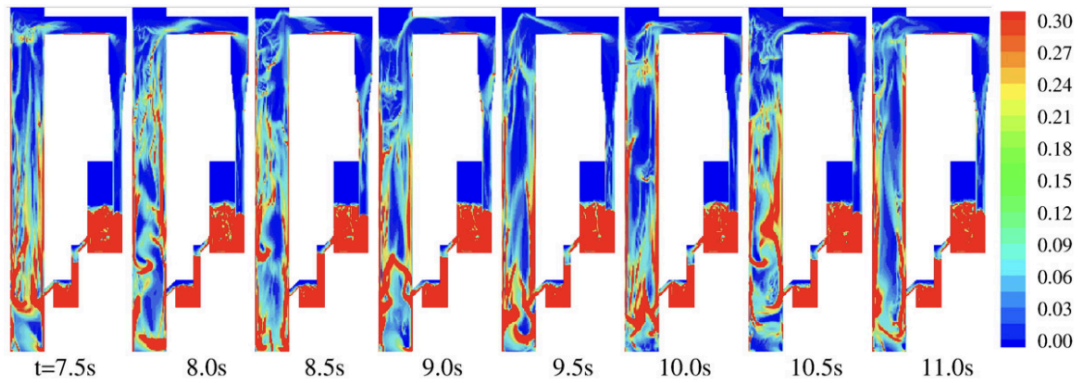


Figure 8. Instantaneous solids concentration contour in the CLC reactor [16].

Reducing gas leakage in the system was the researchers' primary goal. Since CO_2 leakage to the AR and FR directly impacts the system's carbon capture efficiency, it is a crucial factor to take into account when designing a rig. From their simulations, authors pointed out that the degree of gas leakage is influenced by the air velocity in the pot-seal. Figure 9a) displays

that the percentage of gas that leaks from the pot-seal to the FR increases as the pot-seal's velocity increases [16].

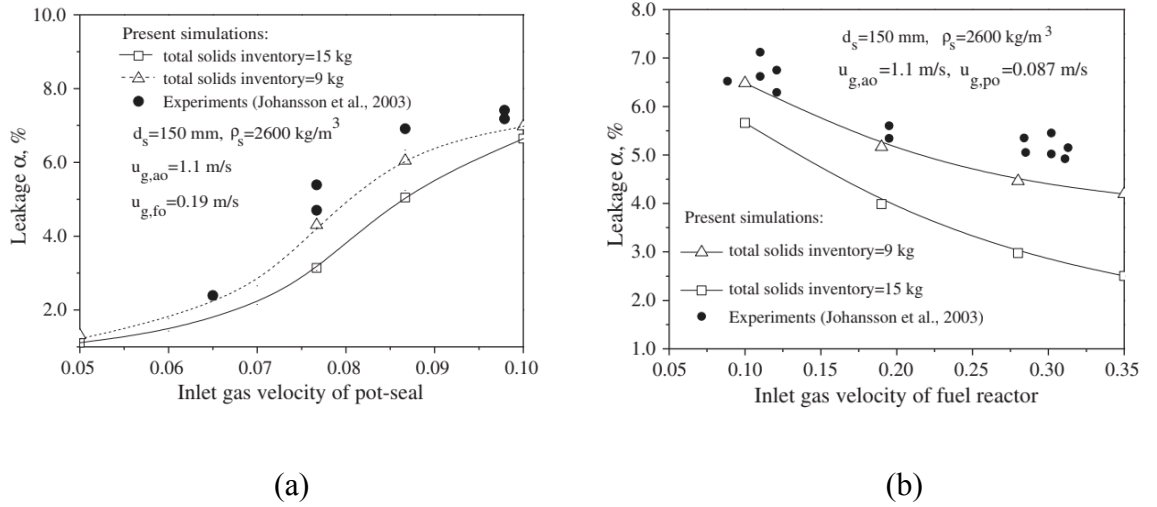


Figure 9. a) Leakage from the pot-seal to FR vs. gas velocity in the pot-seal; b) Leakage from the pot-seal to FR vs. velocity in FR [16].

Nevertheless, this can be mitigated by either adding more solid material to the system or raising the gas velocity in the FR (Figure 9b)). Moreover, the researchers concluded that properly managing the solid inventory ensures a particle seal in the diplegs, minimizing gas leakage [16].

Another aspect discussed in the article is bubble dynamics. It is also largely influenced by the solid bed mass: the higher inventories result in larger bubble formation and coalescence in the FR and pot-seal, which enhances the circulation in the system [16].

The study by Bischi et al. (2011) investigates the hydrodynamic performance of a full-scale CFM of a second-generation CLC reactor system. The experimental setup consists of a double loop circulating fluidized bed (DLCFB) with two interconnected reactors, which slightly differs from the systems described in previous studies. The CFM was constructed as a 1:1 scale replica of a 150 kW plant, with both reactors having a height of 5 m, while the AR and FR diameters are 0.23 m and 0.144 m, respectively (Figure 10). The system contains two cyclones and L-type loop seals [3].

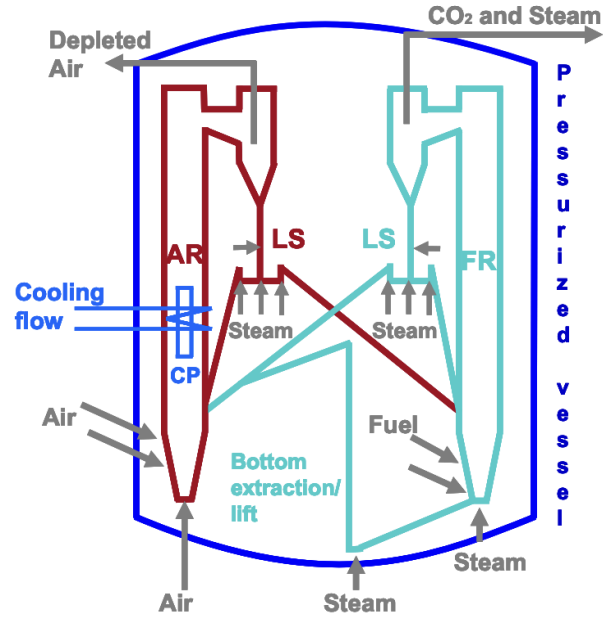


Figure 10. CFM done by Bischi et al. [3].

Unlike the two articles discussed above, much higher solid inventory mass was tested. Researchers utilized 55 kg and 65 kg of Ferro-silicon particles with 34 μm diameter to replicate realistic fluidization behavior. Air is used as the fluidizing medium, with superficial velocity of around 2.2 m/s, ensuring operation in the fast fluidization regime. Differential pressure transmitters scattered along the reactors provided pressure readings. Moreover, a filter box was implemented to accumulate particles discharged from cyclones, thus, monitor their efficiency. Similar to Pröll et al., authors also investigated the effect of loop seal fluidization on the system performance [3].

As showcased in Figure 11, in both reactors, pressure decreased with height, while higher pressure gradients were observed near the bottom due to increased solids concentration; this finding shows consistency with the previous articles [3].

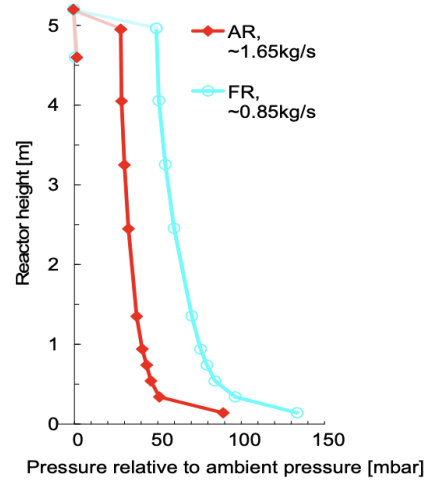


Figure 11. Relative height vs. pressure profiles of AR and FR [3].

Moreover, the results demonstrate that pressure profiles and solids circulation strongly depend on superficial gas velocity and total solids inventory. Increasing air flow and available solid mass leads to higher solids entrainment rates. Additionally, the study highlights that loop-seal operation plays a critical role in controlling solids circulation, where optimized fluidization (5 NL/min) improves mass transfer, but excessive air supply (10 NL/min) causes instability and pressure fluctuations (Figure 12).

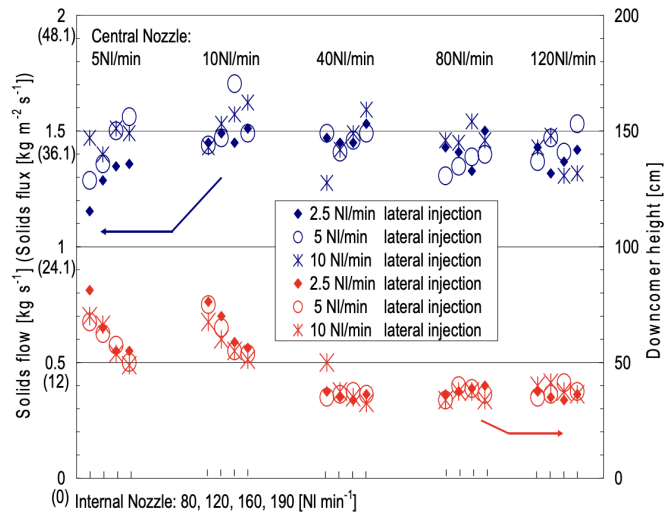


Figure 12. Effect of loop seal operation on solid flux for 55 kg [3]

The authors pointed out the importance of proper cyclone dimensioning. High mass losses due to poor cyclone performance will substantially affect the total solid inventory and consequently the reactor performance.

Table 1 summarizes the findings regarding components dimensioning and system layout of the above-mentioned research papers.

Table 1.

Summary of CFM dimensions presented in the literature review

Authors	Dimensions of AR	Dimensions of FR	Number of cyclone	Loop seal type	Solid bed material
Pröll et al. (2009)	H = 1.6 m D = 0.050 m	H = 1.2 m D = 0.054 m	2	U-type	Bronze
Shuai et al. (2011)	H = 1.9 m D = 0.19 m	H = 0.5 m D = 0.19 m	1	Pot-seal	Silica sand
Bischi et al. (2011)	H = 5 m D = 0.23 m	H = 5 m D = 0.144 m	2	L-type	Ferro-Silicon alloy

2.2 Oxygen Carriers

The choice of oxygen carrier is one of the most important factors, as it determines the efficiency of chemical looping combustion. The performance of metal oxides is characterized by various properties, including reactivity, oxygen transport capacity, lifetime, and melting temperature. These properties are critical because reactivity determines the rate of fuel conversion, oxygen transport capacity regulates the amount of oxygen transported per cycle, while stability and melting temperature determine the material's resistance to deactivation, sintering, and attrition during repeated redox cycles at elevated temperatures [2, 5]. Moreover, cost and impact on the environment are also considered when selecting an oxide for experiments. Oxygen transport capacity, often abbreviated as OTC, is defined as the relative loss of mass when the material changes from its fully oxidized state to its reduced state [9]. The table with OTC values for the main oxygen carriers is shown in Figure 13. Five of the widely used classes of oxygen carriers are Fe-, Mn-, Cu-, Ni-, and Co-based oxides [5].

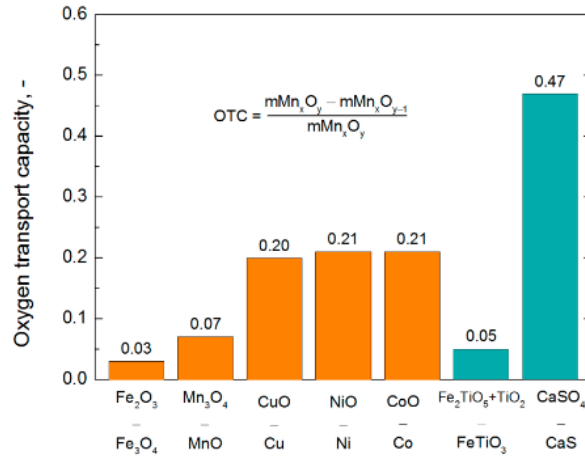


Figure 13. Oxygen transport capacity of metal oxides [5]

2.2.1 Nickel-based oxygen carriers

Nickel-based oxygen carriers are well known for their good efficiency in CLC systems. These carriers have high reactivity and work effectively at high temperatures (900–1100°C). Moreover, Ni-based materials have a fairly high OTC (about 0.21), which helps to transfer oxygen and achieve a high fuel conversion efficiency. In CLC systems, they can achieve near-complete oxidation of CH₄ to CO₂ and H₂O, but thermodynamic limitations still lead to small amounts of carbon monoxide and dihydrogen emerging in the exhaust gas exiting the fuel reactor. Despite its better qualities, nickel is also more expensive than most metal oxides. This disadvantage could be reduced by using particles with a lower Ni concentration but good reactivity. In addition, nickel-based carriers may require the implementation of security measures due to their toxicity. To enhance reactivity and improve regeneration behavior, numerous Ni-based particles produced through different synthesis routes and employing various support materials have been examined. Alumina-based supports have been the focus of extensive research. In comparison with the rest of the metal oxides, Al₂O₃ - supported oxygen carriers typically exhibit very high reactivity with a wide range of fuel gases, resistance to agglomeration, low attrition in fluidized-bed operation, and effective suppression of carbon deposition under CLC conditions [2] [22].

