Novel integrated mechanical biological chemical treatment (MBCT) systems for the production of levulinic acid from fraction of municipal solid waste: A comprehensive techno-economic analysis

Jhuma Sadhukhan*[1], Kok Siew Ng[1] and Elias Martinez-Hernandez[2]

[1]Centre for Environmental Strategy, University of Surrey, GU2 7XH, UK

[2]Department of Engineering Science, University of Oxford, Oxford, OX1 3PJ, UK

Abstract:

This paper, for the first time, reports integrated conceptual MBCT/biorefinery systems for unlocking the value of organics in municipal solid waste (MSW) through the production of levulinic acid (LA by 5 wt%) that increases the economic margin by 110-150%. After mechanical separation recovering recyclables, metals (iron, aluminium, copper) and refuse derived fuel (RDF), lignocelluloses from remaining MSW are extracted by supercritical-water for chemical valorisation, comprising hydrolysis in 2 wt% dilute H_2SO_4 catalyst producing LA, furfural, formic acid (FA), via C_5/C_6 sugar extraction, in plug flow (210−230°C, 25 bar, 12 s) and continuous stirred tank (195−215°C, 14bar, 20mins) reactors; char separation and LA extraction/purification by methyl isobutyl ketone solvent; acid/solvent and by-product recovery. The by-product and pulping effluents are anaerobically digested into biogas and fertiliser. Produced biogas (6.4 MWh/t), RDF (5.4 MWh/t), char (4.5 MWh/t) are combusted, heat recovered into steam generation in boiler (efficiency: 80%); on-site heat/steam demand is met; balance of steam is expanded into electricity in steam turbines (efficiency: 35%).

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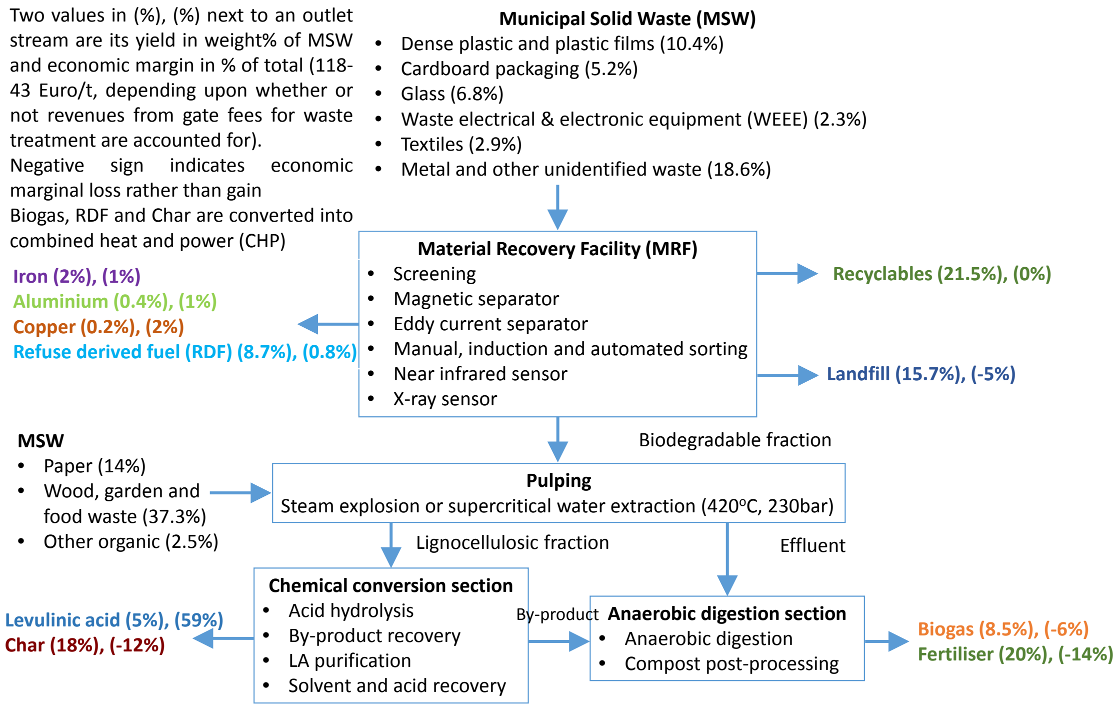
* Author/s to whom correspondence should be addressed: E-mail: jhumasadhukhan@gmail.com; Phone: +44 (0)1483 686642.

Figure for Graphical Abstract:



Two values in (%), (%) next to an outlet stream are its yield in weight% of MSW and economic margin in % of total (118-43 Euro/t, depending upon whether or not revenues from gate fees for waste treatment are accounted for).
Negative sign indicates economic marginal loss rather than gain
Biogas, RDF and Char are converted into combined heat and power (CHP)

Keywords: Circular economy; metal and energy and material resource recovery from MSW; municipal solid waste management; combined heat and power (CHP), refuse derived fuel (RDF) and compost like output (CLO); biorefinery process simulation in Aspen Plus$^{©}$, value analysis and discounted cash flow (DCF) analysis

1. Introduction

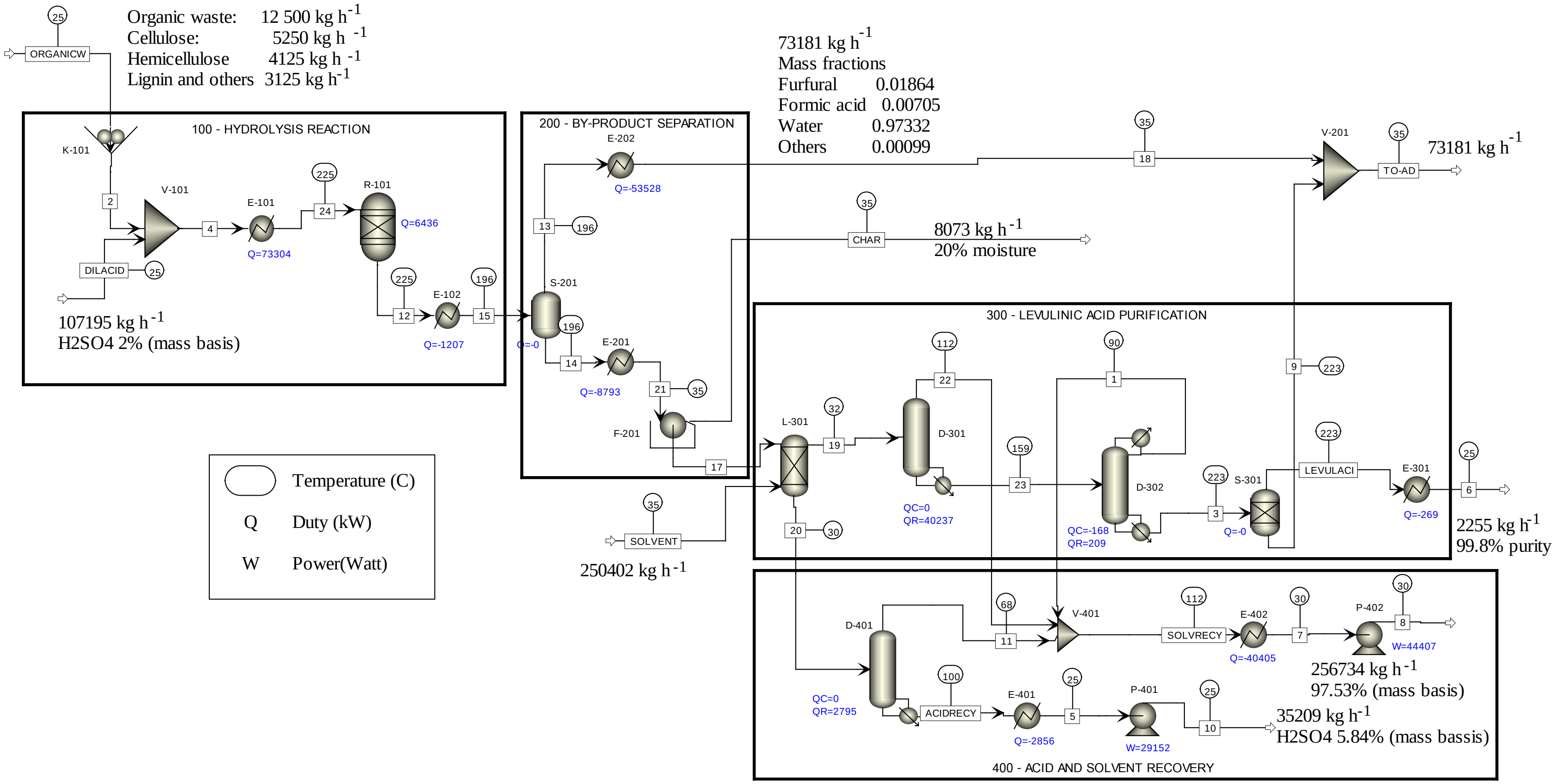
Biomass is the single source of functional organic chemicals and materials (Sadhukhan et al. 2014). Biomass is essentially made up of the same chemical elements (carbon, hydrogen and oxygen) as crude oil. This opens the possibility of producing biomass-based products that can directly replace chemically identical crude oil derivatives as well as chemically different crude oil derivatives having similar functionality. Levulinic acid (LA) falls in the second category that offers many functionalities of petrochemicals and can be a precursor to numerous added value products (Sadhukhan et al., 2014; Hayes et al., 2006). LA is a keto acid and one of the few molecules that can be used as both chemical and energy vector. The most important functional products derived from LA as a building block chemical in the decreasing order from the highest value / lowest volume to the lowest value / highest volume products are: Pharmaceutical (e.g. δ-aminolevulinic acid) > Specialty chemical (e.g. γ-valerolactone) > Agricultural (e.g. diphenolic acid) > Solvent and bulk chemical (e.g. pyrrolidones) > Chemical (e.g. succinic acid as antifreeze agent) > Fuel and additive (e.g. levulinate esters, 2-methyl tetrahydrofuran), respectively (Sadhukhan et al., 2014).

