Dynamic CFD Modelling of Calcination in a Rotary Lime Kiln

Jarod Ryan¹, Markus Bussmann¹, Nikolai DeMartini²

¹Department of Mechanical & Industrial Engineering, University of Toronto, Toronto, Canada ²Department of Chemical Engineering & Applied Chemistry, University of Toronto, Toronto, Canada

Abstract— A dynamic 2D axisymmetric CFD model, coupled to a 1D bed model, has been developed to simulate the production of lime in a rotary kiln. The model simulates heat transfer caused by radiation, convection and conduction between the gas, wall, and bed to determine the axial bed temperature in the kiln. The calcination reaction is modelled using a shrinking core model, where the simulation predicts the start of calcination in the kiln, as well as the degree of calcination at the end of the kiln. The dynamic model will help develop a better understanding of the impact that fluctuations in the calcination start location have on mid-kiln ringing. The solid motion within the kiln is modelled using Kramer's equation, modified for transient response. The 2D CFD gas model is implemented in ANSYS Fluent; the 2D gas and 1D bed models are coupled by mass and heat sinks to simulate heat transfer and the calcination reaction. Steady-state and dynamic simulation results are compared to data from an industrial dry lime kiln, and good agreement is found.

Rotary kiln, Kiln modelling, Heat transfer, Numerical simulation, Calcination

I. INTRODUCTION

Rotary kilns are a type of industrial heat exchanger and are widely used in various chemical production industries, including pulp and paper, for processing of granular materials [1]. The majority of rotary kilns are straight, cylindrical vessels that typically operate at 2-5° from the horizontal and rotate between 1-5 rotations per minute [2, 3]. Typically, kilns utilize direct contact heat transfer between the gas phase and the bed material in a counter-current flow configuration (see Fig. 1) [2]. Raw granular material enters on the elevated side of the kiln and travels through the vessel due to gravity as well as tumbling due to rotation. The main function of a lime kiln is to calcine lime mud (CaCO₃) to lime (CaO) by driving off CO₂ through the transfer of heat from the hot combustion gases to the solid bed material. Rotary lime kilns can either be wet kilns, where the bed material enters with moisture, or dry kilns, where an external mud dryer is used to dry and preheat the bed material before entering the kiln.

While rotary kilns allow for long residence time of the material (greater than one hour) to assist in achieving good product quality, operators still face problems in trying to obtain



Figure 1: Rotary kiln with counter flow of the gas and bed

a uniform product [2]. One of the main issues in rotary lime kilns is ringing, where bed material hardens on the inner wall of the kiln. As a ring grows thicker, it forms a dam in the freeboard region, hindering the flow of solids and gas in the kiln. If the ring grows rapidly or becomes too large, the kiln must shut down in order to remove the ring, resulting in production loss. The focus of this work is on mid-kiln rings, which form in the middle of the kiln and are thought to be due to fluctuations of the start of the calcination zone, which in turn is a function of the axial temperature profile of the bed. While a kiln operator will strive to maintain steady-state conditions in the kiln, changes in production are inevitable. Due to the high gas temperatures and moving bed present within kilns, direct measurement of internal conditions is impractical, making control dependent on inlet and outlet measurements. This motivates the study of the transient behaviour of lime kilns, to better understand how to control the quality of the product material, and to investigate the role of fluctuations of the start of the calcination zone on ringing.

While various steady-state models have been developed to study the different processes that occur within rotary kilns, only a few of these models study the dynamics of a kiln [4, 5]. Due to the large difference between the residence time of the gas and bed, being able to model dynamic changes in the bed is vital for trying to keep temperature profiles, and therefore internal processes, constant. Therefore, the objective of this work is to develop a detailed heat transfer model of a dry rotary lime kiln, including the modelling of the calcination reaction, to determine where calcination starts in the kiln. The results of the model are compared to industrial data, to validate the model.

II. METHODOLOGY

A. Gas Model

The gas model is based on the following assumptions:

 The kiln is 2D axisymmetric around the kiln centerline, and so the effects of gravity and buoyancy are neglected.

- There is no physical bed in the model; instead mass and heat sinks are implemented to account for the effects of the bed.
- The burner is symmetric and concentric with the kiln.
- Rotational velocity components of the fuel and air flow are negligible.