2.2.2 Copper-based oxygen carriers

Copper-based oxygen carriers have fast reaction rates and high OTC. Unlike Ni-based oxides, Cu-based oxygen carriers are not strongly limited by thermodynamics during the complete oxidation of the fuel to CO₂ and H₂O. Also, copper oxides are generally less expensive and more environmentally friendly than nickel- or cobalt-based oxygen carriers.

Previous research also demonstrated that CuO has better reduction than NiO and Fe₂O₃ with similar Gibbs free energy values as reported for CoO [9]. To improve the stability and reactivity of CuO, it could be supported with different materials such as aluminum oxide, bentonite, copper aluminate, magnesium oxide, magnesium aluminate, silicon dioxide, titanium dioxide, and others. Various synthesis methods, including spray drying, impregnation, mechanical mixing, and co-precipitation have also been used to enhance its efficiency. But a major drawback of using copper-based oxygen carriers is the agglomeration tendency, which is related to the relatively low melting point of copper (1085°C) [2]. The type of support plays a crucial role in the redox performance and reaction rate of Cu-based oxygen carriers. Carriers such as Al₂O₃ and TiO₂ usually slow down the redox process, while materials like MgAl₂O₄, zirconium oxide, silicon oxide, or magnesium oxide, on the contrary, accelerate the reactions [9].

2.2.3 Iron-based oxygen carriers

Iron oxides (Fe₂O₃, Fe₃O₄) are the most studied and promising oxygen carriers for CLC as they have exceptional economic as well as environmental advantages [2]. The primary benefits of this oxygen carrier are its low cost, abundant availability in nature, and non-toxicity, which is preferable for industrial utilization on a large scale [2, 6]. The reactivity of this carrier strongly depends on the composition of the surrounding gas; for example, it reacts very effectively with synthesis gas (H₂, CO). However, it exhibits a lower reaction rate with methane (CH₄), which suggests the need to use auxiliary materials or promoters to increase productivity [2, 7]. Iron-based oxygen catalysts also have a high melting point. Proper support for the oxide leads to excellent thermal and chemical stability and resistance to abrasion over many thousands of redox cycles [11].

2.2.4 Manganese-based oxygen carriers

Considerable interest has been directed toward developing manganese-based oxygen carriers, since manganese oxides are non-toxic, inexpensive, and provide higher OTC than iron compounds. Above 800 °C, Mn₃O₄ becomes the dominant phase, meaning the central redox cycle for CLC involves the Mn₃O₄ ↔ MnO transition. However, pure manganese oxide particles have poor reactivity with methane or coal, which has led to the addition of inert carriers to increase the effectiveness of this oxide. For example, Mn-based carriers stabilized with zirconium oxide (ZrO₂) and supported by magnesium oxide (MgO) demonstrate high reactivity with synthetic gas components such as H₂ and C. At the same

time, with this stabilization, they have a weaker reaction capacity with CH₄, which makes them more suitable for synthetic gas combustion. Manganese oxide can achieve very high combustion efficiency (>99.9%) at 800–950°C in a CLC unit [2, 22].

2.2.5 Cobalt-based oxygen carriers

Cobalt oxides (Co₃O₄) are recognized as high-performance oxygen carriers. Their primary advantage lies in their very high reactivity with all common fuel gases, including methane and their OTC is roughly double that of iron oxides (OTC $\approx$ 6.6% during the Co₃O₄ $\rightarrow$ CoO transition) [12, 22]. However, despite their strong performance, cobalt carriers present numerous disadvantages. For example, sintering and agglomeration occur due to their lower melting point compared to iron oxides, which can lead to deactivation and loss of surface area over time [22, 15]. Economic and environmental factors are significant deterrents that hinder their utilization. Due to its integral role in lithium-ion batteries, it is also one of the world's most expensive metals. Cobalt is also an incredibly toxic, carcinogenic, detrimental to the environment, and expensive metal [22].

Based on this information, a comparison table of oxide carriers was compiled (Table 2).

Table 2.

Comparison of oxygen carriers

	Metal Oxide	Oxygen Transport Capacity	Reactivity / Stability	Lifetime	Safety / Environmental	Cost	Problems
1	Ni-oxide	~0.21	Excellent reactivity at 900–1100 °C	high	Not specified (nickel toxicity issues known)	High	High price; susceptible to agglomeration without supports
2	Cu-oxide	~0.20	Good	average	Less toxic than NiO/CoO, but requires protective equipment	average	Low melting points $\rightarrow$ sintering, agglomeration
3	Mn-oxide	~0.07	Low CH ₄ /coal reactivity	low	Non-toxic, environmentally friendly	average	Limited for methane; needs support

							materials
4	Fe-oxide	~0.03	High syngas reactivity, stable with support oxides	average	Non-toxic, environmentally friendly, biocompatible	low	Prone to sintering; requires support design
5	Co-oxide	~0.21	High reactivity with CH ₄ and gaseous fuels	not studied	Toxic and hazardous handling risks	very high	Prone to sintering; expensive, toxic

3 Methodology

Methodology was divided into three phases to approach the project more systematically. Design and assembly of the CFM consists of system dimensioning, component sizing, and final assembled version of the model. Experimental setup and testing outlines the preparation process, input variables, procedure, and parameters to be measured. Lastly, computational fluid dynamics (CFD) simulation details how the simulations were carried out in ANSYS software.

3.1 Design and Assembly of the Cold Flow Model

Major components of the CFM are two reactors (AR and FR), two cyclones, L-type loop seals, and air distribution system. The design was initially developed as a CAD model using the specified input dimensions. Subsequently, the materials required for manufacturing and assembly were identified. The model contains 3D printed parts; they were produced using PLA 1.75 mm plastic on BambuLab printers.



Figure 14. 3D model of the AR and the cyclone.

AR has a height of 2 m and cross section of $0.1 \times 0.1 \text{ m}^2$, whereas FR is 1.2 m tall with a $0.15 \times 0.15 \text{ m}^2$ cross sectional area. The reactors were made by gluing six millimeter thick polycarbonate sheets with 3D printed angles using a silicon sealant for airtightness. The AR was split into sections and connected using flanges to enable easy assembly. Additional flanges were installed at the bottom of the reactors to form an air plenum for the air distribution system. Moreover, adapters were implemented to link reactors with loop seals and cyclones.

The cyclone core diameter is taken as 0.14 m (outside diameter is 0.15 m). The rest of the cyclone dimensions can be found in relation to the main diameter as displayed in Figure 15 provided by Patil et al. [19]. Table 3 presents the dimensioning.

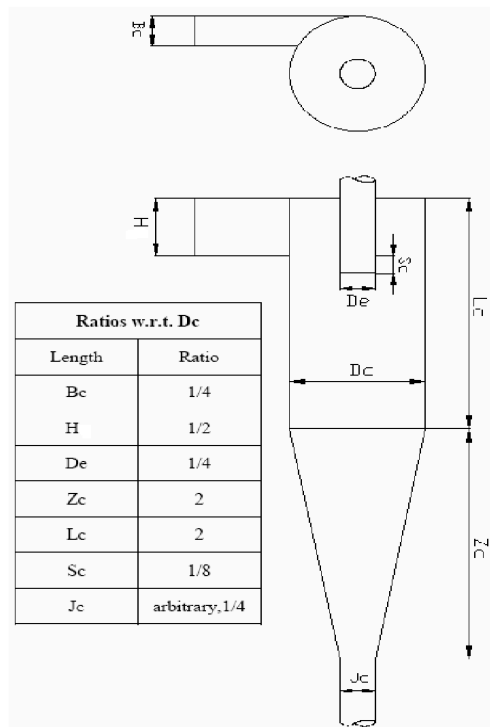


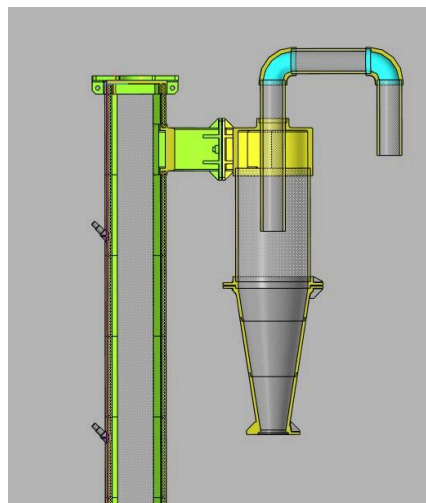
Figure 15. Cyclone dimensioning [19].

Table 3.

Cyclone dimensioning

Cyclone part	Dimension, m
D_c	0.14
B_c	0.035
H	0.07
D_e	0.035
Z_c	0.28
L_c	0.28
S_c	0.0175
J_c	0.035

The cyclone consists of a polycarbonate cylinder with 150 mm outside diameter and 3D printed parts such as cyclone cap, flanges for connecting upper and lower parts, and three-layer conical section (Figure 16). These parts are not easily available in the market, therefore, it was decided to print them to fit our exact dimensions. Polycarbonate pipes of 50 mm diameter and an air duct were attached to the cap to direct escaped sand particles from the cyclone into the collection system, preventing contamination of the surrounding. Both of the cyclones are identical in dimensions and manufacturing.

**Figure 16.** Cross section of a cyclone.

L-type loop seals were implemented in the model due to ease of manufacturing and commonness. Three loop seals were built between Cyclone 1 and FR, Cyclone 2 and FR, and FR and AR. They act as a non-mechanical valve that allows solids to pass from a low-pressure zone (the cyclone) to a higher-pressure zone (the riser) without gas bypassing or flowing in the opposite direction. Without the seals, the gas will blow up through the cyclone, resulting in complete failure of the experiments. The vertical length of the standpipe is 0.4 m, the overall horizontal length is 0.085 m (0.04 meters for both sides and 0.005 m gap) (Figure 17). They are manufactured the same way as the reactors; polycarbonate sheets are glued with 3D printed angles. 50 mm pipes are used here as well to connect loop seals to reactors and cyclones.

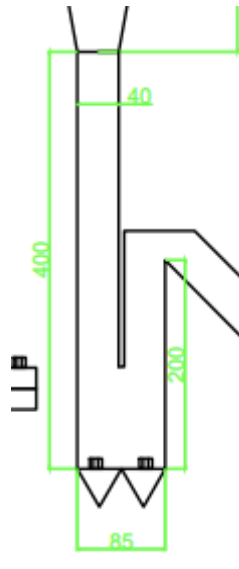


Figure 17. Loop seal dimensions.

An air distribution system based on a bubble cap design was utilized in the project to evenly distribute air in the reactors and loop seals, which were also 3D printed as can be seen in Figure 18. Nine (3 by 3) bubble caps were placed in the AR. One cap has six holes, and the open area of the caps should be 7-8% of the reactor's cross-sectional area. Hence, the diameter of the single hole was calculated as:

$$A_{all} = A_{AR} * 0.08 = 0.01 * 0.08 = 0.0008 \text{ m}^2$$

$$A_{one \text{ bubble cap}} = \frac{A_{all}}{9} = 8.89 * 10^{-5} \text{ m}^2$$

$$A_{one \text{ hole}} = \frac{A_{one \text{ bubble cap}}}{6} = 1.48 * 10^{-5} \text{ m}^2$$

$$D_{one\ hole} = \sqrt{\frac{4A_{one\ hole}}{\pi}} = 0.004\ m = 4\ mm$$

The same procedure was carried out for the FR, taking into account that there is 16 bubble caps:

$$A_{all} = A_{AR} * 0.08 = 0.0225 * 0.08 = 0.0018\ m^2$$

$$A_{one\ bubble\ cap} = \frac{A_{all}}{16} = 1.125 * 10^{-4}\ m^2$$

$$A_{one\ hole} = \frac{A_{one\ bubble\ cap}}{6} = 1.875 * 10^{-5}\ m^2$$

$$D_{one\ hole} = \sqrt{\frac{4A_{one\ hole}}{\pi}} = 0.00489\ m = 5\ mm$$

Moreover, two bubble caps were installed in each of the loop seals (Figure 18b)). Supplying air to them enhances fluidization, as seen during literature review.