Lignocelluloses consist of cellulose (38-50%), hemicellulose (23-32%) and lignin (15-25%), and inorganic ashes and can be extracted from municipal solid waste (MSW) or urban or household waste. Paper, wood, garden, food and other organic wastes comprise the lignocellulosic or biodegradable or organic fraction of MSW. Lignocelluloses can be thermochemically degraded (Thallada and Steele, 2015) or (bio)chemically degraded (Batalha et al., 2015). Thermochemical processing, such as incineration (energy product) and gasification and pyrolysis (chemical and energy products via syngas or bio-oil (Ng and Sadhukhan, 2011a and 2011b)), allows all components' simultaneous valorisation.

Biomass pretreatment followed by biochemical degradation allows recovery of targeted molecules (Sindhu et al., 2016). Pretreatment for decomposition of biomass into cellulose, hemicellulose and lignin is needed for lignocellulosic or second generation feedstock, such as MSW, due to its heterogeneous nature. The various methods of pretreatment broadly fall into two categories: addition of extraneous agent and application of energy (Sadhukhan et al., 2016a). The former incurs higher cost of chemical and downstream separation and purification and the latter incurs higher cost of energy and capital cost of pretreatment. Hydrolysis (acid or alkali), organosolv (extraction using organic solvent) and ionic liquid extraction use extraneous agents for biomass decomposition (Mathew et al., 2016), while ultrasonication and microwave irradiation technologies make use of energy for biomass decomposition (Singh et al., 2016). Steam explosion and supercritical water extraction technologies are a flexible method for biomass decomposition, because moisture is naturally present in biomass reducing the amount of steam requirement. The process liberates hemicelluloses first because these are hydrolysed at a faster rate. Hemicelluloses consist of xylose monomers and C_5 sugars. Hemicelluloses cover the celluloses which are a linear polymeric material consisting of glucose monomers connected by β-(1–4)-glycosidic linkages. These lead to a fibrous and crystalline structure recalcitrant to hydrolysis. Due to the application of supercritical water (420^{o}C and 230 bar), β-(1–4)-glycosidic linkages are broken down, liberating glucose. Acid hydrolysis is then applied, with any of the following acids, sulphuric, hydrochloric, phosphoric, maleic and oxalic acids, in dilute condition 1-2 weight% (Morone et al., 2015). The liberated glucose can be used to target high value chemical products, e.g. 5-hydroxymethylfurfural (HMF) and LA (Mukherjee et al., 2015). These building block chemicals are referred as ‘sleeping giants’ owing to their vast potentials in the emerging bio-based economy due to their key positions in the production of biomass-derived intermediates and transition from fossil based to bio-based economy. In spite of intensive works on these two target chemicals, there is no study on their extraction from MSW. Investigation of individual catalytic conversion steps for MSW valorisation is being undertaken at laboratory scales by Fiberight LLC and CPI (Centre for Process Innovation), UK. There is no research however on how the various processes can be integrated in the form of biorefineries utilising MSW in totality for extracting resources present in MSW, metal, refuse derived fuel (RDF), chemical (e.g. LA), fertiliser and energy, i.e. combined heat and power (CHP). Resource recovery from MSW calls for integrated mechanical biological chemical treatment (MBCT) system. Process systems engineering tools enable integration between multi-processes for maximisation of energy and resource recovery efficiency, mitigation of emission and waste generation and minimisation of cost by optimal tradeoffs. Thus, to fill this gap in research, this paper offers a comprehensive process integration and techno-economic analyses of plausible biorefinery schemes of MSW. The paper proceeds by detailing the analysis methods used and describing the biorefinery process design and simulation framework, yield models of processes in MBCT systems, mass and energy analyses, cost parameters and economic value analysis methodology. Discussion of results includes identification of major parameters that decide techno-economic feasibility and independence for successful commercialization.

2. Methods

Chemical conversion section for production of LA from lignocellulosic fraction of MSW: Fig. 1 shows the Aspen Plus$^{©}$ simulation flowsheet of the chemical conversion section for the production of LA from lignocellulosic fraction of MSW. The process uses 12500 kg h^{-1} (dry basis) or 17 857 kg h^{-1} (wet basis) lignocellulosic fraction extracted from MSW, as feedstock. This moisture mainly comes from the steam inputted in the pulping process prior to the chemical conversion section. The chemical conversion section has four main sub-sections: Hydrolysis reaction (100), By-product separation (200), LA purification (300) and Acid and solvent recovery (400). Input and output mass flowrates and energy requirements of the unit operations are shown in Fig. 1. Throughout the process both the solids and fluids are handled. The stream class MIXCIPSD in Aspen Plus© is suitable to handle such streams and hence, selected in the global set up specifications. The moisture present in the organic waste feedstock is a substream of MIXCIPSD, while the rests are substreams of CIPSD, suitable to handle solids. The mass composition of the lignocellulosic fraction of MSW is as follows: cellulose 42%, hemicellulose 33% and lignin and others 25%. Cellulose, hemicellulose, lignin and char are modelled as $C_6H_{10}O_5$, $C_5H_8O_4$, $C_{7.3}H_{13.9}O_{1.3}$ and C, respectively, in Aspen Plus©. The property method used is NRTL-RK (non-random two liquids – Redlich Kwong) due to the presence of polar components. Considering the water content in the lignocellulosic fraction of MSW and the acid solution, 2% H_2SO_4 by mass are adjusted to get a reactor feed with 10% solid content. The design specifications and types and purposes of the various unit operations feature in the LA production process flowsheet are shown in Table 1. These specifications originated from engineering fundamentals of unit operations (Sadhukhan et al., 2014) are needed in Aspen Plus© simulation models to evaluate the performance of the flowsheet in terms of the product yields and raw material and energy inventories.