The CFD model was generated in ANSYS Fluent version 19.2, and consists of a 2D CFD axisymmetric gas model combined with a 1D mathematical bed model. The realizable $k-\varepsilon$ turbulence model was implemented over the standard $k-\varepsilon$ model due to its improved capability for simulating round jets as well as flows with recirculation [6], both of which are important in the burner section of the kiln. The standard wall functions are utilized as they have often been used for industrial flows and work well for a variety of wall-bounded flows [6].

While industrial kilns generally use natural gas as a fuel source, methane gas is modelled for simplicity. A mixture of methane-air is specified in ANSYS Fluent. While non-premixed combustion is the ideal choice for combustion modelling due to fast computation time, it does not allow for modelling a mass source for individual species. This is an issue as CO₂ from the bed is released to the gas phase during the calcination reaction. Therefore, combustion is modelled using finite rate chemistry. Radiation is modelled using the Discrete Ordinates (DO) model, which is more computationally expensive than the P1 radiation model, but the P1 model may overpredict radiation from localized heat sources [6]. Since this CFD model includes the combustion of methane, the DO model was determined to be the best choice.

Since the gas mixture contains both H_2O and CO_2 which absorb radiation in distinct bands, the use of a gray-gas approximation will not be accurate. Therefore, the weightedsum-of-gray-gases model (WSGGM) was used in ANSYS Fluent. Since the height of the bed is small compared to the kiln diameter, the path length chosen for the WSGGM is the diameter of the kiln.

The burner consists of fuel, primary air and secondary air inlets on the hot end of the kiln. The fuel inlet is centered in the middle of the kiln, with the primary air inlet surrounding the fuel inlet; the secondary air inlet spans from the outer edge of the burner to the refractory wall (see Fig. 2).



B. Bed Model

The bed model is based on the following assumptions:

- The bed material is 100% calcium carbonate (CaCO₃)
- Calcination occurs as a shrinking core reaction
- Bed particles are assumed to be mono-sized and spherical

The steady-state height and volumetric flow rate of the bed is determined using Kramer's equation [7]. The volumetric flow rate through a cross-section is determined as follows [8]:

$$Q = \frac{R^3 \sin^3 \phi}{C_A} \left(R \sin \phi \frac{d\phi}{dz} + C_B \right)$$
(1)

Rearranging the equation gives a non-linear first order ordinary differential equation (ODE) with constant coefficients for the half angle of the bed:

$$\frac{d\phi}{dz} = \frac{C_A Q}{R^4 \sin^4 \phi} - \frac{C_B}{R \sin \phi}$$
(2)

In (2), Q is the volumetric flow rate of the bed material in the kiln, ϕ is the half angle of the bed, R is the inner radius of the kiln, z is the length down the kiln, and C_A is a constant coefficient given as:

$$C_{A} = \frac{3\tan\gamma}{4\pi n}$$
(3)

where γ is the static angle of repose of the bed material in radians, and n is the rotation speed of the kiln in revolutions per second. Finally, C_B is another constant coefficient given as:

$$C_{\rm B} = \frac{\tan\beta}{\tan\gamma} \tag{4}$$

where β is the inclination of the kiln axis from the horizontal in radians. The single boundary condition required for the solution is obtained from the assumption that the bed height at the discharge end is equal to the discharge dam height.

For the dynamic bed height, Spurling, Davidson, and Scott developed an unsteady bed height equation in the form of an unsteady heat diffusion equation [8]:

$$\rho C \frac{\partial \phi}{\partial t} = \frac{\partial}{\partial z} \left(k \frac{\partial \phi}{\partial z} \right) + S \tag{5}$$

The coefficients and source term in (5) are given by:

$$\rho C = 2R^2 \sin^2 \phi \tag{6}$$

$$\mathbf{k} = \frac{\mathbf{R}^4 \sin^4 \phi}{\mathbf{C}_{\mathrm{A}}} \tag{7}$$

$$\mathbf{S} = \left(\frac{3C_{\rm B}R^3}{C_{\rm A}}\right) \cos\phi\sin^2\phi\left(\frac{\partial\phi}{\partial z}\right)_{\rm t} \tag{8}$$

For the single initial condition, the steady-state bed height is used for the initial bed height profile. Two boundary conditions are required; as stated earlier for the steady-state equation, the bed height at the discharge end is equal to the discharge dam height, and the volumetric feed rate at the inlet is equal to the load coming into the kiln. The unsteady bed height was solved numerically using a discretization method derived from a control-volume formulation, as described by Patankar [9].