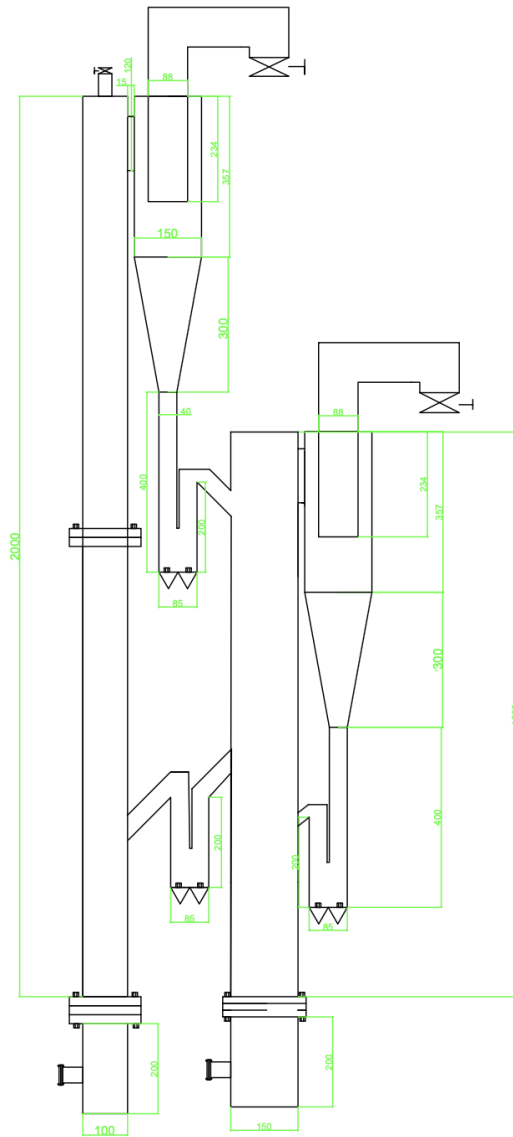


(a)

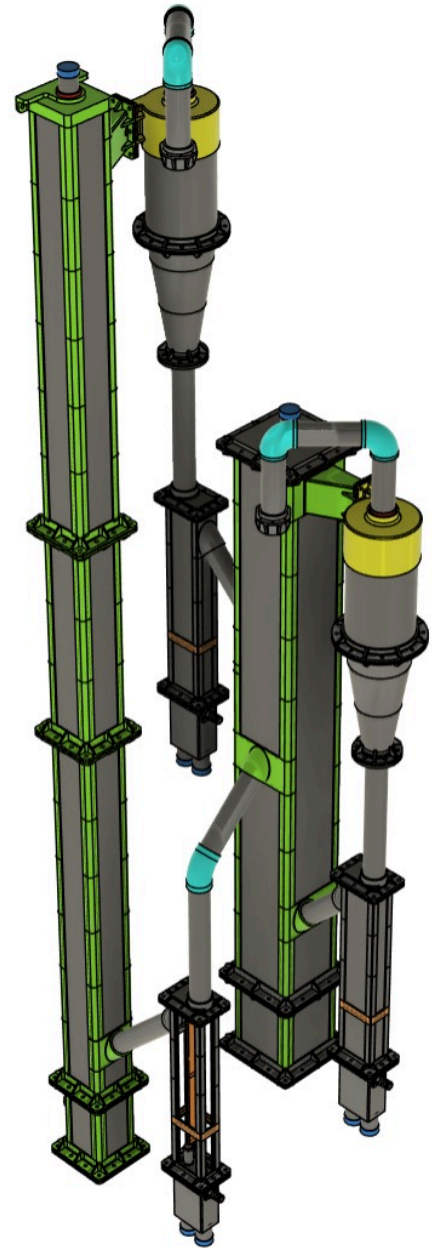
(b)

Figure 18. a) Bubble caps in the FR; b) Bubble caps in loop seals

Figure 19a) details the drawing of the system in AutoCAD software, where Figure 19b) shows the CAD model of the CFM.



(a)



(b)

Figure 19. a) Drawing of the CFM; b) CAD of the CFM

The system is assembled by connecting the parts using a silicon sealant, and further securing with nuts and bolts. Additional support in terms of stools were manufactured for reactors, and the structure was fixed to a vertical beam in the place of assembly. Figure 20 presents the assembled version of the CFM. Holes along the heights of the reactors were drilled to install pressure taps, which were connected to U-type monometers, for pressure drop measurement for further analysis (Figure 22). Moreover, available pipes in the boiler house were used to connect the compressor to the air plenum for air supply.



Figure 20. Assembled CFM.

3.2 Experimental Setup and Testing

3.2.1 Material preparation

The experiments were conducted at ambient temperature of 25°C and at atmospheric pressure (1 atm). A blend of silica sand and river sand with the same density was used as the bed material. This decision was made due to the fact that the available silica sand particles were relatively large and alone were too heavy to achieve the desired fluidization behavior. Therefore, river sand, which has a smaller particle diameter, was mixed with silica sand to obtain a more appropriate particle size distribution for stable operation of the system.

Silica sand and river sand were sieved to obtain particles in the 200 - 500 μm and 150 - 400 μm size range, respectively, with an average particle diameter of approximately 300 μm . Particle size separation was carried out using a mechanical shaker with stacked sieves (Figure 21). They were arranged in descending mesh size (first 500 μm , then 200 μm), with a lid placed on top and a collection pan at the bottom. The sand was placed in the top sieve, the stack was secured in the shaker, and turned on to provide vibrational motion. This allowed the particles to pass through the sieves according to their size. The finer particles accumulated in the collection pan and larger ones left in the bigger sieve were disposed of. The same procedure was carried out for river sand.



Figure 21. Mechanical shaker with sieves

3.2.2 Input velocity and flow rate

After determining the average size of sand particles, minimum fluidizing velocity U_{mf} was calculated using the formula and tables provided in *Fluidization Engineering* by Kunii and Levenspiel.

$$\frac{1.75}{\epsilon_{mf}^3 \phi_s} \left(\frac{d_p U_{mf} \rho_g}{\mu} \right)^2 + \frac{150(1-\epsilon_{mf})}{\epsilon_{mf}^3 \phi_s^2} \left(\frac{d_p U_{mf} \rho_g}{\mu} \right) = \frac{d_p^3 \rho_g (\rho_s - \rho_g) g}{\mu^2} \quad (1)$$

d_p - diameter of solid particle, m

ρ_s - density of solid particle, kg/m³

ρ_g - density of the fluidizing gas, kg/m³

μ - kinematic viscosity of fluidizing gas, Pas

ϵ_{mf} - voidage at minimum fluidizing conditions

ϕ_s - sphericity of solid particle

Voidage and sphericity values were extracted from Tables in *Fluidization Engineering*, presented in Appendix A (Figures A.1 and A.2). Moreover, air was used as a fluidizing medium. Table 4 presents the values for the variables in the equation.

Table 4.

Parameters

Average solid particle diameter	350 μ m
Density of solid particle	2600 kg/m ³
Density of the fluidizing gas	1.2 kg/m ³
Kinematic viscosity of fluidizing gas	1.8*10 ⁻⁵ Pas
Voidage	0.42
Sphericity	0.86

U_{mf} was calculated to be 0.1 m/s. Table 5 outlines the relationships between U_{mf} and velocities in AR, FR, and standpipe and gate sections of the loop seals, and their velocity and flow rate values.

Table 5.

Velocity and flow rate profiles in the system

Parameters	AR	FR	Standpipe section of loop seal	Gate section of loop seal
Relationship with U_{mf}	$V_{AR} = 3 * U_{mf}$	$V_{FR} = 1.5 * U_{mf}$	$V_{standpipe} = U_{mf}$	$V_{gate} = 3 * U_{mf}$
Velocity, m/s	0.3	0.15	0.1	0.3
Flow rate, m ³ /hr	10.8	12.2	1.8	5.4

The flow rate is found as a product of cross sectional area (A) of the geometry and the gas velocity (V). The air flow was regulated using valves and measured with flow meters.

3.2.3 Experimental procedure

The experimental procedure began by loading 55 kg of sand into the system as the initial bed material. After, the air supply was turned on using control valves. The system was first brought to stable operating conditions, and then the pressure drops across the compressors as well as the manometers connected to the reactors were recorded for further analysis. Two operating strategies were tested. In the first approach, the flow rate in the FR was kept constant while the air flow rate in AR was gradually increased to observe its effect on pressure drop and circulation behavior. In the second approach, the air flow rate in the AR was maintained at a constant level, while the FR flow was varied. After completing the tests with 55 kg of sand, the same methodology was repeated using a higher bed mass of 65 kg in order to investigate the influence of increased solid inventory on reactor performance, pressure drop, and circulation stability. The air supply in loop seals was kept constant.

The pressure drops across the AR and FR were measured by placing six pressure taps each along the reactor heights. These taps were connected in pairs to three U-tube manometers,

making three vertical sections: Section 1 (lower region near the distributor), Section 2 (middle region), and Section 3 (upper region).



Figure 22. U-type manometers

SCR was determined by using pressure drop measurements between the two highest pressure taps on the reactor. The measured pressure difference was expressed in terms of water gauge, where 1 mm H₂O corresponds to approximately 1 kg/m². Then, the time it takes for the sand to travel the distance between pressure taps (Δh) with a certain velocity V is estimated using:

$$t = \frac{\Delta h}{V} \quad (2)$$

Finally, SCR is computed by:

$$G_s = \frac{m}{t} \quad (3)$$

3.2.4 Computational Fluid Dynamics (CFD) simulation

Computational Fluid Dynamics simulation is conducted using ANSYS Fluent using 2D recreation of both air and fuel reactors separately to study the phase interaction of air and sand within the boundaries of the reactors. Recreation of full system geometry is time consuming and geometrically complex, thus separate reactor analysis was adopted as the

preferred approach, allowing extraction of key hydrodynamic parameters such as pressure drop, solid volume fraction, and solid particle velocity [30]. For both simulations, the Eulerian multiphase model was applied, treating fluid and solid phases as being continuously interacting with each other within a fixed mesh. This model observes fluid flow and particle transport by solving momentum and continuity equations for each phase separately. This model is widely used for fluidized bed simulations due to high computational efficiency and ability to adjust necessary flow behavior. For accurate modeling of solid-fluid interactions, the sand phase is assumed to be pseudo-fluid [23].

Gas-solid interaction is modeled through momentum exchange between particles expressed through drag force. An accurate drag coefficient model is a critical factor affecting hydrodynamic behavior of particles with the most common to be used are Gidaspow, Syamlal-O'Brien and Wen-Yu [24]. The Gidaspow and the Wen-Yu models are shown to have better correlation with experimental data compared to the Syamlal-O'Brien model that tend to underestimate bubble formation during bubbling fluidized bed expansion. Moreover, for 2D fool loop CFB simulations the Gidaspow model provides more exact pressure profile distribution data along rig height [26]. Comparative studies of 3D fluidized bed modeling using different drag models confirm that Gidaspow drag model is revealed to have adequate correlation with experimental data in terms of mass circulation rate and computational accuracy [27, 28]. Thus, for this study the Gidaspow drag model coefficient is selected.

The Eulerian multiphase model is selected with two fluid phases and an implicit scheme for volume fraction parameters as shown in Table 7 according to the approach by [18]. Air is chosen as a primary phase, while sand is selected to be a secondary phase pseudo-fluid material with a granular diameter of 350 micrometers with granular properties as indicated in Table 5. The solid phase behavior was modeled using the Kinetic Theory of Granular Flow (KTGF), which captures particle interactions, collisions, and stress development in granular systems using Boltzmann equation [24]. KTGF captures granular temperature, frictional stresses at high packing fractions, and radial distribution functions describing the likelihood of particle contact. The granular properties of the sand phase are shown in Table 6 below.

