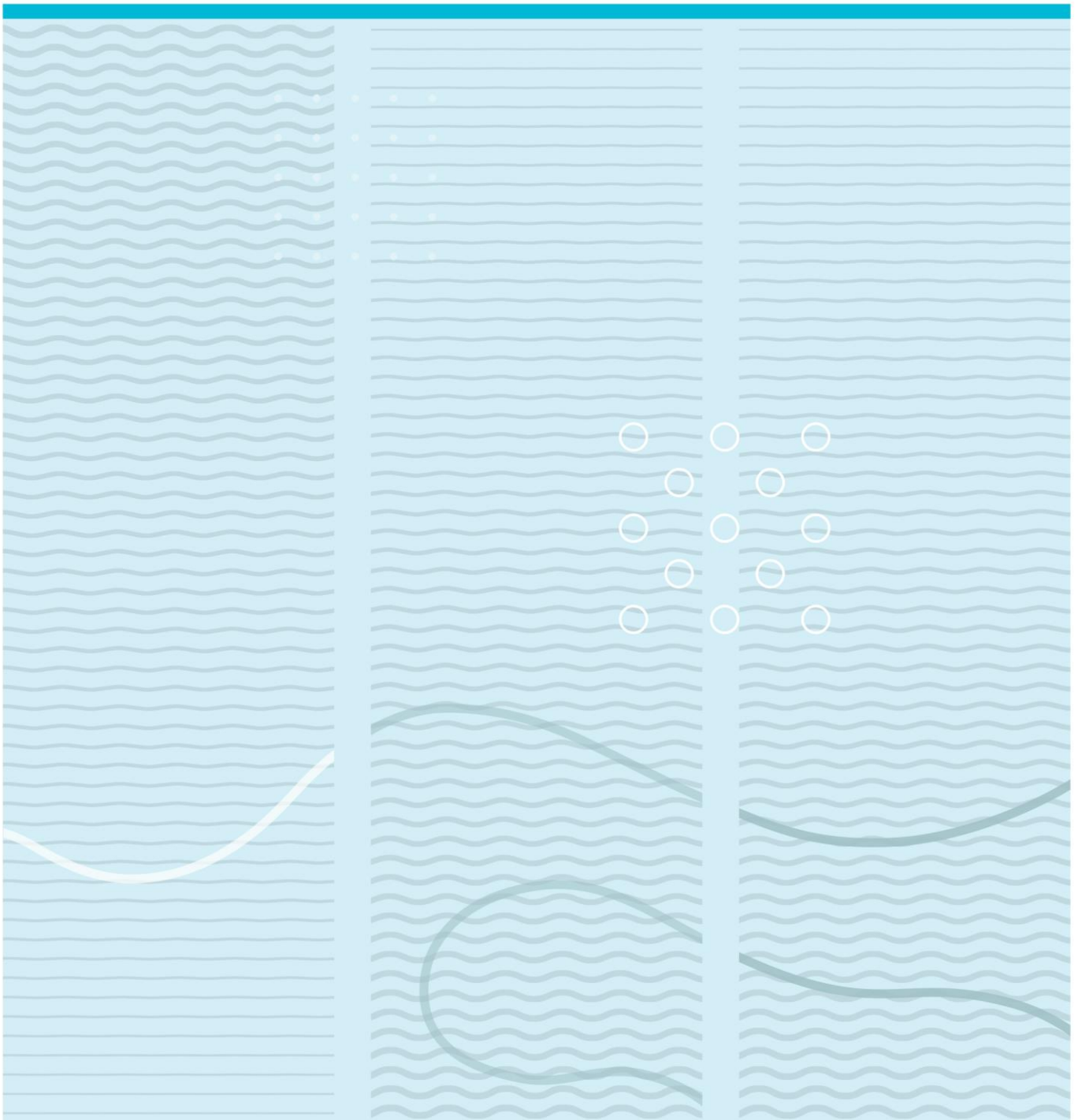


Fatemeh Asadi

Drying of Refuse-Derived Fuel (RDF)



University College of Southeast Norway
Faculty of Technology

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MASTER'S THESIS, COURSE CODE FMH606

Student: Fatemeh Asadi
 Thesis title: Drying of Refuse-Derived fuel (RDF)
 Signature:
 Number of pages: -
 Keywords: RDF drying, steam drying, air drying, belt dryers,
 rotary dryers, dryer cost, inversion temperature...

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Abstract:

Refuse-Derived fuel (RDF) application in high energy demanding plants as a secondary fuel has risen over the past decades, causing CO₂ emission reduction and fuel cost saving. Lowering the moisture content of this solid fuel is essential to improve RDF heating value (LHV) and to avoid more steam generation in the combustor. Therefore, RDF drying is needed.

A feasibility study on RDF direct & indirect steam drying vs. conventional air drying systems was conducted in this project by taking a cement plant data as a case study.

Indirect steam drying option was eliminated mainly due to poor drying efficiency for heterogeneous and varying characteristic solids like RDF, in practical tests. The system suggested for direct steam drying of RDF showed that almost 70t/h superheated steam has to flow in a drying loop to evaporate almost 3.2t/h of moisture. Release of this amount of moisture resulted in RDF heating value increase by 32%. However, there are some uncertainties regarding the mentioned system due to lack of pilot results for RDF steam drying. Inherent controllability and operability challenges for steam processing set-ups and further equipment cost for superheated steam generation and temperature rise via an air/steam heat exchanger, left air drying option as the best solution for the studied case.

Available dry and almost dust free hot air in the plant could serve as the drying gas in the air dryer. A comparison between belt dryers and rotary drum dryers was done through several received quotations from vendors. Low temperature belt dryer technology outweighed rotary types in terms of capital cost (0.5-1.5 M€ for quoted belt dryers and 2-3 M€ for rotary dryer) and operational cost. In addition the amount of environmental footprint like dust, VOC emissions and exhaust air temperature was less in belt dryers. However, low-temperature fluid-bed dryers were excluded from the comparison list which should be considered as a potential competitor for belt counterparts.

University College of Southeast Norway accepts no responsibility for results and conclusions presented in this report.

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Foreword

This master thesis is submitted to Høgskolen i Sørøst-Norge (HSN) as the writer's final project for completion of master degree in process technology. The subject was initiated by Professor Lars Andre Tokheim, as the supervisor.

In this master thesis, different process options for refuse-derived fuel (RDF) drying have been investigated. RDF, which can be used as a supplementary fuel for cement plants, should be dried to gain a higher heat value, increasing the thermal efficiency of the plant. The reader can find technical and commercial comparisons, thermodynamic calculations and sketches of typical dryers in the text. The thesis contains some commercial figures which have been confidentially submitted to the supervisor and they are excluded from the public copy.

I would like to express my gratitude to Professor Lars Andre Tokheim for his supports during the project and to Stela GmbH, ANDRITZ GmbH and Solution Bring DA, which fully supported me with technical know-how and business expertise.

. . . and my sincerest thanks to my husband, Saman, who helped in all moments of the project.

HSN, Porsgrunn/03.06.2016
Fateme Asadi

1. Introduction

In the past few decades, the need for alternative fuels has risen to substitute the use of conventional fossil fuels in energy intensive industries such as cement production and power generation plants. This could be due to environmental regulations and/or fuels costs. The focus of this report is on Refuse-Derived Fuels (RDFs) substitution as a sub-category of municipal wastes. However, fossil fuel replacement, partly or completely, is not as simple as it seems and RDFs need some treatment to become ready to burn as a fuel.

The lower the moisture content is, the more heating value RDF has which is favorable for a supplementary fuel. Therefore, RDF drying is essential to reach the desired moisture content which is the main target of this project.

This report includes general information about solid drying concepts and methods, in the beginning chapters. Various drying classifications exist in which hot air or (superheated) steam act as the drying medium. This could happen in different types of drying systems in direct or indirect modes of heat transfer, although only commercial methods of RDF drying are more described. However, steam drying as an option for RDF drying and the main concept behind are brought more into discussion. Steam drying pros and cons are also given to have a better comparison between air drying and steam drying systems.

In order to perform a more sensible feasibility study on RDF steam drying vs. air drying method, Norcem cement plant data is taken as a case study for this project. A discussion on various RDF drying systems to be utilized in this plant could be found from chapter 4 on. Available excess heat utilization in the plant for drying purposes and making the system more energy efficient and environmental-friendly are the sub-targets of this thesis.

Mass & energy balance, based on plant data and suggested design basis, for the drying system will be also given. In addition; selection of a drying method will be found in chapter 5.

Several efforts were made during the project to contact different vendors to receive more accurate and practical information about commercial RDF dryers size and price; as well as their operational conditions which could be found in chapter 6. Several budgetary quotes were received which include preliminary designs of suggested dryers that can be used in Norcem plant. Albeit, a hand sizing-calculation is also done for one dryer type and

results are compared with the given size recommended by the vendor. Due to legal terms and conditions, vendors' quotations are not attached to the public copy of this report.

2. Refuse-Derived Fuel (RDF)

In the past few decades, the need for alternative fuels has risen to substitute conventional fossil fuels, more specifically in the power generation industry and co-firing cement production plants. Environmental regulations and/or fuel cost could be the key reasons to think about fossil fuel substitutions. Co-processing of alternative fuels, especially in the cement kiln industry, leads to dramatic fall in CO₂ emission as the main contributing gas to global warming[1].

Refuse-Derived Fuels (RDFs) are types of derived materials from Municipal Solid Wastes (MSW) through various processes such as: separation at source, sorting, shredding, screening, blending, drying and pelletizing. Screening process is to separate ferrous compounds and glass which are recyclable; as well as bio wastes (e.g. food) which are highly wet. Pelletizing process can be also ignored depending on the location of RDF production systems and combustion facilities[2]. As shown in Figure 2-1, RDF may consist of degradable/non-degradable wastes, with decent calorific values, such as: paper, textiles, cardboards, plastics, woods and etc.



Figure 2-1 RDF as an alternative fuel[3]

RDF composition highly depends on its origin; as a result, it may vary considerably[2]. However, a typical composition of RDF is presented in Table 2-1.

Several properties of RDF such as calorific value, moisture content, and the amount of chlorine, sulfur and alkali compounds and ash content are important when it is used in cement production plants, as an example. Compared to coal, RDF has lower calorific value and sulfur content and higher chlorine amount[3]. The lower amount of sulfur is favorable to reduce sulfur dioxide emission to meet environmental regulations. However, there are some difficulties with the high chlorine content in the RDF and its effect on the strength

of concrete. Specialists resolve this deficiency by limiting the portion of RDF to be used. Further information about the effect of RDF on concrete quality (not the scope of this project) could be found in the research done by *Mustafa Kara* [1].

Table 2-1 Typical RDF Composition[2]

RDF Compositions	Amount (%)
Plastics	31
Textiles	14
Paper/Cardboard	13
Wood fractions	12
Others	30

Finding invariant values for the mentioned properties is difficult because they are highly dependent on the waste source. For instance, calorific values for various sources of RDF are shown in Table 2-2.

Table 2-2 Calorific values for different RDF sources[2]

RDF Source	Calorific Value(MJ/kg)
Household waste	12-16
Commercial waste	16-20
Industrial waste	18-21
Demolition waste	14-15

Among the mentioned specifications, the main purpose of this project is to lower the water content of the solid fuel to enhance the heating value. So that, proper drying method should be applied to make the excess water evaporate from RDF, which will be more explained in the next chapter.

3. Solid drying

Studies on solid wastes drying have developed increasingly and new techniques together with the conventional types of drying have been recently brought more into discussions. However, drying of solid wastes is still challenging because of process complexities and some variations in solid heterogeneity and physical/chemical and biological properties. As a result, many process variables need to be considered to efficiently dry materials to the desired levels. This means that there may be no single drying method suitable for solid wastes[4].

In this chapter, some characteristics of solid wastes together with the concept of drying mechanism and common classifications for drying methods as well as different dryer types will be discussed followed by an overview on steam drying concept and functionality.

3.1 Characteristics of solid wastes

Solid wastes could have variety of characteristics and properties that may be different, from solid to solid. Heterogeneity is one of them, meaning that the material properties are highly dependent on spatial position. It is due to different origins of wastes which can cause complexity in the process[4].

Another solid characteristic is hydrophilic property that is not the same in different solid wastes[4]. *Merriam Webster dictionary* defines hydrophilic as an adjective meaning having a strong affinity for water. So it shows how solid components like to dissolve in water or water solvents.

Size of solids, their shape and porosity and surface area as well as specific mass and heat capacities are other important elements in selecting a type of dryer. In addition, the heat and mass transfer ability of solids should be taken into account[4].

Moreover, being anisotropic is another important property of solids[4], which means that its properties are not the same in any directions, based on *Merriam Webster* definition. Momentum, heat and mass transfer coefficient of anisotropic solids are dependent on the transfer direction. This property is important in waste drying to some extent especially when the goal is to reduce the energy cost and increase the drying efficiency. For instance, in fixed bed dryers, the permeability of the waste material bed might be more in axial direction drying than radial direction[4].

Hygroscopic or non-hygroscopic property of solid wastes which are based on water/solid interaction are other features to be paid attention. Non-hygroscopic solid wastes do not keep the ambient moisture by pore adsorption and surface adsorption; as a result, the internal/surface vapor pressure of these materials is as equal as pure component and is a function of temperature only. On the contrary, hygroscopic solid wastes retain the water in their structures and the vapor pressure is less than pure water at the same temperature. In hygroscopic materials the solid moisture diffuses out to the surface where it evaporates to the gas phase. Organic materials like fruit and vegetable wastes and some inorganic ones like paper, textile and rubbers are in the category of hygroscopic and other inorganic wastes like polymeric ones are categorized as non-hygroscopic group. Generally, moisture removal from hygroscopic materials is more difficult than materials with free moisture (non-hygroscopic ones)[4].

These sorts of information are required principally to choose a proper dryer and suitable operating conditions. Apart from them, the level of drying and operating cost are also needed to be considered in selecting the most appropriate method. Drying operating cost can be substantial for hygroscopic solid wastes, using air as the drying medium. Because mostly high temperature gas with low humidity is a requirement which is energy consuming. Non-hygroscopic wastes may also need high operating temperature and gas flow rate when the initial moisture is high and the air temperature is as equal as wet bulb temperature while surface moisture is removed[4].

3.2 Drying mechanism

Drying mechanism can be explained by the amount of evaporated moisture in time. Figure 3-1 illustrates a drying-rate curve for a hygroscopic product. However, this curve could be different for various solids depending on their water content.

As Figure 3-1 shows, there are three stages of drying; although, the transition between stages is not as sharp as what is depicted, in reality. The rate of drying is constant in the beginning when the free moisture is released from the surface. In this period, the rate of drying is controlled by water diffusion to the air-water interface or by the evaporation rate. As approaching to the end of this stage, almost all surface moisture evaporates and internal moisture diffuses outward from the solid inside to the surface. The rate is

constant until the critical moisture content (X_{cr}) is reached and more drying creates dry spots on the surface.

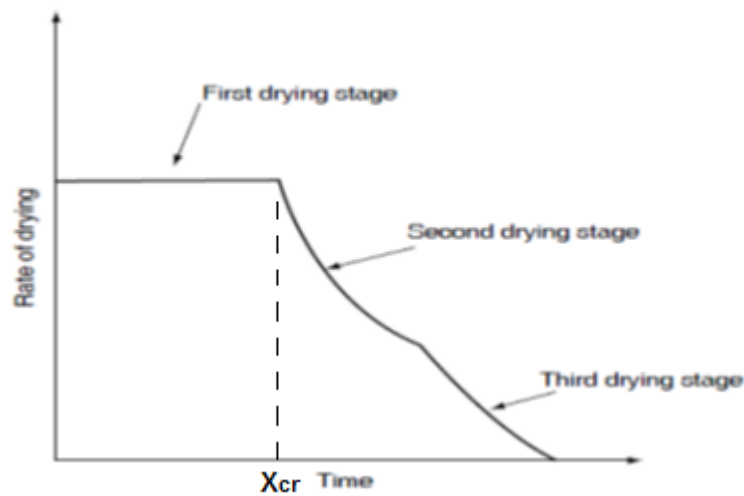


Figure 3-1 typical drying-rate curve[5]

Then, the first falling drying rate period or the second drying stage starts. This period lasts until the moisture film on the surface evaporates completely. Some solids may contain this part as the whole falling rate stage and some may miss this part. As drying proceeds, the second falling rate period or the third drying stage appears. At this stage, concentration gradient makes the moisture diffuse through the solid to the surface. Heat is transferred first to the surface and then conduction occurs in the solid. So that, heat conduction affects the drying rate a lot. However, if the solid density is high and if it consists of small pores, the drying rate is more controlled by the mass diffusion through the solid and the mass transfer from the surface to the air. At this stage, the drying rate drops faster than the last stage and it continues until the moisture content drops to the equilibrium value which is the air humidity content (in drying with hot air) and then it stops.

The solid material may have high amount of moisture or the product may need to have more amount of residual water content. In that case drying will probably take place in the first constant stage. However, both constant and falling rate periods may exist in most cases. Materials which are slow-drying are dried mainly in the falling-rate period. Falling-rate period is also important when the initial moisture is low. The time of drying process is also long for this type of solids.

If the heat transfer rate is increased by the air velocity, air velocity becomes an important factor. Other parameters such as air temperature, air humidity, thickness of the solid and the depth of bed play essential roles[6].

Anyway, a drying rate curve for RDF is only achieved via experiments and it could vary as a result of various RDF sources and characteristics. However, performing an experiment to find a proper drying rate curve did not happen in this project.

3.3 Drying method classifications

There are different dryer classifications to help selecting a proper drying method. This diversity could stem from various solid waste types, their origins and challenges in working with non-homogeneous systems. Some categories may be traditionally based on heat transfer methods and some are based on solid characteristics. Heat transfer from the drying gas to the solid could be via conduction, convection, radiation and microwaving. Thus, it also offers a wide range of drying technologies[4].

Handbook of Perry's chemical engineering suggests a basic classification, taking the heat transfer modes into account, as presented in Figure 3-2. In this sort of classification, the three main categories of dryers are direct dryers, indirect dryers and Infrared radiant heat dryers which are briefly discussed in the next parts[5]. Dryers are divided into batch and continuous mode of processing among which continuous process with the drying mode of direct and indirect is more focused in this report.