Organic waste: 12 500 kg h-1
Cellulose: 5250 kg h -1
Hemicellulose 4125 kg h -1
Lignin and others 3125 kg h-1
ORGANICW
25
100 - HYDROLYSIS REACTION
K-101
2
V-101
4
E-101
Q=73304
225
24
R-101
Q=6436
DILACID
25
225
12
E-102
Q=-1207
196
15
107195 kg h-1
H2SO4 2% (mass basis)
200 - BY-PRODUCT SEPARATION
E-202
Q=-53528
13
196
S-201
Q=0
196
14
E-201
Q=-8793
21
35
F-201
17
73181 kg h-1
Mass fractions
Furfural 0.01864
Formic acid 0.00705
Water 0.97332
Others 0.00099
35
18
V-201
35
TO-AD
73181 kg h-1
35
CHAR
8073 kg h-1
20% moisture
300 - LEVULINIC ACID PURIFICATION
112
22
90
1
9
223
L-301
32
19
D-301
159
23
D-302
QC=0
QR=40237
QC=-168
QR=209
223
S-301
3
Q=0
223
LEVULACI
E-301
Q=-269
25
6
2255 kg h-1
99.8% purity
20
30
35
SOLVENT
250402 kg h-1
Temperature (C)
Q Duty (kW)
W Power(Watt)
D-401
QC=0
QR=2795
68
11
V-401
112
SOLVRECY
E-402
Q=-40405
30
7
P-402
30
8
W=44407
256734 kg h-1
97.53% (mass basis)
100
ACIDRECY
E-401
Q=-2856
25
5
P-401
W=29152
25
10
35209 kg h-1
H2SO4 5.84% (mass bassis)
400 - ACID AND SOLVENT RECOVERY

Stream name	ORGANICW	DILACID	15	18	21	17	19	23	3	LEVULACI	20	ACIDRECY	SOLVRECY	CHAR	SOLVENT
To unit	K-101	V-101	S-201		F-201	L-301	D-301	D-302	S-301	E-301	D-401	E-401	E-402		L-301
From unit			E-102	E-202	E-201	F-201	L-301	D-301	D-302	S-301	L-301	D-401	V-401	V-201	
Mass Flow kg/hr															
CELLULOS	5250.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMICELL	4125.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
LIGNIN	3125.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2SO4	0.0	2143.9	2143.9	12.2	2131.7	2055.9	0.0	0.0	0.0	0.0	2055.9	2055.9	0.0	75.8	0.0
LEVULINA	0.0	0.0	2415.0	60.8	2354.2	2270.5	2263.0	2257.5	2253.0	2253.0	7.5	7.5	10.0	83.7	0.0
FORMICAC	0.0	0.0	945.0	515.7	429.3	414.1	412.7	0.0	0.0	0.0	1.4	1.4	412.7	15.3	0.0
FURFURAL	0.0	0.0	1650.0	1364.0	286.0	275.8	275.8	202.9	2.4	2.4	0.0	0.0	273.4	10.2	0.0
CHAR	0.0	0.0	6458.7	0.0	6458.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	6458.7	0.0
SOLVENT	0.0	0.0	0.0	0.0	0.0	0.0	249646.0	998.6	0.1	0.1	755.9	15.1	250387.0	0.0	250402.0
WATER	5357.1	105051.0	111439.0	71228.4	40210.7	38780.9	5603.9	0.0	0.0	0.0	33177.0	33124.9	5656.0	1429.8	0.0
Total Flow kmol/hr	297.4	5853.1	6266.2	3979.8	2286.3	2205.1	2834.8	31.5	19.4	19.4	1870.2	1859.9	2825.7	81.3	2500.0
Total Flow kg/hr	17857.1	107195.0	125052.0	73181.1	51870.6	43797.2	258201.0	3458.9	2255.5	2255.5	35997.7	35204.8	256739.0	8073.4	250402.0
Temperature C	25.0	25.0	196.0	35.0	35.0	35.0	31.7	159.3	223.5	223.5	30.4	99.6	112.1	35.1	35.0
Pressure bar	1.0	1.0	14.0	1.0	5.0	4.8	3.0	1.0	0.4	0.4	3.0	1.0	1.0	1.0	5.0

Fig. 1. Aspen Plus© simulation flowsheet for LA production process (for the Chemical section in Fig. 2).

Table 1 Design specifications and types and purposes of the various unit operations feature in levulinic acid production process flowsheet.

Unit name and model	Type or purpose of unit operation	Process specifications
K-101 Crusher	Crusher	Maximum particle diameter = 1 cm CIPSD Bond work index = 400 Inlet Particle Size Distribution (PSD) 5-10 mm: 0.05, 10-20 mm: 0.2, 20-30 mm: 0.3, 30-40 mm: 0.3, 40-50 mm: 0.1 50-60 mm: 0.05
V-101, 201, 401 Mixers	Mixing of streams	Pressure drop = 0
R-101RYield	Overall yield model for both 1st and 2nd acid hydrolysis reactors	Temperature = 225°C Pressure = 25 bar Mass yield mixed stream: Levulinic Acid 0.019648849 Formic Acid 0.00768868 Furfural 0.013424679 Water 0.906688308 Char (CIPSD) 0.052549483 Inerts: H_2SO_4

S-201 Flash2	2-phase flash separator of organic vapours and acidic liquid phase	Temperature = 196ºC Pressure = 14 bar
F-201 Filter	Filter to separate char formed by the humins	Max pressure drop = 0.2 bar Revolutions = 1200 rpm Solids mass fraction in cake = 0.8
L-301 Extract	Solvent extraction of levulinic acid using methyl isobutyl ketone (MIBK)	Number of stages = 10 Key components: 1st liquid phase: Water 2nd liquid phase: Solvent (MIBK) Pressure at stage 10: 3 bar
D-301 RadFrac	Solvent stripper column	Number of stages = 10 Condenser = NONE Top pressure = 1 bar Distillate rate adjusted for a solvent recovery by 0.996 (by mass fraction)
D-302 RadFrac	Levulinic acid distillation	Number of stages = 10 Condenser = TOTAL Feed stage = 5 LA purity = 99.8% by mass Bottom flow rate adjusted for a LA recovery by 0.998 (by mass fraction) Pressure = 0.4 bar
S-301 Sep	Separator removes any residual solid impurities	Split fractions (on mass basis): Levulinic acid 1, Formic acid 0.999 Furfural 0.999, Char 0, Solvent 0.999
D-401 RadFrac	H_2SO_4 recovery column	Number of stages =10 Condenser = NONE Top pressure = 1 bar Distillate rate adjusted for a solvent recovery by 0.98 by mass fraction
P-401 Pump	Pumping acid solution to reactor	Outlet pressure = 25 bar Efficiency = 0.8
P-402 Pump	Pumping solvent to extraction column	Outlet pressure = 5 bar Efficiency = 0.8

The chemical conversion section begins with feedstock shredding to reduce particle size to 0.5–1 cm. The particles are then conveyed by a high pressure steam injection system to a mixing tank. The feedstock is mixed with dilute sulphuric acid (concentration of 2 weight%) and pumped into the acid hydrolysis reactors. A RYield model is used to simulate the reactor system to set the production limits as shown in Table 1 take account of the following conversion rates (Hayes, 2009):

a) 46% LA, 18% FA and the balance being primarily char of the initial cellulose mass: 2415 kg/h LA and 945 kg/h FA.

b) 40% furfural, 35% char and the balance being primarily water of the initial hemicellulose mass: 1650 kg/h furfural and 1443.75 kg/h char.

c) Majority of the initial lignin mass being char in the product stream.

In addition, sulphuric acid and water entered to the system are also present in the outlet product of the acid hydrolysis process, i.e. stream 15 in Fig. 1.

Conditions in the flash separator correspond to the second reactor of the acid hydrolysis reaction system, 196°C and 14 bar. At these reactor conditions, furfural, FA and water are flashed into the vapour phase. The liquid outlet from the flash separator is cooled down before entering the filter to separate the cake containing the char, tar and remaining solids from the LA stream. A press filter can be used to obtain a cake with 80% solid content by mass, with the rest being moisture. Then, the LA is extracted from the filtrate liquid using Methyl isobutyl ketone (MIBK) as solvent in an extraction column having 10 stages. The solvent mass flowrate is adjusted to recover pure LA at the end. The solvent is then stripped in a column with 10 stages to separate LA. The column operates at the atmospheric pressure and the distillate rate can be adjusted for a mass fractional recovery of solvent by 0.996. Due to lower volatility, LA is obtained in the bottom stream.

This is then refined in a distillation column with 10 stages and the bottom mass flowrate is maintained to recover pure LA. This column operates at vacuum (0.4 bar) to avoid LA decomposition. An additional separator is placed just to make sure that the purified LA stream does not contain any other organic residues. The organic residual stream having a high chemical oxygen demand (COD) is a suitable feedstock for anaerobic digestion (AD). The acidic stream resulting from the extraction column passes through a column to recover the acid in the bottom and any remaining solvent in the distillate. This acid recovery column has 10 stages. Its distillate rate is adjusted for a mass fractional recovery of solvent by 0.98. The solvent from the acid recovery column and the solvent stripping column are combined. This recovered solvent at 112°C is cooled down and pumped at 5 bar for recycling back to the extraction column. The recovered sulphuric acid is pumped at 25 bar for recycling back to the reactor. The purified LA at 223°C is cooled down to room temperature for storage.