Table 6.

Granular Properties of sand

Parameter	Value/Model
Granular Diameter	350 μm
Restitution Coefficient	0.9
Granular Bulk Viscosity [$\text{kg/m}^*\text{s}$]	Lun-et al. model
Solids Pressure [Pa]	Lun-et al. model
Granular Temperature [m^2/s^2]	Algebraic closure
Frictional Viscosity [$\text{kg}/(\text{m}^*\text{s})$]	None (dilute assumption)
Packing Limit	0.63
Radial Distribution	Lun-et al. model
Elasticity Modulus [Pa]	Derived

The AR with CFB regime is generated with the dimensions of $0.1 \text{ m} \times 0.1 \text{ m} \times 2 \text{ m}$ by generating 2D surface body as shown in Figure 23a). The FR was recreated with the dimensions of $0.15 \text{ m} \times 0.15 \text{ m} \times 1.2 \text{ m}$ as can be seen in Figure 23b). Both reactors were initialized with 20 kg of sand in each reactor with an initial volume fraction of 0.6 using patch function. The fluid domain is discretized using quadrilateral mesh with coarse, medium and fine mesh element sizes of 0.008 m, 0.005 m and 0.003 m, respectively to study relationship between enhanced mesh and hydrodynamical parameters. Boundary conditions included inlet air velocity of 2 m/s at the bottom edge of the air reactor for circulating fluidization and 1.2 m/s at the bottom edge of the fuel reactor for bubbling fluidization regimes, in accordance with fluidization regimes described in CFD circulating bed simulations [4]. A pressure outlet boundary condition with ambient pressure of 0 Pa is set at the top as solid-particles are assumed to be suspended by air induced drag force. Transient pressure-based solver is used to follow the unsteady nature of fluidized beds due to altering configurations. Solver parameters are shown in Table 7.

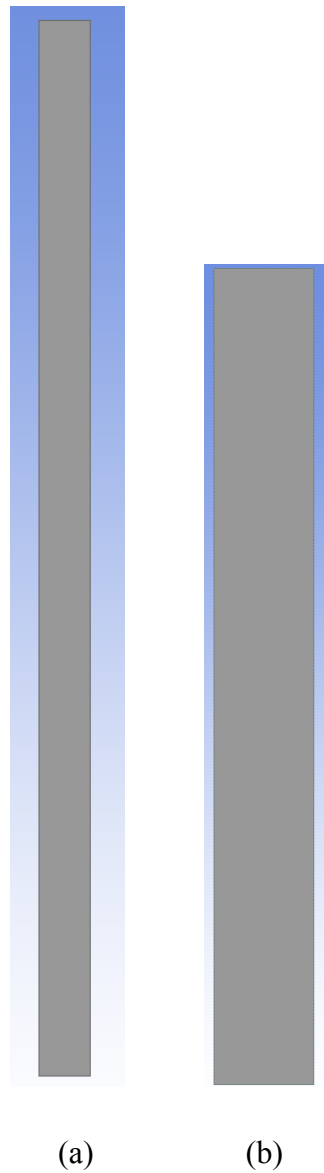


Figure 23. ANSYS Geometry of Reactors: a) Air reactor; b) Fuel reactor

Table 7.

Key solver parameters

Parameter	Value/Model
Solver type	Transient, pressure-based
Time step size	0.003-0.005 s
Number of time steps	2000
Volume fraction formulation	Implicit
Phase interaction	Gidaspow drag coefficient model, KTFG

As outlined in the literature review above, the main parameter for assessment of hydrodynamic performance is static pressure drop across the reactors. For comparison with the experimental values, static pressure values along reactors' heights were extracted at discrete locations corresponding to pressure tap positions on the laboratory-scale rig. Iso-surfaces were created along Y-axis at each tap height, and area-weighted average static pressure values of air-sand mixture was computed at these iso-surfaces. This approach allows correct estimation of average force exerted across the iso-surfaces, which is useful as the pressure is not uniform. The pressure drop along reactor heights is calculated as the difference between adjacent taps for reconstruction of pressure profile over time:

$$\Delta P = P_i - P_{i+1} \quad (4)$$

This methodology allows extraction of the axial pressure profile for comparison with experimental results [30].

4 Results and Discussion

4.1 Stable Solid Circulation

A series of trial tests was conducted, and stable solid circulation was successfully achieved during 10 days of experiments. However, the theoretically calculated gas velocity was insufficient to initiate circulation in the system. The reactor operated in a bubbling fluidized bed regime at $V_{AR} = 0.3$ m/s and $V_{FR} = 0.15$ m/s rather than a fast fluidized bed regime for circulation. To establish stable solid circulation, the gas velocity was increased until continuous particle entrainment and recirculation were observed. This was reached at approximately $V_{AR} = 1.86$ m/s and $V_{FR} = 1.23$ m/s. Moreover, flow rate in the gate part of the loop seals had to be increased to 18 m³/hr.



Figure 24. Operation of the CFM

4.2 Pressure Drop

The measurements were recorded at constant air flow conditions of $Q_{AR} = 67 \text{ m}^3/\text{hr}$ (1.86 m/s) and $Q_{FR} = 100 \text{ m}^3/\text{hr}$ (1.23 m/s). The results are summarized in Tables 7 and 8, and the graphs are presented in Figures 25 and 26, where the pressure drop is plotted against the three sections for two solid mass inventory (55 kg and 65 kg). 1 mm H₂O is equal to 9.81 Pa. Section 1 is the dense zone, Section 2 is the transition zone, and Section 3 is the freeboard zone.

The findings indicate a clear decreasing trend in pressure drop in AR and FR with increasing height for both sand inventories (55 kg and 65 kg). The highest pressure drop was consistently observed in Section 1, while the lowest values were recorded in Section 3. The sudden drop in pressure between Sections 2 and 3 indicates a change from a turbulent fluidization regime to a more dilute transport regime, where there are fewer particles that resist the flow of gas. This behavior can be explained by the distribution of solids within the reactor. A dense bed region forms near the bottom section because of particle build-up and intense gas-solid interactions. This makes it harder for air to flow through, causing a higher pressure drop. The concentration of particles decreases as the gas rises, changing into a more dilute phase in the upper zones. This leads to lower pressure drop values.

Table 8.

Pressure drop in three sections of the AR at different masses of solid inventory

AR, m = 55 kg			AR, m = 65 kg		
ΔP_1	ΔP_2	ΔP_3	ΔP_1	ΔP_2	ΔP_3
430 mm	200 mm	40 mm	530 mm	250 mm	130 mm
4.2 kPa	2 kPa	0.4 kPa	5.2 kPa	2.5 kPa	1.3 kPa

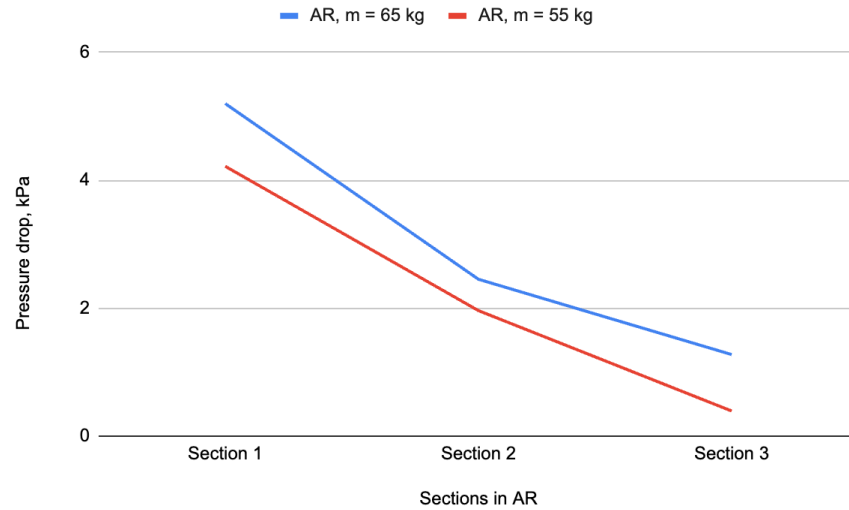


Figure 25. Pressure drop vs. Geometrical height for AR

Table 9.

Pressure drop in three sections of the FR at different masses of solid inventory

FR, m = 55 kg			FR, m = 65 kg		
ΔP_1	ΔP_2	ΔP_3	ΔP_1	ΔP_2	ΔP_3
290 mm	260 mm	60 mm	420 mm	310 mm	110 mm
2.8 kPa	2.6 kPa	0.6 kPa	4.1 kPa	3.1 kPa	1.1 kPa

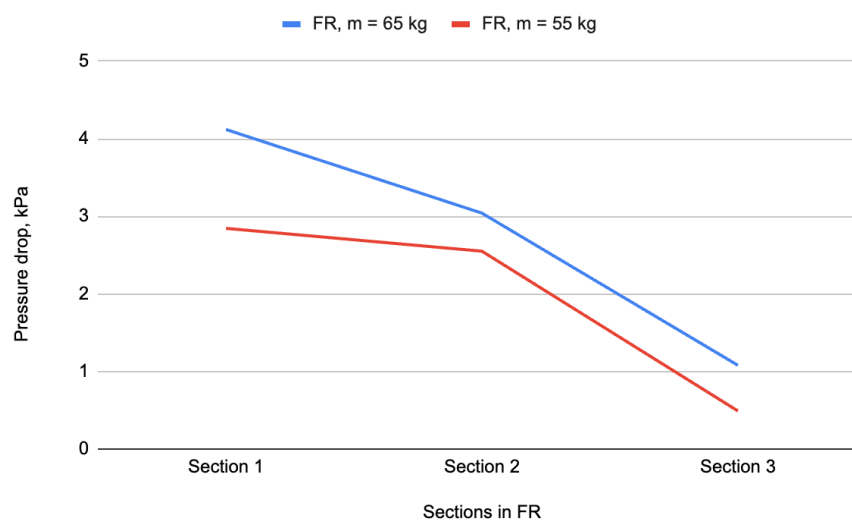


Figure 26. Pressure drop vs. Section in FR

Additionally, mass of the solid inventory has a significant effect on the pressure drop. When the bed inventory was raised from 55 kg to 65 kg, the pressure drop across all sections went up by a noticeable amount. For example, in Section 1 of AR the pressure drop increased from nearly 4.2 kPa to 5.2 kPa, whereas 1.3 kPa increase (from 2.8 kPa to 4.1 kPa) was noticed in the FR in the same segment. This is to be expected because a higher solid loading makes the bed denser and enhances interparticle interactions, which increases flow resistance. Overall, the observed pressure profile aligns with the standard behavior in DCFB systems identified during the literature review. This confirms that the system is operating under conditions suitable for solid circulation.

Moreover, the effect of inlet air velocity on pressure drop was also investigated. It was observed that increasing the airflow in a given reactor led to a corresponding increase in pressure drop within that same reactor as seen in graphs in Figures 27 and 28. The biggest increase was noticed in Section 1, which is explained by its closest location to the air distribution system. This is to be expected because higher gas velocities improve gas-solid interactions, increase drag forces, and keep particles suspended, all of which contribute to a higher resistance to flow.

