

CZECH TECHNICAL UNIVERSITY IN PRAGUE

FACULTY OF MECHANICAL ENGINEERING

DEPARTMENT OF PROCESS ENGINEERING

**THERMAL-EXPANSIONARY PRETRERATMENT OF
LIGNOCELLULOSE WASTES IN INDUSTRIAL SCALE**

DIPLOMA THESIS

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Annotation - Czech: Projektování zařízení na výrobu bioplynu s a bez předupravou. Termicko-expanzní předuprava. Schéma s výstavby celé technologie, se začíná z odpadů a končí s výrobou elektrické energie a její výpočtu hmoty, energie a ekonomického analýzu pro dvě zařízení na výrobu bioplynu. Diskuse o proveditelnosti tvrzení pohledu, a výsledek závěrečné zkoušky.

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substrate and ending with electricity production, and its calculation of mass, energy and economic analysis for two biogas plants. Discussion about feasibility point of view, and final result examination.

Keywords: Biogas, wheat straw, pretreatment.

Utilization: Designing of technology. Project works in biorefineries.

Declaration

I confirm that the diploma work was disposed by myself and independently, under leading of my thesis supervisor. I stated all sources of the documents and literature.

In Prague

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Name and Surname

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I dedicate this work to my mother, Vera, and grandmother, Nadezhda. They are my everything.

I would like to thank my supervisor, Ing. Lukáš Krátký, Ph.D., for willingness to help me at any time, and for all these hours of dedicating time to me.

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Bc. Andrey Kutsay

“The nation behaves well if it treats the natural resources as assets which it must turn over to the next generation increased and not impaired in value. Conservation means development as much as it does protection.”

Theodore Roosevelt (1858–1919)

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1. INTRODUCTION

The master thesis was initiated according to the needs of energy production and heat production. The energy carrier to take, in this thesis, was biogas, which is being accounted as the alternative to fossil fuel. The feedstock used to produce biogas is wheat straw (in this work only), which is agricultural waste that is burned in fields without any energy recovery. Before anaerobic digestion, the straw could be treated to enhance the production, especially of methane, which is taking crucial point in biogas plant.

It is not the news, all of us know that shortage of nowadays energy sources will appear in coming 50 years. The current research and future predictions say, the crude oil will use up within 40 to 70 years, and natural gas would be exhausted within the 50 years [1]. Also, the global average temperature would be increased by 1.4 to 5.8 °C by year 2100 and will only continue to grow [2]. The reason of the following temperature change is greenhouse effect, the emitted CO₂ is staying in the atmosphere and it somehow absorb the radiation from the Sun.

At this point it is clear, that in the closest future we need to find new solutions of energy demands for human needs. Nevertheless, these solutions should be effective, differ in pollution rates and be covered by national standards. There are several methods of processing substrate for methane production, but, unfortunately, none of them is effective from the biodegradability point of view. This paper would be dedicated only to special pretreatment of a substrate (wheat straw), which will increase biodegradability. Nevertheless, the following pretreatment could be applied for other types of feedstock.

1.1 Small History

Roman scholar Pliny, first noticed around 50 BC some glimmering lights appearing underneath the layers of swamps, he later called them flammable air.

Later, in the 17th century, Jan Baptita Van Helmout said that flammable gasses can appear after putrescent organic matters. Alessandro Volta, in 1776, collected gas from the lake Como to test it. He proved that formation of the strongly depends on fermentation process, as well he said that gas and air can create an explosive mixture [3].

The final chemical structure of CH₄ was explained by Avogadro in 1821.

Later, in the second half of the 19th century the France starts investigations, researches to implement the biogas in human life. So, for example, Louis Pasteur tried in 1884 to produce biogas from horse dung, which he collected from Paris roads. After, he and his student Gavon, managed to obtain 100l of methane from 1m³ of dung fermented at 35 °C. As the sum up of it, they assumed that this production rate should enough to insure the energy requirements for street lighting in Paris [3].

1.2 Attempts at using Biogas

Starting from 1897, the first lamps in Exeter, England, were driven by gas obtained from anaerobic digestion treatment of wastewater. It was the first time when biogas was not totally wasted to the atmosphere, during wastewater treatment, but even partly used [3].

Not so many time after, the technician Imhoff has constructed anaerobic wastewater treatment units in the Ruhr, Germany. The “Imhoff tank” (Figure 1) was installed into existing treatment plants to separate the sludge. The residence for such a unit was around 60 days [3].

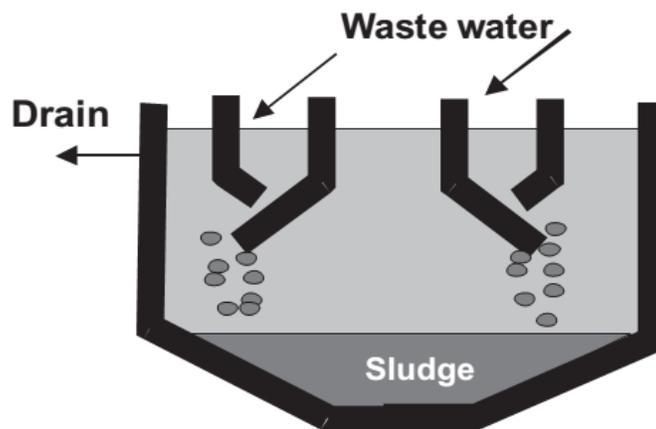


Figure 1. Imhoff tank – a sedimentation tank for mechanical sewage treatment [3].

In 1955, the importance of biogas was significantly decreased, because biogas no longer profitable due to excess of oil [3].

Around 1970, the necessity of biogas grew up, due to oil crisis.

In 2000, the Law of Renewable Energies became effective, which stated the rules for the subsidization of the power supplied by biogas facilities in Germany. The result of this was rapid growing of biogas plant, especially after implementing even higher subsidies. Approximately 1500 biogas facilities were built in Germany. The average amount of electric power per one plant raised from around 300 to 500 kW. The largest one was built with more than 10 MW of electric power [3].

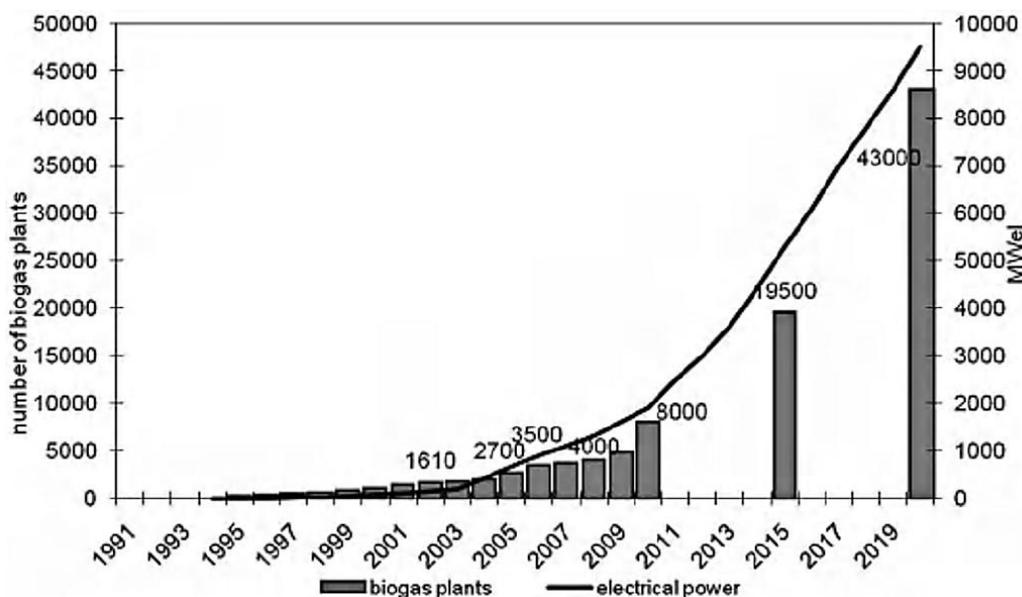


Figure 2. Expansion of biogas production in Europe [3].

Other European countries are booming in production of biogas. Such as, in Germany there already 8000 plants counted in 2014, in Czech Republic 500 plants were built with generated power 400 MW [32], in Austria 300 plants are already working.

The typical biogas plant is described on the Fig. 3. After the manure is collected in some tank, it is preferably to be pretreated. The biogas is taken from digester and burned in some gas engine, where heat and electricity generated. The residuals of fermentation process could be used as fertilizer for growing new plants.

Biogas could be produced from different kinds of substrate: straw, maize, sorghum etc. However, the following work utilize wheat straw only, as it is reach of carbon content in cellulose and hemicellulose.

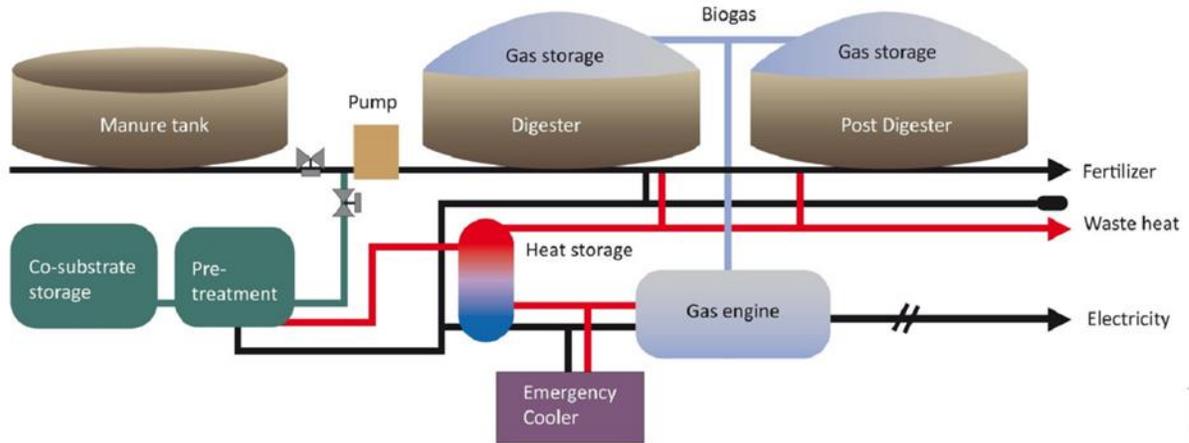


Figure 3. Biogas plant, working principle [24].

1.3 Biology – raw material composition

Lignocellulosic biomass, like wood, grass, wheat straw etc., is generally composed of cellulose, hemicellulose, lignin (Figure 4) and other different organic and inorganic compounds. The long-chain carbohydrates present in insoluble structure such as cellulose, hemicellulose, and starch cannot be simply broken into short chain-chain, because of consisting inside lignin [3]. The lignin serves as the glue between chains, and that is why some pretreatment should exist in any bio-fuel plant. Without pretreatment the hydrolysis would not be effective, means that the most potential part of a substrate would not be released for further fermentation.

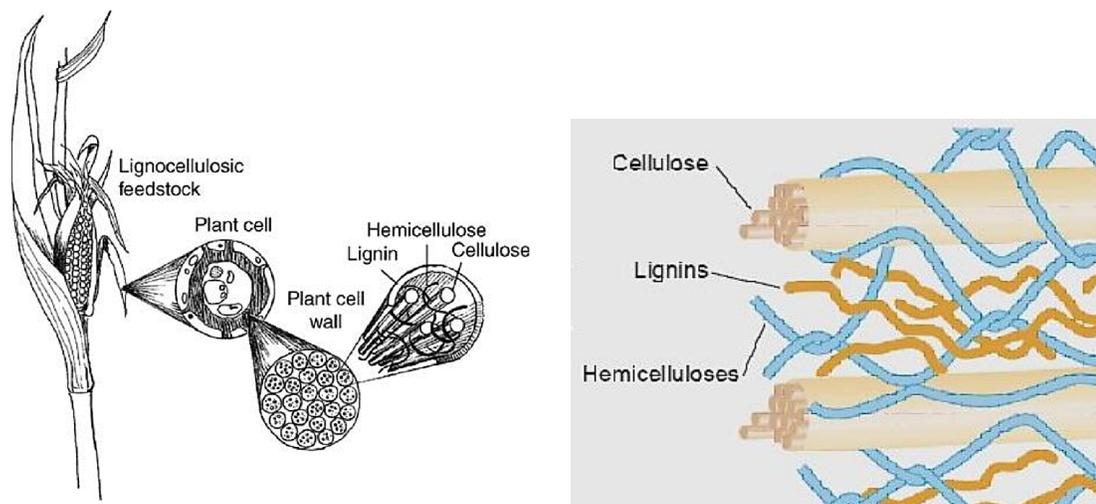


Figure 4. Lignocellulosic biomass [26].

The composition of some lignocellulosic biomasses could be found on the next table.

	Residues generated [t. 10 ⁶ . year ⁻¹]	Composition [wt.%]				Biogas yield [m ³ . kg ⁻¹ DM]
		<i>Cellulose</i>	<i>Hemicellulose</i>	<i>Lignin</i>	<i>Ash</i>	
Sugar cane bagasse	317-380	32-44	27-32	19-24	4.5-9	0.4-1.0
Wheat straw	154-185	29-35	26-32	16-21	12-29	0.5-0.7
Barley straw	35-42	31-34	24-29	14-15	5-7	0.75-0.99
Rye straw	4.3-5.2	33-35	27-30	16-19	2-5	0.57-0.79

Table 1. Some lignocellulosic materials in numbers [3, 33].

2. CRITICAL LITERATURE SEARCH

2.1 Anaerobic digestion

Anaerobic digestion (AD) is a microbiological process, therefore organic matter is processed in the absence of oxygen. This process is common to many natural environments such as swamps or the stomach of ruminants. Using an engineering approach and controlled design, the AD process is applied to process organic biodegradable matters in airproof reactor tanks, commonly named digesters, to produce biogas. Various groups of microorganisms are involved in the anaerobic degradation process which generates two main products: energy-rich biogas and a nutritious digestate.

Anaerobic digestion of organic waste provides many benefits. This includes the generation of renewable energy, a reduction of greenhouse gases, less consumption of fossil fuel, new job positions, and completion of nutrient cycle. It transforms organic waste material into valuable resources while at the same time reducing solid waste volumes and thus waste disposal cost. Biogas as a renewable energy source not only improves the energy balance of a country but also contributes to the preservation of natural resources by reducing deforestation, and to environmental protection by reducing pollution from waste and use of fossil fuels [4].

The calorific value of biogas is around $6.0 - 6.5 \text{ kWh.Nm}^{-1}$, depending on the percentage of methane, which on average is in the range of $55 - 70 \%$ vol. [3]. The net calorific value depends on the efficiency of the biogas burners or other appliances used to process the biogas. A gas generator, for example, can convert about 2 kWh into usable electricity, while the remaining energy is emitted as heat.

2.1.1 Process chain of anaerobic digestion of biowaste

The following figure based on a systematic supply chain perspective [5],

- **Substrate chain:** Waste generation, collection, transportation and supply to the digestion facility and necessary pretreatment of waste before feeding the digester.

- **Transformation process:** Biological and chemical transformation processes of feedstock in the digester which leads to value products.
- **Product chain:** Post-treatment of outflows from the digester that refine these into improved value products, and their distribution and utilization.

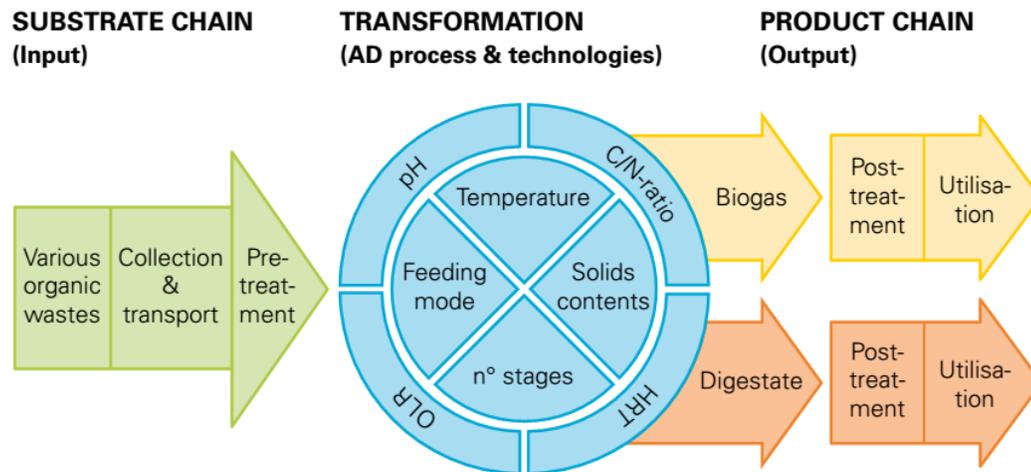


Figure 5. Process chain of anaerobic digestion [5].