Fig. 1 also shows the streams' analyses in terms of mass flowrates, connections between unit operations and temperature and pressure conditions in the LA production flowsheet. It can be observed that 2255 kg/h of LA are produced at 99.8% purity from 12500 kg/h of lignocellulosic fraction of MSW (dry basis). The quantity of sulphuric acid entered the reactors under the set of conditions, is 2144 kg/h, but 2056 kg/h can be recovered, thus only 88 kg/h of pure sulphuric acid makeup is needed by the system. Similarly, the amount of solvent required by the extraction process is 250402 kg/h, but 250392 kg/h can be recovered, thus only 9.3 kg/h of solvent makeup is needed by the system. The total heat requirement by the process is 123 MW, while the cooling duty is 107.2 MW. Note that the heat or steam demand also includes the share by the pulping process, as the lignocellulosic fraction of MSW contained some steam injected in the pulping process.

However, these utility requirements can be reduced to 62 MW of heating and 46 MW of cooling (cooling water), respectively, by heat integration assisted by pinch analysis. Table 2 lists the heat exchangers present in Aspen Plus© simulation flowsheet, their hot and cold streams' names, supply and target temperatures, duties and the type of exchangers (process to process heat recovery or duty supplied by hot or cold utility). The heat integration approach is detailed elsewhere (Sadhukhan et al., 2014).

Table 2

CHP system: The energy in the char can be used to provide CHP required by the process. The CHP process configuration consists of a biomass boiler that is an integrated unit with biomass fuel combustion zone, steam drum and heat recovery steam generator inside the unit, and steam turbines (Wan et al., 2016). In the case of the MBCT system, the CHP system can be a total site utility system, providing electricity and heat to the material recovery facilities (MRF), pulping and chemical conversion and AD sections. The fuels to the CHP system also include RDF from MRF, biogas from AD and char from the chemical conversion section. The fuel is combusted in the combustion zone and the heat of combustion from the resulting flue gas is recovered into high pressure steam generation in the steam drum and heat recovery steam generator. The energy efficiency of the biomass boiler from fuel to steam is at least 80% (Sadhukhan et al., 2014). A part of the steam is used to supply the heat demand by the MBCT system. The residual steam is then expanded into electricity generation in back pressure and condensing steam turbines. The energy efficiency of the steam turbines from steam to electricity generation is 35% (Sadhukhan et al., 2014). The boiler feed water (BFW) after electricity generation from the steam turbines is returned to the biomass boiler. Any excess electricity after fulfilling the demand by the MBCT system can be exported.

MRF system: The state-of-the-art MRF are configured to recover recyclables, metals and RDF from MSW. MRF consist of mechanical unit operations: screening, magnetic separator, Eddy current separator, manual, induction and automated sorting, near infrared sensor, X-ray sensor, etc. Individual mechanical unit operations needed for resource recovery from various streams of urban waste are illustrated in Fig. 2. Usually, source separated MSW consists of the following streams diverted into various lines for recycling: paper and cardboard packaging; glass; dense plastic and plastic films (container, plastic packaging); wood, garden and food waste; textiles; WEEE (waste electrical and electronic equipment). Other than these, metals and unidentified wastes are present in these streams. Also, the source segregation is not perfect; hence, MRF are essential for recycling these materials back to value chains. The constituents of MSW on the basis 1 kt/d or 45.6 t/h (8000 operating hours per year of MRF) are shown in Table 3.

Paper and cardboard packaging are separated after conveying by air classifier fitted with a digital camera and a weighing machine; the air flowrate is adjusted to separate paper and cardboard packaging according to their images and weights, into two separate compartments and bailed for transporting to milling sites. Alternatively, paper and cardboard packaging may not need to be separated, but can be used as mixed substrates in pulping process for recovering organic fraction for conversion into fuel or chemical. The latter option is evaluated for techno-economic feasibility in this study.

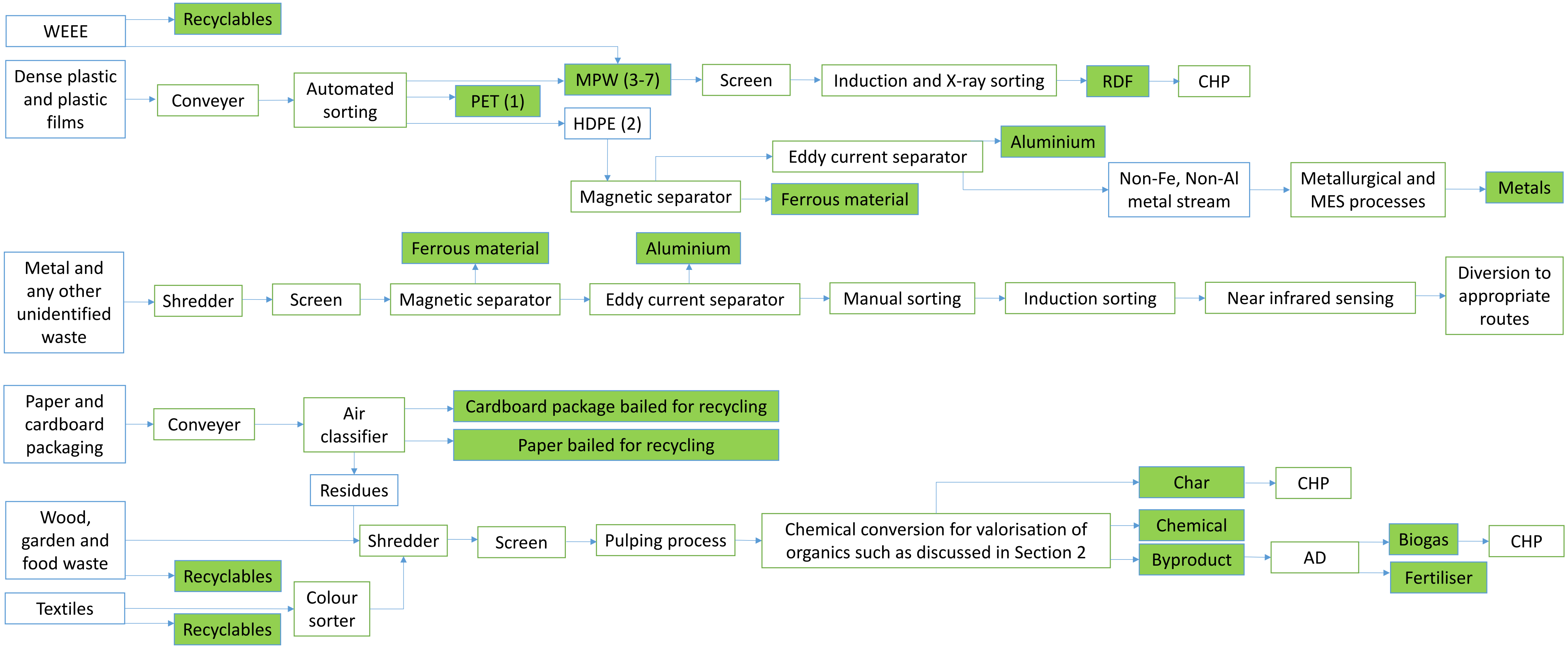


Fig 2. MBCT plant configuration for deep valorisation of MSW into recyclables, PET (1), metals, biogas, RDF and char to CHP, building block chemicals and fertiliser. Potential end products with market demand and prices are shown in shaded blocks. The chemical section is illustrated in Fig. 1.

Table 2

Heat exchanger names, duties, hot and cold streams' names, supply and target temperatures and type of exchangers (process to process heat recovery or duty supplied by hot or cold utility) deduced by heat integration within the chemical section for LA production.