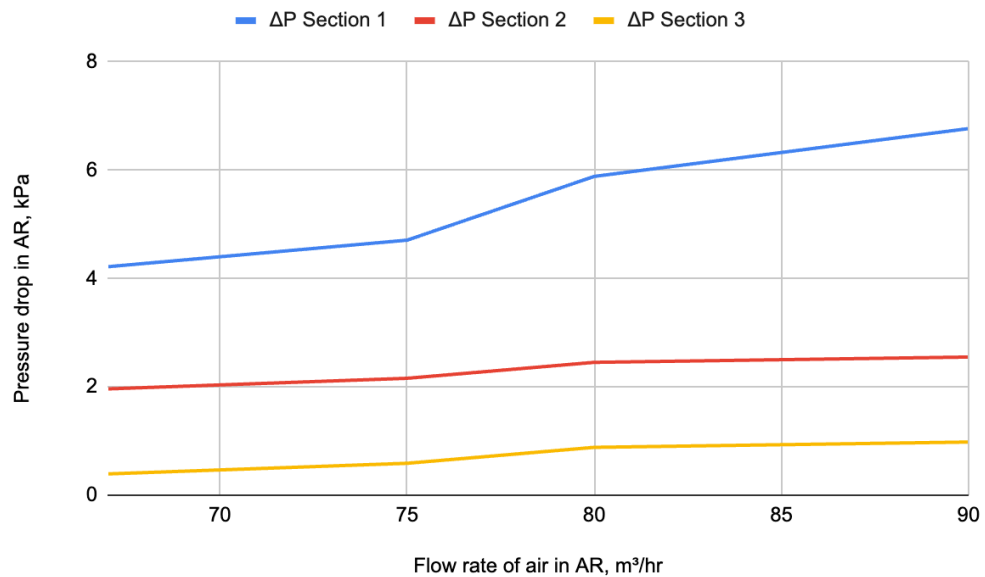


Figure 27. Pressure profiles of different sections of AR at various flow rates of AR

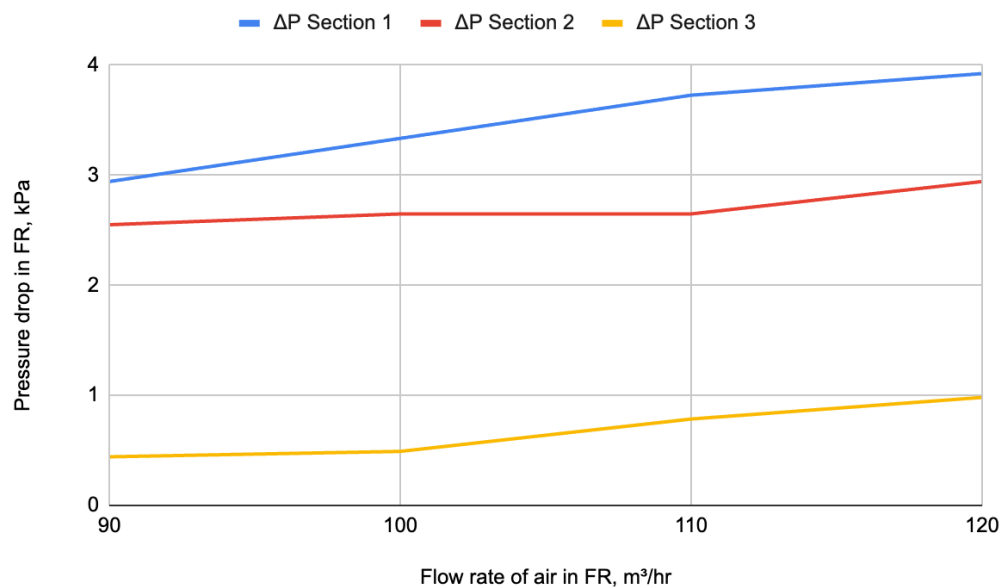


Figure 28. Pressure profiles of different sections of FR at various flow rates of FR

However, changes in airflow in one reactor had a negligible effect on the pressure drop in the other reactor. For instance, increasing the air supply to the AR did not significantly influence the pressure profile in the FR, and vice versa (Figures 29 and 30). It can be inferred that each reactor primarily responds to its own gas flow conditions.

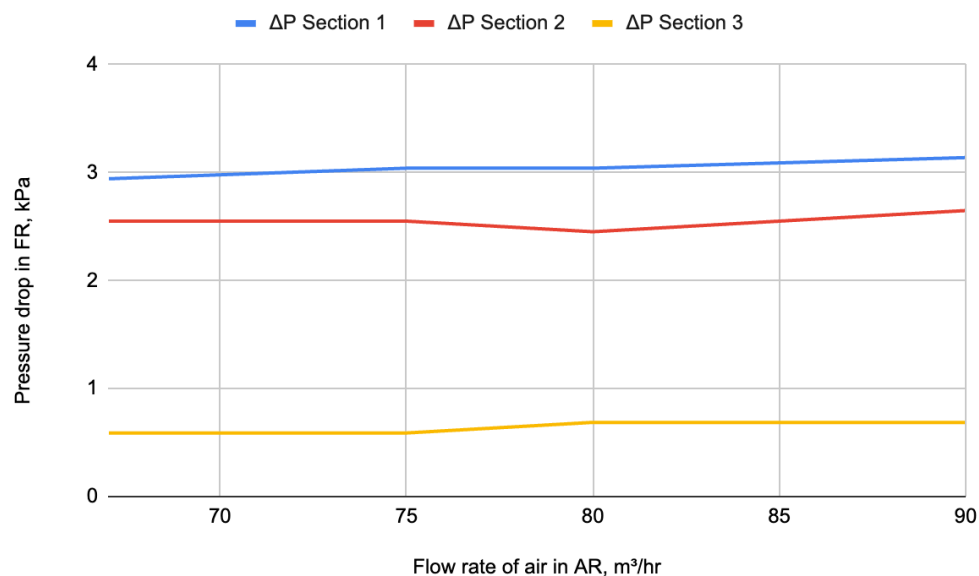


Figure 29. Pressure profiles of different sections of FR at various flow rates of AR

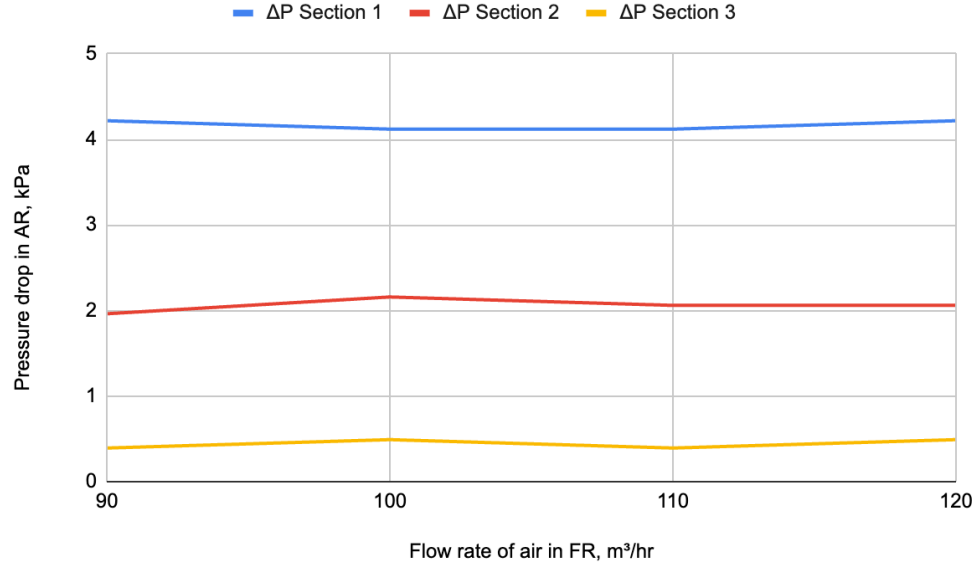


Figure 30. Pressure profiles of different sections of AR at various flow rates of FR

4.3 Solid Circulation Rate

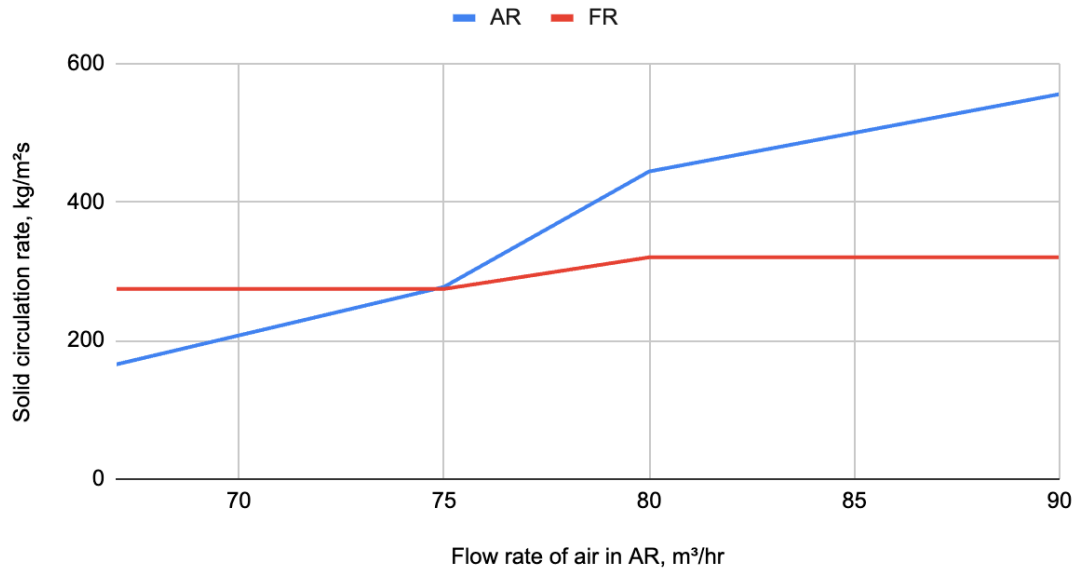
First, the experiments were carried out with 55 kg of solid inventory mass. The air velocity in the AR was varied (67, 75, 80, 90 m³/hr), while the operating conditions in the FR were kept constant (60 mm H₂O, $Q = 100$ m³/hr, $V = 1.23$ m/s). In the second set of experiments the airflow in the FR was varied, while the AR was maintained at constant operating conditions (500 mm H₂O, $Q = 67$ m³/hr, $V = 1.86$ m/s). Then, the same procedure was repeated for 65 kg.

In the experiment with 55 kg solid inventory and constant FR operating conditions, the SCR in the AR significantly increased with an increase in the airflow rate in the AR. The SCR in the AR increased from 165.33 kg/m²s to 555.56 kg/m²s, with the gas velocity in the AR increasing from 1.86 m/s to 2.5 m/s (Table 8). This suggests that increasing the gas velocity in the AR improved particle entrainment and solids transport due to increased drag forces and enhanced gas-solid mixing. However, the SCR in the FR only slightly increased from 274.35 kg/m²s to 320.07 kg/m²s, which indicates that the effect of airflow in the AR on the solid circulation in the FR was weak. The visual changes can be observed in Figure 31.

Table 10.

Solid flux in AR and FR at different air velocities in AR (55 kg solid bed material)

Pressure drop in compressor in AR, mm H ₂ O	Flow rate of air in AR, m ³ /hr	Velocity of air in AR, m/s	Pressure drop in AR manometers, mm H ₂ O	Pressure drop in FR manometers, mm H ₂ O	G _s in AR, kg/m ² s	G _s in FR, kg/m ² s
500	67	1.86	40	60	165.33	274.35
600	75	2.08	60	60	277.33	274.35
700	80	2.22	90	70	444.00	320.07
800	90	2.5	100	70	555.56	320.07

**Figure 31.** SCR in AR and FR at different AR flow rates (55 kg)

The same trend was evident with the 65 kg solid inventory at the same operating conditions in the FR. The particle circulation in the AR increased significantly as the air flow rate in the AR increased, whereas in the FR, a more moderate increase in particle circulation was observed (Figure 32). However, in comparison to the 55 kg solid inventory case, the SCR in both reactors were much higher for 65 kg. For instance, during the same operating conditions, (80 m³/hr) solid flux in the AR made up 986.67 kg/m²s and 444 kg/m²s in 55 kg and 65 kg cases, respectively. This suggests that the increased solid inventory promoted

better particle availability and facilitated solid circulation in the system, leading to more intense and stable circulation.

Table 11.