2.2 Biochemical process of anaerobic digestion

The anaerobic decomposition of organic matter occurs in four – step process.

- **Hydrolysis**

Hydrolysis is usually the slowest of the four degradation steps. The bacteria transform complex organic materials into liquefied monomers and polymers, for example, proteins, carbohydrates and lipids (fats) are transformed to amino acids, monosaccharides and fatty acids respectively [6]. This extracellular enzyme-mediated transformation of higher mass organic molecules into basic structural building blocks is very important as particulate organic materials are simply too large to be directly absorbed and used by microorganisms as substrate.

The hydrolysis of carbohydrates takes place within a few hours and the hydrolysis of proteins and lipids within a few days. Lignocellulose and lignin are degraded only slowly and incompletely [3].

- **Acidogenesis**

The monomers formed in the hydrolytic phase are taken up by different facultative and obligatorily anaerobic bacteria and are degraded in the second, acidogenic, phase to short-chain organic acids, C₁-C₅ molecules, alcohols, nitrogen oxide, hydrogen sulfide, hydrogen and carbon dioxide. The concentration of the intermediately formed hydrogen ions affects the kind of fermentation products. The higher the partial pressure of the hydrogen, the fewer reduced compounds, such as acetate, are produced. The degradation of amino acid also leads to production of ammonia [3].

- **Acetogenesis**

At this point long chain fatty acids and volatile fatty acids and alcohols are transformed by acetogenic bacteria into hydrogen, carbon dioxide and acetic acid. During this reaction the BOD and the COD are both reduced and the pH decreased [7]. Hydrogen plays an important intermediary role in this process, as the reaction will only occur if the partial pressure is low enough to thermodynamically allow the conversion of all the acids. Hydrogen scavenging bacteria lead to a lower partial pressure. Thus the hydrogen concentration in a digester is an indicator of its “health” [8].

- **Methanogenesis**

In the last stage, methanogenic bacteria convert the hydrogen and acetic acid to methane gas and carbon dioxide. Methanogenesis is affected by conditions in the reactor such as temperature, feed composition and organic loading rate [9].

The gaseous product, biogas, consists mainly of methane and carbon dioxide, but also contains several other gaseous “contaminants” such as hydrogen sulphide (easily detected by its smell of rotten eggs), nitrogen, oxygen and hydrogen. Biogas with methane content higher than 45 % is flammable; the higher the CH₄ content the higher the energy value of the gas [3].

Components	Symbol	Concentration [vol. %]
Methane	CH ₄	55 – 70
Carbon dioxide	CO ₂	35 – 40
Water	H ₂ O	2 (20°C) – 7 (40°C)
Hydrogen sulphide	H ₂ S	20 – 20000 ppm (2%)
Nitrogen	N ₂	<2
Oxygen	O ₂	<2
Hydrogen	H ₂	<1
Ammonia	NH ₃	<0.05

Table 2. Typical composition of biogas from biowaste [11].

On the following Fig. 6, you could see 4 steps in motion.

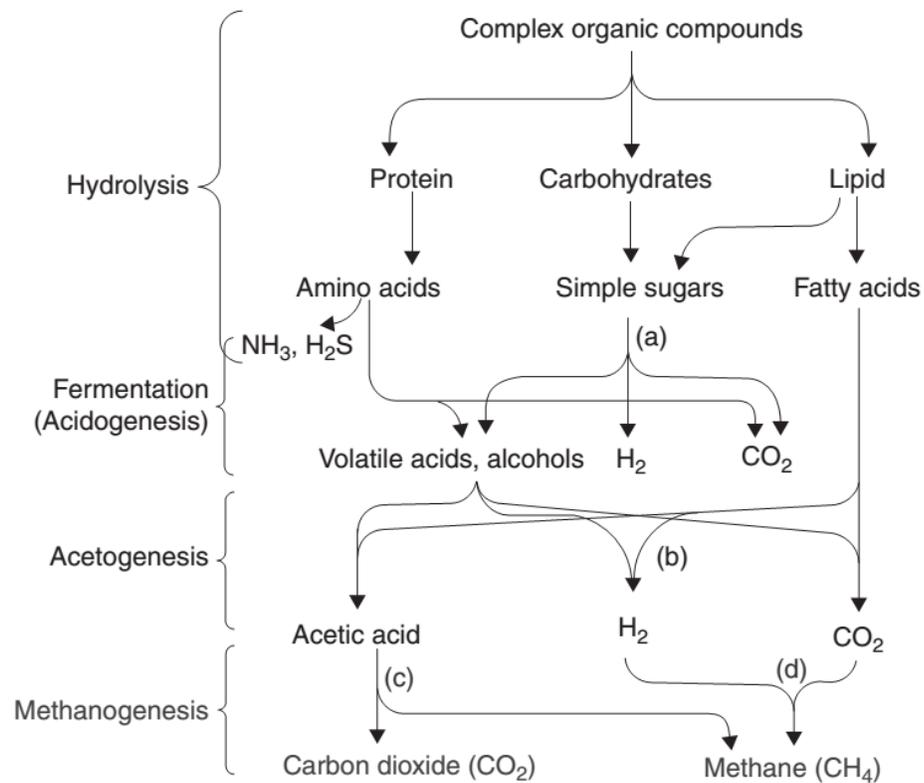


Figure 6. Scheme of biodegradation steps [25].

2.3 Process parameters

The grow rate of microorganisms is at high importance for the anaerobic digestion. Even temperature change or changes in substrate or its concentration can lead to shutdown of the biogas production. It could last up to 3 weeks or sometimes more, when the new ecological system will adapt to a new conditions and the biogas production would be started again. However, the human factor could prolong the time for 3-4 weeks more.

The environmental requirements for fermentative bacteria, with which the first and second (hydrolysis and acidification) steps occur, differ from requirements of the other steps. Best environmental conditions for all microorganisms, which involved into degradation that can only be set in two-stage plant with one-stage for hydrolysis/acidogenesis and one-stage for acetogenesis/methanogenesis. It said, that the complete degradation process has to take place in the same reaction system (one-stage process), the environmental requirements of methanogenesis must be fulfilled with priority, because these would otherwise have no chance of survival within the mixed culture, because of their lower growth rate and higher sensitivity to environmental factors. However, for lignocellulosic biomasses are in higher priority with hydrolysis limits [3].

2.3.1 Temperature

The anaerobic degestion is principally feasible under almost all climatic conditions, at low temperature (below 15 °C) the digestion process does not work properly [10]. In cool climatic conditions, either a heating system has to be installed or a larger digester has to be built in order to increase retention time [11]. Heating system and good insulation can provide proper digestion temperature even in cold climates or season, however the required additional invests makes it economically unviable. Among temperatures like, mean temperature, also large temperature variations are important, such as temperatures between day and night, or seasonal variations. Digesters built underground provide minimization of temperature changes by buffer capacity of the soil. For household digesters the design should ensure the gas production even in unfavorable season of the year. Given the additional investments required in colder climates, a general rule is that the cost of biogas production increase as the temperature decrease.

Mainly, there are two ideal temperature ranges for the performance of anaerobic bacteria. The one ranges from 30 °C to 40 °C, it is mesophilic microorganisms, and one ranges from 45 °C to 60 °C, it is thermophilic microorganisms [3]. Operation of a digester in the mesophilic range is more stable, as these microbial communities can tolerate greater changes in environmental parameters and consume less energy. Interruption by ammonium is less critical in the mesophilic range as compared to thermophilic conditions, because of lower content of free ammonia at lower temperatures. On the other hand, the mesophilic microorganisms are slower and thus a longer fermentation time in digester is needed to reach the highest yield of biogas production [5].

On the Fig. 7, you could see fermentation time for different microorganisms.

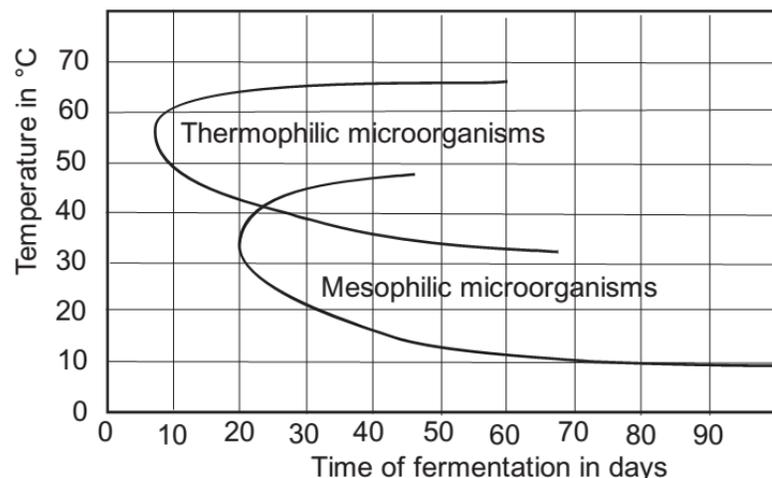


Figure 7. Influence of the temperature on fermentation time [3].

2.3.2 Specific surface area of material

To keep the biochemical reaction, the surface area of material should better be kept as large as possible. In order to increase the surface area of material, milling or comminution of the substrate is preferred in many cases before processing [12].

2.3.3 pH value

The situation with regard to pH value is similar to that for temperature. Those microorganisms, which involved in the different stages of decomposition, require different pH value. The pH optimum for hydrolysis and acid-forming bacteria is in a range from 5.2 to

6.3. They are not fully reliable on this, however, and are still capable of converting substrates at a slightly higher pH value. The only problem is that their activity is slightly reduced. In contrast, the pH value in the neutral range from 6.5 to 8 is absolutely essential for acetogenesis and methanogenesis. Thus, if the fermentation process occur in single digester, this pH range must be maintained [13].

Regardless of whether process is single-stage or multi-stage, the pH value is established automatically in the system with help of alkaline and acid metabolic products formed in the course of anaerobic digestion [13].

If too much organic matter is fed into the process within too short period of time, for example, or if methanogenesis is inhibited for some other reason, the acid metabolic products of acidogenesis will accumulate. Normally the pH value is established in the neutral range (6.5-8) by the carbonate and ammonia buffer. If the system's buffer capacity is exhausted, for example, if too many organic acid have built up, the pH value decreases. This, in the same way, increase the inhibitory effect of hydrogen sulphide and propionic acid, to extent that process in the digester comes to halt within a very short space of time. With regard to process control, it must be known that because of its inertia although the pH value is of only limited use for controlling the plant, in view of its great importance it should always be measured [13].

2.3.4 Carbon to nitrogen ratio

The difference between carbon and nitrogen amount in organic materials is represented by C:N ratio. The C:N ration is an important parameter in estimating nutrient deficiency and ammonia inhibition [5]. Optimal C:N ratio in anaerobic digesters are between 16 and 25 [3]. High C:N ration is the result of rapid consumption of nitrogen by methanogens, which then results in reduced gas production. On the other side, a low C:N ratio causes ammonia accumulation and pH value can grow more than 8.5. Such conditions can be toxic to methanogenic bacteria. Although methanogenic bacteria can adapt to high ammonia concentration this only occur if concentrations are increased gradually allowing time for adaptation. Proper C:N ratios can be ensured by mixing various feedstock materials, with high

(organic solid wastes) and low (sewage or animal manure) C:N ratios to obtain a perfect ratio level [5].

2.3.5 Organic loading rate and fermentation time of the digester

Whenever a biogas plant is being designed and constructed, high attention is normally paid to economic considerations. Consequently, when the size of digester is being chosen the focus is not necessarily on highest gas yield or on full decomposition of the organic matter contained in the substrate. If it was said to reach full decomposition of the organic material, sometimes quite long fermentation time would be necessary for the substrate in digester, together with correspondingly large tank volumes, because some substances take very long time to break down – if at all. Thus, to find optimum degradation performance at acceptable economic cost is in privilege [13].

At this point, the organic loading rate is a crucial operating parameter. It shows how many kilograms of volatile solids (VS, or organic dry matter – ODM) can be fed into digester per every cubic meter of working volume per unit time. The organic loading rate is expressed in the units $\text{kg}_{\text{vs}}\cdot\text{m}^{-3}\cdot\text{d}^{-1}$ [13]. The overloading leads to significant rise in volatile fatty acids, which can result in acidification and system failure. In industrialized countries the OLR is in the range $4\text{-}8 \text{ kg}_{\text{vs}}\cdot\text{m}^{-3}\cdot\text{d}^{-1}$, which result in VS removal in the range of 50-70 % [14]. This is ideal for CSTR. However, for non-stirred AD systems which are dominant in developing countries, an OLR below $2 \text{ kg}_{\text{vs}}\cdot\text{m}^{-3}\cdot\text{d}^{-1}$ is preferred and considered suitable.

In this work we will assume that organic loading is in the range $1\text{-}3 \text{ kg}_{\text{vs}}\cdot\text{m}^{-3}\cdot\text{d}^{-1}$, because in reality the mixing within the digester occurs only once in 5-10 minutes [3]. The result of not continuous mixing will lead to the smaller OLR, as the substrate will spread less uniformly, for every cubic meter, inside the digester.

2.4 Intensification process pretreatments

The technical expertise required to maintain industrial scale of anaerobic digesters connected with high investment cost and low process efficiencies had limited the level of its industrial application as a waste treatment technology. However, the anaerobic digestion technology was recognized as one of the most decentralized source of energy supply. There are many

benefits associated anaerobic digestion, which imply mass reduction, pathogen reduction, odor removal, less energy use etc. [15].

Anaerobic digesters originally designed for processing of sewage sludge and animal waste (manure). However, both of them are not well biodegradable substrates with the most potential for anaerobic digestion, there plenty of other different substrates for processing. The substrate composition is a major factor in defining the methane yield and methane production rates, the better biodegradable substrate the higher the gas yields it possible to obtain [15].

Each substrate has own yield and kinetics possibilities of methane of methane production. Pretreatments by mechanical size reduction, heat treatment and/or chemical treatment could improve its digestibility. The aim of pretreatments stage is to modify the structure of complex materials (usually cellulose) with decreasing degree of polymerization, the weakening of the bonds of lignin with carbohydrates and increased surface area of particles constitute. Lignocellulosic substrates are considered as the most abundant renewable resource, the major biodegradable fraction in municipal solid waste, with the potential of making substantial difference in the supply of biofuel. The biodegradability of cellulose depends on their amorphous or crystalline form and the range lignin content [15].

The pretreatment methods of substrates objective at improving anaerobic digestion processes by altering physical or chemical properties and shows the next advantages: increased final handling cost; a trend towards lower nitrogen limits, which is driving up sludge ages and decreasing degradability of activated substrates streams and increased legislative requirements for stabilization performance and pathogen removal. Therefore, there is an increased need of overviewing and analyzing the different pretreatment options in terms of mechanism, cost, and performance [15].

2.4.1 Physical pretreatment

The application of physical pretreatment can increase the rate of hydrolysis and improve the anaerobic biodegradability of substrate, especially of municipal solid waste.

2.4.1.1 Mechanical pretreatment

The idea of mechanical pretreatment is to break the cell walls of a substrate to make them more accessible to microorganisms, or in another words to increase the specific surface area. The reduction of a size increase the accessibility of microorganisms to the residue particles due to breaking of large structures in shorter chains, which improve the speed and efficiency of hydrolysis. The methods used for mechanical pretreatment: mechanical jet, high-pressure homogenizer, mechanical ball mill, etc. Although, various type of mechanical pretreatment showed a significant impact on biogas production improvement, the main challenge of using mechanical pretreatment is high-energy requirements [15].

Based on Kratky and Jirout [23], the mechanical disintegration enhance the biofuel yield up to 25% and also shortens the digestion time by 23-59 %.