Exchanger name	Enthalpy or duty (MW)	Hot stream	Supply temperature (°C)	Target temperature (°C)	Cold stream	Supply temperature (°C)	Target temperature (°C)	Type (Process heat recovery or Hot or Cold utility)
E-101A	29.5	SOLVRECY	109	49	4 - Reaction mixture	25	99	Process heat recovery
E-101B	28.9	13	196	109	4 - Reaction mixture	99	172	Process heat recovery
E-101C	21.3	STEAM	250		4 - Reaction mixture	172	225	Hot utility
E-102	1.2	12 - reactor effluent	225	196	COOLING WATER	20	40	Cold utility
E-201A	2.0	14	196	160	COOLING WATER	20	40	Cold utility
D-401 REBOILER	2.8	14	160	109	D-401 Bottoms liquid A	99	100	Process heat recovery
E-201B	4.0	14	109	35	COOLING WATER	20	40	Cold utility
E-202	24.6	13	109	35	COOLING WATER	20	40	Cold utility
D-301 REBOILER	40.2	STEAM	250		D-301 Bottoms liquid	158	159	Hot utility
D-302 REBOILER	0.2	STEAM	250		D-302 Bottoms liquid	222	223	Hot utility
D-302 CONDENSER	0.2	D-302 distillate vapour	91	90	COOLING WATER	20	40	Cold utility
E-301	0.3	LEVULACI	223	25	COOLING WATER	20	40	Cold utility
E-401	2.9	ACIDRECY	100	25	COOLING WATER	20	40	Cold utility
E-402A	1.5	SOLVRECY	112	109	COOLING WATER	20	40	Cold utility
E-402B	9.4	SOLVRECY	49	30	COOLING WATER	20	40	Cold utility
Total Cold Utility	46	MW						
Total Hot utility	62	MW						

The stream containing dense plastic and plastic films (container, plastic packaging) after conveyance is separated by automated sorting system employing various types of sensing systems into three streams: Al cartons with HDPE (high density polyethylene) (according to the numbering of plastic, it is numbered as 2), PET (polyethylene terephthalate numbered as 1) and mixed plastic waste (MPW numbered as 3-7). Magnetic and Eddy current separators are used downstream to Al cartons with HDPE stream to first isolate ferrous and non-ferrous streams and then to separate Al cans from the non-ferrous stream. Other streams if manually or automatically detected to be containing Al are also diverted to the Eddy current separator. An 'eddy current' occurs when a conductor is exposed to a changing magnetic field. It is an electromagnetic way of dividing ferrous and non-ferrous metals. MPW can also be recovered from WEEE. This stream with low PVC (poly vinyl chloride) can be further screened to remove traces of metals and recover polymer in purer form to give rise to RDF. RDF is used as an alternative to fossil fuel, specifically coal. It uses materials which are not otherwise possible to recycle. To make RDF useful in industrial incineration and energy generating plant, it is important to ensure the quality of RDF, when it comes to heating values, ingredients, and contaminants like metals, stones and chemicals. Therefore, in some plants, induction sorting systems and x-ray sorting systems are installed to detect and remove these components (Kohaupt, 2009; Capel, 2008). In induction sorting, material is sent along a conveyor belt with a series of sensors underneath. These sensors locate different types of metal which are then separated by a system of fast air jets which are linked to the sensors. X-rays can be used to distinguish between different types of materials based on their density.

Wood, garden and food wastes are the primary source of organics. This mixed stream can be treated by steam explosion or supercritical hot water extraction, called pulping process that separates the curbside-type recyclables from the lignocellulosic fraction of MSW. The lignocellulosic fraction of MSW goes through a primary wash for ash removal and cellular disruption for yield maximization combined with a sterilization stage – fractionation of this lignocellulosic fraction of MSW is then carried out by the controlled acid hydrolysis process for eventually producing LA in the chemical conversion section, simulation of which has been discussed earlier.

The effluent containing residual organics can be a substrate in AD process producing biogas and fertilizer, with yields and environmental incentives in optimal set of indicators discussed elsewhere (Sadhukhan, 2014). For more advanced and intense valorization, emerging technologies such as microbial electrosynthesis (MES) can be applied for further recovery of organics from the effluent. The process is versatile in terms of the ability to process mixed stillage streams, containing metals, organics, inorganics, (Nancharaiah et al., 2015) e.g. stillage streams from MRF, into the recovery of metals, bioplastics, biofuel and biochemical (Sadhukhan et al., 2016b).

Near infrared sensors (NIR) are used further downstream for the recovery of any lost metal in the MRF, before any unrecovered waste may be discarded. In NIR, materials are illuminated they mostly reflect light in the near infrared wavelength spectrum. The NIR sensor can distinguish between different materials based on the way they reflect light.

The new configuration in Figs. 2-3 essentially employs a novel concept of MBCT system, including MRF liberating recyclables, metals and RDF from MSW, followed by a pulping process for extracting lignocellulosic fraction of MSW, a chemical conversion section sproducing LA from the lignocellulosic fraction of MSW, a biological section to release fertiliser and biogas from the effluent and a total site CHP system utilising RDF, char and biogas. Table 3 shows yields and quantities of constituents in recyclables and MSW free of recyclables; the latter is aggregation of RDF, metal and chemical conversion section and AD feedstock streams, constituents of which are also shown in the table. Recyclables are recovered at relatively lower efficiencies compared to MRF (that can achieve upto 95% recovery efficiency), because intensive valorization of MSW is the aim of this work for economic independence of future generations MBCT systems. Additionally, this concept can address the challenge of declining demand and profitability of the milling operations that give rise to a greater quantity of lignocellulosic waste. An MBCT system can be a cluster across various locations; in which case transportation cost has to be added to the cost of production precursors moving to various locations. The recovery efficiencies on mass basis are WEEE: 86%; glass: 69%; paper and cardboard packaging: 54%; metals: 27%; textiles: 17%; dense plastic and plastic film: 17% and other materials: 8%, respectively (DEFRA, 2015). The balance of each stream is subjected to intensive valorisation.

Table 3

MSW composition on mass basis; separated into recyclables and MSW free of recyclables; the latter is further recovered into RDF, metal stream, chemical and AD sections' feedstocks and the balance goes to landfill.

	MSW	Recyclables	MSW free of recyclables	RDF	Metal stream	Chemical feedstock	AD feedstock	Landfill
Food waste	170.0		170.0			85.0	85.0	
Garden waste	165.0		165.0			132.0	33.0	
Other waste	149.0	12.2	136.8					136.8
Paper	140.0	75.9	64.1			64.1		
Glass	68.0	47.1	20.9					20.9
Dense plastic	66.0	11.0	55.0	55.0				
Card packaging	52.0	28.2	23.8			23.8		
Plastic films	38.0	6.3	31.7	31.7				
Wood	38.0		38.0			38.0		
Metals	37.0	9.9	27.1		27.1			
Textiles	29.0	5.0	24.0			23.5		0.6
Other organic	25.0		25.0			25.0		
WEEE	23.0	19.8	3.2					
Total, t/d	1000.0	215.4	784.6	86.7	27.1	391.4	118.0	158.3
Total, t/h	45.6	9.8	35.8	4.0	1.2	17.9	5.4	7.2

Policy relevance: The European Parliament and the Council of the European Union have regulations to limit the heavy metal emissions to the atmosphere, water and land. The EC Regulation No. 166/2006 sets the thresholds for releasing zinc to the atmosphere, water and land are 200, 100 and 100 kg/year, respectively. For copper and chromium, these thresholds are 100, 50 and 50 kg/year, respectively. For nickel, the thresholds are 50, 20 and 20 kg/year respectively. The Directive on the Limitation of Emissions of Certain Pollutants into the Air from incinerators (2001/80/EC) - has acted to limit heavy metal emissions via dust control and absorption of heavy metals. For MBT plants, effective extraction of valuable metals from wastes is urgently needed to meet or exceed the regulations for higher environmental goals. In addition to specific unit operations for metal separation and recovery as discussed earlier (Kohaupt, 2009; Capel, 2008), unit operations used in metal and mineral industry (e.g. pyrometallurgical and hydrometallurgical techniques globally applied at 70% and 30%, respectively) can be employed for recovery of certain metals, e.g. zinc, copper, nickel, chromium, etc.

Value analysis methodology: Value analysis (Martinez-Hernandez et al., 2014; Sadhukhan et al., 2008; 2004; 2003) has been adopted in this study to evaluate the economic margins of individual products recovered in the MBT system. Aggregation of economic margins of all output flows gives the overall economic margin of the system. Thus, maximising positive economic margins of profitable products and minimising or eliminating negative economic margins of non-profitable products and outlet streams can ensure overall highest economic margin of the system. Economic margin of a stream i, EM_i is calculated by multiplying the flowrate of the stream, F_i with the difference between its value on processing (VOP) and its cost of production (COP), shown in equation 1. The unit of F is t/h and that of COP and VOP is Euro/t and EM is Euro/h.

$$EM_i = F_i \times \left(VOP_i - COP_i\right) \tag{1}$$

As defined by Martinez-Hernandez et al. (2014), the VOP of a stream is the prices of products that are ultimately produced from it, subtracted by the costs of auxiliary raw materials, utilities and annualised capital cost of equipment that contribute to its further processing into these final products.