Solid flux in AR and FR at different air velocities in AR (65 kg solid bed material)

Pressure drop in compressor in AR, mm H ₂ O	Flow rate of air in AR, m ³ /hr	Velocity of air in AR, m/s	Pressure drop in AR manometers, mm H ₂ O	Pressure drop in FR manometers, mm H ₂ O	G _s in AR, kg/m ² s	G _s in FR, kg/m ² s
500	67	1.86	130	110	537.33	501.11
600	75	2.08	160	115	739.56	523.89
700	80	2.22	200	120	986.67	546.67
800	90	2.5	230	120	1277.78	546.67

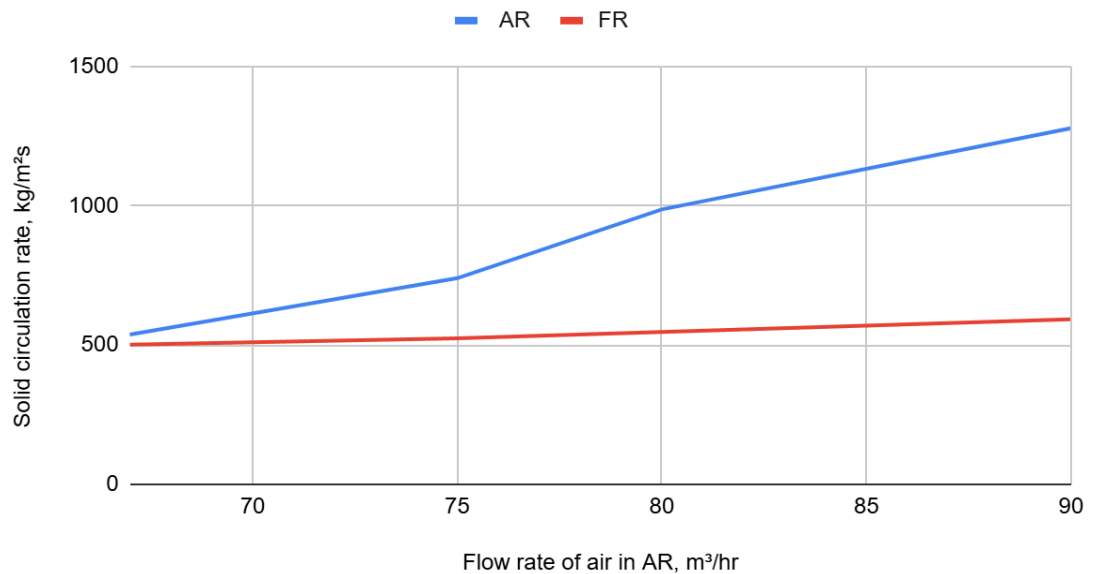


Figure 32. SCR in AR and FR at different AR flow rates (65 kg)

In the experiment with a 55 kg solid inventory and a constant operating condition in the AR, a significant increase in the SCR in the FR was observed as the airflow in the FR increased (Figure 33). The circulation rate in the FR increased from 205.74 kg/m²s to 548.15 kg/m²s as the gas velocity in the FR increased from 1.11 m/s to 1.48 m/s. The SCR in the AR, on the other hand, increased from 165.33 kg/m²s to 227.33 kg/m²s (Table 10). This confirms that

the gas flow rate in the FR had a significant effect on the circulation in the FR, but only a marginal effect on the AR, coinciding with the previous results.

Table 12.

Solid flux in AR and FR at different air velocities in FR (55 kg solid bed material)

Pressure drop in compressor in FR, mm H ₂ O	Flow rate of air in FR, m ³ /hr	Velocity of air in FR, m/s	Pressure drop in AR manometers, mm H ₂ O	Pressure drop in FR manometers, mm H ₂ O	G _s in AR, kg/m ² s	G _s in FR, kg/m ² s
50	90	1.11	40	50	165.33	205.74
60	100	1.23	40	60	165.33	274.35
70	110	1.36	50	80	206.67	402.96
80	120	1.48	55	100	227.33	548.15

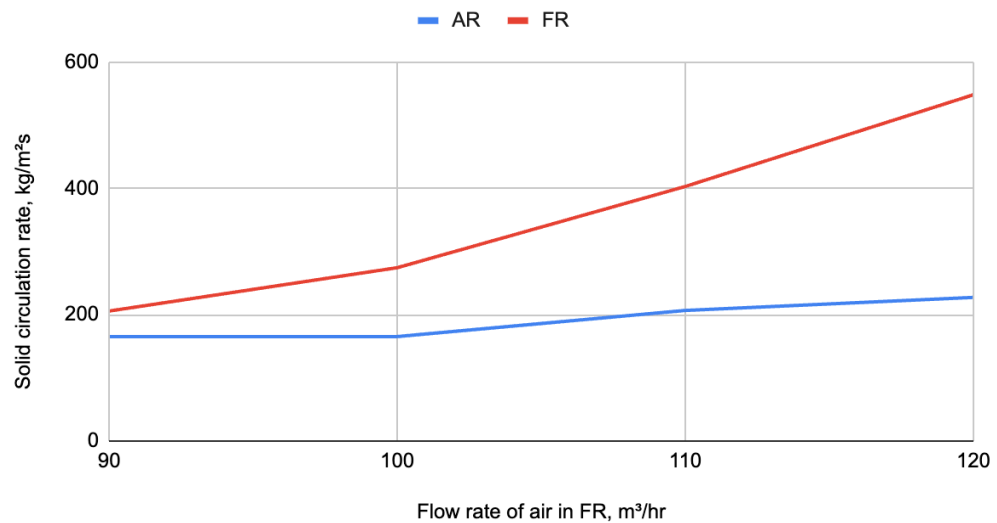


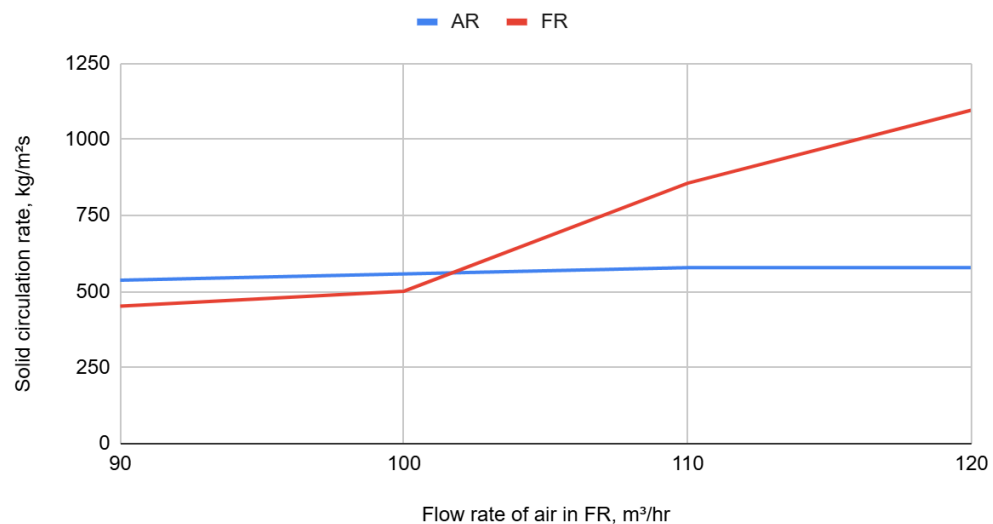
Figure 33. SCR in AR and FR at different FR flow rates (55 kg)

A similar trend was also observed for the 65 kg solid inventory. The SCR in the FR increased considerably for increasing airflow in the FR, while the circulation rate in the AR remained almost unaffected. Again, the circulation rates obtained with 65 kg solid inventory were significantly higher than those from the 55 kg inventory, further showing that increasing the solid inventory leads to better solids circulation and overall better hydrodynamic behaviour of the system.

Table 13.

Solid flux in AR and FR at different air velocities in FR (65 kg solid bed material)

Pressure drop in compress or in FR, mm H ₂ O	Flow rate of air in FR, m ³ /hr	Velocity of air in FR, m/s	Pressure drop in AR manometers, mm H ₂ O	Pressure drop in FR manometers, mm H ₂ O	G _s in AR, kg/m ² s	G _s in FR, kg/m ² s
50	90	1.11	130	110	537.33	452.22
60	100	1.23	135	130	558.00	501.11
70	110	1.36	140	170	578.67	856.30
80	120	1.48	140	200	578.67	1096.30

**Figure 34.** SCR in AR and FR at different FR flow rates (65 kg)

The findings show that increasing the airflow in one reactor had a significant impact on the SCR in that reactor as it improved gas-solid mixing and particle movement. On the other hand, changes in airflow in one reactor only had a very limited effect on the other reactor. Furthermore, increasing the solid inventory from 55 kg to 65 kg had a positive effect on the SCR in both reactors, suggesting that the solid inventory is an important factor in enhancing the SCR in circulating fluidized bed systems.

4.4 CFD simulations

4.4.1 Air Reactor

Static pressure values were extracted at six iso-surface positions as shown in Table 14.

Table 14.

Pressure taps locations

Pressure tap order	Tap 1	Tap 2	Tap 3	Tap 4	Tap 5	Tap 6
Position along Y-axis, m	0.05	0.4	0.8	1.2	1.6	1.85

The coarse mesh of 0.008 m creates a steep exponential-like pressure between pressure taps. As can be seen in figure 43, most of the pressure is lost at the bottom of the air reactor with 4.6 kPa decreasing by half at the next point. The values for pressure profile are lower than should be expected for circulating bed with 20 kg of sand, which could indicate that the coarse mesh accuracy in solving dense phase interaction is poor with the solid particles dispersing artificially in the upward direction. Also, the value close to zero at the last tap shows a lack of particles at the top of the reactor, which is inaccurate for fluidized bed operating at the velocity much higher than minimum fluidization velocity.

Table 15.

Pressure values for AR with 0.008 m element size coarse mesh

Pressure tap order	Tap 1	Tap 2	Tap 3	Tap 4	Tap 5	Tap 6
Static pressure value, kPa	4.6	2.3	1.5	0.7	0.1	0.11
Pressure drop, kPa	2.3		0.8		0.01	

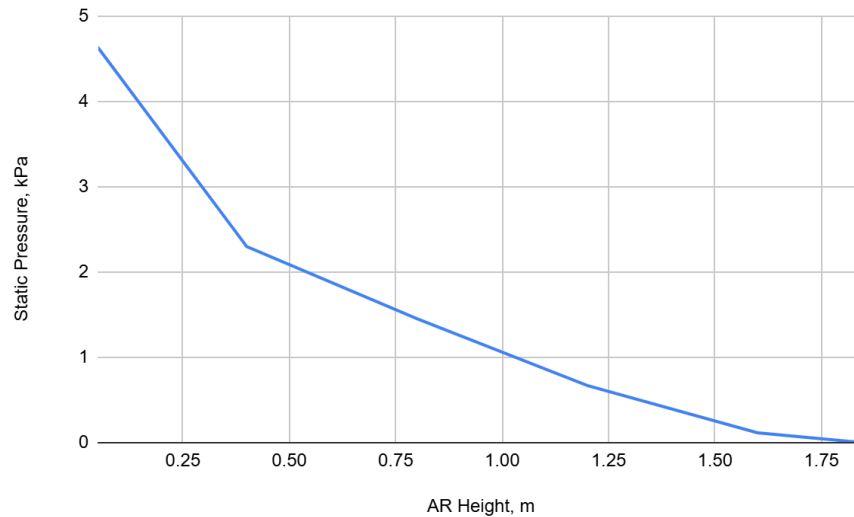


Figure 35. Pressure vs. AR Height for coarse mesh

Static pressure values produced by the medium mesh have much higher magnitude and more linear pressure profile, with Tap 1 value of approximately 50.2 kPa declining to Tap 6 at 3.9 kPa in a fairly consistent gradient. Such linear axial pressure gradient is typical for well mixed and uniformly suspended solid loading. The overall pressure drop of approximately 46.3 kPa is much higher than the coarse mesh indicating higher solid retention and lower distribution of the particles within the fluid domain. The remaining pressure at Tap 6 of around 3.9 kPa shows that the discharge area was continuously particle loaded, which is physically compatible with CFB.

Table 16.