2.4.1.2 Thermal pretreatment

The thermal pretreatment is the step that enhances the effectiveness of anaerobic process, due to the fact that thermal hydrolysis leads to partial solubilisation of substrates, the organic and inorganic compounds are efficiently solubilized during thermal pretreatment, which reduce digester mixer volume and improve biogas production [15].

Bougrier and co-workers [18] suggest the thermal treatments classification according the impact on biogas production: 1) treatments at temperatures from 70 °C to 121 °C which led to 20-30 % of biogas production increase, 2) treatments from 160 °C to 180 °C led to a 40-100 % biogas production increase. As the sum up, the 160-180 °C pretreatments are therefore most efficient to improve sludge anaerobic digestion but these evaluations are not favorable to increase the biogas production (40-100 %).

2.4.1.3 Ultrasound – assisted pretreatments

The idea of high intensity ultrasound to a liquid system may lead to physical and chemical reactions that can significantly upgrade the structure presented in liquid. Sound propagates in the form of waves (sound waves) with a given height (intensity) and distance between waves (frequency). Ultrasound is the notation used to any wave higher than 20 kHz, which is higher frequency that human ear can hear. Nevertheless, quite high frequency cause a phenomenon

known as cavitation. When ultrasonic waves are passing through a medium, they generate gas bubbles, which are subsequently compressed and expanded by the passing sound waves until achieving the critical point where they eventually explode producing local extreme conditions of temperature and pressure [15].

Ditl and co-workers [27], experimentally proved that there is an increase of biogas yield approximately 20 % in comparing to the untreated version [27].

2.4.2 Chemical pretreatment

Pretreatment of agriculture residuals by size reduction (physical pretreatment), heat treatment and/or chemical pretreatment usually enhance its digestibility. Chemical pretreatment methods include alkali treatments; organosolv process; wet oxidation (treatment with water and air or at temperature above 120 °C for a period of e.g. 30 min); ozonolysis pretreatment (treatment with ozone); acid hydrolysis pretreatment (sulfuric acid is the most applied acid). In the case of anaerobic pretreatment of the substrates, the methods of alkali are the mostly used for increase at performer in terms of solubilization of COD and biogas production [15].

As the example, the procedure of alkaline treatment of substrate in anaerobic digestion consists of adding the portion of NaOH on the substrate at room temperature (25 °C) with contact of 24 hours, finally the samples are filtered through a nylon sieve of 20-micron size or centrifugal obtaining a liquid fraction (supernatant) and solid (precipitate). The alkaline environment causes swelling of the organic matters, making them more acceptable to enzymatic attack by enhancing the biodegradability of the solid phase in contrast to acid hydrolysis, characterized first for a mechanism of solubilization of organic material. Therefore, the additional buffer capacity, resulting from the addition of alkali, could help the neutralization of organic acids (VFA) during the digestion stage, smoothing the inhibitory effect [15].

There were several investigations on treatment by alkali. The increase of yield within the anaerobic digestion could rise up to 27.3-64.5 % for rice straw by using 6 % NaOH for 3 weeks at ambient temperature [26].

2.4.3 Biological pretreatment

The idea of biological pretreatment is to prepare the substrate for the enzymatic degradation and the best way and conditions of pretreatment depends mainly on the type of substrate. Bacteria can be used for biological pretreatment of substrates in anaerobic digestion. Kurakake et al. [19] studied the biological pretreatment of office paper with bacterial strains (*Sphingomonas paucimobilis* and *Bacillus circulans*) and obtained enhance of the enzymatic hydrolysis; the sugar recovery was improved up to 94 % for office paper only.

Low energy requirement, no chemical requirement, and mild environmental conditions are the main advantages of biological pretreatment. Nevertheless, the efficiency is very low in mos biological pretreatments. If the following pretreatment is used it is usually applied after physical or chemical pretreatments [15].

It was achieved an additional biogas yield of 21 % at a hydraulic retention time of approximately 30 days. This caused by higher degradation through increase hydrolytic enzyme activity [26].

2.4.4 Combined pretreatment

Pretreatment by physical, chemical or biological means is well investigated in the anaerobic digestion technology. There are many ways in pretreatment of waste materials, both to improve biodegradability and increase the biogas production. According to Taherzadeh and Karimi [20], an effective pretreatment should meet the next requirements: (1) avoiding formation of possible inhibitors for hydrolytic enzymes and fermenting microorganisms, (2) minimizing the energy demand, (3) decreasing the cost of material for construction of pretreatment reactors, (4) producing less residuals and (4) consumption of small or no chemical and using a cheap chemicals. When the pretreatment by physical, chemical or biological does not have efficient results, combined pretreatment can be applied: physical – chemical, chemical – biological, etc.

2.5 Comparison of substrate pretreatments

The main effects that pretreatments have on various type of substrate could be divided into (1) particle-size, (2) solubilisation, (3) biodegradability improvement, (4) formation of refractory compounds, (5) loss of organic material.

According to Carlsson and co-workers [21], the amount of substrates used for processing within the production of different types of biofuel will look like Fig. 8.

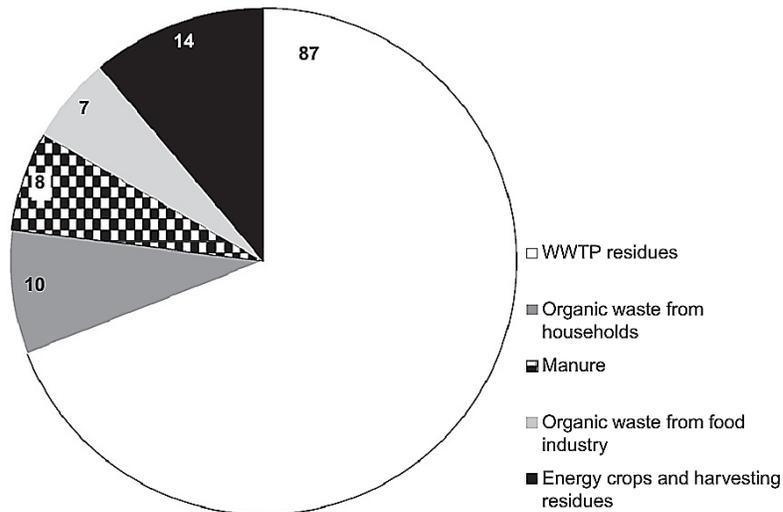


Figure 8. Substrate circle [21].

The most abundant part is of course those from WW, because the water used in enormous amounts on our planet. The 2nd one is agricultural products and residuals, and the 3rd is organic wastes.

Carlsson and co-workers [21] made a lot of researches about, which exactly the best pretreatment method to process. It is obvious that every other type of substrate going to have different data for treatment. Nevertheless, the Fig. 9 was obtained during Carlsson and co-workers work.

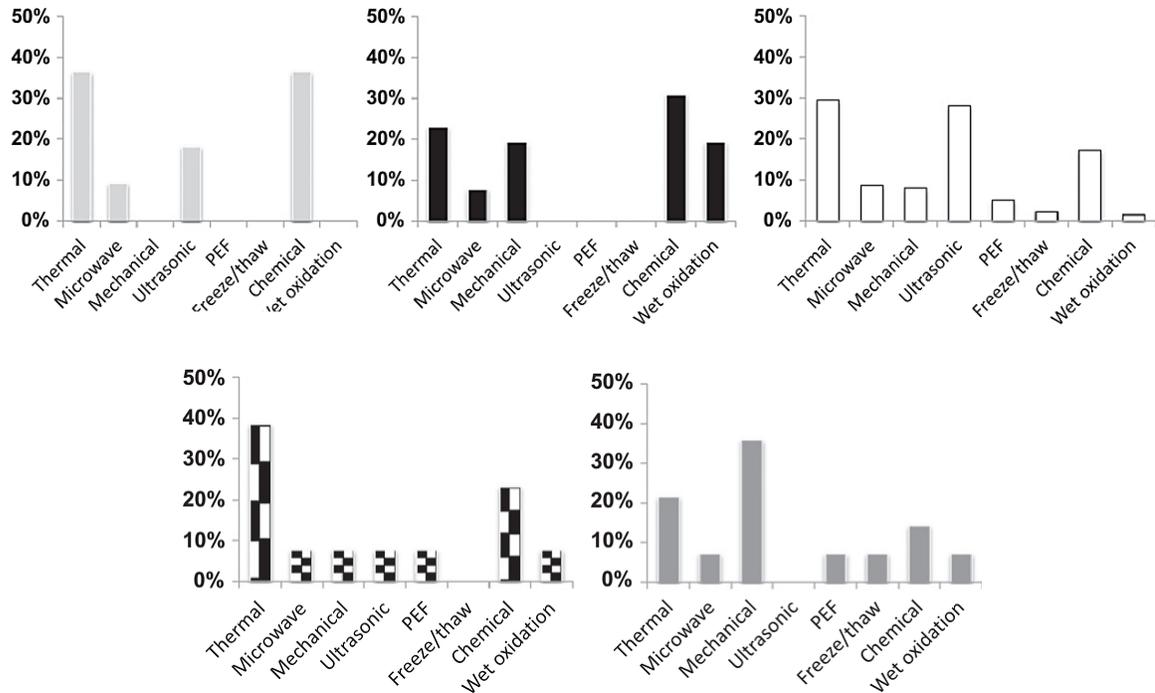


Figure 9. Different pretreatments on different substrates. Vertical – the preferred treatment for the plant; Horizontal – types of treatments [21].

From the Fig. 9 could be seen that mainly physical (thermal, mechanical, ultrasonic) and chemical pretreatments are the most effective. Taking into account these two methods, we can say that the physical one is more effective. Additionally, the chemical pretreatment needs extra step and extra invests for eliminating of chemicals after treatment.

Usually, combined pretreatments were done it the following combination physical-chemical or chemical-biological. Previously we spoke about physical pretreatment that it is most effective from the rest, but still for individual substrates it could be not the best.

Thermal – expansionary pretreatment is the combination of physical pretreatments. So, thermal – expansionary pretreatment is combination of thermal, mechanical and “explosion” pretreatments.

2.6 Thermal – expansionary pretreatment

Thermal – expansionary pretreatment is modern technology that can be used with different types of lignocellulosic biomasses. During this pretreatment lignocellulosic biomass is held

in the reactor, where high temperature steam and pressure is applied for a few minutes. When residence time is over, the batch is quickly depressurized. Liquid, which is sucked into biomass pores, changes its phase from liquid to vapor and the associated volumetric change causes a sudden disruption of the lignocellulosic bonds. The effectiveness of a steam explosion/ expansion particularly depends on the particle size, its humidity and composition, on steam temperature and on residence time. The equivalent hemicellulose removal can be obtained at a high steam temperature with a short residence time (270 °C, 1 min) or at low temperature with a long residence time (190 °C, 10 min) [16]. Nevertheless, it was found that the treatment at low temperature with a long residence time is more preferable [17].

Thermal – expansionary pretreatment is a novel trend in processing of lignocellulosic biomass in industrial biofuel technology. This double-stage process combines the boiling of an aqueous biomass suspension within the high pressure followed by rapid decompression. Until now, there were no information about the effectiveness of the pretreatment. This work was dedicated to estimate the effectiveness of usage of the following pretreatment on the plant-scale.

2.6.1 Lab – scale

Thermal – expansionary pretreatment was used in batch mode , see the next figure. The experiment was done by Kratky and Jirout [16]. Briefly, the hydrolyser is batch double-jacketed pressure vessel, which could be filled maximum by 8 l. Wheat straw is heated by oil (O) in double-jacketed vessel. The expansion vessel (2) is used to keep the expanded substrate. The dump valve (V2) is used to discharge the hydrolysate (expanded substrate). Expansion vessel is equipped with water (W) cooling system. The ball valve (V1) keeps the pressure in hydrolyser, which is gained during heating. At first, the hydrolyser is filled with the mixture of substrate and water (95 wt. %). The mixture is heated by oil up to some desired temperature, as the hydrolyser is closed, the pressure is increased. The ball valve opens and the mixture is expanded in expansion vessel due to different pressure difference in both vessels. The products of sudden expansion is vapor and hydrolysate.

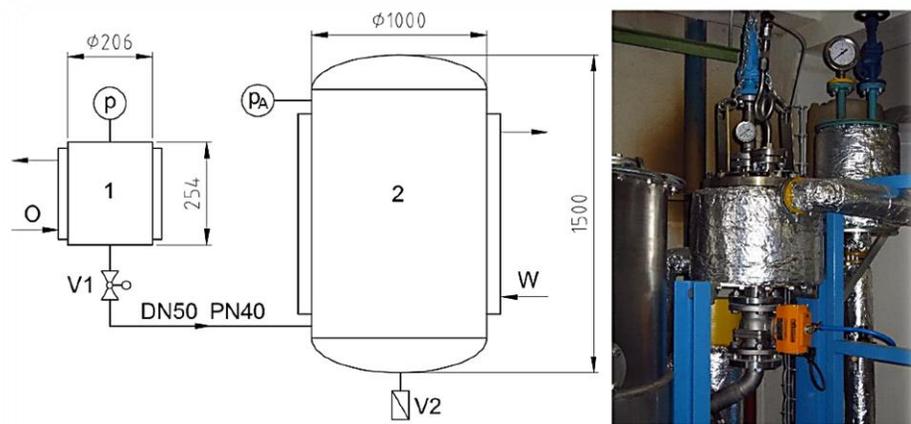


Figure 10. Laboratory unit for the thermal-expansive treatment. 1-hydrolyser, 2-expansion vessel, V1-ball valve with pneumatic actuator, V2-dump valve, W-water cooling [16].

As we heat the mixture up to some desired temperature and pressure, we process it for some residence time after the depressurizing the temperature is rapidly decreased. The dependence of P, T parameters to time could be seen on the Fig.11.

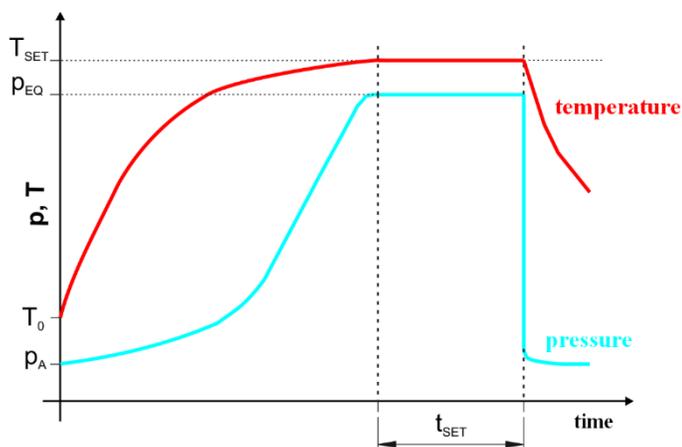


Figure 11. P, T parameters [16].

2.6.2 Lab – results

Kratky and Jirout [16] found that, the best yield in according to degradation of raw material, was reached at 185 °C and 40 minutes of residence time. On the following figure you could see the results of the experiment, which are very important for further calculations.