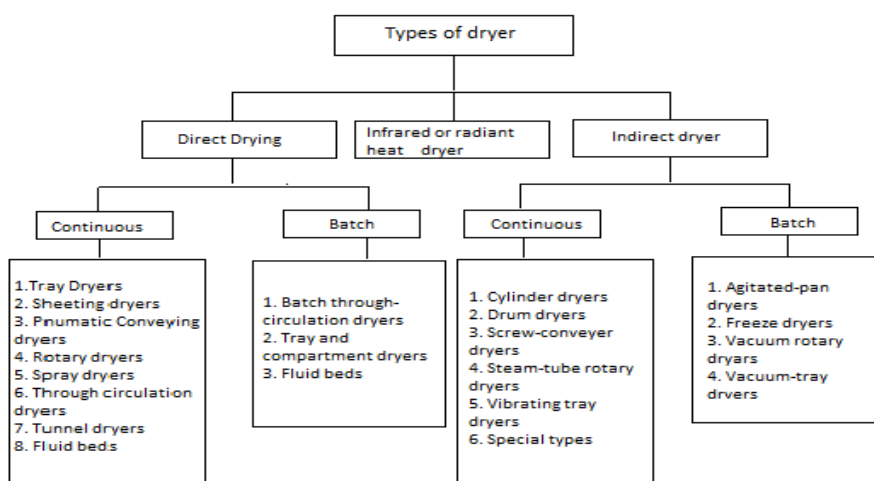


Figure 3-2 Dryer classification based on modes of heat transfer[5]

3.3.1 Direct drying method

In direct drying, heat is transferred through a direct contact between moist solid and the hot drying gas. The hot gas acts as a drying medium and the moisture carrier. Direct drying may be also named as convective drying method. There are some operational characteristics suggested by *Perry's handbook* while working with this type of method and some of them are described as follows:

- When the temperature of the gas is below the boiling point of the moisture, the drying rate and final water content of product are affected by the gas vapor pressure. However, if the temperature is above the boiling point, gas vapor content is not as effective as the other case. That is why superheated vapor of the liquid to be dried could be also applied for drying purposes.
- In case of drying with air, air needs to be dehumidified especially for low-temperature drying.
- Efficiency enhancement can be achieved by increasing the inlet gas temperature.
- Since the drying medium is in direct contact with the fuel to be dried, lots of contaminations and dust may entrain to the drying gas. So that, the need for filtration and exhaust gas treatment is vital.
- More energy is consumed in direct dryers while the final moisture content is lower[5].

3.3.2 Indirect drying method

In indirect drying, there is a wall between drying gas and the wet material. Therefore, heat transfer is through the wall or via a hot surface. As a result, drying gas does not play the role of moisture carrying anymore. Conduction dryer or contact dryer are other names referring to this type of drying system. Some of the operating characteristics associated with indirect dryers are presented as follows:

- The wall temperature can vary from below freezing in freeze drying to more than 800K.
- Indirect dryers are applicable when solvent recovery is of interest and/or to avoid oxidation of materials and explosion risks. Drying of materials in low pressure condition and inert environment are also suited to be done in indirect dryers.

- When drying medium can be condensed easily, indirect dryer is an energy efficient choice.
- Solid entrainment is not the case in this dryer type, so that it is more convenient to work with dusty materials in indirect dryers [5].

3.3.3 Infrared or radiant heat dryers

In this type of dryer heat transfer is via radiation. The source of radiant heat could vary. It can be from electricity by infrared lamps or from electric resistance elements and other sources. This method is not as applicable as direct and indirect methods in chemical processing industry. However, its application in drying paint films and thin layer heating of materials is worth-mentioning[5].

Turning back to the drying method classification offers another dryer group by *Perry's chemical engineering handbook* which is based on solids to be handled. However, it seems difficult to put RDF in one category to choose a suitable group of dryer for. Although, it may be quit useful for other solid drying systems.

Another sort of classification is to differentiate between dewatering and thermal drying of solid wastes. In some literatures, dewatering and liquid separation are also named as drying. That is why, drying and thermal drying are two distinct concepts in environmental literatures. Dewatering or wastes dehydration is regarded as drying and thermal drying is the process of moisture evaporation by means of hot gas as a drying medium[4]. Drying and thermal drying are identical in this report, to avoid confusion.

No particular classification for solid waste drying could be found in the literatures. However, another type of classification is suggested by *Hugo Perazzini et al.* as shown in Figure 3-3 [4]. This figure gives a common set of wastes drying techniques from which thermal drying in traditional dryers with hot air are more related to our case.

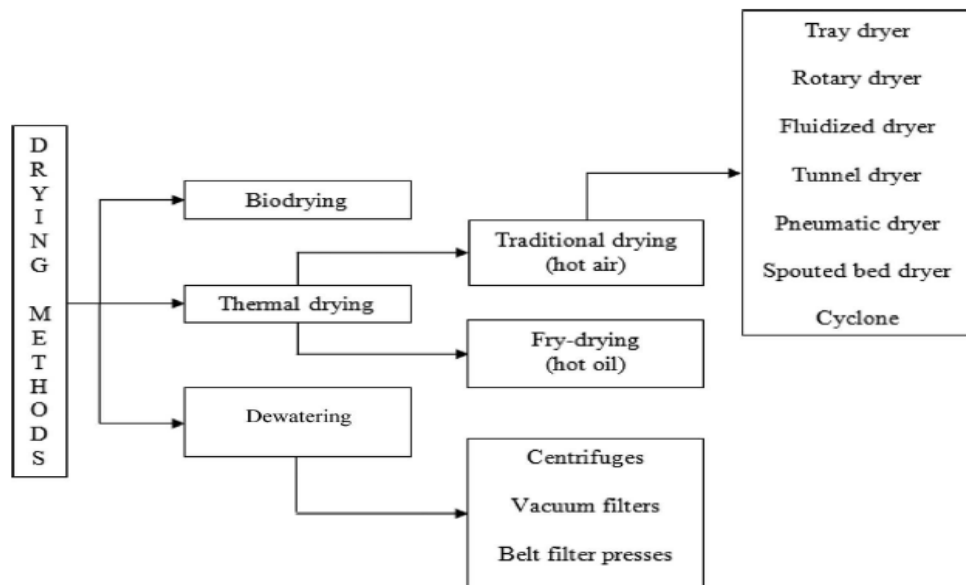


Figure 3-3 Common solid waste drying methods[4]

As shown in Figure 3-3, in immersion fry-drying boiling oil is used instead of a hot gas as the heat source. Waste and oil contact pushes some physical and chemical reactions to take place in the temperature range of 100-180°C with considerable heat and mass transfer rate. Moisture is removed from the solid by oil diffusion to the pores making the water move out. So that, moisture evaporates. This type of drying can be applied in processes like sludge drying and wood wastes drying [4].

Bio drying, in Figure 3-3, is the moisture reduction method through aeration at elevated temperature and it can be done by wastes organic degradation[4].

However, among the types of dryers suggested in Figure 3-3, and Figure 3-2, rotary dryers, fluid bed dryers and belt dryers (i.e. could be a sub-category of tunnel dryer group/through circulation group) are the most applicable groups for RDF drying techniques commercially used. This claim is based on various dryer manufacturers' reference lists given in the appendix D.

3.4 Dryer types

Among different classifications suggested in the previous sub-chapter, more emphasis is on rotary dryers', fluid bed dryers' and belt dryer's' functionalities in this part. These are the most commercially dryers for RDF treatment. Following that, steam drying will be explained in more details.

3.4.1 Belt conveyor dryer

Belt conveyors and screen-type conveyors are of the types of dryers which are used in a completely continuous process. Screen-type belt dryers could be in the group of continuous through circulation dryers[5].

Belt dryers consist of some sections with circulating fans and heating coils. They are all placed inside a housing or a tunnel in which a conveyor belt is running through. Solid bed which is 2-15cm deep, is conveyed on the belt and transported in the tunnel while being in contact with the drying medium[5].

Drying gas flow could be countercurrent, co-current or combined flow (cross flow) among which air recirculation or cross-flow design is the most common flow configuration. In that case the conveyor belt is perforated and they are mostly applicable for free flowing materials. Through circulation of the gas could improve the heat and mass transfer rates. The hot gas flows upward or downward through the permeable bed and moisture can be driven off as the hot air infiltrates to the solid bed. Figure 3-4 schematically shows a section of a belt dryer where the hot gas passes through the wet bed and reheated before being in contact with the solid material again. It is common to have upward flow of the drying gas where the wet materials enters and downward flow in the dry side. Some amount of hot gas is also exhausted through fans during the process. Drying process could be operated at constant temperature as a result of heating coils installation inside the chamber as well as some fans for recirculation. Air could enter and leave the chamber at any favorable parts[6]. Maximum pressure drop for the hot gas through the bed with no gas bypassing and leakage is almost 5mbar[5].

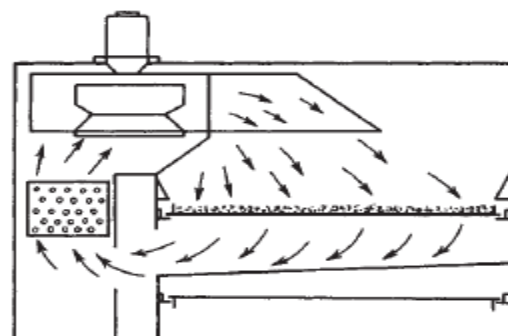


Figure 3-4 Drying gas circulation in a conveyor belt dryer[5]

Belt dryers are simple in design and operation and quite flexible for various air flow and temperature [6].

Belt dryer fabrications include sections of conveyor screen with widths of 0.3-4.4m and length of 1.6-2.5m. Each section have a metal enclosure insulated walls and roof as well as heating coils and fan, baffles for air distributing and some parts to catch fines underneath of the conveyor[5].

Different parts of a belt dryer and its functionality are well-depicted in Figure 3-5 In this figure the dryer parts together with solid and gas flows are labelled by numbers from 1 to 17.



Figure 3-5 Belt dryer configuration for RDF drying[7]

As seen in Figure 3-5, wet solid material with some moisture content enters the dryer by passing through two-part distribution screw (i.e. labeled as 1 in the figure). The solid bed on the conveyor belt can be well-spread via a distributing screw (i.e. labeled as 3). Drying air (number 13) flows through the bed to dry the wet material and carry out the evaporated moisture. The dried solid is finally discharged by a discharged screw, labeled as 4 in the figure below. If the fresh air is going to be used as the hot gas, it should be

heated up to a certain level by passing through the heat-exchangers installed in the dryer chamber. In the Figure 3-5, the layer of solid bed and a web belt could play the role of filtration in the dryer to reduce the exhaust air dust emission to the atmosphere[8].

3.4.2 Direct rotary drum dryer

A rotary dryer is a proper choice for heterogeneous solid materials for a continuous process. Moist solids with various shapes and sizes can be processed homogeneously, as a result of a proper mixing, and give a more uniform product at the end[4].

Rotary dryers have a cylindrical drum with a bearing to rotate upon. These dryers are mostly inclined to some degree to ease the movement of the moist solids forward. Moist materials enter the cylinder from one side and move ahead as a result of head impact, rotational movement and cylinder slope to exit from the other side. Drying gas flow in the cylinder could be either co-current or counter-current with the solid flow which may enhance the solid flow or slow it down, respectively. **Countercurrent flow improves the heat transfer rate** while co-current flow may be more applicable for heat sensitive materials because the hot gas temperature decreases significantly as a result of surface moisture initial evaporation[5].

Rotary dryers do not have simple operations and the process is controlled by different mechanisms such as momentum, heat and mass transfer. Many independent parameters like solid flow rate, solid type, moisture content, hot gas flow rate, hot gas temperature, drum diameter and length, angle of slope, rotation speed and residence time play key roles while working with rotary dryers [4].

Direct rotary dryer is a simple cylindrical drum with or without flights as shown in Figure 3-6. Flights application inside the drum is to lift and showering of the solids for a better distribution during the drying process. They can be used in different shapes depending on the materials to be dried. For instance, a radial 90 degree lip is more suitable for light materials or free flowing ones and flat radial flights without lip suits sticky solids. Figure 3-7 depicts the two mentioned types of flights. However, there are other arrangements, as well.

Having some elements inside the rotary drum may enhance the solid mixing; although it makes cleaning and maintenance more difficult.

Depending on the metal strength applied in the fabrication, rotary dryers can be used for low and medium operating temperatures[5].

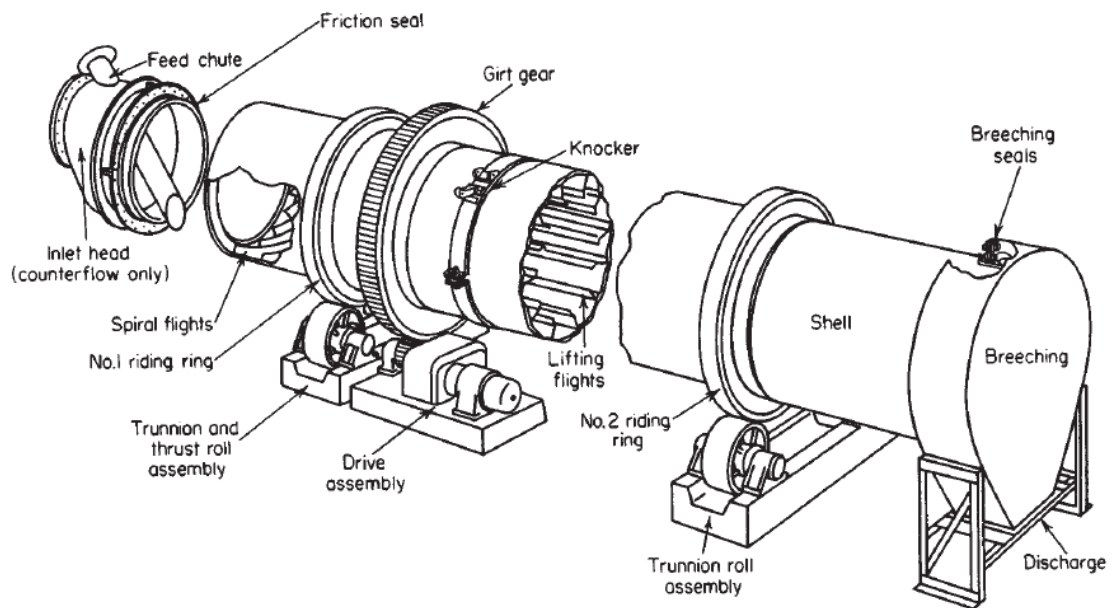


Figure 3-6 Different components of a direct rotary dryer[5]

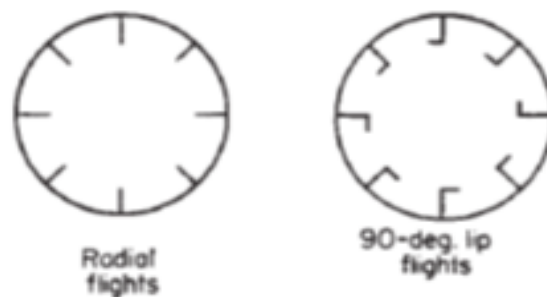


Figure 3-7 Radial flight without lip and with 90 degree lip arrangement[5]

Direct rotary dryer is simpler and more cost-effective to manufacture than the indirect counterpart. However, it is only applicable when there is no restriction of direct contact between solid and drying gas. In direct rotary dryers, a large amount of drying gas volume and gas velocities are needed. The drying gas velocity is mostly higher than 0.5m/s for an effective design. So that, when the solid material has the tendency to produce dust, solid particle entrainment to the gas stream is inevitable. As a result, a filter or a cyclone may be needed to catch the dust from the exhaust gas[5].

How the solid materials are fed in the rotary dryer depends on solid characteristics and location as well as upstream equipment. However, in order to seal the feeding system and dryer properly and if the gravity feed is not practical (gravity feed needs a chute through which solid is fed to the dryer) a screw conveyor is applied[5].

The carrier gas flow is usually countercurrent with respect to the solid flow and it drives off the produced vapor at the feed end of the dryer[5].

Based on the study done by *Pinacho et al.* drying of heterogeneous solids is more efficient using a continuous rotary dryer than the tray ones[4]. So that it can be expected to have an effective RDF drying as it is heterogeneous.

According to the analysis done by *Zabanitou*, different design parameters like angle of slope, rotation speed and hot gas condition could have profound impacts on forestry woody materials drying. For instance, it was observed that air temperature reduction led to increased constant drying rate period and reduced falling drying rate period. Other outcome of their analysis was the effect of residence time in a way that short residence time may result in less drying efficiency[4].

Figure 3-8, shows a flowsheet of a drying system using direct rotary system. As seen, having a cyclone and/or a bag filter is inevitable.

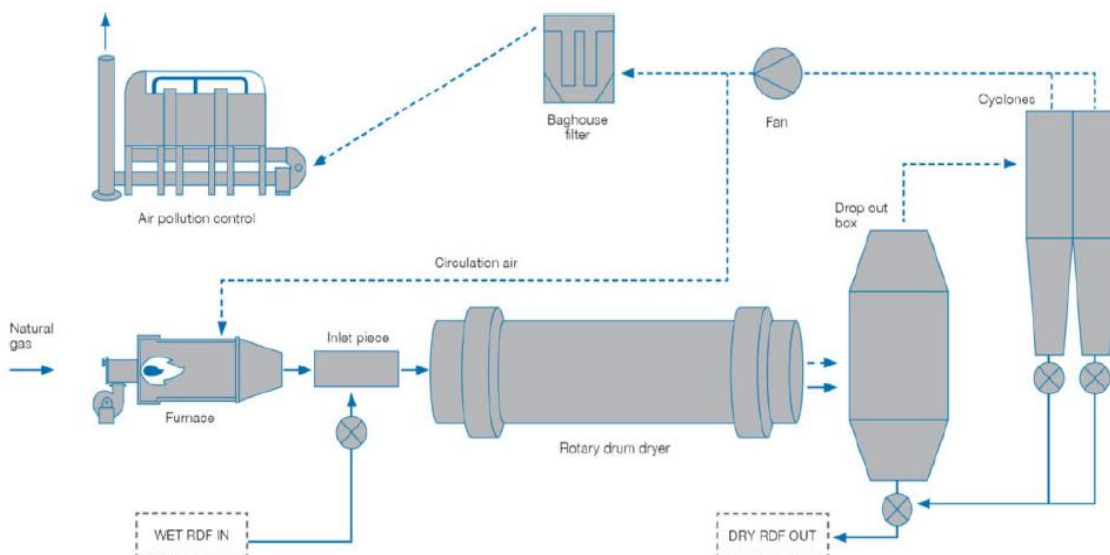


Figure 3-8 Rotary drum dryer system overview[image courtesy of ANDRITZ Separation

]

3.4.3 Indirect rotary dryer

Steam tube rotary dryer is one of the most applicable indirect rotary dryers as shown in Figure 3-9. In this type of dryer, steam tubes are installed radially along the drum length in one or several rows. Drying occurs when solids are in contact with the heated tubes through conduction and radiation heat transfer. Drying is controlled by steam temperature or pressure in the tubes and also by the solids residence time. In case of dealing with sticky solids, one tube row is suitable. Steam tubes can be normal pipes having condensate drainage or they can be bayonet-types. Flight installation inside the cylindrical drum assists material mixing and enhances turbulence in the system[5].

In this dryer type, solid materials are fed into the dryer via a screw conveyer or a chute and leave the dryer through openings on the shell at the other end. These openings are also used for a carrier gas intake to remove the moisture vapor. Vacuum pressure in the dryer could also drive the moisture off the system [5].