Pressure values for AR with 0.005 m element size medium mesh

Pressure tap order	Tap 1	Tap 2	Tap 3	Tap 4	Tap 5	Tap 6
Static pressure value, kPa	50.2	41.2	30.9	20.6	10.3	3.9
Pressure drop, kPa	9.7		10.3		6.4	

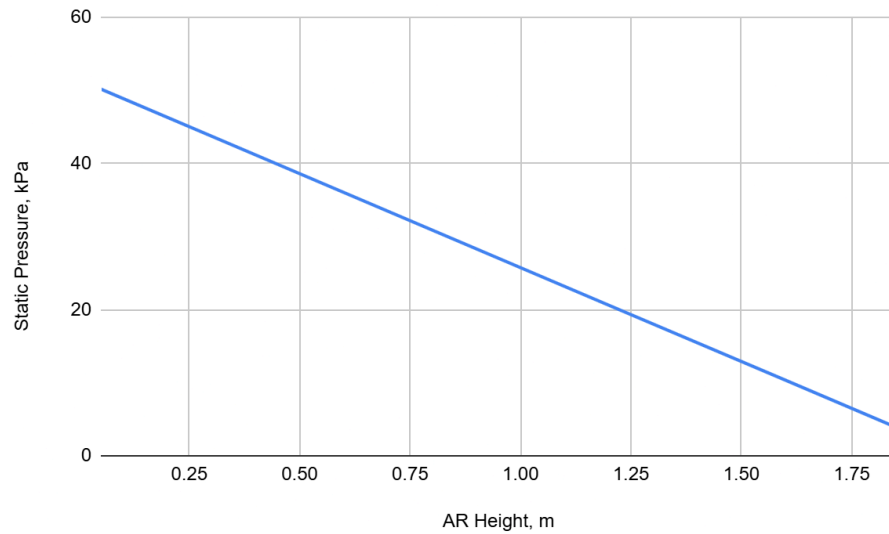


Figure 36. Pressure vs. AR Height for medium mesh

The fine mesh yields the lowest static pressure values of the three configurations, and most of the pressure is concentrated in the bottom half of the AR with nearly zero values above 1.0 m. This trend indicates that higher mesh resolution is associated with improved control of numerical diffusion and results in localized solid clustering around the air distributor. The sharp pressure gradient with exponential decay suggests a particle dense bottom and dilute upper layer, which follows circulating bed behavior but the magnitude is unexpectedly low. This can either indicate insufficient simulation time, time-step sensitivity at 0.003 s or necessity for model adjustments for finer mesh.

Table 17.

Pressure values for AR with 0.003 m element size fine mesh

Pressure tap order	Tap 1	Tap 2	Tap 3	Tap 4	Tap 5	Tap 6
Static pressure value, Pa	1.2	1	0.5	0.07	0.02	0.00008
Pressure drop, kPa	0.2		0.43		0.01992	

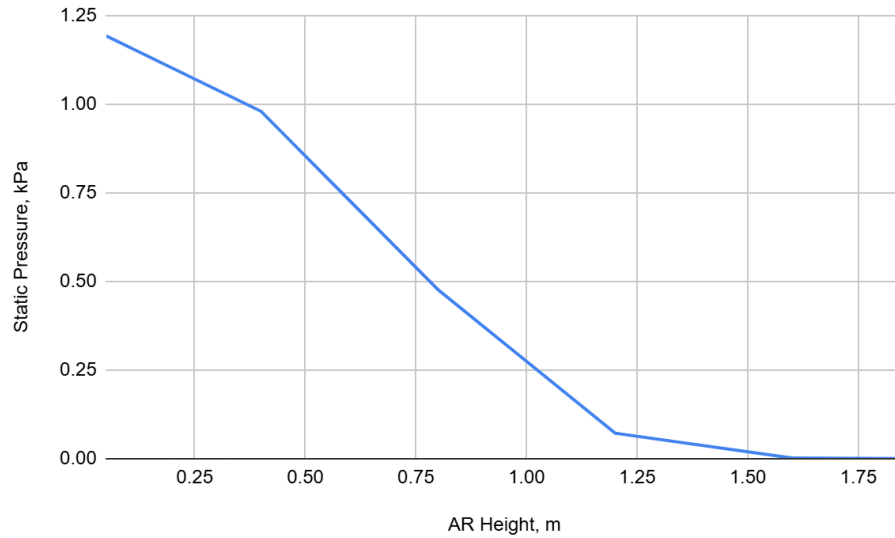


Figure 37. Pressure vs. AR Height for fine mesh

Pressure profile was shown to have substantially different behavior for the three mesh configurations, which implies that the AR simulation is highly mesh-dependent with key observations being the following: the medium mesh configuration with element size of 0.005 m represents the highest pressure distribution with linear pressure gradient. On the other hand, the coarse mesh leads to a high solid reduction rate which is consistent with CFB hydrodynamics in which particles are transported along the column length. It is more physically realistic with lower computational complexity. The fine mesh results in high concentration of solids at the bottom of the AR. Convergence of particles may require longer simulation time and lower time step since CFB needs some time to enter hydrodynamic steady state.

4.4.2 Fuel Reactor

Similar to the approach for the air reactor, iso-surfaces were created at the approximate position of pressure taps in the laboratory rig, as shown in Table 18.

Table 18.

Pressure taps location

Pressure tap order	Tap 1	Tap 2	Tap 3	Tap 4	Tap 5	Tap 6
Position along Y-axis, m	0.05	0.25	0.5	0.75	0.95	1.1

The coarse mesh simulation of FR indicates a focused dense bed in the lower part in the region between taps 1 and 2, followed by a sharp pressure drop between tap 2 (4.3 kPa) and Tap 3 (1 kPa), and a near-zero pressure on top. This profile fits a bubbling fluidized bed (BFB) regime behavior in which there is a strong dense phase and bubble phase (dilution). The abrupt decrease in pressure above tap 3 indicates a high solid concentration region, and above it is freeboard with a small amount of solids, as can be expected for BFB conditions.

Table 19.

Pressure values for FR with 0.008 m element size coarse mesh

Pressure tap order	Tap 1	Tap 2	Tap 3	Tap 4	Tap 5	Tap 6
Static pressure value, kPa	5.1	4.3	1	0.00013	0.00002	0
Pressure drop, kPa	0.8		0.99987		0.00002	

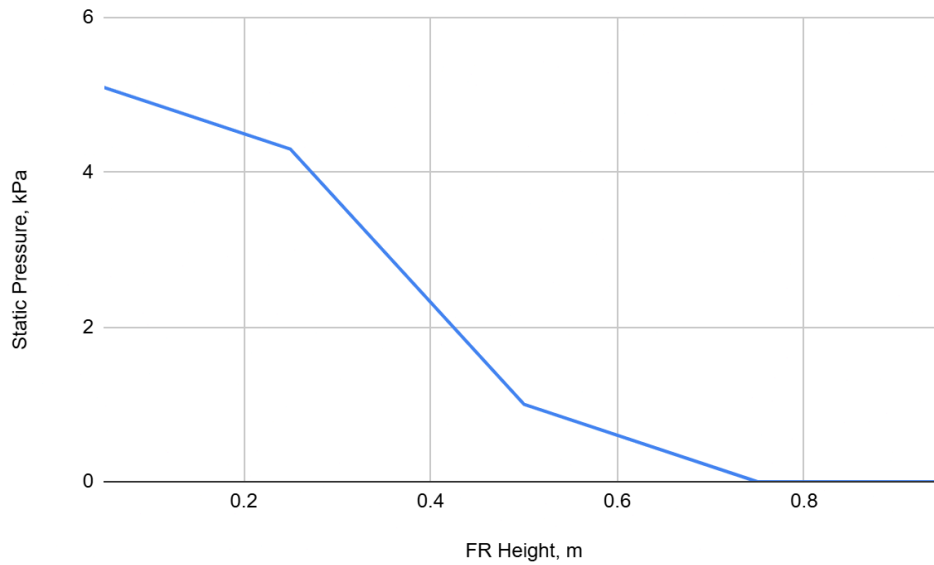


Figure 38. Pressure vs. FR Height for coarse mesh

The medium mesh bed performance is characterized by an exponentially declining profile of pressure from the very beginning. The difference in total static pressure between two extreme taps is approximately 3.7 Pa. The medium mesh resolution is more accurate for representation of continuous bubble rise and phase interactions in BFB than the coarse mesh due to the smoother decay. A presence of pressure value at tap 4 (0.2 kPa) distinguished from zero value for the coarse mesh suggests solid entrainment in the upper part of the reactor. This is in line with bubbling activity which causes particle discharge above nominal solid bed area.

Table 20.

Pressure values for FR with 0.005 m element size medium mesh

Pressure tap order	Tap 1	Tap 2	Tap 3	Tap 4	Tap 5	Tap 6
Static pressure value, kPa	3.7	1.6	0.9	0.2	0.00058	0
Pressure drop, kPa	2.1		0.7		0.00058	

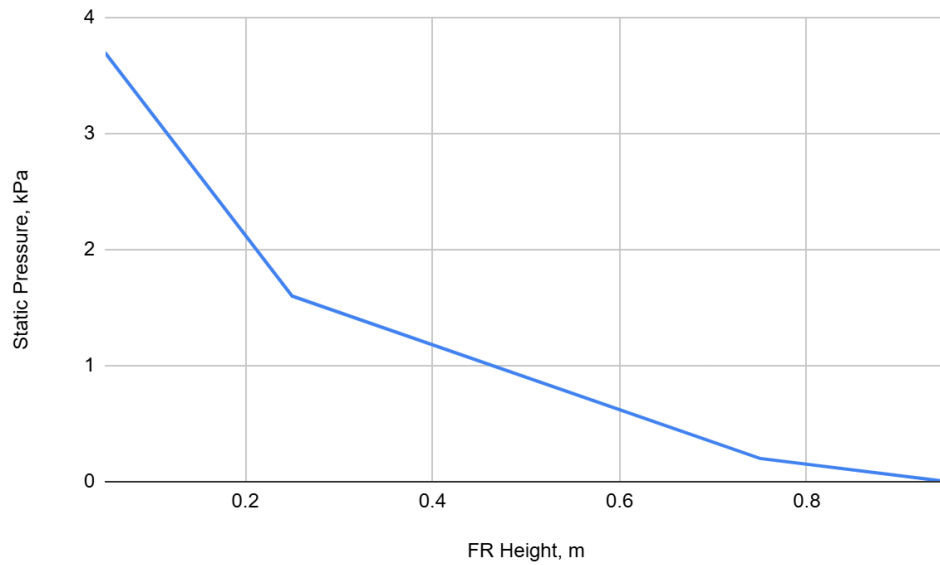


Figure 39. Pressure vs. FR Height for medium mesh

Static pressure values for the fine mesh FR simulation were revealed to have extremely low magnitude having essentially zero pressure profile higher than 0.6 m. The bed is shown to be diluted in all tap positions which can be caused by the small bubble structures exiting the fluid domain quickly or due to an inadequate time for reaching hydrodynamic equilibrium. The values beginning with 0.3 Pa at tap 1 are significantly smaller than coarse and medium results, indicating that the fine mesh might require modified time-step settings.

Table 21.

Pressure values for FR with 0.003 m element size fine mesh

Pressure tap order	Tap 1	Tap 2	Tap 3	Tap 4	Tap 5	Tap 6
Static pressure value, kPa	0.26	0.09	0.009	0.00047	0.00017	0
Pressure drop, kPa	0.17		0.0043		0.00017	

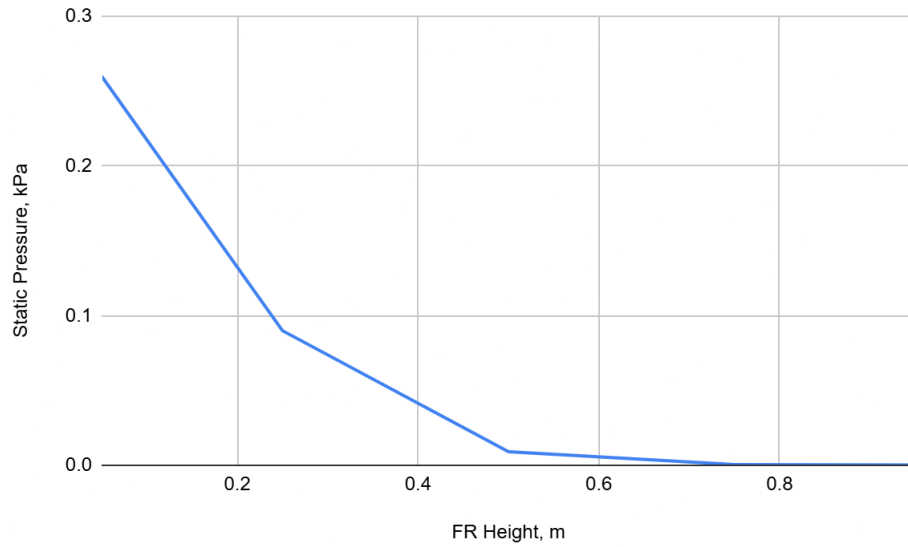


Figure 40. Pressure vs. FR Height for fine mesh

There is a pattern of pressure concentration in the lower reactor section for all three FR mesh levels, as can be expected in BFB behavior where the dense mixture phase would be found in the lower bed section. The coarse mesh gives the broadest pressure profile, while the medium mesh results in a smoother and more realistic profile of particle mixing. The fine mesh needs adjustment including more simulation time and lower time step to achieve more reasonable quantitative results. The medium mesh can be chosen for comparison with experimental data since this configuration most closely aligns with the exponential BFB pressure profile [31].