The mass balance of the bed at an axial position of $z + \Delta z$ is:

$$\dot{\mathbf{m}}_{\mathbf{b},z+\Delta z} = \dot{\mathbf{m}}_{\mathbf{b},z} - \dot{\mathbf{m}}_{\mathrm{CO}_{2},z} \tag{9}$$

In (9), $\dot{m}_{b,z}$ is the mass flow of bed material, and $\dot{m}_{CO_2,z}$ is the amount of CO₂ that is released due to decomposition. The energy balance of the bed is given as:

$$\rho_{b}V_{b}C_{p,b}\frac{dT_{b}}{dt} + \dot{m}_{b}C_{p,b}\frac{dT_{b}}{dz} = \dot{Q}_{bed} - \dot{Q}_{dec}$$
(10)

In (10), ρ_b is the density of the bed material, V_b is the volume of the bed, $C_{p,b}$ is the specific heat of the bed, T_b is the bulk bed temperature, \dot{Q}_{bed} is the total amount of heat transferred to the bed, and \dot{Q}_{dec} is the heat required for decomposition.

C. Heat Transfer

1) Radiation

Radiation from the gas to the exposed bed is calculated using the equation developed by Hottel and Sarofim [10]:

$$\dot{Q}_{g \to eb}^{r} = A_{eb} \sigma \frac{(\varepsilon_{b} + 1)}{2} \left(\varepsilon_{g} T_{g}^{4} - \alpha_{g} T_{b}^{4} \right)$$
(11)

In (11), A_{eb} is the area of the exposed bed, σ is Stefan's constant, ε_b is the emissivity of the bed, ε_g is the emissivity of the gas, α_g is the absorptivity of the gas, and T_g is the average axial temperature of the gas. This equation is derived from an expression for radiative heat transfer from a gas to a black surface, which is then multiplied by a correction factor to account for the emissivity of the surface. Hottel and Sarofim [10] determined that using this equation to account for radiative heat transfer from the gas to surrounding surfaces results in a maximum error of 10%, as long as the emissivity of the bed and wall are above 0.8. However, in the hot end of the kiln, both the bed and wall emissivity are estimated to drop below 0.8. Nevertheless, due to the lack of further equations to estimate the radiative heat transfer, the equation developed by Hottel and Sarofim is used in the simulation.

Radiation between the exposed wall and exposed bed is determined by the following equation:

$$\dot{Q}_{ew \to eb}^{r} = A_{eb} \sigma \epsilon_{w} \epsilon_{b} \Omega \left(T_{w}^{4} - T_{b}^{4} \right)$$
(12)

In (12), ε_w is the emissivity of the inner refractory wall, T_w is the temperature of the inner refractory wall, and Ω is the form factor for radiation which is calculated as:

$$\Omega = \frac{L_{eb}}{2(\pi - \phi)R} \tag{13}$$

where L_{eb} is the length of the exposed bed.

2) Conduction

The conductive heat transfer from the covered wall to the bed is as follows:

$$\dot{Q}_{cw\to cb}^{cd} = A_{cb}h_{cw\to cb}(T_w - T_b)$$
(14)

where A_{cb} is the area of the bed in contact with the wall. Schlünder and Mollekopf [11] estimate the heat transfer coefficient from the wall to a particle bed as:

$$\frac{1}{h_{cw\to cb}} = \frac{1}{h_{con}} + \frac{1}{h_{pen}}$$
(15)

where h_{pen} is the penetration heat transfer coefficient and is derived through the contact of two surfaces:

$$h_{pen} = 2\sqrt{\frac{k_b \rho_b C_{p,b}}{2\pi \phi}}$$
(16)

where k_b is the thermal conductivity of the bed. In (15), h_{con} is the contact heat transfer coefficient between the wall and bed, which accounts for: conduction through wall-particle contact, radiation between the wall and particle, and heat transfer through the gas. Details of this equation can be found in [11].