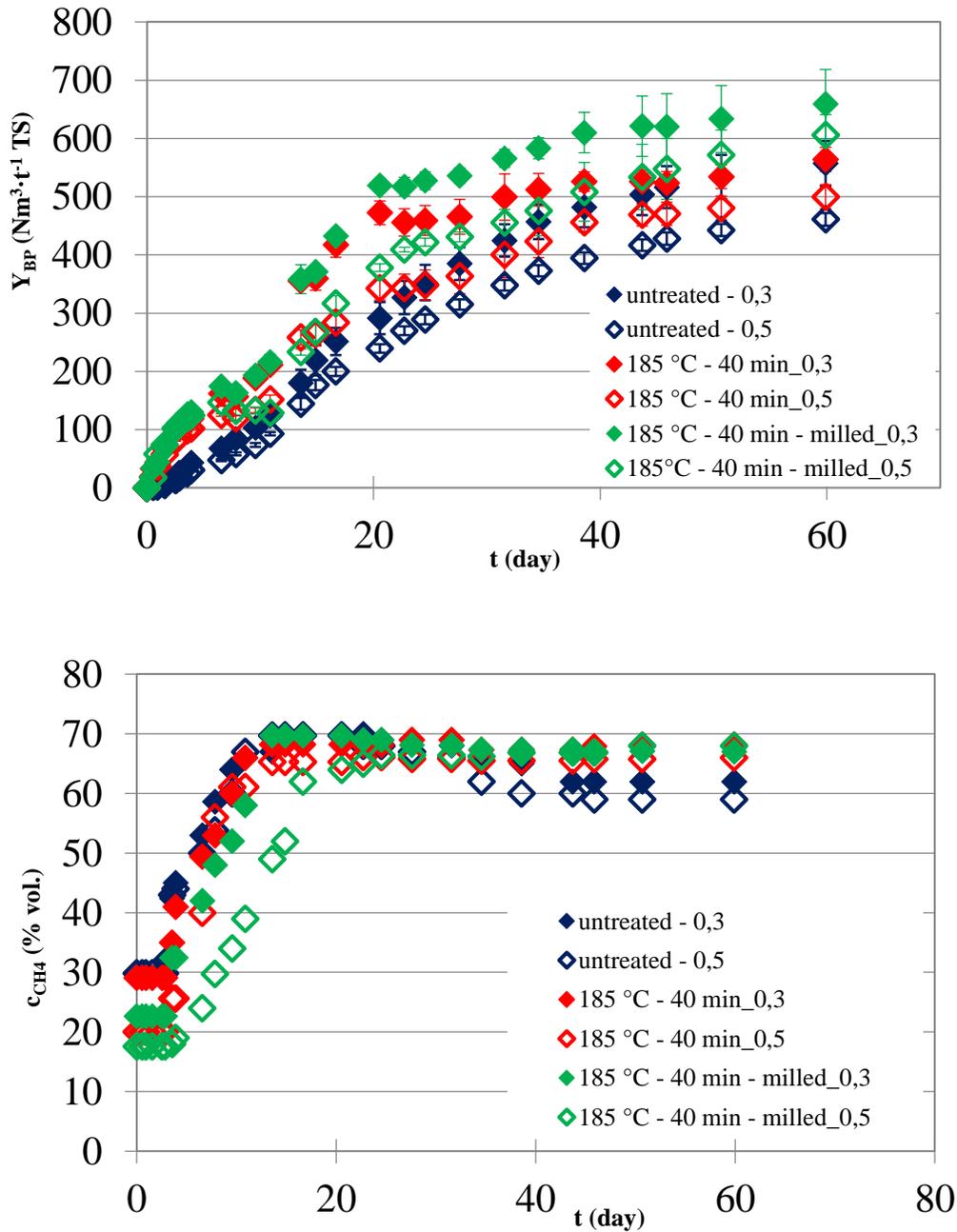


Figure 12. Biogas yield (1st) and volume concentration of methane in biogas (2nd) [16].

Obtained data depicted on the Table 3. Those results would be used in the practical calculations. In the Table 3 there are 3 different samples: untreated straw, and two samples at the same temperature and residence time, but the one was premilled. For the practical part would be selected untreated one and premilled, as the premilled has the best increase in methane production and untreated is the reference. However these are only experimental

results, they would be compared to those from real practical point view. This procedure will justify the truth of the experimental data.

Sample	Specific production [Nm ³ .t ⁻¹ TS]				
	<i>Y_{BG}</i> <i>average</i>	<i>Biogas</i> <i>formation</i> [wt.%] - <i>C^m_{BG}</i>	<i>Y_{CH4}</i> <i>average</i>	<i>CH4 formation</i> [wt.%] - <i>C^m_{CH4}</i>	<i>Increase in</i> <i>CH4</i> <i>production</i>
untreated straw	509 ± 58	60	243 ± 49	28	-
185 °C /40 min	532 ± 52	64	299 ± 29	33	23 %
Milled + 185 °C /40 min	633 ± 52	76	362 ± 43	34	49 %

Table 3. Table of important data [16].

3. OBJECTIVES

Until now, there were no public accessible work, where you can find detailed description of mass, energy balance and economic analysis for the biogas plant. The aim of these work is to show the detailed calculation of the plant, however the brand new pretreatment would be utilized in parallel.

3.1 General objectives

- Design of intensified and non-intensified model of biogas plant with full energy – economy balance.

3.2 Specific objectives

- To review of non-intensified biogas plant working principle.
- To make a critical literature search about utilization of biogas in the past and nowadays substrate pretreatments.
- To create based on literature search, a flowsheets of non-intensified and intensified biogas plants. In the same way to create the best heat regeneration cycle for each type of the plant.
- To make, based on flowsheet, an excel file of mass and energy balance for non-intensified and intensified technology.
- To perform the economical calculations of non-intensified and intensified biogas plant. It will include the detail calculation of total investment costs, operating costs.
- To calculate based on economical calculation, the payback period for two types of biogas plants.
- To compare non-intensified and intensified technology from utilization and economical point of view.

4. INTENSIFIED PLANT

4.1 Flowsheet of thermal-expansive pretreatment technology

In chapter 2.6 we discussed about lab-scale and lab-results of thermal-expansive (intensified) pretreatment. From here starts practical part. The idea is to create the flowsheet of intensified pretreatment, which is efficient and has regeneration of energy within the system.

After studying Kratky and Jirout [16, 23], the author proposed own mass and energy balanced flowsheet of thermal-expansive pretreatment technology.

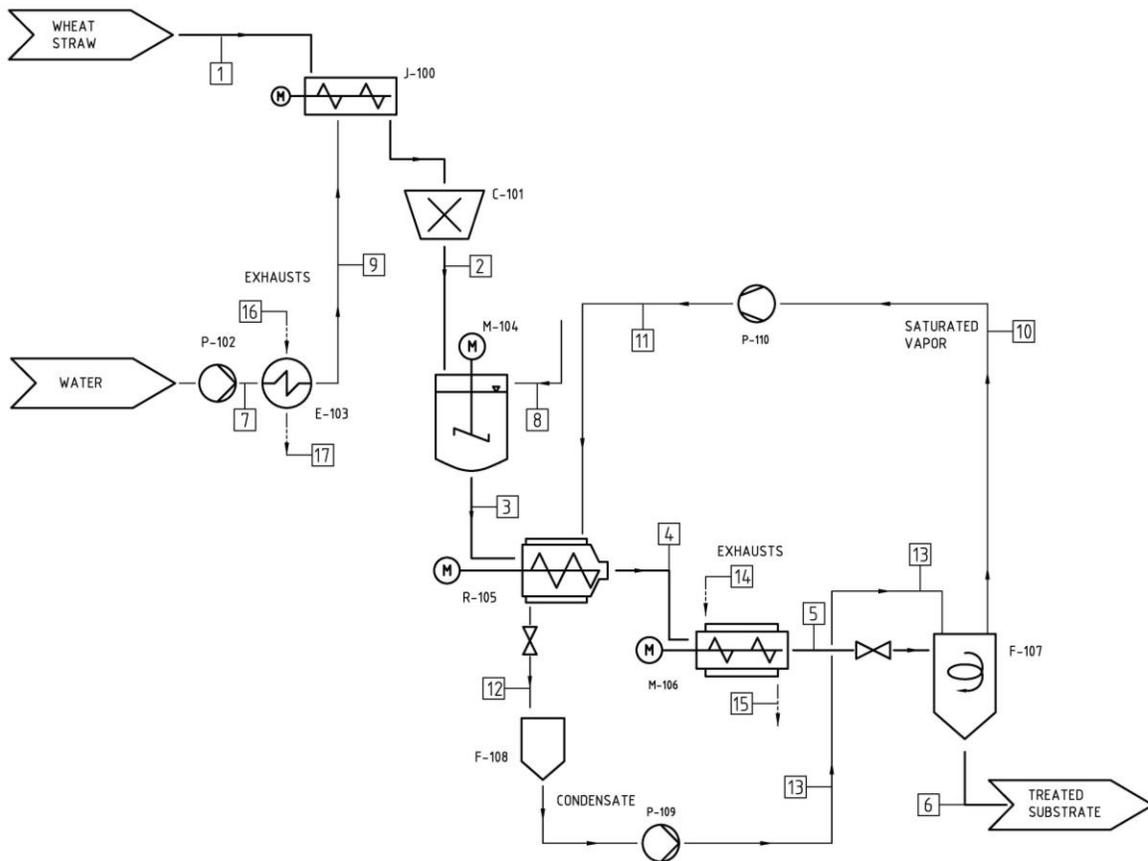


Figure 13. Thermal – expansive pretreatment flowsheet.

The flowsheet of thermal-expansionary pretreatment technology was constructed. The following flowsheet is adapted, for 5 % wt. of substrate and 95 % wt. of water, for treated substrate [16].

The stream 1 (substrate) is mixed with stream 9 (heated water) and sent to milling machine (C-101) by screw conveyor (J-100). The mixing of heated water and substrate is done for intensification of the milling [10]. The stream 2 is sent to vessel (M-104) and mixed with stream 8 (heated water). The stream 8 heated by heat exchanger, which is used for cooling of treated substrate (not illustrated). The mass flow rate of these two streams (2, 8) is essential, so that the mass fraction of water within two streams is equal to 95 %. After the mixing, the stream 3 sent to extruder (R-105). Extruder playing the crucial part in this technology. The inlet mixture (stream 3) is pressurized from 1 to 11.2 bar, at the same time it is heated by compressed steam (stream 11), which was the outlet from expansion vessel (F-107). The stream 4 processed after in jacketed vessel with screw conveyor (M-106). This jacketed vessel is used for heating of pressurized mixture up to 185 °C, as it was discussed in “Lab – results” chapter. Jacketed vessel supplied by exhaust gases, which were created within the combustion of biogas in cogeneration unit. The surplus of exhaust gases energy that is left after heating of vessel (M-106), is used for heating of stream 7. Finally, after all, the stream 5 rapidly expanded inside the expansion vessel, where the soaked water by the substrate immediately evaporated, due to difference of pressure. The fibers of substrate disrupted to the powder level, due to evaporation of water which was inside the substrate. The treated substrate is now ready for subsequent processing.

4.2 Substrate mass flow rate

For better understanding of intensified pretreatment technology effectiveness, the author proposed to calculate the mass flow rate for non-intensified (untreated wheat straw) case, and then to use the same number for both cases.

Medium biogas plants have the power 500 kW of cogeneration unit. Usually, when we combust 100 % of biogas in cogeneration unit, 35 – 42 % of stored energy in the fuel is converted to electricity, 50 – 43 % is converted to heat [3]. So, let us take the average 38 % of converted electricity, and for calculation in the next chapters 45 % of produced heat.

Using formula below, we are able to estimate mass flow rate with already given conditions.

$$\dot{m}_{TS} = \frac{\dot{Q}}{\eta_E \times q_{CH_4} \times Y_{CH_4_UNTREATED}} \quad (1)$$

$$\dot{m}_{TS} = \frac{500}{0.38 \times 9.94 \times 243}$$

$$\dot{m}_{TS} = 0.545 t_{TS} \cdot h^{-1} = 0.152 kg_{TS} \cdot s^{-1}$$

As the feasibility prove of substrate mass flow rate, the following figure was found. It describes the amount of added energy (y-axis) to produce the corresponding electricity (x-axis) within the cogeneration unit [22].

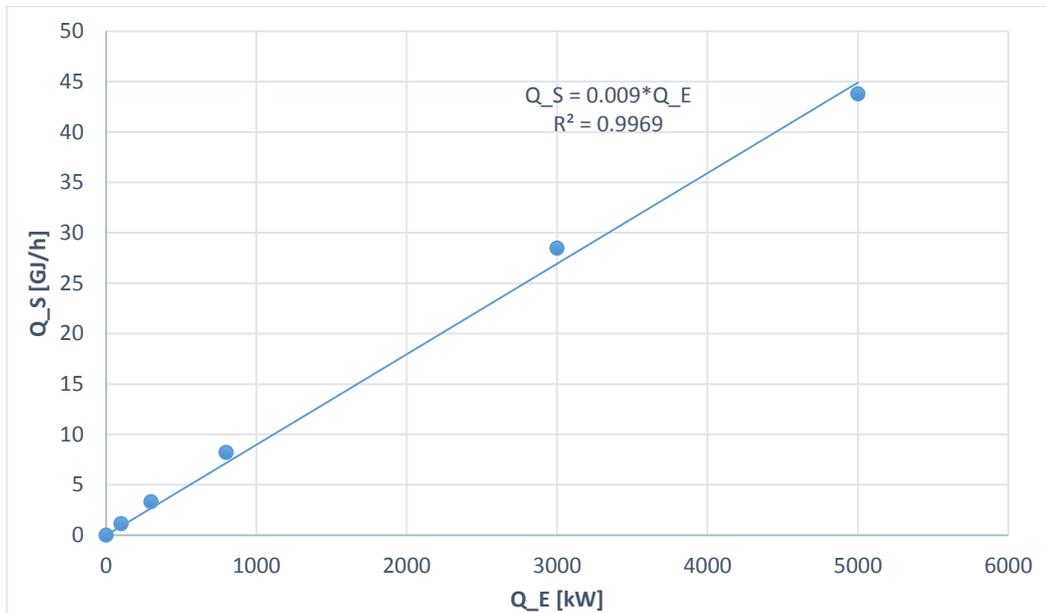


Figure 14. Energy vs. Power, linear regression [22].

If we want to estimate the supplied power to cogeneration unit, we can use linear equation (obtained from linear trend on the figure), and for Q_E value we will use 500 kW, as before.

$$\dot{Q}_S = 0.009 \times \dot{Q}_E \quad (2)$$

$$\dot{Q}_s = 0.009 \times 500 = 4.5 \text{GJ.h}^{-1}$$

The LHV of methane is 50 MJ/kg, the biogas formation from substrate is 60% wt. (chapter 2.6.2 – untreated straw) and the methane formation from biogas is 28 % wt. (chapter 2.6.2 – untreated straw). At this point, we are able to calculate the amount of methane in the estimated mass flow rate of the substrate.

$$\dot{m}_{CH_4} = (\dot{m}_{TS} \times C_{BG}^m) \times C_{CH_4}^m \quad (3)$$

$$\dot{m}_{CH_4} = (0.152 \times 0.6) \times 0.282 = 0.026 \text{kg.s}^{-1} = 92.6 \text{kg.h}^{-1}$$

Therefore, the supplied methane energy will look like,

$$\dot{Q}_s = \dot{m}_{CH_4} \times LHV \quad (4)$$

$$\dot{Q}_s = 92.6 \times 50 = 4629 \text{MJ.h}^{-1} = 4.6 \text{GJ.h}^{-1} \approx 4.5 \text{GJ.h}^{-1}$$

Now you could see that obtained experimental data correspond to practical life of biogas plant.

4.3 Power of cogeneration unit

Previously, we found the amount of substrate for supplying of cogeneration unit with power 500 kW. The mass flow rate was considered for the case of untreated substrate biodegradation, as the author propose, now we will calculate the power of cogeneration unit with supplying of treated substrate.

The following formula will help us to estimate the power of cogeneration unit.

$$\dot{Q} = \dot{m}_{TS} \times \eta_E \times q_{CH_4} \times Y_{CH_4_TREATED} \quad (5)$$

$$\dot{Q} = (0.152 \times 3600/1000) \times 0.38 \times 9.94 \times 362 \cong 750 \text{kW}$$

Comparing two different cases, one for untreated and another for treated substrate, we can see that for the same amount of mass flow it is possible to obtain 50 % more power. Unfortunately, there is no 750 kW power cogeneration unit, propose of author is to use one with power 250 kW and another one with 500 kW, so that in total they will have 750 kW power.

4.4 Flowsheet of intensified plant

Based on literature search and Kratky, Jirout [16, 23] the author proposed balanced flowsheet (appendix A). However, the flowsheet of intensified plant is quite big. Thus, in here you can see only figure of main streams.