The COP of a stream is the summation of all associated cost components, i.e. the costs of feedstocks, auxiliary raw materials, utilities and annualised capital cost that have contributed to the production of the stream. This means that only those fractional costs involved with the stream's production are included in its COP.

Further, the concise equations 2-3 for representation of VOP and COP of a stream are given. VOP of a feed f to a process unit k is calculated from the known values of the product streams p and the total costs of the process unit k, shown in equation 2.

$$VOP_f = \left[\sum_{p=1}^{q} VOP_p \, P_p - \bar{O}_k\right] / \sum_{f=1}^{g} F_f \tag{2}$$

where q is the number of products (excluding emissions / wastes), g is the number of feedstocks considered as main material streams (excluding auxiliary raw materials). P_p and F_f correspond to the mass flowrates of product and feedstock, respectively.

COP of a product p from a process unit k is calculated from the known prices or costs of the feed streams f and the total costs of the process unit k, shown in equation 3.

$$COP_f = \left[\sum_{f=1}^{q} COP_f \, P_f + \bar{O}_k\right] / \sum_{f=1}^{g} F_f \tag{3}$$

Capital cost consists of direct and indirect capital costs. The direct capital cost comprises the costs of equipment, installation, instrumentation and control, piping, electrical systems, building, yard improvements and service facilities. The total capital cost is the summation of direct costs, indirect costs and working capital, a total CAPEX of 5.03 times the delivered cost of equipment for a solid-fluid processing system (Sadhukhan et al., 2014). An annual capital charge of 13% corresponding to a discount rate of 10%, a plant life of 15 years and a start-up period of 2 years (capital expenditures of 25% and 75% on the 1st and 2nd year) is then applied to the total CAPEX, to estimate the annual capital cost. The delivered cost of equipment can be estimated using cost and size correlation, shown in equation 4, at first, and thereafter updating that cost from reported year to the current year, by applying the Chemical Engineering Plant Cost Index (CEPCI), shown in equation 5.

$$NEW\ COST\ AT\ THE\ BASE\ YEAR = BASE\ COST \times \left(\frac{NEW\ SIZE}{BASE\ SIZE}\right)^{SCALING\ FACTOR} \quad (4)$$

$$NEW\ COST\ AT\ THE\ CURRENT\ YEAR = \ NEW\ COST\ AT\ THE\ BASE\ YEAR \times \frac{CEPCI\ AT\ THE\ CURRENT\ YEAR}{CEPCI\ AT\ THE\ BASE\ YEAR} \quad (5)$$

Table 4 shows the base sizes, base costs, estimated scaling factors, base or reporting years and CEPCI at the base years of the various process units in the MBCT system, and thus, the estimation of the delivered cost of equipment, total CAPEX and annual capital cost. The recent most year for cost update is taken 2015, when CEPCI has been stabilised at 576.73. The delivered cost of equipment is then multiplied by 5.03 to obtain the total CAPEX (Sadhukhan et al., 2014), which is then factored by the annual capital charge (0.13 in this case) to obtain the annual capital cost. For shredder, screen, magnetic and Eddy current separators and manually sorting cabin, the scaling factors shown in Table 4 are calculated using the parameters given in (Āriṇa et al., 2014); for induction sorting and near infrared sensors, the scaling factors are calculated using the values given by 4R Sustainability, Inc. (2011). The capital cost parameters in Table 4 for the pulping, AD and chemical conversion sections are collated from the work by Sadhukhan et al. (2014).

Table 4

Parameters used for estimation of delivered cost of equipment and the estimated delivered cost of equipment, total CAPEX and annual capital cost of each unit in the MBCT system.

	Base size (t/h)	Base cost (million Euro)	Scaling factor	Year	CEPCI of base year
MRF with CHP					
Shredder	10	0.27	0.60	2014	576.10
Screen	10	0.16	0.97	2014	576.10
Magnetic separator	10	0.06	0.58	2014	576.10
Eddy current separator	10	0.12	0.33	2014	576.10
Manually sorting cabin	10	0.12	0.19	2014	576.10
Induction sorting	7	0.28	0.81	2011	585.70
Near infrared sensors	1.8	0.08	0.94	2011	585.70
RDF CHP	2.23	0.38	0.61	2002	395.60
Pulping section	83.3	1.41	0.78	2003	402.00
AD section with CHP					
Anaerobic digestion	12.5	11.62	0.92	2005	468.20
Biogas CHP	2.2	0.38	0.61	2002	395.60
Compost post-processing	6	0.05	0.44	2007	525.40
Chemical section with CHP	4	11.28	0.78	2003	402.00

	Size (t/h)	Delivered cost of equipment (million Euro)	Total CAPEX (million Euro)	Annual capital cost (million Euro/y)
MRF with CHP				
Shredder	35.80	0.58	2.94	0.38
Screen	35.80	0.55	2.77	0.36
Magnetic separator	35.80	0.13	0.63	0.08
Eddy current separator	35.80	0.18	0.92	0.12
Manually sorting cabin	35.80	0.15	0.77	0.10
Induction sorting	35.80	1.04	5.23	0.68
Near infrared sensors	35.80	1.32	6.64	0.86
RDF CHP	3.95	0.80	4.00	0.52
Pulping section	23.39	0.75	3.78	0.49
AD section with CHP				
Anaerobic digestion	13.04	14.89	74.88	9.73
Biogas CHP	3.91	0.79	3.98	0.52
Compost post-processing	9.13	0.06	0.30	0.04
Chemical section with CHP	18.00	54.44	273.84	35.60
Total		75.68	380.69	49.49

The annual capital cost of a unit must be added to its annual operating cost to obtain the total annual cost of the unit. The annual operating cost consists of the fixed and variable (raw materials and utilities) costs. The parameters for estimating fixed operating costs such as maintenance, laboratory, supervision and plant overheads, etc. are given in Sadhukhan et al. (2014). A brief overview of the correlations to calculate the various operating cost items (Sadhukhan et al. 2014) is as follows.

Fixed operating cost items are as follows 1-3.

1. Costs of maintenance, capital charges, insurance, local taxes and royalties = 24% of indirect capital cost
2. Personnel cost = 0.595 million Euro/100 MWth LHV (low heating value)
3. Laboratory, supervision and plant overhead costs = 90% of personnel cost

Direct Production Cost (DPC) is then calculated as the summation of the variable and fixed operating costs: DPC = Variable operating cost (e.g. raw materials and utilities, etc.) + Fixed operating cost. The DPC is then increased by 30% (or 1.3 times the DPC) to account for miscellaneous items: sales expense, general overheads and research and developments. Fig. 3 illustrates the results of value analysis using the following bases. Revenues from recyclables have not been accounted for in the value analysis. Hence, the starting feedstock is the MSW without recyclables, 35.8 t/h.