3) Convection

The convective heat transfer from the gas to the exposed bed is as follows:

$$\dot{Q}_{g\to eb}^{ev} = A_{eb} h_{g\to eb} (T_g - T_b)$$
(17)

Tscheng and Watkinson [12] derived the convective heat transfer coefficient as:

$$h_{g\to eb} = \frac{k_g}{D_e} 0.46 Re_g^{0.535} Re_{\omega}^{0.104} f^{-0.341}$$
(18)

where Re_{g} and $\operatorname{Re}_{\omega}$ are defined as:

$$\operatorname{Re}_{g} = \frac{\operatorname{V}_{g} \operatorname{D}_{e}}{\operatorname{v}_{g}}$$
(19)

$$\operatorname{Re}_{\omega} = \frac{D_{e}^{2}\omega}{v_{g}}$$
(20)

with V_g being the velocity of the gas, D_e is the equivalent diameter in the gas phase, ω is the kiln rotational velocity, v_g is the kinematic viscosity of the gas, and f is the filling degree in the kiln. Equation (18) is valid for $1600 < \text{Re}_g < 7800$ and $20 < \text{Re}_{\omega} < 800$ [12]. Although Re_g can be greater than 7800 in large kilns, due to the lack of other equations to estimate the convective heat transfer, (18) is used in the simulation.

4) Decomposition

The decomposition or calcination reaction is simulated using a shrinking core model. The rate of reaction is proportional to the area of the reaction front, the difference between the CO_2 concentration at the reaction front and in the gas, diffusion resistance, and boundary layer resistance. The rate of reaction is given by [13]:

$$\dot{m}_{CaCO_3} = 4\pi r_{p,z}^2 N_p M_{CaCO_3} \frac{C_{CO_2,s1} - C_{CO_2,\infty}}{R_{1,c} + R_{2,c}}$$
(21)

with $r_{p,z}$ being the radius of the unreacted particle, N_p the total number of particles, $C_{CO_2,s1}$ the CO₂ concentration at the reaction front, $C_{CO_2,\infty}$ the CO₂ concentration in the gas phase, $R_{1,c}$ the diffusion resistance, and $R_{2,c}$ the resistance due to the boundary layer between the particle and gas phase. Details of this equation can be found in [13]. The heat required for decomposition can then be determined by:

$$\dot{Q}_{dec} = \dot{m}_{CaCO_3} \Delta h \tag{22}$$

where the reaction enthalpy, Δh , is specified as 1630 kJ/kg CaCO₃ [14]. The percent calcination, or lime availability, at the end of the kiln can be calculated based on the remaining CaCO₃ in the bed:

$$PC = \left(1 - \frac{r_{p,2}^3}{r_{p,0}^3}\right) \cdot 100\%$$
 (23)

where $r_{p,0}$ is the initial radius of the particle.

5) Heat Loss

Heat loss from the outer shell to the ambient environment can occur by both convection and radiation. The following equation is used to calculate the convective heat transfer coefficient [15]:

$$h_{sh\to a} = 0.11 \frac{k_a P r^{0.36}}{D_{sh}} \left(0.5 R e_{\omega}^2 + R e_a^2 + G r \right)^{0.35}$$
(24)

where D_{sh} is the diameter of the outer shell, Pr is the Prandtl number, Re_a is the gas Reynolds number, and Gr is the Grashof number, all for the ambient environment. The total heat loss from the shell to the ambient environment can then be calculated as:

$$\dot{Q}_{loss} = A_{sh}h_{sh\to a}(T_{sh} - T_a) + A_{sh}\sigma\epsilon_{sh}(T_{sh}^4 - T_a^4) \qquad (25)$$

where A_{sh} is the surface area of the shell, T_{sh} is the temperature of the shell, T_a is the temperature of the ambient environment, and ε_{sh} is the emissivity of the shell.

6) Coupling of the Bed and Gas Models

In order to simulate the bed in the CFD model, mass and energy equations must be coupled. Therefore, the CFD model includes one mass source as well as one energy source. The CO₂ mass source and energy source at an axial position z are given as the following, respectively:

$$S_{m,CO_2,z} = \frac{\dot{m}_{CO_2,z}}{\pi R^2 \Delta z}$$
(26)

$$\mathbf{S}_{e,z} = \frac{\dot{\mathbf{Q}}_{bed,z} + \dot{\mathbf{Q}}_{CO_2,z}}{\pi R^2 \Delta z}$$
(27)

where $Q_{CO_{2,z}}$ represents the energy added to the gas phase by the addition of the mass. In Fluent, mass sources enter the domain with no momentum or thermal heat, and so the energy source is required in order to add the mass to the gas domain at the proper temperature. It is assumed that when CO₂ enters the gas phase, it is simultaneously transferred to the total area at an axial position z. The equation for $\dot{Q}_{CO_{2,z}}$ is:

$$\dot{Q}_{CO_{2},z} = \dot{m}_{CO_{2},z} \left(H_{CO_{2},T_{b,z}} - H_{CO_{2},T_{ref}} \right)$$
 (28)

where $H_{CO_2,T_{b,z}}$ is the enthalpy of the CO_2 at the bed temperature, and $H_{CO_2,T_{ref}}$ is the enthalpy of the CO_2 at the reference temperature, defined in Fluent as 298.15 K. The mass and energy source terms are volumetric, so they are divided by the internal volume of the kiln at an axial position *z*.