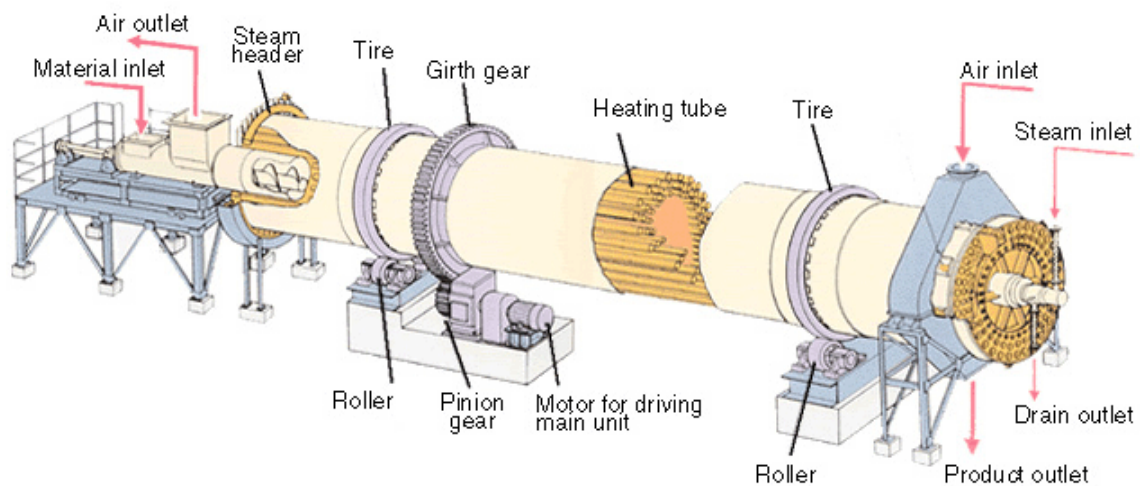


Figure 3-9 Steam-tube rotary dryer components[9]

3.4.4 Fluid-bed dryer

Fluid-bed dryers or fluidized-bed dryers are quite well-known in the industry because of their vast applications and broad operating conditions. A system of a fluidized bed consists of cylindrical body as the fluidized bed column filled with a bed of solid particles which provides a proper contact with the gas phase coming into the cylinder[4]. In case of using air as the drying gas, its temperature may be in the range of 100°C to 450°C

which depends on the solid to be dried; however, drying at inlet low temperature can reduce fire risk in the dryer[10]. This type of dryer also includes a gas blower to assist the gas flow, a heater and a gas cleaning system to separate some fine particles that entrained to gas phase[4, 6]. Figure 3-10 is a simple set up of a fluidized bed drying system. In this type of dryer, a bed of moist solids requires a greater minimum fluidization velocity than the bed of same dried materials. So that, only the upper levels of the bed can become fluidized and the lower levels may not move during the first stages of drying [6].

This type of dryer offers a great solid mixing, considerable heat and mass transfer, uniformity of the dried solids and easy solid transport [4].

The behavior of the solid bed in the fluidized bed column is partly shown in the upper right side of the Figure 3-10. As seen, two zones are generated, named as dense phase zone and freeboard zone. The former is at the bottom and the latter is located on top. In freeboard zone, solid hold-up and density are inversely proportional to the bed height. Fine solids that have terminal velocities smaller than the gas velocity can be carried by fluidizing gas flow in the freeboard region.

Solid hold-up is reduced with the increase of freeboard height to a level above which solid hold-up does not change. This level is called Transport disengagement height (TDH). The gas discharge point in the column should be above the TDH level to avoid solid entrainment to the gas stream[6].

Despite of the advantages coming with this dryer type, some restrictions such as high pressure drop, increased power consumption, inability to proper fluidization behavior of some solids, non-uniformity of solids in some dryer types, pipe and vessel erosion, particle entrainment and etc. also exist[6].

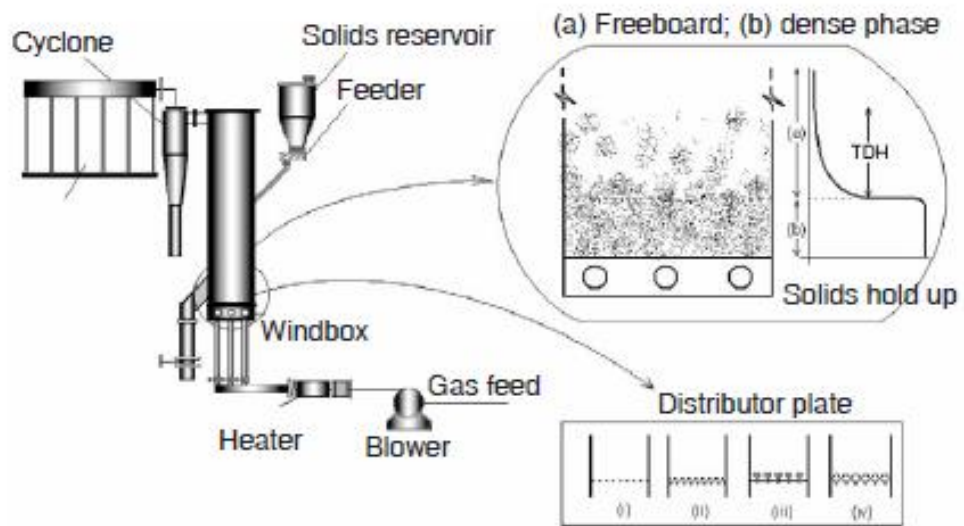


Figure 3-10 Typical fluidized bed drying set-up[6]

According to the research done by *Moreno et al.* elevated temperature as high as 187°C resulted in some advantageous in forestry wastes drying by a fluidized bed. This conclusion was based on some experiments, showed that higher gas temperature led to a more energy-efficient drying system as a result of higher rate of drying[4].

3.5 Steam (super-heated steam) drying

Steam drying, as an alternative to air drying methods is divided into direct steam drying and indirect steam drying. The indirect type is similar to what is already explained in previous sub-chapter as indirect drying method. It means using steam as the heating medium to provide the energy needed for solid drying. The type of dryer for this purpose can be steam tube rotary dryer. Taking advantage of steam latent heat energy which is quite high, would reduce the size of dryer considerably compared to hot air dryers.

On the other hand, direct steam drying or super-heated steam drying (SSD), is one of the technologies attracted attentions over the past decades. Albeit, the concept goes back to more than 100 years ago and the first application was found in Germany around 70 years ago. It is the method of superheated steam utilization in convective dryers. Superheated steam could be principally used in any direct or the combination of direct and indirect drying methods. In this approach steam act as the drying medium and the carrier to remove evaporated moisture as the excess steam, just like hot air in conventional dryer types. SSD technique is more complicated than the air system, so that replacement of steam in dryers needs special consideration and criteria[6].

3.5.1 Principle of super-heated steam drying (SSD)

In SSD, the super-heated steam gives its sensible heat to the wet material. During the drying process, the steam should be above its condensing temperature. However, the condensation only occurs in initial phase of drying. Most of the steam would be circulated in the system after being heated and the excess steam could be utilized as the heating medium for other applications. Figure 3-11 illustrates a scheme of the SSD system.

As seen in the figure, slightly superheated steam as the dryer exhaust passes through compressor after removal of excess steam. High pressure steam enters a super-heater before being fed to the dryer. Purged steam could be possibly sent to the heat recovery unit where steam is either directly used or heat recovered elsewhere in the plant. Afterwards, condensate having pollutants could be taken out from the system.

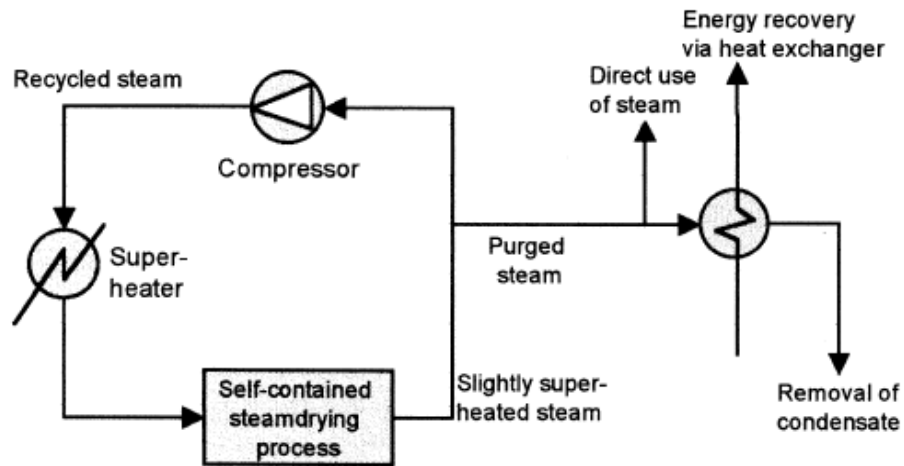


Figure 3-11 Steam flowsheet in an SSD system[11]

Three phases exist to describe the drying kinetics of solid materials, in SSD as following:

a) Condensation-evaporation period

In this phase, steam is condensed initially to heat-up the solid material by the energy of its latent heat. After that, the moisture layer on the sample evaporates at constant temperature of 100°C. In this stage, heat transfer coefficient between the solid and drying gas is dominating.

b) Constant drying rate period

Moisture evaporation continues at the surface of the solid bed and the drying rate is constant at this stage. The solid temperature would be constant at 100°C. Heat transfer rate controls the process.

c) Falling drying rate phase

Solid temperature exceeds 100 °C in this stage and the moisture content drops exponentially. Moisture diffusion from solid inside to the surface controls the drying process[12].

Superheated steam has higher heat transfer properties than air at the same temperature. In SSD the rate of drying in constant rate period depends only on heat transfer rate, not on mass transfer, since no resistance exists for the moisture to diffuse into the steam.

Equation 3-1 gives a simplified relationship for the rate of surface moisture evaporation into steam.

$$W_D = \frac{q}{\Delta H} = \frac{h(T_{ss} - T_s)}{\lambda} \quad (3-1)$$

Different terms in equation 3-1 are as following:

W_D : Rate of evaporation ($\frac{\text{kg water}}{\text{m}^2\text{s}}$)

T_s : Surface temperature corresponds to the saturation temperature at operating pressure

T_{ss} : Superheated steam temperature

h : Convective heat transfer coefficient ($\frac{\text{W}}{\text{m}^2\text{K}}$)

λ : Latent heat of vaporization ($\frac{\text{kJ}}{\text{kg}}$)

The effect of sensible heat, heat losses and other heat transfer methods are neglected in equation 3-1[13].

The surface temperature in hot air drying corresponds to wet bulb temperature. As a result, temperature difference (ΔT) is higher in air drying system. However, the heat transfer coefficient (h) is less than steam. These two effects introduce a new term called inversion. Inversion is the temperature above which drying with SSD is faster than drying with air[13]. In another explanation by *Yoshida and Hyodo*, inversion temperature is a point at which the evaporation rate does not depend on drying gas humidity[12]. Figure 3-12 is more illustrative of this phenomenon.

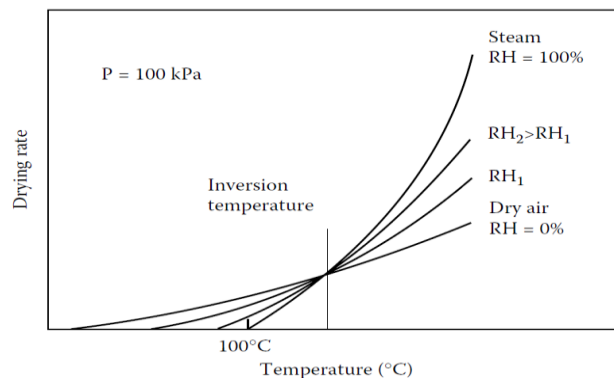


Figure 3-12 Variation of drying rate with temperature using steam/air [13]

Inversion temperature in the order of 160-200 °C for the water evaporation in superheated steam for different flow regimes, flow configurations and geometries is observed. Above the inversion temperature, superheated steam has higher rate of drying for direct drying methods when surface moisture is evaporated. So, the inversion temperature is only related to the surface moisture evaporation and not to the internal moisture[13].

Sometimes, the rate of drying in the falling rate period is more in SSD compared to air dryers, which is because of higher solid temperature in SSD. This leads to higher water mobility in the solid material. In addition, case hardening does not exist in SSD; as a consequent, there should be fewer obstacles for more severely drying and solid is probably more porous afterwards[13].

3.5.2 SSD advantages and dis-advantages

Pros and cons always exist with any methods and these are process-related. Some advantages may not work for some systems or may even turn to disadvantages. Consequently, major consideration needs to be done to select a proper system with paying attention to positive and negatives aspects of the selected method.

The biggest advantage of SSD technology is having steam as the exhaust dryer gas which could be heat recovered more easily than the off-gas in air dryers. In fact, the whole latent heat of exhaust gas (i.e. excess steam) in SSD is recoverable if the system is sealed enough to prevent any air penetration into or decrease it to the regulated level. In that case, condensing the excess steam or mechanical vapor recompression could provide a pretty reasonable heat recovery[13].

SSD is more interesting and energy-efficient if this excess steam is applied within the drying system or elsewhere for other purposes [13]. For instance, the excess steam energy could be utilized for deinking of recycled paper in the paper industry or for steamer fixation in textile industry. Some studies has shown a potential energy saving up to 75% when excess steam is reused in paper and textile industry [11]. As such, according to *Kudra & Mujumdar*, a figure for the net energy consumption of SSD is 1000-1500kJ/kg

of evaporated moisture compared to 4000-6000kJ/kg of evaporated water in the corresponding hot air dryer[13].

Some emissions from drying process in SSD could be rejected when steam is condensed which is an advantage; however, some pollutants may remain in the gas [11].

Other advantages of SSD technology over air drying techniques are as following:

- Since there is no free oxygen in the system, except those which infiltrates to the system, oxidation and combustion reaction is prevented in SSD. So that, no explosion or fire risks occur.
- The quality of products can be improved in some cases when SSD is applied. This is more applicable for food drying, for instance.
- There is a higher rate of drying in both constant and falling rate periods when the steam temperature is above the inversion temperature in SSD. While, below that, drying with air is more quickly. This is due to higher thermal conductivity and heat capacity of the superheated steam that results in increase of drying rate for the surface moisture. In addition, higher product temperature and no diffusion resistance for evaporated water into steam enhances the drying rate.
- Case-hardening phenomenon and formation of water-impermeable shell, which occurs in fast air drying does not exist in drying with superheated steam.
- When the solid wastes have toxic compounds, they can be recovered via exhaust gas condensation[13].
- SSD system could be pressurized if high pressure steam is going to be reused in other parts of process or when high temperature products are of interest[11].

On the contrary, there are some restrictions for using SSD as following:

- As mentioned earlier, there should be no air penetration to the SSD system if the heat recovery by condensation or compression is favorable. This is due to the challenges with existence of non-condensable parts. As a result, the input and the dryer discharge should be sealed enough. However, non-condensable compounds can also enter together with the products to be dried.

- The process in SSD technique is more difficult to start-up and shutdown compared to conventional air dryers.
- Having the feed at ambient temperature results in initial condensation of steam in the dryer. So that, about 10-15% should be added to the dryer residence time. This can be avoided by a preheater installation for wet solids.
- Since the temperature is high in SSD method, materials which could melt or thermally damaged should be avoided to enter such dryers. However, lowering the pressure can be an alternative for such compounds and it increases the drying rate.
- Products which may need oxygen for quality improvement should not be dried in superheated steam dryers like in food industry.
- The cost of auxiliary systems accompanying SSD such as feeding part, product collection system, and steam recovery is often more expensive than the dryer itself. Consequently, superheated steam drying may be more feasible and cost-effective for continuous operation and high capacities.
- Another restriction of using SSD is the lack of experience; however, it has been increasing more and more over the past years[13].
- Another limitation of SSD is the necessity for having a proper insulation of dryer because all parts of dryer interior should be above steam condensation temperature to avoid droplet formation[11].

There are many publications about superheated steam drying method for food drying as well as paper and wood industry. However, no commercial application for solid waste drying (not bio wastes) was found by the writer. Having no experimental set-up was another limitation to more accurately evaluate steam drying of RDF over conventional dryers with hot air. Next chapter would give some plant data as a case study where RDF should be dried to be applied as a fuel in a cement plant.

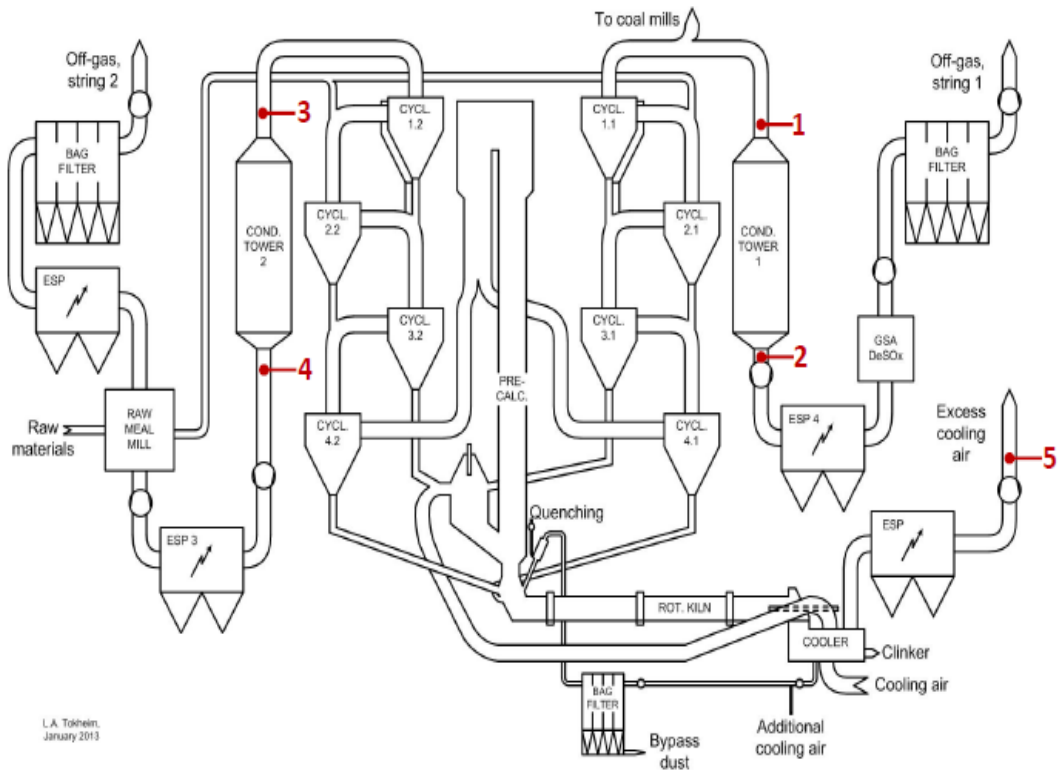
4. Norcem cement plant as a case study

As stated earlier, RDF utilization, especially in cement industry is increasing. However, drying of RDF is quite vital to increase its calorific value before being used as a fuel.

Norcem cement kiln plant located in Brevik, southern Norway, is taken as a case study plant for this project. Coal is the main fuel being used in Norcem as the only cement producer in Norway; however, municipal wastes like RDF is also applied as the alternative fuels. In terms of using secondary fuels, Norcem Brevik is of pioneering plants in Europe[14].

A process flow diagram (PFD) for a part of process in the plant from feed entry to clinker production (Ref. appendix B), is presented in Figure 4-1. This figure is a basis for potential waste heat utilization to be later used for necessary calculations.

As seen in Figure 4-1, raw materials enter a Raw Meal Mill where they can be preheated and then pass through a set of cyclones in parallel to take some more heat from the hot exhaust stream coming from the kiln. These raw materials enter a rotary kiln and heated-up to react and produce an intermediate material named clinker which is then used to produce cement. Coal and alternative fuels are combusted in the kiln to supply energy for combustion. Exhaust gas leaving the rotary kiln divided into 3 strings and they all have to pass an Electro-Static Precipitator (ESP) before venting to the atmosphere. Two of three strings can heat exchange with the raw materials by passing through the cyclones as two parallel flows. However, these two streams need to be conditioned in condition towers 1 & 2 before entering ESP 3 & ESP 4. The third string of the exhaust gas is directly vented to the ambient after being filtered at point 5 in the figure. The gas characteristics of these three strings such as flowrate, pressure, temperature and composition at different points in the process, are also presented in Table 4-1 (Ref. appendix B).