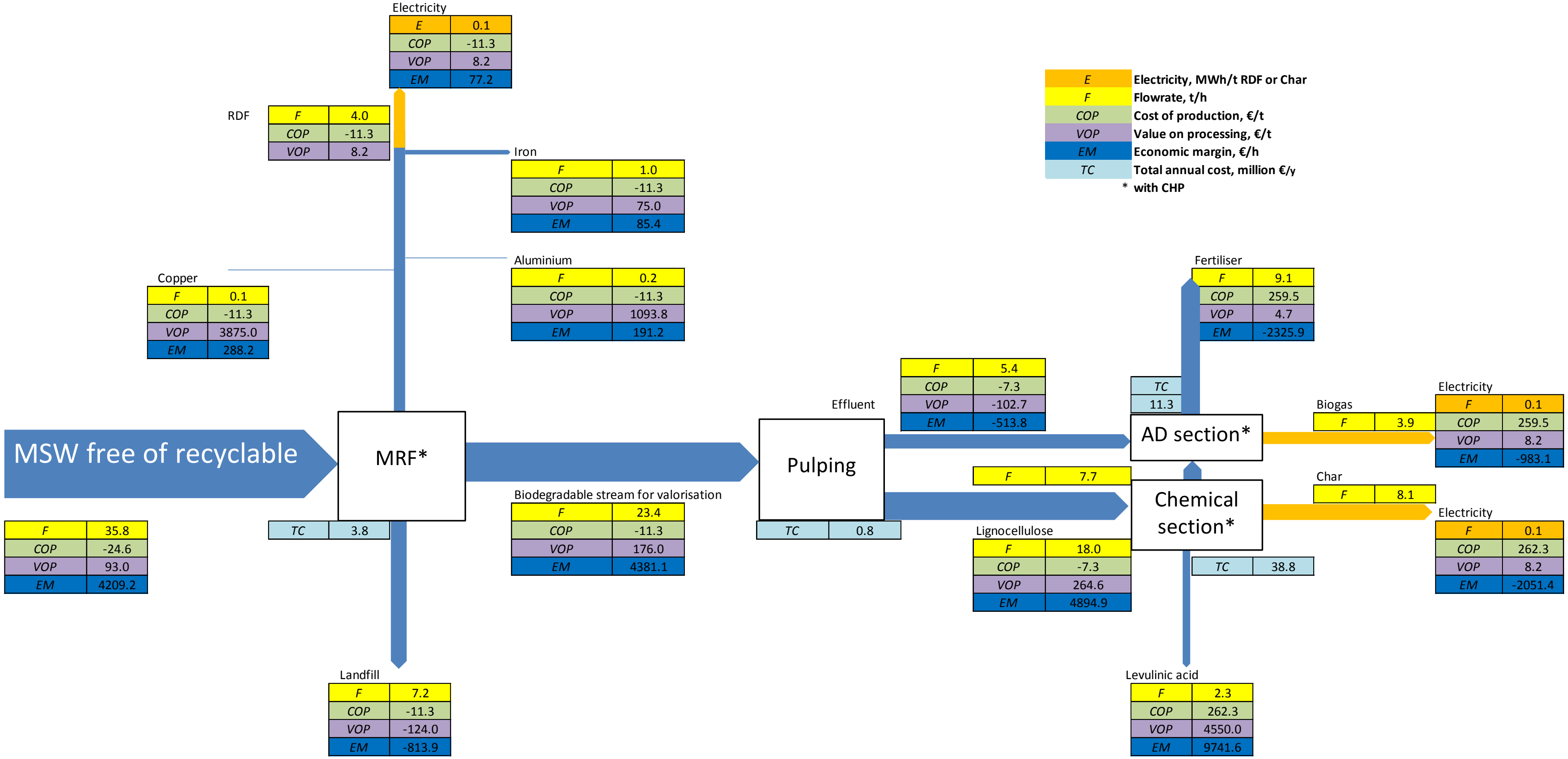


Fig. 3. Value analysis of MBCT system.

Energy balance of the MBCT system: As discussed earlier, RDF from MRF, biogas from AD and char from the chemical conversion section are the combined fuel to the boiler in the total site CHP system (Wan et al., 2016). Char from the chemical conversion section also has minor quantities of LA, FA and furfural (Char in Fig. 1). Their calorific values (CVs) are 18, 5.4 and 16 MJ/kg, respectively. The CV of pure char is 20 MJ/kg or 5.6 MWh/t. Water present in the char is an inert. Hence, by taking account of the CVs of pure char, LA, FA and furfural, according to their mass compositions (Char in Fig. 1), the CV of the overall char stream (8.0734 t/h) becomes 16.2 MJ/kg or 4.5 MWh/t. The CVs of biogas and RDF are: 23 (Sadhukhan, 2014) and 19.5 (Sadhukhan et al., 2014), MJ/kg or 6.4 and 5.4, MWh/t, produced at 3.9122 (Sadhukhan, 2014) and 3.9535 t/h, respectively. Thus, the following energy input and output calculations for the total site CHP system can be performed:

$$Energy\ input\ to\ biomass\ boiler = \frac{(16.2\times8073.4+23\times3912.2+19.5\times3953.5)}{3600} = 82.7\text{ MW}$$

$$Steam\ generation\ from\ biomass\ boiler = 82.7\times0.8 = 66.2\text{ MW}$$

$$Hot\ utility\ (steam)\ demand\ by\ the\ pulping\ and\ chemical\ conversion\ section$$
$$= 62\text{ MW}$$

$$Net\ steam\ available\ for\ electricity\ generation\ via\ steam\ turbines = 66.2-62$$
$$= 4.2\text{ MW}$$

$$Electricity\ generation\ from\ steam\ turbines = 4.2\times0.35 = 1.5\text{ MW}$$

$$Electricity\ demand\ by\ MRF = 10\text{ kWh/t MSW}$$

$$Electricity\ demand\ by\ MRF = \frac{10\times35.8}{1000} = 0.4\text{ MW}$$

$$Net\ electricity\ export = 1.5-0.4 = 1.1\text{ MW} = \frac{1.1}{(8.0734+3.9122+3.9535)}$$
$$= 0.07\text{ MWh/t}$$

$$Market\ price\ of\ electricity = 0.118\ \text{Euro/kWh}\ \ \text{(DECC 2015)}$$

$$Price\ of\ net\ electricity\ export\ per\ unit\ mass\ of\ fuel = 0.118 \times 0.07 \times 1000$$

$$= 8.2\ \text{Euro/t}$$

The net electricity export (0.07 MWh/t) and price of net electricity export per unit mass of fuel (8.2 Euro/t) are applied to each fuel type, char, biogas and RDF for the value analysis calculations.

Total annual costs of various sections in the MBCT system: Table 4 gives the estimated delivered cost of equipment, total CAPEX and annual capital cost of each unit in the MBCT system. Total annual cost of MRF with CHP section = 3.8 million Euro / y (Fig. 3). The delivered cost of equipment is 4.75 million Euro at the recent most year corresponding to the CEPCI of 576.73 in 2015 for a mass flowrate of 35.8 t/h MSW free of recyclables (Table 4). The annual capital cost (3.11 million Euro / y) is then obtained by multiplying 4.75 by 5.03 (the total capital cost is 5.03 times the delivered cost of equipment) and 0.13 (annual capital charge), as discussed earlier. These factors are applied to calculate the annual capital cost of each section in the MBCT system. The operating costs are contributed by the personnel, fixed and miscellaneous items, 0.35, 0.19 and 0.16 million Euro / y, respectively. There is no additional utility cost for this section, because its electricity demand is met by the total site CHP system. For the pulping section, the annual capital cost and personnel, miscellaneous and fixed operating costs are 0.49, 0.17, 0.06 and 0.03 million Euro / y, respectively. As discussed earlier, its demand for steam is met by the total site CHP system. Total annual cost of chemical conversion section with CHP = 38.8 million Euro / y (Fig. 3), 92% of which are due to the annual capital cost. The cost of makeup sulphuric acid and solvent is negligible. As discussed earlier, its demand for steam is met by the total site CHP system. Total annual cost of AD section with CHP = 11.3 million Euro / y (Fig. 3), 91% of which are due to the annual capital cost. 12.9 t/h AD feedstock are made up of 5.4 t/h effluent from the pulping process and 7.5 t/h effluent from the chemical conversion section. Thus, the various sections in the MBCT system in the decreasing order of their capital investments are Chemical conversion section > AD section > MRF > Pulping, respectively.

Cost of MSW: An average waste collection fee of 84.5 Euro/t MSW is paid by the MBCT plant owner to the local authority (Hogg, 2002). At the same time, the MBCT plant owner is eligible to receive a gate fee from the local authority, for treating MSW. This rate is 109.12 Euro/t MSW (WRAP, 2015). Therefore, the *COP* of MSW is estimated to be (84.5 – 109.1) = −24.6 Euro/t. This implies that the current business model allows 24.6 Euro/t revenue guaranteed for the MSW treatment plant owner. This is a strong economic incentive for waste valorisation and thereby mitigation of environmental impacts of wastes and landfilling. Valorisation of organics of MSW into chemicals seeds the route to economic independence of MSW treatment systems. For e.g. if MSW is priced at 50 Euro/t, the economic margin of the integrated MBCT system still stays positive at 43 Euro/t. Hence, by integration of chemical conversion section to an MBT system, the need for gate fee allowance from the local authority to the MSW treatment industry can be eliminated.

Cost of residues to landfill: The median rate of gate fee for landfilling (including Landfill Tax) in 2014/2015 at 124 Euro/t is specified (WRAP, 2015).

Value of LA: Market price of LA is 4550 Euro/t.

Value of metals: The metal containing stream has 80% iron, 14% aluminium and 6% copper, respectively. Their prices in £/t metal vary 40-80, 250-1500 and 2600-3600, respectively (ScrapSales UK). So, an average price is assumed to create the base case in Euro/t (assumed 1 Euro = £0.8): 75, 1093.8 and 3875, respectively.