Sand phase distribution in the fuel reactor for each mesh resolution with bubble formation is shown in Figure 41.

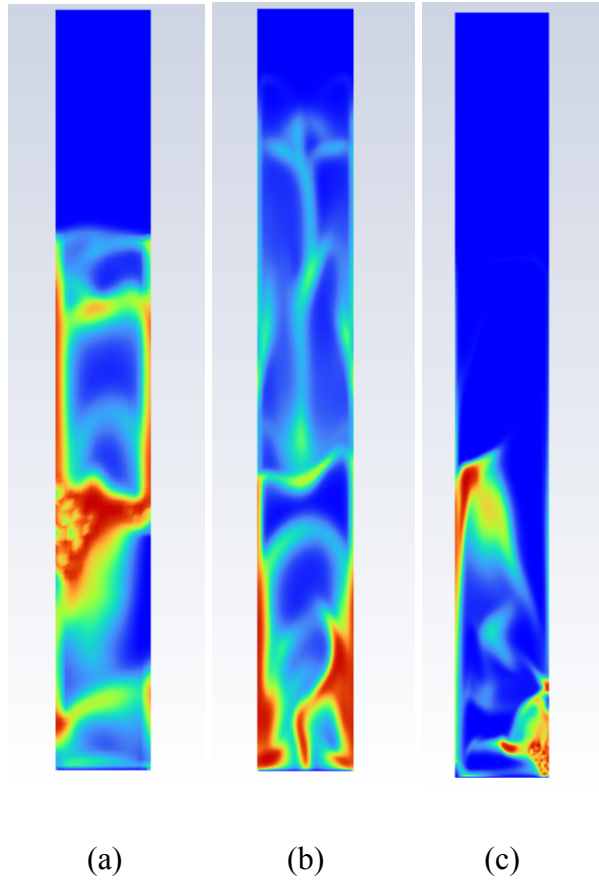


Figure 41. Sand volume fraction profiles for FR: a) 0.008 m element size; b) 0.005 m element size; c) 0.003 m element size

4.4.3 Comparison of CFD simulations and experimental results

For comparison between the CFD simulations and the experimental rig air reactors the 55 kg configuration is used, as the sand mass is considered to be distributed between two reactors and loop seals. The pressure drop values are summarized in Tables 22. The most close values are achieved by the coarse mesh, with the resulting pressure drop being lower due to lower sand mass in the simulation, resulting in underestimation. The medium and fine mesh AR simulation yield high differences with the experimental values, because of the solid retention and inadequate simulation time.

In the case of the fuel reactor, 55 kg configuration is also used with the results shown in Table 23. The medium mesh demonstrated better correlation with the experimental values. The experiment as well as CFD exhibit the same qualitative behavior: the maximum pressure drop is in the lower dense zone and then disappears strongly towards the upper section, which is a confirmation that the simulations are able to recreate the basic hydrodynamic regime structure of a bubbling fluidized bed.

Obtained results verify the drop of pressure with height across all conditions and that the dense zone is limited to the lower part of the reactor - as expected with well-established DCFB behavior. The experiment result that the airflow manipulation in one of the reactors causes no significant change to the pressure profile of the other reactor also justifies the independent reactor modelling strategy used in the CFD model.

The main causes of quantitative discrepancy include possible solid inventory difference in the CFD model, the 2D geometric simplification and the small amount of simulation time - especially with the fine mesh. The most effective steps toward achieving quantitative agreement would be to construct a full CFD model and adjust it to the actual sand loading used in the experiment (55 kg or 65 kg) with extended simulation time.

Table 22.

Comparison of experimental results with CFD simulation pressure drop for AR

Section number	Experimental, kPa	Coarse, kPa	Medium, kPa	Fine, kPa
Section 1	4.2	2.3	9.7	0.2
Section 2	2	0.8	10.3	0.43
Section 3	0.4	0.01	6.4	0.01992

Table 23.

Comparison of experimental results with CFD simulation pressure drop for FR

Section number	Experimental, kPa	Coarse, kPa	Medium, kPa	Fine, kPa
Section 1	2.8	0.8	2.1	0.17
Section 2	2.6	0.99987	0.7	0.0043
Section 3	0.6	0.00002	0.00058	0.00017

4.5 Limitations and Recommendations

During operation, both of the reactors experienced a localized structural failure, where a section of the reactor casing opened, likely due to elevated internal pressure and high gas velocity (Figure 42). This required immediate shutdown of the system, followed by complete removal of the bed material and partial disassembly of the reactors. The damaged sections were reinforced and reassembled as can be seen in Figure 43.



Figure 42. Structural failure on the FR.



Figure 43. Reinforcement on the FR.

The need to upkeep high gas velocities to achieve circulation puts high pressure on the system components, making the system less stable and more prone to breakage. This issue shows that future designs should incorporate stronger materials that can withstand higher pressure to keep the structure stable under high-velocity fluidization conditions.

Additionally, a minor limitation observed was the leakage of sand particles through the holes of bubble caps into the air plenum, where they accumulated over time (Figures 44 and 45). While this did not cause any significant damage to the system, periodic removal of the collected material was carried out to maintain proper operation. To mitigate this issue, it is recommended to implement longer bubble caps with smaller hole diameters and increase the number of holes to improve air distribution and reduce particle loss.



Figure 44. Sand leakage to the air plenum of AR.



Figure 45. Sand leakage to the air plenum of the lower loop seal.

A limitation was identified in the cyclone performance, whereby not only fine particles but also a fraction of larger particles were getting carried away through the cyclone. This unintended separation may have affected the inaccuracy of the experimental results. Therefore, further optimization of the cyclone design and dimensioning is required to achieve more efficient and selective particle separation.

Furthermore, the experiments demonstrated the importance of high air supply capacity. Working with particles with larger diameters, higher densities, or more bed material require increased gas velocities for fluidization. Insufficient air flow may limit fluidization quality and solid circulation, particularly in circulating fluidized bed systems. Therefore, future setups should incorporate a higher-capacity compressor to ensure consistent and stable operation under such conditions.

5 Conclusion

In this project, the design, construction, and testing of a laboratory-scale dual circulating fluidized bed CFM for use in chemical looping combustion were demonstrated. The prototype allowed the study of important hydrodynamic properties, such as pressure drop and SCR, under ambient conditions. The results indicated that pressure drop declined with increasing reactor height, decreasing from 5.2 to 1.28 kPa in the air reactor and from 2.22 to 0.39 kPa in the fuel reactor, reflecting the transition from the dense particle concentration zone to the dilute particle concentration region. Furthermore, the increase in solid inventory from 55 kg to 65 kg resulted in a significant increase in pressure drop and a stronger SCR in both reactors. The research also showed that increasing the gas velocity in the air reactor from 1.86 m/s to 2.5 m/s increased the SCR from 165.33 to 555.56 kg/m²s, corresponding to approximately 200% rise, while having only a relatively minor effect on the interconnected reactor. A stable circulation was achieved only at greater gas velocities than expected theoretically. The study also demonstrated that CFD models were able to simulate pressure distributions and gas-solid flow patterns observed in the reactors. The CFM has been proven to be an effective tool to investigate the DCFB hydrodynamics safely and cost-effectively. However, its limitations were identified in the project, including the structural instability at high pressure and the leakage of particles through the bubble caps and cyclones. In general, the results obtained will be useful for the future design and scale-up of chemical looping combustion systems for efficient and sustainable energy generation.

6 Contributions and Acknowledgements

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Dana Amangulova contributed to the literature review on oxygen carriers. She compared the five most common oxides that are most widely studied and used in chemical looping combustion. The comparison was based on criteria such as oxygen transport capacity, reactivity, lifetime, safety, and cost. She performed calculations for the distributor plate and prepared drawings for a cold flow model assembly. In addition, she analyzed and interpreted the results of experiments concerning solid circulation rates at various solids inventory masses and gas flow rates in the reactors. In the report, she contributed to the following sections: Abstract; Chapter 2. Literature Review, Section 2.2 Oxygen Carriers; Chapter 3. Methodology, Section 3.1 Design and Assembly of the Cold Flow Model; Chapter 4. Results and Discussion, Section 4.3 Solid Circulation Rate; Chapter 5. Conclusion.

Gulshat Bekkaliyeva contributed to the literature review on cold flow models, with a focus on identifying key system components and common design practices. She performed the dimensioning of the cold flow model, including calculating the required cyclone dimensions, reactor cross-sections, and air distribution system. In addition, she investigated the parameters relevant for analyzing and characterizing system performance, such as pressure drop and solid flux, and established their relationships with input variables (air velocity and solid inventory mass). In the report, she contributed to the following sections: Chapter 2. Literature Review, Section 2.1 Cold Flow Model; Chapter 3. Methodology, Section 3.1 Design and Assembly of the Cold Flow Model, Section 3.2 Experimental Setup and Testing; Chapter 4. Results and Discussion, Section 4.1 Stable Solid Circulation, Section 4.2 Pressure Drop.

Yerzhan Konysbayev contributed to the introduction of the topic by showing the importance of the topic to the current global energy supply and presenting the working principle of CLC. The challenges that may be potentially encountered with the proposed solution are presented. After reviewing the literature related to the CLC cold flow model performance, he analyzed

the data related to the expected results of solid circulation rate and pressure drop. In the discussion part, he presented the future analysis of the relationship between key parameters and problems that should be addressed. In the report, he contributed to the following sections: Chapter 1. Introduction and Problem Statement; Chapter 3. Methodology, Section 3.3 Computational Fluid Dynamics (CFD) simulations; Chapter 4. Results and Discussions, Section 4.4 CFD simulations, Section 4.5 Limitations and Recommendations.

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Appendix A

<i>Particles</i>	<i>Size, d_p (mm)</i>						
	<i>0.02</i>	<i>0.05</i>	<i>0.07</i>	<i>0.10</i>	<i>0.20</i>	<i>0.30</i>	<i>0.40</i>
Sharp sand, $\phi_s = 0.67$	—	0.60	0.59	0.58	0.54	0.50	0.49
Round sand, $\phi_s = 0.86$	—	0.56	0.52	0.48	0.44	0.42	—
Mixed round sand	—	—	0.42	0.42	0.41	—	—
Coal and glass powder	0.72	0.67	0.64	0.62	0.57	0.56	—
Anthracite coal, $\phi_s = 0.63$	—	0.62	0.61	0.60	0.56	0.53	0.51
Absorption carbon	0.74	0.72	0.71	0.69	—	—	—
Fischer-Tropsch catalyst $\phi_s = 0.58$	—	—	—	0.58	0.56	0.55	—
Carborundum	—	0.61	0.59	0.56	0.48	—	—

Figure A.1. Voidage at minimum fluidizing conditions [8]

<i>Type of Particle</i>	<i>Sphericity ϕ_s</i>	<i>Source</i>
Sphere	1.00	(a)
Cube	0.81	(a)
Cylinder		
$h = d$	0.87	(a)
$h = 5d$	0.70	(a)
$h = 10d$	0.58	(a)
Disks		
$h = d/3$	0.76	(a)
$h = d/6$	0.60	(a)
$h = d/10$	0.47	(a)
Activated carbon and silica gels	0.70–0.90	(b)
Broken solids	0.63	(c)
Coal		
anthracite	0.63	(e)
bituminous	0.63*	(e)
natural dust	0.65	(d)
pulverized	0.73	(d)
Cork	0.69	(d)
Glass, crushed, jagged	0.65	(d)
Magnetite, Fischer-Tropsch catalyst	0.58*	(e)
Mica flakes	0.28	(d)
Sand		
round	0.86*	(e)
sharp	0.66*	(e)
old beach	as high as 0.86	(f)
young river	as low as 0.53	(f)
Tungsten powder	0.89	(d)
Wheat	0.85	

Figure A.2. Sphericity of particles [8]