L.A. Tokheim,
January 2013

Figure 4-1 A part of cement production PFD in Norcem, Brevik

Table 4-1 Typical gas characteristics of red points shown in Figure 4-1

Points	Gas flow rate [Nm ³ /h]	Temperature [°C]	Pressure [mbarg]	N ₂ [Vol%]	CO ₂ [Vol%]	H ₂ O [Vol%]	O ₂ [Vol%]	Dust [mg/Nm ³]
1	132250	386	-80	61	23	9	7	50000
2		120						
3	132250	386	-80	61	23	9	7	50000
4		224						
5	180000	220 ^a	5	79	0	0	21	5

^a Reduction to 100°C would be OK (relevant if a waste heat recovery unit is installed)

In order to dry RDF before combustion in the kiln, it is always preferred to utilize excess heat in the plant to make this process more efficient instead of introducing new sources of energy. The amount of plant excess heat is presented in the following part. Moreover, a mass and energy balance around a potential dryer are given later in this chapter.

4.1 Available Excess Heat in the Plant

Based on Figure 4-1 and plant data in Table 4-1, there is some heat available in the plant which can be implemented for drying purposes.

The first available heat source is the energy of the exhaust gas taken in the condition tower 1 in Figure 4-1. In this unit, the temperature of the gas is reduced from 386°C to 120°C. This temperature difference for a gas flowrate of 132250 Nm³/h could have a significant amount of energy which can be used elsewhere in the system.

Another excess energy source could be from condition tower 2 in which the gas temperature drops from 386°C at point 3 to 224°C at point 4. The reason of having different outlet temperature in condition towers 1& 2 is that, the gas leaving tower 2 is going to preheat the raw materials in raw meal mill.

The third available energy is the excess cooling air at point 5 in Figure 4-1. Air flowrate of 180000 Nm³/h with a temperature of 220°C offers a considerable amount of energy when the temperature could drop to 100°C for the waste heat recovery unit.

The energy amount of the three possible heat sources could be calculated using equations (4-1) or (4-2) and results are summarized in Table 4-2. However, more detailed calculation steps and data could be found in appendix C.

$$\dot{Q} = \dot{n}\bar{c}_p\Delta T \quad (4-1)$$

$$\dot{Q} = \dot{m}\bar{c}_p\Delta T \quad (4-2)$$

Different notations in equations (4-1) & (4-2) are described as following:

\dot{Q} : Energy of the gas per time (Power) [MW]

\dot{n} : Gas molar flowrate [$\frac{\text{mol}}{\text{s}}$]

\dot{m} : Gas mass flowrate [$\frac{\text{kg}}{\text{s}}$]

\bar{c}_p : Average specific heat capacity of the gas, molar based [$\frac{\text{J}}{\text{mol.K}}$]

\bar{c}_p : Average specific heat capacity of the gas, mass based [$\frac{\text{J}}{\text{kg.K}}$]

ΔT : Temperature difference [K]

Table 4-2 Amounts of three available energy sources in Norcem plant

Available power from point 1 to 2		Available power from point 3 to 4		Available power at point 5	
$\dot{n}_1 [\frac{mol}{h}]$	$\sim 5.9 \times 10^6$	$\dot{n}_3 [\frac{mol}{h}]$	$\sim 5.9 \times 10^6$	$\dot{n}_5 [\frac{mol}{h}]$	$\sim 8 \times 10^6$
$\bar{c}_{p1-2} [\frac{J}{mol \cdot K}]$	33.943	$\bar{c}_{p3-4} [\frac{J}{mol \cdot K}]$	34.51	$\bar{c}_{p5} [\frac{kJ}{kg \cdot K}]$	1.017
$\Delta T_{1-2} [K]$	266	$\Delta T_{3-4} [K]$	162	$\Delta T_5 [K]$	120
$\dot{Q}_{1-2} [MW]$	14.8	$\dot{Q}_{3-4} [MW]$	9.2	$\dot{Q}_5 [MW]$	7.9

Depending on the amount of energy needed for drying process, each of the above energy streams can be chosen. The next step is to calculate how much energy is needed to dry RDF to a certain level. However, it is important to know about RDF characteristics being applied in Norcem and use them for further calculations.

4.2 RDF characteristics used in Norcem

In order to find out the level of drying, total thermal energy for this purpose and to do related further calculations, RDF characteristics data and detail analysis are necessary. Table 4-3 gives RDF characteristics being utilized in Norcem plant which consists of moisture content, volatiles, char, ash and calorific value. It should be noted that all values in Table 4-3 are wet basis (Ref. appendix B).

Table 4-3 RDF characteristics used in Norcem plant

Parameter	Unit	Value
Moisture content	wt %	29
Volatiles	wt%	49
Char	wt%	5
Ash	wt%	17
Lower heating value (LHV)	MJ/kg	12

Detailed RDF component analysis is also given in Table 4-4 (Ref. appendix B). These data may be useful to calculate RDF combustion temperature. Compositions of different components is also wet basis.

Table 4-4 Detailed RDF component analysis

Component	Moisture	C	H	O	N	S	Ash	sum
Unit	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%
Composition	29.0	28.8	4.1	20.4	0.5	0.2	17.0	100.0

Table 4-5, presents some values for the ash components in an RDF sample. All weight fractions are dry basis (Ref. appendix B). This data is applicable when the effect of ash on absorbing some heat of combustion is considered. Details about this effect is more discussed in the following sub-chapters.

Table 4-5 Ash component analysis of an RDF sample (dry basis)

Ash component	CaO	Al ₂ O ₃	Fe ₂ O ₃	K ₂ O	Na ₂ O	Sum
Composition wt%	25	33	12	13	17	100

Annual consumption of RDF and the kiln run time in a year are also given in table 4-6 (Ref. appendix B).

Table 4-6 Typical operational values of the plant

RDF consumption	t/y	100000
Typical kiln run time	h/y	7000

4.3 Final RDF moisture content

In order to find out the amount of moisture to be evaporated and the amount of required energy for this purpose, a mass and energy balance should be performed which can be utilized later to size a dryer. However, before doing that, it is important to know about the final moisture content of RDF after drying. Reaching to about zero moisture level may seem a doable job in theory but it may not be an optimum level in practice. Performing a drying process beyond the optimum level could result in having a huge dryer and an infinite residence time for the solid in the dryer. The only way to understand this is to

perform some tests on RDF samples from Norcem or to take advantage of some dryer manufacturers' advice. Set up an experimental test to reach the optimum level of RDF drying was not possible for this project. This resulted in consulting with some vendors to take their advice regarding optimum RDF final moisture content. They all recommended that around 10wt% moisture content in the dryer outlet is optimal, as it seems a common limit in commercial RDF drying (Ref. appendix D).

Since, the ultimate goal is to increase the heating value of RDF, it can be useful to see how RDF calorific value differs with various level of moisture content. This can be done through a series of calculation, as presented in appendix C. In this calculation, initial moisture content of RDF and its heating value are taken from Table 4-3 which shows the typical RDF characteristics in Norcem Brevik. Variation of RDF lower heating value (LHV) with moisture content is well depicted in Figure 4-2. This graph was achieved as a result of calculation done in appendix C. As seen, LHV of RDF increases with decreasing of RDF moisture content, as expected.

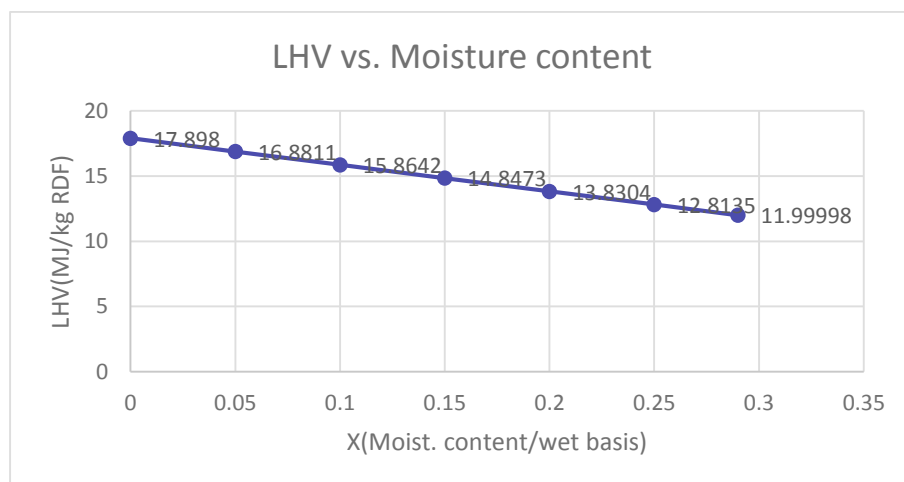


Figure 4-2 RDF heating value variation vs. moisture content

If we assume that suggested value by vendors which is 10wt% moisture content is the final level of drying, Figure 4-2 indicates that we can improve the heating value of RDF by almost 32% which sounds a pretty reasonable estimate.

$$\text{LHV increase} = \frac{15.86 - 12}{12} \times 100 = 32.2\%$$

Another quantitative evaluation to see if 10wt% is a proper estimate, can be done through adiabatic flame temperature comparison. In order to calculate flame temperature for RDF combustion in different moisture content, detailed component analysis of RDF is applied, taken from Table 4-4. Table 4-5 which gives a more detailed ash compositions is also used to calculate RDF flame temperature. The first step to find out flame temperature of RDF is knowing about the amount of air needed for combustion. Detailed calculation steps and taken assumptions are presented in appendix C.

The output of this calculation is summed up in Figure 4-3.

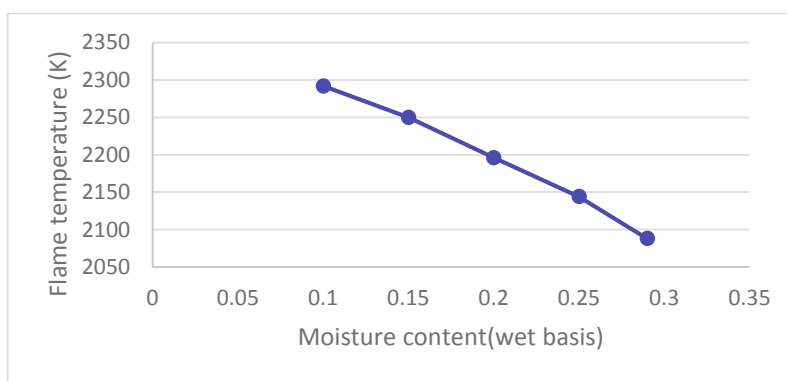


Figure 4-3 Flame temperature in RDF combustion vs. moisture content

As seen in Figure 4-3, lowering the RDF moisture content results in combustion temperature rise. By assuming that the initial RDF temperature is 283K, and no excess air is used, its flame temperature could increase from around 2088K to around 2292K, if RDF with 10% moisture content is combusted. Meaning that, we could have an increase of almost 200K in flame temperature by decreasing moisture content from 29wt% to 10wt% which seems a considerable rise.

However, it should be noted that the values and graphs, presented in this part, are just to show the trend and considered as an initial simplified estimation for flame temperature, because they are based on some assumptions that are well mentioned in appendix C. Errors because of neglecting excess air in the combustion, taking an average specific heat capacity for the product, different reactant and ambient temperatures may cause some changes in the mentioned values and figure, in the real process.

4.4 Mass balance around a potential dryer

Based on preliminary estimation done for RDF calorific value and flame temperature improvement in the previous sub-chapter, vendors' suggestion for having 10wt% final moisture content for RDF sounds a quite good guess which is further used in mass balance calculation.

Figure 4-4, schematically shows a simplified flow diagram for a dryer. Equations 4-3 and 4-4 are also used as total and component mass balance equations. Explanation of the Symbols and notations used in Figure 4-4 and equations 4-3 and 4-4, are as following:

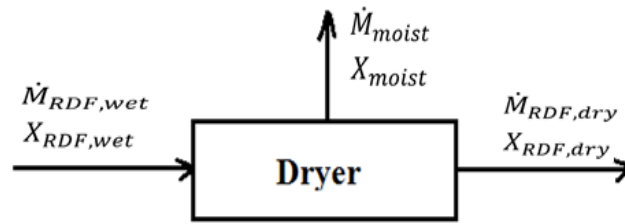


Figure 4-4 Flow streams around dryer

$\dot{M}_{RDF,wet}$: Mass flowrate of wet RDF coming into the dryer $[\frac{kg}{h}]$

$\dot{M}_{RDF,dry}$: Mass flowrate of dried RDF leaving the dryer $[\frac{kg}{h}]$

$X_{RDF,wet}$: Moisture content of wet RDF, wet basis $[\frac{kg \text{ water}}{kg \text{ wet RDF}}]$

$X_{RDF,dry}$: Moisture content of dried RDF, wet basis $[\frac{kg \text{ water}}{kg \text{ wet RDF}}]$

\dot{M}_{moist} : Mass flowrate of evaporated moisture $[\frac{kg}{h}]$

X_{moist} : Moisture content of evaporated moisture $[\frac{kg \text{ water}}{kg \text{ moist}}]$

$$\dot{M}_{RDF,wet} = \dot{M}_{RDF,dry} + \dot{M}_{moist} \quad (4-3)$$

$$X_{RDF,wet} \cdot \dot{M}_{RDF,wet} = X_{RDF,dry} \cdot \dot{M}_{RDF,dry} + X_{moist} \cdot \dot{M}_{moist} \quad (4-4)$$

As concluded in the last sub-chapter, $X_{RDF,dry} = 0.1 \frac{kg \text{ water}}{kg \text{ wet RDF}}$

By substitution of right values in above formulas, \dot{M}_{moist} which is the moisture to be evaporated in the dryer, can be achieved as below:

Wet RDF amount can be calculated via equation 4-5 in which RDF consumption and kiln run time are taken from Table 4-6.

$$\dot{M}_{\text{RDF,wet}} = \frac{\text{RDF consumption}}{\text{Kiln run time}} \quad (4-5)$$

$$= \frac{10^5 \left(\frac{\text{t}}{\text{y}}\right)}{7000 \left(\frac{\text{h}}{\text{y}}\right)} \approx 14.286 \frac{\text{t}}{\text{h}} \approx 14286 \frac{\text{kg}}{\text{h}} \approx 15000 \frac{\text{kg}}{\text{h}}$$

The amount of wet RDF obtained from above formula, is rounded up to 15t/h to compensate the variation of RDF input to the dryer in different time of the year.

From equation 4-3:

$$15000 = \dot{M}_{\text{RDF,dry}} + \dot{M}_{\text{moist}}$$

From equation 4-4:

$$0.29 \times 15000 = 0.1 \times \dot{M}_{\text{RDF,dry}} + 1 \times \dot{M}_{\text{moist}}$$

As a result:

$$\dot{M}_{\text{moist}} \approx 3167 \frac{\text{kg}}{\text{h}}$$

$$\dot{M}_{\text{RDF,dry}} \approx 11833 \frac{\text{kg}}{\text{h}}$$

So that, the dryer should be able to evaporate about 3.2t/h of moisture from RDF. How much energy is required for this amount of evaporation will be calculated in the next sub-chapter.

4.5 Energy balance for the potential dryer

From the mass balance done in the last sub-chapter, the amount of moisture to be released is 3167 kg/h. This amount of water could evaporate using either direct or indirect drying method. Energy needed for the drying process can be supplied by hot air or steam as drying medium. However, for any of these cases, performing a general energy balance for the dryer is vital to obtain the minimum amount of energy required.

In order to proceed energy balance, some assumptions are taken into account which will be stated meanwhile the calculation.

Energy balance calculation could initiate from equation 4-6, in which the heat duty to the dryer is to provide energy to a) heat-up the feed to vaporization temperature, b) water phase change (vaporization), c) solid temperature rise to its final temperature and d) produced vapor temperature rise to its final temperature[15].

Different symbols and terms as well as taken values in equation 4-6 are listed in Table 4-7.

$$\dot{Q}_{\min,drying} = Q_{RDF\ temp.rise} + Q_{Moist.temp.rise} + Q_{evaporation} + Q_{vap.temp.rise}$$

$$\dot{Q}_{\min,drying} = \dot{M}_s C_{ps}(T_{sf} - T_{si}) + \dot{M}_{RDF} X_i C_{pL}(T_v - T_{si}) + \dot{M}_{moist} \cdot \lambda + \dot{M}_{RDF} X_f C_{pL}(T_{sf} - T_v) + \dot{M}_{moist} C_{pv}(T_{vf} - T_v) \quad (4-6)$$

Table 4-7 Necessary data for energy balance calculation

Parameter	Statement	Value
X_i	Inlet moisture content (wet basis)	0.29(kg water/kg wet RDF)
X_f	Final moisture content of RDF (assumption)	0.1 (kg water/kg wet RDF)
\dot{M}_{RDF}	Mass flowrate of wet RDF entering the dryer	15000 kg/h
\dot{M}_s	Mass flowrate of dried RDF leaving the dryer	11833 kg/h
\dot{M}_{moist}	Mass flowrate of evaporated moisture	3167 kg/h
T_{si}	RDF initial temperature (assumption)	10°C
T_{sf}	RDF final temperature after drying (Note 1)	70°C
T_{vf}	Produced vapor temperature leaving the dryer (assumption)	110°C
P_{dryer}	Dryer pressure (assumption)	1 atm
T_v	Water saturated temperature at dryer pressure	100°C
C_{ps}	RDF specific heat capacity (assumed to be constant)[16]	1.4 (kJ/kg.K)
C_{pL}	Water specific heat capacity (assumed to be constant)	4.183(kJ/kg.K)
C_{pv}	Steam specific heat capacity at average temperature of 105°C, (Note 2)	1.89(kJ/kg.K)
λ	Water latent heat at dryer pressure	2256.1(kJ/kg)
<i>Note1: RDF final temperature is assumed to be 70°C based on a quotation received for a drum dryer(refer to appendix D)</i>		
<i>Note2: steam specific heat capacity was taken from ASPEN HYSYS V8.6 data base at the linear average temperature between 110°C &100°C</i>		

It should be noted that there are some assumptions taken for some of the values in Table 4-7. For instance, RDF specific heat capacity is taken 1.4 (kJ/kg.K) and is assumed

to be constant, RDF outlet temperature is assumed to be 70°C and vapor temperature is also assumed to be 110 °C. It means that the vapor leaves the dryer at slightly superheated condition to make sure that no condensation occurs in the ending part of dryer. These guessed values could be improved by consulting different RDF dryer manufacturers or by experiments; however, they would not make a major error in calculations since the dominant energy for drying is spent on moisture phase change. Substitution of all given data in Table 4-7 in equation 4-6 gives the following results:

$$\begin{aligned}
 Q_{\text{RDF temp.rise}} &= \dot{M}_s C_{ps} (T_{sf} - T_{si}) \\
 &= 11833 \left(\frac{\text{kg}}{\text{h}}\right) \times 1.4 \left(\frac{\text{kJ}}{\text{kg.K}}\right) \times (70 - 10)(\text{K}) \times \frac{1\text{h}}{3600\text{s}} \\
 &\approx 276.1\text{kW}
 \end{aligned}$$

$$\begin{aligned}
 Q_{\text{Moist.temp.rise}} &= \dot{M}_{\text{RDF}} X_i C_{pL} (T_v - T_{si}) + \dot{M}_{\text{RDF}} X_f C_{pL} (T_{sf} - T_v) \\
 &\approx 455 - 52.3 = 402.7\text{kW}
 \end{aligned}$$

$$\begin{aligned}
 Q_{\text{evaporation}} &= \dot{M}_{\text{moist}} \cdot \lambda \\
 &= 3167 \left(\frac{\text{kg}}{\text{h}}\right) \times 2256.1 \left(\frac{\text{kJ}}{\text{kg}}\right) \times \frac{1\text{h}}{3600\text{s}} \\
 &\approx 1984.5 \text{ kW}
 \end{aligned}$$

$$\begin{aligned}
 Q_{\text{vap.temp.rise}} &= \dot{M}_{\text{moist}} C_{pv} (T_{vf} - T_v) \\
 &= 3167 \left(\frac{\text{kg}}{\text{h}}\right) \times 1.89 \left(\frac{\text{kJ}}{\text{kg.K}}\right) \times (110 - 100)(\text{K}) \times \frac{1\text{h}}{3600\text{s}} \\
 &\approx 16.6\text{kW}
 \end{aligned}$$

As a result:

$$Q_{\text{min,drying}} = 276.1 + 402.7 + 1984.5 + 16.6 = 2.73\text{MW}$$

The minimum required energy for drying purposes can be fully covered by available excess heat in the plant as calculated in sub-chapter 4-1. This amount of energy is far lower than three energy sources previously discussed.