3. Results and discussions

The results of value analysis using the bases discussed in the earlier section are shown in Fig. 3. The MBCT system produces excess electricity from the total site CHP system utilising RDF, biogas and char; metals: iron, aluminium and copper; and LA. In Fig. 3, the cost of the MSW free of recyclables feedstock is −24.6 Euro/t as it enters the MBCT plant at 35.8 t/h. The first operation it undergoes is MRF, which incurs total annual capital and operating cost of 3.8 million Euro/y. Then, the COP of the outlet streams from this section, i.e. iron, aluminium, copper, electricity, biodegradable stream going to the chemical and AD sections via the pulping process, and landfill is the result of the sum of the total cost of feedstock (−24.6 × 35.8 × 8000 Euro/y) (assuming there are 8000 operating hours in a year) and the total annual capital and operating cost of the mechanical separation and material recovery unit (3800000 Euro/y) and divided by the mass flowrate of the feedstock (35.8 × 8000 t/y): (−24.6 × 35.8 × 8000 + 3800000) / (35.8 × 8000) = −11.3 Euro/t. The product mass flowrates (t/h) from MRF are RDF (4), iron (1), aluminium (0.2), copper (0.1), biodegradable stream (23.4) and the flow going to landfill (7.2), respectively. All these outlet streams have a COP of −11.3 Euro/t. The negative sign implies no real cost and rather a guaranteed profit (even before consideration of the values of products) of the MRF system by the obtainment of gate fees.

The market prices of 75, 1093.8 and 3875 Euro/t of iron, aluminium and copper when applied give economic margins of 1 × (75 – (–11.3)) = 85.4, 0.2 × (1093.8 – (–11.3)) = 191.2 and 0.1 × (3875 – (–11.3)) = 288.2 Euro/h, respectively. The stream going to landfill thus incurs an economic loss: 7.2 × (−124 – (–11.3)) = −813.9 Euro/h.

As the pulping process incurs an annual total cost of 0.8 million Euro / y, the COP of its outlet streams, lignocellulosic and effluent streams to chemical and AD sections, respectively, is: ((−11.3) × 23.4 × 8000 + 800000) / (23.4 × 8000) = −7.3 Euro/t.

For the end products, the economic margins calculated using equation 1 are as follows.

1) Electricity from RDF in the total site CHP section: 4 × (8.2 – (–11.3)) = 77.2 Euro/h

   Electricity from biogas CHP: 3.9 × (8.2 – 259.5) = –983.1 Euro/h

   Electricity from char CHP: 8.1 × (8.2 – 262.3) = –2051.4 Euro/h
2) Metals from the MRF section: 85.4, 191.2 and 288.2 Euro/h, for iron, aluminium and copper, respectively
3) LA from the chemical conversion section: 2.3 × (4850 – 262.3) = 9741.6 Euro/h
4) Fertiliser from the AD section: AD system has two effluent streams as the feedstock, one from the pulping process (5.4 t/h) and the other is from the chemical conversion section (7.5 t/h). Thus, the COP of its overall feedstock is: $\frac{((-7.3)\times 5.4\times 8000+262.3\times 7.5\times 8000+11.3\times 1000000)}{(5.4+7.5)} = 259.5$ Euro/h. Thus, the fertiliser from the AD section: 9.1 × (4.7 – 259.5) = –2325.9 Euro/h (The yields of fertiliser and biogas products from the AD section are obtained from Sadhukhan, 2014)
5) Loss to landfill: 7.2 × (–124 – (–11.3)) = –813.9 Euro/h
6) Total economic margin of the MBCT system: 77.2 + (–983.1) + (–2051.4) + 85.4 + 191.2 + 288.2 + 9741.6 + (–2325.9) + (–813.9) = 4209.2 Euro/h

   This way, the overall economic margin of the plant can be calculated from the total marginal contributions of the outlet streams, which is also equal to the economic margin of MSW, i.e. 35.8 × (93− (−24.6)). The overall economic margin of the MBCT system is thus equal to 118 Euro/t MSW.

7) The analysis also gives an estimate of the payback time, as follows $\left(\frac{Total\ CAPEX\ in\ Table\ 4}{\sum_{all\ outlet\ streams} VOP \times mass\ flowrate - COP \times mass\ flowrate\ of\ MSW\ free\ of\ recyclables}\right)$:

$$\frac{(2.94 + 2.77 + 0.63 + 0.92 + 0.77 + 5.23 + 6.64 + 4 + 3.78 + 74.88 + 3.98 + 0.3 + 273.84) \times 10^6}{[4550 \times 2.3 + 3875 \times 0.1 + 1093.8 \times 0.2 + 75 \times 1 + 8.2 \times 4 + (-124) \times 7.2 + 8.2 \times 3.9 + 8.2 \times 8.1 - (-24.6) \times 35.8] \times 8000}$$

= 4.3 years

Considering 20 years of life time and 13% annual charge on 380.69 million Euro of total CAPEX of the MBCT system, the discounted cash flow (DCF) analysis gives 6.1% internal rate of return (IRR), shown as follows:

$$DCF\ between\ 0 - 20\ years = \frac{4209.2 \times 8000}{1000000 \times \left(1 + \frac{IRR}{100}\right)^{year}} \text{ million Euro/y}$$

$$Net\ present\ value\ (NPV) = \sum_{year=1}^{20} DCF$$

$$= \sum_{year=1}^{20} \frac{4209.2 \times 8000}{1000000 \times \left(1 + \frac{IRR}{100}\right)^{year}} = 380.69$$

Thus, $IRR = 6.1\%$

The streams with negative economic margins include electricity and fertiliser products and residues to the landfill. These must be reduced to zero or made positive by added value productions to enhance economic independence of waste treatment industry. The value analysis framework is dynamic, wherein uncertain parameters, e.g. prices, costs and flowrates and any other uncertain parameters can be varied to examine their sensitivity on the techno-economic feasibility. For example, if the gate fee is eliminated and MSW is priced at 50 Euro/t, the economic margin of the MBCT system is reduced from 118 Euro/t (Fig. 3) to 43 Euro/t. The capital investment of the highest capital intensive section in the MBCT system, i.e. chemical conversion section, if reduced to 50% and 25% of its present capital investment, which can be due to the learning curve effects (Sadhukhan, et al., 2014), the economic margin of the MBCT system increases to 180 and 210, Euro/t MSW, respectively. Similarly, if the capital investment of the second highest capital intensive section, i.e. AD section, is reduced to 50% and 25%, the economic margin of the MBCT system increases to 135 and 145, Euro/t MSW, respectively. Metal prices are also variables. For the two price ranges of metals noted in this study, i.e. 40-80, 250-1500 and 2600-3600 £/t of iron, aluminium and copper, respectively, the total economic margin of the MBCT system varies between 110-119 Euro/t MSW. The economic margin of the MBCT system is very sensitive to the value of the fertiliser product. If its VOP is equal to COP, i.e. 259.5 Euro/t, or it incurs zero economic margin, which is better than the loss it's presently incurring, the economic margin of the MBCT system increases to 183 Euro/t MSW. If the organic constituents in it can be recovered worthy 1 Euro/kg (VOP), the economic margin of the MBCT system increases to 371 Euro/t MSW. It is clear that the diversion of biodegradable waste currently exploited in making CLO, to valorisation into high value chemicals is the key to increased economic feasibility of the system. For long-term sustainability of the waste sector and creating a circular economy, it is imperative that the sector becomes dynamic and economically competitive by employing progressive upgrading strategies as shown in this paper. To date, MBT plants or MRF do not utilise organic fraction of MSW other than to produce CLO. Increasing product portfolio enhances economic competitiveness of the waste management sector. In this highly competitive market, business leadership requires breakthrough game changer technologies and financial value chain analysis, both of which are demonstrated here.

4. Conclusions

A novel concept of MBCT systems for extensive fractionation of urban waste into added value products has been introduced. The study demonstrated that integrated MBCT systems result in the highest economic margin attributed to the revenues from chemical, metal, energy and fertiliser, in decreasing values respectively. Resource recovery and product generation (without the inclusion of gate fees) is more than enough to outweigh waste collection fees, annual capital and operating costs. The work involves process design and integration using yield based and Aspen Plus$^{©}$ simulation models and techno-economic feasibility and sensitivity analyses through a powerful value analysis tool.

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