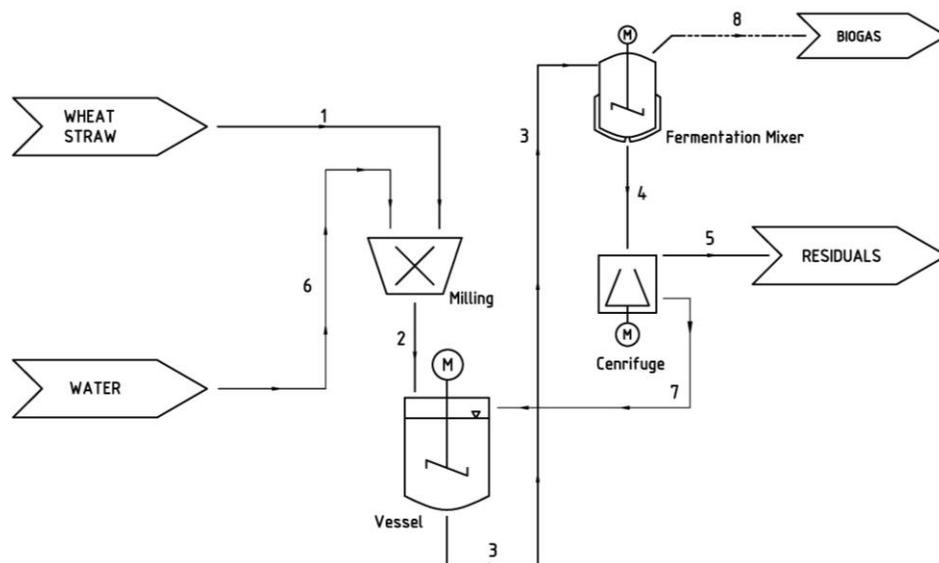


Figure 15. Main streams for intensified plant.

Briefly, the substrate is milled together with soaked water and sent to a vessel. In the vessel, the substrate is mixed with the water in certain amount. The mixture is ready for fermentation, it sent to fermentation mixer (digester) where it is held for certain residence time. Within the fermentation process the biogas is produced. The residuals after fermentation process are used as the fertilizers for different plants, and the water is reused in the vessel.

Nevertheless, the explanation of the flowsheet is in must. See the flowsheet in appendix A.

In the chapter 4.1 we discussed about flowsheet for thermal expansionary pretreatment technology. As this flowsheet was taken from the intensified biogas plant, proposed by author, we will start from the point, when the substrate is already treated and ready for fermentation.

- So, the treated suspension (stream 8) is cooled down by water (stream 14), what is the residual after fermentation. As the result, heat exchanger (E-108) heats the water (stream 14) up to 92 °C (stream 15). The heated water is sent to vessel (M-104). The following recirculating of water inside the plant creates low necessity of supplied water.
- Further, the water cooled down up to 75 °C (stream 9). Nevertheless, this water still carrying energy and it could be used for heating the water (stream 32, 33) which is used for keeping the suspension inside digester (M-113) at temperature 35 °C. The temperature wasn't chose randomly, the idea of keeping 35 °C creates good conditions for mesophilic microorganisms (chapter 2.3.1). The suspension is cooled down (E-111) up to 37.5 °C and sent to fermentation mixer (digester – M-113).
- From this point the fermentation process starts. As was discussed before (chapter 2.2), the fermentation process consist mainly of 4 steps: hydrolysis, acidogenesis, acetogenesis and methanogenesis. It was found by Kratky and Jirout [16], that it takes approximately 40 days (residence time), for treated substrate, to reach the highest yield within the fermentation process. It is very important to remember the residence time for fermentation, because in further calculations we will use this number to calculate the volume of a digester, and capital cost.
- Going back, during the fermentation, biogas (stream 16) is produced and it is sent to purification step (H-501). Produced biogas has usually water vapor content of 5 % vol. [3]. Thus, some purification should be installed to the process, otherwise water vapor decrease efficiency of methane burning. The installed biogas cleaner (H-501) decrease moisture content, now the biogas could be burned.
- The cogeneration unit of the power 750 kW has the methane supply (stream17). The cogeneration unit equipped with spark – ignition (SI) engine, where the methane content is combusted. Whenever the inlet demand to the cogeneration unit would be higher, it would be directly burned in the burner (B-503).

- The products of combustion, exhausts, are produced (stream 19). The exhaust amount carrying still high enough energy to be utilized, temperature of the stream is 530 °C. So, these exhausts are sent to screw vessel (M-106). They heat the suspension to reach the desired temperature 185 °C. The outlet of these exhausts have temperature 350 °C. The residuals of energy are used for heating (E-103) water (stream 2), which is mixed with substrate.
- The residuals of fermentation are separated in the centrifuge (H-115), into substrate (stream 12) and water (stream 13). The separated substrate (stream 12) still has some amount of water, and it is preferably to dry it before sending it to a field as the fertilizer. The separated water (stream 13), at first, is heated by the hot water (stream 26), after it is heated by hot suspension (stream 8) up to 92 °C and eventually sent to the vessel (M-104). The series usage of heat exchangers (E-120, E-108) is used for preheating of water (stream 13), as the result, decreasing the heat transfer area of the second heat exchanger (E-108).
- The heated water has the temperature 80 °C. It was estimated, that some amount of water is used for heating of water (stream 13) and another amount would be used for the needs of a plant (tap water, cleaning, etc.).
- During production could appear bad smell that is why we will remove it (stream 34).

During the construction of intensified flowsheet, author tried to create efficient, balanced and good regeneration cycles within all operations of intensified plant.

For more detail look into appendix A, C.

4.5 Mass balance

Before describing each stream, we need to define mass flow rate of biogas. As was discussed before, the biogas consist of: methane, carbon dioxide and water vapor. Using before calculated mass flow rate of substrate (chapter 4.2) and the yield of biogas, methane (chapter 2.6.2) for treated and milled substrate, we can create the following table. The density of biogas was defined by Deublin and Steinhauser [3].

	density at 0°C [kg.m ⁻³]	Vol. flow [m ³ .s ⁻¹]	by vol. [%]	Mass. Flow [kg.s ⁻¹]	by wt. [%]
Biogas	1.20	0.096	100	0.115	100
CH ₄	0.71	0.055	57.2	0.039	33.8
CO ₂	1.91	0.036	37.8	0.074	63.7
H ₂ O	0.59	0.005	5	0.003	2.5

Table 4. Biogas composition.

For more details please look into appendix B.

The biogas and methane formation, from Table 3, was calculated using Table 4.

For calculation of mass balance was used only main streams, because we are interested in those streams where the concentration is changing.

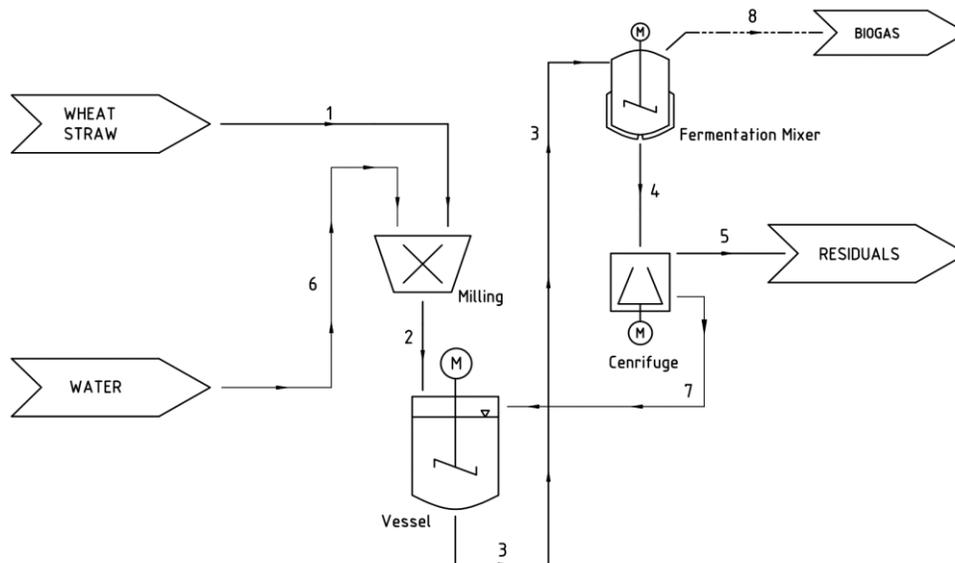


Figure 16. Main stream for mass flow rate.

Based on Kratky and Jirout [16, 23] the critical concentrations were chosen. The suspension after milling (stream 2) should have 40 % wt. concentration of water. After, when the suspension is mixed with residual water from centrifuge (stream 7) the concentration of water in new suspension (stream 3) must be 95 % wt. As was discussed in lab – results (chapter 2.6.2), the conversion of substrate to biogas is 76 % wt, so it means that the rest (residuals)

are going out from the fermentation mixer (stream 4). After separation there is 1 % wt. [3] of water in the residuals of substrate (stream 5) from the water which was taken out of fermentation mixer (stream 4). If we consider all the previous concentrations we would be able to estimate all the individual mass flow rates.

As the example here you could see the calculation of suspension from the milling machine (stream 2).

$$C_{water}^m = \frac{\dot{m}_{water}}{\dot{m}_{water} + \dot{m}_{substrate}} \quad (6)$$

$$\dot{m}_{water} = \frac{C_{water}^m \times \dot{m}_{substrate}}{(1 - C_{water}^m)} = \frac{0.4 \times 0.152}{(1 - 0.4)}$$

$$\dot{m}_{water} = 0.101 \text{ kg} \cdot \text{s}^{-1}$$

On Table 5 you can see the results of such a calculation. The number, which is standing first after letter “m” refers to stream number, and second number refers to dry substrate (index “1”) or water (index “2”).

m1;1	0.152	kg.s ⁻¹
m2;1	0.152	kg.s ⁻¹
m2;2	0.101	kg.s ⁻¹
m3;1	0.152	kg.s ⁻¹
m3;2	2.888	kg.s ⁻¹
m4;1	0.039	kg.s ⁻¹
m4;2	2.882	kg.s ⁻¹
m5;1	0.039	kg.s ⁻¹
m5;2	0.029	kg.s ⁻¹
m6;2	0.101	kg.s ⁻¹
m7;2	2.853	kg.s ⁻¹
m8;1	0.113	kg.s ⁻¹
m8;2	0.006	kg.s ⁻¹

Table 5. Results of mass balancing.

However, there is small problem appears. The separated water from centrifuge (stream 7) has more water than we need. The crucial point is to obtain exactly 95 % wt. of water content in suspension (stream 3). So, the author suggest, to put an extra vessel between centrifuge and vessel, which will send only needed amount of water.

4.6 Energy balance

The calculation of energy balance based: mass flow rates, temperatures of inlet and outlet streams, and properties of the fluid. The proper and efficient utilization of the energy for each stream, will make the plant, simply, much cheaper. Nevertheless, you will see the calculation of energy balance for each point, where it was needed. According to flowsheet (appendix A) there are 6 points, where the energy balance should be applied.

- *Expansion vessel (F-107)*

There are two outlets in the expansion vessel, the reason is that when the suspension (stream 7) is depressurized from 11.2 bar to 1 bar the liquid and vapors are appeared. For the calculation we will assume the suspension as the water only, as there is only 5 % wt. of the substrate inside.

Let us repeat, the suspension (stream 7) exactly on saturated liquid level (11.2 bar, 185 °C) is depressurized to 1 bar, and result is some vapor and liquid. All used enthalpy values are tabulated. So, the following enthalpy balance could be applied.

$$\dot{m}_{liq_in(7)} \times h_{SAT_LIQ(7)} = \dot{m}_{vap_out(28)} \times h_{VAPOR(28)} + \dot{m}_{liq_out(8)} \times h_{LIQUID(8)} \quad (7)$$

However, we have 2 unknowns: \dot{m}_{vap_out} and \dot{m}_{liq_out} . So, we need to apply mass balance.

$$\dot{m}_{liq_in(7)} = \dot{m}_{vap_out(28)} + \dot{m}_{liq_out(8)} \quad (8)$$

$$\dot{m}_{vap_out(28)} = \dot{m}_{liq_in(7)} \times \frac{h_{SAT_LIQ(7)} - h_{LIQUID(8)}}{h_{VAPOR(28)} - h_{LIQUID(8)}} = 3.04 \times \frac{785.3 - 417.4}{2674.9 - 417.4}$$

$$\dot{m}_{vap_out(28)} = 0.5 \text{ kg} \cdot \text{s}^{-1}$$

$$\dot{m}_{liq_out(8)} = \dot{m}_{liq_in(7)} - \dot{m}_{vap_out(28)} = 3.04 - 0.5 = 2.54 \text{ kg.s}^{-1}$$

Nevertheless, we consider steady-state case. So, it means that, according to the flowsheet, the saturated vapor (stream 28) is returning back to the expansion vessel (F-107) as the condensed vapor (stream 31). So, the suspension outlet is equal to the inlet (stream 7) to the expansion vessel (F-107).

- *Extruder (R-105)*

The energy balance based on heat transfer from hot stream (saturated vapor at 11.2 bar) to cold stream (suspension). The suspension is considered to be water, as the content of water is 95% wt.

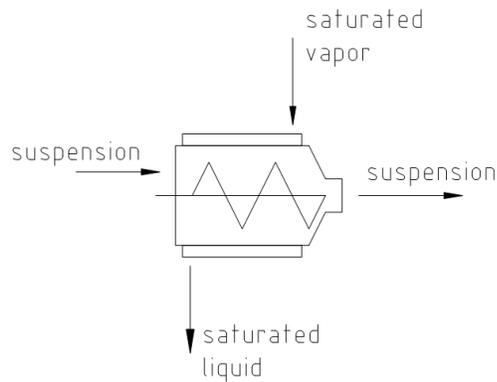


Figure 17. Extruder (R-105).

In previous calculation we calculated the mass flow rate of saturated vapor, but saturated vapor at 1 bar has around 100 °C, still we need to heat the suspension up to 185 °C. The author proposes to put the compressor (P-202) which will pressurize the saturated vapor up to 11.2 bar (corresponds to saturated liquid at 185 °C). In the following formula, used enthalpy values and fluid properties are tabulated.

Nevertheless, let us assume that the inlet temperature of suspension is 92 °C. After you will see the feasibility proof of such a selection.

$$Q = \dot{m}_{vapor} \times (h_{SAT_VAP(29)} - h_{SAT_LIQ(30)}) = \dot{m}_{suspension} \times C_{P_WATER} \times (T_6 - T_5) \quad (9)$$

$$0.5 \times (2781.4 \times 10^3 - 785.3 \times 10^3) = 3.04 \times 4200 \times (T_6 - 92)$$

$$T_6 \cong 169^\circ C$$

Knowing all the temperatures and estimating the overall heat transfer coefficient we are able to calculate heat transfer area. The overall heat transfer was selected for the case steam-water heat transfer through stainless steel wall [34].

$$Q = \dot{m}_{vapor} \times (h_{SAT_VAP(29)} - h_{SAT_LIQ(30)}) = k \times S \times \Delta T_{ln} \quad (10)$$

$$0.5 \times (2781.4 \times 10^3 - 785.3 \times 10^3) = 680 \times S \times 43.3$$

$$S = 33.6m^2$$

- *Jacketed vessel (M-106)*

From the previous calculation we calculated that energy of the vapor is not enough to achieve the desired temperature level. At this point, we utilize the energy of exhaust gases, which were the result of methane combustion in the chamber of cogeneration unit.

The mass flow rate of exhausts was calculated (see appendix C) and the temperature was estimated to be 530 °C [3].

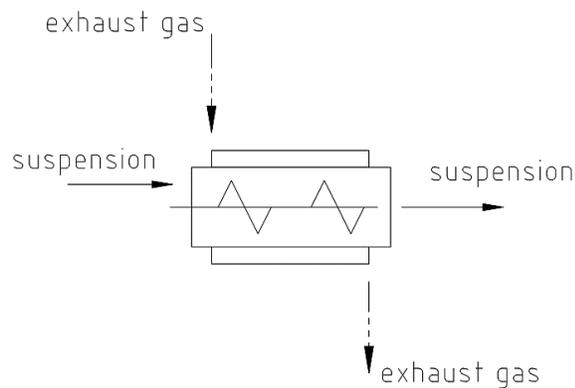


Figure 18. Jacketed vessel (M-106).

$$Q = \dot{m}_{exhausts} \times C_{P_EXHAUST_GAS} \times (T_{20} - T_{21}) = \dot{m}_{suspension} \times C_{P_WATER} \times (T_7 - T_6) \quad (11)$$

$$1 \times 1100 \times (530 - T_{21}) = 3.04 \times 4200 \times (185 - 169)$$

$$T_{21} \cong 350^\circ C$$

Knowing all the temperatures and estimating the overall heat transfer coefficient we are able to calculate the heat transfer area. The overall heat transfer was selected for the case gas-liquid heat transfer [36].

$$Q = \dot{m}_{exhausts} \times C_{P_EXHAUST_GAS} \times (T_{20} - T_{21}) = k \times S \times \Delta T_{ln} \quad (12)$$

$$1 \times 1100 \times (530 - 350) = 60 \times S \times 253.7$$

$$S = 13m^2$$

- *Heat exchanger (E-103)*

The energy residuals of exhaust can be utilized for heating the water, which is soaked by substrate and therefore milled more efficient. The propose is to heat it up to 92 °C, due to temperature level necessities of vessel (M-104). Having all needed temperatures we can estimate the outlet temperature of exhaust gases.

$$Q = \dot{m}_{exhausts} \times C_{P_EXHAUST_GAS} \times (T_{21} - T_{22}) = \dot{m}_{water} \times C_{P_WATER} \times (T_3 - T_2) \quad (13)$$

$$1 \times 1100 \times (350 - T_{22}) = 0.101 \times 4200 \times (92 - 15)$$

$$T_{22} \cong 320^\circ C$$

- *Heat exchanger (E-108)*

As the suspension is depressurized, the liquid part (suspension) is ready to be fermented. Nevertheless, we assume, that at 1 bar the liquid will have 100 °C (on the level of saturated liquid). As was discussed before (chapter 2.3.1), the mesophilic microorganisms more adapted

to smaller temperature range. So, we can use the carrying energy of suspension for other propose.