The excess cooling air stream shown as stream 5 in Figure 4-1, can be chosen to provide energy needed for drying. Reasons of this selection are that, the amount of excess energy in stream 5 (7.9MW) is closer in value to the minimum energy needed for drying (2.73MW) and air stream 5 is almost dust free and no additional filtration is needed.

5. Possible RDF drying methods in Norcem plant

There are three different theoretical possibilities to use excess energy in the plant to provide RDF drying energy as following:

- 1) Direct steam drying: producing super-heated steam from available hot air in the plant and its utilization as drying medium in a dryer with direct mode
- 2) Indirect steam drying: producing steam using available hot air energy in the plant and applying that indirectly in the dryer
- 3) Hot air drying: taking available hot air in the plant as drying medium and applying that directly into the dryer

Based upon the survey conducted by the writer, item 3 is the most recommended process for this purpose. The process flow diagrams and further calculations for all three alternatives are prepared and elaborated in this chapter. The reasons for process selection is also provided afterwards.

5.1 Direct steam drying of RDF in Norcem plant

As discussed in chapter 3, Super-Heated Steam (SHS) should be applied as drying medium if direct mode of heat transfer is used in a dryer. Before suggesting a process flow diagram, it is necessary to calculate the amount of SHS needed. It is assumed that SHS condition is 200°C and 1.25bars at the dryer inlet and it leaves the dryer at 1.2bar. Not having a basis for inlet conditions, the assumption applied in an experimental set-up done for carrot steam drying was used in this study[11]. Lower SHS temperature is not recommended as it will increase the process steam requirement. On the other hand, high temperature may increase Volatile Organic Compounds (VOCs) emissions from RDF [17]. However, the optimum value should be found through experiments. Other values which may be used to calculate the amount of superheated steam and a basic design basis for the drying system are listed in Table 5-1.

As calculated in sub-chapter 4-5, minimum amount of energy to dry RDF from 0.29 to 0.1 moisture content is 2.73 MW, which has to be supplied by SHS in the direct steam dryer. If we assume that 5% of heat loss occurs in the drying loop, the amount of drying energy changes to almost 2.9MW. Equation 5-1 provides the amount of SHS for this purpose.

$$\dot{Q}_T = \dot{m}_{SHS} C_{pSHS} (T_{SHS,in} - T_{SHS,out}) \quad (5-1)$$

Where:

\dot{Q}_T : Total amount of drying energy pr time [MW]

\dot{m}_{SHS} : Mass flowrate of superheated steam [$\frac{kg}{s}$]

C_{pSHS} : Average specific heat capacity of SHS at average temperature of SHS inlet and outlet temperatures [$\frac{kJ}{kg.K}$]

$T_{SHS,in}$: Inlet temperature of SHS to the dryer [K]

$T_{SHS,out}$: Outlet temperature of SHS from the dryer [K]

Substitution of right values in equation 5-1 results in:

$$2.9MW = \dot{m}_{SHS} \left(\frac{kg}{s}\right) \times 1.91 \left(\frac{kJ}{kg.K}\right) \times (200 - 120)(K)$$

$$\dot{m}_{SHS} = 19.24 \frac{kg}{s} = 69.3 \frac{t}{h} \approx 70 \frac{t}{h}$$

If we assume that we are in the continuous mode of process, meaning that steam is already generated in the plant and it is being used in the drying loop, available energy in cooling air exhaust from the plant could be applied to compensate superheated steam energy reduction in the dryer. Figure 5-1, shows a preliminary process flow diagram (PFD) of this system, in which flow conditions are shown for all streams. This PFD is based on some presumed values suggested by the writer and some of the values in that were achieved via ASPEN HYSYS V8.6 simulator and the sketch is drawn using Microsoft Visio. As seen in Figure 5-1, SHS enters the dryer at 200°C and 1.25bar as stream 7 together with wet RDF as stream 1. Steam leaves the dryer at slightly superheated condition at 120°C and 1.2bar shown as stream 3. The amount of steam in stream 3 is different from stream 7 as it has additional evaporated moisture in itself. Since direct drying method is used, entrainment of some dust particles to the steam is inevitable. So that, a filter might be needed. There is also the possibility of VOC emission from RDF, due to gas high temperature [17], that is assumed to be treated in the filtration unit. Pressure drop of 0.1bar is assumed for the steam when it passes through the filter. Excess steam is taken out from the loop in stream 5 which equals to the amount of evaporated moisture. The rest of steam remains in the loop and should pass a blower to be pressurized by 0.2 bar. As a result of pressure rise, steam temperature would also increase to around 141.4°C in stream 6 before entering an air/steam heat exchanger. Hot air from the plant is fed to

the heat exchanger to give its energy to the steam. An approach temperature of 20°C is assumed for the heat exchanging unit so that, the air outlet temperature should be above 161°C. Two flow transmitters as FT1 and FT2 are installed to measure the flowrate in stream 4 and 7 respectively. The signals from these two flow transmitters control the valve in line 5 in a way that the difference between these two signals indicates how open the valve should be to let the excess steam release. Detailed control philosophy is not included, since Figure 5-1 is a basic PFD of the process. However, a very complex control system is needed to keep the system tight enough to avoid air entering into the system or steam loss.

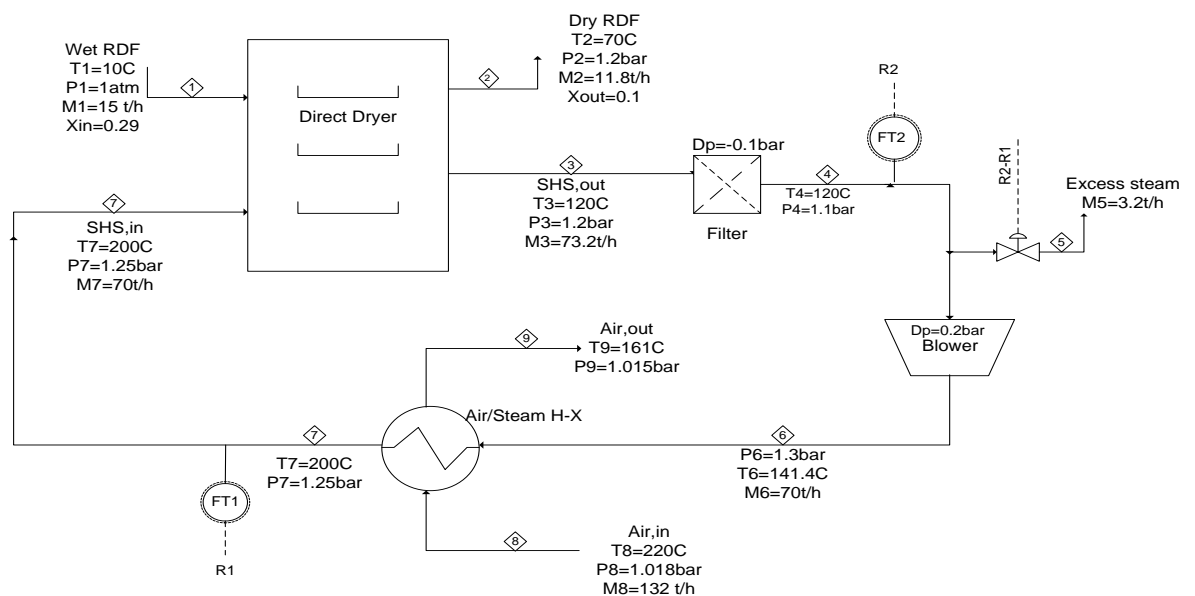


Figure 5-1 Basic PFD of drying process using a direct steam dryer

Table 5-1 Design basis for direct steam drying method

X_i	Inlet moisture content (wet basis) (kg water/kg wet RDF)	0.29
X_f	Final moisture content of RDF;(assumption)	0.1
\dot{M}_{RDF}	Mass flowrate of wet RDF entering the dryer	15000 kg/h
\dot{M}_s	Mass flowrate of dried RDF leaving the dryer	11833 kg/h
\dot{M}_{moist}	Mass flowrate of evaporated moisture	3167 kg/h
T_i	RDF initial temperature (assumption)	10°C
T_f	RDF final temperature after drying (assumption)	70°C
T_{vf}	Vapor temperature leaving the dryer (assumption)	120°C
$T_{SHS,in}$	Superheated steam temperature in the dryer inlet(assumption)	200°C

$T_{SHS,out}$	Superheated steam temperature in the dryer outlet(assumption)	120°C
P_{dryer}	Dryer pressure	1.2bara
T_v	Water saturated temperature at dryer pressure	104.8°C
C_{ps}	Constant RDF specific heat capacity (assumption)	1.4 (kJ/kg.K)
C_{pL}	Water specific heat capacity (assumed to be constant)	4.183(kJ/kg.K)
C_{pv}	Vapor specific heat capacity at average temperature of 112.4 °C (Note1)	1.89(kJ/kg.K)
C_{pSHS}	SHS specific heat capacity at average temperature of 160°C (Note 2)	1.91(kJ/kg.K)
$C_{p,air}$	Air specific heat capacity at average temperature of 180.5°C (Note 3)	1.022(kJ/kg.K)
λ	Water latent heat at dryer pressure	2265(kJ/kg)
Q_{loss}	Heat loss in the drying loop (assumption)	5%
Dp_{filter}	Pressure drop in the filter	-0.1bar
Dp_{blower}	Pressure difference in the blower	0.2 bar
$\Delta T_{approach}$	Approach temperature in air/steam heat exchanger	20°C
<p>Note 1: The average temperature in the vapor side $\bar{T} = \frac{120+104.8}{2} = 112.4^\circ\text{C}$</p> <p>Note 2: The average temperature in the SHS side $\bar{T} = \frac{200+120}{2} = 160^\circ\text{C}$</p> <p>Note 3: The average temperature of air stream in the heat exchanger $\bar{T} = \frac{220+161}{2} = 180.5^\circ\text{C}$</p>		

The output of suggested system in Figure 5-1 is summarized in Table 5-2. The values in the table were achieved by ASPEN HYSYS V8.6 in which blower efficiency was set to 75% as the default value and data from design basis were applied.

Table 5-2 Output of system suggested for direct steam drying

Steam flowrate	~70t/h
Excess steam flowrate	3.2 t/h
Energy consumption for steam temperature rise in steam/air H-X	2213kW Or 2515 kJ/kgH ₂ O
Energy consumption for the blower	795kW
Energy produced from excess steam	2048kW
Net thermal energy consumption	165 kW Or 185.6 kJ/kg H ₂ O

It should be noted that, summation of thermal energy consumption in the heat exchanger and the blower energy consumption should be almost equal to the required energy in the dryer. This pattern is not surprising as we have a closed loop system in Figure 5-1.

5-2 Indirect steam drying of RDF in Norcem plant

Indirect steam drying is the second theoretical alternative to dry RDF in Norcem plant which could be performed in an indirect dryer like steam tube rotary dryer. Just like what is suggested for direct steam drying, we need to know the amount of steam. Since no direct contact between steam and solid materials takes place, steam condensation is allowed in the dryer. As a result, latent heat of steam could be perfectly used to provide the energy for drying process. A basic design basis and relevant data to calculate the amount of steam and hot air are presented in Table 5-3.

By assuming that 10% heat loss takes place in the drying loop and saturated steam is used in the dryer, the amount of steam could be obtained using equation 5-2.

$$\dot{Q}_T = \dot{m}_{\text{steam}} \cdot \lambda_{120^\circ\text{C}} + \dot{m}_{\text{steam}} \cdot C_{p_w} \cdot (120^\circ\text{C} - T_{w,\text{out}}) \quad (5-2)$$

In which:

$\lambda_{120^\circ\text{C}}$: Latent heat of steam at 120°C , $[\frac{\text{kJ}}{\text{kg}}]$

C_{p_w} : Specific heat capacity of water $[\frac{\text{kJ}}{\text{kg}\cdot\text{K}}]$

$T_{w,\text{out}}$: Sub-cooled water temperature in the dryer outlet $[\text{C}]$

The amount of heat needed for drying, taking heat loss into account, would be:

$$\dot{Q}_T = 1.1 \times 2.8 = 3.07\text{MW}$$

$$3.07(\text{MW}) = \dot{m}_{\text{steam}} \times 2222 \left(\frac{\text{kJ}}{\text{kg}}\right) + \dot{m}_{\text{steam}} \cdot 4.183 \times (120^\circ\text{C} - 110^\circ\text{C})$$

$$\dot{m}_{\text{steam}} = 4.9 \frac{\text{t}}{\text{h}} \approx 5 \frac{\text{t}}{\text{h}}$$

This amount of steam should be generated in an air/steam heat exchanger. Available cooling air stream in the plant serves as the heating medium in the heat exchanger. So that, the amount of hot air (\dot{m}_{air}) producing about 5 t/h steam can be achieved by equation 5-3 in which \dot{Q}_{air} is the required heat provided by air that equals to the total amount of required energy for drying process.

It should be noted that an approach temperature of 20°C is assumed in the air/steam heat-exchanger. As a result, air outlet temperature should be above 140°C as given in the design basis.

$$\dot{Q}_{air} = \dot{Q}_T = \dot{m}_{air} \cdot C_{p,air} \cdot (T_{air,in} - T_{air,out}) \quad (5-3)$$

$$3.07\text{MW} = \dot{m}_{air} \times 1.022 \left(\frac{\text{kJ}}{\text{kg} \cdot \text{K}} \right) \times (220 - 140)(\text{K})$$

$$\dot{m}_{air} \approx 135.2 \frac{\text{t}}{\text{h}}$$

Based on all the calculated values, a basic process flow diagram of the drying process using an indirect drying method is suggested in Figure 5-2. This PFD may not be applicable for the real process, as some values and operating conditions are guessed.

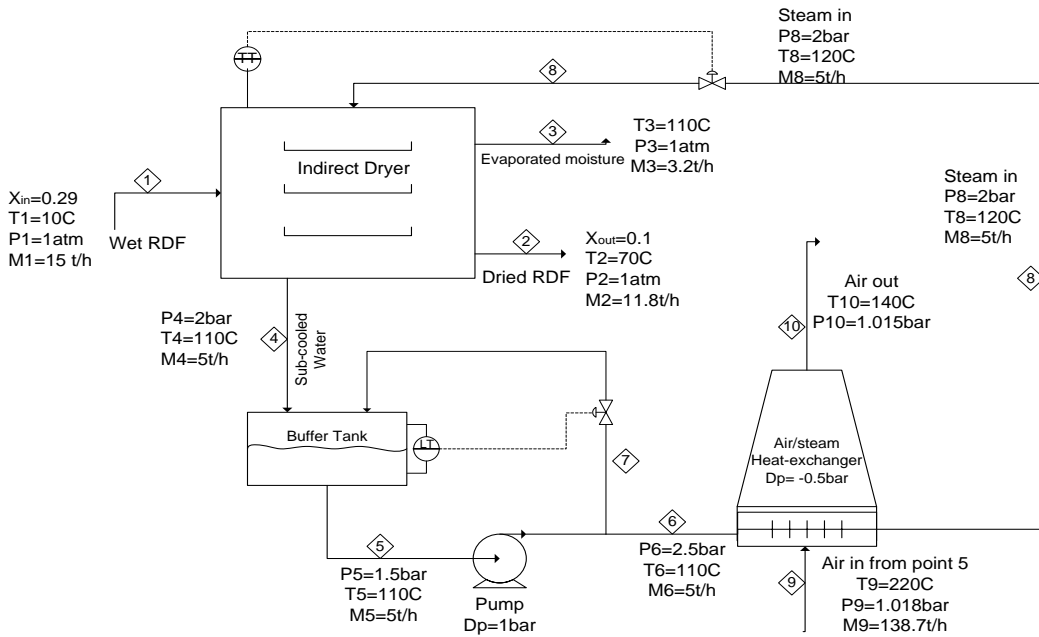


Figure 5-2 Basic PFD of drying process using an indirect dryer

As seen in Figure 5-2, wet RDF enters the dryer and is heated up indirectly by saturated steam having 2bar pressure. Dried RDF leaves the dryer at 70°C to be further used as a fuel. However, it could be a wrong guess to assume the same value for RDF outlet temperature in direct & indirect steam drying. Because, solid temperature could be higher in SSD method, as earlier discussed in sub-chapter 3-5.

Anyway, drying medium which is a slightly subcooled liquid, leaving the dryer at 2bar pressure enters a buffer tank before being pumped. It is assumed that maximum 0.5 bar

pressure drop for water could occur from the tank to the pump. To compensate the pressure drop in the loop, water passes a pump and its pressure raised by 1bar. To make sure that there is always a certain liquid level in the tank (to avoid pump cavitation), a branch of pump discharge flow is sent back to the buffer tank measured by a valve controlled by the level transmitter installed in the tank. Pump discharge flow (having pressure of 2.5bar and temperature of 110°C) enters air/steam heat exchanger to turn to steam. The required energy is supplied by hot air having a temperature of 220°C. Air outlet temperature should be at least 140°C to meet the approach temperature of 20°C for the heat exchanger. It is assumed that maximum 0.5bar pressure drop exists for the water stream in the heat exchanger. Finally, generated steam sent back to the dryer to complete the loop. A temperature sensor installed in the dryer also controls the steam flowrate into the dryer.