III. RESULTS AND DISCUSSION

A. Steady-State Results

Fig. 3 shows the bed and average axial gas temperatures along the industrial dry kiln that was modelled. A volume average at each axial location is used to determine the axial gas temperature. The simulated peak average gas temperature is slightly above 1800°C, which agrees well with previous literature on gas temperatures in rotary lime kilns [16, 17]. In addition, since the adiabatic flame temperature of methane is around 2000°C [18], a peak average gas temperature slightly above 1800°C is reasonable. Going from left to right, the gas temperature first rises when combustion occurs, then reaches a peak 16 meters into the kiln (or 7 meters from the tip of the burner), then slowly decreases during the rest of the length of the kiln as heat is transferred from the hot gas to the bed. Going



Figure 3: Simulated bed and average axial gas temperature

from right to left, the bed enters at 327° C and then the temperature increases almost linearly until calcination begins at 57.2 meters, where the slope changes due to the calcination reaction and heat is now absorbed by the reaction instead of heating the bed. The product bed temperature is around 1000°C, which agrees with other model outputs ranging between 900°C to 1100°C [19]–[20], which depend on: the kiln, the calcination model being used, and whether calcination is assumed to be fully complete or not. Finally, the simulated outlet gas temperature of 750°C exactly matches the measured outlet gas temperature of 750°C.

Fig. 4 shows good agreement between the measured and simulated outer shell temperatures along the length of the kiln, which helps to validate the heat loss calculation in the model. There is some variability in the measured outer shell temperature, which may be due to bearings and other components on the kiln since the temperature is measured by thermal cameras.

B. Dynamic Results

Fig. 5 shows the load entering the kiln and the measured fuel flow rate over a period of one week. During this time period, a ring near the cold end of the kiln grew rapidly, as seen in Fig. 6. Fig. 6 (top) displays a thermal scan of the outer shell during regular operation, while Fig. 6 (bottom) shows the thermal scan of the outer shell when a ring has formed, as indicated in blue on the right of the kiln (a ring appears colder on a thermal scan as it insulates the kiln in that location.)

Fig. 7 shows the measured and simulated outlet gas temperatures over the seven day span. As seen in Fig. 7, there is very good agreement between the measured and simulated data.



Figure 4: Measured and simulated outer shell temperature



Figure 5: Measured load and fuel flow rate over one week in the kiln



Figure 6: Outer shell thermal scan during regular operation (top) and when a ring has grown (bottom)

In Fig. 5, a drop in both load and fuel occurs on the 5th of January, and this drop in production can be seen in the outlet gas temperature as the measured and simulated gas temperatures drop significantly over that period. As seen in Fig. 5, the load and fuel are slowly increased after this drop in production, in order to bring the kiln back up to regular operation. During this time, the outlet gas temperature, measured and simulated, also begin to increase slowly during this time period.

Fig. 8 shows the location of the start of calcination in the kiln over the seven day span. During the first two days, the simulation shows little fluctuation in the calcination location. However, beginning on January 2^{nd} , the calcination start location begins to move towards the cold end of the kiln. The total shift in the calcination location is roughly 6 meters over the span of a couple of days. During the shutdown, the calcination start location start location is a cut-off in the data. However, when the kiln starts up again, the calcination start location is shifted towards the hot end of the



Figure 7: Measured and simulated outlet gas temperature over one week in the kiln



Figure 8: Calcination start location from the hot end of the kiln over one week

kiln, and slowly increases back to regular operation by the end of the simulation.

As seen in Fig. 8, there is a large shift in the calcination location, during the same time that ringing is a concern in the kiln. While there is a correlation between these two variables for the given time period, more data will have to be examined in order to validate this correlation. Future work will involve modelling the kiln over longer time periods to attempt to correlate fluctuations in the calcination location with ring growth.