The water residuals after fermentation are more likely to be reused. The author propose to heat these residuals up to 92 °C and send them back to vessel (M-104). However, temperature difference after separation of water (H-115) is big enough, around 77 °C, and we expect high enough heat transfer area. But let us assume, that cold stream has the inlet temperature 35 °C (see later *Heat exchanger E-120*).

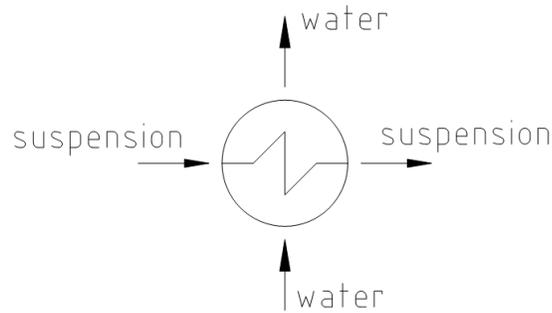


Figure 19. Heat exchanger (E-108).

$$Q = \dot{m}_{suspension} \times C_{P_WATER} \times (T_8 - T_9) = \dot{m}_{water} \times C_{P_WATER} \times (T_{15} - T_{14}) \quad (14)$$

$$3.04 \times 4200 \times (100 - T_9) = 2.8 \times 4200 \times (92 - 35)$$

$$T_9 \cong 48^\circ C$$

Knowing all the temperatures and estimating the overall heat transfer coefficient we are able to calculated heat transfer area. The overall heat transfer was selected for the case liquid-liquid heat transfer [36].

$$Q = \dot{m}_{suspension} \times C_{P_WATER} \times (T_8 - T_9) = k \times S \times \Delta T_{ln} \quad (15)$$

$$3.04 \times 4200 \times (100 - 48) = 1800 \times S \times 10$$

$$S = 37m^2$$

- Heat exchanger (E-111)

The suspension (stream 9) still need to be cooled down, to reach the adaptable temperature for mesophilic microorganisms.

The fermentation mixer (M-113) has big volume, because of high residence time. The crucial point is to keep the temperature inside at around 35 °C. Thus, we need to supply heat to the fermenter. It was found that every cubic meter of fermenter needs 12.5 W.m⁻³ [28]. The choice of outlet temperature for cold stream (stream 33) is randomly reasonable in order to obtain mass flow rate.

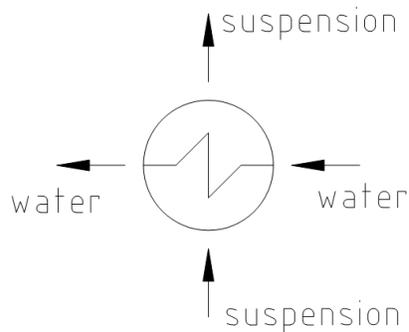


Figure 20. Heat exchanger (E-111).

$$Q = V \times 12.5 = \dot{m}_{water} \times C_{P_WATER} \times (T_{33} - T_{32}) \quad (16)$$

In the previous formula we can see the volume involved calculation. The calculation of volume is rather simple, it is simply based on volumetric flow into vessel and the residence time.

$$V = \frac{\dot{m}_{SUSPENSION}}{\rho_{WATER}} \times t = \frac{3.04}{1000} \times 40 \times 24 \times 3600 \cong 10500 m^3$$

$$10500 \times 12.5 = \dot{m}_{water} \times 4200 \times (42 - 34.5)$$

$$\dot{m}_{water} = 4.2 kg.s^{-1}$$

Knowing all the temperatures and estimating the overall heat transfer coefficient we are able to calculate heat transfer area. The overall heat transfer was selected for the case liquid-liquid heat transfer [36].

$$Q = \dot{m}_{water} \times C_{P_WATER} \times (T_{33} - T_{32}) = k \times S \times \Delta T_{ln} \quad (17)$$

$$4.2 \times 4200 \times (42 - 34.5) = 1800 \times S \times 4.1$$

$$S = 17.6 m^2$$

- *Heat exchanger (E-120)*

Previously, we assumed that water residuals will have temperature 35 °C. To obtain such a temperature we will use the heated water from cogeneration unit. Let us assume that we have mass flow rate of 6 kg.s⁻¹ and temperature 80 °C (for more details look into appendix C – “CHP” part).

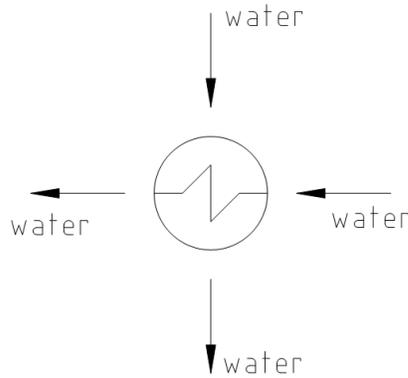


Figure 21. Heat exchanger (E-120).

$$Q = \dot{m}_{water} \times C_{P_WATER} \times (T_{26} - T_{27}) = \dot{m}_{water} \times C_{P_WATER} \times (T_{14} - T_{13}) \quad (18)$$

$$6 \times 4200 \times (80 - T_{27}) = 2.8 \times 4200 \times (35 - 15)$$

$$T_{27} \cong 71^\circ C$$

Knowing all the temperatures and estimating the overall heat transfer coefficient we are able to calculate heat transfer area. The overall heat transfer was selected for the case liquid-liquid heat transfer [36].

$$Q = \dot{m}_{water} \times C_{P_WATER} \times (T_{26} - T_{27}) = k \times S \times \Delta T_{ln} \quad (19)$$

$$6 \times 4200 \times (80 - 71) = 1800 \times S \times 50.5$$

$$S = 2.5m^2$$

4.7 Economic analysis

Even if the production line seems to be good and energy efficient, without economic analysis of the project it will never be affirmative. Thus, the total invests and payback time are in high priority.

In previous calculations, we obtained all the necessary, or so called “key parameters”, for preliminary estimation of capital cost for every individual apparatus.

4.7.1 Capital cost

The capital cost of a plant is estimated using factorial method, which has error $\pm 30\%$. The factorial method of cost estimation based on purchased equipment costs and therefore requires good estimates for equipment costs. In this work the preliminary estimates were made with help of correlations and formula [29].

$$C_e = a + b \times S^n \quad (20)$$

It would be too long to write every calculation. However, you can see some calculations, which were mainly used.

For more details about economic analysis look into appendix D.

- *Liquid pump (P-300)*

In the calculation of heat exchanger E-111, it was found the mass flow rate. Using the previous formula and defining mass flow rate as the key parameter we are able to calculate capital cost.

$$C_e = 3300 + 48 \times 4.2^{1.2}$$

$$C_e = 3569 \text{ \$}$$

- *Fermentation mixer (M-113)*

Unfortunately, fermentation mixer is not tabled value, hence it is not the same calculations as e.g. for liquid pumps etc. However, the price estimate was based on literature search.

Deublin and Steinhauser [3] say, that estimation of total investment cost could be calculated from the volume of fermenter for non-intensified plant, because nowadays practice has only this possibility. The propose is to take the cost from 300 to 500 \$.m⁻³ and based on this estimation recalculate the price for fermenter alone. Nevertheless, it was found 110 \$.m⁻³ for fermenter price alone.

$$C_e = 10500 \times 110$$

$$C_e = 1.16 \text{ mln. \$}$$

Such a way was chosen, because fermenter varies from 30 % to 50 % of total capital cost for biogas plant. So, it is mean that calculation should be in deep enough to decrease the error of calculation.

Moving from one apparatus to another, we are able to estimate the total capital cost. In the table below you could see all apparatuses with their prices.

Some costs like: burner, biogas purification, screw pump etc. they were taken from real biogas plants in Czech Republic, and they are in czech korunas. Unfortunately, it is not possible to calculate them directly using factorial method, thus they were signed as reference (ref.).

Number	Name	Key parameter	a	b	n	key parameter	capital cost [\$]
B-203	Burner	duty, MW	180000			ref.	9268
C-101	Milling machine	kg/h	3000	390	0.5	900	14700
E-103	Heat exchanger	area, m ²	1100	850	1	1	1950
E-108	Heat exchanger	area, m ²	1100	850	1	37	32550
E-111	Heat exchanger	area, m ²	1100	850	1	18	16400
E-120	Heat exchanger	area, m ²	1100	850	1	3	3650
F-107	Expansion vessel	capacity,m ³	5700	700	0.7	10	9208
F-200	Conical vessel	capacity,m ³	5700	700	0.7	10	9208
F-117	Conical vessel	capacity,m ³	5700	700	0.7	250	39093
F-118	Vessel	capacity,m ³	5700	700	0.7	10	9208
F-301	Vessel	capacity,m ³	5700	700	0.7	10	9208
F-400	Vessel	capacity,m ³	5700	700	0.7	10	9208
F-402	Distribution tank	capacity,m ³	5700	700	0.7	10	9208
G-500	Ventilator	m ³ /h	4200	27	0.8	360	7195
G-600	Ventilator	m ³ /h	4200	27	0.8	300	6789
G-601	Ventilator	m ³ /h	4200	27	0.8	300	6789
G-701	Ventilator	m ³ /h	4200	27	0.8	200	6071
H-400	Biofilter	capacity,m ³	76000	54000	0.5	0.4	110153
H-115	Centrifuge	power, kW	37000	1200	1.2	20	80694
H-501	Biogas Purification	ref.	600000			ref.	30894
J-100	Screw conveyor	ref.	90000			ref.	4634
J-109	Screw conveyor	ref.	90000			ref.	4634
M-104	Agitated vessel	volume,m ³	14000	15400	0.7	0.5	18784
M-106	Jacketed vessel with scr. con	volume,m ³	14000	15400	0.7	3	467558
M-110	Vessel	volume,m ³	14000	15400	0.7	0.6	19816
M-113	Fermentation Vessel	volume,m ³	110		\$/m ³	10500	1155000
P-102	Liquid pump	flow l/s	3300	48	1.2	0.2	3307
P-109	Screw pump	ref.	80000			ref.	4119
P-112	Screw pump	ref.	80000			ref.	4119
P-114	Screw pump	ref.	80000			ref.	4119
P-110	Compressor	power, kW	8400	3100	0.6	40	36753
P-119	Liquid pump	l/s	3300	48	1.2	2.853	3469
P-201	Liquid pump	l/s	3300	48	1.2	0.41	3316
P-300	Liquid pump	l/s	3300	48	1.2	4.2	3569
P-401	Liquid pump	l/s	3300	48	1.2	7	3796
R-105	Extruder	volume,m ³	14000	15400	0.7	6	461337
Q-502	Cogeneration unit	kW	1200		\$/kW	750	900000
Total Capital Cost [mln. \$]							3.56

Table 6. Total Capital Cost intensified plant.

After calculation of total capital cost we need to obtain total investment. The extra costs like construction (10%) and engineering (10%) should be accounted as well.

$$Tot. invst. cost = Tot. cap. cost + 0.1 \times Tot. cap. cost + 0.1 \times Tot. cap. cost \quad (21)$$

$$Tot. invst. cost = 3.56 + 0.1 \times 3.56 + 0.1 \times 3.56$$

$$Tot. invst. cost = 4.27 \text{ mln. \$}$$

4.7.2 Payback period

In order to estimate the payback time, we need to calculate a lot of different parameters: direct operating cost (raw material, laboratory etc.), indirect operating cost (insurance, corporate directions), distribution cost (transport), revenue from sales.

On the Table 7 you can see the results.

Profit from the sale of electricity & residuals (revenues from sales)	\$.year⁻¹	995,778
Operating costs excluding depreciation and interest payments	\$.year⁻¹	433,261
Direct operating costs	\$.year⁻¹	349,687
Raw materials	\$.year ⁻¹	110,574
Personal costs of employees of the operation	\$.year ⁻¹	41,362
Supervision	\$.year ⁻¹	2,068
Maintenance costs	\$.year ⁻¹	170,833
Consumables	\$.year ⁻¹	21,354
Laboratory	\$.year ⁻¹	0
Reserve	\$.year ⁻¹	3,497
Indirect operating costs	\$.year⁻¹	81,362
Insurance	\$.year ⁻¹	42,708
Corporate directions	\$.year ⁻¹	38,654
Distributional costs	\$.year⁻¹	2,211
Transport	\$.year ⁻¹	2,211
Taxable income	\$.year⁻¹	562,518

Table 7. Expenses and revenues.

Nevertheless, to obtain the previous results, we need to have project parameters. This project parameters or all economic calculations could be found in appendix D.

After estimating revenue, operating cost and other parameters we are able to find payback period. On the Fig. 22, 23 you could see these periods.

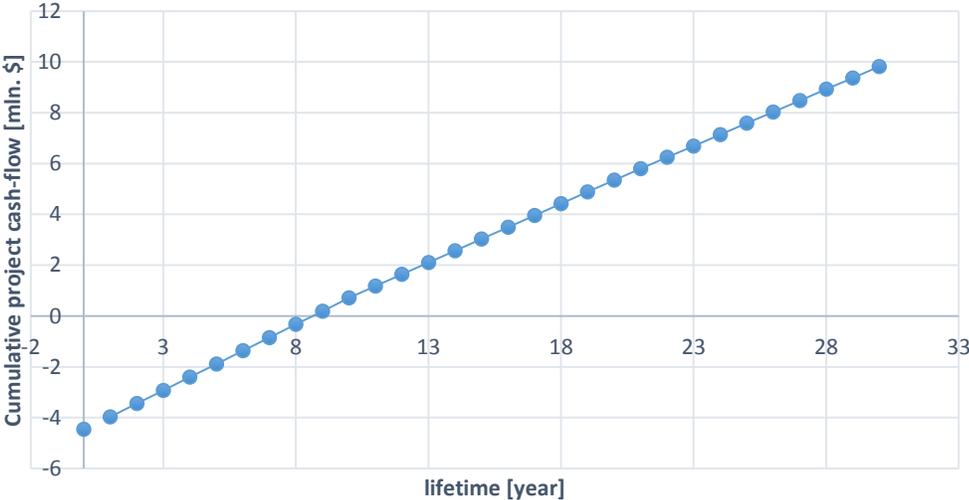


Figure 22. Payback period for 0 % discount rate.