Table 5-3 Design Basis for indirect drying method

X_i	Inlet moisture content (wet basis) (kg water/kg wet RDF)	0.29
X_f	Final moisture content of RDF;(assumption)	0.1
\dot{M}_{RDF}	Mass flowrate of wet RDF entering the dryer	15000 kg/h
\dot{M}_s	Mass flowrate of dried RDF leaving the dryer	11833 kg/h
\dot{M}_{moist}	Mass flowrate of evaporated moisture	3167 kg/h
T_i	RDF initial temperature (assumption)	10°C
T_f	RDF final temperature after drying (assumption)	70°C
T_{vf}	Vapor temperature leaving the dryer (assumption)	110°C
$T_{st,in}$	Steam temperature in the dryer inlet(assumption)	120°C
$T_{st,out}$	Water temperature in the dryer outlet(assumption)	110°C
P_{st}	Steam pressure at the dryer inlet	2bar
P_{dryer}	Dryer pressure in the RDF side	atmospheric
T_v	Water saturated temperature at dryer pressure in RDF side	100°C
$T_{air,in}$	Inlet air temperature into the heat exchanger	220°C
$T_{air,out}$	Outlet air temperature out of heat exchanger	140°C
C_{ps}	RDF specific heat capacity (assumption)	1.4 (kJ/kg.K)
C_{pL}	Water specific heat capacity (assumed to be constant)	4.18(kJ/kg.K)
C_{pv}	Vapor specific heat capacity at average temperature of 105°C (Note 1)	1.89(kJ/kg.K)
$C_{p,air}$	Air specific heat capacity at average temperature of 180°C (Note 2)	1.02(kJ/kg.K)
λ	Water latent heat at steam pressure	2222(kJ/kg)
Q_{loss}	Heat loss in the drying loop (assumption)	10%

Dp_{dryer}	Pressure drop from the dryer to the pump entry	-0.5bar
Dp_{pump}	Pressure difference in the pump	0.6 bar
$\Delta T_{approach,HX}$	Approach temperature in air/steam heat exchanger	20°C
ΔT_{dryer}	Approach temperature in the dryer	10°C
<p>Note 1: The average temperature in the vapor side $\bar{T} = \frac{110+100}{2} = 105^{\circ}\text{C}$</p> <p>Note 2: The average temperature of air $\bar{T} = \frac{220+140}{2} = 180^{\circ}\text{C}$</p>		

The output of the system suggested in Figure 5-2 is presented in Table 5-4

Table 5-4 Output of system suggested for indirect steam drying

Steam flowrate	~5t/h
Evaporated moisture flowrate	3.2 t/h
Thermal energy consumption in heat exchanger	3149kW Or 3578 kJ/kg H ₂ O
Electricity consumption in the pump	0.12kW
Energy produced from excess steam	2041kW
Net thermal energy consumption	1108 kW Or 1246.5 kJ/kg H ₂ O

As seen in Table 5-4, thermal energy consumption in indirect steam drying is more than SSD method, as a result of expected more heat loss in indirect system which is already explained in sub-chapter 3.4.3.

5.3 Direct RDF drying with available hot air in Norcem plant

As stated already, apart from steam drying, there could be a third alternative for RDF drying in Norcem. The reason of suggesting this is abundance of available dry and almost dust-free hot air in the plant. As a result, if the hot cooling air, as stream 5 in Figure 4-1, serves as drying medium, there will be no need to process steam (i.e. an intermediate fluid) and utilize it as drying gas. So that, some equipment presented in PFD for direct and indirect steam drying system like steam/air heat exchanger could be removed.

Figure 5-3 displays a preliminary PFD for direct air drying system.

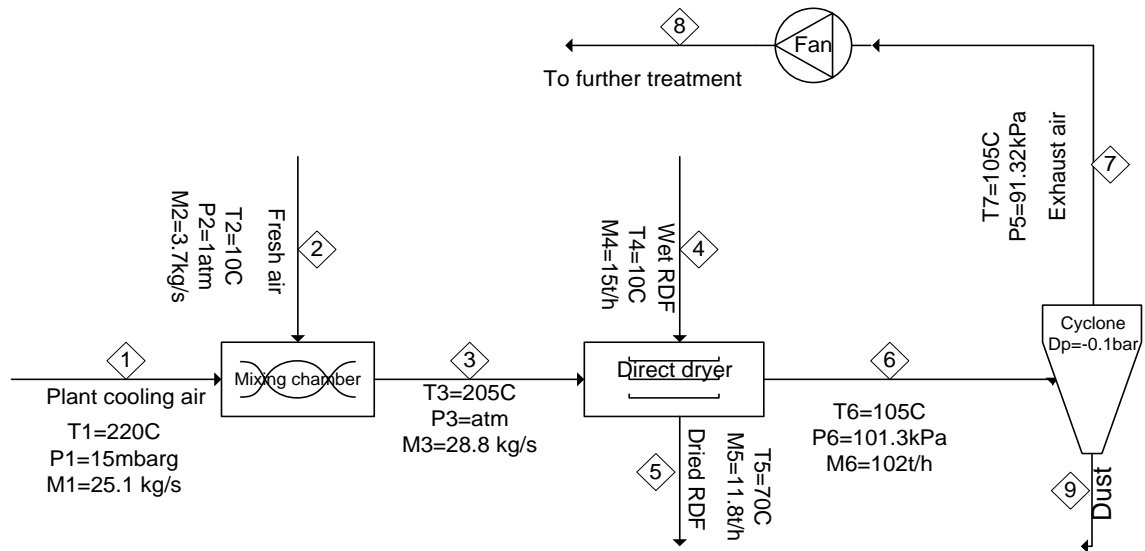


Figure 5-3 Direct air drying system PFD for RDF

Values for the gas inlet and outlet temperature in the dryer of Figure 5-3 and RDF outlet temperature were taken from one of the vendors' quotations for a drum dryer (Ref. appendix D). Mass flowrate of air was achieved through mass and energy balance calculation (could be found in sub-chapter 6-1). It is assumed that there is no pressure drop for the gas in the dryer and the pressure drop in the cyclone is assumed to be 0.1 bar based on some design tips for cyclones[18].

As seen in Figure 5-3, some of the plant cooling air is supplied to the drying system at 220°C and 5mbarg, as stream 1. It has to be cooled to 205°C in a mixing chamber where fresh air (at 10°C & 1atm) serves as the cooling medium. Exhaust air from the mixing chamber, stream 3, has the mass flowrate of 28.8 kg/s which was achieved from energy balance calculation around dryer (see sub-chapter 6-1). As a result, mass flowrates of stream 1&2 were obtained through mass and energy balance around the mixing chamber. So that, 25.1kg/s of cooling air from the plant is used for the drying, as seen in the PFD. A direct dryer that could be a rotary dryer as an example, is fed by stream 3 and RDF as stream 4. Most of the heat in dryer is consumed for the moisture evaporation rather than the solid temperature rise. However, in the ending part of the dryer, some heat could be consumed to raise the RDF temperature to 70°C (Ref. appendix D). Dryer outlet flow, stream 6, would be wet air at 105°C having the mass flowrate of 102 t/h. Dryer exhaust air may have considerable amount of dust, so it passes through the cyclone

having 0.1bar pressure drop. Exhaust wet air, stream 7, is pressurized by passing through a fan and is sent to further treatment system if necessary. The pressure of stream 8 is not specified as it depends of downstream facilities for further treatment.

The output of the system suggested in Figure 5-3 is presented in Table 5-5. It should be noted that energy consumption for the fan is not specified as it depends on the downstream process.

Table 5-5 Output of system suggested for direct air drying

Air mass flowrate to the dryer	~98.6t/h
Moisture amount	3.2 t/h
Thermal energy consumption for drying Assuming 5%heat loss	2835kW Or 3222 kJ/kgH ₂ O
Energy consumption for the exhaust fan	Not specified
Energy produced from excess steam	2048kW
Net thermal energy consumption	787 kW Or 894 kJ/kg H ₂ O

5.4 RDF drying system selection for Norcem plant

In order to select the most appropriate drying method from all three suggested systems, different factors should be taken into account, such as, energy consumption of the system, existence of auxiliary equipment, operability and controllability of the process, cost of the drying system and etc. However, drying system cost play an important role among all selection factors.

Based on data acquired for any of three alternatives and different vendors' advice, hot air drying system is selected as the best solution for the case in Norcem plant. Reasons for this selection are described more as following:

➤ Reasons to eliminate indirect steam drying option

The second option of RDF drying system in Norcem plant was indirect steam drying. In this method, based on suggested PFD of Figure 5-2, almost 5t/h saturated steam serves as heating medium in the dryer. Although, the amount of steam is much lower than the direct steam drying case, the necessity to have an additional heat exchanger is a

drawback which would have cost impact. However, heat exchanger in this system is smaller than direct system.

A pump in this system has the same role as blower in the direct steam drying which is an additional rotating equipment, the pump is of lower price compared to the blower in direct drying though.

Being in contact with a dryer manufacturer company, *Büttner Energie- und Trocknungstechnik GmbH*, resulted in collecting useful data and advice about steam tube bundle rotary dryers as one of the indirect types of dryers. *Büttner* is one of the supplier of drying systems internationally and their indirect drying systems could be applied in drying of woodchips, biomass and bulk solids and etc. Application of indirect rotary dryer was not recommended by this vendor. *“Indirect dryers properly work when the solid materials are homogeneous like solid drying in food industry. RDF is so heterogeneous for this type of dryer and feed properties are not invariant so that this dryer type is not as efficient as direct dryers. Indirect systems are mainly applicable when there is a restriction for direct contact of drying medium with solid materials or when the size matters, as indirect dryers are smaller than direct ones. Both of these limitations are not the case in your plant. So that, we don’t recommend tube bundle dryer for your case.”* Said Carsten Otto, sales director at *Büttner*, on a telephone conversation on 03/05/2016.

From the energy consumption viewpoint, as shown in Table 5-4, net thermal energy consumption is about 3578 kJ/kg H₂O without vapour heat recovery which is much higher than direct steam air drying systems.

➤ **Reasons to eliminate direct steam drying option**

Direct steam drying method uses superheated steam (SHS) as drying medium. The amount of SHS needed for this type is calculated as almost 70t/h which has to flow in the loop, as shown in Figure 5-1. It means that larger pipes, valves and fittings will be required, having a profound cost impact on the whole system. The size of pipes to let 70t/h of SHS flow through, could be found via equation 5-4 and 5-5.

$$\dot{m} = \rho \cdot U \cdot A \quad (5-4)$$

$$A = \pi \cdot \frac{D^2}{4} \quad (5-5)$$

In which:

\dot{m} : Mass flowrate of gas [$\frac{\text{kg}}{\text{s}}$]

ρ : Gas density [$\frac{\text{kg}}{\text{m}^3}$]

U : Gas velocity [$\frac{\text{m}}{\text{s}}$]

A : Pipe cross sectional area [m^2]

D : Internal diameter of the pipe [m]

To calculate the pipe diameter, a design steam velocity should be chosen. Steam velocity in a pipeline is dependent on steam properties, pressure and the pipe diameter. Greater velocity in the pipe leads to more pressure drop and noise and erosion problems. However, higher velocity can be anticipated for dry steam (i.e. SHS) than wet steam (i.e. saturated steam) because wet steam is usually more erosive[19].

A rough estimate of the allowable SHS velocity in a pipeline is chosen to be almost 50m/s based on suggested values in *Piping calculation manual* [19]. The reason behind this selection is that we have dry steam at almost atmospheric pressure so that the lower speed limit was taken. Based on stream conditions in Figure 5-1, steam density is between 0.576-0.688kg/m³ for different flow streams. This limit was achieved through Aspen HYSYS V8.6 data base for SHS in different conditions. Applying equations 5-4 and 5-5 would give the value for pipe diameter in the range of DN800-DN900 which is a substantially large pipe size.

Another reason is the additional heat exchanger required to make the steam more superheat, adding to the equipment cost. On the contrary, if air is directly applied in the dryer, the need for having an additional heat exchanger is eliminated. Instead, a mixing chamber would play the role of heat exchanging unit in air drying system that is simpler and less expensive device, normally supplied as an integrated part of the drying package by vendors. Additional heat exchanger may also cause some heat waste, lowering the thermal efficiency of the plant.

Since exhaust gas in SSD technique is steam, excess steam heat recovery is more straightforward than air drying methods. So that thermal energy consumption could

decrease substantially if there is an application for the purged steam either directly or indirectly in the system.

As seen in Table 5-2, net thermal energy is 185.6 kJ/kg H₂O which is quite low compare to other options. However, this energy amount is based on the assumption of excess steam heat recovery. Since there will be no potential use of the excess heat in the plant, according to a conversation with *L. A. Tokheim* in the meeting on *April 2016*, purged steam energy is not recovered. Although there may be a possibility to use this energy in the drying loop itself, having substantial excess energy in the plant is not an encouraging factor to think about this option. Consequently, thermal energy consumption may substantially increase to 2489.6kJ/kg H₂O without excess heat recovery which does not seem a well more energy efficient system (compared to air drying options).

Since the drying temperature is high, the whole system should be well insulated to avoid heat loss due to conduction and radiation from the system. This item is not a drawback compared to suggested air drying method also operating at high temperature of 200°C. However, it is definitely a disadvantage when lower temperature air dryers like belt and fluid-bed dryers are applied which are more explained in discussion chapter.

As explained in chapter 3, steam drying system has to be tight enough to avoid air penetration into, that could be quite challenging to control if excess steam heat recovery is favourable. A more complex control system which is generally inherent to steam processing system is also required for SSD technique.

Due to high-temperature steam utilization as drying medium, Volatile Organic Carbons (VOCs) may emit considerably[17] and the need for an exhaust air treatment system is vital for both dust and VOCs removal.

➤ **Reasons to choose direct drying with hot air**

First of all, availability of cooling air stream with the temperature of 220°C does not introduce major process equipment to first producing steam (steam generator) and later heating that up (heat exchanger). So that, the total equipment cost may reduce to some extent.

Secondly, no commercial dryer was found to use steam to dry RDF. Lack of experience in commercial steam drying application for RDF could be another reason of choosing air dryers, as they are extensively used to dry solid wastes in the industry.

From the energy point of view, we don't need to have an intermediate heating medium so, we avoid some heat loss in this case. In addition, thermal energy consumption for both air drying and SSD technique does not differ substantially. Although, SSD is thermally more efficient than air drying system.

On the other hand, the dryer will have no steam equipment, results in lower capital and operating cost, lowering maintenance requirement and simpler control system.

Air drying method could have some drawbacks, as well. Exhaust air treatment may be required as there is a direct contact between solid materials and the hot air (just like SSD). Air dryer may not be as efficient as direct steam dryer, since drying rate in steam drying is higher as explained in sub-chapter 3.5.

Some air dryers require higher heat source (i.e. higher air temperatures) like rotary drum dryers, while others could operate at lower temperature ranges like belt dryers. Since higher air temperatures cause higher VOC emissions[17], dryer systems with lower temperatures are preferred, if applicable.

6. Dryer sizing and cost estimation

This chapter is giving a rough estimate for the size and price of air drying system, as the most proper solution for the case in Norcem plant. Although it may be possible to calculate dryers dimensions and total costs by hand and using online sources, more accurate information could be acquired through vendors' quotations.

6.1 Example of hand calculated size of RDF air dryer

Following the mass and energy balance done in the previous chapters, dryer size could be calculated via some equations suggested in the literatures. However, more accurate design could be accomplished by experiments and tests on pilot scales.

Equation 6-1, shows the governing formula used to calculate dryer size and/or heat transfer coefficient.

$$\dot{Q}_T = U \cdot A \cdot \overline{\Delta T} \quad (6-1)$$

Where:

\dot{Q}_T : Total heat of drying per time [W]

U: Overall heat transfer coefficient [$\frac{W}{m^2.K}$]

A: Heat transfer area [m^2]

$\overline{\Delta T}$: Average temperature difference (logarithmic mean temperature difference in some cases)

The overall heat transfer coefficient (U) can be attained either by calculation or experiments for most of the dryers. Area of the heat transfer (A) could be easy to determine for some dryers (e.g. in belt dryers); however, finding the area in rotary dryers is not simple. To overcome this problem, a volumetric heat transfer coefficient is defined to calculate the dryer volume instead of area. As a consequence, equation 6-1 turns to equation 6-2.

$$\dot{Q}_T = Ua \cdot V \cdot \overline{\Delta T} \quad (6-2)$$

In equation 6-2, "a" is defined as heat transfer area per unit dryer volume which is an unknown variable.

Ua: Volumetric heat transfer coefficient [$\frac{W}{m^3.K}$]

V: Volume of the dryer [m³]

Turning back to the case of Norcem plant, different dryer types which were explained in chapter 3 such as belt dryers, rotary dryers and fluid bed dryers have been found widely applied as commercial solid wastes and RDF processing facilities.

In this part, a rotary air dryer is chosen as an example to do the hand calculation to achieve the dryer size.

As stated, equation 6-2 is the governing formula for rotary dryers heat transfer. To estimate the value of volumetric heat transfer coefficient, an empirical correlation as equation 6-3 could be used[15].

$$Ua = \frac{0.5 G^{0.67}}{D} \quad (6-3)$$

Substitution of equation 6-3 in 6-2 would give equation 6-4 which is being used to design a rotary dryer[15].

$$\dot{Q}_T = 0.125 \pi DL G^{0.67} \overline{\Delta T} \quad (6-4)$$

In which:

\dot{Q}_T : Heat transfer rate [$\frac{\text{Btu}}{\text{h}}$]

D: Dryer diameter [ft]

L: Dryer length [ft]

G: Mass velocity of the drying gas [$\frac{\text{lb}}{\text{ft}^2 \cdot \text{h}}$]

$\overline{\Delta T}$: Logarithmic mean temperature of wet bulb depression in the dryer inlet and outlet as average temperature difference [°F]

In direct dryers, rotary dryer more specifically, the gas outlet temperature plays an important role on the economy of the dryer. It could be obtained through equation 6-5 which defines the number of heat transfer unit in a water/air dryer systems[15].

$$N_t = \ln \frac{T_1 - T_w}{T_2 - T_w} \quad (6-5)$$

Different variables in equation 6-5 are as follows:

N_t : Number of heat transfer units in a dryer based on the gas

T_1 : Gas initial temperature

T_2 : Gas outlet temperature

T_w : Wet-bulb temperature

N_t should be between 1.5 to 2.5, which is the range of heat transfer unit in industrial rotary dryers with efficient operation. As a result, gas outlet temperature could be achieved [15].

It is a common assumption to take $T_v = T_w$, as T_w is usually unknown [15].

In order to find the dryer size for the case in Norcem, some assumptions are taken as following:

- 1) Dryers works in atmospheric pressure
- 2) Inlet gas temperature of 205°C or 401°F, recommended by vendors
- 3) N_t is assumed to be 2.5.
- 4) Solid outlet temperature is 70°C, recommended by a vendor
- 5) 5% heat loss in the dryer

Based on aforementioned assumptions, outlet temperature of the drying gas through equation 6-5 is:

$$2.5 = \ln\left(\frac{205 - 100}{T_2 - 100}\right)$$

$$T_2 = 108^\circ\text{C} = 227^\circ\text{F}$$

This temperature value seems reasonable as it is in the range of vendor's data for rotary dryer (see appendix D).