IV. CONCLUSION

A comprehensive 2D axisymmetric CFD model combined with a 1D bed model was developed to simulate calcination and heat transfer in a rotary lime kiln. Steady-state and dynamic results for the kiln are in good agreement with mill data, indicating that the assumptions used for calcination and heat transfer in the model are reasonable. The calcination start location shifts towards the cold end of the kiln when rapid ring growth appears in the kiln. However, more data will be required in order to correlate any fluctuation in the calcination location as the cause for ring growth in a lime kiln.

REFERENCES

- P. R. Davies, M. J. S. Norton, D. I. Wilson, J. F. Davidson, and D. M. Scott, "Gas flow in rotary kilns," *Particuology*, vol. 8, no. 6, pp. 613– 616, Dec. 2010
- [2] A. A. Boateng, "1 The Rotary Kiln Evolution and Phenomenon," in *Rotary Kilns (Second Edition)*, A. A. Boateng, Ed. Boston: Butterworth-Heinemann, 2016, pp. 1–11

- [3] C. Csernyei and A. G. Straatman, "Numerical modeling of a rotary cement kiln with improvements to shell cooling," *International Journal of Heat and Mass Transfer*, vol. 102, pp. 610–621, Nov. 2016
- [4] H. A. Spang, "A dynamic model of a cement kiln," Automatica, vol. 8, no. 3, pp. 309–323, May 1972
- [5] T. Ginsberg and M. Modigell, "Dynamic modelling of a rotary kiln for calcination of titanium dioxide white pigment," *Computers & Chemical Engineering*, vol. 35, no. 11, pp. 2437–2446, Nov. 2011
- [6] "ANSYS FLUENT 12.0 User's Guide," 2009.
- [7] H. Kramers and P. Croockewit, "The passage of granular solids through inclined rotary kilns," 1952
- [8] R. J. Spurling, J. F. Davidson, and D. M. Scott, "The Transient Response of Granular Flows in an Inclined Rotating Cylinder," *Chemical Engineering Research and Design*, vol. 79, no. 1, pp. 51–61, Jan. 2001
- [9] S. V. Patankar, Numerical Heat Transfer and Fluid Flow. Boca Raton: CRC Press, 2018
- [10] E. R. G. Eckert, "Radiative transfer," *AIChE Journal*, vol. 15, no. 5, pp. 794–796, 1969
- [11] E.-U. Schlünder and N. Mollekopf, "Vacuum contact drying of free flowing mechanically agitated particulate material," *Chemical Engineering and Processing: Process Intensification*, vol. 18, no. 2, pp. 93–111, Mar. 1984
- [12] S. H. Tscheng and A. P. Watkinson, "Convective heat transfer in a rotary kiln," *The Canadian Journal of Chemical Engineering*, vol. 57, no. 4, pp. 433–443, 1979
- [13] T. Bluhm-Drenhaus, E. Simsek, S. Wirtz, and V. Scherer, "A coupled fluid dynamic-discrete element simulation of heat and mass transfer in a lime shaft kiln," *Chemical Engineering Science*, vol. 65, no. 9, pp. 2821–2834, May 2010
- [14] P. C. G. Sandaka, Calcination Behavior of Lumpy Limestones from Different Origins. docupoint Verlag, 2016.
- [15] S.-Q. Li, L.-B. Ma, W. Wan, and Q. Yao, "A Mathematical Model of Heat Transfer in a Rotary Kiln Thermo-Reactor," *Chemical Engineering & Technology*, vol. 28, no. 12, pp. 1480–1489, 2005
- [16] H. Shahin, S. Hassanpour, and A. Saboonchi, "Thermal energy analysis of a lime production process: Rotary kiln, preheater and cooler," *Energy Conversion and Management*, vol. 114, pp. 110–121, Apr. 2016
- [17] H. F. Elattar, E. Specht, A. Fouda, and A. S. Bin-Mahfouz, "Study of Parameters Influencing Fluid Flow and Wall Hot Spots in Rotary Kilns using CFD," *The Canadian Journal of Chemical Engineering*, vol. 94, no. 2, pp. 355–367, Feb. 2016
- [18] R. Pupo, "Adiabatic Flame Temperature for Combustion of Methane II," Undergraduate Journal of Mathematical Modeling: One + Two, vol. 5, Jan. 2013
- [19] A. Agrawal and P. S. Ghoshdastidar, "Computer Simulation of Heat Transfer in a Rotary Lime Kiln," *Journal of Thermal Science and Engineering Applications*, vol. 10, Feb. 2018
- [20] P. V. Barr, "Heat transfer processes in Rotary kilns," University of British Columbia, 1986