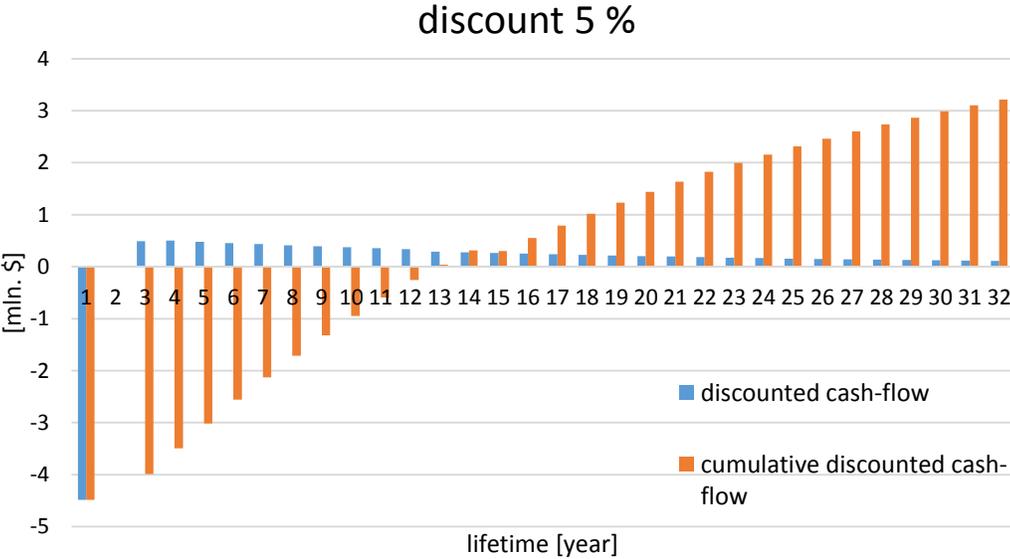


Figure 23. Payback period for 5 % discount rate.

Mainly, to obtain more realistic payback period, it is preferred to account discount rate. The discount rate in discounted cash-flow analysis takes into account not only the time value of money, but also the risk or uncertainty of future cash flow. So, for 0 % discount rate, or in other words the ideal case, we will get our money back in approximately 9 years. However, for 5 % discount rate the payback period is approximately 15 years.

The obtained years could seem to be unfeasible, but we shouldn't forget that the assumption is to take 100 % bank credit. In developed countries, government usually giving subsidies to build biogas plant or other plants with processing different kinds of renewable energy sources, also subsidies for electricity profit. These subsidies could vary from 30 % to 60 % of total investment cost [30]. The electricity profit, with subsidies (green electricity), could increase electricity price from 2 to 4 times. Assuming subsidies of 50 % for total investment cost only, we can decrease the payback period from 15 years to 6 - 7 years.

5. NON-INTENSIFIED PLANT

5.1 Flowsheet of non-intensified plant

Based on literature search and Kratky, Jirout [16, 23] the author proposed balanced flowsheet (appendix E). On the figure below you could see the flowsheet of main streams. Main streams are those streams, where the substrate takes part.

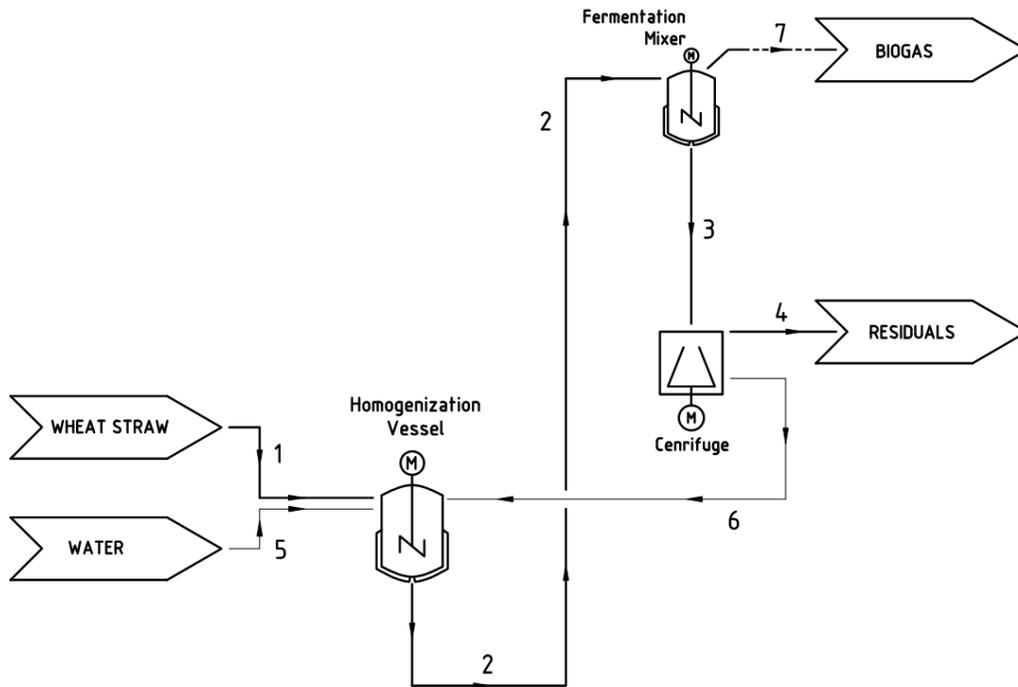


Figure 24. Main streams for non-intensified plant.

Shortly, the substrate (wheat straw) is mixed with water, with certain proportion, in homogenization vessel. After short residence time the suspension is sent to fermentation mixer (fermenter) for biodegradation. The production of biogas within the biodegradation takes time. Nevertheless, there are some residuals after fermentation process. Residuals are separated in the centrifuge. So, the one outlet of centrifuge is substrate residuals, which could be used as the fertilizers, another outlet stream is water, this water is reused in homogenization vessel.

As previously, the detail description of the flowsheet is in must (see Appendix E).

- Wheat straw (stream 1) is brought from the fields to the biogas plant and stored in the vessel (F-100). After, wheat straw is transported via screw conveyor (J-101) to homogenization vessel (M-104).
- Coming water is pumped to the same homogenization vessel. The concentration of water in the vessel is essential, as it should be 90 % wt. of water and 10 % wt. of substrate [16, 23]. As was said, the mesophilic microorganisms are more adapted on certain range of temperature (chapter 2.3.1). Nevertheless, there is another upcoming stream to the homogenization vessel, it came from as the residual of fermentation process. This water was heated up to 40 °C with help of heated water from cogeneration unit. The jacketed construction is used to keep the temperature level inside homogenization vessel at 35 °C (stream 21).
- The homogenized suspension is transported (stream 3) to the fermentation vessel (M-106). Kratky and Jirout [16] found the residence time, which is approximately 50 days, for untreated straw to achieve the best biogas yield (chapter 2.6.2).
- As biogas is produced (stream 8), it is purified (H-201). The water concentration of water in biogas is 5 % wt. For better combustion of methane it should be removed.
- The cogeneration unit (Q-202) has the power 500 kW (chapter 4.2). The inlet mass flow rate of methane is essential, because as the production is of methane is higher, the surplus is burned in the safety burner (B-203).
- Cogeneration unit heats the water (stream 20) for the needs of production (E-303, E-113) or for social usage inside the biogas plant.
- Some part of substrate is converted to biogas, another is residual. The residuals are approximately 40 % wt. of the substrate (chapter 2.6.2).
- The residuals consist of water and substrate (stream 4). They are separated in centrifuge (H-108).
- The separated substrate (stream 5) is transported (J-109) to the storage (F-110). This substrate still has some amount of water, and preferably it should be dried before it could utilized, e.g. as the fertilizers for the fields.
- The separated water (stream 6) heated by hot water from cogeneration unit (stream 16) and reused (stream 7) in homogenization vessel.
- The bed smell from vessel, could be removed (stream 13).

5.2 Mass balance

Before calculating each stream, we will calculate fractions of methane, carbon dioxide and water in the biogas. We already calculated mass flow rate of substrate (chapter 4.2) and yield of biogas, methane (chapter 2.6.2) for untreated substrate. Now we can construct the following Table 8.

	density at 0 °C [kg.m ⁻³]	Vol. flow [m ³ .s ⁻¹]	by vol. [%]	Mass. Flow [kg.s ⁻¹]	by w.t [%]
BioGas	1.20	0.077	100	0.093	100
CH4	0.71	0.037	47.7	0.026	28.2
CO2	1.91	0.037	47.3	0.064	69.3
H2O	0.59	0.004	5	0.002	2.5

Table 8. Biogas composition.

For more details please look into appendix F.

For calculation of mass balance was used only main streams. The Fig. 24 describes these streams.

Based on Kratky and Jirout [16, 23] the critical concentrations were chosen. The suspension (stream 2) after homogenization has 90 % wt. of water and 10 % wt. of substrate. As was discussed (chapter 2.6.2), the conversion of substrate to biogas is 60 % wt. The residuals has 1 % of water from the water which was taken out of fermentation mixer (stream 3).

As the example, here you can see mass flow rate calculation of water into the fermentation mixer.

$$C_{water}^m = \frac{\dot{m}_{water}}{\dot{m}_{water} + \dot{m}_{substrate}} \quad (22)$$

$$\dot{m}_{water} = \frac{C_{water}^m \times \dot{m}_{substrate}}{(1 - C_{water}^m)} = \frac{0.9 \times 0.152}{(1 - 0.9)}$$

$$\dot{m}_{water} = 1.368 \text{ kg.s}^{-1}$$

On the following table you can see the results of such calculation. The number, which is standing first after letter “m” refers to stream number, and second number refers to dry substrate (index “1”) or water (index “2”).

m1;1	0.152	kg.s ⁻¹
m2;1	0.152	kg.s ⁻¹
m2;2	1.368	kg.s ⁻¹
m3;1	0.061	kg.s ⁻¹
m3;2	1.363	kg.s ⁻¹
m4;1	0.061	kg.s ⁻¹
m4;2	0.014	kg.s ⁻¹
m5;2	0.018	kg.s ⁻¹
m6;2	1.350	kg.s ⁻¹
m7;2	0.005	kg.s ⁻¹
m7;3	0.091	kg.s ⁻¹

Table 9. Results of mass balancing.

5.3 Energy balance

The calculation of energy balance based on: mass flow rates, temperatures of inlet and outlet streams, and properties of the fluid. In comparing to intensified version, there is less points where we need to apply energy balance.

For more details please look to the appendix G.

- *Heat Exchanger (E-303)*

The following heat exchanger is supplied by the hot water (stream 17) from cogeneration unit. The hot water will transfer energy to the cold water. The heated water will feed fermentation vessel (M-106). It was found that energy loss is around 12.5 W.m⁻³ [28]. Now having volumetric flow and residence time of the vessel we are able to estimate the volume of each.

$$V = \frac{\dot{m}_{SUSPENSION}}{\rho_{WATER}} \times t = \frac{1.52}{1000} \times 50 \times 24 \times 3600 \cong 6600 m^3 \quad (23)$$

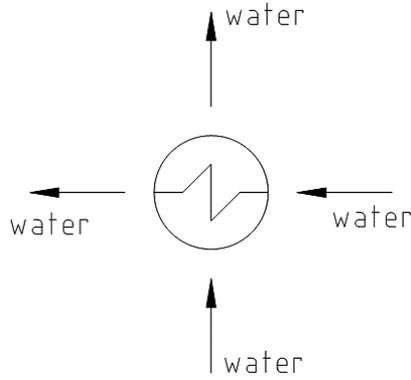


Figure 25. Heat exchanger (E-303).

$$Q = V \times 12.5 = \dot{m}_{water} \times C_{P_WATER} \times (T_{21} - T_{22}) \quad (24)$$

$$6600 \times 12.5 = \dot{m}_{water} \times 4200 \times (52 - 41)$$

$$\dot{m}_{WATER} = 1.8 \text{ kg} / \text{s}$$

Knowing all the temperatures and estimating the overall heat transfer coefficient we are able to calculate heat transfer area. The overall heat transfer was selected for the case liquid-liquid heat transfer [36].

$$Q = \dot{m}_{water} \times C_{P_WATER} \times (T_{21} - T_{22}) = k \times S \times \Delta T_{ln} \quad (25)$$

$$1.8 \times 4200 \times (52 - 41) = 1800 \times S \times 17$$

$$S = 3 \text{ m}^2$$

- *Heat Exchanger (E-113)*

The cold water (stream 6) heated by the hot water (stream 16), which is coming from cogeneration unit. It was estimated the hot water has temperature 70 °C and the mass flow rate is 2 kg.s⁻¹.

For more details please look into appendix G (“CHP” section).

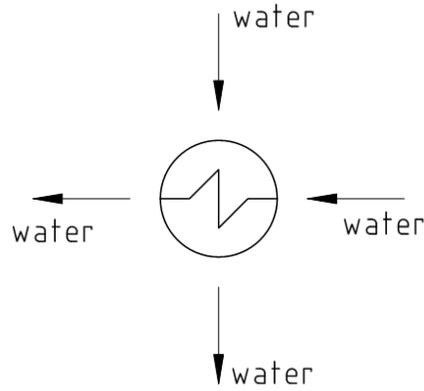


Figure 26. Heat exchanger (E-113).

$$Q = \dot{m}_{water} \times C_{P_WATER} \times (T_{16} - T_{19}) = \dot{m}_{water} \times C_{P_WATER} \times (T_7 - T_6) \quad (26)$$

$$2 \times 4200 \times (70 - T_{19}) = 1.35 \times 4200 \times (40 - 15)$$

$$T_{19} \cong 53^\circ C$$

Knowing all the temperatures and estimating the overall heat transfer coefficient we are able to calculate heat transfer area. The overall heat transfer was selected for the case liquid-liquid heat transfer [36].

$$Q = \dot{m}_{water} \times C_{P_WATER} \times (T_{16} - T_{19}) = k \times S \times \Delta T_{ln} \quad (27)$$

$$2 \times 4200 \times (70 - 53) = 1800 \times S \times 34$$

$$S = 2.5m^2$$

5.4 Economic analysis

The calculation of total investment cost based on the same principle, as for intensified plant. All the obtained values in mass and energy balance are used to calculate total capital cost.

5.4.1 Capital Cost

For instance, it would be taken some equipment for better understanding of the calculation procedure.

- *Fermentation mixer (M-106)*

At first, the calculation of fermenter cost based on volume. So, having the mass flow rate into the fermenter and residence time we are able to volume (see chapter 5.3). The cost of each cubic meter for concrete fermenter was found during literature search [3]. However, the cost found is only for estimation of total investment cost, which is 300-500 \$ [3]. Fortunately, other equipment were possible to calculate. Thus, the estimation of capital cost for fermenter is just fitting procedure to obtain more or less realistic value of total investment cost.

The price of 135 \$.m⁻³ was found. The smaller volume, the higher price for every cubic meter should be. In comparing to fermenter in intensified plant that one was 110 \$.m⁻³.

$$C_e = 6600 \times 135$$

$$C_e = 0.89 \text{ mln.}\$$$

The fermenter's capital cost is 47 % of total capital cost (see Appendix H).

- *Cogeneration unit (Q-202)*

The mass flow rate of the substrate was calculated for 500 kW cogeneration unit (chapter 4.2, 4.3). The cost of every kW of cogeneration unit is around 1200 \$.kW⁻¹ [31].

$$C_e = 500 \times 1200$$

$$C_e = 0.6 \text{ mln.}\$$$

For more details about capital cost estimation of other equipment please look into next Table 10 or appendix H.

Number	Name	Key parameter	a	b	n	key parameter	capital cost [\$]
B-203	Burner	duty, MW	180000			ref.	9268.085
E-303	Heat exchanger	area,m ²	1.1	850	0.4	3	1320.169
E-113	Heat exchanger	area,m ²	1.1	850	0.4	3	1320.169
F-100	Conical vessel	capacity,m ³	5700	700	0.7	250	39093.05
F-102	Conical vessel	capacity,m ³	5700	700	0.7	20	11399.27
F-305	Vessel	capacity,m ³	5700	700	0.7	10	9208.311
F-111	Vessel	capacity,m ³	5700	700	0.7	10	9208.311
F-306	Vessel	capacity,m ³	5700	700	0.7	15	10359.74
F-110	Conical vessel	capacity,m ³	5700	700	0.7	250	39093.05
F-300	Distrib. tank	capacity,m ³	5700	700	0.7	1	6400
G-401	Ventilator	m ³ /h	4200	27	0.8	200	6071.491
G-200	Ventilator	m ³ /h	4200	27	0.8	490	8032.839
H-108	Centrifuge	power, kW	37000	1200	1.2	17	72951.74
H-400	Biofilter	capacity,m ³	76000	54000	0.5	0.4	110152.6
H-201	Biogas Purification	ref.	600000			ref.	30893.62
J-101	Shredded screw conveyor	ref.	120000			ref.	6178.723
J-109	Screw conveyor	ref.	90000			ref.	4634.043
M-104	Homogen. vessel	volume,m ³	350		\$/m ³	20	7000
M-106	Ferment. vessel	volume,m ³	135		\$/m ³	6600	891000
P-103	Liquid pump	l/s	3300	48	1.2	0.1	3303.029
P-304	Liquid pump	l/s	3300	48	1.2	3.2	3493.83
P-301	Liquid pump	l/s	3300	48	1.2	1.5	3378.082
P-302	Liquid pump	l/s	3300	48	1.2	3.5	3515.836
P-112	Liquid pump	l/s	3300	48	1.2	3.7	3530.719
P-105	Screw pump	ref.	80000			ref.	4119.149
P-107	Screw pump	ref.	80000			ref.	4119.149
Q-202	Cogeneration unit	kW	1200		\$/kW	500	600000
Total Capital Cost [mln. \$]							1.90

Table 10. Total Capital Cost for non-intensified plant.