The next step is to calculate drying energy rate. By inserting right data (some taken from mass balance and design basis data in chapter 5) in the equation 4-6, drying energy rate is as following:

$$Q_{\min, \text{drying}} = \dot{M}_s C_{ps} (T_{sf} - T_{si}) + \dot{M}_{\text{RDF}} X_i C_{pL} (T_v - T_{si}) + \dot{M}_{\text{moist}} \cdot \lambda + \dot{M}_{\text{RDF}} X_f C_{pL} (T_{sf} - T_v) + \dot{M}_{\text{moist}} \cdot C_{pv} (T_{vf} - T_v) \quad (4-6)$$

$$\begin{aligned} Q_{\min, \text{drying}} &= 11833 \times 1.4 \times (70 - 10) + 15000 \times 0.29 \times 4.183 \times (100 - 10) + 3167 \\ &\quad \times 2256 + 15000 \times 0.1 \times 4.183 \times (70 - 100) + 3167 \times 1.89 \times (108 - 100) \\ &= 2.7\text{MW} \end{aligned}$$

Assuming 5% heat loss would give:

$$Q_T = 1.05 \times Q_{\min, \text{drying}} = 2.835\text{MW}$$

Next step is the logarithmic mean temperature calculation which can be achieved through equation 6-6.

$$\overline{\Delta T} = \frac{(T_2 - T_w) - (T_1 - T_w)}{\ln\left(\frac{T_2 - T_w}{T_1 - T_w}\right)} \quad (6-6)$$

Inserting right values in equation 6-6 in °F gives:

$$\overline{\Delta T} = \frac{(227 - 212) - (401 - 212)}{\ln\left(\frac{227 - 212}{401 - 212}\right)} = 68.7^\circ\text{F}$$

Mass velocity of the drying gas could be achieved using equation 6-7, as below:

$$G = \frac{\dot{m}_{\text{air}}}{A} \quad (6-7)$$

In which, G is the mass velocity of the air, \dot{m}_{air} is the mass flowrate of the air and A is the cross-sectional area of the dryer.

$$\dot{Q}_T = \dot{m}_{\text{air}} \cdot C_{p_{\text{air}}} (T_1 - T_2)$$

$C_{p_{\text{air}}}$ can be achieved at the average temperature between T_1 and T_2 . Using Aspen HYSYS V8.6 gives the value of 1.015 kJ/kg at 155°C.

So that,

$$2835 = \dot{m}_{\text{air}} \times 1.015 \times (205 - 108)$$

$$\dot{m}_{\text{air}} = 28.8 \text{ kg/s}$$

Mass velocity of the air in rotary dryers is mostly between 0.5 (kg/m².s) to 5(kg/m².s).

However, the upper range is usually preferred if dust entrainment is not considerable.

The value of 0.5(kg/m².s) is assumed for this case. So that, "A" could be obtained via equation 6-7 as below:

$$0.5 \frac{\text{kg}}{\text{m}^2 \cdot \text{s}} = \frac{28.8 \frac{\text{kg}}{\text{s}}}{A}$$

$$A = 57.6 \text{ m}^2 = 620\text{ft}^2$$

As a result, diameter of the dryer could be achieved as following:

$$A = \frac{\pi}{4} \cdot D^2$$

$$D = 28.1 \text{ ft}$$

Substitution of the values with the correct dimensions into equation 6-4 results in:

$$L = 243.4\text{ft}$$

In order to verify the guessed value for the air mass velocity, L/D ratio checking suggested in some literatures. This ratio lies in the range of 4 to 10 for most efficient commercial dryers [5, 6]. Based on so far calculations, L/D ratio is 8.7 which falls in the range; however, taking the lower range value for mass velocity of the air (i.e. 0.5 kg/m².s)

resulted in a very large dryer. On the other hand, since RDF materials are not so dusty, taking a middle range value (e.g. 2.5 kg/m².s) seems more reasonable. So that, the dryer size from re-calculation will be:

$$D = 12.6 \text{ ft} = 3.84\text{m}$$

$$L = 185.2 \text{ ft} = 56.4\text{m}$$

As a result, L/D ratio will be 14.7 which is out of suggested range; although, the dryer is more became smaller. However, the value for the dryer length (even the new value based on the new air mass velocity) is so far from the value suggested by the vendor. The reason could stem from applying different experimental equation and assumptions and taking different value for the average temperature.

6.2 Dryers size and price by vendors

Over the period of the project, some European dryer manufacturers were contacted to acquire more accurate data of the size and cost of RDF air dryers. The information was obtained through sending inquiries to different vendors. This attempt was made in assistance with a recently established company called *SolutionBring D/A* which has the focus on finding innovative and novel solutions for different business areas.

For this purpose, a technical specification was prepared, covering all required data for vendors. Several tele-communications were done for clarifications, letting vendors have a closer insight to the case. The exact data in the budget quotes received by *SolutionBring D/A* are confidential and is avoided to be included in the main report body due to legal terms and conditions. However they are accessible in appendix D for the writer, supervisor and the external assessor.

The extensive attempt to contact vendors resulted in receiving three (3) quotations: two belt dryer solutions and one rotary drum dryer for comparison purposes, as it was not recommended by the corresponding vendor. The size of these dryers are presented in Table 6-1.

The fourth vendor contacted via *Solution Bring*, was the Dutch company of *Tema Process B.V* which offers a fluid bed dryer solution. However, despite several phone calls, no quotation was received at the requested date.

As seen in Table 6-1, ANDRITZ rotary dryer size differs from what is estimated by hand calculation.

Table 6-1 Air dryers size suggested by vendors

Dryer type	Belt dryer	Rotary drum dryer
Actual drying surface/volume	84m ² – 140m ²	450m ³ – 490m ³
Drying package price range	(0.5 –1.5) M€	(2 –3) M€
Vendors	-Stela GmbH -ANDRITZ Fließbett Systeme GmbH	ANDRITZ Fließbett Systeme GmbH

Following vendors budgetary quotations, the estimated cost of dryer packages could be also seen in Table 6-1 which are between 0.5 to 1.5 M€ for belt dryers and 2 to 3 M€ for rotary drum dryer. Vendors' prices are available in Appendix D, accessible for the supervisor and external assessor only. These prices are non-committal based, and shall be confirmed by vendors if necessary.

Apart from the capital cost of the drying package, utility cost as a part of OPEX could also be acquired from vendors' data. Utility consumption consists of electricity and water for these drying packages. Water consumption for belt dryer varies between 1.5-200 m³/h in which upper value shows the consumption where the exhaust gas condenser is applied (see vendors' quotes in appendix D). Water supply has been considered as an internal process (treatment and recycling) within Norcem area so, no additional considerable cost is assumed to be imposed to the plant.

As per *Statistisk sentralbyrå*, electricity price for the energy intensive manufacturing, was 31.8 (Øre/kW.h), in 4th quarter of year 2015[20]. Taking this number and multiplying that to the electricity consumption of the drying process (taken from vendors' quotations) will give the total annual electricity cost. Table 6-2 shows the calculated range of electricity consumption and cost.

Table 6-2 Power consumption and cost for various drying packages

Dryer type	Power consumption [kW]	Electricity cost [kNOK/yr]
Belt dryer	180-270	400-601
Rotary drum dryer	300-330	668-735

7. Discussion

RDF as a supplementary fuel may need drying for LHV improvement before entering to the combustion chamber and rises the flame temperature inside combustor. Meanwhile, drying of RDF could reduce the required volume of combustion chamber, adding more capacity to the existing combustor at the plant. This is due to removal of moisture off RDF which avoids steam generation inside the combustor.

Several drying methods exist in which steam or hot air/gas serve as drying medium (see chapter 3). However, the focus of this project was on evaluation of steam drying approach over air drying systems and its functionality to dry RDF being used in Norcem cement plant, as a case study.

No commercial application of RDF steam drying was found while working on this project. Although, extensive studies have been done for application of steam drying in food and paper industries and drying of bulky solids like in the wood industry. Lack of enough data to treat RDF with steam, led to taking some assumptions and contacting some dryer manufacturers to take advice. A process flow diagram for direct and indirect steam drying, as suggested in chapter 5, may not (efficiently) work in practice, although they sound reasonable in theory. To examine suggested systems, an experimental set-up is needed, specifically for RDF, to verify taken assumptions. Because RDF drying characteristic with steam is not well-known.

Direct and indirect steam drying as well as air drying method were extensively discussed in chapter 5, as three options that could be theoretically applied in Norcem. However, air drying method outweighed the other two alternatives.

The main reason of direct steam drying elimination was the necessity to handle substantial amount of steam (i.e. almost 70t/h) in a drying loop. This results in huge pipe size of DN800-DN900 and larger valves and fittings installations which could have large cost impact. A heat exchanger installation to compensate the decreased energy amount of superheated steam by available hot air was another reason of excluding steam drying as the best option due to its additional cost to the system. This heat exchanger could be a large one as a result of high steam flowrate.

Although, steam drying generally offers faster drying rate, experimental test results on RDF drying will be also required to find out drying characteristics and the inversion temperature (discussed in chapter 3). This will show the exact temperature where the

steam drying would be faster than the air drying method. In addition, controllability of a direct steam drying system is not an easy task compared to air drying system (see sub-chapter 3.5).

It should be noted that, superheated steam drying (SSD) might be generally a good option for drying of RDF where energy consumption really matters. Because, easier excess steam energy recovery could reduce the amount of needed energy by up to 75% which could be used elsewhere in the process plant (see sub-chapter 3.5). This advantage of SSD technique is not applicable for the Norcem case as no application for excess heat could be found.

Indirect steam drying method was also not a perfect option as it has shown lower drying efficiency for heterogeneous solids with invariant properties like RDF (Ref. chapter 5). Furthermore, steam generation and necessity for having a water/air heat-exchanger to turn water to saturated steam is still a disadvantage. Moreover, indirect dryers are mostly more difficult to manufacture and more expensive than the direct counterparts[5]. From the energy consumption point of view, indirect dryer showed the least energy efficient system among the three options due to the more heat loss in the dryer (see sub-chapter 5.2)

As a whole, direct air drying system founded to be the best solution for the Norcem case study. The main reason for this selection could be the availability of hot, dry and almost dust free air in the plant which can be directly used as the drying gas and do not introduce an intermediate medium (i.e. steam) to the drying system. Although, some fresh air may be mixed with hot cooling air from the plant to regulate the drying air temperature. Among different types of air dryers, three of them are commercially used to dry RDF, such as: belt dryers, rotary dryers and fluid-bed dryers. Efforts to contact vendors for these three dryer types, resulted in receiving three quotations for belt and rotary dryers which are more discussed.

Despite of several contacts with the fluid-bed dryer manufacturer, no quotation received from them. So that, this option will not be discussed in this part.

Evaluation of three received quotations (i.e. one for rotary dryer and two for belt dryers) from two vendors, ANDRITZ and Stela, showed that belt dryers are more cost-effective than the rotary dryers. However, other factors than the price should be also taken into account which are presented in this part.

One important and noticeable difference between belt and rotary dryers is the operation temperature. Drying temperature is lower in both quoted belt dryers than the rotary type. ANDRITZ rotary dryer, takes the hot air at high temperature of almost 200°C and discharge it at almost 100°C (Ref. appendix D). Due to high temperature, more volatile organic compounds (VOCs) could emit from the RDF materials which has to be treated in downstream facilities [17]. Moreover, fire hazard/ dust explosion has more probability to occur in the rotary drum dryer as a result of higher temperature. However, these are not a problem in belt dryers as they work in lower temperature. In addition, as ANDRITZ documents show (in appendix D), rotary dryer has more tendency to dust carry over. That is why, a cyclone or a bag filter downstream of the exhaust gas is recommended by the vendor. As seen in appendix D, thermal energy consumption of rotary dryer is more than belt dryers meaning that rotary drum dryer is less energy efficient for our case.

As already mentioned, the two quoted belt dryers operate at temperature as low as 100°C which leads to lower VOC emission and fire risks. Dust emission with exhaust air has been reported to be less than 10mg/Nm³ of the wet gas for both belt dryers, corresponding to 10.06mg/Nm³ dry gas for the rotary case for instance. This limit is well below the allowable range of dust emission in Norway which is maximum 30mg/Nm³ of dry gas[21].

Exhaust gas temperature in both belt dryers is about 40°C. This amount has the minimum environmental footprint. However, if exhaust cooling air temperature from the plant (i.e. drying gas) is allowed to be lower than 100°C, belt dryers could be applicable.

Operational cost for drum dryers vs. belt dryers were presented in chapter 6. It showed that operating cost for belt dryers were lower thanks to lower power consumption. Other costs, like maintenance and services, however, shall be evaluated in detail with vendors. Although, drum dryers seem to have a higher maintenance requirements than belt dryers because of being a heavy rotating mechanical item.

8. Conclusion

RDF, as a supplementary fuel is widely used in some energy intensive industries like cement production plants. The lower RDF moisture content, the more heating value it has; therefore, RDF drying is an essential fuel treatment process.

Various drying methods consist of conventional direct air dryers and steam dryers (direct and indirect) have been evaluated to check their applicability for RDF drying. In order to make this comparison more sensible, drying of RDF being used in Norcem cement plant was taken as a case study. Available exhaust cooling air (having $T=220^{\circ}\text{C}$) was assumed to supply required energy for drying purposes in this plant.

Mass and energy balance of Norcem plant indicated that approximately 3.2t/h moisture should be evaporated in an optimum drying process, reducing the RDF moisture content from 29 to 10 wt%. According to a suggested design basis, almost 70 t/h of superheated steam with a temperature of 200°C could thermodynamically serve as the hot gas in a drying loop, however, no commercial application was found utilizing dry steam for RDF drying. Indirect steam drying was also taken as an option. Even though, only 5t/h of saturated steam was thermodynamically needed in the suggested system, it was understood that indirect dryers do not efficiently work for heterogeneous & varying characteristic feeds as RDF is.

The last alternative for the Norcem case was direct RDF drying utilizing the plant hot air as drying gas. This option considered as the best solution. Among various types of air dryers, belt dryers, rotary drum dryers and fluid-bed dryers were found extensively used in commercial industry. Collecting of practical data and specifications from different vendors of belt and rotary dryers, proved that belt dryers overtake the rotary type in terms of CAPEX and OPEX, emission and energy efficiency. No comparison was done with fluid-bed dryer as no vendor feedback was received for this type.

Some further attempts are recommended for process development prior to any decision made for plant procurement and installation. For instance, collecting more practical data on RDF composition and characteristics is required to minimize uncertainties in calculations. In addition, setting up a pilot unit, may help to investigate different processes, including steam drying as no commercial steam drying was found. Furthermore, fluid-bed dryer seems to be a practical alternative to belt dryers for this purpose so, it would be reasonably important to evaluate this process as an option.

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Appendices

Appendix A: Task description

Appendix B: Norcem plant data for the case study

Appendix C: Calculations

Appendix D: Technical specification and vendors' quotations (confidential)

Appendix A: Task description



Telemark University College

Faculty of Technology

FMH606 Master's Thesis

Title: Steam drying of wet waste fuels

TUC supervisor: Lars-André Tokheim

External partner: -

Task background:

Solid waste fuels, which are often partly biogenic, are used increasingly in the industry as a replacement for coal or other fossil fuels, for example in the cement industry. The purpose is partly to reduce net CO₂ emissions, partly to reduce fuel costs.

Replacement of coal by waste fuels is, however, not always trivial, since these fuels in many cases have characteristics that may give negative operational impacts. One such characteristic is higher moisture content, which reduces the heating value of the fuel.

Pre-drying of wet waste fuels would increase the heating value and thus make it an even more attractive coal replacement fuel. Drying requires thermal energy, but in some cases, waste heat may be available, and this energy could be used for drying purposes. Different process schemes could be used, and steam drying could be an interesting alternative.

Task description:

The task should include the following subtasks:

- Literature study on waste drying methods
- Steam drying concept description
- Design basis
- Mass and energy balance
- Specification of steam characteristics
- Potential drier design (size, number of units, dimensions, material choice etc)
- Process flow diagram
- Cost estimation



Student category:

PT or EET students

Practical arrangements:

-

Signatures:

Student (date and signature): 01.02.16 Fatimah Anueli

Supervisor (date and signature): 01.02.16 Len Andre Tahlin

Appendix B: Norcem plant data for the case study
**B1: Norcem Brevik, Kiln #6: Basis for potential waste
heat utilization**

Table 1: Typical gas characteristics

Point	Gas flow rate [Nm ³ /h]	Temperature [°C]	Pressure [mbarg]	N ₂ [vol%]	CO ₂ [vol%]	H ₂ O [vol%]	O ₂ [vol%]	Dust [mg/Nm ³]
1	132 250	386	-80	61	23	9	7	50 000
2		120						
3	132 250	386	-80	61	23	9	7	50 000
4		224						
5	180 000	220 ^a	5	79	0	0	21	5

^a Reduction to ~100 °C would be ok (relevant if a WHRU is installed).

Lars-André Tokheim,

USN, 2016-02-12

Appendix B: Norcem plant data for the case study

B2: Typical RDF characteristics in Norcem

Typical RDF characteristics

L.A. Tokheim, updated 2016-05-20

Typical proximate analysis

Parameter	Unit	Value
Moisture	wt%	29
Volatiles	wt%	49
Char	wt%	5
Ash	wt%	17
Sum	wt%	100

Typical heating value

Parameter	Unit	Value
Lower heating value	MJ/kg	12

Typical ultimate analysis

Parameter	Unit	Value
Moisture	wt%	29.0
C	wt%	28.8
H	wt%	4.1
O	wt%	20.4
N	wt%	0.5
S	wt%	0.2
Ash	wt%	17.0
Sum	wt%	100.0

Example of (normalized) RDF ash composition (dry basis)

Parameter	Unit	Value
CaO	wt%	25
Al ₂ O ₃	wt%	33
Fe ₂ O ₃	wt%	12
K ₂ O	wt%	13
Na ₂ O	wt%	17
Sum	wt%	100

Typical operational values

RDF consumption	t/y	100 000
Typical kiln run time	h/y	7 000

Appendix C: Calculations

C1: Available excess heat calculation in Norcem Cement Plant

C1: Available excess heat calculation in Norcem Cement Plant

A list of necessary equations for this calculation is as following:

$$\dot{Q} = \dot{n} \bar{c}_p \Delta T \quad (C-1)$$

$$\dot{Q} = \dot{m} \bar{c}_p \Delta T \quad (C-2)$$

$$P \cdot \dot{V} = \dot{n} \cdot R \cdot T \quad (C-3)$$

$$\dot{m} = \dot{n} \cdot MW \quad (C-4)$$

$$\bar{c}_p = \sum x_i \cdot \bar{c}_{p_i} \quad (C-5)$$

Different notations in the mentioned formulas:

\dot{Q} : Energy of the gas per second (Power) [MW]

\dot{n} : Gas molar flowrate [$\frac{\text{mol}}{\text{h}}$]

\dot{m} : Gas mass flowrate [$\frac{\text{kg}}{\text{h}}$]

\bar{c}_p : Average specific heat capacity of the gas, molar based [$\frac{\text{J}}{\text{mol.K}}$]

\bar{c}_p : Average specific heat capacity of the gas, mass based [$\frac{\text{J}}{\text{kg.K}}$]

ΔT : Temperature difference [K]

P: Pressure [Pa]

\dot{V} : Gas volumetric flowrate [$\frac{\text{m}^3}{\text{s}}$]

R: Universal gas constant [$\frac{\text{J}}{\text{mol.K}}$]

T: Temperature [K]

MW: Molecular weight [$\frac{\text{gr}}{\text{mol}}$]

x_i : Mole fraction of component i

Calculation of molar flowrate for 3 available sources:

In order to calculate molar flowrate of the gas, normal condition should be applied to be used in the gas law formula (C-3). Plant data (given in appendix B) should be also used for this purpose.