After we have total capital cost we need to obtain total investment. The extra costs like construction (10%) and engineering (10%) should be accounted as well.

$$Tot. invst. cost = Tot. cap. cost + 0.1 \times Tot. cap. cost + 0.1 \times Tot. cap. cost \quad (28)$$

$$Tot. invst. cost = 1.90 + 0.1 \times 1.90 + 0.1 \times 1.90$$

$$Tot. invst. cost = 2.28 \text{ mln. \$}$$

5.4.2 Payback period

Calculation of payback period depends on different parameters like: operating cost, profit from sales, income etc. All assumed parameters are depicted on the Table 11.

Profit from the sale of electricity & residuals (revenues from sales)	\$.year⁻¹	671,412
Operating costs excluding depreciation and interest payments	\$.year⁻¹	310,845
Direct operating costs	\$.year⁻¹	259,143
Raw materials	\$.year ⁻¹	110,574
Personal costs of employees of the operation	\$.year ⁻¹	41,362
Supervision	\$.year ⁻¹	2,068
Maintenance costs	\$.year ⁻¹	91,154
Consumables	\$.year ⁻¹	11,394
Laboratory	\$.year ⁻¹	0
Reserve	\$.year ⁻¹	2,591
Indirect operating costs	\$.year⁻¹	49,491
Insurance	\$.year ⁻¹	22,789
Corporate directions	\$.year ⁻¹	26,702
Distributional costs	\$.year⁻¹	2,211
Transport	\$.year ⁻¹	2,211
Taxable income	\$.year⁻¹	360,566

Table 11. Expenses and revenues.

Summarizing all the obtained values we are able to estimate the payback period for non-intensified plant. They are depicted on Fig. 27, 28.

From the obtained data you could see the payback period for two types of analysis: 0 % discount and 5 % discount. The 0 % discount approximate payback time at around 8 years, and the 5 % discount approximate it around 12 years.

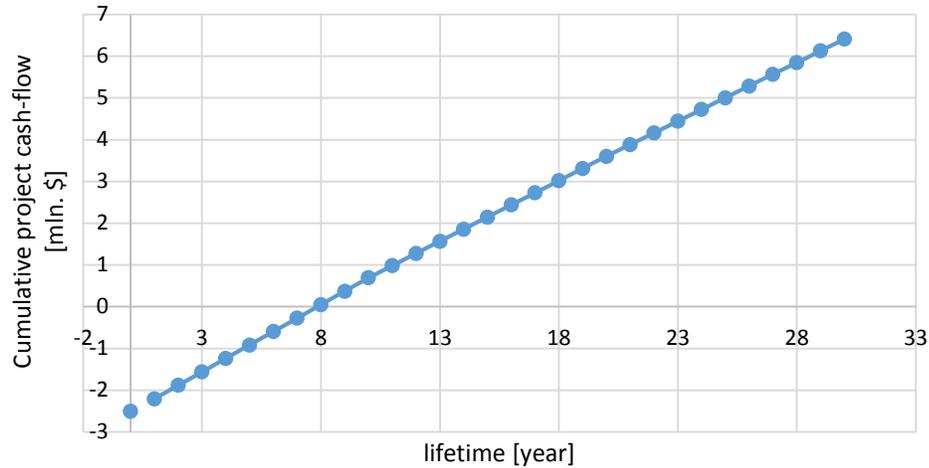


Figure 27. Payback period for 0 % discount rate.

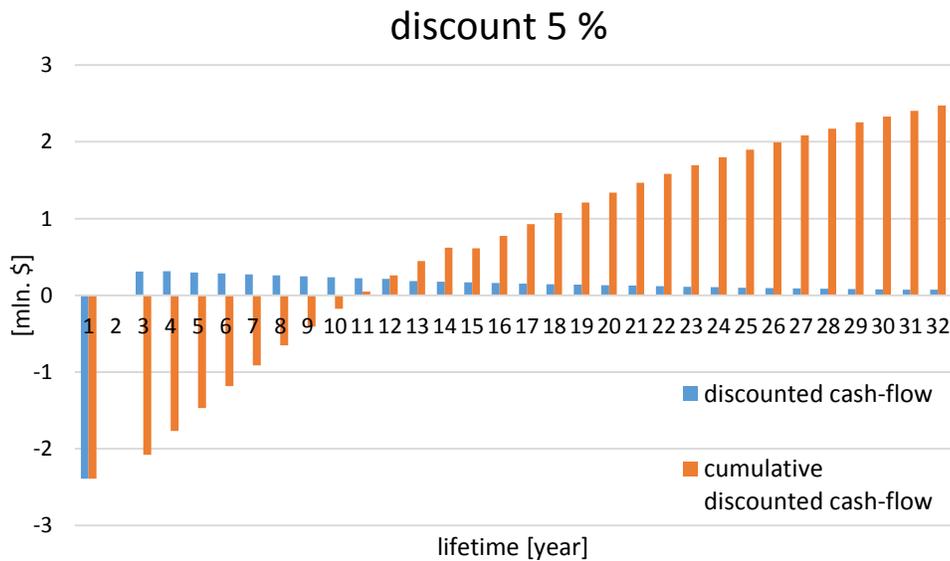


Figure 28. Payback period for 5 % discount rate.

We shouldn't forget these values are for assumption 100 % of bank credit. Nevertheless, the government gives subsidies for construction of such a plants. These subsidies vary from 30% to 60% of total investment cost. Also, the profit from sold electricity (green electricity) is 2-4 times higher [30]. With assumption of granted subsidies, only assuming 50 % of total investment cost, the payback period reduces from 12 years to 6 - 7 years, such a number you could see more likely during some literature search.

6. DISCUSSION

Quite fare and true comparison was used for both biogas plants, intensified and non-intensified. The author tried to estimate mass, energy and economic study as precise as student access allow to do that. On the Table 12 you can see the output of the work.

TITLE		NON-INTENSIFIED	INTENSIFIED
PROCESS	Substrate mass flow [$\text{kg}_{\text{TS}}\cdot\text{s}^{-1}$]	0.152	0.152
	Biogas yield [$\text{Nm}^3\cdot\text{t}^{-1}_{\text{TS}}$]	509 ± 58	633 ± 52
	Electric power [kW]	500	750
ECONOMIC ANALYSIS	Total Capital Cost (TCC) [mln. \$]	1.90	3.56
	No. of equipment	27	37
	Fermenter volume [m^3]	6600	10500
	Fermenter cost [mln. \$]	0.891	1.155
	Fermenter percentage of TCC [%]	46.9	32.4
	Pretreatment percentage of TCC [%]	–	30.3
	Total Investment Cost [mln. \$]	2.28	4.27
	Specific Investment [$\text{\$.kW}^{-1}$]	4560	5690
	Profit [mln. $\text{\$.year}^{-1}$]	0.67	1
	Payback period [year]	12	15
	Payback, including subsidies [year]	6 – 7	6 – 7
Lifetime [year]	30	30	
	ORL value [$\text{kg}_{\text{vs}}\cdot\text{m}^{-3}\cdot\text{d}^{-1}$]	2.0	1.2

Table 12. Table of important outputs.

In the beginning of “practical part” the calculations were done for intensified plant. As the repetition, the suspension was composed of wheat straw and water. The wheat straw is 5 % wt. and the rest, 95 % wt., is water. For simplification of the case, the parameters of solution was said to be parameters of water, because the solution mainly consist of water. Nevertheless, the porous wheat straw absorbs the water by its volume, and therefore makes the solution procedure more close to water. Energy estimation is crucial point in here. The author proposed such a flowsheet, which makes the plant independent of external energy supply. All the electricity demands should be covered by the energy produced in cogeneration unit. It was assumed that approximately 6 % of energy produced spent on the equipment supply.

The calculations for non-intensified were rather friendly, in comparing to previous one. The same assumptions were done as for intensified plant. However, the fractions were a bit different, 10 % wt. of wheat straw and 90 % wt. of water. After calculation of mass and energy balance, finally, the economic analysis was estimated and the payback period was found.

Total investment cost for intensified plant is 4.3 mln. \$, and the payback period is approximately 15 years. For non-intensified plant, total investment cost is 2.3 mln. \$, and payback period is 12 years. The error of such a calculation is $\pm 30\%$. For all this, the reality is different. Building of such a plant is not quite profitable business, thus the government is taking part in here and giving subsidies. As the example, green electricity (electricity produced from processing of renewable energy sources) has higher price on market in comparing with regularly generated electricity. The price vary from 2 to 4 times exciding the price of conventional electricity. Also, in developed countries the government gives opportunity for the investors, who want to build plant with green electricity on the output. They are covered by the subsidies, and usually they are vary from 30 % to 60 % of total investment cost.

In between, investment-costs per 1 m³ reactor volume showed, that for non-intensified plant is equal to 345 \$.m⁻³, and for intensified is 406 \$.m⁻³. Deublin and Steinhauser [3] proposed the range of such calculation should range in 300-500 \$.m⁻³. Another, investment-costs per 1 kW installed power should vary in 2400-6000 \$.kW⁻¹. Taking mean value 4200 \$.kW⁻¹, and comparing it to our 4560 \$.kW⁻¹, gives us a feasibility prove of performed calculation.

Going back to payback estimation, assuming 50 % subsidies only, we immediately obtain payback period in the range 6 to 7 years for both biogas plants. Numbers like these you are more likely to see during some literature search. Even there is a gap between payback periods, 15 and 12 years, with the subsidies and due to higher profit from electricity sales, intensified plant slightly, but winning the game called “shorter payback period”. We can say, including possible errors for both cases, the payback would be approximately in the same range, at this point another factor is coming, if there is a need to have more electricity for locality.

OLR is quite important parameter in designing the fermenter. It is said that the order of 2 $\text{kg}_{\text{vs}}\cdot\text{m}^{-3}\cdot\text{day}^{-1}$, is appropriate for continuous anaerobic digestion. However, if we estimate our two examined cases, intensified technology has 1.2, and non-intensified has 2 exactly. Such difference was done by the different fractions of wheat straw content. To remind you for intensified technology was used 5 % wt. of wheat straw, but for non-intensified 10 % wt. Nevertheless, the experiment data showed that, for pretreated wheat straw we can get maximum biogas yield in 40 days, for untreated wheat straw we get in 50 days. It is possible to make some assumptions for intensified plant, with optimized OLR. Reaching the order of 2 (by increasing the wheat straw fraction up to 10 % wt.), we can obtain even higher biogas yield and shorter residence time. However, these are only predictions, and for proving the appropriate experiments better be done.

Some preliminary summary could be done. Building bigger biogas plants is good investing, because, selling green electricity in big amounts would be positively reflected on payback period. Nevertheless, this is just estimations, and as for reality, the bigger plants should be investigated in order to prove the above statement.

Some suggestions for future work, the carbon dioxide is actually, also the source of methane. The content of it in biogas could exceed 50 % vol. With the Sabatier reaction we can get methane. However, the problem is where to take hydrogen? With the excess of electricity we can perform water electrolysis, in order to obtain hydrogen. Another suggest, is the fermentation of digestate. The digestate still has some amount of carbon, so why to avoid it.

7. CONCLUSION

During the work the following sections were accomplished, for non- and intensified biogas plant:

- ✓ Literature search showed, that mechanical disintegration (milling) and thermal-expansive pretreatments have a high potential, as an intensification step in currently operating or new built biogas plant. They offer a possibility to increase biogas yield by 50 % towards to untreated straw.
- ✓ Detailed flowsheet of non-intensified and intensified biogas plant were designed. For non-intensified biogas plant, 500 kW of electric power was chose, as the reference. After examining of possible released potential energy of wheat straw, which passed through pretreatments (mechanical disintegration and thermal-expansion pretreatment), the electric power was recalculated, for the same amount of wheat straw, to 750 kW.
- ✓ Mass and energy balances were calculated for each stream, suitable machines and equipment were chosen for non- and intensified biogas technologies. Description of process streams, machines and equipment dispositions are indispensable parts of flowsheet.
- ✓ Energy optimization of proposed intensified technology was done. The optimization include: recovery of biogas combustion exhausts, heated water from CHP unit on the needs of stream and human needs, vapor from batch decompression, utilization of energy stored in heated suspension.
- ✓ Based on economic analysis, capital costs, investment costs and payback periods for non- and intensified technology were determined. It was found out, that only fermenter price is 32.4 % of total capital cost for intensified technology, specific investment costs are, 4560 \$/kW for non- and 5690 \$/kW for intensified biogas plant.
- ✓ Payback period of both technologies are the same, i.e. 6-7 years with included subsidies (without green electricity even!), 12 years for non- and 15 years for intensified biogas technology without subsidies. Profit of intensified technology is 33% higher and makes it more attractive.

- ✓ Based on results, it was also found out, that intensified technology is not fully optimized. OLR of fermenter batch is equal to $1.2 \text{ kg}_{\text{vs}} \cdot \text{m}^{-3} \cdot \text{d}^{-1}$. This value is lower than recommended values, which is ranging in $2\text{-}3 \text{ kg}_{\text{vs}} \cdot \text{m}^{-3} \cdot \text{d}^{-1}$. Thus, fermenter volume is higher than preferred. Optimized OLR value will lead to decreasing of fermenter volume, lower investment costs and payback period, and, to all these, probability of higher biogas yield. Therefore, optimization of process parameters is another challenge.
- ✓ All these calculations are programmed in Microsoft Excel software. Modifying input data (biogas yield of material, required electric power), mass-energy-economic balance of new technology is immediately and easily counted up.

SYMBOLS

Basic quantities

a, b, n	Correlation factors	—
C_i^m	Concentration of i component	—
C_p	Specific heat capacity	$kJ. (kg. K)^{-1}$
C_e	Capital cost	\$
h_i	Enthalpy of i component	$kJ. kg^{-1}$
k	Overall heat transfer coefficient	$W. (m^2. K)^{-1}$
LHV	Low heating value	$MJ. kg^{-1}$
l	Liter	l
\dot{m}_i	Mass flow rate of i component	$kg. s^{-1}$
η_E	Efficiency for electricity production in cogeneration unit	—
P	Pressure	bar
P_A	Atmospheric pressure	bar
P_{EQ}	Equivalent pressure	bar
\dot{Q}_S	Supplied energy	$GJ. h^{-1}$
\dot{Q}_E	Electric power	kW
\dot{Q}	Power	kW
q_{CH_4}	Combustion heat	$kWh. Nm^{-3}$
S	Heat transfer surface	m^2
T	Temperature	$^{\circ}C$

T_{SET}	Desired temperature	$^{\circ}C$
T_0	Initial temperature	$^{\circ}C$
t	Time	s
t_{SET}	Set time	s
V	Volume	m^3
Y_{CH_4}	Methane yield	$Nm^3.t_{TS}^{-1}$
Y_{BG}	Biogas yield	$Nm^3.t_{TS}^{-1}$
ΔT_{ln}	Logarithmic mean temperature	$^{\circ}C$

Subscripts

AD	Anaerobic digestion
BOD	Biological oxygen demand
CH_4	Methane
$C:N$	Carbon:Nitrogen
COD	Chemical oxygen demand
$CSTR$	Continuous stirred tank reactor
CO_2	Carbon dioxide
$NaOH$	Sodium Hydroxide
OLR	Organic loading rate
VS	Volatile solids
WW	Waste water

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APPENDIX

- Appendix A. Flowsheet of intensified biogas plant.
- Appendix B. Mass balance of intensified biogas plant.
- Appendix C. Energy balance of intensified plant.
- Appendix D. Economic analysis of intensified plant.
- Appendix E. Flowsheet of non-intensified plant.
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