Normal conditions:

$$T = 0^{\circ}\text{C} = 273.15\text{K}$$

$$P = 1\text{atm} = 101325\text{ Pa}$$

$$\dot{n}_1 = \frac{P_N \cdot \dot{V}_{N1}}{R \cdot T_N} = \frac{101325(\text{Pa}) \times 132250\left(\frac{\text{Nm}^3}{\text{h}}\right)}{8.314\left(\frac{\text{Pa} \cdot \text{m}^3}{\text{mol} \cdot \text{K}}\right) \times 273.15(\text{K})} = 5900666.5 \frac{\text{mol}}{\text{h}}$$

$$\dot{n}_3 = \frac{P_N \cdot \dot{V}_{N3}}{R \cdot T_N} = \frac{101325(\text{Pa}) \times 132250\left(\frac{\text{Nm}^3}{\text{h}}\right)}{8.314\left(\frac{\text{Pa} \cdot \text{m}^3}{\text{mol} \cdot \text{K}}\right) \times 273.15(\text{K})} = 5900666.5 \frac{\text{mol}}{\text{h}}$$

$$\dot{n}_5 = \frac{P_N \cdot \dot{V}_{N5}}{R \cdot T_N} = \frac{101325(\text{Pa}) \times 180000\left(\frac{\text{Nm}^3}{\text{h}}\right)}{8.314\left(\frac{\text{Pa} \cdot \text{m}^3}{\text{mol} \cdot \text{K}}\right) \times 273.15(\text{K})} = 8031152.87 \frac{\text{mol}}{\text{h}}$$

Mass flowrate of cooling air at point 5:

$$MW_{\text{air}} = 28.97 \frac{\text{g}}{\text{mol}}$$

$$\dot{m}_5 = \dot{n}_5 \cdot MW_{\text{air}} = 8031152.87 \times 28.97 \times 0.001 = 232662.5 \frac{\text{kg}}{\text{h}}$$

Calculation of average \bar{c}_p or $\bar{\hat{c}}_p$:

Equation (C-5) can be applied to calculate heat capacity of the mixture.

NB! All specific heat values are taken from ASPEN HYSYS V8.6 data base for each component at the wanted temperature.

Compositions and specific heat capacities for each component in stream 1 are shown in table C-1.

NB! Specific heat values are taken at the average temperature between point 1 & 2 in Norcem PFD which is:

$$\bar{T}_{1-2} = \frac{386 + 120}{2} = 253^{\circ}\text{C} = 526.15\text{K}$$

From table C-1 and equation C-5, the average \bar{c}_p value is:

$$\bar{c}_{p_{1-2}} = \sum x_i \cdot \bar{c}_{p_i} = 33.943 \frac{\text{J}}{\text{mol.K}}$$

Table C-1: Gas composition and \bar{c}_p values of each component in condition tower 1.

Component	N ₂	CO ₂	H ₂ O	O ₂
x _i	0.61	0.23	0.09	0.07
\bar{c}_{p_i} at $\bar{T}[\frac{\text{J}}{\text{mol.K}}]$	29.708	45.364	35.455	31.36

Same procedure is done to calculate average specific heat value over condition tower 2. Gas compositions and average specific heat values for each component in stream 2 can be seen in table C-2:

$$\bar{T}_{1-2} = \frac{386 + 224}{2} = 305^{\circ}\text{C} = 578.15\text{K}$$

Table C-2: Gas composition and \bar{c}_p values of each component in condition tower 2.

Component	N ₂	CO ₂	H ₂ O	O ₂
x _i	0.61	0.23	0.09	0.07
\bar{c}_{p_i} at $\bar{T}[\frac{\text{J}}{\text{mol.K}}]$	29.98	46.73	36.02	31.87

$$\bar{c}_{p_{3-4}} = \sum x_i \cdot \bar{c}_{p_i} = 34.51 \frac{\text{J}}{\text{mol.K}}$$

Average specific heat value for air at point 5 at average temperature of 160°C:

$$\bar{T}_5 = \frac{220 + 100}{2} = 160^{\circ}\text{C}$$

$$\bar{c}_{p5} = 1.017 \frac{\text{kJ}}{\text{kg.K}}$$

ΔT Values:

$$\Delta T_{1-2} = 386 - 120 = 266\text{K}$$

$$\Delta T_{1-2} = 386 - 224 = 162\text{K}$$

$$\Delta T_5 = 220 - 100 = 120\text{K}$$

Power Calculation:

$$\begin{aligned}\dot{Q}_{1-2} &= \dot{n}_1 \cdot \bar{c}_{p_{1-2}} \cdot \Delta T_{1-2} = 5900666.5 \left(\frac{\text{mol}}{\text{h}}\right) \times 33.943 \left(\frac{\text{J}}{\text{mol.K}}\right) \times 266(\text{K}) \\ &= 53276.2 \frac{\text{MJ}}{\text{h}} = 14.8 \text{ MW}\end{aligned}$$

$$\begin{aligned}\dot{Q}_{3-4} &= \dot{n}_3 \cdot \bar{c}_{p_{3-4}} \cdot \Delta T_{3-4} = 5900666.5 \left(\frac{\text{mol}}{\text{h}}\right) \times 34.51 \left(\frac{\text{J}}{\text{mol.K}}\right) \times 162(\text{K}) \\ &= 32988.4 \frac{\text{MJ}}{\text{h}} = 9.16 \text{ MW}\end{aligned}$$

$$\begin{aligned}\dot{Q}_5 &= \dot{m}_5 \bar{c}_{p5} \Delta T_5 = 232662.5 \left(\frac{\text{kg}}{\text{h}}\right) \times 1.017 \left(\frac{\text{kJ}}{\text{kg.K}}\right) \times 120(\text{K}) \\ &= 28394.13 \frac{\text{MJ}}{\text{h}} = 7.887 \text{ MW}\end{aligned}$$

NB! It is assumed that the energy carried by dust is negligible in streams 1 and 3.

Appendix C: Calculations

C2: RDF heating value variation vs. moisture content

C2: RDF heating value variation vs. moisture content

Lower heating values for various moisture contents can be obtained via equations C-6 and (C-7)[22]

$$\text{LHV} = \text{HHV}_d(1 - X) - 2.44X \quad (\text{C-6})$$

$$\text{HHV}_d = \frac{\text{HHV}}{1-X} \quad (\text{C-7})$$

LHV: Lower heating value of the fuel, wet basis $\left[\frac{\text{MJ}}{\text{kg}}\right]$

HHV_d: Higher heating value of the fuel, dry basis $\left[\frac{\text{MJ}}{\text{kg}}\right]$

X: Moisture content of the fuel, wet basis

The value of 2.44 is the latent heat of water in MJ/kg at 25°C.

Calculation steps:

Known data: at 0.29wt% moisture content, LHV=12

Substitution in equation C-6 gives:

$$12 = \text{HHV}_d \times (1 - 0.29) - 2.44 \times 0.29$$

$$\text{HHV}_d = 17.9$$

As a result LHV could be plotted vs moisture content following below correlation and results are shown in figure C-1:

$$\text{LHV} = 17.9(1 - X) - 2.44X$$

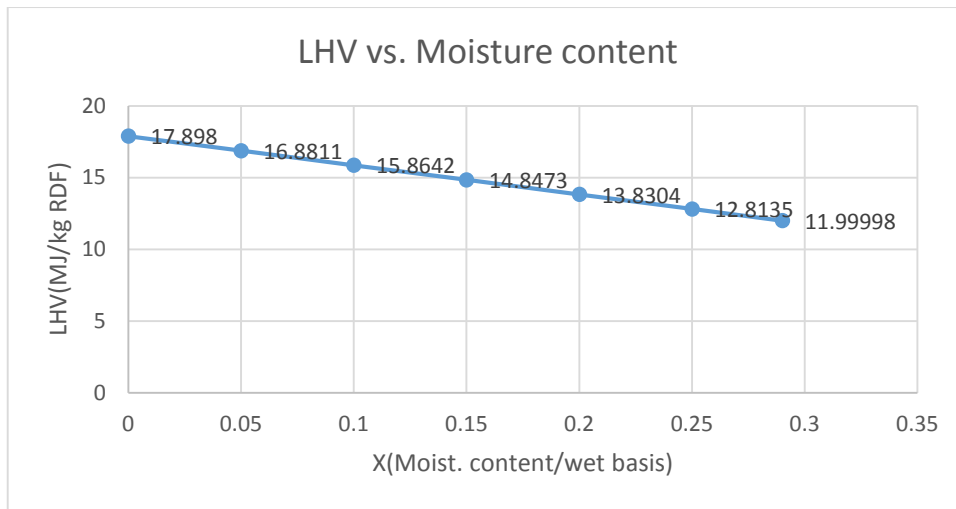


Figure C-1 RDF heating value variation vs. moisture content

Appendix C: Calculations

C3: Adiabatic flame temperature calculation in various RDF moisture content

C3: Adiabatic flame temperature calculation in various RDF moisture content

Combustion temperature could vary with moisture content.

Main formulas to calculate flame temperature:

Flame temperature can be achieved using equation (C-8) which is coming from an energy balance over a combustor [23], as below:

$$T_P \approx T_R + \frac{LHV \cdot N_{fuel} \cdot M_{fuel}}{\sum_i N_{i,P} \bar{C}_{Pi}} \quad (C-8)$$

However, this formula does not take the amount of energy for ash heating into account. In such case, some of the heat of combustion is consumed by the ash component. It consequently reduces the flame temperature. If ash contribution to the flame temperature calculation is considered, one additional term would be added to the equation (C-8) which could come from an energy balance around a combustor. Figure C-2 shows flow streams around a combustor to assist deriving of equation C-9.

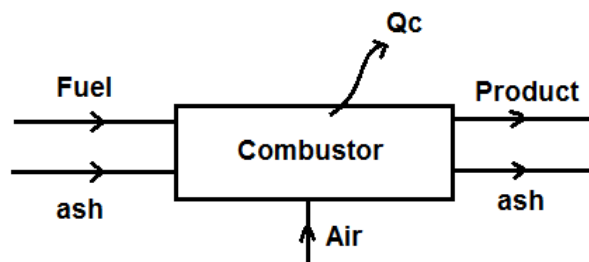


Figure C-2 Flow streams around a combustor

$$\dot{m}_p \bar{C}_{pp} \cdot (T_p - T_R) = LHV \cdot \dot{m}_f - \dot{m}_{ash} \cdot \bar{C}_{pash} \cdot (T_p - T_R) \quad (C-9)$$

Equation C-9 mathematically shows that heat of combustion would heat up combustion product and ash contents in the fuel.

Another form of equation C-9 which has an additional term to equation C-8 could be:

$$T_P \approx T_R + \frac{LHV \cdot m_f}{\sum_i N_{i,P} \bar{C}_{Pi}} + \frac{m_{ash} \cdot \bar{C}_{pash}}{\sum_i N_{i,P} \bar{C}_{Pi}} (T_R - T_p) \quad (C-10)$$

Equation C-11 is the conversion of wet basis property to dry basis.

$$x_d = \frac{x_w}{100 - x_{moist}} \times 100 \quad (C-11)$$

Different notations in above formulas:

T_P : Adiabatic flame temperature [K]

T_R : Reactant temperature (RDF temperature) [K]

LHV: Lower Heating Value of the fuel [J/kg]

N_{fuel} : Number of fuel moles [mole]

M_{fuel} : Molecular weight of the fuel [gr/mol]

$N_{i,P}$: Number of produced moles product [mole]

\hat{C}_{p_i} : Average specific heat capacity of component "i" in the product [J/mol.K]

\dot{m}_p : Product mass flowrate [kg/s]

\overline{C}_{p_p} : Average specific heat capacity of product at average temperature of reactant and product temperatures [J/kg.K]

\dot{m}_f : Mass flowrate of fuel [kg/s]

\dot{m}_{ash} : Mass flowrate of ash [kg/s]

$\overline{C}_{p_{ash}}$: Average specific heat capacity of ash at average temperature of reactant and product temperatures [J/kg.K]

x_d : Dry basis composition, wt% $\left[\frac{\text{kg component}}{\text{kg dry fuel}} \right]$

x_w : Wet basis composition, wt% $\left[\frac{\text{kg component}}{\text{kg wet fuel}} \right]$

x_{moist} : Composition of the moisture, wet basis $\left[\frac{\text{kg water}}{\text{kg wet fuel}} \right]$

In order to calculate the number of moles produced after combustion, the amount of air for burning should be known.

Air amount calculation for combustion:

In order to unify calculations for all moisture contents, dry basis compositions are used as seen in table below. Dry basis composition could be achieved using equation (C-11). Dry basis composition of the fuel components are listed in table C-3.

Table C-3 Dry basis composition of fuel components

Component	Moisture	C	H	O	N	S	Ash	Sum
Unit	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%
Composition(Wet basis)	29.0	28.8	4.1	20.4	0.5	0.2	17.0	100.0
Composition(Dry basis)	40.845	40.56	5.77	28.73	0.704	0.282	23.94	140.8

In order to simplify further calculation, 1kg of dried RDF is taken as the basis. Since air consists of Oxygen and Nitrogen, the amount of this two components has be identified.

As a result, the amount of air to burn 1kg of dried RDF can be obtained based on dry basis compositions as following:

1000gr of dried RDF consists of:

$$405.6 \text{ gr C} \times \frac{1 \text{ mol}}{12 \text{ grC}} = 33.8 \text{ molC}$$

This amount of C would give 33.8mol CO₂. It needs 33.8mol O₂.

$$57.75 \text{ gr H} \times \frac{1 \text{ mol H}_2}{2 \text{ gr}} = 28.9 \text{ mol H}_2$$

This amount of H₂ gives 28.9mol H₂O. So that it needs 28.9mol "O". Consequently, it needs 14.44mol O₂.

Fuel Oxygen:

$$287.32 \text{ gr} \times \frac{1 \text{ mol O}_2}{32 \text{ grO}_2} = 8.98 \approx 9 \text{ mol O}_2$$

Total Oxygen required for burning of 1kg RDF= 33.8 + 14.44 - 9=39mol O₂

It is known that each mole of air includes 79% Nitrogen and 21% Oxygen. As a result, the amount of Nitrogen will be:

$$39 \text{ mol O}_2 \times \frac{79}{21} \left(\frac{\text{mol N}_2}{\text{mol O}_2} \right) = 146.64 \text{ mol N}_2$$

Total air moles required= 146.64+39=185.64 mole air

Mass of air required= (number of air moles)×(air molecular weight)

$$m_{\text{air}} = n_{\text{air}} \times M_{\text{air}} = 185.64(\text{mol}) \times 28.78 \left(\frac{\text{gr}}{\text{mol}} \right) = 5.34 \text{kg air}$$

Note1: Sulphur amount of RDF would change to Sulphur dioxide. However, Oxygen amount for this conversion is assumed to be low enough to neglect its effect on what is calculated so far.

Note2: Ash content in the fuel does not contribute to the air amount for burning.

Flue gas calculation:

The main produced gases from complete combustion of 1kg of dried RDF are CO₂, H₂O, O₂ and N₂. If we assume that fuel burns in stoichiometric condition, the flue gas composition is as following:

CO₂: As calculated already, the amount of CO₂ is 33.8 mole.

H₂O: As already calculated, 28.9 mole H₂O would be produced; however, the moisture of the fuel should be also taken into account.

$$\text{Fuel moisture amount for 1kg of dried RDF: } 408.45 \text{gr} \times \frac{1 \text{mol}}{18 \text{gr}} = 22.7 \text{mol H}_2\text{O}$$

As a result, the total amount of H₂O in the flue gas is: 22.7+28.9= 51.6 mole

N₂: From previous calculations, the amount of N₂ would be 146.64 mole.

O₂: Since stoichiometric amount of fuel and air is assumed, all amount of Oxygen is consumed in the combustion.

Flame temperature calculation:

In order to use equation C-9 or C-10, a try and error procedure should be done to estimate specific heat values for different temperatures, since average C_p value method would be used for this purpose. However, the accuracy of this method highly depends on variation of C_p with temperature. It should be also noted that \hat{C}_{p_i} is assumed to be constant at the average temperature (between reactant and product temperature) and that the average heat capacity of the reactants and products are approximately the same ($m_p \cdot \overline{C}_{p_p} = m_R \cdot \overline{C}_{p_R}$ or $\sum_i N_{i,R} \hat{C}_{p_{i,R}} \approx \sum_i N_{i,P} \hat{C}_{p_{i,P}}$) [23].

Below calculations give the procedure to achieve flame temperature for RDF with 29wt% and results are summarized in table C-6.

Since the contribution of ash content to absorb some combustion heat is going to be considered we need to find out the amount of ash and its average specific heat capacity. Table C-4 gives relevant data for this sort of calculation.

Table C-4 Ash composition and Cp value

Ash component	CaO	Al ₂ O ₃	Fe ₂ O ₃	K ₂ O	Na ₂ O	Sum
Mass fraction (dry basis)	0.25	0.33	0.12	0.13	0.17	1
Mw (gr/mol)	56.08	101.96	159.69	94.2	61.98	-
gr of component kg of dried fuel (Note 1)	59.85	79	28.73	31.12	40.7	239.4
Moles of component in 1kg of dried fuel	1.067	0.775	0.18	0.33	0.66	3
Mole fraction (x_i)	0.35	0.26	0.06	0.11	0.22	1
\hat{C}_{p_i}(J/mol.K) (Note 2)	54.87[24]	89.7[25]	103.7[26]	83.62[27]	69.1[28]	$\sum x_i \cdot \bar{C}_{p_i}$ = 85.03
<p><i>Note 1: As seen from table C-3, there is 239.4gr ash in 1000gr of dried fuel.</i></p> <p><i>Note 2: It is assumed that C_p values for the above solid oxides are almost constant for different temperatures and can be further used in equation C-10. Some of the value are taken at 298K.</i></p>						

First guess for the average temperature between reactant and product temperature is 1500K.

So that, the heat specific values of flue gas component for this temperature can be achieved and results are listed in table C-5.

NB! It should be noted that C_p values for each flue gas component at the wanted temperatures was achieved using available property tables and linear interpolation was applied wherever it was needed [23].

For each temperature guess, C_p values are estimated and they are used in equation C-10 to get a new value for product temperature. If the average temperature between reactant (assumed to be 10°C) and product temperatures differs by 0 to 20K in 2 iterations in row [23], calculation stops.

NB! When using equation C-10 to calculate the product temperature, dry basis LHV has been used since 1kg of dried fuel is taken as the basis for this set of calculation.

Table C-5 T_p calculation for RDF with 29wt% moisture content (wet basis)

component	CO2	H2O	O2	N2	sum	Ash	T_p
Mole number (based on 1kg of dried fuel) (n_i)	33.8	51.6	0	146.64	232.04	3	-
Guess 1: $T_{ave} = 1500K$							
C_{p_i} at T_{ave} (J/mol.K)	58.29	47.1	-	34.8	$\sum n_i C_{p_i}$ = 9113	85.03	2017 K
Guess 2: $T_{ave} = 1150K$							
C_{p_i} at T_{ave} (J/mol.K)	55.78	43.26	-	33.49	$\sum n_i C_{p_i}$ = 9028	85.03	2103
Guess 3: $T_{ave} = 1193K$							
C_{p_i} at T_{ave} (J/mol.K)	56.2	43.87	-	33.71	$\sum n_i C_{p_i}$ = 9106	85.03	2088
Guess 4: $T_{ave} = 1185K$ stop!							

As seen in table C-5, the difference between average temperature of guess 4 and 3 are less than 20K. As a result, calculation stops and 2088K is taken as the final product temperature.

Same procedure could be done for other values of RDF moisture content. Number of moles for flue gas components remain constant except for H₂O which is needed to be renewed for different moisture content, just like what was done in flue gas calculation part for H₂O.

Results for flame temperature of RDF in various moisture contents are presented in table C-6.

Table C-6 Flame temperature of RDF in various moisture content

Moisture content (wet basis)	Flame temperature (K)
0.29	2088
0.25	2144
0.2	2196
0.15	2250
0.1	2292